FIFTH EDITION

Wastewater Engineering

Treatment and Resource Recovery

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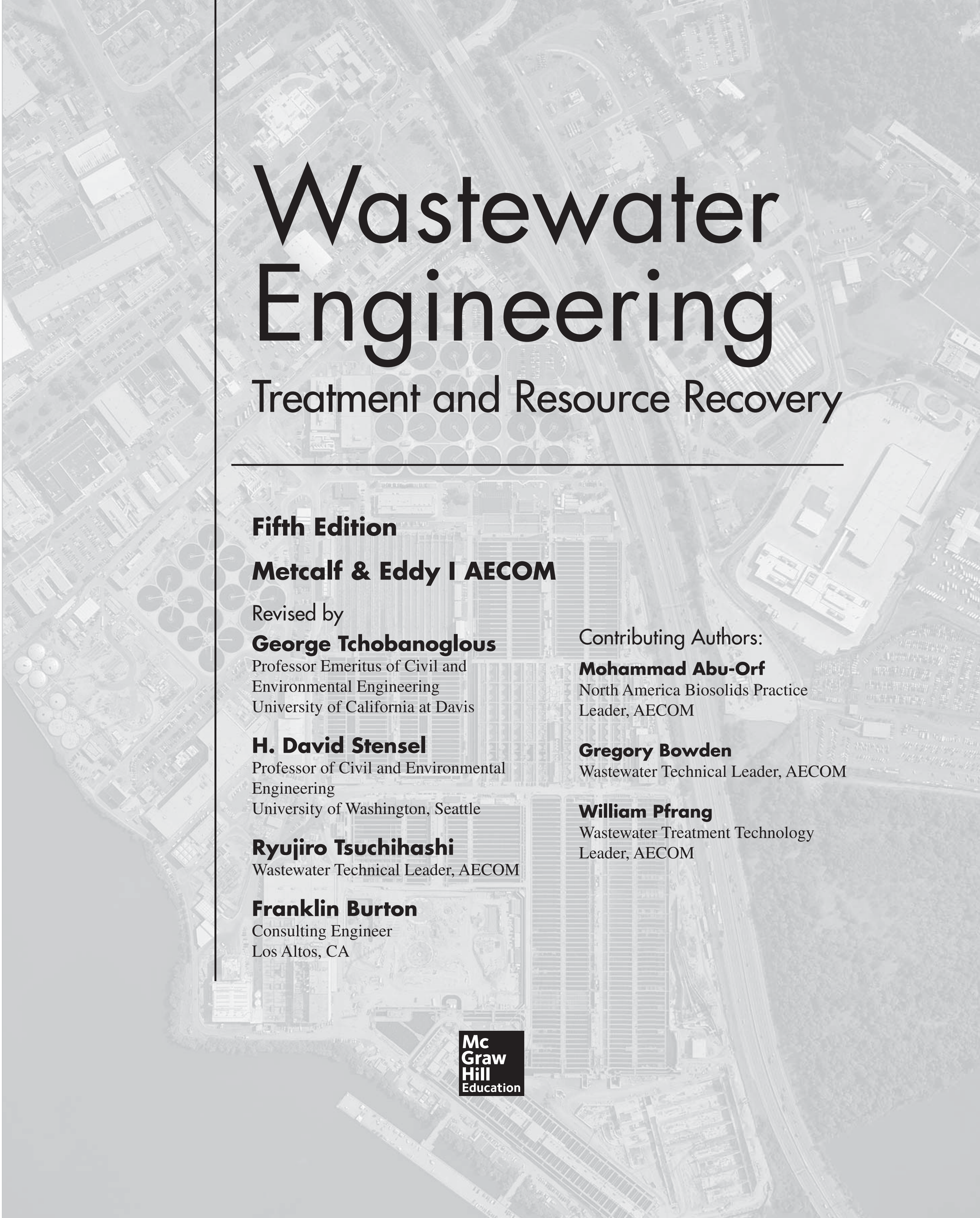

Wastewater Engineering

Treatment and Resource Recovery

Fifth Edition

Metcalf & Eddy I AECOM

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Preface

Since completion of the fourth edition of this textbook, the field of wastewater engineering has evolved at a rapid pace. Some of the more significant changes include:

1. A new view of wastewater as a source of energy, nutrients, and potable water.
2. More stringent discharge requirements related to nitrogen and phosphorus;
3. Enhanced understanding of the fundamental microbiology and physiology of the microorganisms responsible for the removal of nitrogen and phosphorus and other constituents;
4. An appreciation of the importance of the separate treatment of return flows with respect to meeting more stringent standards for nitrogen removal and opportunities for nutrient recovery,
5. Increased emphasis on the treatment of sludge and the management of biosolids; and
6. Increased awareness of carbon footprint impacts and greenhouse gas emissions, and an emphasis on the development of energy-neutral or energy-positive wastewater plants through more efficient use of chemical and heat energy in wastewater.

The 5th edition of this textbook has been prepared to address the significant changes cited above. Increased understanding of the importance of pre-treatment processes is addressed in Chap. 5. Advances in biological treatment are addressed in Chaps. 7 through 10. New developments in disinfection are considered in Chap. 12. The management of sludge and biosolids is now covered in Chaps. 13 and 14. Return flow treatment is considered in Chap. 15. Energy management is considered in Chap. 17. An emphasis of this fifth edition is to present practical design and operational data, while maintaining a solid theoretical discussion of the technologies and applications. Input from AECOM's process engineers and outside reviewers was sought to provide the user with a source of real-world practical information, the likes of which is not available in any single source.

IMPORTANT FEATURES OF THIS BOOK

In the 4th edition of this book, a separate chapter was devoted to the fundamentals of process analysis, including an introduction to the preparation of mass balances and reaction kinetics. Because introductory courses on process analysis and modeling are now taught at most colleges and universities, the material on the fundamentals of process analysis from the 4th edition has been condensed and is now included in Secs. 1–7 through 1–11 in Chap. 1. The material on process analysis has been retained as a reference source for students that have already had a separate course on modeling and as an introduction to the subject for students who may not have had an introductory course.

Following the practice in the 4th edition, more than 150 example problems have been worked out in detail to enhance the readers' understanding of the basic concepts presented in the text. To aid in the planning, analysis, and design of wastewater management systems, design data and information are summarized and presented in more than 400 tables, most of which are new. To illustrate the principles and facilities involved in the field of wastewater management, over 850 individual illustrations, graphs, diagrams, and photographs are included. An additional 120 drawings are included in tables. More than 375 homework problems and discussion topics are included to help the readers of this textbook hone their analytical skills and enhance their mastery of the material. Extensive references are also provided for each chapter.

The International System (SI) of Units is used in the 5th edition. The use of SI units is consistent with teaching practice in most US universities and in most countries throughout the world. In general, dual sets of units (i.e., SI and US customary) have been used for the data tables. Where the use of double units was not possible, conversion factors are included as a footnote to the table.

To further increase the utility of this textbook, several appendixes have been included. Conversion factors from International System (SI) of Units to US Customary Units and the reverse are presented in Appendixes A–1 and A–2, respectively. Conversion factors used commonly for the analysis and design of wastewater management systems are presented in Appendix A–3. Abbreviations for SI and US customary units are presented in Appendixes A–4 and A–5, respectively. Physical characteristics of air and selected gases and water are presented in Appendixes B and C, respectively. The statistical analysis is reviewed in Appendix D. Dissolved oxygen concentrations in water as a function of temperature are presented in Appendix E. Carbonate equilibrium is considered in Appendix F. Moody diagrams for the analysis of flow in pipes are presented in Appendix G. The analysis of nonideal flow in reactors is considered in Appendix H. Modeling nonideal flow in reactors is addressed in Appendix I.

USE OF THIS BOOK

Enough material is presented in this textbook to support a variety of courses for one or two semesters, or three quarters at either the undergraduate or graduate level. The book can be used both as a class textbook or class reference to supplement instructors' notes. The specific topics to be covered will depend on the time available and the course objectives. Suggested course outlines are presented below.

For a one semester introductory course on wastewater treatment, the following material is suggested.

Topic	Chapter	Sections
Introduction to wastewater treatment	1	1–1 to1–6
Wastewater characteristics	2	All
Wastewater flowrates and constituent loadings	3	All
Physical unit processes	5	5–1 to 5–8
Chemical unit processes	6	6–1 to 6–3
Introduction to biological treatment of wastewater	7	All
Disinfection	12	12–1 to 12–5, 12–9
Biosolids management	13, 14	All
Process selection, design, and implementation	4	All
Advanced treatment processes (optional)	6, 11	6–7, 6–8, 11–5 to 11–7

For a two semester course on wastewater treatment, the following material is suggested.

Topic	Chapter	Sections
Introduction to wastewater treatment	1	1–1 to1–6
Wastewater characteristics	2	All
Wastewater flowrates and constituent loadings	3	All
Process selection, design, and implementation	4	4–1 to 4–5
Physical unit operations	5	All
Chemical unit operations	6	All
Introduction to biological treatment of wastewater	7	All
Suspended growth biological treatment processes	8	All
Attached growth and combined biological treatment processes	9	9–1 to 9–5
Anaerobic treatment processes	10	10–1 to 10–5
Disinfection	12	All
Sludge Management	13	All
Biosolids management	14	All
Treatment of return flows	15	All

For a one semester course on biological wastewater treatment, the following material is suggested.

Topic	Chapter	Sections
Introduction to wastewater treatment	1	1–1 to1–6
Wastewater characteristics	2	All
Process selection, design, and implementation	4	4–2, 4–4, 4–5
Introduction to biological treatment of wastewater	7	7–1 to 7–8
Suspended growth processes	8	8–1 to 8–3
Attached growth biological treatment processes	9	All
Anaerobic treatment processes	10	10–1 to 10–5
Anaerobic sludge treatment	13	13–9, 13–10

For a one semester course on physical and chemical unit processes, the following material is suggested. It should be noted that material listed below could be supplemented with additional examples from water treatment.

Topic	Chapter	Sections
Process selection, design, and implementation	4	4–1 to 4–4
Introduction to physical unit processes		
Mixing and flocculation	5	5–3
Sedimentation	5	5–4, 5–6, 5–7,
Gas transfer	5	5–10, 5–11
Filtration (conventional depth filtration)	11	11–3, 11–4, 11–6
Membrane filtration	11	11–7
Adsorption	11	11–9
Gas stripping	11	11–10
UV disinfection	12	12–9
Introduction to chemical unit processes		6–2
Coagulation	6	6–2
Chemical precipitation	6	6–3, 6–4, 6–6
Ion exchange	11	11–11
Water stabilization	6	6–10
Chemical oxidation (conventional)	6	6–7
Advanced oxidation processes	6	6–8
Photolysis	6	6–9

Acknowledgments

This textbook is a tribute to the engineers and scientists who continue to push forward the practice and technologies of the wastewater industry. These advances continue to offer the world cleaner water resources and sustainable water supplies. The book could not have been written without the efforts of numerous individuals including the primary writers, contributing authors, individuals with specialized skills, technical reviewers, outside reviewers, and practitioners who contributed real life experiences.

Contributing authors from AECOM included: Dr. Mohammad Abu-Orf who revised and updated Chaps. 13 and 14, Dr. Gregory Bowden who wrote Chap. 15, and Mr. William Pfrang who revised and updated Chap. 5. Their assistance is acknowledged gratefully. Dr. Harold Leverenz of the University of California at Davis, is singled out for special acknowledgment for extraordinary contributions to the development of the graphics used throughout the text, the revision of Chap. 6, and individual section write ups. Others deserving special acknowledgment, in alphabetical order, are: Mr. Russel Adams an environmental consultant provided comprehensive reviews of Chaps. 3, 11, and 12; Dr. Heidi Gough of the University of Washington wrote the molecular biology section of Chap. 7; Dr. April Gu of Northeastern University who helped write and provided material for Chap. 9; Ms. Emily Legault of HDR Engineers provided thoughtful and comprehensive reviews of Chaps. 2, 3, 7, 8, 11, and 12; Mr. Mladen Novakovic of AECOM contributed to the development of Chap. 5; Mr. Terry Goss of AECOM contributed extensively to the development of Chaps. 13 and 14; and Mr. Dennis Totzke of Applied Technologies had significant involvement in the development of Chap. 10.

The review of the manuscript was critical to maintain the quality of the text. Outside reviewers, arranged alphabetically, who provided critical reviews included: Dr. Onder Caliskaner of Kennedy/Jenks Consultants reviewed portions of Chap. 11; Dr. Robert Cooper of BioVir laboratories reviewed the section on microbiology in Chap. 2; Ms. Libia Diaz of the University of California at Davis reviewed the homework problems; Dr. Robert Emerick of Stantec Engineers, reviewed the section on UV disinfection in Chap. 12; Dr. David Hokanson of Trussell Technologies reviewed portions of Chap. 11; Ms. Amelia Holmes of University of California at Davis reviewed the homework problems; Dr. Kurt Ohlinger of Sacramento Regional County Sanitation District provided review for phosphorus recovery. Dr. Edward Schroeder professor emeritus of the University of California at Davis reviewed portions of Chaps. 1 and 2.

A number of current and former AECOM engineers contributed to the development of the manuscript by providing design information and by reviewing specific portions of the text. Listed in alphabetical order they are:

- Mr. Michael Adkins
- Mr. David Ammerman
- Ms. Jane Atkinson
- Mr. Simon Baker
- Dr. William Barber
- Mr. David Bingham
- Mr. Joerg Blischke
- Mr. Gary Breitwisch
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- Mr. Nathan Cassity
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- Mr. William Clunie
- Dr. Patrick Coleman
- Mr. Nicholas Cooper
- Mr. Grant Davies
- Mr. Daniel Donahue
- Mr. Ralph Eschborn
- Mr. Bryce Figdore

Mr. Steven Freedman
Mr. Lee Glueckstein
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Mr. Thomas Weber
Mr. Simon Wills

Finally, the production of this textbook could not have been completed without the guidance and assistance of the following individuals. Mr. William Stenquist, Executive Editor and Ms. Lorraine Buczek, Development Editor of the McGraw Hill Book Company. Ms. Rose Kernan and Ms. Erin McConnell of RPK Editorial Services, Inc. provided service above and beyond in working with the authors to produce the textbook. The collective efforts of these individuals were invaluable and greatly appreciated.

George Tchobanoglous
H. David Stensel
Ryujiro Tsuchihashi
Franklin Burton

Foreword

One hundred years have passed since the three-volume "American Sewerage Practice" treatise was published in 1914–1915 by Leonard Metcalf and Harrison P. Eddy. The initial publication quickly became the standard of care and established the foundation for modern wastewater treatment. The original concept of combining theory with a strong compliment of practical data and design guidance continues on in the fifth edition. The wealth of practical information continues to be a cornerstone of Metcalf & Eddy publications, and has led to its reputation as the number one wastewater practice textbook. In this fifth edition over 150 example problems and over 375 homework problems are provided.

The textbook has become a widely used teaching resource for universities and colleges and a reference for engineering firms throughout the world and is now published in Chinese, Greek, Italian, Japanese, Korean, and Spanish.

New advances in technology continue to occur at a record pace in all fields including wastewater treatment. As a result this fifth edition includes numerous advances and represents the current state of the art information. AECOM takes great pride in presenting this Metcalf & Eddy textbook, a comprehensive compilation of the best wastewater practices in use today.

The manuscript was developed by a team of primary writers including Dr. George Tchobanoglous, Dr. H. David Stensel, Dr. Ryujiro Tsuchihashi, Dr. Mohammad Abu-Orf, Mr. William Pfrang and Dr. Gregory Bowden. In addition to our primary authors, over 55 AECOM employees and outside technical specialists contributed in reviews and provided practical data and guidance.

I would also like to acknowledge Mr. Bill Stenquist, Executive Editor, McGraw-Hill, who was instrumental in bringing the resources of McGraw-Hill to this project.

The fifth edition textbook could not have been developed without the enthusiastic support of AECOM. I thank Mr. John M. Dionisio, Chairman and Chief Executive Officer, Mr. Robert Andrews, Chief Executive, Water, and Mr. James T. Kunz, Senior Vice President—Program Director.

Jekabs P. Vittands
Senior Vice President
AECOM

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1

Introduction to Wastewater Treatment and Process Analysis

WORKING TERMINOLOGY

Term	**Definition**
Batch reactor	A vessel in which flow is neither entering nor leaving during the reaction time.
Biosolids	Sludge from wastewater treatment processes that has been stabilized to meet the criteria in the U.S. EPA's 40 CFR 503 regulations and, therefore, can be used beneficially.
Complete-mix reactor (CMR)	A reactor in which complete mixing occurs instantaneously and uniformly throughout the reactor as fluid particles enter the reactor.
Characteristics, wastewater	General classes of wastewater constituents such as physical, chemical, and biological constituents.
Homogeneous reaction	Reactions that occur uniformly throughout the fluid so that the potential for reaction at any point within the fluid is the same.
Heterogeneous reaction	Reactions that occur between one or more constituents that can be identified with specific sites.
Ideal flow	A flow regime in which all fluid particles are retained in a reactor for a time period equivalent to the theoretical detention time.
Mass-balance analysis	An accounting of mass within a defined boundary before and after reactions and conversions have taken place.
Molecular diffusion, coefficient of	The movement of molecules from a region of higher concentration to a region of lower concentration.
Nonideal flow	A flow regime in which a portion of the fluid particles are held in a reactor for a time other than the theoretical detention time.
Plug-flow reactor (PFR)	A vessel intended to transport fluid particles such that they leave in the same sequence as they entered.
Reaction rate	The rate of change (decrease or increase) in the number of moles of a reactive substance per unit volume per unit time (for homogeneous reactions), or per unit surface area or mass per unit time (for heterogeneous reactions).

Term	Definition
Reaction order	An empirical coefficient used to describe the relationship between reaction rate and concentration for a compound in a given reaction.
Reactor	A vessel or tank in which physical, chemical, or biological reactions occur.
Sludge	Any material (i.e., sludge) produced during primary, secondary, or advanced wastewater treatment that has not undergone any process to reduce pathogens or vector attraction (see also Biosolids).
Steady-state	The condition that occurs when the concentration of a constituent does not change with time.
Temperature coefficient	A coefficient used to correct reaction rate constants for the effect of temperature.
Stoichiometry	The proportions in which chemical elements combine or are produced and the weight relations in any chemical reaction.
Stormwater	Part of the precipitation that appears in surface streams. It is the same as streamflow unaffected by artificial diversions, storage, or other works of man in or on the stream channels.
Wastewater	Used water discharged from homes, businesses, cities, industry, and agriculture. Various synonymous uses such as municipal wastewater (sewage), industrial wastewater, and stormwater.
Wastewater treatment	The removal of constituents so the treated effluent can be returned to the environment or reused safely.
Water reuse	The use of treated wastewater for a beneficial use such as agricultural and landscape irrigation and for indirect and direct potable use.
Unit process	A subset of a complete treatment process in which constituent conversion or removal is accomplished by physical forces (such as gravity sedimentation) or by chemical or biological reactions.

Wastewater is essentially the water supply of the community after it has been used in a variety of applications and which now contains constituents that render it unsuitable for most uses without treatment. When untreated wastewater accumulates and is allowed to go septic, the decomposition of the organic matter it contains will lead to nuisance conditions, including the production of malodorous gases. In addition, untreated wastewater contains numerous pathogenic microorganisms that dwell in the human intestinal tract. Wastewater also contains nutrients, which can stimulate the growth of aquatic plants, and may contain toxic compounds or compounds that potentially may be mutagenic or carcinogenic. For these reasons, the immediate and nuisance-free removal of wastewater from its sources of generation (see Fig. 1–1), followed by treatment, reuse, or dispersal (disposal) into the environment, is necessary to protect public health and the environment.

Wastewater engineering is that branch of environmental engineering in which the basic principles of science and engineering are applied to solving the issues associated with the collection, treatment, and reuse of wastewater. The ultimate goal of wastewater engineering is the protection of public and environmental health in a manner commensurate with economic, social, and political concerns. The objective of this chapter is twofold. The first objective is to introduce and briefly discuss (1) the evolution of wastewater treatment, (2) the evolution of regulations pertaining to wastewater management, (3) the general characteristics of wastewater, (4) the classification of wastewater treatment methods, (5) the application of treatment methods, and (6) the status of wastewater

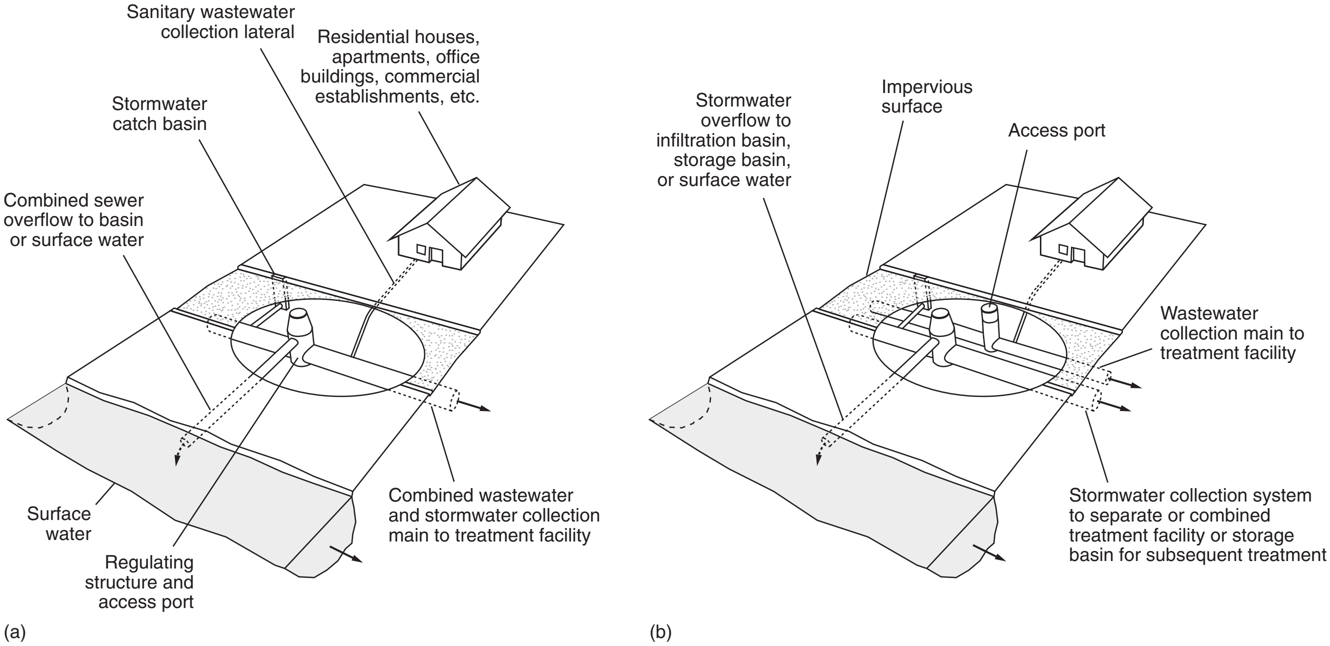


Figure 1–1

Schematic diagram of wastewater collection system infrastructure: (a) combined (wastewater and stormwater) collection system and (b) separate wastewater and stormwater collection systems. (Courtesy of H. Leverenz.)

treatment in the United States. The material presented in the first five sections of this chapter is intended to serve as an introduction to the subject of wastewater treatment and to provide a basis for the analysis of the unit processes that will be presented in subsequent chapters.

The second objective of this chapter is to present a review of and introduction to (1) the fundamentals of process analysis, (2) the reactors used in wastewater treatment, (3) the modeling of ideal flow reactors, (4) process kinetics, and (5) treatment process modeling. The material presented in the last five sections of this chapter is meant to serve as a review for those readers who have already taken courses in process analysis and to provide an introduction to the subject for those who are new to process analysis. By dealing with the basic concepts (e.g., mass balance and reactor analysis) first, it will be possible to apply them (without repeating the details) in the remaining chapters.

1–1 EVOLUTION OF WASTEWATER TREATMENT

The evolution of the field of wastewater treatment is the story of health and environmental concerns, especially as the size of cities increased. Methods of wastewater treatment were first developed in response to the concern for public health and the adverse conditions caused by the discharge of wastewater to the environment. Also important, as cities became larger, the land required for wastewater treatment and disposal, principally by irrigation and intermittent filtration (methods used commonly in the early 1900s) was no longer available. Thus, it became necessary to develop other methods of treatment that could be used to accelerate the forces of nature, under controlled conditions, in engineered treatment facilities that required less land (i.e., a smaller footprint).

Treatment Objectives

In general, from about 1900 to the early 1970s, treatment objectives were concerned with (1) the removal of suspended and floatable material, (2) the treatment of biodegradable organics, and (3) the elimination of pathogenic organisms. Unfortunately, these objectives were not uniformly met throughout the United States, as is evidenced by the many plants that were discharging partially treated wastewater well into the 1980s.

Since 1980, the water-quality improvement objectives of the 1970s have continued, but the emphasis has shifted to the definition and removal of constituents that may cause long-term health effects and environmental impacts. Consequently, while the early treatment objectives remain valid today, the required degree of treatment has increased significantly, and additional treatment objectives and goals have been added. Therefore, treatment process design must go hand-in-hand with the water quality objectives or standards established by the federal, state, and regional regulatory authorities.

Current Health and Environmental Concerns

As research into the characteristics of wastewater has become more extensive and as the techniques for analyzing specific constituents and their potential health and environmental effects have become more comprehensive, the body of scientific knowledge has expanded significantly. Many of the new treatment methods being developed are designed to deal with health and environmental concerns associated with constituents found as a result of advanced detection methods. However, the advancement in treatment technology effectiveness has not kept pace with the enhanced constituent detection capability. Constituents can be detected at lower concentrations than can be attained by available treatment technology. Therefore, careful assessment of health and environmental effects and community concerns has become increasingly important in wastewater management. The need to establish a dialog with the community is important to assure that health and environmental issues are being addressed.

Sustainability Considerations

The need to be more efficient with the use of resources and the dispersal of anthropogenic constituents in the environment has become a central issue in nearly all aspects of society. Some important considerations include the rapid extraction of remaining fossil carbon supplies and subsequent transfer of this carbon to the atmosphere, the consumption of nonrenewable resources, and the release of gases to the atmosphere, which may have an influence on climate. Some notable examples of problematic current and past practices with respect to wastewater treatment include the discharge of nutrients and trace constituents, excessive headloss and pumping as a result of poor hydraulic design, inefficient aeration system design, lack of consideration for the importance of primary treatment systems, limited use of anaerobic processes for BOD removal and energy recovery, limitations in sludge reuse and ultimate disposal options, placement of wastewater treatment facilities without regard to water reuse, life cycle pumping energy implications, and the potential impacts of sea level rise.

Sustainability issues such as the overall energy balance, process-related greenhouse gas emissions, total chemical usage and carbon footprint associated with those chemicals, and the fate of the constituents present in effluent and process byproducts are now of critical importance in the design, construction, and operation of wastewater management facilities. A number of developments have taken place recently that have made the prospect of sustainability a reality, including the development of new or improved technologies for the recovery of heat and chemical energy from wastewater, advanced processes capable of full-scale water purification, new technologies for satellite and decentralized wastewater management infrastructure, and process models that also compute resource consumption and greenhouse gas emissions. The refinement and optimization of existing processes, along with the implementation of new approaches and technologies, will create the foundation for a sustainable approach to wastewater management.

1–2 EVOLUTION OF REGULATIONS OF SIGNIFICANCE TO WASTEWATER ENGINEERING

The establishment of the U.S. Environmental Protection Agency (U.S. EPA) and the enactment of federal regulations have brought about significant changes in the planning and design of wastewater treatment facilities in the United States. The creation of the U.S. EPA and the important regulations that govern the management of wastewater are considered briefly in the following text.

Establishment of Environmental Protection Agency

The U.S. EPA was established on December 2, 1970. The objective was to consolidate in one agency a variety of federal research, monitoring, standard-setting, and enforcement activities to ensure environmental protection. Although examples of pollution were visible for decades prior to 1970, the seminal event in coalescing concerns about the environment and public opinion was the publication of *Silent Spring* by Rachael Carson in 1962. The book appeared first in serial form in *The New Yorker* in June 1962 and in book form later that year. The impacts resulting from the indiscriminate use of pesticides as documented by Carson helped rally public support for the federal government to take action. The pesticide DDT was banned in 1972.

Important Federal Regulations

Following the formation of the U.S. EPA, a number of enabling federal regulations brought about changes in planning and design of wastewater treatment facilities in the United States. The principal regulations are summarized in Table 1–1 and discussed below.

Public Law 92-500. Clearly, the most significant event in the field of wastewater management in the last 100 y was the passage of the Federal Water Pollution Control Act Amendments of 1972 (Public Law 92-500), often referred to as the Clean Water Act (CWA). Before that date, there were no specific national water pollution control goals or objectives. The CWA not only established national goals and objectives—“to restore and maintain the chemical, physical, and biological integrity of the Nation’s waters”—but also marked a change in water pollution control philosophy. No longer was the classification of the receiving stream of ultimate importance, as it had been before.

National Pollution Elimination Discharge System (NPDES). The CWA law decreed that the quality of the nation’s waters was to be improved by the imposition of specific effluent limitations. A National Pollution Elimination Discharge System (NPDES) program was established based on uniform technological minimums with which each point source discharger had to comply. It is interesting to note that the Clean Air Acts of 1970 and 1990 have also had a significant impact on industrial and municipal wastewater programs, primarily through the implementation of treatment facilities for the control of emissions.

Table 1–1

Summary of significant U.S. Federal regulations that affect wastewater management

Regulation	Description
The Federal Water Pollution Control Act (FWPCA) (P.L. 80-845, 62 Stat. 1155) of 1948	The first major law enacted by Congress to address the problems of water pollution in the United States.
Clean Water Act (CWA)(P.L. 92-500, Federal Water Pollution Control Act Amendments of 1972)	Establishes the National Pollution Discharge Elimination System (NPDES), a permitting program based on uniform technological minimum standards for each discharger.
Water Quality Act of 1987 (WQA) (Amendment of the CWA)	Strengthens federal water quality regulations by providing changes in permitting and adds substantial penalties for permit violations. Amends solids control program by emphasizing identification and regulation of toxic pollutants in sewage sludge.
40 CFR Part 503 (1993) (Sewage Sludge Regulations)	Regulates the use and disposal of biosolids from wastewater treatment plants. Limitations are established for items such as contaminants (mainly metals), pathogen content, and vector attraction.
National Combined Sewer Overflow (CSO) Policy (1994)	Coordinates planning, selection, design and implementation of CSO management practices and controls to meet requirements of CWA. Nine minimum controls and development of long term CSO control plans are required to be implemented immediately .
Clean Air Act of 1970 and 1990 Amendments	Establishes limitations for specific air pollutants and institutes prevention of significant deterioration in air quality. Maximum achievable control technology is required for any of 189 listed chemicals from "major sources," i.e., plants emitting at least 60 kg/d.
40 CFR Part 60	Establishes air emission limits for sludge incinerators with capacities larger than 1000 kg/d (2200 lb/d) dry basis.
Total Maximum Daily Load (TMDL) (2000) Section 303(d) of the CWA	Requires states to develop prioritized lists of polluted or threatened waterbodies and to establish the maximum amount of pollutant (TMDL) that a waterbody can receive and still meet water quality standards (U.S. EPA, 2000).

Standards for Secondary Treatment. Pursuant to Section 304(d) of Public Law 92-500 (see Table 1–1), the U.S. EPA published its definition of minimum standards for secondary treatment. The definition of secondary treatment is reported in Table 1–2 and includes three major effluent parameters: biodegradable organics, expressed in terms of five-day biochemical oxygen demand (5-d BOD); total suspended solids (TSS); and the hydrogen ion concentration expressed as pH. These standards provided the basis for the design and operation of most wastewater treatment plants. The secondary treatment regulations were amended further in 1989 to clarify the percent removal requirements during dry periods for treatment facilities served by combined sewers.

Water Quality Act of 1987. In 1987, Congress enacted the Water Quality Act of 1987 (WQA), the first major revision of the CWA. Important provisions of the WQA were (1) the strengthening federal water quality regulations by providing changes in permitting and adding substantial penalties for permit violations, (2) significantly amending the CWA's formal solids control program by emphasizing the identification and regulation of toxic pollutants in the solids (sludge), (3) providing funding for state and U.S. EPA studies for defining nonpoint and toxic sources of pollution, (4) establishing new deadlines for compliance including priorities and permit requirements for stormwater, and (5) phasing out the construction grants program as a method of financing publicly owned treatment works (POTWs).

Table 1–2

Minimum national standards for secondary treatment[a, b]

Characteristic of discharge	Unit of measurement	Average 30-d concentration[c]	Average 7-d concentration[c]
BOD_5	mg/L	30[d]	45
Total suspended solids	mg/L	30[d]	45
Hydrogen-ion concentration	pH units	Within the range of 6.0 to 9.0 at all times[e]	
$CBOD_5$[f,g]	mg/L	25	40

[a] Federal Register (1988, 1989).

[b] Present standards allow stabilization ponds and trickling filters to have higher 30-d average concentrations (45 mg/L) and 7-d average concentrations (65 mg/L) BOD/suspended solids performance levels as long as the water quality of the receiving water is not adversely affected. Exceptions are also permitted for combined sewers, certain industrial categories, and less concentrated wastewater from separate sewers. For precise requirements of exceptions, Federal Register (1988) should be consulted.

[c] Not to be exceeded.

[d] Average removal shall not be less than 85 percent.

[e] Only enforced if caused by industrial wastewater or by in-plant inorganic chemical addition.

[f] Carbonaceous 5-d biochemical oxygen demand.

[g] May be substituted for BOD_5 at the option of the permitting authority.

Biosolids Regulations. Regulations that affect wastewater facilities design include those for the treatment, disposal, and beneficial use of biosolids (40 CFR Part 503). In the biosolids regulation promulgated in 1993, national standards were set for pathogen and heavy metal content and for the safe handling and use of biosolids. The standards are designed to protect human health and the environment where biosolids are applied beneficially to land. The rule also promotes the development of a "clean sludge" (U.S. EPA, 1992). The regulations describing the reuse and disposal of solids are found in Chap. 13.

Total Maximum Daily Load. The total maximum daily load (TMDL) program was promulgated in 2000 but did not take effect until 2002. The TMDL rule is designed to protect ambient water quality. A TMDL represents the maximum amount of a pollutant that a water body can receive and still meet water quality standards. A TMDL is the sum of (1) the individual waste load allocations for point sources, (2) load allocations for nonpoint sources, (3) natural background levels, and (4) a margin of safety (U.S. EPA, 2000). To implement the rule, a comprehensive watershed-based water quality management program must be undertaken to find and control nonpoint sources in addition to conventional point source discharges. With implementation of the TMDL rule, the focus on water quality shifts from technology-based controls to preservation of ambient water quality. The end result is an integrated planning approach that transcends jurisdictional boundaries and forces different sectors, such as agriculture, water, and wastewater utilities and urban runoff managers, to cooperate. Implementation of the TMDL rule will vary depending on the specific water quality objectives established for of each watershed and, in some cases, will require the installation of advanced levels of treatment.

Air Emissions. Wastewater treatment facilities are potential sources of odors and other air emissions. Volatile organic compounds (VOCs) are of a particular concern because many of them are believed to be carcinogenic. Regulations have been developed for hazardous air pollutants (HAPs), a set of 188 chemicals regulated specifically by the U.S. EPA that are known or thought to cause human health effects in excess of specified levels. The regulations are defined in 40 CFR61.

Other Federal Regulations

In addition to the requirements established under the 1987 Water Quality Act and enforced by the U.S. EPA, other federal, state, and local regulations have to be considered in the planning, design, construction, and operation of wastewater treatment plants. Significant federal regulations include those prescribed by the Occupational Safety and Health Act (OSHA), which deals with safety provisions to be included in the facility's design. State, regional, and local regulations may include water quality standards for the protection of public health and the beneficial uses of the receiving waters, air quality standards for the regulation of air emissions (including odor) from treatment facilities, and regulations for the disposal and reuse of solids and biosolids.

State and Regional Regulations

Many of the state and regional regulatory agencies not only establish the permit requirements for wastewater discharges but also issue design guidelines for specific processes. Well-known design standards include the so-called "Ten States Standards" published by the Great Lakes-Upper Mississippi River Board of State Sanitary Engineers (2004) and the "Manual TR-16, Guides for the Design of Wastewater Treatment Works" published by the New England Interstate Water Pollution Control Commission (1998). For each state, the requirements of the regulatory agencies, including those having jurisdiction for public health, air quality, and solid waste management, have to be investigated carefully.

1–3 CHARACTERISTICS OF WASTEWATER

An understanding of the nature of wastewater is essential in the design and operation of collection, treatment, and reuse facilities and in the management of environmental quality. The sources of wastewater and the constituents and their sources that are found in wastewater are introduced in this section. These subjects are discussed in greater detail in Chaps. 2 and 3.

Sources of Wastewater

As received at the wastewater treatment facility, the characteristics of wastewater from a community depend on the components that make up the wastewater flow and changes occurring in the collection system used. Common sources of wastewater may include:

Domestic wastewater. Wastewater discharged from residences and from commercial, institutional, and public facilities. Domestic wastewater is also known as *sanitary wastewater.*

Industrial wastewater. Wastewater in which industrial wastes predominate.

Infiltration/inflow (I/I). Water that enters the collection system through indirect and direct means. Infiltration is extraneous water that enters the collection system through leaking joints, cracks and breaks. Inflow is stormwater that enters the collection system from storm drain connections (catch basins), roof leaders, foundation and basement drains, or through access port (manhole) covers.

Stormwater. Runoff resulting from rainfall and snowmelt.

Types of Collection Systems

Three types of collection systems are used for the removal of wastewater and stormwater: (1) sanitary wastewater collection systems, (2) combined wastewater and stormwater collection systems, and (3) stormwater collection systems (see Fig. 1–1). Where separate sanitary wastewater collection systems are used, the wastewater flow is composed of domestic wastewater, industrial wastewater, and infiltration/inflow. Where a combined collection system is used, the wastewater flow is composed of domestic wastewater, industrial wastewater, and infiltration/inflow plus stormwater and the constituents carried in the stormwater. In both cases, the percentage of the wastewater components will vary with local conditions and the time of the year. Where separate stormwater collection systems are used, it has been found that varying amounts of wastewater may be present in the stormwater, depending on local conditions.

Wastewater Constituents

The constituents in wastewater can be characterized in terms of their physical, chemical, and biological properties. The principal constituents of wastewater, and their sources, are reported in Table 1–3. It should be noted that many of the physical properties and chemical and biological constituents listed in Table 1–3 are interrelated. For example, temperature, a physical property, affects both the amounts of gases dissolved in the wastewater as well as the biological activity in the wastewater.

Constituents of Concern. Although the list of constituents presented in Table 1–3 is extensive, secondary treatment standards for wastewater, as reported in Table 1–2, are based on the removal of biodegradable organics, total suspended solids, and pH. More stringent standards may apply, depending on the use that will be made of the treated wastewater. For example, when wastewater is to be reused as a potable water supply, it may be necessary to remove nutrients, heavy metals, pathogens, and priority pollutants to significantly lower levels than would normally be required. Depending on local conditions, it may even be necessary to remove dissolved inorganic solids.

Recoverable Resources in Wastewater. As noted previously, the focus of the developments in wastewater treatment in the twentieth century was on the removal of constituents considered to be contaminants when discharged to the environment. In the twenty first century, there has been a paradigm shift in the view of wastewater. Wastewater is now considered a *renewable recoverable source of energy, resources, and water* (Tchobanoglous, 2011). In this new view of wastewater, the focus of wastewater treatment is changing. New processes are being developed that will make it possible to recover energy, resources, and water more effectively. In fact, it is reasonable to assume that in the not-so-distant future, wastewater treatment plants could become net exporters as opposed to consumers of energy. The heat and chemical energy in wastewater are considered in Chap. 2.

1–4 CLASSIFICATION OF WASTEWATER TREATMENT METHODS

The constituents found in wastewater are removed by physical, chemical, and biological means. The individual methods usually are classified as physical, chemical, and biological unit processes. Although these processes occur in a variety of combinations in treatment systems, it has been found advantageous to study their scientific basis separately because the principles involved do not change.

Physical Unit Processes

Treatment methods in which the application of physical forces predominate are known as physical unit processes. Because most of these methods evolved directly from man's

Table 1–3

Physical, chemical, and biological characteristics of wastewater and their sources

Characteristic	Sources
Physical properties:	
Color	Domestic and industrial wastes, natural decay of organic materials
Odor	Decomposing wastewater, industrial wastes
Solids	Domestic water supply, domestic and industrial wastes, soil erosion, inflow/infiltration
Temperature	Domestic and industrial wastes
Chemical constituents	
Organic	
Carbohydrates	Domestic, commercial, and industrial wastes
Fats, oils, and grease	Domestic, commercial, and industrial wastes
Pesticides	Agricultural wastes
Phenols	Industrial wastes
Proteins	Domestic, commercial, and industrial wastes
Priority pollutants	Domestic, commercial, and industrial wastes
Surfactant	Domestic, commercial, and industrial wastes
Volatile organic compounds	Domestic, commercial, and industrial wastes
Other	Natural decay of organic materials
Inorganic:	
Alkalinity	Domestic wastes, domestic water supply, groundwater infiltration
Chlorides	Domestic wastes, domestic water supply, groundwater infiltration
Heavy metals	Industrial wastes
Nitrogen	Domestic and agricultural wastes
Potassium	Domestic, commercial, and industrial wastes
pH	Domestic wastes, domestic water supply, groundwater infiltration
Phosphorus	Domestic, commercial, and industrial wastes; natural runoff
Priority pollutants	Domestic, commercial, and industrial wastes
Sulfur	Domestic water supply; domestic, commercial, and industrial wastes
Gases	
Hydrogen sulfide	Decomposition of domestic wastes
Methane	Decomposition of domestic wastes
Oxygen	Domestic water supply, surface-water infiltration
Biological constitutes	
Animals	Open watercourses and treatment plants .
Helminths (worms)	Domestic wastes
Plants	Open watercourses and treatment plants
Protists	
Eubacteria	Domestic wastes, surface-water infiltration, treatment plants
Archaebacteria	Domestic wastes, surface-water infiltration, treatment plants
Viruses	Domestic wastes

direct observations of nature, they were the first to be used for wastewater treatment. Screening, mixing, flocculation, sedimentation, flotation, filtration, and adsorption are typical physical unit process. For example, adsorption involves the removal of specific compounds from the wastewater on solid surfaces using the forces of attraction between bodies.

Chemical Unit Processes

Treatment methods in which the removal of constituents is brought about by the addition of chemicals or by other chemical reactions are known as chemical unit processes. Precipitation, gas transfer, adsorption, and disinfection are the most common examples used in wastewater treatment. In chemical precipitation, treatment is accomplished by producing a chemical precipitate that can be removed by settling, filtration, or membrane processes. In most cases, the separated precipitate will contain both the constituents that may have reacted with the added chemicals and the constituents that were swept out of the wastewater as the precipitate settled. The addition of oxygen to water to support aerobic reactions is the most common example of gas transfer. Another common chemical unit process is the use of chlorine for wastewater disinfection, which has been practiced for more than a century.

Biological Unit Processes

Treatment methods in which the removal of constituents is brought about by biological activity are known as biological unit processes. Biological treatment is used primarily to remove the colloidal or dissolved biodegradable organic substances found in wastewater. Basically, these substances are converted into (a) gases that can escape to the atmosphere and (b) biological cell tissue that can be removed by settling or another solid's separation process. Biological treatment is also used to remove the nitrogen and phosphorus from wastewater. With proper environmental control, wastewater can be treated biologically in most cases. Therefore, it is the responsibility of the engineer to ensure that the proper environment is produced and controlled effectively to achieve all treatment objectives.

1–5 APPLICATION OF TREATMENT METHODS

The principal methods now used for the treatment of wastewater and process residuals are identified in this section. Detailed descriptions of each method are not presented because the purpose here is only to introduce the many different ways in which treatment can be accomplished. The detailed descriptions are presented throughout the remainder of this book.

Wastewater Processing

To achieve the removal of constituents, a number of unit processes are grouped together to provide what is known as primary, secondary, tertiary, and advanced treatment (see Table 1–4). In general, the term *primary* refers to application of physical unit processes; *secondary* refers to chemical and biological unit processes; and *tertiary* refers to combinations of all three. It should be noted that these terms are arbitrary and in most cases of little value. A more rational approach is first to establish the degree of constituent removal (treatment) required before the wastewater can be reused or dispersed in the environment. The unit processes necessary to achieve the required degree of treatment can then be grouped together on the basis of fundamental considerations. The constituents of major

Table 1–4

Levels of wastewater treatment

Treatment level	Description
Preliminary	Removal of wastewater constituents such as rags, sticks, floatables, grit, and grease that may cause maintenance or operational problems with the treatment operations, processes, and ancillary systems.
Primary	Removal of a portion of the suspended solids and organic matter from the wastewater.
Advanced primary	Enhanced removal of suspended solids and organic matter from the wastewater. Typically accomplished by chemical addition or filtration.
Secondary	Removal of biodegradable organic matter (in solution or suspension) and suspended solids. Disinfection is also typically included in the definition of conventional secondary treatment.
Secondary with nutrient removal	Removal of biodegradable organics, suspended solids, and nutrients (nitrogen, phosphorus, or both nitrogen and phosphorus).
Tertiary	Removal of residual suspended solids (after secondary treatment), usually by granular medium filters, cloth filters, or microscreens. Disinfection is also typically a part of tertiary treatment. Nutrient removal is often included in this definition.
Advanced	Removal of dissolved and suspended materials remaining after normal biological treatment when required for various water reuse applications.

interest in wastewater and the unit processes or methods applicable to the removal of these constituents are shown in Table 1–5.

Secondary treatment, as defined by the U.S. EPA, is directed principally toward the removal of readily biodegradable organics and suspended solids. To further protect the environment in some critical areas, more stringent standards have been directed toward the removal of nutrients and pathogens and achieving lower levels of oxygen demand than are now possible with secondary treatment techniques. When wastewaters are to be reused, standards may include removal requirements for refractory organics, heavy metals, and dissolved inorganic solids. In general, the complexity of the treatment-process flow diagram will depend on which constituents need to be removed and the required levels of removal.

Residuals Processing

For the most part, the methods and systems reported in Table 1–5 are used to treat the liquid portion of the wastewater. Of equal if not more importance in the overall design of treatment facilities are the corresponding processes or systems used to treat the sludge removed from the liquid portion of the wastewater. The principal methods now in use are reported in Table 1–6.

Typical Treatment Process Flow Diagrams

When treatment processes are grouped together to achieve a specific treatment objective, the grouping is identified as a *treatment process flow diagram* or as a *treatment train*. Representative treatment process flow diagrams are presented on Fig. 1–2.

Conventional Secondary Treatment. A typical flow diagram for secondary treatment is shown on Fig. 1–2(a). As noted above, conventional secondary treatment is used primarily for the removal of BOD_5 and TSS. The primary treatment portion of the

(*continued*)

Table 1–5

Unit processes used to remove constituents of concern found in wastewater

Constituent	Unit process	See Chap.
Suspended solids	Screening	5
	Grit removal	5
	Sedimentation and high-rate clarification	5
	High-rate clarification	5
	Flotation	5
	Chemical precipitation with settling, flotation or filtration	6
	Depth filtration	11
	Surface filtration	11
	Membrane filtration	11
Biodegradable organics	Aerobic suspended growth variations	8
	Aerobic attached growth variations	9
	Anaerobic suspended growth variations	10, 13
	Anaerobic attached growth variations	10
	Physical-chemical systems	6, 11
	Chemical oxidation	6
	Advanced oxidation	6
	Membrane filtration	11
Nutrients		
Nitrogen	Chemical oxidation (breakpoint chlorination)	12
	Suspended-growth nitrification and denitrification variations	8
	Fixed-film nitrification and denitrification variations	9
	Air stripping	11, 15
	Ion exchange	11
Phosphorus	Chemical precipitation	6
	Biological phosphorus removal	8, 9
Nitrogen and phosphorus	Biological nutrient removal variations	8, 9
Pathogens	Chlorine compounds	12
	Chlorine dioxide	12
	Ozone	12
	Ultraviolet (UV) radiation	12
	Heat treatment (Pasteurization)	12
Colloidal and dissolved solids	Membranes	11
	Chemical treatment	6, 11
	Carbon adsorption	11
	Ion exchange	11
Volatile organic compounds	Air stripping	11, 15
	Carbon adsorption	11
	Advanced oxidation	6
Odors	Chemical scrubbers	16
	Carbon adsorption	11, 16
	Bio-trickling filters	16
	Compost filters	16

Table 1–6

Residuals processing and disposal methods

Processing or disposal process	Unit process or treatment method	See Chap.
Preliminary operations	Sludge pumping	13
	Sludge grinding	13
	Sludge blending and storage	13
	Sludge degritting	13
Thickening	Gravity thickening	13
	Flotation thickening	13
	Centrifugation	13
	Gravity belt thickening	13
	Rotary drum thickening	13
Stabilization	Lime stabilization	13
	Heat treatment	13
	Anaerobic digestion	13
	Aerobic digestion	13
	Composting	14
Conditioning	Chemical conditioning	13
	Heat treatment	13
Disinfection	Pasteurization	13
	Long term storage	14
Dewatering	Centrifuge	14
	Belt press filter	14
	Rotary press	14
	Screw press	14
	Filter presses	14
	Elecro-dewatering	14
	Sludge drying beds	14
	Reed beds	14
	Lagoons	14
Heat drying	Dryer variations	14
Thermal reduction	Multiple hearth incineration	14
	Fluidized bed incineration	14
	Co-incineration with solid wastes	14
Resource recovery	Nutrients	15
Energy recovery	Anaerobic digestion	13
	Thermal oxidation	14, 17
	Production of oil and liquid fuels	14, 17
Ultimate disposal	Land application	14
	Landfill	14
	Lagooning	14

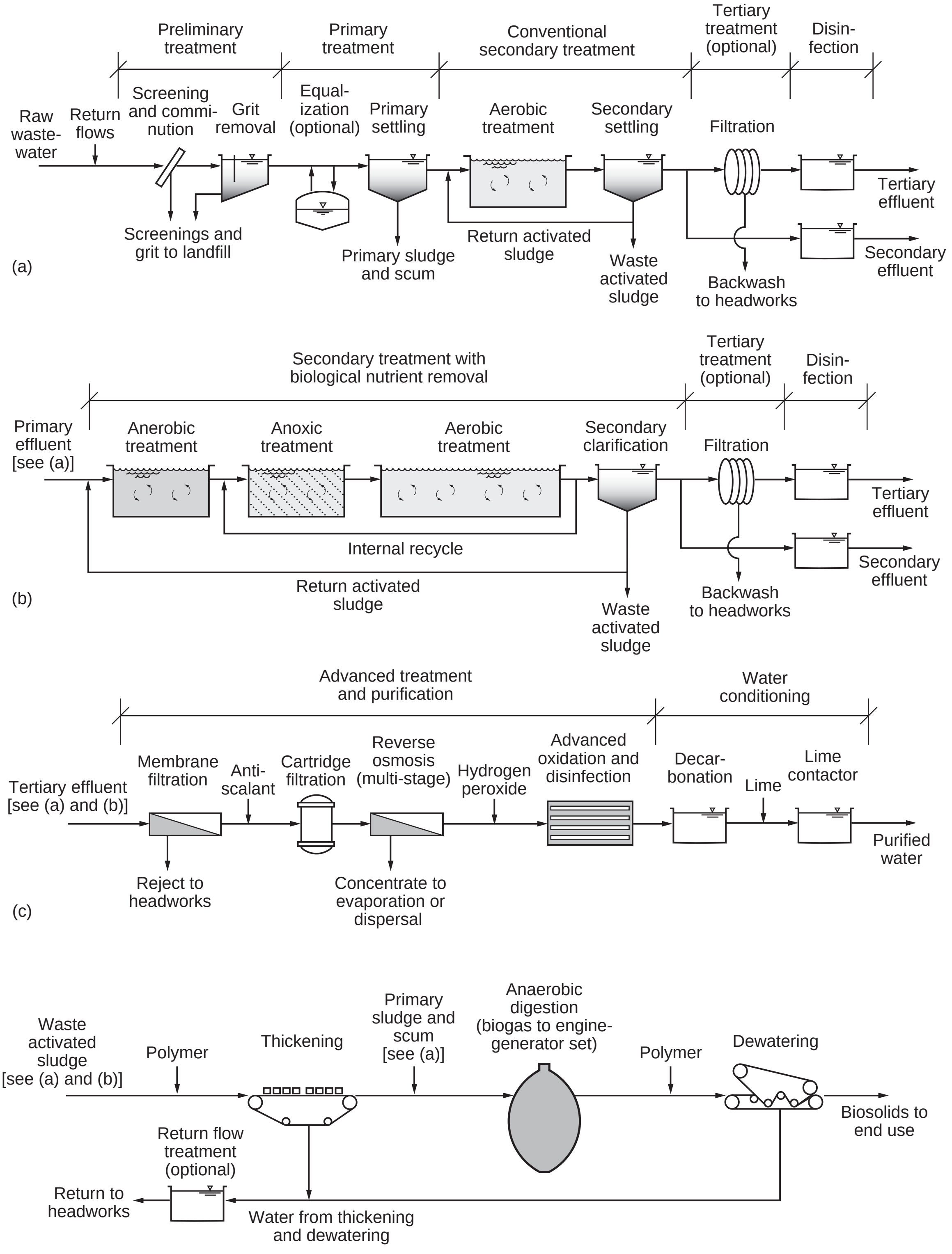


Figure 1–2

Typical flow diagrams for the treatment of wastewater and biosolids: (a) conventional biological treatment, (b) biological nutrient removal, (c) advance treatment following conventional or nutrient removal treatment, and (d) anaerobic treatment of sludge from primary sedimentation and excess biological sludge.

flow diagram is used for the removal of large, coarse solids by screening and those particles of sufficient density to be removed by gravity settling. Biological treatment is used to remove BOD_5, and TSS disinfection is used for the control of microorganisms. Tertiary treatment typically involves the filtration of settled secondary effluent to remove any residual suspended solids and thus enhance the disinfection process. Tertiary treatment is required for a number of reuse applications.

Conventional Treatment Plus Nutrient Removal. In many locations, especially around the Great Lakes in the United States and in communities surrounding inland lakes, the removal of nutrients is needed to limit eutrophication. A typical nutrient removal treatment process flow diagram is shown on Fig. 1–2(b). The flow diagram shown on Fig. 1–2(b) is similar to that shown on Fig. 1–2(a) with the exception that the biological treatment process is more complex.

Conventional Treatment with Advanced Treatment for the Production of Potable Water. Where potable water is to be produced from wastewater, advanced treatment processes beyond those used for secondary treatment or nutrient removal will be needed to remove residual suspended and colloidal and dissolved constituents. A typical advanced wastewater treatment process flow diagram is shown on Fig. 1–2(c). The process flow diagram shown on Fig. 1–2(c) is similar to that used at the Orange County Water District in California to produce potable water, which is injected into a groundwater aquifer from which it is withdrawn for use as a public water supply after a suitable residence time (Asano et al., 2007).

Processing the Coarse Solids, Primary Sludge, and Secondary Sludge from Wastewater Treatment. *Coarse solids* are produced from raw wastewater by screening and grit removal, *primary sludge* is produced from the primary sedimentation of raw wastewater, and *secondary sludge* is produced from the biological treatment of wastewater. Each of these solid fractions requires further processing. Coarse solids and grit are typically landfilled. Sludge obtained from primary sedimentation are combined with secondary solids from biological treatment and treated further by a number of different processes, with anaerobic digestion being the most common. Stabilized sludges are termed *biosolids*. A typical solids processing flow diagram is shown on Fig. 1–2(d). It should be noted that the processes shown on Fig. 1–2(d) are representative of the many different processes used for the treatment of sludge as reported in Table 1–5. Also shown on Fig. 1–2(d) is the option of treating the return flows from various solids processing equipment. Treatment of return flows is of importance where restrictive nutrient standards must be met to avoid returning excess nutrients that must be treated again (see Chap. 15).

1–6 STATUS OF WASTEWATER TREATMENT IN THE UNITED STATES

Up until the late 1980s, conventional secondary treatment was the most common method of treatment in the United States for the removal of BOD and TSS. Nutrient removal was used in special circumstances, such as in the Great Lakes area, Florida, and the Chesapeake Bay, where sensitive nutrient-related water quality conditions were identified. Because of nutrient enrichment that has led to eutrophication and water quality degradation (due in part to point source discharges), nutrient removal processes have evolved and now are used extensively in other areas as well. As a result of implementation

Table 1–7

Number of wastewater treatment facilities in the United States by flow range in 1996[a] and 2008[b]

Flow ranges		Number of facilities		Flowrate, Mgal/d		Flowrate, m^3/s	
Mgal/d	m^3/s	1996 survey	2008 survey	1996 survey	2008 survey	1996 survey	2008 survey
0.000–0.100	0.000–0.00438	6444	5703	287	257	12.6	11.3
0.101–1.000	0.0044–0.0438	6476	5863	2323	2150	101.8	94.2
1.001–10.000	0.044–0.438	2573	2690	7780	8538	340.9	374.0
10.001–100.00	0.44–4.38	446	480	11,666	12,847	511.1	562.8
>100.00	>4.38	47	38	10,119	8553	443.3	374.7
Other[c]		38	6	—	—	—	—
Total		16,204	14,780	32,175	32,345	1409.7	1417.0

[a] Adapted from U.S. EPA (1997).

[b] Adapted from U.S. EPA (2008); Alaska, North Dakota, Rhode Island, American Samoa, and the Virgin Islands are not included in the data.

[c] Flow data unknown.

of the Federal Water Pollution Control Act Amendments, significant data have been obtained on the numbers and types of wastewater facilities used and needed in accomplishing the goals of the act.

Recent Survey Results

The most recent survey results on the number of facilities, published in 2008 (U.S. EPA, 2008), are reported in Tables 1–7 and 1–8 along with the data from the 1996 survey (U.S. EPA, 1998). The comparative data have been aggregated by treatment plant capacity in Table 1–7 and by the level of treatment in Table 1–8. These data are useful in forming an overall view of the current status of wastewater treatment in the United States. In 2008, the municipal wastewater treatment enterprise comprised more than 15,000 plants (including those not included in the survey) that are used to treat a total flow of 1417 cubic meters per second (m^3/s) [32,345 Mgal/d (million gallons per day)]. Approximately 93 percent of the total existing flow is handled by plants having a capacity of 0.044 m^3/s [1 million gallons per day (Mgal/d)] and larger. More than one-half of the present design capacity is situated in plants providing greater than secondary treatment.

Trends

Comparing the total flowrate between 1996 and 2008, the per capita flowrate has gone down because in the same period the population increased from about 266.5 million to 305 million persons. It is also interesting to note the shifts that have occurred in the level of treatment. The number of treatment plants providing less than secondary treatment has decreased from 176 to 30. Also, the number of treatment plants providing greater than secondary treatment has increased from 4428 to 5071. It is anticipated that the trends observed in these tables will continue, especially the decreasing flowrate per capita (discussed further in Chap. 3) and the shift in the level of treatment toward greater than secondary.

Table 1–8

Number of wastewater treatment facilities in the United States by level of treatment in 1996[a] and 2008[b]

Level of treatment	Number of facilities 1996 survey	Number of facilities 2008 survey	Flowrate, Mgal/d 1996 survey	Flowrate, Mgal/d 2008 survey	Flowrate, m^3/s 1996 survey	Flowrate, m^3/s 2008 survey
Less than secondary	176	30	3054	422	133.8	18.5
Secondary	9388	7302	17,734	13,142	777.0	575.7
Greater than secondary[c]	4428	5071	20,016	16,776	877.0	734.9
No discharge[d]	2032	2251	1421	1815	62.3	79.5
Total	16,024	14,780	42,225	32,345	1850.1	1408.6

[a] Adapted from U.S. EPA (1997).

[b] Adapted from U.S. EPA (2008); Alaska, North Dakota, Rhode Island, American Samoa and the Virgin Islands are not included in the data.

[c] Treatment plants that meet effluent standards higher than those given in Table 1–3.

[d] Plants that do not discharge to a water body and dispose of wastewater via methods such as industrial reuse, irrigation, or evaporation.

1–7 INTRODUCTION TO PROCESS ANALYSIS

Most of the physical, chemical, and biological unit processes, identified in Tables 1–5 and 1–6, are carried out in vessels or tanks commonly known as "reactors." The fundamental approach used to describe the changes that take place when a reaction is occurring in a reactor or in some definable portion of a body of liquid is the mass-balance analysis. The basic aspects involved in such an analysis along with the types of reactors that are available and their applications are described in this section. Although, the material presented in this and in the subsequent sections that follow may be known to some, it is included here as background reference material that can be referred to as needed. Similarly, the material presented in Appendixes H and I on dispersion and tracer analysis serves the same purpose.

Mass-Balance Analysis

The fundamental basis for the analysis of the physical, chemical, and biological unit processes used for wastewater treatment is the *materials mass balance principle*, based on the principle that mass is neither created nor destroyed, but the form of the mass can be altered (e.g., liquid to a gas). The mass balance analysis affords a convenient way of defining what occurs within treatment reactors as a function of time. To illustrate the basic concepts involved in the preparation of a mass-balance analysis, consider the reactor shown on Fig. 1–3. For a given reactant, the general mass balance analysis is given by:

1. General word statement:

$$\begin{array}{ccccccc} \begin{array}{c}\text{Rate of accumulation}\\ \text{of reactant within the}\\ \text{system boundary}\end{array} & = & \begin{array}{c}\text{rate of flow of}\\ \text{reactant into the}\\ \text{system boundary}\end{array} & - & \begin{array}{c}\text{rate of flow of}\\ \text{reactant out of the}\\ \text{system boundary}\end{array} & + & \begin{array}{c}\text{rate of generation}\\ \text{of reactant within the}\\ \text{system boundary}\end{array} \\ (1) & & (2) & & (3) & & (4) \end{array} \tag{1–1}$$

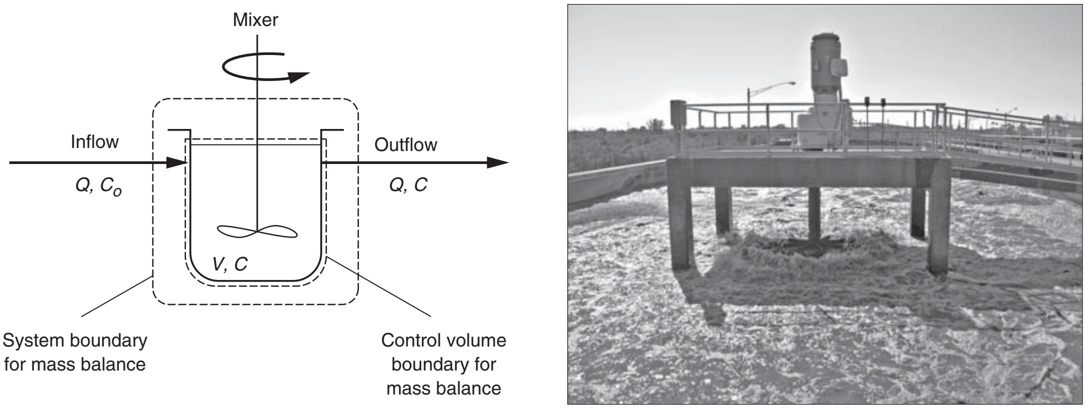


Figure 1–3

Definition sketch for the application of materials mass-balance analysis for a complete-mix reactor with inflow and outflow. The presence of a mixer is used to represent symbolically the fact that the contents of the reactor are mixed completely.

2. The corresponding simplified word statement is

$$\underset{(1)}{\text{Accumulation}} = \underset{(2)}{\text{inflow}} - \underset{(3)}{\text{outflow}} + \underset{(4)}{\text{generation}} \tag{1–2}$$

The material's mass balance is made up of the four terms cited above. Depending on the flow regime or treatment process, one or more of the terms can be equal to zero. For example, in a sealed batch reactor (i.e., no inflow or outflow) the second and third terms will be equal to zero. In the analysis of the hydraulic characteristics of reactors as well as in the analysis of separation processes discussed in Sec. 1–4, the fourth term, the rate of generation, r_c, will be equal to zero. In Eq. (1–2), a positive sign is used for the rate of generation term because the necessary sign for the operative process is part of the rate expression (e.g., $r_c = -kC$ for a decrease in the reactant or $r_c = +kC$ for an increase in the reactant).

Preparation of Mass Balances. In preparing mass balances it is helpful if the following steps are followed, especially as the techniques involved are being mastered:

1. Prepare a simplified schematic or flow diagram of the system or process for which the mass balance is to be prepared.
2. Draw a system boundary to define the limits over which the mass balance is to be applied. Proper selection of the system boundary is extremely important because, in many situations, it may be possible to simplify the mass balance computations.
3. List all of the pertinent data and assumptions that will be used in the preparation of the materials balance on the schematic or flow diagram.
4. List all of the rate expressions for the biological or chemical reactions that occur in the process.
5. Select a convenient basis on which the numerical calculations will be based.

It is recommended that these steps be followed routinely to avoid the errors that are often made in the preparation of mass balance analyses.

Application of the Mass-Balance Analysis

To illustrate the application of the mass balance analysis, consider the complete-mix reactor shown on Fig. 1–3. First, the system boundary must be established so that all the flows of mass into and out of the system can be identified. On Fig. 1–3, the boundary is shown by the outer dashed line.

To apply a mass balance analysis to the liquid contents of the reactor shown on Fig. 1–3, it will be assumed that

1. the volumetric flowrate in to and out of the control volume is constant;
2. the liquid within the control volume is not subject to evaporation (constant volume);
3. the liquid within the control volume is mixed completely;
4. a chemical reaction involving a reactant A is occurring within the reactor; and
5. the rate of change in the concentration of the reactant A that is occurring within the control volume is governed by a first order reaction ($r_c = -kC$).

Using the above assumptions, the mass balance can be formulated as follows:

1. Simplified word statement:

   Accumulation = inflow − outflow + generation

2. Symbolic representation (refer to Fig. 1–3):

$$\frac{dC}{dt}V = QC_o - QC + r_cV \tag{1–3}$$

Substituting $-kC$ for r_c yields

$$\frac{dC}{dt}V = QC_o - QC + (-kC)V \tag{1–4}$$

where dC/dt = rate of change of reactant concentration within the control volume, $ML^{-3}T^{-1}$
V = volume contained within control volume, L^3
Q = volumetric flowrate into and out of control volume, L^3T^{-1}
C_o = concentration of reactant entering the control volume, ML^{-3}
C = concentration of reactant leaving the control volume, ML^{-3}
r_c = first order reaction, ($-kC$), $ML^{-3}T^{-1}$
k = first order reaction rate coefficient, T^{-1}

Before attempting to solve any mass-balance expression, a unit check should always be made to assure that units of the individual quantities are consistent. For example, if the following units are substituted into Eq. (1–4)

V = m^3, L
dC/dt = $g/m^3 \cdot s$, mg/L·s
Q = m^3/s, L/s
C_o, C = g/m^3, mg/L
k = 1/s

the resulting unit check yields

$$\frac{dC}{dt}V = QC_o - QC + (-kC)V$$

$$(g/m^3 \cdot s)\ m^3 = m^3/s\ (g/m^3) - m^3/s\ (g/m^3) + (-1/s)(g/m^3)\ m^3$$

$$g/s = g/s - g/s - g/s \text{ (units are consistent)}$$

The analytical procedures that are adopted for the solution of mass balance equations usually are governed by (1) the nature of the rate expression, (2) the type of reactor under consideration, (3) the mathematical form of the final materials balance expression (i.e., ordinary or partial differential equation), and (4) the corresponding boundary conditions. The mass balance for a plug-flow reactor, as will be illustrated in the following section, results in a partial differential equation. A variety of solution procedures for mass balances in both ordinary and partial differential equation formats are readily available in the literature.

Steady-State Simplification. Fortunately, in most applications in the field of wastewater treatment, the solution of mass balance equations, such as the one given by Eq. (1–4), can be simplified by noting that the steady-state (i.e., long-term) concentration is of principal concern. If it is assumed that only the steady-state effluent concentration is desired, then Eq. (1–4) can be simplified by noting that, under steady-state conditions, the rate accumulation is zero ($dC/dt = 0$). Thus, Eq. (1–4) can be written as:

$$0 = QC_o - QC - r_c V \tag{1–5}$$

When solved for r_c, Eq. (1–5) yields the following expression

$$r_c = \frac{Q}{V}(C - C_o) \tag{1–6}$$

The solution to the expression given by Eq. (1–5) will depend on the nature of the rate expression (e.g., zero, first, or second order).

1–8 REACTORS USED IN WASTEWATER TREATMENT

As noted previously, the physical, chemical, and biological unit processes are carried out in vessels or tanks commonly known as “reactors.” The types of reactors that are available and their applications are introduced in this section.

Types of Reactors

The principal types of reactors used for the treatment of wastewater, illustrated on Fig. 1–4, are (1) the batch reactor, (2) the complete-mix reactor [also known as the continuous flow stirred-tank reactor (CFSTR) in the chemical engineering literature], (3) the plug-flow reactor (also known as a tubular-flow reactor), (4) complete-mix reactors in series, (5) the packed-bed reactor, and (6) the fluidized-bed reactor.

Batch Reactor. In the batch reactor [see Fig. 1–4(a)], flow is neither entering nor leaving the reactor (i.e., flow enters, is treated, and then discharged and the cycle repeats). The liquid contents of the reactor are mixed completely. For example, the BOD test discussed in Chap. 2 is carried out in a batch reactor (i.e., BOD bottle as shown on Fig. 2–21 in Chap. 2), although it should be noted that the contents are not mixed completely during the incubation period. Batch reactors are often used to blend chemicals or to dilute concentrated chemicals.

Complete-Mix Reactor (CMR). In the complete-mix reactor [see Fig. 1–4(b)], it is assumed that complete mixing occurs instantaneously and uniformly throughout the reactor as fluid particles enter the reactor. Fluid particles leave the reactor in proportion to their statistical population. Complete mixing can be accomplished in round or square reactors if the contents of the reactor are uniformly and continuously redistributed. The actual time required to achieve completely mixed conditions will depend on the reactor geometry and the power input.

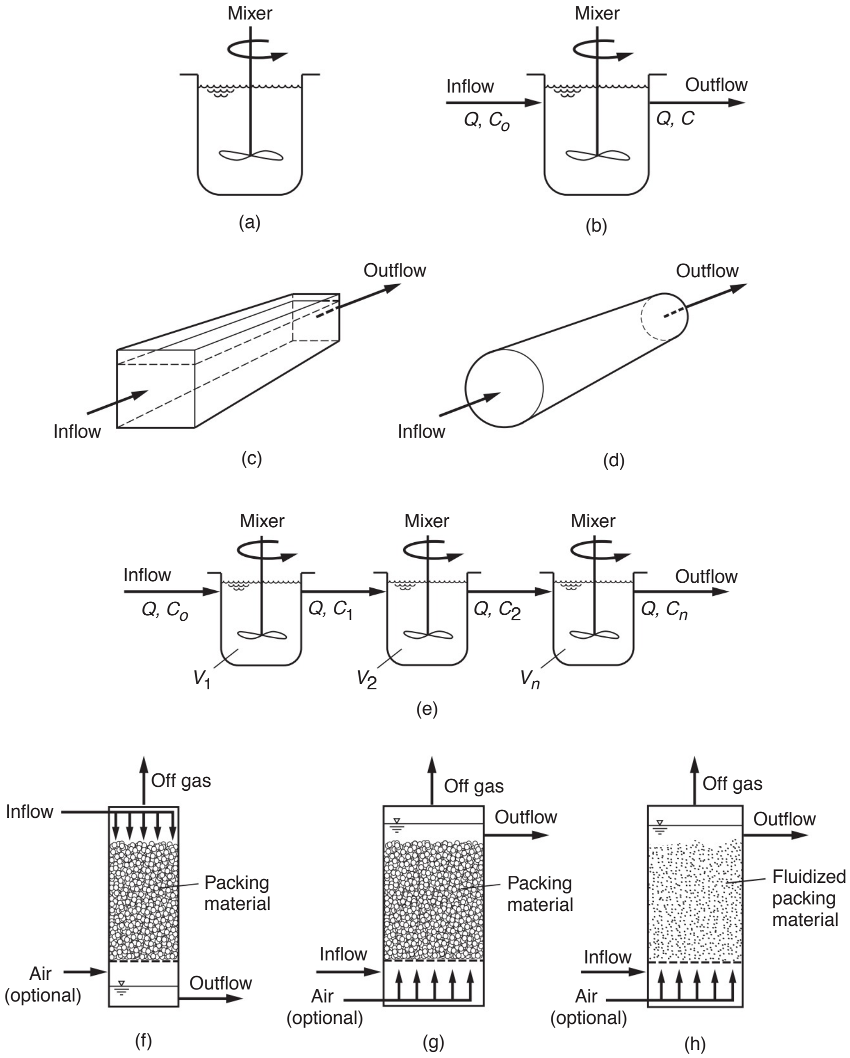


Figure 1–4

Definition sketch for the different types of reactors used for wastewater treatment: (a) batch reactor; (b) complete-mix reactor; (c) plug-flow open reactor; (d) plug-flow closed reactor, also known as a tubular reactor; (e) complete-mix reactors in series; (f) packed bed downflow reactor; (g) packed bed upflow reactor; and (h) expanded bed upflow reactor.

Plug-Flow Reactor (PFR). Fluid particles pass through the reactor with little or no longitudinal mixing and exist from the reactor in the same sequence in which they entered. The particles retain their identity and remain in the reactor for a time equal to the theoretical detention time. This type of flow is approximated in long open tanks with a high length-to-width ratio in which longitudinal dispersion is minimal or absent [see Fig. 1–4(c)] or closed tubular reactors [e.g., pipelines, see Fig. 1–4(d)].

Complete-Mix Reactors in Series. The series of complete-mix reactors [see Fig. 1–4(e)] is used to model the flow regime that exists between the ideal hydraulic flow patterns corresponding to the complete-mix and plug-flow reactors. If the series is composed of one reactor, the complete-mix regime prevails. If the series consists of an infinite number of reactors in series, the plug-flow regime prevails.

Packed-Bed Reactor. The packed-bed reactor is filled with a type of packing medium, such as rock, slag, ceramic, or, now more commonly, plastic. With respect to flow, the packed-bed reactor can be operated in either the downflow or upflow mode.

Dosing can be continuous or intermittent (e.g., trickling filter). The packing medium in packed bed reactors can be continuous [see Fig. 1–4(f)] or arranged in multiple stages, with flow from one stage to another. A packed-bed upflow anaerobic (without oxygen) reactor is shown on Fig. 1–4(g).

Fluidized-Bed Reactor. The fluidized-bed reactor is similar to the packed-bed reactor in many respects, but the packing medium is expanded by the upward movement of fluid (air or water) through the bed [see Fig. 1–4(h)]. The expanded porosity of the fluidized-bed packing medium can be varied by controlling the flowrate of the fluid.

Hydraulic Characteristics of Reactors

Complete-mix and plug-flow reactors are the two reactor types used most commonly in the field of wastewater treatment. The hydraulic flow characteristics of complete-mix and plug-flow reactors can be described as varying from ideal and nonideal, depending on the relationship of the incoming flow to outgoing flow. Ideal and nonideal flow, as well as application of reactors in wastewater treatment, are described in the following discussion.

Ideal Flow in Complete-Mix and Plug-Flow Reactors. The ideal hydraulic flow characteristics of complete-mix and plug-flow reactors are illustrated on Fig. 1–5 in which dye tracer response curves are presented for pulse (slug injection) and step inputs (continuous injection). On Fig. 1–5, t is the actual time and τ is equal to the theoretical hydraulic detention time defined as follows:

$$\tau = \frac{V}{Q} \tag{1–7}$$

where τ = the hydraulic detention time, T
V = volume of the reactor, L^3
Q = volumetric flowrate, L^3T^{-1}

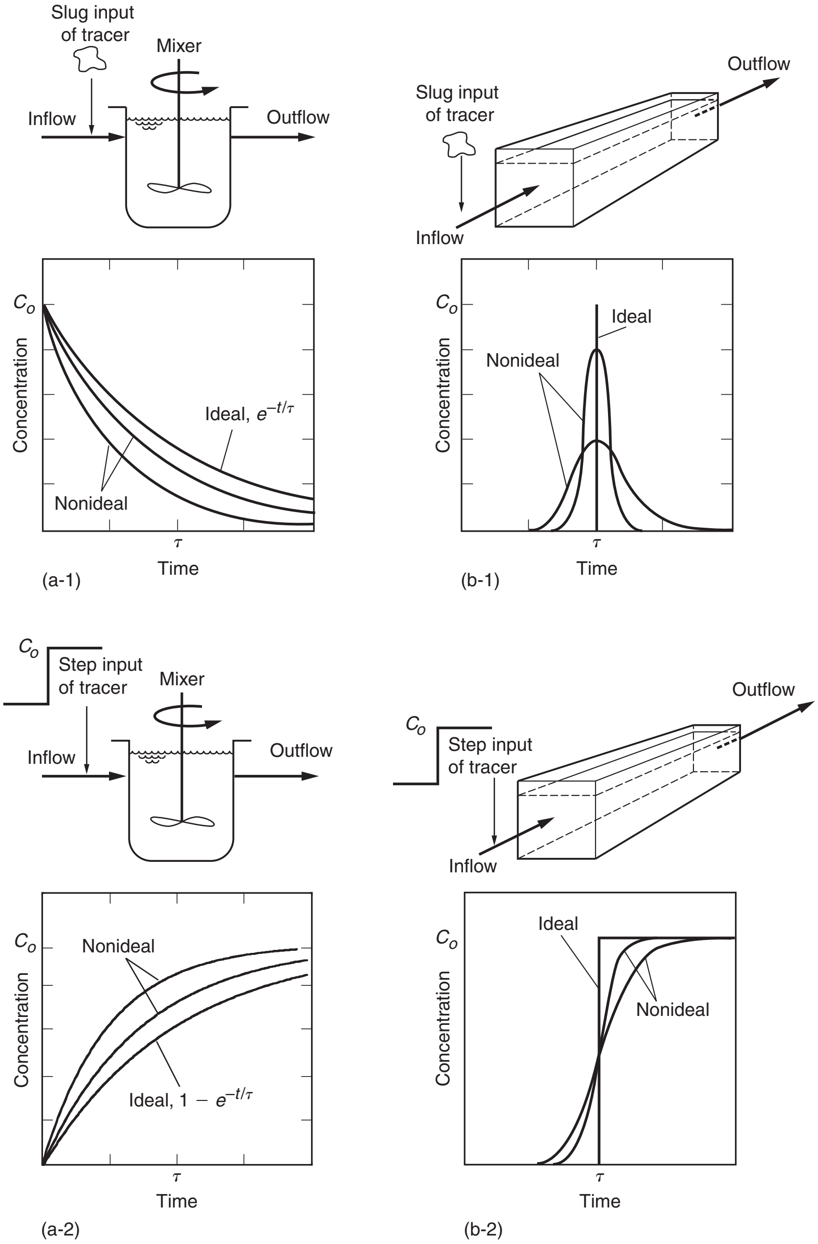
If a pulse input of a conservative (i.e., nonreactive) tracer is injected and dispersed instantaneously in an ideal flow complete-mix reactor, with a continuous inflow of clear water, the output tracer concentration would appear as shown on Fig. 1–5(a-1). If a continuous step input of a conservative tracer at concentration C_o is injected into the inlet of an ideal complete-mix reactor, initially filled with clear water, the appearance of the tracer at the outlet would occur as shown on Fig. 1–5(a-2).

In the case of an ideal plug-flow reactor, the reactor is initially filled with clear water before being subjected to a pulse or a step input of tracer. If an observer were positioned at the outlet of the reactor, the appearance of the tracer in the effluent for a pulse input, distributed uniformly across the reactor cross-section, would occur as shown on Fig. 1–5(b-1). If a continuous step input of a tracer were injected into such a reactor at an initial concentration C_o, the tracer would appear in the effluent as shown on Fig. 1–5(b-2).

Nonideal Flow in Complete-mix and Plug-Flow Reactors. In practice, the flow in complete-mix and plug-flow reactors is seldom ideal. For example, when a reactor is designed, how is the flow to be introduced to satisfy the theoretical requirement of instantaneous and complete dispersion? In practice, there is always some deviation from ideal conditions, and it is the precautions taken to minimize these effects that are important. Nonideal flow occurs when a portion of the flow that enters the reactor during a given time period arrives at the outlet before the bulk of the flow that entered the reactor during the same time period arrives. Nonideal flow is illustrated on Figs. 1–5(a-2) and 1–5(b-2).

Figure 1–5

Output tracer response curves from reactors subject to slug (pulse) and step inputs of a tracer: (a-1) and (a-2) complete-mix reactor and (b-1) and (b-2) plug-flow reactor.



The important issue with nonideal flow is that a portion of the flow will not remain in the reactor as long as may be required for a biological or chemical reaction to go to completion.

Application of Reactors

The principal applications of reactor types used for wastewater treatment are reported in Table 1–9. Operational factors that must be considered in the selection of the type of reactor or reactors to be used in the treatment process include (1) the nature of the wastewater to be treated, (2) the nature of the reaction (i.e., homogeneous or heterogeneous),

Table 1–9

Principal applications of reactor types used for wastewater treatment

Type of reactor	Applications in wastewater treatment
Batch	Conduct of BOD test, sequencing batch reactor activated sludge biological treatment, mixing of concentrated solutions into working solutions
Complete-mix without recycle	Aerated lagoons, aerobic sludge digestion, anaerobic digestion
Complete-mix with recycle	Activated sludge biological treatment
Plug-flow	Chlorine contact basin, reaeration basin, natural treatment systems
Plug-flow with recycle	Activated sludge biological treatment, aquatic treatment systems
Complete-mix reactors in series	Lagoon treatment systems, used to simulate nonideal flow in plug flow reactors
Packed bed	Nonsubmerged and submerged trickling filter biological treatment units, depth filtration, membrane filtration, adsorption, ion exchange, air stripping, natural treatment systems
Fluidized bed	Fluidized bed reactors for aerobic and anaerobic biological treatment, upflow sludge blanket reactors, air stripping, thermal oxidation of sludge

(3) the reaction kinetics governing the treatment process, (4) the process performance requirements, and (5) local environmental conditions. Homogeneous and heterogeneous reactions and reaction kinetics are discussed in Sec. 1–9. In practice, the construction costs and operation and maintenance costs also affect reactor selection. Because the relative importance of these factors varies with each application, each factor should be considered separately when the type of reactor is to be selected.

1–9 MODELING IDEAL FLOW IN REACTORS

Modeling of the hydraulic characteristics of reactors is important because the results can be used to determine the actual amount of time a given volume of water will remain in the reactor and its average age. In turn, the average ages can be related to the degree of treatment achieved, based on the applicable kinetics. The coupling of reactor hydraulic characteristics and reaction rates to determine treatment process performance is considered in Sec. 1–9.

Comparison of actual hydraulic characteristics of a reactor, measured using tracers, to the expected theoretical response can be used to assess the degree to which the design ideal has been achieved. The complete-mix and plug-flow reactors, as noted previously, are the reactor types used most commonly in the field of wastewater treatment. The mathematical analysis of ideal flow in complete-mix and plug-flow reactors is considered below. The modeling of nonideal flow is considered in Appendix I.

Ideal Flow in Complete-Mix Reactor

Analytically, using the mass-balance approach introduced above, the effluent tracer concentration from an ideal flow complete-mix reactor as a function of time for a pulse input of tracer, which is mixed instantaneously and is purged with clear water [see Fig. 1–5(a–1)] can be determined by writing a mass balance around the reactor.

1. **General word statement:**

$$\begin{matrix}\text{Rate of accumulation} & & \text{rate of flow of} & & \text{rate of flow of} \\ \text{of tracer within} & = & \text{tracer into} & - & \text{tracer out of} \\ \text{the reactor} & & \text{the reactor} & & \text{the reactor}\end{matrix} \qquad (1\text{–}8)$$

2. **Simplified word statement:**

   Accumulation = inflow − outflow (1–9)

3. **Symbolic representation [refer to Fig. 1–5(a-1)]:**

$$\frac{dC}{dt}V = QC_o - QC \qquad (1\text{–}10)$$

Rewriting Eq. (4–11) and simplifying by noting that $C_o = 0$ yields:

$$\frac{dC}{dt} = -\frac{Q}{V}C \qquad (1\text{–}11)$$

Integrating between the limits of $C = C_o$ to $C = C$, and $t = 0$ to $t = t$ yields

$$\int_{C_o}^{C}\frac{dC}{C} = -\frac{Q}{V}\int_0^t dt \qquad (1\text{–}12)$$

The resulting expression after integration is

$$C = C_o e^{-t(Q/V)} = C_o e^{-t/\tau} = C_o e^{-\theta} \qquad (1\text{–}13)$$

where C = concentration of the tracer in the reactor at time t, ML^{-3}
C_o = initial concentration of the tracer in the reactor, ML^{-3}
t = time, T
Q = volumetric flowrate, $\mathrm{L}^3\mathrm{T}^{-1}$
V = reactor volume, L^3
τ = the theoretical detention time, V/Q, T
θ = the normalized detention time, t/τ, unitless

The corresponding response for a continuous step input of tracer [see Fig. 1–5(a-2)] which is mixed instantaneously is given by

$$C = C_o(1 - e^{-t(Q/V)}) = C_o(1 - e^{-t/\tau}) = C_o(1 - e^{-\theta}) \qquad (1\text{–}14)$$

It will be noted that Eq. (1–14) has the same form as the BOD equation [see Chap. 2, Eq. (2–60)].

Ideal Plug-Flow Reactor

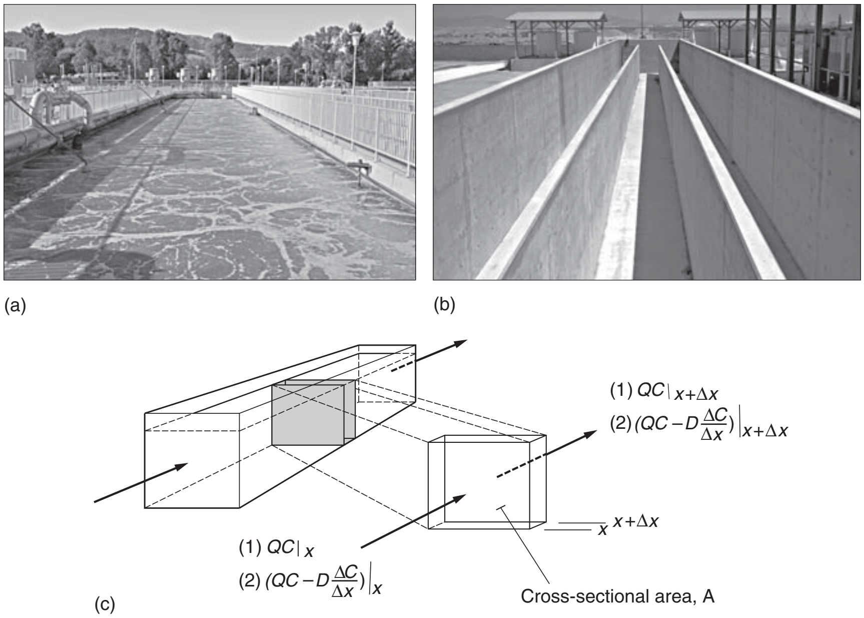
Under ideal plug-flow conditions, t, the measured detention time, should be the same as τ, the theoretical detention time (V/Q). To verify the form of the plot given previously on Fig. 1–5(b-2), it will be instructive to prepare a materials balance for an ideal plug-flow reactor (no axial dispersion) in which the concentration, C, of a nonreactive tracer is distributed uniformly across the cross-sectional area of the control volume. The materials balance for a nonreactive tracer for the differential volume element shown on Fig. 1–6 can be written as follows:

1. **General word statement:**

$$\begin{matrix}\text{Rate of accumulation of} & & \text{rate of flow of tracer} & & \text{rate of flow of tracer} \\ \text{tracer within differential} & = & \text{into differential} & - & \text{out of differential} \\ \text{volume element} & & \text{volume element} & & \text{volume element}\end{matrix} \qquad (1\text{–}15)$$

Figure 1–6

Views of plug-flow reactors and definition sketch: (a) view of plug-flow activated sludge process reactor, (b) view of empty plug-flow chlorine contact basin with long narrow channels, and (c) definition sketch for the hydraulic analysis of a plug-flow reactor with (1) advection only and (2) with advection and axial dispersion.



2. Simplified word statement:

Accumulation = inflow − outflow (1–16)

3. Symbolic representation (refer to Fig. 1–6)

$$\frac{\partial C}{\partial t}\Delta V = QC|_x - QC|_{x+\Delta x} \tag{1–17}$$

where $\partial C/\partial t$ = change in constituent concentration with time, $ML^{-3}T^{-1}$, (g/m^3·s)
ΔV = differential volume element, L^3, (m^3)
t = time, T, (s)
Q = volumetric flowrate, L^3T^{-1}, (m^3/s)
x = some point along the reactor length, L, (m)

The change in concentration with time term ($\partial C/\partial t$) is written as partial differential because the concentration is also changing with distance (i.e., the change in concentration is a function of both time and distance). Substituting the differential form for the term $QC|_{x+\Delta x}$ in Eq. (1–17) results in

$$\frac{\partial C}{\partial t}\Delta V = QC - Q\left(C + \frac{\Delta C}{\Delta x}\Delta x\right) \tag{1–18}$$

Substituting $A\Delta x$ for ΔV, where A is the cross-sectional area in the x direction, and simplifying yields

$$\frac{\partial C}{\partial t}A\Delta x = -Q\frac{\Delta C}{\Delta x}\Delta x \tag{1–19}$$

Dividing by A and Δx yields

$$\frac{\partial C}{\partial t} = -\frac{Q}{A}\frac{\Delta C}{\Delta x} \tag{1–20}$$

Taking the limit as Δx approaches zero yields

$$\frac{\partial C}{\partial t} = -\frac{Q}{A}\frac{\partial C}{\partial x} = -\upsilon\frac{\partial C}{\partial x} \tag{1–21}$$

where υ = the velocity of flow, LT^{-1}, (m/s)

Because both sides of the equation are the same (note $\partial t = \partial x/\upsilon$), except for the minus sign, the only way that the equation can be satisfied is if the change in concentration with distance is equal to zero. Thus, the effluent concentration must be equal to the influent concentration, which is consistent with the depiction on Fig. 1–5(b-2).

1–10 INTRODUCTION TO PROCESS KINETICS

From the standpoint of process selection and design, the controlling stoichiometry and the rates of the reaction for chemical and biological unit processes are of principal concern. The number of moles of a substance entering into a reaction and the number of moles of the substances produced are defined by the stoichiometry of a reaction. The stoichiometry of reaction refers to the definition of the quantities of chemical compounds involved in a reaction. The rate at which a substance disappears or is formed in any given stoichiometric reaction is defined as the rate of reaction. These and other related topics are discussed in this section. The rate expressions discussed in this section will be integrated with the hydraulic characteristics of the reactors, discussed previously, to define treatment kinetics.

Types of Reactions

The two principal types of reactions that occur in wastewater treatment are classified as homogeneous and heterogeneous (nonhomogeneous).

Homogeneous Reactions. In homogeneous reactions, the reactants are distributed uniformly throughout the fluid so that the potential for reaction at any point within the fluid is the same. Homogeneous reactions are usually carried out in the batch, complete-mix, and plug-flow reactors [see Figs. 1–4 (a), (b), (c), and (d)]. Homogeneous reactions may be either irreversible or reversible.

Examples of irreversible reactions are

1. Simple reactions

$$\mathrm{A \rightarrow B} \tag{1–22}$$

$$\mathrm{A + A \rightarrow C} \tag{1–23}$$

$$\mathrm{aA + bB \rightarrow C} \tag{1–24}$$

2. Parallel reactions

$$\mathrm{A + B \rightarrow C} \tag{1–25}$$

$$\mathrm{A + B \rightarrow D} \tag{1–26}$$

3. Consecutive reactions

$$\mathrm{A + B \rightarrow C} \tag{1–27}$$

$$\mathrm{A + C \rightarrow D} \tag{1–28}$$

Examples of reversible reactions are

$$\mathrm{A} \rightleftarrows \mathrm{B} \tag{1–29}$$

$$\mathrm{A} + \mathrm{B} \rightleftarrows \mathrm{C} + \mathrm{D} \tag{1–30}$$

As will be discussed subsequently, for both irreversible and reversible reactions, the rate of reaction will be an important consideration in the design of the treatment facilities in which these reactions will be carried out. Special attention must be given to the design of mixing facilities, especially for reactions that are rapid.

Heterogeneous Reactions. Heterogeneous reactions occur between one or more constituents that can be identified with specific sites, such as those on an ion-exchange resin in which one or more ions is replaced by another ion. Reactions that require the presence of a solid-phase catalyst are also classified as heterogeneous. Heterogeneous reactions are usually carried out in packed and fluidized bed reactors [see Fig. 1–4(f), (g), and (h)]. These reactions are more difficult to study because a number of interrelated steps may be involved. The typical sequence of these steps, as quoted from Smith (1981) is as follows:

1. Transport of reactants from the bulk fluid to the fluid-solid interface (external surface of catalyst particle)
2. Intraparticle transport of reactants into the catalyst particle (if it is porous)
3. Adsorption of reactants at interior sites of the catalyst particle
4. Chemical reaction of adsorbed reactants to adsorbed products (surface reaction)
5. Desorption of adsorbed products
6. Transport of products from the interior sites to the outer surface of the catalyst particle

Rate of Reaction

The rate of reaction is the term used to describe the change (decrease or increase) in the number of moles of a reactive substance per unit volume per unit time (for homogeneous reactions), or per unit surface area or mass per unit time (for heterogeneous reactions) (Denbigh and Turner, 1984).

For homogeneous reactions, the rate of reaction r is given by

$$r = \frac{1}{V}\frac{d[N]}{dt} = \frac{\text{moles}}{(\text{volume})(\text{time})} \tag{1–31}$$

If N is replaced by the term VC, where V is the volume and C is the concentration, Eq. (1–31) becomes

$$r = \frac{1}{V}\frac{d(VC)}{dt} = \frac{1}{V}\frac{VdC + CdV}{dt} \tag{1–32}$$

If the volume remains constant (i.e., isothermal conditions, no evaporation), Eq. (1–32) reduces to

$$r = \pm\frac{dC}{dt} \tag{1–33}$$

where the plus sign indicates an increase or accumulation of the substance, and the minus sign indicates a decrease of the substance.

For heterogeneous reactions where S is the surface area, the corresponding expression is

$$r = \frac{1}{S}\frac{d[N]}{dt} = \frac{\text{moles}}{(\text{area})(\text{time})} \tag{1–34}$$

For reactions involving two or more reactants with unequal stoichiometric coefficients, the rate expressed in terms of one reactant will not be the same as the rate for the other reactants. For example, for the reaction

$$aA + bB \rightarrow cC + dD \tag{1–35}$$

the concentration changes for the various reactants are given by

$$-\frac{1}{a}\frac{d[A]}{dt} = -\frac{1}{b}\frac{d[B]}{dt} = \frac{1}{c}\frac{d[C]}{dt} = \frac{1}{d}\frac{d[D]}{dt} \tag{1–36}$$

Thus, for reactions in which the stoichiometric coefficients are not equal, the rate of reaction is given by

$$r = \frac{1}{c_i}\frac{d[C_i]}{dt} \tag{1–37}$$

where the coefficient term $(1/c_i)$ is negative for reactants and positive for products.

The rate at which a reaction proceeds is an important consideration in wastewater treatment. For example, in some cases the operative reaction may take too long to go to completion. In such cases, treatment processes are designed on the basis of the rate at which the reaction proceeds rather than the equilibrium position of the reaction. Often, quantities of chemicals in excess of the stoichiometric, or exact reacting amount, may be used to accomplish the treatment step in a shorter period of time by driving the reaction to completion.

Specific Reaction Rate

From the law of mass action it can be shown that the rate of reaction for a given reaction is proportional to the remaining concentration of the reactants. Thus, for a reaction involving a single component *A*, the rate of reaction is given by

$$r = \pm k C_A \tag{1–38}$$

Where *k* is a constant of proportionality formally defined as the specific reaction rate (also known as the reaction-rate constant, velocity constant, and the rate coefficient). The specific reaction rate has the units of the specific reaction and concentration. For Eq. (1–38), the units of the specific reaction-rate constant are

$$k = \frac{r}{C} = \frac{1}{V}\frac{dN}{dt}\frac{1}{C} = \frac{\text{mole}}{\text{L}\cdot\text{s(mole/L)}} = \frac{1}{s} \tag{1–39}$$

In application, the rate of reaction, *r*, takes into account the effects of concentration, and the specific reaction-rate constant, *k*, takes into account the effects of all the other variables that may affect the reaction. Of the many variables in any given situation, temperature is usually the most important.

Effects of Temperature on Reaction Rate Coefficients

The temperature dependence of the specific reaction rate constants is important because of the need to adjust for other temperatures. The temperature dependence of the rate constant is given by the van't Hoff-Arrhenius relationship.

$$\frac{d(\ln k)}{dT} = \frac{E}{R^2} \tag{1–40}$$

where k = reaction rate constant at temperature T
T = temperature, K = 273.15 + °C
E = activation energy (a characteristic value for a reaction (e.g., J/mole)
R = ideal gas constant, 8.314 J/mole·K (1.99 cal/mole·K)

Integration of Eq. (1–40) between the limits T_1 and T_2 gives

$$\ln\frac{k_2}{k_1} = \frac{E(T_2 - T_1)}{RT_1T_2} = \frac{E}{RT_1T_2}(T_2 - T_1) \tag{1–41}$$

With k_1 known for a given temperature and with E known, k_2 can be calculated.

Activation Energy. The activation energy, E, can be calculated using Eq. (1–41) by determining the k at two different temperatures as illustrated in Example 1–1. Common values of E for wastewater treatment processes are in the range of 8400 to 84,000 J/mole (2000 to 20,000 cal/mole).

EXAMPLE 1–1 **Determination of Activation Energy** For a given chemical reaction it has been observed that the rate of reaction doubles for each 10°C increase in temperature. If the initial temperature was 10°C, estimate the activation energy for the reaction.

Solution

1. Solve Eq. (1–41) for the activation energy. The required equation is

$$E = \frac{R\ln(k_2/k_1)}{(1/T_1 - 1/T_2)}$$

2. Substitute known values and solve for E:

$T_1 = (273 + 10°\text{C}) = 283\text{ K}$

$T_2 = (273 + 20°\text{C}) = 293\text{ K}$

$k_2 = 2k_1$

$R = 8.314\text{ J/mole·K}$

$$E = \frac{(8.314\text{ J/mole·K})(\ln 2k_1/k_1)}{(1/283\text{K} - 1/293\text{K})} = 48{,}024\text{ J/mole}$$

Comment Although used as a constant, the value of the activation energy, E, will vary somewhat with temperature according to the above equation. However, the temperature range in which wastewater treatment process operate is relatively limited. There is much greater variability in the measured reaction rates.

Temperature Coefficient, θ. Because most wastewater treatment processes are carried out over a relatively narrow temperature range, the term, E/RT_1T_2, in Eq. (1–41) may be assumed to be constant for all practical purposes. If the term E/RT_1T_2 is designated by C, then Eq. (1–41) can be written as

$$\ln\frac{k_2}{k_1} = C(T_2 - T_1) \tag{1–42}$$

$$\frac{k_2}{k_1} = e^{C(T_2 - T_1)} \tag{1–43}$$

Replacing the term, e^C, in Eq. (1–43) with a temperature coefficient, θ, yields the following expression:

$$\frac{k_2}{k_1} = \theta^{(T_2 - T_1)} \tag{1–44}$$

Equation (1–44) is used commonly in the sanitary engineering field to adjust the value of the operative rate constant to reflect the effect of temperature. It should be noted, however, that although the value of θ is assumed to be constant, it can often vary considerably with temperature. Therefore, caution must be used in selecting appropriate values for θ for different temperature ranges. Typical values for various processes for different temperature ranges are given, where available, in the sections in which the individual topics are discussed. Values for θ for some biological treatment systems vary from about 1.020 to 1.10.

Reaction Order

The rate at which reactions occur is determined usually by measuring the concentration of either a reactant or product as the reaction proceeds to completion. The measured results are then compared to the corresponding results obtained from various standard rate equations by which the reaction under study is expected to proceed.

The order of a reaction with respect to a specified compound is equal to the stoichiometric coefficient for that compound. For example, in the following reaction, the reaction order for compound A is a, compound B is b, and so on.

$$\text{aA} + \text{bB} + \ldots \rightarrow \text{pP} + \text{qQ} + \ldots \tag{1–45}$$

If the rate is experimentally found to be proportional to the first power of the concentration of A (i.e., a = 1), then the reaction is said to be first order with respect to A.

When the mechanism of reaction is not known, the reaction rate for Eq. (1–45) may be approximated with the following expression:

$$r = kC_A^a C_B^b C_C^c \cdots C_P^p = kC_A^n \tag{1–46}$$

where a and b are the reaction orders with respect to reactants A and B, and n is the overall reaction order ($n = a + b + \ldots p$). The sum of the exponents to which the concentration(s) are raised is known as the order of the reaction. Several reaction rate expressions with different reaction orders are as follows.

$$r = \pm k \qquad \text{(Zero order)} \tag{1–47}$$

$$r = \pm kC \qquad \text{(First order)} \tag{1–48}$$

$$r = \pm k(C - C_s) \qquad \text{(First order)} \tag{1–49}$$

$$r = \pm kC^2 \qquad \text{(Second order)} \tag{1–50}$$

$$r = \pm kC_A C_B \qquad \text{(Second order)} \tag{1–51}$$

$$r = \pm \frac{kC}{K + C} \qquad \text{(Saturation or mixed order)} \tag{1–52}$$

$$r = \pm \frac{kC}{(1 + r_t t)^n} \qquad \text{(First order retarded)} \tag{1–53}$$

The application of various rate expressions in wastewater treatment are described in the following discussion.

Rate Expressions Used in Wastewater Treatment

The physical, chemical, and biological processes that are applied in wastewater treatment for the conversion or separation of constituents are numerous and varied. Important constituent treatment processes, along with the constituents affected, are reported in Table 1–10. The various processes listed in Table 1–10 will be referred to throughout this text.

Table 1–10

Common constituent conversion and separation processes (i.e., fate processes) in the environment and the constituents affected

Process	Comments	Constituents effected
Adsorption/ desorption	Many chemical constituents tend to attach or sorb onto solids. The implication for wastewater discharges is that a substantial fraction of some toxic chemicals are associated with the suspended solids in the effluent. Adsorption combined with solids settling results in the removal from the water column of constituents that might not otherwise decay.	Metals, trace organics, NH_4^+, PO_4^{3-}
Algal synthesis	The synthesis of algal cell tissue using the nutrients found in wastewater.	NH_4^+, NO_3^-, PO_4^{3-}, pH, etc.
Bacterial conversion	Bacterial conversion (both aerobic and anaerobic) is the most important process in the transformation of constituents released to the environment. The exertion of BOD and nitrogenous oxygen demand (NOD) are the most common examples of bacterial conversion encountered in water-quality management. The depletion of oxygen in the aerobic conversion of organic wastes is also known as deoxygenation. Solids discharged with treated wastewater are partly organic. Upon settling to the bottom, they decompose bacterially either anaerobically or aerobically, depending on local conditions. The bacterial transformation of toxic organic compounds is also of great significance.	BOD_5, nitrification, denitrification, sulfate reduction, anaerobic fermentation (in bottom sediments), conversion of priority organic pollutants, etc.
Chemical reactions	Important chemical reactions that occur in the environment include hydrolysis, photochemical, and oxidation-reduction reactions. Hydrolysis reactions occur between contaminants and water.	Chemical disinfection, decomposition of organic compounds, specific ion exchange, element substitution
Filtration	Removal of suspended and colloidal solids by straining (mechanical and chance contact), sedimentation, interception, impaction, and adsorption.	TSS, colloidal particles
Flocculation	Flocculation is the term used to describe the aggregation of smaller particles into larger particles that can be removed by sedimentation and filtration. Flocculation is brought about by Brownian motion, differential velocity gradients, and differential settling in which large particles overtake smaller particles and form larger particles.	Colloidal and small particles
Gas absorption/ desorption	The process whereby a gas is taken up by a liquid is known as absorption. For example, when the dissolved oxygen concentration in a body of water with a free surface is below the saturation concentration in the water, a net transfer of oxygen occurs from the atmosphere to the water. The rate of transfer (mass per unit time per unit surface area) is proportional to the amount by which the dissolved oxygen is below saturation. The addition of oxygen to water is also known as reaeration. Desorption occurs when the concentration of the gas in the liquid exceeds the saturation value, and there is a transfer from the liquid to the atmosphere.	O_2, CO_2, CH_4, NH_3, H_2S

(*continued*)

Table 1–10 (*Continued*)

Process	Comments	Constituents effected
Natural decay	In nature, contaminants will decay for a variety of reasons, including mortality in the case of bacteria and photooxidation for certain organic constituents. Natural and radioactive decay usually follow first-order kinetics.	Plants, animals, algae, fungi, protozoa, eubacteria (most bacteria), archaebacteria, viruses, radioactive substances, plant mass
Photochemical reactions	Solar radiation is known to trigger a number of chemical reactions. Radiation in the near-ultraviolet (UV) and visible range is known to cause the breakdown of a variety of organic compounds.	Oxidation of inorganic and organic compounds
Photosynthesis/ respiration	During the day, algal cells in water bodies will produce oxygen by means of photosynthesis. Dissolved oxygen concentrations as high as 30 to 40 mg/L have been measured. During the evening hours algal respiration will consume oxygen. Where heavy growths of algae are present, oxygen depletion has been observed during the evening hours.	Algae, duckweed, submerged macrophytes, NH_4^+, PO_4^{3-}, pH, etc.
Sedimentation	The suspended solids discharged with treated wastewater ultimately settle to the bottom of the receiving water body. This settling is enhanced by flocculation and hindered by ambient turbulence. In rivers and coastal areas, turbulence is often sufficient to distribute the suspended solids over the entire water depth.	TSS
Sediment oxygen demand	The residual solids discharged with treated wastewater will, in time, settle to the bottom of streams and rivers. Because the particles are partly organic, they can be decomposed anaerobically as well as aerobically, depending on conditions. Algae which settle to the bottom will also be decomposed, but much more slowly. The oxygen consumed in the aerobic decomposition of material in the sediment represents another dissolved oxygen demand in the water body.	O_2, particulate BOD
Volatilization	Volatilization is the process whereby liquids and solids vaporize and escape to the atmosphere. Organic compounds that readily volatilize are known as VOCs (volatile organic compounds). The physics of this phenomenon are very similar to gas absorption, except that the net flux is out of the water surface.	VOCs, NH_3, CH_4, H_2S, other gases

For example, bacterial conversion is considered in Chap. 2 in the analysis of the BOD reaction, and in greater detail in the chapters dealing with biological treatment. Because all of the processes summarized in Table 1–10 are rate dependent, representative rate expressions used to model these processes are presented in Table 1–11. The important thing to note about Table 1–11 is the variety of different rate expressions that have been used to model constituent conversion and separation processes.

Conversion Processes. Rate expressions have been used to describe the conversion of wastewater constituents in treatment processes and the fate of constituents released in the environment. For example, the first order reaction, Eq. (1–48), expressed as ($r_c = -kC$) is used to model the exertion of BOD and bacterial decay, as discussed subsequently in Chap. 2. Although Eq. (1–51) is second order overall, it is first order with respect to C_A and C_B, individually. Equation (1–52), known as a saturation type of equation (also known as a Monod type equation), is illustrated on Fig. 1–7. As shown on Fig. 1–7, when the concentration, C, is large the rate of reaction is zero order, and when the concentration is low, the rate of reaction is first order.

Table 1–11

Examples of rate expressions for selected conversion and separation processes given in Table 1–10[a]

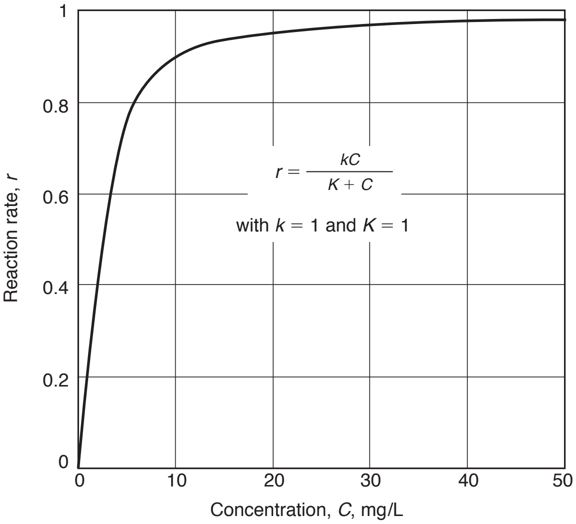
Process	Rate expression	Comments
Conversion processes		
Bacterial conversion	$r_c = -kC$	r_c = rate of conversion, M/L^3T k = first order reaction rate coefficient, $1/T$ C = concentration of organic material remaining, M/L^3
Chemical reactions	$r_c = \pm k_n C^n$	r_c = rate of conversion, M/L^3T k_n = reaction rate coefficient, $(M/L^3)^{n-1}/T$ C = concentration of constituent, $(M/L^3)^n$ n = reaction order (e.g. for second order $n = 2$)
Natural decay	$r_d = -k_d N$	r_d = rate of decay, no./T k_d = first order reaction rate coefficient, $1/T$ N = amount of organisms remaining, no.
Separation processes		
Gas absorption/ desorption	$r_{ab} = k_{ab}\frac{A}{V}(C_s - C)$ $r_{de} = -k_{de}\frac{A}{V}(C - C_s)$	r_{ab} = rate of absorption, M/L^3T r_{de} = rate of desorption, M/L^3T k_{ab} = coefficient of absorption, L/T k_{de} = coefficient of desorption, L/T A = area, L^2 V = volume, L^3 C_s = saturation concentration of constituent in liquid, M/L^3 (see Eq. 2–49) C = concentration of constituent in liquid, M/L^3
Sedimentation	$r_s = \frac{v_s}{H}(SS)$	r_s = rate of sedimentation, $1/T$ v_s = settling velocity, L/T H = depth, L SS = settleable solids, L^3/L^3
Volatilization	$r_v = -k_v(C - C_s)$	r_v = rate of volatilization per unit time per unit volume, M/L^3T k_v = volatilization constant, $1/T$ C = concentration of constituent in liquid, M/L^3 C_s = saturation concentration of constituent in liquid, M/L^3 (see Eq. 2–49)

[a] Adapted in part from Ambrose et al. (1988), Tchobanoglous et al. (2003).

The rate expression given by Eq. (1–53) is known as a retarded first order rate expression because the rate constant changes with distance or time, as illustrated on Fig. 1–8, or with the degree of treatment which, in turn, can be related to distance or time. The term, r_t, in the denominator is the retardation factor. In wastewater treatment applications, the exponent n in Eq. (1–53) is related to the particle size distribution (see Fig, 1–8). For example, if all of the particles are the same size and composition, the value of the exponent n is equal to one, and the retardation factor r_t is equal to zero. The retarded rate expression is also applied to the removal of organic matter from mixtures where the

Figure 1–7

Rate of reaction versus concentration for a saturation type expression. Beyond about 20 mg/L the rate of reaction is essentially zero order.



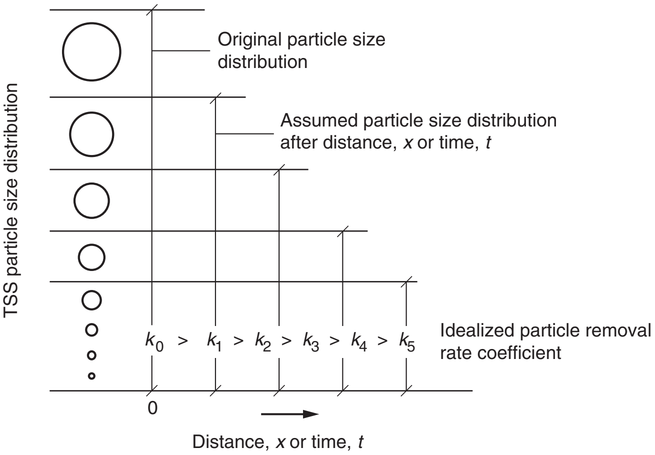
biodegradability of the individual constituents comprising the organic matter is different (Tchobanoglous et al., 2003).

Separation Processes. Unlike conversion processes where constituents are removed through transformation, separation processes bring about removal by the physical transfer of constituents from a diluted state to a concentrated state. Separation processes exploit particular characteristics of constituents to bring about removal. The removal methods for particulate and dissolved constituents are considered below.

Particulate Constituents. The removal of particulate constituents depends on the nature and size of the constituent but is brought about primarily through the application of gravity and pressure forces. For example, large coarse solids in wastewater, greater than about 6 mm (0.25 in.), are removed by screening (i.e., sieving). The force of gravity is used to bring about the separation (removal) of grit and other settleable material. Very light constituents such as oils and grease are also removed by the force of gravity and by flotation because their density is less that that of water. Smaller particulate constituents that cannot

Figure 1–8

Definition sketch to illustrate the change that can occur in the removal rate coefficient with distance or time when an influent wastewater with a particle size distribution such as shown is applied to a granular medium filter or a constructed wetland.



be removed by gravity can be removed by filtration, in which wastewater is passed through a filtering medium by the application of force in the form of pressure.

Dissolved Constituents. Dissolved constituents can also be removed from water by concentration on a solid surface (e.g., activated carbon adsorption and ion exchange). An important consideration in the modeling of adsorption-type separation processes is that because the reaction is assumed to be instantaneous after the constituent reaches the relevant surface, the reaction rate is controlled by the transport of the constituent to the point of reaction. The transfer of mass by molecular diffusion in stationary systems can be represented by the following expression, known as Fick's first law:

$$r = -D_m \frac{\partial C}{\partial x} \tag{1–54}$$

where r = rate of mass transfer per unit area per unit time, $\mathrm{ML^{-2}T^{-1}}$
D_m = coefficient of molecular diffusion in the x direction, $\mathrm{L^2T^{-1}}$
C = concentration of constituent being transferred, $\mathrm{ML^{-3}}$
x = distance, L

The negative sign in Eq. (1–54) is used to denote the fact that diffusion takes place in the direction of decreasing concentration. Also, it should be noted that the concentration gradient ($\partial C/\partial x$) is assumed to be constant. In the chemical engineering literature the symbol J is used to denote mass transfer in concentration units whereas the symbol N is used to denote the transfer of mass expressed as moles.

The coefficient of molecular diffusion is related to the frictional coefficient of a particle as given by the Stokes-Einstein law of diffusion. For spherical particles the coefficient of diffusion is given by the following expression (Shaw, 1966).

$$D = \frac{kT}{6\pi\mu r_p} = \frac{RT}{6\pi\mu r_p N} \tag{1–55}$$

where D = coefficient of diffusion, $\mathrm{m^2/s}$
k = Boltzmann constant 1.3805×10^{-23} J/K
T = temperature, K = 273.15 + °C
R = universal gas law constant, 8.3145 J/mole·K
μ = dynamic viscosity, $\mathrm{N{\cdot}s/m^2}$
r_p = radius of particle, m
N = Avogadro's number, 6.02×10^{23} molecules/g·mole

The terms in the denominator in Eq. (1–55) correspond to the coefficient of friction for a particle as defined by Stokes law. The coefficient of diffusion for a particle with a radius of 10^{-7} m (0.01 μm), which corresponds to the size of the smallest bacteria, for the following conditions is:

$T = 20°\mathrm{C}$

$\mu = 1.002 \times 10^{-3}\ \mathrm{N{\cdot}s/m^2}$

$$D = \frac{kT}{6\pi\mu r_p A} = \frac{(8.3145\ \mathrm{J/mole{\cdot}K})(293\ \mathrm{K})}{6(3.14)(1.002 \times 10^{-3}\ \mathrm{N{\cdot}s/m^2})(10^{-7}\mathrm{m})(6.02 \times 10^{23}/\mathrm{mole})}$$

$$= 21.43 \times 10^{-13}\ \mathrm{m^2/s} = 2.143 \times 10^{-8}\ \mathrm{cm^2/s}$$

From the above computation it is easy to see that as the particles get smaller the coefficient of molecular diffusion increases. Depending on the fluid regime, the coefficient of molecular diffusion in Eq. (1–55) will be replaced by the turbulent coefficient of dispersion, as described further in Appendix I.

Many important separation processes used in wastewater treatment involve mass transfer across the gas-liquid interface (e.g., aeration) or the removal of undesirable constituents (e.g., stripping). For example, the rate of flux of a slightly soluble gas from the gas to the liquid phase (liquid film controls transfer rate; see discussion in Sec. 5–10), based on Fick's first law, can be approximated as follows:

$$r = K_L(C_s - C_t) \tag{1–56}$$

where r = rate of mass transferred per unit area per unit time, $ML^{-2}T^{-1}$
K_L = overall liquid mass transfer coefficient, LT^{-1}
C_s = concentration in equilibrium with gas as given by Henry's Law, ML^{-3}
C_t = concentration in liquid bulk phase at time t, ML^{-3}

The mass transfer coefficient depends on the characteristics of the wastewater and the treatment process design and is, therefore, unique for each situation. The application of mass transfer for aeration is considered in Secs. 5–10 and 5–11. Other treatment processes that depend on mass transfer including carbon adsorption, gas stripping, and ion exchange are considered in Chap. 11.

Analysis of Reaction Rate Coefficients

Typically, reaction rate coefficients are determined using the results obtained from batch experiments (i.e., no inflow or outflow), from continuous flow experiments, and from pilot and field scale experiments. Using the data from batch experiments, the coefficients can be determined using a variety of methods including (1) the method of integration and (2) the differential method (see Table 1–12).

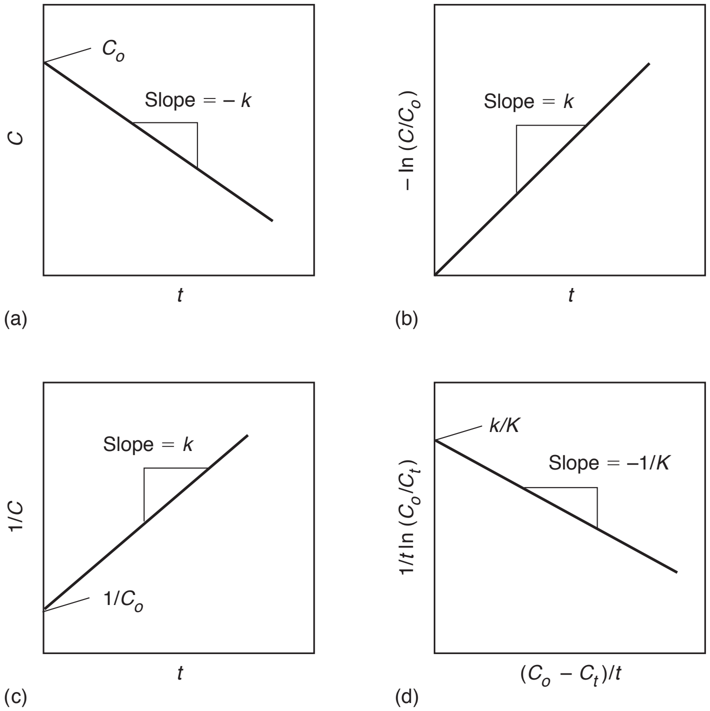
Table 1–12

Integration and differential methods used to determine reaction rate coefficients

Rate expression		Method used to determine the reaction rate coefficient
Integration method		
Zero order reaction	Integrated form	
$r_c = \frac{dC}{dt} = -k$	$C - C_o = -kt$	Graphically, by plotting C versus t [see Fig. 1–9(a)]
First order reaction		
$r_c = \frac{dC}{dt} = -kC$	$\ln\frac{C}{C_o} = -kt$	Graphically, by plotting $-\ln(C/C_o)$ versus t [see Fig. 1–9(b)]
Second order reaction		
$r_c = \frac{dC}{dt} = -kC^2$	$\frac{1}{C} - \frac{1}{C_o} = kt$	Graphically, by plotting $1/C$ versus t [see Fig. 1–9(c)]
Saturation reaction		
$r_c = \frac{dC}{dt} = -\frac{kC}{K + C}$	$kt = K\ln\frac{C_o}{C_t} + (C_o - C_t)$	Graphically, by plotting $1/t \ln(C_o/C_t)$ versus $(C_o - C_t)/t$ [see Fig. 1–9(d)]
Differential method		
$r_c = \frac{dC}{dt} = -kC^n$		Analytically, by solving for $n = \frac{\log[-d(C_1/dt)] - \log[-(dC_2/dt)]}{\log(C_1) - \log(C_2)}$ Once the order of the reaction is known, the reaction rate coefficient can be determined by substitution

Figure 1–9

Graphical analysis for the determination of reaction order and reaction-rate coefficients: (a) zero-order reaction, (b) first-order reaction, (c) second-order reaction, and (d) saturation type reaction.



As summarized in Table 1–12, the method of integration involves the substitution of the measured data on the amount of reactant remaining at various times into the integrated form of the rate expression. Plots of the integrated forms of the reaction rate expressions used to determine the reaction rate coefficients are shown on Fig. 1–9. In the differential method, where the order of the reaction is unknown, the concentrations remaining at two different times are used to solve the differential form of the rate expression for the order of the reaction. Once the reaction order is known, the reaction rate coefficient is determined by substitution using the test data. The application of these two methods is illustrated in Example 1–2.

EXAMPLE 1–2 **Determination of the Reaction Order and the Reaction Rate Coefficient** Given the following set of data obtained using a batch reactor [see Fig. 1–4(a)], determine the order of the reaction and the reaction rate coefficient using the integration and differential methods.

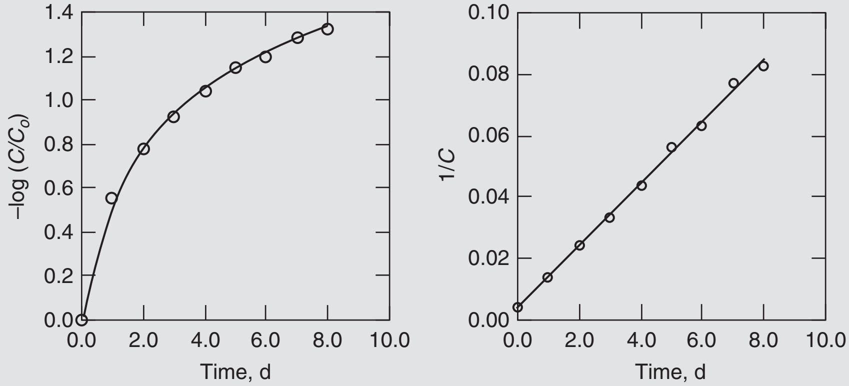
Time, d	Concentration, C, mole/L
0	250
1	70
2	42
3	30
4	23
5	18
6	16
7	13
8	12

Solution—Part 1 Integration Method

1. Determine the reaction order and the reaction rate constant using the integration method. Develop the data needed to plot the experimental data functionally, assuming the reaction is either first or second order.

Time, d	C, mole/L	$-\log(C/C_o)$	$1/C$
0	250	0	0.004
1	70	0.553	0.014
2	42	0.775	0.024
3	30	0.921	0.033
4	23	1.036	0.044
5	18	1.143	0.056
6	16	1.194	0.063
7	13	1.284	0.077
8	12	1.319	0.083

2. To determine whether the reaction is first or second order, plot $-\log(C/C_o)$ and $1/C$ versus t as shown below. Because the plot of $1/C$ versus t is a straight line, the reaction is second order with respect to the concentration C.



3. Determine the reaction rate coefficient.

Slope $= k$

The slope from the plot $= \dfrac{0.084 - 0.024}{8\text{d} - 2\text{d}} = 0.010/\text{d}$

$k = 0.010/\text{d}$

Solution—Part 2 Differential Method

1. Determine the reaction order and the reaction rate constant using the differential method.

$$n = \frac{\log[-(dC_1/dt)] - \log[-(dC_2/dt)]}{\log(C_1) - \log(C_2)}$$

a. Use the experimental data obtained at day 3 and 6.

Time, d	C, mole/L	$\frac{C_{t+1} - C_{t-1}}{2}$	$\approx \frac{dC_t}{dt}$
0	250		
1	70		
2	42		
3	30	(23 − 42)/2	−9.5
4	23		
5	18		
6	16	(13 − 18)/2	−2.5
7	13		
8	12		

b. Substitute and solve for n.

$$n = \frac{\log(9.5) - \log(2.5)}{\log(30.0) - \log(16.0)} = 2.07 \quad \text{use } n = 2$$

c. The reaction is second order.

d. The reaction rate constant is

$$\frac{1}{C} - \frac{1}{C_o} = kt$$

$$\frac{1}{42} - \frac{1}{250} = k(2)$$

$$k = 0.0103/\text{d}, \quad \text{use } k = 0.010/\text{d}$$

In the applications described above, the initial concentration of a constituent is generally known. However, in the conventional BOD test, described in Chap. 2, both UBOD (ultimate biological oxygen demand) and k_1 are unknown. To determine these values, the usual procedure is to run a series of BOD measurements with time. Using these measurements, the UBOD and k_1 values can be determined using a number of methods including the method of least-squares, the method of moments, the daily-difference method, the rapid-ratio method, the Thomas method, and the Fujimoto method, as discussed in Sec. 2–6 in Chap. 2.

1–11 INTRODUCTION TO TREATMENT PROCESS MODELING

In wastewater treatment, the chemical and biological reactions that are needed to bring about the treatment of wastewater are carried out in the reactors described previously in Sec. 1–7. Treatment process kinetics involves the coupling of reactors and reaction rates to determine treatment process performance. In this section, the focus is on modeling the reactions that occur in the reactors used for wastewater treatment. The reactors considered include (1) batch, (2) complete-mix, (3) complete-mix reactors in series, (4) ideal plug-flow, (5) ideal plug-flow with retarded reaction rate, and (6) plug-flow with axial dispersion.

Batch Reactor with Reaction

The derivation of the materials mass balance equation for a batch reactor [see Fig. 1–7 (a)] for a reactive constituent is written as follows:

Accumulation = inflow − outflow + generation

$$\frac{dC}{dt}V = QC_o - QC + r_C V \tag{1–57}$$

Because $Q = 0$ the resulting equation for a batch reactor is

$$\frac{dC}{dt} = r_C \tag{1–58}$$

Before proceeding further, it will be instructive to explore the difference between the rate of change term that appears as part of the accumulation term and the rate of generation or utilization or decay term. In general, these terms are not equal, except in the special case of a batch reactor in which there is no inflow or outflow from the control volume. The key point to remember is that when flow is not occurring, the concentration per unit volume is changing according to the applicable rate expression. On the other hand, when flow is occurring, the concentration in the reactor is also being modified by the inflow and outflow from the reactor.

If the rate of reaction is defined as first order (i.e., $r_c = -kC$), integrating between the limits $C = C_o$ and $C = C$ and $t = 0$ and $t = t$ yields

$$\int_{C=C_o}^{C=C} \frac{dC}{C} = -k\int_{t=0}^{t=t} dt = kt \tag{1–59}$$

The resulting expression is

$$\frac{C}{C_o} = e^{-kt} \tag{1–60}$$

Equation (1–60) is the same as the BOD equation Eq. (2–59) considered subsequently in Chap. 2.

Complete-Mix Reactor with Reaction

The general form of the mass-balance equation for a complete-mix reactor as shown on Figs. 1–4(b) and 1–5(a-1), in which the liquid in the reactor is mixed completely, follows:

Accumulation = inflow − outflow + generation

$$\frac{dC}{dt}V = QC_o - QC + r_c V \tag{1–61}$$

Assuming first order removal kinetics ($r_c = -kC$), Eq. (1–61) can be rearranged and written as follows

$$C' + \beta C = \frac{Q}{V}C_o \tag{1–62}$$

where $C' = dC/dt$

$\beta = k + Q/V$

To solve Eq. (1–62) both sides of the expression are multiplied by the integrating factor $e^{\beta t}$:

$$e^{\beta t}(C' + \beta C) = \frac{Q}{V}C_o e^{\beta t} \tag{1–63}$$

The left side of the above expression can be written as a differential as follows:

$$(Ce^{\beta t})' = \frac{Q}{V}C_o e^{\beta t} \tag{1–64}$$

The differential sign is removed by integrating the above expression

$$Ce^{\beta t} = \frac{Q}{V}C_o \int e^{\beta t} \tag{1–65}$$

Integration of Eq. (1–65) yields

$$Ce^{\beta t} = \frac{Q}{V}\frac{C_o}{\beta}e^{\beta t} + K \tag{1–66}$$

Dividing by $e^{\beta t}$ yields

$$C = \frac{Q}{V}\frac{C_o}{\beta} + Ke^{-\beta t} \tag{1–67}$$

But when $t = 0$, $C = C_o$ and K is equal to

$$K = C_o - \frac{Q}{V}\frac{C_o}{\beta} \tag{1–68}$$

Substituting for K in Eq. (1–68) and simplifying yields the following expression, which is the non-steady state solution of Eq. (1–61):

$$C = \frac{Q}{V}\frac{C_o}{\beta}(1 - e^{-\beta t}) + C_o e^{-\beta t} \tag{1–69}$$

The solution to Eq. (1–61) under steady-state conditions (i.e., the rate accumulation term is equal to zero [$dC/dt = 0$)] is given below.

$$C = \frac{C_o}{[1 + k(V/Q)]} = \frac{C_o}{(1 + k\tau)} \tag{1–70}$$

It should also be noted that when $t \rightarrow \infty$, Eq. (1–69) becomes the same as Eq. (1–67).

Complete-Mix Reactors in Series with Reaction

When complete-mix reactors are used in series, the steady-state solution is of concern as it is used for design. Two approaches are presented for the analysis of reactors in series: (1) analytical and (2) graphical. The graphical approach also applies to cascades of reactors, as discussed for mass transfer equilibria.

Analytical Solution. The steady-state form of the mass balance for the second reactor of the three reactor system (see Fig. 1–10), is given by

Accumulation = inflow − outflow + generation

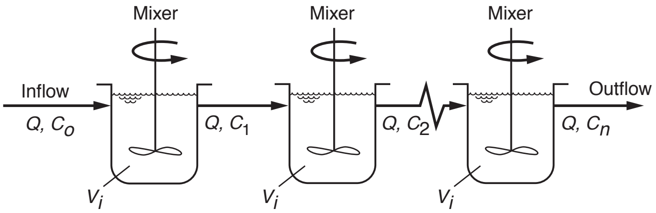
$$\frac{dC_2}{dt}\frac{V}{2} = 0 = QC_1 - QC_2 + r_c\frac{V}{2} \tag{1–71}$$

Assuming first order removal kinetics ($r_c = -kC_2$), Eq. (1–71) can be rearranged and solved for C_2 yielding

$$C_2 = \frac{C_1}{[1 + (kV/2Q)]} \tag{1–72}$$

Figure 1–10

Definition sketch for the analysis of complete-reactors in series.



But from Eq. (1–70), the value of C_1 is equal to

$$C_2 = \frac{C_o}{[1 + (kV/2Q)]} \tag{1–73}$$

Combining the above two expressions yields

$$C_2 = \frac{C_o}{[1 + (kV/2Q)]^2} \tag{1–74}$$

For n reactors in series the corresponding expression is

$$C_n = \frac{C_o}{[1 + (kV/nQ)]^n} = \frac{C_o}{[1 + (k/2\tau)]^n} \tag{1–75}$$

For example, consider three 1000 m^3 complete-mix reactors in series with a flowrate of 100 m^3/d and first order kinetics with a reaction rate coefficient value of $k = 0.1/d$. Using Eq. (1–75), the effluent concentration from the third reactor, assuming the starting concentration was 100 mg/L is

$$C_3 = \frac{C_o}{[1 + (kV/3Q)]^3} = \frac{100}{\left\{1 + \left[\frac{(0.1/\text{d})(3000\ \text{m}^3)}{(3 \times 100\ \text{m}^3/\text{d})}\right]\right\}^3} = 12.5\ \text{mg/L}$$

Solving Eq. (1–75) for the detention time yields

$$\tau = \frac{V}{Q} = \left[\frac{1}{(C_n/C_o)^{1/n}} - 1\right]\left(\frac{n}{k}\right) \quad \text{or} \quad \tau = \left[\left(\frac{C_o}{C_n}\right)^{\frac{1}{n}} - 1\right]\left(\frac{n}{k}\right) \tag{1–76}$$

Graphical Solution. The graphical solution for 3 (or for n) reactors in series is obtained as follows. For a single reactor, Eq (1–71) can be written as follows:

Accumulation = inflow − outflow + generation

$$0 = QC_o - QC_1 - r_cV \tag{1–77}$$

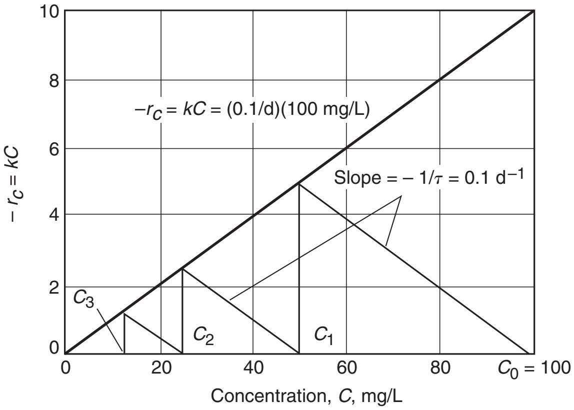
The first step in developing a graphical solution is to draw a graph of r_c versus C (see Fig. 1–10). To plot r_c versus C, Eq. (1–77) is now rewritten as follows:

$$r_c = -\frac{Q}{V}(C_1 - C_o) = -\frac{1}{\tau}(C_1 - C_o) \tag{1–78}$$

The above equation can be represented graphically by a straight line drawn from the point $r_c = 0$ and $C = 100$ mg/L with a slope of $-1/\tau$. The line drawn will intersect the graph of r_c versus C at $r_c = 5.0$ and $C_1 = 50$ mg/L as shown on Fig. 1–11. The value $C_1 = 50$ mg/L is the solution of Eq. (1–78) for a single rector for the stated conditions used to derive the

Figure 1–11

Graphical analysis used to determine the effluent concentration from a series of complete-mix reactors.



analytical solution, presented previously. If the procedure is repeated for a second and a third reactor, the final effluent concentration from the third reactor is found to be 12.5 mg/L, which is the same as the analytical solution determined above. The graphical approach is especially useful in solving phase separation processes as described previously in Sec. 1–9. To use the graphical procedure, the reaction rate coefficient must be a function of a single variable (e.g., *C*). Use of both the analytical and graphical methods of analysis are illustrated in Example 1–3. Additional details on the graphical solution of design equations may be found in Eldridge and Piret (1950) and Smith (1981).

EXAMPLE 1–3 **Analysis of Reactors in Series Using Both an Analytical and Graphical Approach** Two 1000 m^3 complete-mix reactors are to be used in series with a flowrate of 500 m^3/d and second order kinetics with a *k* value of 0.01/d. Determine the effluent concentration from the second reactor assuming the starting concentration is 100 mg/L.

Solution

1. Determine the effluent concentration from the series of two complete-mix reactors analytically.
   a. At steady-state, the mass balance for the first complete-mix reactor is:

$$0 = QC_o - QC_1 - k_c C_1^2 V$$

Substituting the given values and solving for C_1 yields

$$0 = \frac{(500\ m^3/d)}{1000\ m^3}(100\ mg/L) - \frac{(500\ m^3/d)}{1000\ m^3}C_1 - (0.01/d)C_1^2$$

$$C_1 = 50\ mg/L$$

   b. At steady-state, the mass balance for the second complete-mix reactor is

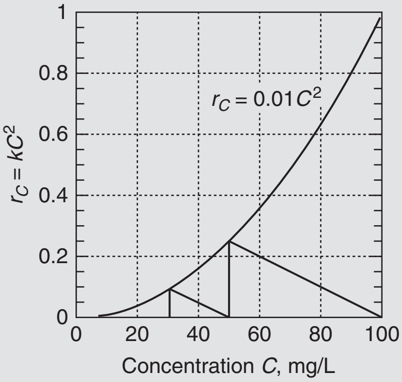
$$0 = QC_1 - QC_2 - kC_2^2 V$$

Substituting the given values and solving for C_2 yields

$$0 = \frac{(500\ m^3/d)}{1000\ m^3}(50\ mg/L) - \frac{(500\ m^3/d)}{1000\ m^3}C_2 - (0.01/d)C_2^2$$

$$C_2 = 30\ mg/L$$

2. Determine the effluent concentration from the series of two complete-mix reactors graphically.
   a. Prepare a plot of r_c versus C as shown below:



   b. Linearize the mass balance equation from Step 1a

$$r_c = -\frac{Q}{V}(C_1 - C_o)$$

   c. On the plot prepared above draw a straight line from the point $r_c = 0$ and $C =$ 100 mg/L with a slope of $-Q/V$ equal to -0.5/d [$-(500\ \text{m}^3/\text{d})/1000\ \text{m}^3$]. The line drawn intersects the graph of r_c versus C at $r_c = 0.25$ and $C_1 = 50$ mg/L. Repeating the above procedure the final effluent concentration from the second reactor is found to be 30 mg/L, which is the same as the analytical solution determined in Step 1.

Ideal Plug-Flow Reactor with Reaction

The derivation of the materials balance equation for an ideal plug-flow reactor, in which the concentration, C, of the constituent is uniformly distributed across the cross-sectional area of the control volume, and there is no longitudinal dispersion, can be illustrated by considering the differential volume element shown on Fig. 1–6. For the differential volume element ΔV shown on Fig. 1–6, the materials balance on a reactive constituent C is written as follows:

Accumulation = inflow − outflow + generation

$$\frac{\partial C}{\partial t}\Delta V = QC|_x - QC|_{x+\Delta x} + r_c\Delta V \tag{1–79}$$

Accumulation = inflow − outflow + generation

where $\partial C/\partial t$ = change in constituent concentration with time, $\text{ML}^{-3}\text{T}^{-1}$ ($\text{g/m}^3\cdot\text{s}$)
C = constituent concentration, ML^{-3} (g/m^3)
ΔV = differential volume element, L^3 (m^3)
Q = volumetric flowrate, L^3T^{-1} (m^3/s)
r_c = reaction rate for constituent C, $\text{ML}^{-3}\text{T}^{-1}$, ($\text{g/m}^3\cdot\text{s}$)

Substituting the differential form for the term $QC|_{x+\Delta x}$ in Eq. (1–79) results in

$$\frac{\partial C}{\partial t}\Delta V = QC - Q\left(C + \frac{\Delta C}{\Delta x}\Delta x\right) + r_c\Delta V \tag{1–80}$$

Substituting $A\Delta x$ for ΔV and dividing by A and Δx yields

$$\frac{\partial C}{\partial t} = -\frac{Q}{A}\frac{\Delta C}{\Delta x} + r_c \tag{1–81}$$

Taking the limit as Δx approaches zero yields

$$\frac{\partial C}{\partial t} = -\frac{Q}{A}\frac{\partial C}{\partial x} + r_c \tag{1–82}$$

If steady-state conditions are assumed ($\partial C/\partial t = 0$) and the rate of reaction is defined as $r_c = -kC^n$, integrating between the limits $C = C_o$ and $C = C$ and $x = 0$ and $x = L$ yields

$$\int_{C_o}^{C}\frac{dC}{C^n} = -k\frac{A}{Q}\int_0^L dx = -k\frac{AL}{Q} = -k\frac{V}{Q} = -k\tau \tag{1–83}$$

Equation (1–83) is the steady-state solution to the materials balance equation for a plug-flow reactor without dispersion. If it is assumed that n is equal to one, Eq. (1–83) becomes

$$\frac{C}{C_o} = e^{-k\tau} \tag{1–84}$$

which is equivalent to Eq. 1–60, derived previously for the batch reactor.

Comparison of Complete-Mix and Plug-Flow Reactors with Reaction

The combined effect of reactor type (e.g., complete-mix versus plug-flow) and kinetics is also of interest. The total volume required for various removal efficiencies for first order kinetics, using 1, 2, 4, 6, 8, or 10 reactors in series is reported in Table 1–13 and shown graphically on Fig. 1–12. The corresponding volume required for a plug-flow reactor is also reported in Table 1–13 As shown in Table 1–13, as the number of reactors in series is increased, the total reactor volume required approaches that of a plug-flow reactor. A comparison of reactor types for second-order kinetics is examined in Example 1–4.

It should be noted, however, that for zero order kinetics the volume of the two reactors will be the same. It is also important to note that biological processes do not obey the results presented in Table 1–13 (i.e., plug-flow is more efficient than complete-mix) because biological process are modeled using BOD and COD which includes microbial products in addition to any residual substrate. As a result, the volumes required for the two reactors will be the same. The use of a plug-flow reactor, or mixed cells in series, is often favored to help control the growth of filamentous organisms (see discussion in Chap. 7).

Table 1–13

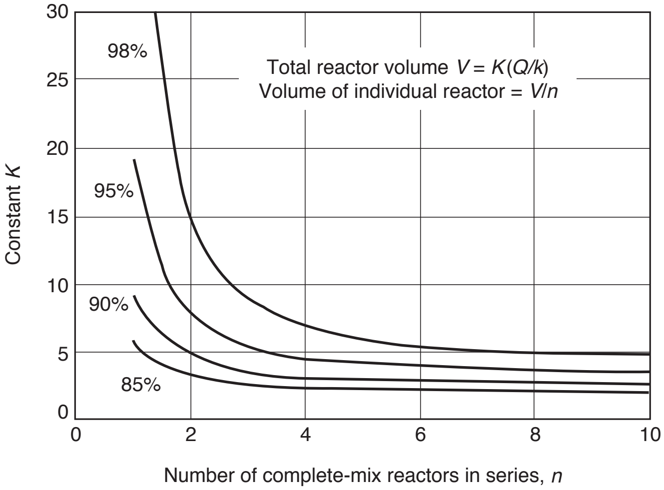
Required reactor volumes expressed in terms of Q/k for a series of complete-mix reactors and for a plug-flow reactor for various removal efficiencies for first-order kinetics[a]

	Reactor volume $V = K(Q/k)$			
	Removal efficiency, %			
No of reactors in series	85	90	95	98
1	5.67	9.00	19.00	49.00
2	3.16	4.32	6.94	12.14
4	2.43	3.11	4.46	6.64
6	2.23	2.81	3.89	5.52
8	2.14	2.67	3.63	5.05
10	2.09	2.59	3.49	4.79
Plug flow	1.90	2.30	3.00	3.91

[a] Volume of individual reactors equals value in table divided by the number of reactors in series.

Figure 1–12

Definition sketch for the total volume required versus the number of complete-mix reactors in series for various removal efficiencies. The K value on the vertical axis is multiplied by the flowrate and divided by the reaction coefficient to obtain the total volume required. The volume of an individual reactor is equal to the total volume divided by the number of reactors in series.



EXAMPLE 1–4 **Comparison of Required Reactor Volumes for Second-order Kinetics**

Assuming that second-order kinetics apply ($r_c = -kC^2$), compare the required volume of a complete-mix reactor to the volume of a plug-flow reactor to achieve a 90 percent reduction in the concentration ($C_o = 1$ and $C_e = 0.1$).

Solution

1. Compute the required volume for a complete-mix reactor in terms of Q/k.
   a. At steady state, a mass balance for a complete-mix reactor yields

$$0 = QC_o - QC_e - kC_e^2V$$

   b. Simplify and substitute the given data.

$$V = \frac{Q}{k}\left(\frac{C_o - C_e}{C_e^2}\right) = \frac{Q}{k}\frac{1 - 0.1}{(0.1)^2} = 90\frac{Q}{k}$$

2. Compute the required volume for a plug-flow reactor in terms of Q/k.
   a. At steady state, a mass balance for a plug-flow reactor yields

$$0 = -Q\frac{dC}{dx}dx + Adx(-kC^2)$$

   b. The integrated form of the steady-state equation is

$$V = \frac{Q}{k}\int_{C_o}^{C_e}\frac{dC}{C^2} = \frac{Q}{k}\frac{1}{C}\bigg|_{C_o}^{C_e} = \frac{Q}{k}\left(\frac{1}{C_e} - \frac{1}{C_o}\right)$$

   c. Substituting the given concentration values yields

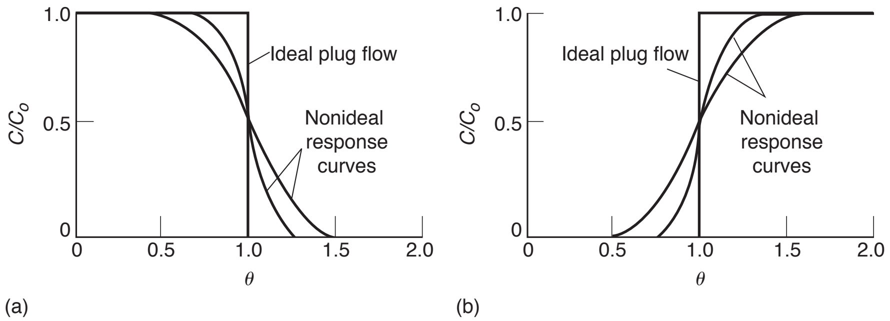
$$V = \frac{Q}{k}\left(\frac{1}{0.1} - \frac{1}{1}\right) = \frac{9Q}{k}$$

3. Determine the volume ratio.

$$\frac{V_{\text{CMR}}}{V_{\text{PFR}}} = \frac{(90\,Q/k)}{(9\,Q/k)} = 10$$

Figure 1–13

Theoretical and generalized nonideal response curves for a plug-flow reactor with axial dispersion.



Plug-Flow Reactor with Axial Dispersion and Reaction

In most full-scale plug-flow reactors, the flow usually is non-ideal because of entrance and exit flow disturbances, axial dispersion, and dispersion caused by advection (see Appendix I for an expanded discussion of dispersion and advection). Depending on the magnitude of these effects, the ideal effluent-tracer curves may look like the curves shown on Fig. 1–13. Using first order removal kinetics, Wehner and Wilhelm (1958) have developed a solution for a plug-flow reactor with dispersion numbers varying from complete-mix ($d = \infty$) to ideal plug-flow ($d = 0$). The equation developed by Wehner and Wilhelm is as follows:

$$\frac{C}{C_o} = \frac{4a\exp(1/2\mathrm{d})}{(1 + a)^2\exp(a/2\mathrm{d}) - (1 - a)^2\exp(-a/2\mathrm{d})} \qquad (1\text{–}85)$$

where C = effluent concentration, ML^{-3}

C_o = influent concentration, ML^{-3}

$a = \sqrt{1 + 4k\tau\mathrm{d}}$

d = dispersion factor = D/vL [see Eq. (I–9), Appendix I]

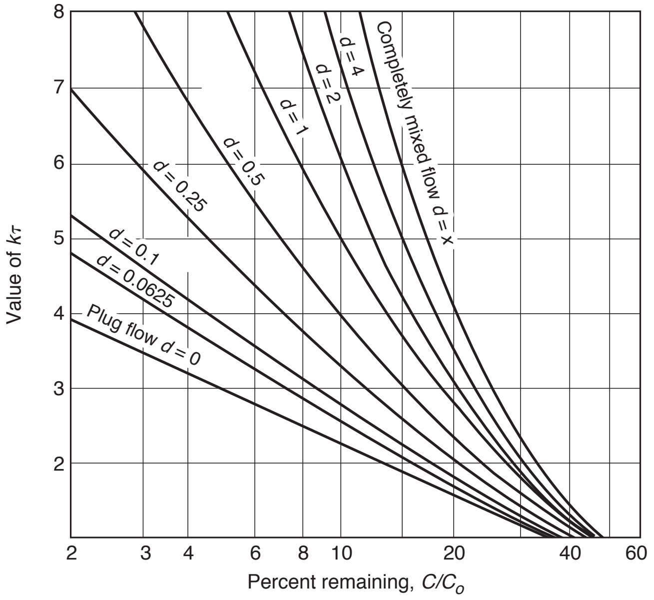
k = first order reaction constant, T^{-1}, (1/h)

τ = hydraulic detention time, V/Q, T, (h)

To facilitate the use of Eq. (1–85) for the design of treatment processes such as stabilization ponds and natural systems, Thirumurthi (1969) developed Fig. 1–14, in which the

Figure 1–14

Values of kt in the Wehner and Wilhelm equation (Eq. 1–85) versus percent remaining for various dispersion factors and first order kinetics for a plug flow reactor. (Adapted from Thirumurthi, 1969.)



term $k\tau$ is plotted against C/C_o for dispersion factors varying from zero for an ideal plug-flow reactor to infinity for a complete-mix reactor. The application of Fig. 1–14 is illustrated in the following example.

EXAMPLE 1–5 **Comparison of the Performance of a Treatment Process Occurring in a Plug-flow Reactor Without and with Axial Dispersion** A treatment process reactor was designed assuming ideal plug-flow with a first order BOD removal rate constant of 0.5/d at 20°C and a detention time of 5 d. Once in operation, a considerable amount of axial dispersion was observed in the reactor. What effect will the observed axial dispersion have on the performance of the treatment process? The dispersion factor for the reactor, *d*, has been estimated to be about 0.5. Determine how much longer the detention time must be for a reactor with a dispersion factor of 0.5 to achieve the same degree of treatment as expected initially with the ideal plug-flow reactor.

Solution

1. Estimate the percentage removal for an ideal plug-flow reactor using Eq. (1–84).
    a. The BOD remaining is:

$$\frac{C}{C_o} = e^{-k\tau}$$

$$\frac{C}{C_o} = e^{-0.5\times 5} = 0.082 = 8.2\%$$

    b. The percentage removal is

Percentage removal $100 - 8.2 = 91.8\%$

2. Determine the percentage removal for the reactor using Fig. 1–14.
    a. The value of $k\tau$ equals

$$k\tau = (0.5/\text{d} \times 5\ \text{d}) = 2.5$$

    b. The percent remaining from Fig 1–14 is equal to

$$C/C_o = 0.20 = 20\%$$

Percentage removal $100 - 20 = 80.0\%$

3. Determine the required detention time to achieve 91.8 percent removal
    a. The value of $k\tau$ from Fig. 1–14 for a C/C_o value of 8.2% is 4.6.
    b. The required detention time is

$$k\tau = 4.6$$

$$\tau = 4.6/0.5 = 9.2\ \text{d}$$

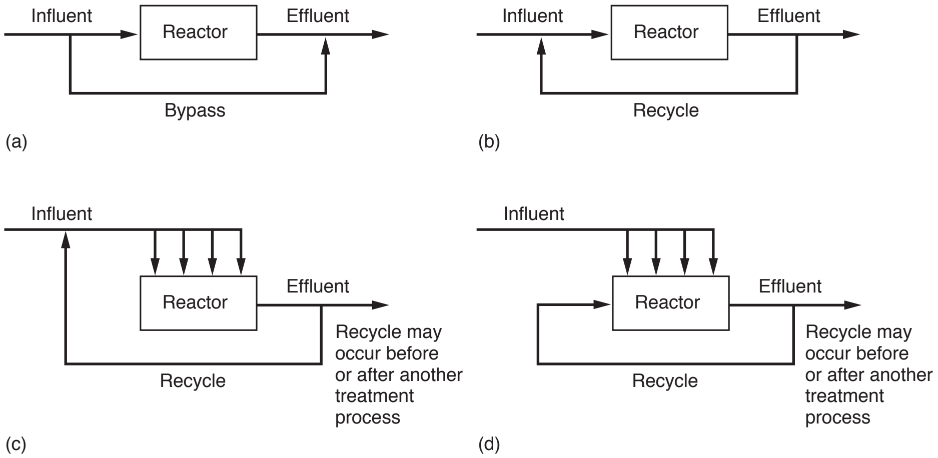
Comment Clearly, axial dispersion can affect the predicted performance of a treatment process designed to function as an ideal plug-flow reactor. Because of axial dispersion and temperature effects, the actual performance of the treatment process will generally be less than expected.

Other Reactor Flow Regimes and Reactor Combinations

In the previous discussions of complete-mix and plug-flow reactors, a single-pass straight-through flow pattern has been used for the purpose of analyses. In practice, other flow regimes and reactor combinations are also used. Some of the more common alternative

Figure 1–15

Flow regimes commonly used in the treatment of wastewater: (a) direct input with bypass flow (plug-flow or complete-mix reactor), (b) direct input with recycle flow (plug-flow or complete-mix reactor), (c) stepped input with recycle (recycle flow mixed with influent, recycle type 1), and (d) stepped input with recycle (recycle flow introduced at influent end of reactor, recycle type 2).

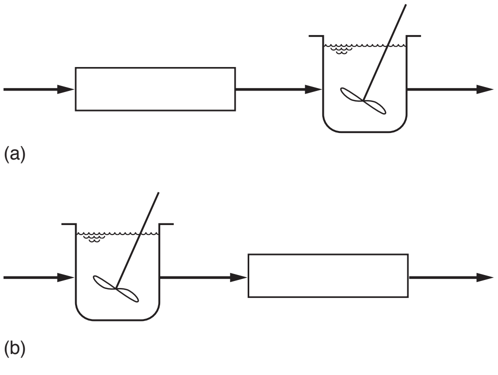


flow regimes are shown schematically on Fig. 1–15. The flow regime shown on Fig. 1–15(a) is used to achieve intermediate levels of treatment by blending various amounts of treated and untreated wastewater. The flow regime used on Fig. 1–15(b) is often adopted to achieve greater process control and will be considered specifically in Chaps. 9 and 10 which deal with biological wastewater treatment. The flow regime shown on Fig. 1–15(c) is used to reduce the loading rate applied to the process. On Fig. 1–15(d), the return flow is not mixed with the influent, but is introduced at the entrance of the reactor to achieve greater initial dilution of the wastewater to be treated. Each of these hydraulic regimes is considered further in the following chapters.

Among the numerous types of reactor combinations that are possible and that have been used, two combinations using a plug-flow reactor and a complete-mix reactor are shown on Fig. 1–16. In the arrangement shown on Fig. 1–16(a), complete-mixing takes place second; in the arrangement shown on Fig. 1–16(b), it occurs first. If no reaction takes place and the reactors are used only to equalize temperature, for example, the result will be identical. If a reaction is occurring, however, the product yields of the two reactor systems can be different. The use of such hybrid reactor systems will depend on the specific product requirements. Additional details on the analysis of such processes may be found in Denbigh and Turner (1965), Kramer and Westererp (1963), and Levenspiel (1972).

Figure 1–16

Hybrid reactor systems: (a) plug-flow reactor followed by complete-mix reactor and (b) complete-mix reactor followed by plug-flow reactor.



PROBLEMS AND DISCUSSION TOPICS

1–1 A water storage tank receives a constant feed rate of 0.2 m^3/s and the demand varies according to the relationship $0.2[1 - \cos \pi t/(43{,}200)\ m^3/s]$. The tank is cylindrical with a cross-sectional area of 1000 m^2. If the depth at $t = 0$ is 5 m, plot the water depth as a function of time.

1–2 Solve Problem 1–1 assuming the feed rate is 0.33 m^3/s and that the storage tank is a square with a cross-sectional area of 1600 m^2.

1–3 A large tank having a floor area of 1000 m^2 and a sidewall depth of 10 m is used as an equalization reservoir. Flow out of the basin is 0.3 m^3/s, while flow into the basin is $0.3[1 + \cos \pi t/(43{,}200)\ m^3/s)]$. Plot the hourly values of water depth versus time, assuming $h = h_o = 5$ m at $t = 0$.

1–4 Solve Problem 1–3 assuming the feed rate is $0.35[1 + \cos \pi t/(43{,}200)\ m^3/s]$, the flow out of the basin is 0.35 m^3/s, and the floor area for the storage tank is 2000 m^2. Plot the hourly values of water depth versus time, assuming $h = h_o = 2$ m at $t = 0$.

1–5 Wastewater is being pumped into a 4.2 m diameter tank at the rate of 0.5 m^3/min. At the same time, water leaves the tank at a rate that is dependent on the height of the liquid in the tank. The relationship governing the flow from the tank is $q = [2.1\ (m^2/min) \times h(m)]$. If the tank was initially empty, develop a relationship that can be used to define the height of the liquid in the tank as a function of time. What is the steady state height of the liquid in the tank?

1–6 Solve Problem 1–5 assuming the feed rate is 0.75 m^3/min and the tank outflow is $q = [2.7\ (m^2/min) \times h(m)]$.

1–7 The following data were obtained for four different reactants for the reaction $A \rightarrow B + C$. Determine the order of the reaction for one reactants (to be selected by instructor) and the value of the reaction rate constant *k*.

Time, *t*, min	Concentration, mg/L			
	Reactant number			
	1	2	3	4
0	90	1.9	240	113
10	72	1.55	150	80
20	57	1.31	110	56
40	36	0.99	70	28
60	23	0.8	51	14

1–8 A bimolecular reaction $A + B \rightarrow P$ is 10 percent complete in 10 min. If the initial concentration of A and B is equal to 1 mole/L, determine the reaction rate constant and how long it will take for the reaction to be 90 percent complete.

1–9 A bimolecular reaction $A + B \rightarrow P$ is 8 percent complete in 12 min. If the initial concentration of A and B is equal to 1.33 mole/L, determine the reaction rate constant and how long it will take for the reaction to be 96 percent complete.

1–10 The reaction rates at 10 and 25°C for a given reaction were found to differ by a factor of 2.75. Estimate the activation energy, *E*, for this reaction.

1–11 If two reaction rates differ by a factor on 2.4 and the activation energy, *E*, is 58,000 J/mole what is the temperature difference if the temperature at which the lowest reaction rate is observed is 15°C?

1–12 What is the difference in the reaction rates when the activation energy, *E*, is equal to 52,000 J/mole, the temperature difference is 15°C, and the temperature at the higher reaction rate is 27°C?

1–13 The following values have been obtained for the rate constant for the reaction A + B → P. Using these data, determine the activation energy E and the value of the rate constant at 15°C.

$k_{25°\text{C}} = 1.5 \times 10^{-2}$ L/mole·min

$k_{45°\text{C}} = 4.5 \times 10^{-2}$ L/mole·min

1–14 Solve Problem 1–13 for the following rate constant values.

$k_{20°\text{C}} = 1.25 \times 10^{-2}$ L/mole·min

$k_{35°\text{C}} = 3.55 \times 10^{-2}$ L/mole·min

1–15 An aqueous reaction is being studied in a laboratory-sized complete-mix reactor with a volume of 5 L. The stoichiometry of the reaction is A → 2R, and reactant A is introduced into the reactor at a concentration of 1 mole/L. From the results given in the following table, find the rate expression for this reaction. Assume steady-state flow.

Run	Feed rate, cm^3/s	Temperature, °C	Concentration of R in effluent, mole/L
1	2	13	1.8
2	15	13	1.5
3	15	84	1.8

1–16 The rate of reaction for an enzyme-catalyzed substrate in a batch reactor can be described by the following relationship.

$$r_c = \frac{kC}{K + C}$$

where k = maximum reaction rate, mg/L min
C = substrate concentration, mg/L
K = constant, mg/L

Using this rate expression, derive an equation that can be used to predict the reduction of substrate concentration with time in a batch reactor. If k equals 40 mg/L·min and K = 100 mg/L, determine the time required to decrease the substrate concentration from 1000 to 100 mg/L.

1–17 Solve Problem 1–16 for the following values: k equals 28 mg/L·min and K = 116 mg/L.

1–18 A wastewater is to be treated in a complete-mix reactor. Assuming that the reaction is irreversible and first-order ($r = -kC$) with a reaction rate coefficient equal to 0.15 d, determine the flowrate that can be treated if the reactor has a volume of 20 m^3 and 98 percent treatment efficiency is required. What volume would be required to treat the flowrate determined above if the required treatment efficiency is 92 percent?

1–19 For first-order removal kinetics, demonstrate that the maximum treatment efficiency in a series of complete-mix reactors occurs when all the reactors are the same size.

1–20 Determine the number of completely mixed chlorine contact chambers each having a detention time of 30 min that would be required in a series arrangement to reduce the bacterial count of a treated effluent from 10^6 to 14.5 organisms/mL if the first-order removal rate constant is equal to 6.1 h^{-1}. If a plug-flow chlorine contact chamber were used with the same detention time as the series of completely mixed chambers, what would the bacterial count be after treatment?

1–21 Derive the integrated expression for a plug-flow reactor assuming the removal of the constituent in question can be described by retarded first order reaction [Eq. (1–53)] for the following conditions $n = 1$ and $n \neq 1$.

1–22 Plot the ratio of required tank volume for a plug-flow reactor to that of a complete-mix reactor (V_{PFR}/V_{CMR}) versus the fraction of the original substrate that is converted for the following reaction rates.

$r = -k$

$r = -kC^{0.5}$

$r = -kC$

$r = -kC^2$

What is the value of the required volume ratio for each of these rates when $C = 0.25$ mg/L and $C_o = 1.0$ mg/L?

1–23 Solve Problem 1–22 for the following values: $C = 0.17$ mg/L and $C_o = 1.25$ mg/L.

1–24 If second-order reaction kinetics are applicable ($r = -kC^2$), determine the effluent concentration for each of the reactor systems shown on Fig. 1–16. To simplify the computations, assume that the following data apply.

$k = 1.0\ \text{m}^3/\text{kg}\cdot\text{d}$

$Q = 1.0\ \text{m}^3/\text{d}$

$V_{\text{PFR}} = 1.0\ \text{m}^3$

$V_{\text{CMR}} = 1.0\ \text{m}^3$

$C_o = 1.0\ \text{kg/m}^3$

Explain your results. What would happen if first- or zero-order kinetics are applicable?

1–25 A portion of the outflow, αQ, from an ideal plug-flow reactor is recycled around the reactor where $\alpha \geq 0$. Assume that the rate of conversion can be defined as $r_c = -kC$.

a. Sketch the generalized curve of conversion versus the recycle ratio.

b. Sketch a family of curves showing the effect of the recycle ratio α on the longitudinal concentration gradient.

c. If a complete-mix reactor were substituted for the plug-flow reactor, what effect would the recycle have on conversion?

1–26 Determine the effect of recycle on the performance of a complete-mix reactor for first and second order reactions.

1–27 Derive an expression that can be used to compute the effluent concentration from a free-surface constructed wetland treatment system, designed as an ideal plug-flow reactor, assuming the removal of the constituent in question can be described by a retarded second order equation. Assume the exponent n in the retardation term is equal to one. If the value of the retardation coefficient r_t is equal to 0.2, compare the effluent concentration with and without retardation. Assume the following values also apply: $C_o = 1$, $k = 0.1$, and $t = 1$.

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2

Wastewater Characteristics

WORKING TERMINOLOGY

Term	Definition
Absorbance	A measure of the amount of light, of a specified wavelength, that is absorbed by the constituents in solution.
Antibodies	Proteins used by the immune system to neutralize specific foreign material such as bacteria and viruses.

Term	Definition
Antigen	Any substance that stimulates the immune system to produce antibodies to protect against it.
Anthropogenic compounds	Chemical compounds created by humans, often resistant to biodegradation.
Bacteria	Microscopic organisms typically ranging from approximately 0.5 to 5 μm in length. Municipal wastewater can contain a wide variety and concentration of bacteria, including those pathogenic to humans.
Coliform group of bacteria	Coliforms include several genera of bacteria belonging to the family *Enterobacteriaceae*. Coliform organisms are common in the environment and in the feces of humans and warm blooded animals. Members of the coliform group include *Escherichia, Enterobacter, Klebsiella, and Citrobacter* among others.
Disinfection byproducts (DBPs)	A variety of organic compounds that are formed in treated wastewater when a strong oxidant for the purpose of disinfection such as chlorine or ozone is combined with residual organic matter.
Emerging contaminants	Constituents, which have been identified in water, that are considered for regulatory action pending the development of additional information on health and environmental impacts.
Endocrine disrupting compounds (EDCs)	Synthetic and natural compounds that mimic, block, stimulate, or inhibit natural hormones in the endocrine systems of animals, including humans. The origins of EDCs include pesticides, pharmaceutically active chemicals (PhACs), personal care products (PCPs), herbicides, industrial chemicals, and disinfection byproducts.
Enteric virus	An intracellular parasite that is an obligate human pathogen, which means it can only replicate in the human host.
Escherichia coli (E coli)	A species of the total coliform group that is specific to the intestinal tract of humans and warm blooded animals, commonly associated with fecal contamination. Although most *E. coli* are non-pathogenic, some strains (serotypes) such as *E. coli* O157:H7 can cause serious disease.
Fecal coliforms	A thermo-tolerant sub-group of total coliform group found in the intestinal tract of humans and warm blooded animals.
Helminths	A group of parasitic worms; worldwide, helminths are one of the principal causative agents of human disease. Helminths and helminth ova (eggs) are found increasingly in untreated municipal wastewater in the United States.
Higher heating value (HHV)	The amount of heat produced by complete combustion of a unit quantity of fuel.
Lower heating value (LHV)	The lower heating value is obtained by deducting the latent heat of vaporization of water vapor formed by combustion from the HHV.
Nanoparticles	Small objects or particles, ranging in size from 1 to 100 nm, which behave as an entire unit with respect to their properties and transport.
Pathogens	Microorganisms capable of causing diseases of varying severity.
Personal care products	Products such as shampoo, hair conditioner, deodorants, and body lotion.
Pharmaceutically active compounds (PhACs)	Chemicals synthesized for medical purposes (e.g., antibiotics).
Priority pollutants	Constituents, both inorganic and organic, that have been identified by the Environmental Protection Agency as known or suspected carcinogens, mutagens, teratogens, or highly toxic are to be regulated by categorical discharge standards.
Protozoa	Protozoa are single-celled organisms that lack a cell wall. They are common in fresh and marine water and some can grow in soil and other locations.
Trace constituent	A diverse classification of constituents found in low concentrations in untreated wastewater and not readily removed by conventional secondary treatment.
Trace organics	Organic compounds detected at at nanogram or microgram per liter concentrations in untreated and treated wastewater by the means of sophisticated laboratory instrumentation.
Viruses	Viruses are infectious agents able to multiply only within a host cell.

An understanding of the nature of wastewater is essential in the design and operation of collection, treatment, and reuse facilities, and in the engineering management of environmental quality. To promote this understanding, the information in this chapter is presented in ten sections dealing with (1) an introduction to the characterization of wastewater, (2) sampling and analytical procedures, (3) physical properties, (4) inorganic nonmetallic constituents, (5) metallic constituents, (6) aggregate organic constituents, (7) individual organic constituents and compounds, (8) microbial characteristics, (9) radionuclides, and (10) toxicity tests. The material in this chapter has been organized in a manner similar to that used in Standard Methods (2012), the standard reference work for the characterization of wastewater in the field of environmental engineering.

2–1 WASTEWATER CHARACTERIZATION

The principal constituents in wastewater, derived from domestic, municipal, and industrial sources, are: human excreta (i.e., feces and urine), shower/bath water, food waste, personal and household maintenance products, along with a wide variety of other inorganic and organic compounds in trace amounts. Given the wide variety of constituents that may be found in wastewater, it is common practice to characterize wastewater in terms of its physical properties and its chemical and biologic constituents. The physical properties and constituents found in wastewater and the constituents of concern in wastewater treatment are introduced briefly in the following discussion.

Wastewater Properties and Constituents

The principal physical properties and the chemical and biological constituents of wastewater and their sources are reported in Table 2–1. It should be noted that many of the physical properties and chemical and biological characteristics listed in Table 2–1 are interrelated. For example, temperature, a physical property, affects both the amounts of gases dissolved in the wastewater as well as the biological activity in the wastewater. Another distinction that can be made about the constituents reported in Table 2–1 is whether they are aggregate or individual constituents.

Constituents of Concern in Wastewater Treatment

The important constituents of concern in wastewater treatment are listed in Table 2–2. Secondary treatment standards, as reported in Table 1–2 in Chap. 1, are concerned with the removal of biodegradable organics, total suspended solids, and pathogens. Many of the more stringent standards that have been developed recently deal with the removal of nutrients, heavy metals, and priority pollutants. When wastewater is to be reused, standards normally include additional requirements for the removal of refractory organics, heavy metals, and in some cases, dissolved inorganic solids.

2–2 SAMPLING AND ANALYTICAL PROCEDURES

Proper sampling and analytical techniques are of fundamental importance in the characterization of wastewater. Sampling techniques, the methods of analysis, the units of measurement for chemical constituents, and some useful concepts from chemistry are considered in the next sections.

Table 2–1

Common analyses used to assess the constituents found in wastewater[a]

Test[a]	Abbreviation/ definition	Use or significance of test results
Physical characteristics		
Total solids	TS	To assess the reuse potential of a wastewater and to determine the most suitable type of operations and processes for its treatment
Total volatile solids	TVS	
Total fixed solids	TFS	
Total suspended solids	TSS	
Volatile suspended solids	VSS	
Fixed suspended solids	FSS	
Total dissolved solids	TDS (TS − TSS)	
Volatile dissolved solids	VDS	
Total fixed dissolved solids	FDS	
Settleable solids	SS	To determine those solids that will settle by gravity in a specified time period
Particle size	PS	To asses the performance of treatment processes, especially disinfection
Particle size distribution	PSD	To asses the performance of treatment processes
Turbidity	NTU[b]	Used to asses the quality of treated wastewater
Color	Light brown, grey, black	To assess the condition of wastewater (fresh or septic)
Transmittance	%T	To assess the suitability of treated effluent for UV disinfection
Odor	TON	To determine if odors will be a problem
Temperature	°C or °F	Important in the design and operation of biological processes in treatment facilities
Thermal energy content	J/g·°C	Important parameter in the recovery of heat from wastewater
Density	ρ	
Conductivity	EC	Used to assess the suitability of treated effluent for agricultural applications
Inorganic chemical characteristics		
Ammonia	NH_3	Used as a measure of the nutrients present and the degree of decomposition in the wastewater; the oxidized forms can be taken as a measure of the degree of oxidation.
Ammonium	NH_4^+	
Nitrite	NO_2^-	
Nitrate	NO_3^-	
Organic nitrogen	Org N	
Phosphorus, inorganic	Inorg P	Includes orthophosphates and polyphosphates
Orthophosphate	PO_4^{3-}	Simplest of the phosphoric acids salts
Organic phosphorus	Org P	
pH	$pH = -\log [H^+]$	A measure of the acidity or basicity of an aqueous solution

(*continued*)

Table 2–1 (*Continued*)

Test[a]	**Abbreviation/ definition**	**Use or significance of test results**
Inorganic chemical characteristics (continued)		
Alkalinity	$\Sigma(HCO_3^- + CO_3^{2-} + OH^- - H^+)$	A measure of the buffering capacity of the wastewater
Chloride	Cl^-	To assess the suitability of wastewater for agricultural reuse
Sulfate	SO_4^{2-}	To assess the potential for the formation of odors and may impact the treatability of the waste sludge
Metals	As, Cd, Ca, Cr, Co, Cu, Pb, Mg, Hg, Mo, Ni, Se, Na, Zn	To assess the suitability of the wastewater for reuse and for toxicity effects in treatment. Trace amounts of metals are important in biological treatment
Specific inorganic elements and compounds		To assess presence or absence of a specific constituent
Various gases	O_2, CO_2, NH_3, H_2S, CH_4	The presence or absence of specific gases
Organic chemical characteristics		
Five-day biochemical oxygen demand	BOD_5	A measure of the amount of oxygen required to stabilize a waste biologically over a 5-d period
Five-day carbonaceous biochemical oxygen demand	$CBOD_5$	A measure of the amount of oxygen required to stabilize a waste biologically, over a 5-d period, in which nitrogen oxidation is suppressed
Ultimate carbonaceous biochemical oxygen demand	UBOD (also BOD_u, BOD_L)	A measure of the amount of oxygen required to stabilize a waste biologically
Nitrogenous oxygen demand	NOD	A measure of the amount of oxygen required to oxidize biologically the nitrogen in the wastewater to nitrate
Chemical oxygen demand	COD	Often used as a substitute for the BOD test
Total organic carbon	TOC	Often used as a substitute for the BOD test
Specific organic compounds and classes of compounds	MBAS[c], CTAS[d]	To determine presence of specific organic compounds and to assess whether special design measures will be needed for removal
Chemical energy content	MJ/kg COD	To assess the chemical energy in wastewater
Biological characteristics		
Coliform organisms	MPN (most probable number)	To asses potential presence of pathogenic bacteria and effectiveness of disinfection process
Specific microorganisms	Bacteria, protozoa helminths, viruses	To asses presence of specific organisms in connection with plant operation and for reuse
Toxicity	TU_a[e] and TU_c[f]	To assess acute and chronic toxicity of various wastewater samples

[a] Details on the various tests may be found in Standard Methods (2012).

[b] NTU = Nephelometric turbidity unit.

[c] MBAS = Methylene blue active substances.

[d] CTAS = Cobalt thiocyanate active substances.

[e] TU_a = toxic unit acute.

[f] TU_c = toxic unit chronic.

Table 2–2

Principal constituents of concern in wastewater treatment

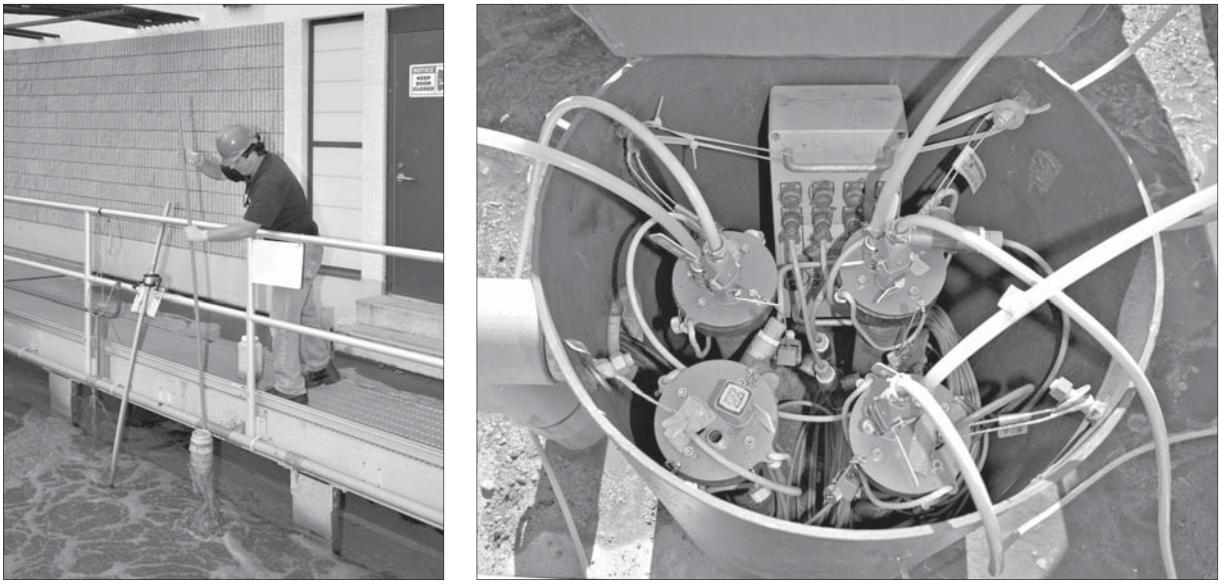
Constituent	Reason for importance
Suspended solids	Suspended solids can lead to the development of sludge deposits and anaerobic conditions when untreated wastewater is discharged in the aquatic environment.
Biodegradable organics	Composed principally of proteins, carbohydrates, and fats, biodegradable organics are measured most commonly in terms of BOD (biochemical oxygen demand) and COD (chemical oxygen demand). If discharged untreated to the environment, their biological stabilization can lead to the depletion of natural oxygen resources and to the development of septic conditions.
Pathogens	Communicable diseases can be transmitted by the pathogenic organisms that may be present in wastewater.
Nutrients	Both nitrogen and phosphorus, along with carbon, are essential nutrients for growth. When discharged to the aquatic environment, these nutrients can lead to the growth of undesirable aquatic life. When discharged in excessive amounts on land, they can also lead to the pollution of groundwater.
Priority pollutants	Organic and inorganic compounds selected on the basis of their known or suspected carcinogenicity, mutagenicity, teratogenicity, or high acute toxicity. Many of these compounds are found in wastewater.
Refractory organics	These organics tend to resist conventional methods of wastewater treatment. Typical examples include surfactants, phenols, and agricultural pesticides.
Heavy metals	Heavy metals are usually added to wastewater from commercial and industrial activities and may have to be removed if the wastewater is to be reused.
Dissolved inorganics	Inorganic constituents such as calcium, sodium, and sulfate are added to the original domestic water supply as a result of water use and may have to be removed if the wastewater is to be reused.

Sampling

Sampling programs are undertaken for a variety of reasons, such as to obtain (1) routine operating data on overall plant performance, (2) data that can be used to document the performance of a given treatment process, (3) data that can be used to implement proposed new programs, and (4) data needed for reporting regulatory compliance. To meet the goals of the sampling program, the data collected must be

1. *Representative:* The data must represent the wastewater or environment being sampled.
2. *Reproducible:* The data obtained must be reproducible by others following the same sampling and analytical protocols.
3. *Defensible:* Documentation must be available to validate the sampling procedures. The data must have a known degree of accuracy and precision.
4. *Useful:* The data can be used to meet the objectives of the monitoring plan (Pepper et al., 1996).

Because the data from the analysis of the samples will ultimately serve as a basis for implementing wastewater management facilities and programs, the techniques used in a



(a) (b)

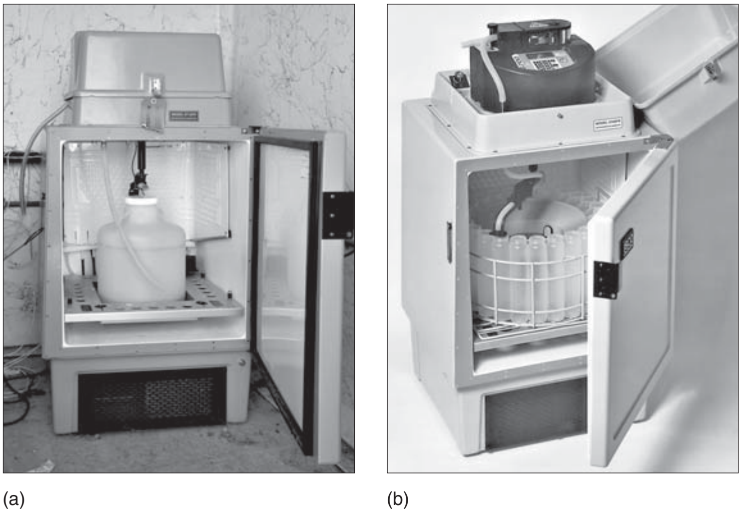
Figure 2–1

Collection of samples for analysis: (a) collection of a mixed liquor sample from end of plug flow activated sludge reactor and (b) view of an uncapped monitoring well equipped with sampling outlets for four different well depths. Samples are collected from each depth to monitor a groundwater injection system.

wastewater sampling program must be such that representative samples are obtained. There are no universal procedures for sampling; sampling programs must be tailored individually to fit each situation (see Fig. 2–1). Special procedures are necessary to handle sampling problems that arise when wastes vary considerably in composition.

Before a sampling program is undertaken, a detailed sampling protocol must be developed along with a quality assurance project plan (QAPP) (known previously as quality assurance/quality control, QA/QC). As a minimum, the following items must be specified in the QAPP (Pepper et al., 1996). Additional details on the subject of sampling may be found in Standard Methods (2012).

1. *Sampling plan:* Number of sampling locations, number and type of samples, time intervals (e.g., real time and/or time-delayed samples).
2. *Sample types and size:* Catch or grab samples, composite samples, or integrated samples, separate samples for different analyses (e.g., for metals). Sample size (i.e., volume) required.
3. *Sample labeling and chain of custody:* Sample labels, sample seals, field log book, chain of custody record, sample analysis request sheets, sample delivery to the laboratory, receipt and logging of sample, and assignment of sample for analysis.
4. *Sampling methods:* Specific techniques and equipment to be used (e.g., manual, automatic, or sorbent sampling).
5. *Sampling storage and preservation:* Type of containers (e.g., glass or plastic), preservation methods, maximum allowable holding times.
6. *Sample constituents:* A list of the parameters to be measured.
7. *Analytical methods:* A list of the field and laboratory test methods and procedures to be used and the detection limits for the individual methods.



(a) (b)

Figure 2–2

Typical refrigerated automatic composite samplers used to collect process and effluent samples over a 24-h period: (a) sampler with single sample bottle in place used to obtain a 24-h composite sample and (b) sampler used to collect individual hourly samples throughout the day. The individual samples can be composited in proportion to flow to obtain flow-weighted mass loading rates. (Courtesy of Teledyne Laboratory & Field Instruments.)

If the physical, chemical, or biological integrity of the samples is not maintained during interim periods between sample collection and sample analysis, a carefully performed sampling program will become worthless. Considerable research on the problem of sample preservation has failed to perfect a universal treatment or method or to formulate a set of fixed rules applicable to samples of all types. Prompt analysis is undoubtedly the most positive assurance against error due to sample deterioration. When analytical and testing conditions dictate a lag between collection and analysis, such as when a 24-h composite sample is collected, provisions must be made for preserving samples (see Fig. 2–2). Current methods of sample preservation for the analysis of properties subject to deterioration must be used (Standard Methods, 2012). Probable errors due to deterioration of the sample should be noted in reporting analytical data.

Methods of Analysis

The analyses used to characterize wastewater vary from precise quantitative chemical determinations to the more qualitative biological and physical determinations. The quantitative methods of analysis are gravimetric, volumetric, or physicochemical. In the physicochemical methods, properties other than mass or volume are measured. Instrumental methods of analysis such as turbidimetry, colorimetry, potentiometry, polarography, adsorption spectrometry, fluorometry, spectroscopy, and nuclear radiation are representative of the physicochemical analyses. Details concerning the various analyses may be found in Standard Methods (2012), the accepted reference that details the conduct of water and wastewater analyses.

Regardless of the method of analysis used, the detection level must be specified. Several detection limits are defined and are listed here in order of increasing levels (Standard Methods, 2012).

1. *Instrumental detection level (IDL).* Constituent concentration that produces a signal greater than five times the signal/noise ratio of the instrument.
2. *Lower level of detection (LLD).* Constituent concentration in reagent water that produces a signal ($2 \times 1.645s$) above the mean of blank analyses where s is the standard deviation.
3. *Method detection level (MDL).* Constituent concentration that, when processed through the complete method, produces a signal with a 99 percent probability that it is different from the blank.

4. *Level of quantification (LOQ).* Constituent concentration that produces a signal sufficiently greater than the blank that can be detected within specified levels by good laboratories during routine operating conditions. Typically it is the concentration that produces a signal 10*s* above the reagent blank signal.
5. *Minimum reporting level (MRL).* Minimum constituent concentration that can be reported as a quantitative value.

Units of Expression for Physical and Chemical Parameters

The results of the analysis of wastewater samples are expressed in terms of physical and chemical units of measurement. The most common units for these measurements are reported in Table 2–3. Measurements of chemical parameters are usually expressed in the physical unit of milligrams per liter (mg/L) or grams per cubic meter (g/m^3). The concentration of trace constituents is usually expressed as micrograms per liter (μg/L) or nanograms per liter (ng/L). As noted in Table 2–3, the concentration can also be expressed as parts per million (ppm), which is a mass-to-mass ratio. The relationship between mg/L and ppm is

$$\text{ppm} = \frac{\text{mg/L}}{\text{specific gravity of fluid}} \tag{2–1}$$

For dilute systems, such as those encountered in natural waters and wastewater in which one liter of sample weighs approximately one kilogram, the units of mg/L or g/m^3 are interchangeable with ppm. The terms *parts per billion* (ppb) and *parts per trillion* (ppt) are used interchangeably with μg/L and ng/L, respectively. Dissolved gases, considered to be chemical constituents, are expressed in parts per million by volume (ppm$_v$), μg/m^3, or mg/L. Conversion of gas concentrations between ppm$_v$ and μg/m^3 is given by Eq. (2–45), the universal law. Gases that evolve as by-products of wastewater treatment, such as carbon dioxide and methane (from anaerobic decomposition), are measured in terms of L or m^3 (ft^3). Parameters such as temperature, odor, hydrogen ion, and biological organisms are expressed in other units.

Useful Chemical Relationships

Other useful relationships from chemistry used in the analysis and evaluation of wastewater test results and in the design of treatment facilities include mole fraction, electroneutrality, chemical equilibrium, activity coefficient, ionic strength, and solubility product.

Mole Fraction. The ratio of the number of moles of a given solute to the total number of moles of all components in solution is defined as the *mole fraction.* Along with its importance in solution chemistry, the mole fraction is of importance in the mass transfer of gases into and out of liquids. In equation form,

$$x_B = \frac{n_B}{n_A + n_B + n_C \ldots n_N} \tag{2–2}$$

where x_B = mole fraction of solute B
n_B = number of moles of solute B
n_A = number of moles of solute A
n_C = number of moles of solute C
n_N = number of moles of solute N

Table 2–3

Units commonly used to express analytical results

Basis	Application	Unit
Physical analyses:		
Density	$\frac{\text{Mass of solution}}{\text{Unit of volume}}$	$\frac{\text{kg}}{\text{m}^3}$
Percent by volume	$\frac{\text{Volume of solute} \times 100}{\text{Total volume of solution}}$	% (by vol)
Percent by mass	$\frac{\text{Mass of solute} \times 100}{\text{Combined mass of solute} + \text{solvent}}$	% (by mass)
Volume ratio	$\frac{\text{Milliliters}}{\text{Liter}}$	$\frac{\text{mL}}{\text{L}}$
Mass per unit volume[a]	$\frac{\text{Picograms}}{\text{Liter of solution}}$	$\frac{\text{pg}}{\text{L}}$
	$\frac{\text{Nanograms}}{\text{Liter of solution}}$	$\frac{\text{ng}}{\text{L}}$
	$\frac{\text{Micrograms}}{\text{Liter of solution}}$	$\frac{\mu\text{g}}{\text{L}}$
	$\frac{\text{Milligrams}}{\text{Liter of solution}}$	$\frac{\text{mg}^a}{\text{L}}$
	$\frac{\text{Grams}}{\text{Cubic meter of solution}}$	$\frac{\text{g}}{\text{m}^3}$
Mass ratio	$\frac{\text{Milligrams}}{10^9 \text{ milligrams}}$	ppb[b]
	$\frac{\text{Milligrams}}{10^6 \text{ milligrams}}$	ppm
Chemical analyses:		
Molality	$\frac{\text{Moles of solute}}{\text{1000 grams solvent}}$	$\frac{\text{mole}}{\text{kg}}$
Molarity	$\frac{\text{Moles of solute}}{\text{Liter of solution}}$	$\frac{\text{mole}}{\text{L}}$
Normality	$\frac{\text{Equivalents of solute}}{\text{Liter of solution}}$	$\frac{\text{eq}}{\text{L}}$
	$\frac{\text{Milliequivalents of solute}}{\text{Liter of solution}}$	$\frac{\text{meq}}{\text{L}}$

[a] mg/L = g/m^3.

[b] ppb = parts per billion, ppm = parts per million, 10^3 ppb = ppm.

Note: 10^{12} pg = 10^9 ng = 10^6 μg = 10^3 mg = 1 gm.

The application of Eq. (2–2) is illustrated subsequently in Example 2–1.

EXAMPLE 2–1 **Determination of Mole Fraction** Determine the mole fraction of oxygen in water if the concentration of dissolved oxygen is 10.0 mg/L.

Solution

1. Determine the mole fraction of oxygen using Eq. (2–2) written as follows:

$$x_{O_2} = \frac{n_{O_2}}{n_{O_2} + n_w}$$

a. Determine the moles of oxygen.

$$n_{O_2} = \frac{(10 \text{ mg/L})}{(32 \times 10^3 \text{ mg/mole } O_2)} = 3.125 \times 10^{-4} \text{ mole/L}$$

b. Determine the moles of water.

$$n_w = \frac{(1000 \text{ g/L})}{(18 \text{ g/mole of water})} = 55.556 \text{ mole/L}$$

c. The mole fraction of oxygen is:

$$x_{O_2} = \frac{3.125 \times 10^{-4}}{3.125 \times 10^{-4} + 55.556} = 5.62 \times 10^{-6}$$

Electroneutrality. The principle of *electroneutrality* requires that the sum of the positive ions (cations) must equal the sum of negative ions (anions) in solution, thus

$$\Sigma \text{ cations} = \Sigma \text{ anions} \tag{2–3}$$

where cations = positively charged species in solution expressed in terms of equivalent weight per liter, eq/L or milliequvalent weight per liter, meq/L
anions = negatively charged species in solution, eq/L or meq/L

The equivalent weight of a compound is defined as:

$$\text{Equivalent weight, g/eq} = \frac{\text{molecular weight, g}}{Z} \tag{2–4}$$

where Z = (1) the absolute value of the ion charge, (2) the number of H^+ or OH^- ions a species can react with or yield in an acid-base reaction, or (3) the absolute value of the change in valence occurring in an oxidation reduction reaction (Sawyer et al., 2003).

Equation (2–3) can be used to check the accuracy of chemical analyses by taking into account the percent difference defined as follows (Standard Methods, 2012):

$$\text{Percent difference} = 100 \times \left(\frac{\Sigma \text{ cations} - \Sigma \text{ anions}}{\Sigma \text{ cations} + \Sigma \text{ anions}} \right) \tag{2–5}$$

The acceptance criteria are as given below.

Σ anions, meq/L	Acceptable difference
0–3.0	±0.2 meq/L
3.0–10.0	±2%
10–800	5%

From Standard Methods (2012).

The application of Eqs. (2–3) and (2–5) is illustrated in Example 2–2.

EXAMPLE 2–2 **Checking the Accuracy of Analytical Measurements** The following analysis has been completed on a filtered effluent, from an extended aeration wastewater treatment plant, that is to be used for landscape watering. Check the accuracy of the analysis to determine if the analysis is sufficiently accurate, based on the criteria given above.

Cation	Conc., mg/L	Anion	Conc., mg/L
Ca^{2+}	82.2	HCO_3^-	220.0
Mg^{2+}	17.9	SO_4^{2-}	98.3
Na^+	46.4	Cl^-	78.0
K^+	15.5	NO_3^-	25.6

Solution

1. Prepare a cation-anion balance.

Cation	Conc., mg/L	mg/meq[a]	meq/L	Anion	Conc., mg/L	mg/meq[a]	meq/L
Ca^{2+}	82.2	20.04[b]	4.10	HCO_3^-	220.0	61.02	3.61
Mg^{2+}	17.9	12.15	1.47	SO_4^{2-}	98.3	48.03	2.05
Na^+	46.4	23.00	2.02	Cl^-	78.0	35.45	2.20
K^+	15.5	39.10	0.40	NO_3^-	25.6	62.01	0.41
	Σ cations		7.99		Σ anions		8.27

[a] Molecular weight in grams/Z

[b] For calcium, eq wt = 40.08/2 = 20.04 g/eq or 20.04 mg/meq

2. Check the accuracy of the cation-anion balance using Eq. (2–5).

$$\text{Percent difference} = 100 \times \left(\frac{\Sigma \text{ cations} - \Sigma \text{ anions}}{\Sigma \text{ cations} + \Sigma \text{ anions}}\right)$$

$$\text{Percent difference} = 100 \times \left(\frac{7.99 - 8.27}{7.99 + 8.27}\right) = -1.72\%$$

For a total anion concentration between 3 and 10 meq/L, the acceptable difference must be equal to or less than 2 percent (see table given above), thus, the analysis is of sufficient accuracy.

Comment If the cation-anion balance is not of sufficient accuracy, the problem may be analytical or a constituent of significant concentration may be missing.

Chemical Equilibrium. A reversible chemical reaction in which reactants *A* and *B* combine to yield products *C* and *D* may be written as

$$aA + bB \rightleftarrows cC + dD \tag{2–6}$$

Where the *stoichiometry* coefficients *a*, *b*, *c*, and *d* correspond to the number of moles of constituents *A, B, C*, and *D*, respectively. The *stoichiometry* of a reaction refers to the definition of the quantities of chemical compounds involved in a reaction (e.g., *a* of *A, b* of *B*, etc.). When the chemical species come to a state of equilibrium, as governed by the law of mass action, the numerical value of the ratio of the products over the reactants is known as the *equilibrium constant K* and is written as

$$\frac{[C]^c[D]^d}{[A]^a[B]^b} = K \tag{2–7}$$

For a given reaction, the value of the equilibrium constant will change with temperature and the ionic strength of the solution. It should also be noted that in Eq. (2–7) it is assumed that that activity of the individual ions is equal to one.

Brackets are used in Eq. (2–7) to denote molar concentrations. The use of molal concentrations (see Table 2–3) is more correct theoretically, but for dilute solutions encountered in wastewater applications, molar concentrations are used. Molal concentrations must be used for brine solutions and sea water. To account for non-ideal conditions encountered due to ion-ion interactions, a new concentration term called *activity* is used. The activity of an ion is defined as follows:

$$a_i = \gamma[C_i] \tag{2–8}$$

where a_i = activity of *i*th ion, mole/L
γ = activity coefficient for the *i*th ion
C_i = concentration of *i*th ion in solution, mole/L

If Eq. (2–7) is written in terms of activity and activity coefficients rather than concentrations, the resulting expression is:

$$\frac{[a_C]^c[a_D]^d}{[a_A]^a[a_B]^b} = \frac{[\gamma_c C]^C[\gamma_D D]^d}{[\gamma_A A]^a[\gamma_B B]^b} = K \tag{2–9}$$

Ionic Strength. The *ionic strength* of a solution is a measure of the concentration of dissolved chemical constituents. The ionic strength of a solution can be estimated using the following expression:

$$I = \frac{1}{2}\Sigma\, C_i Z_i^2 \tag{2–10}$$

where I = ionic strength
C_i = concentration of the *i*th species, mole/L
Z_i = the valance (or oxidation) number of the *i*th species [see Eq. (2–4)]

The ionic strength can also be estimated based on the total dissolved solids concentration using the following expression:

$$I = 2.5 \times 10^{-5} \times \text{TDS} \tag{2–11}$$

where TDS = total dissolved solids, mg/L or g/m^3

Equation (2–11) is often used to estimate the ionic strength of treated wastewater in groundwater recharge applications.

Activity Coefficient. The activity coefficient can be estimated using the following expression, derived from the Debye-Huckel theory, as proposed by Davies (1962). Computation of the activity coefficient is illustrated in Example 2–3 following the discussion of ionic strength and solubility.

$$\log \gamma = -0.5\,(Z_i)^2\left(\frac{\sqrt{I}}{1+\sqrt{I}} - 0.3I\right) \tag{2–12}$$

where Z_i = charge on ith ionic species
I = ionic strength

The above relationship, without the $-0.3I$, term is often used for solutions with an ionic strength that does not exceed 0.1 M. A number of other similar relationships will be found in the literature. Computation of the activity coefficient is illustrated in Example 2–3 following the discussion of ionic strength and solubility.

Solubility Product. The equilibrium constant for a reaction involving a precipitate and its constituent ions is known as the *solubility product.* For example, the reaction for calcium carbonate ($CaCO_3$) is

$$CaCO_3 \rightleftarrows Ca^{2+} + CO_3^{2-} \tag{2–13}$$

Because the activity of the solid phase is usually taken as 1, the solubility product is written as

$$[Ca^{2+}][CO_3^{2-}] = K_{sp} \tag{2–14}$$

where K_{sp} = solubility product constant.

It is important to note that the value of the equilibrium constant will change with the temperature of the solution. Written in terms of activity coefficients, Eq. (2–14) becomes

$$\gamma_{Ca^{2+}}[Ca^{2+}]\gamma_{CO_3^{2-}}[CO_3^{2-}] = K_{sp} \tag{2–15}$$

The application of Eq. (2–15) is illustrated in Example 2–3.

EXAMPLE 2–3 Determine the Activity Coefficients and Solubility of Calcium Carbonate

Determine the activity coefficients for the mono and divalent ions in the wastewater given in Example 2–2. Using the value of the activity coefficient for a divalent ion, estimate the equilibrium concentration of calcium in solution needed to satisfy the solubility product for calcium carbonate ($CaCO_3$) at 25°C. The value of the solubility product constant K_{sp} for $CaCO_3$ at 25°C is 5×10^{-9}.

Solution

1. Determine the ionic strength of the wastewater using Eq. (2–10).
   a. Prepare a computation table to determine the summation term in Eq. (2–10) using the data from Example 2–2.

Ion	Conc., C, mg/L	$C \times 10^3$, mole/L	z^2	$cz^2 \times 10^3$
Ca^{2+}	82.2	2.051	4	8.404
Mg^{2+}	17.9	0.736	4	2.944
Na^+	46.4	2.017	1	2.017
K^+	15.5	0.396	1	0.397
HCO_3^-	220	3.607	1	3.607
SO_4^{2-}	98.3	1.024	4	4.096
Cl^-	78.0	2.200	1	2.200
NO_3^-	25.6	0.413	1	0.413
Sum				23.876

b. Determine the ionic strength of the wastewater.

$$I = \frac{1}{2}\Sigma C_i Z_i^2 = \frac{1}{2}(23.876 \times 10^{-3}) = 11.938 \times 10^{-3}$$

2. Determine the activity coefficients for Ca^{2+} and CO_3^{2-}. Because both species have a valance (charge) of 2, the activity of each will be the same.

   a. For monovalent ions

$$\log \gamma = -0.5(Z_i)^2\left(\frac{\sqrt{I}}{1+\sqrt{I}} - 0.3I\right)$$

$$\log \gamma = -0.5(1)^2\left[\frac{\sqrt{11.938 \times 10^{-3}}}{1+\sqrt{11.938 \times 10^{-3}}} - 0.3(11.938 \times 10^{-3})\right] = -0.0475$$

$$\gamma = 0.896$$

   b. For divalent ions

$$\log \gamma = -0.5(2)^2\left[\frac{\sqrt{11.938 \times 10^{-3}}}{1+\sqrt{11.938 \times 10^{-3}}} - 0.3(11.938 \times 10^{-3})\right] = -0.1898$$

$$\gamma = 0.646$$

3. Determine the minimum solubility of calcium using Eq. (2–15).

   a. Because the molar concentrations of calcium and carbonate ions are the same, Eq (2–15) can be written as follows:

$$\gamma^2 [C^2] = K_{sp}$$

   b. Solve for the concentration C.

$$C = \sqrt{\frac{K_{sp}}{\gamma^2}} = \sqrt{\frac{5 \times 10^{-9}}{(0.646)^2}} = 1.09 \times 10^{-4} \text{ mole/L}$$

   c. Convert the molar concentration of calcium carbonate to mg/L.

$$\text{Ca} = 1.09 \times 10^{-4} \text{ mole/L} \times 40{,}000 \text{ mg/mole} = 4.36 \text{ mg/L}$$

Comment The computed value represents the minimum concentration of calcium that would be required in solution to be in equilibrium with solid calcium carbonate.

2–3 PHYSICAL PROPERTIES

Important physical characteristics, as reported in Table 2–1, include total solids, particle size, particle size distribution, turbidity, color, transmittance, temperature, and conductivity. Total solids includes floating matter, settleable matter, colloidal matter, and matter in solution. Density, specific gravity, and specific weight, of general importance in water, are also important for wastewater.

Sources of Physical Properties

The sources of the physical properties used to characterize wastewater are of both natural and anthropogenic origin. Natural physical properties will depend on the source of the water and what treatment it has received prior to distribution as potable water. For example, the initial temperature of the water will vary depending on whether the source is surface water or groundwater as well as the part of the country. The specific gravity and weight are inherent properties of natural water. The other physical properties of wastewater are derived from the constituents added during usage, commercial and industrial discharges, and constituents found in inflow and infiltrating groundwater.

Solids

Wastewater contains a variety of solid materials varying from rags to colloidal material. In the characterization of wastewater, coarse materials are usually removed before the sample is analyzed for solids. The various solids classifications are identified in Table 2–4. The interrelationship between the various solids fractions found in wastewater is illustrated graphically on Fig. 2–3. The standard test for settleable solids consists of placing a wastewater sample

Table 2–4

Definitions for solids found in wastewater[a]

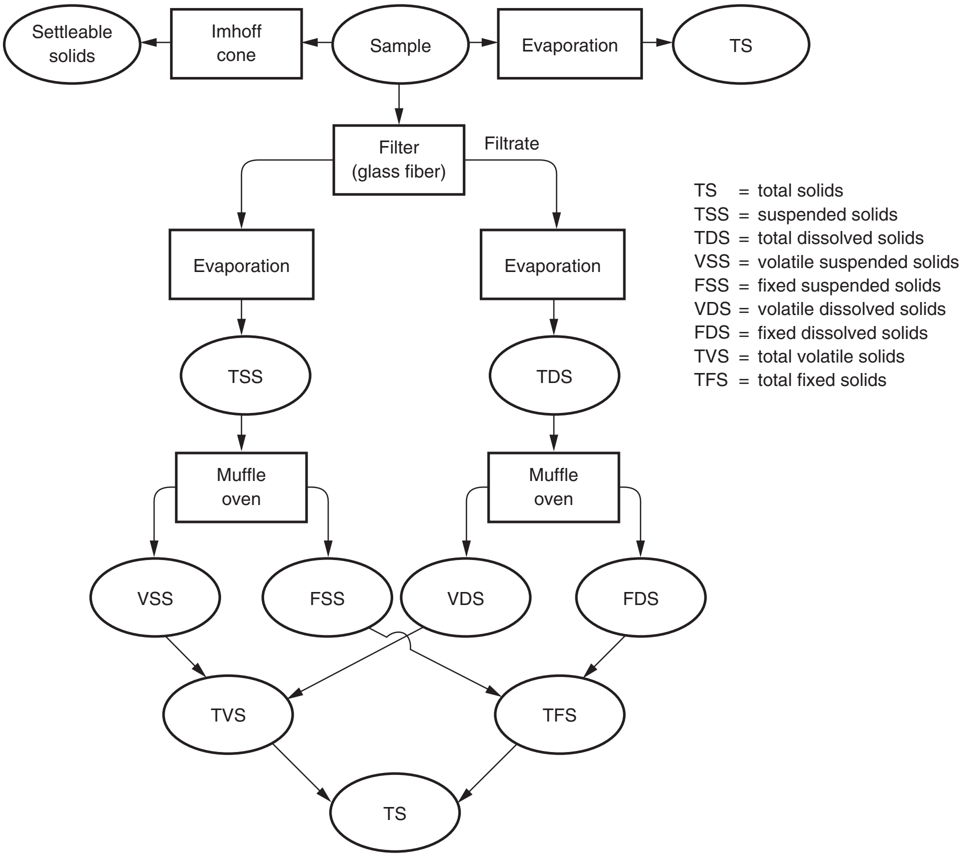
Test[b]	Description
Total solids (TS)	The residue remaining after a wastewater sample has been evaporated and dried at a specified temperature (103 to 105°C)
Total volatile solids (TVS)	Those solids that can be volatilized and burned off when the TS are ignited (500 ± 50°C)
Total fixed solids (TFS)	The residue that remains after TS are ignited (500 ± 50°C)
Total suspended solids (TSS)	Portion of the TS retained on a filter (see Fig. 2–3) with a specified pore size, measured after being dried at a specified temperature (105°C). The filter used most commonly for the determination of TSS is the Whatman glass fiber filter, which has a nominal pore size of about 1.58 μm
Volatile suspended solids (VSS)	Those solids that can be volatilized and burned off when the TSS are ignited (500 ± 50°C)
Fixed suspended solids (FSS)	The residue that remains after TSS are ignited (500 ± 50°C)
Total dissolved solids (TDS) (TS − TSS)	Those soilds that pass through the filter, and are then evaporated and dried at specified temperature. It should be noted that what is measured as TDS is comprised of colloidal and dissolved solids. Colloids are typically in the size range from 0.001 to 1 μm.
Total volatile dissolved solids (VDS)	Those solids that can be volatilized and burned off when the TDS are ignited (500 ± 50°C)
Fixed dissolved solids (FDS)	The residue that remains after TDS are ignited (500 ± 50°C)
Settleable solids	Suspended solids, expressed as milliliters per liter, that will settle out of suspension within a specified period of time (see Fig. 2–4)

[a] Adapted from Standard Methods (2012).

[b] With the exception of settleable solids, all solids values are expressed in mg/L.

Figure 2–3

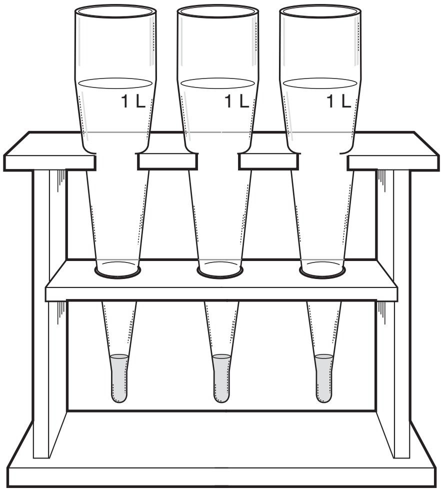
Interrelationships of solids found in water and wastewater. In much of the water quality literature, the solids passing through the filter are called dissolved solids (Tchobanoglous and Schroeder, 1985).



in a 1-liter Imhoff cone (see Fig. 2–4) and noting the volume of solids in millimeters that settle after a specified time period (1 h). Typically, about 60 percent of the suspended solids in a municipal wastewater are settleable. Total solids (TS) are obtained by evaporating a sample of wastewater to dryness and measuring the mass of the residual residue. As shown on Fig. 2–3, a filtration step is used to separate the total suspended solids (TSS) from the total dissolved solids (TDS). The apparatus used to determine TSS is shown on Fig. 2–5.

Figure 2–4

Imhoff cone used to determine settleable solids in wastewater. Solids that accumulate in the bottom of the cone after a 60-min settling time are reported as mL/L.



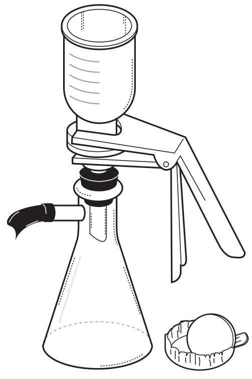


Figure 2–5

Vacuum filtration apparatus used for the determination of total suspended solids. After wastewater sample has been filtered, the preweighed filter paper is placed in an aluminum dish for drying before weighing.

Total Suspended Solids. Because a filter is used to separate the TSS from the TDS, the TSS test is somewhat arbitrary, depending on the pore size of the filter paper used for the test. Filters with nominal pore sizes varying from 0.45 μm to about 2.0 μm have been used for the TSS test (see Fig. 2–6). More TSS will be measured as the pore size of the filter used is reduced. Thus, it is important to note the pore size of the filter paper used when comparing reported TSS values.

It is also important to note that the TSS test itself has no fundamental significance. The principal reasons that the test lacks a fundamental basis are

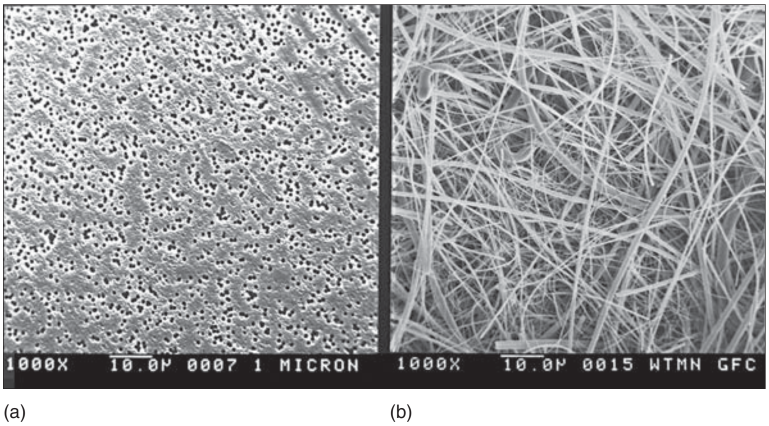
1. The measured values of TSS are dependent on the type and pore size of the filter used in the analysis.
2. Depending on the sample size used for the determination of TSS, auto filtration, where the suspended solids that have been intercepted by the filter also serve as a filter, can occur. Auto filtration will capture smaller particles than otherwise possible and cause an apparent increase in the measured TSS value over the actual value.
3. Depending on the characteristics of the particulate matter, small particles may be removed by adsorption to material already retained by the filter.
4. Because the number and size distribution of the particles that comprise the measured value is unknown, TSS is a *lumped parameter*.

Nevertheless, TSS test results are used routinely to assess the performance of conventional treatment processes and the need for effluent filtration in reuse applications. The TSS test is one of the two universally used effluent standards (along with BOD) by which the performance of treatment plants is judged for regulatory control purposes.

Total Dissolved Solids. By definition, the solids contained in the filtrate that pass through a filter with a nominal pore size of 1.2 μm or less are classified as dissolved (Standard Methods, 2012). Yet it is known that wastewater contains a high fraction of colloidal solids. The size of colloidal particles in wastewater is typically in the range of 0.01 to 1.0 μm. It should be noted that some researchers have classified the size range for colloidal particles as varying from 0.001 to 1.0 μm, others from 0.003 to 1.0 μm. The size range for colloidal particles considered in this text is from 0.01 to 1.0 μm. The number of colloidal particles in untreated wastewater and after primary sedimentation is typically in the range of 10^8 to 10^{12}/mL. Because the distinction between colloidal particles and truly dissolved material has not been made routinely, it has led to confusion in the analysis of treatment plant performance and in the design of treatment processes.

Figure 2–6

Micrographs of two laboratory filters used for the measurement of suspended solids in wastewater: (a) polycarbonate membrane filter with a nominal pore size of 1.0 μm and (b) glass fiber filter with a nominal pore size of 1.2 μm.



Volatile and Fixed Solids. Material that can be volatilized and burned off when ignited at 500 ± 50°C is classified as volatile. In general, volatile solids (VS) are presumed to be organic matter, although some organic matter will not burn, and some inorganic solids break down at high temperatures. Fixed solids (FS) comprise the residue that remains after a sample has been ignited. Thus, TS, TSS, and TDS comprise both fixed solids and volatile solids. The ratio of the VS to FS is often used to characterize the wastewater with respect to amount of organic matter present.

Particle Size and Particle Size Measurement

As noted above, TSS is a lumped parameter. In an effort to understand more about the nature of the particles that comprise the TSS in wastewater, measurement of particle size is undertaken and an analysis of the distribution of particle sizes is conducted (Tchobanoglous, 1995). Information on particle size is of importance in assessing the effectiveness of treatment processes (e.g., secondary sedimentation, effluent filtration, and effluent disinfection). Because the effectiveness of both chlorine, ozone, and UV disinfection is dependent on particle size, the determination of particle size has become more important, especially with greater effluent reuse in the western United States.

Information on the size of the biodegradable organic particles is significant from a treatment standpoint, as the biological conversion rate of these particles is dependent on size (see discussion in Sec. 2–6, which deals with biochemical oxygen demand). Methods that have been used to determine particle size are summarized in Table 2–5. As reported in Table 2–5, the methods can be divided into two general categories: (1) methods based on observation and measurement and (2) methods based on separation and analysis techniques. The methods used most commonly to study and quantify the particles in wastewater

Table 2–5

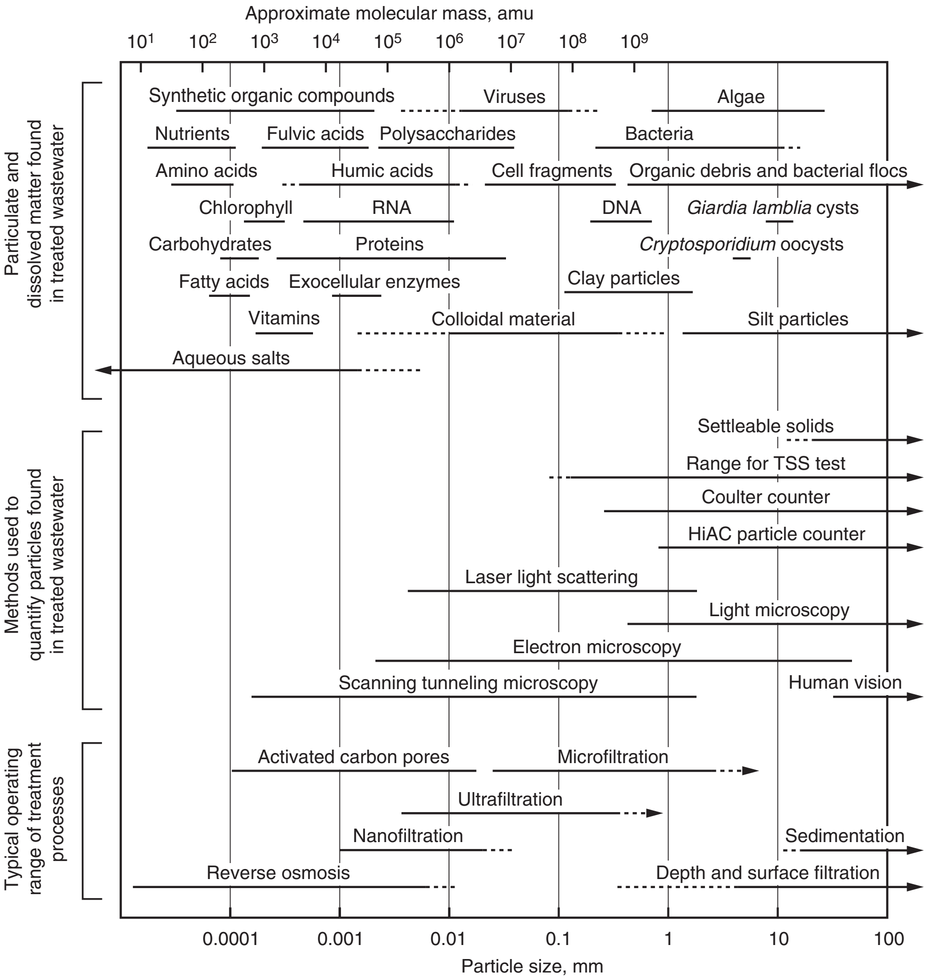
Representative analytical techniques applicable to particle size analysis of wastewater contaminants[a]

Technique	Typical size range, μm
Observation and measurement	
Microscopy	
Light	0.2–>100
Transmission electron (TEM)	0.2–>100
Scanning electron (SEM)	0.002–50
Image analysis	0.2–>100
Particle counters	
Conductivity difference	0.2–>100
Equivalent light scattering	0.005–>100
Light blockage	0.2–>100
Separation and analysis	
Centrifugation	0.08–>100
Field flow fractionation	0.09–>100
Gel filtration chromatography	<0.0001–>100
Sedimentation	0.05–>100
Membrane filtration (see Chap. 11)	0.0001–1

[a] Adapted from Levine et al. (1985).

Figure 2–7

Size ranges of organic constituents in wastewater and size separation and measurement techniques used for their quantification.



are (1) serial filtration, (2) electronic particle counting, and (3) direct microscopic observation. The principal types of materials that comprise the filterable and non-filterable solids in treated wastewater and their approximate size range are reported on Fig. 2–7.

Serial Filtration. Serial filtration may be used to determine an approximate particle size distribution of suspended solids based on mass (Levine et al., 1985). In the serial filtration method, a wastewater sample is passed sequentially through a series of membrane filters (see Fig. 2–8) with circular openings of known diameter (typically 12, 8, 5, 3, 1, and 0.1 μm), and the amount of particulate material retained in each filter is measured. Typical results from such a measurement are shown on Fig. 2–9. What is interesting to note on Fig. 2–9 is that a significant amount of colloidal material will be found between 0.1 and 1.0 μm. Although some information is gained on the size and distribution of the particles in the wastewater sample, little information is gained on the nature of the individual particles.

Electronic Particle Size Analyzers. To understand more about the nature and distribution of particles in wastewater, nondestructive measurement of particle size and particle size distribution is now quite common. However, it should be noted that electronic particle sizing and counting techniques cannot be used reliably for determining the source or type of particle (e.g., distinguishing between a viable cyst, a nonviable cyst, or a similar size silt particle).

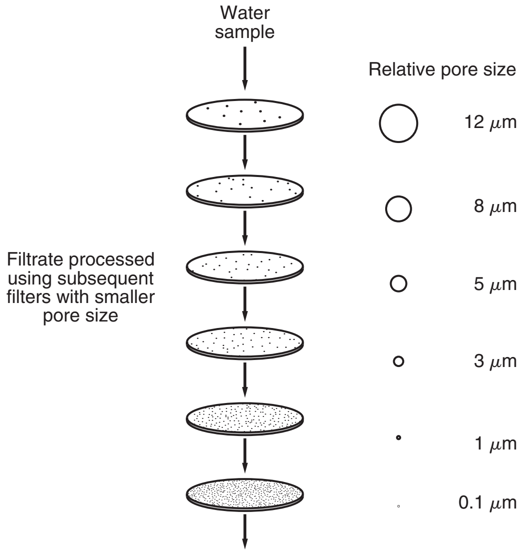


Figure 2–8

Definition sketch for the determination of the particle size distribution (by mass) using serial filtration with membrane filters.

In electronic particle size counting, particles are counted by diluting a treated wastewater sample and then passing the diluted sample through a calibrated orifice or past a laser beam [see Figs. 2–10(a) and (b), respectively]. As the particles pass through the orifice, the conductivity of the fluid changes, due to the presence of the particle. The change in conductivity is correlated to the size of an equivalent sphere. In a similar fashion, as a particle passes by a laser beam, it reduces the intensity of the laser due to light scattering. The reduced intensity is correlated to the diameter of the particle.

The typical size ranges quantifiable with different types of particle size counters were reported previously in Table 2–5. Most particle counters used in wastewater treatment facilities to assess performance have sensors available in different size ranges, such as 1.0 to 60 μm or 1 to 350 μm, depending on the manufacturer and application.

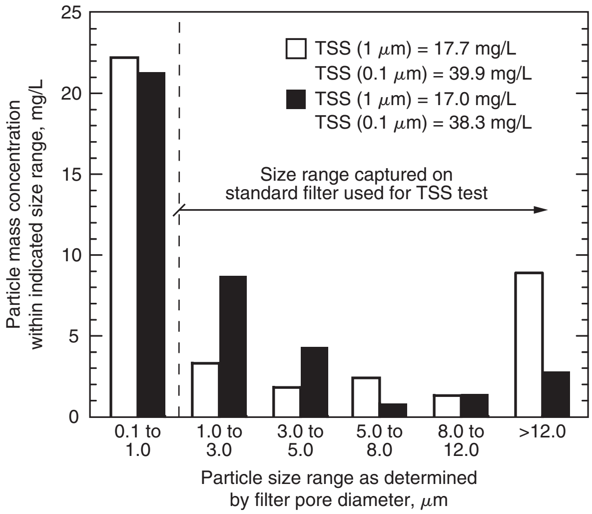


Figure 2–9

Typical data on the distribution of filterable solids obtained in two different tests by serial filtration in trickling filter effluent. Note: the large fraction of unmeasured solids between 0.1 and 1.0 μm using conventional TSS test.

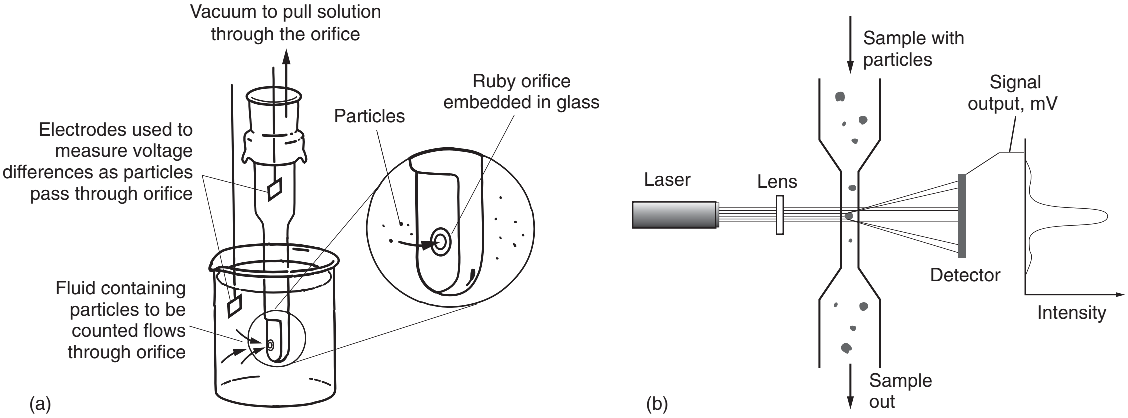


Figure 2–10

Determination of particle size distribution: (a) coulter counter, voltage difference as particle passes through the orifice is used to determine the size of an equivalent spherical particle and (b) laser particle size counter, size of equivalent spherical particle is based on reduced intensity and light scattering as particle passes through light beam.

Particle counters that do not measure particles smaller than 1 μm may be a limitation in some cases. Particle counts are typically measured and recorded in about 10 to 20 size ranges (e.g., 2 to 5 μm) called channels (or bins) of the chosen sensor range. One commonly used particle counter utilizes 128 channels. Channel sizes can be arithmetic, logarithmic, or arbitrary, depending on the measurement objective. Using a logarithmic scale, the upper channel limit is equal to the lower channel limit times a scaling factor. A typical particle size analysis with a laser type counter with 128 channels is shown on Fig. 2–11(a).

For disinfection studies, channel sizes are often selected that correspond to the size ranges of interest, for example, *Cryptosporidium* (2 to 5 μm) and *Giardia* (5 to 15 μm). With particle size counters that use large numbers of small channel sizes, the interpretation of the resulting data is more difficult. Where extremely small channel sizes are used, it is recommended that the data be aggregated into appropriate bin sizes [see Fig. 2–11(b)]. In addition to reporting particle number by size, the data can be reported in terms of surface area and volume; the volume fraction corresponding to each particle size range can also be computed, if needed (Standard Methods, 2012).

Direct Observation. For visualization of particles that are smaller than those visible to the unaided eye, microscopic techniques may be used. The use of microscopic observation allows for the determination of particle size counts and in some cases for more rigorous identification of a particle's origin than is possible with other analysis techniques. In microscopic observation, a measured volume of sample is placed in a particle counting cell, and the individual particles may be counted, often with the use of a stain to enhance the particle contrast. The size range quantifiable using a variety of microscopic techniques is reported in Table 2–5. In general, microscopic counting of particles is impractical on a routine basis, given the number of particles per mL of wastewater. Nevertheless, this method can be used to qualitatively assess the nature and size of the particles in wastewater.

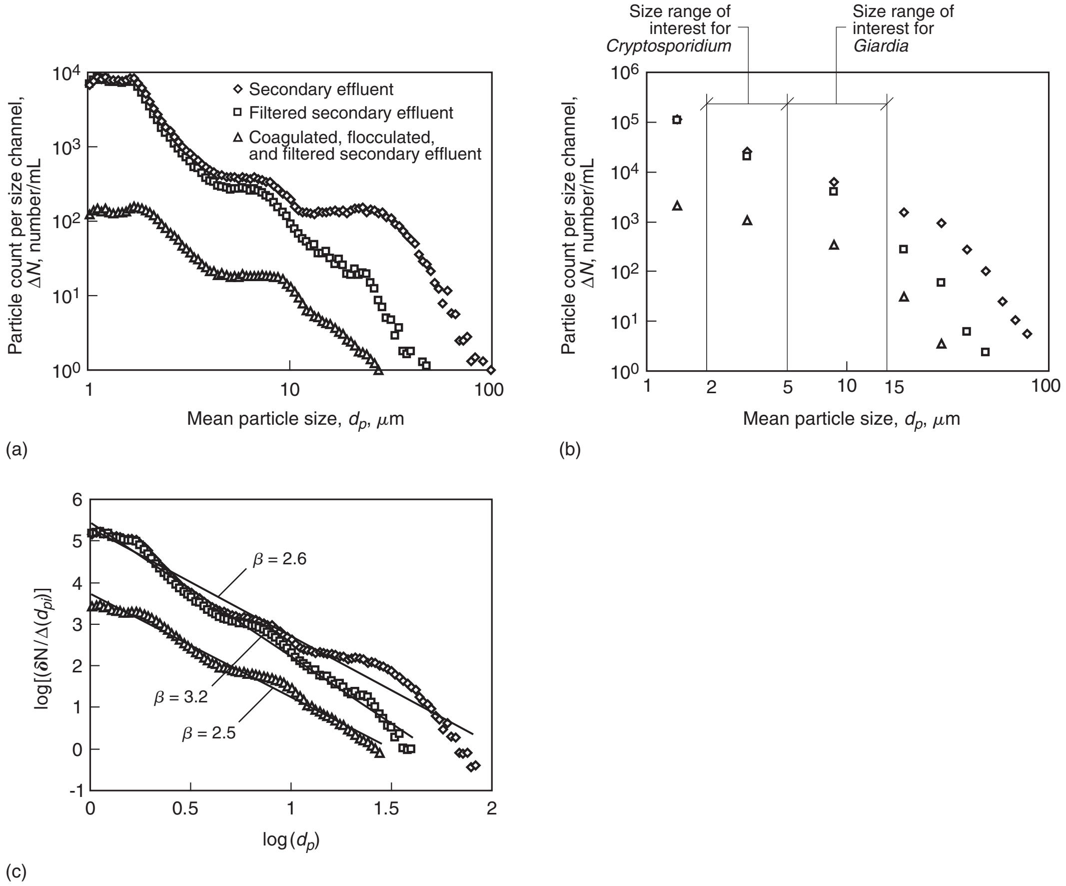


Figure 2–11

Effect the use of chemicals on filter particle size removal performance (a) original data as collected (courtesy of K. Bourgeous, 2005), (b) original data aggregated into selected channel (bin) sizes, and, (c) the original data, plotted functionally according to the power law (see Example 2–4).

Particle Size Distribution

In wastewater, it has been observed that the number of particles increases with decreasing particle diameter and that the frequency distribution typically follows a power law distribution of the form:

$$\frac{dN}{d(d_p)} = A(d_p)^{-\beta} \simeq \frac{\Delta N}{\Delta(d_{pi})} \tag{2–16}$$

where dN = the particle number concentration with respect to the incremental change in particle diameter $d(d_p)$, number/mL·μm
$d(d_p)$ = incremental change in particle diameter, μm
A = power law density coefficient, unitless
d_p = arithmetic (or geometric) mean particle diameter, depending on counter channel configuration, μm

β = power law slope coefficient
ΔN = the particle number concentration in given channel, number/mL
$\Delta(d_{pi})$ = incremental channel size, μm

In effect the right-hand term in Eq. (2–16) is used to normalize the data and allows for comparison between particle size distributions. Taking the log of both sides of Eq. (2–16) results in the following expression, which can be plotted to determine the unknown coefficients A and β:

$$\log\left[\frac{\Delta N}{\Delta(d_{pi})}\right] = \log A - \beta \log(d_p) \tag{2–17}$$

The value of A is determined when $d_p = 1\ \mu$m. As the value of A increases, the total number of particles in each size classification increases. The slope β is a measure of the relative number of particles in each size range. Thus, if β is less than one the particle size distribution is dominated by large particles, if β is equal to one all particle sizes are represented equally, and if β is greater than one the particle size distribution is dominated by small particles (Trussell and Tate, 1979). Because different slope values will be obtained, depending on the selection of the bin sizes, care must be exercised in interpreting the results. The analysis of data obtained from a particle size counter is shown on Fig. 2–11(c); the necessary computational steps are illustrated in Example 2–4.

EXAMPLE 2–4 **Analysis of Particle Size Information** Determine the coefficients A and β in Eq. (2–16) for the following particle size data obtained using a particle counter with arithmetic channel settings.

Channel size, mm	Number
1–2	20,000
2–5	6688
5–10	3000
10–15	1050
15–20	300
20–30	150
30–40	27
40–60	12
60–80	6
80–100	4
100–140	2

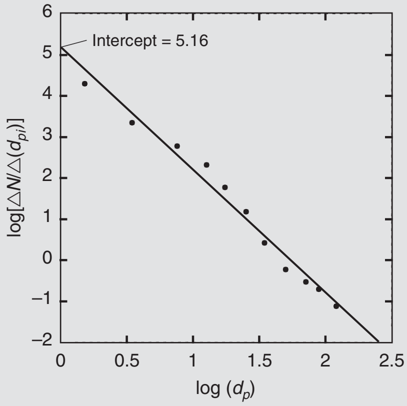
Solution

1. Set up a table to determine the information needed to plot the data

Channel size, mm	Mean diameter[a] d_p, μm	ΔN, number/mL	Channel size interval, $\Delta(d_{pi})$	$\log(d_p)$	$\log[\Delta N/\Delta(d_{pi})]$
1–2	1.50	20,000	1	0.18	4.30
2–5	3.50	6688	3	0.54	3.35
5–10	7.5	3000	5	0.88	2.78
10–15	12.5	1050	5	1.10	2.32
15–20	17.5	300	5	1.24	1.78
20–30	25.0	150	10	1.40	1.18
30–40	35.0	27	10	1.54	0.43
40–60	50.0	12	20	1.70	−0.22
60–80	70.0	6	20	1.85	−0.52
80–100	90.0	4	20	1.95	−0.70
100–140	120.0	2	40	2.08	−1.12

[a] Arithmetic mean diameter, 1.5 = [(1 + 2)/2].

2. Prepare a plot of the log of the geometric mean particle diameter, d_p, versus the normalized number of particles for the corresponding bin size, $\log[\Delta N/\Delta(d_{pi})]$.



3. Determine A and β in Eq. (2–16)
   a. Determine A

      When $\log(d_p) = 0$, $d_p = 1$, and $A = 10^{5.16}$

   b. Determine β

$$-\beta = \frac{3.65 - (-1.15)}{0.5 - 2} = -3.2$$

$$\beta = 3.2$$

Comment As the value of β is greater than one, the distribution is dominated by small particles, which is consistent with the actual data. It is important to note that the slope of the line of best fit through the plotted data will vary depending on the bin sizes selected for analysis. Also, it should be noted that the line used to define β may not be linear depending on the characteristics of the suspension and the minimum and maximum particle sizes measured, a characteristic of the specific instrument used in the analysis. The channel sizes of 2–5 and 5–15 μm were selected to determine if the number of *Cryptosporidium* or *Giardia* determined analytically can be correlated with particle size measurements.

Nanoparticles and Nanocomposites

Nanoparticles, originating from natural and anthropogenic processes, are small objects or particles, ranging in size from 1 to 100 nm, which behave as an entire unit with respect to their properties and transport. Because nanoparticles can form a variety of structures, such as nanospheres, nanotubes, or nanosheets, at least two of the three dimensions must be between 1 and 100 nm. Nanoparticles have also been referred to as ultrafine particles. Nanoparticles form a bridge between bulk materials and molecular or atomic structures. Nanocomposites, formed from two or more dissimilar materials, are developed to produce new structures with differing but controllable properties. For nanocomposites, at lease one of the materials (phases) must have a dimension in the nanoscale.

Common materials used for the production of nanoparticles, arranged alphabetically, include aluminum oxide, cerium oxide, cobalt, gold, iron, iron oxide, nickel, platinum, silica (SiO_2), silver, titanium dioxide (TiO_2), and zinc oxide. In addition to the constituents just mentioned, nanocomposites can include citrate, polyvinyl acetates (PVA), polyvinylpyrrolidone (PVP), tannic acid, and an ever expanding list of compounds. Nanoparticles are formed through natural processes and human industrial (anthropogenic) means. Natural processes include the oxidation of volatile compounds of biogenic origin. Industrially, nanoparticles are formed in the liquid and gas phase through a series of controlled chemical reactions.

Interest in nanoparticles and nanocomposites stems from the fact that they are now used extensively in the manufacture of a wide variety of consumer products such as self-cleaning glasses, clothing, scratch-resistant coatings, swimming pool cleaners, personal care products, and food production. Because of their widespread use, nanoparticles released from different household products and industrial activities are now being found in ever increasing concentrations in untreated and treated wastewater and biosolids.

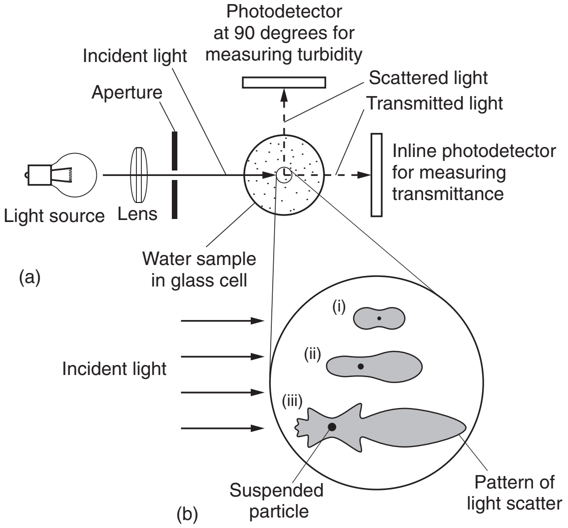
At this time, little is known about the long-term effects of nanoparticles on public health and their impacts when discharged to the environment. Also, there is some concern that nanoparticles may accumulate and that such accumulations may have health implications. In a recently completed study, it was found that the accumulation of silver nanoparticles may have a detrimental effect on nitrification and nutrient removal (Hu, 2010). Because the field of nanotechnology is evolving so rapidly, the current literature should be consulted for the latest developments including production, patterns of utilization, and the potential presence of nanoparticles in wastewater and their implications for treatment, public health, and environment. A comprehensive review of nanotechnology has been prepared by SCENIHR (2006).

Turbidity

Turbidity is a measure of the light scattering properties of a solution containing suspended and colloidal particles. Turbidity measurements require a light source (incandescent or

Figure 2–12

Determination of turbidity by light scattering: (a) schematic of turbidity apparatus and (b) typical light scattering patterns for small (i), intermediate (ii), and large (iii) particles.



light-emitting diode) and a sensor to measure the scattered light. As shown on Fig. 2–12(a), the scattered light sensor is located at 90 degrees to the light source. The measured turbidity increases as the intensity of the scattered light increases. Turbidity is expressed in nephelometric turbidity units (NTU). The spatial distribution and intensity of the scattered light, illustrated on Fig. 2–12(b), will depend on the size of the particle relative to the wavelength of the light source (Hach, 1997). For particles less than one-tenth of the wavelength of the incident light, the scattering of light is fairly symmetrical [see Fig. 2–12(b)(i)].

Limitations of Turbidity Measurements. As the particle size increases relative to the wave length of the incident light, the light reflected from different parts of the particle create interference patterns that are additive in the forward direction [see Figs. 2–12(b)(ii) and (iii)]. Also, the intensity of the scattered light varies with the wavelength of the incident light. For example, the turbidity of a solution of lamp black will essentially be equal to zero. Based on these considerations, turbidity measurements tend to be more sensitive to particles in the size range of the incident light wavelength (0.3 to 0.7 μm for visible light).

Thus, two filtered wastewater samples with nearly identical turbidity values could have very different particle size distributions. A further complication with turbidity measurements is that some particles will essentially adsorb most of the light and only scatter a minimal amount of the incident light. Also, because of the light scattering characteristics of large particles, a few large particles would not be detected in the presence of many smaller particles. Also, some online turbidity meters used to monitor the performance of microfiltration units are affected by the air used to clean the membranes.

Thus, there is no fundamental relationship between turbidity and the concentration of total suspended solids, and turbidity alone is not a good measure of whether wastewater can be disinfected effectively. As a result, it is almost impossible to compare turbidity values reported in the literature. However, turbidity readings at a given facility can be used for process control.

Relationship Between Turbidity and TSS

In general, there is no relationship between turbidity and the concentration of total suspended solids in untreated wastewater. There is, however, a reasonable relationship between turbidity and total suspended solids for the settled and filtered secondary effluent from the activated sludge process. The general form of the relationship is as follows:

$$\text{TSS, mg/L} \approx (\text{TSS}_f)(T) \tag{2–18}$$

where TSS = total suspended solids, mg/L
TSS_f = factor used to convert turbidity readings to total suspended solids, (mg TSS/L)/NTU
T = turbidity, NTU

The specific value of the conversion factor will vary for each treatment plant, depending primarily on the operation of the biological treatment process. The conversion factors for settled secondary effluent and for secondary effluent filtered with a granular-medium depth filter will typically vary from 2.3 to 2.4 and 1.3 to 1.6, respectively.

Color

Historically, the term *condition* was used along with composition and concentration to describe wastewater. Condition refers to the age of the wastewater which is determined qualitatively by its color and odor. Fresh wastewater is usually a light brownish-gray color. However, as the travel time in the collection system increases, and more anaerobic conditions develop, the color of the wastewater changes sequentially from gray to dark gray, and ultimately to black. When the color of the wastewater is black, the wastewater is often described as septic. Some industrial wastewaters may also add color to domestic wastewater. In most cases, the gray, dark gray, and black color of the wastewater is due to the formation of metallic sulfides, which form as the sulfide produced under anaerobic conditions reacts with the metals in the wastewater.

Absorption/Transmittance

The absorbance of a solution is a measure of the amount of light, of a specified wavelength, that is absorbed by the constituents in a solution. Absorbance is measured using a spectrophotometer with a fixed path length (usually 1.0 cm) at a wavelength of 254 nm. Absorbance follows the Beers-Lambert Law as given by Eq. (2–19):

$$\log\left(\frac{I}{I_o}\right) = \varepsilon(\lambda)Cx \tag{2–19}$$

Where I = light intensity at distance x from the light source, mW/cm^2
I_o = light intensity at light source, mW/cm^2
$\varepsilon(\lambda)$ = molar absorptivity (also known as the extinction coefficient) of the light-absorbing solute at wavelength λ, L/mole·cm
C = concentration of light-absorbing solute, mole/L
x = light path-length, cm

When the left-hand side of Eq. (2–19) is expressed as a natural logarithm, the right-hand side of the equation must be multiplied by 2.303 because the absorbance coefficient is determined in base 10. The term on the right-hand side of Eq. (2–19) is defined as the absorbance, $A(\lambda)$, which is unitless but is often reported in units of cm^{-1}, which corresponds to absorptivity $k(\lambda)$. If the length of the light path is 1 cm, absorptivity is equal to the absorbance.

$$k(\lambda) = e(\lambda)C = \frac{A(\lambda)}{x} \tag{2–20}$$

where $k(\lambda)$ = the absorptivity, cm^{-1}
$A(\lambda)$ = absorbance, dimensionless

Absorbance is measured using a spectrophotometer typically using a fixed sample path length of 1.0 cm.

The transmittance of a solution is defined as

$$\text{Transmittance, } T, \% = \left(\frac{I}{I_o}\right) \times 100 \tag{2–21}$$

The transmittance at a given wavelength can also be derived from absorbance measurements using the following relationship:

$$T = 10^{-A(\lambda)} \tag{2–22}$$

The term *percent transmittance,* commonly used in the literature is

$$T, \% = 10^{-A(\lambda)} \times 100 \tag{2–23}$$

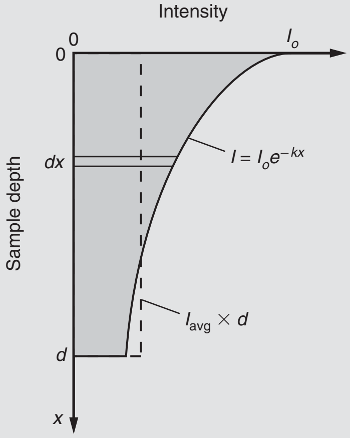
Thus, for a perfectly transparent solution $A(\lambda) = 0$, $T = 1$ and for a perfectly opaque solution $A(\lambda) \rightarrow \infty$, $T = 0$.

The principal water characteristics that affect the percent transmittance include: inorganic compounds (e.g., copper, iron), organic compounds (e.g., organic dyes, humic substances, and aromatic compounds such as benzene and toluene), and small colloidal particles ($\leq 0.45\ \mu$m). Additional details of the compounds that affect transmittance are given in the discussion of UV disinfection in Chap. 12. Typical absorbance and transmittance values for wastewater following various levels of treatment are given in Table 12–29, in Chap. 12.

EXAMPLE 2–5 **Variation of UV Intensity with Depth** If the intensity of the UV irradiation measured at the water surface in a Petri dish is 10 mW/cm^2, determine the average UV intensity to which a sample will be exposed if the depth of water in the Petri dish is 10 mm. Assume the absorptivity, $k(\lambda = 254\text{ nm})$, is equal to 1.0 cm^{-1}.

Solution

1. Use Beers-Lambert Law [Eq. (2–19)] to determine the average intensity.
   a. The definition sketch for this problem is given below.



b. Develop the required equation.

$$I_{\text{avg}} \times d = \int_0^d I_o e^{-kx} dx = -\frac{I_o}{k} e^{-kx} \Big]_0^d = -\frac{I_o}{k} e^{-kd} + \frac{I_o}{k} = \frac{I_o}{k}(1 - e^{-kd})$$

$$I_{\text{avg}} = \frac{I_o}{kd}(1 - e^{-kd})$$

2. Compute the average intensity for a depth of 10 mm (1 cm).
   a. Absorptivity, k, $= 1.0\ \text{cm}^{-1}$ (given)
   b. Solve for I_{avg}

$$I_{\text{avg}} = \frac{I_o}{kd}(1 - e^{-kd}) = \frac{(10\ \text{mW/cm}^2)}{(1/\text{cm})(1\ \text{cm})}[1 - e^{-(1/\text{cm})(1\ \text{cm})}] = 6.32\ \text{mW/cm}^2$$

Comment The equation developed in Step 1 is utilized for the analysis of UV dose in Chap. 12.

Temperature

The temperature of wastewater is commonly higher than that of the local water supply because of the addition of warm water from households and industrial activities. As the specific heat of water is much greater than that of air, the observed wastewater temperatures are higher than the local air temperatures during most of the year and are lower only during the hottest summer months. Depending on the geographic location, the mean annual temperature of wastewater in the United States varies from about 3 to 27°C (37 to 81°F); 15.6°C (60°F) is a representative value. Temperatures as high as 30 to 35°C (84 to 98°F) have been reported for countries in Africa and the Middle East. The variation that can be expected in influent wastewater temperatures is illustrated on Fig. 2–13. Depending on the location and time of year, the effluent temperatures can either be higher or lower than the corresponding influent values.

Effects of Temperature. The temperature of water is a very important parameter because of its effect on chemical reactions and reaction rates, aquatic life, and the suitability of the water for beneficial uses. Increased temperature, for example, can cause a change in the species of fish that can exist in the receiving water body. Industrial establishments that use surface water for cooling-water purposes are particularly concerned with the temperature of the intake water.

In addition, oxygen is less soluble in warm water than in cold water. The increase in the rate of biochemical reactions that accompanies an increase in temperature, combined with the decrease in the quantity of oxygen present in surface waters, can often cause serious depletions in dissolved oxygen concentrations in the summer months. When significantly large quantities of heated water are discharged to natural receiving waters, these effects are magnified. It should also be realized that a sudden change in temperature can result in a high rate of mortality of aquatic life. Moreover, abnormally high temperatures can foster the growth of undesirable water plants and wastewater fungi.

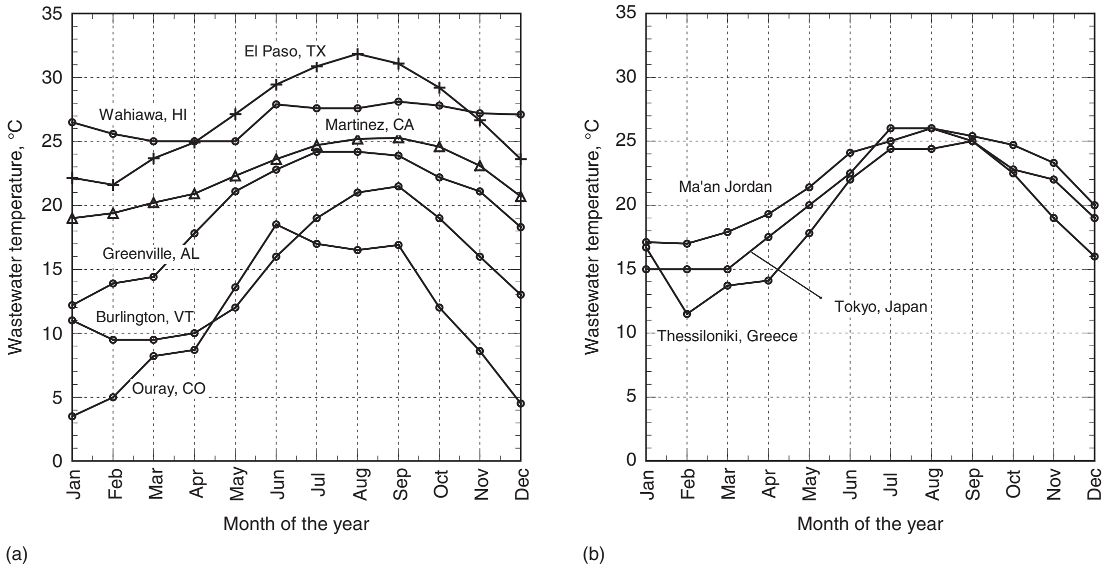


Figure 2–13

Typical variations in monthly influent wastewater temperatures (a) in various parts of the United States and (b) selected countries worldwide.

Optimum Temperatures for Biological Activity. Optimum temperatures for bacterial activity are in the range from 25 to 35°C. Aerobic digestion and nitrification stops when the temperature rises to 50°C. When the temperature drops to about 15°C, methane-producing bacteria become quite inactive, and at about 5°C, the autotrophic-nitrifying bacteria practically cease functioning. At 2°C, even the chemoheterotrophic bacteria acting on carbonaceous material become essentially dormant. The effects of temperature on the performance of biological treatment processes are considered in greater detail in Chaps. 7 and 8.

Estimation of Temperature Effects on Reaction Rates. Equilibrium constants, solubility product constants, and specific reaction-rate constants are all dependent on temperature. The temperature dependence of rate and equilibrium constants is based on the van't Hoff-Arrhenius relationship, as described in Chap. 1.

If the rate coefficient, say k_1, is known for a given temperature, the rate coefficient, k_2, at another temperature can be estimated using Eq. (1–44), repeated here for convenience. The derivation of Eq. (1–44) from the van't Hoff-Arrhenius relationship is presented in Chap. 1, Sec. 1–6.

$$\frac{k_2}{k_1} = \theta^{(T_2 - T_1)} \tag{1–44}$$

It should be noted, however, that although the value of the temperature coefficient, θ, is assumed to be constant, it will vary with temperature. Therefore, caution must be used in selecting appropriate values for θ for different temperature ranges (see Example 1–1). Typical values for various processes for different temperature ranges are given, where available, in the sections in which the individual topics are discussed.

Thermal Energy Content of Wastewater

The total energy content of wastewater is comprised of the heat and the organic constituents found in wastewater. The energy content of wastewater is an important consideration as new technologies and concepts are evaluated. Typically, the total energy content in wastewater is two to four times the energy needed to operate a WWTP. The sources of energy in wastewater are heat energy discussed below and chemical energy considered in Sec. 2–6.

The heat energy in wastewater derives from the energy added to water to heat it for a number of domestic and commercial applications. In domestic usage, the most important sources of heat in wastewater are from clothes washing, dishwashing, and bathing. The temperature of potable water typically varies from 4.5 to 10°C (40 to 50°F). The corresponding temperatures of wastewater as discharged from a residence are typically in the range from 15 to 28°C (60 to 80°F). Wastewater temperatures measured at wastewater treatment plants will vary with the season of the year and the characteristics of the collection system but are in the range from 15 to 24°C (60 to 75°F).

The heat content of wastewater can be approximated by considering the specific heat of water, which at 20°C is 4.1816 J/g·°C. Thus, for a 10°C difference, the heat content of wastewater per 1000 m^3 is equal to 41,816 MJ/10°C·10^3 m^3. Although the heat in wastewater can be recovered using a heat pump (see Chap. 17), the economic feasibility of recovering heat energy from wastewater depends on having a year-round use for the extracted heat. The heat energy extracted from wastewater can be used for drying organic materials removed by fine screening, sludge, and biosolids for subsequent processing.

Conductivity

The electrical conductivity (EC) of water is a measure of its ability to conduct an electrical current. Because the electrical current is transported by the ions in solution, the conductivity increases as the concentration of ions increases. In effect, the measured EC value is used as a surrogate measure of total dissolved solids (TDS) concentration. At present, the EC of a water is one of the important parameters used to determine the suitability of a water for irrigation. The salinity of treated wastewater to be used for irrigation is estimated by measuring its electrical conductivity.

The electrical conductivity in SI units is expressed as millisiemens per meter (mS/m) and in micromhos per centimeter (μmho/cm) in US customary units. It should be noted that 1 mS/m is equivalent to 10 μmho/cm. Equation (2–24) can be used to estimate the TDS of a water sample based on the measured EC value (Standard Methods, 2012).

$$\text{TDS (mg/L)} \cong \text{EC } (\mu\text{S/cm or } \mu\text{mho/cm}) \times (0.55 - 0.70) \tag{2–24}$$

The above relationship does not necessarily apply to raw wastewater or high-strength industrial wastewater. The above relationship can also be used to check the acceptability of chemical analyses (see Standard Methods, 2012).

The electrical conductivity can also be used to estimate the ionic strength of a solution using the following relationship (Russell, 1976)

$$I = 1.6 \times 10^{-5} \times \text{EC } (\mu\text{S/cm or } \mu\text{mho/cm}) \tag{2–25}$$

Equation (2–25) is used to estimate the ionic strength of treated wastewater in groundwater recharge applications (see Chap. 13).

Density, Specific Gravity, and Specific Weight

The density of wastewater, ρ_w, is defined as its mass per unit volume expressed as g/L or kg/m^3 in SI units and as $\text{lb}_\text{m}/\text{ft}^3$ in U.S. customary units. Density is an important physical characteristic of wastewater because of the potential for the formation of density currents in sedimentation tanks, chlorine contact tanks, and other treatment units. The density of domestic wastewater, which does not contain significant amounts of industrial waste, is essentially the same as that of water at the same temperature (see Appendix C).

In some cases the specific gravity of the wastewater s_w, is used in place of the density. The specific gravity is defined as:

$$s_w = \frac{\rho_w}{\rho_o} \tag{2–26}$$

where ρ_w = density of wastewater
ρ_o = density of water

Both the density and specific gravity of wastewater are temperature dependent and will vary with the concentration of total solids in the wastewater.

The specific weight of a fluid, γ, is its weight per unit volume. Specific weight is expressed as kN/m^3 in SI units and as lb_f/ft^3 in U.S. customary units. The relationship between γ, ρ, and the acceleration due to gravity, g, is $\gamma = \rho g$. At normal temperatures γ is about 9.81 kN/m^3 (62.4 lb_f/ft^3). Values for both density and specific weight as a function of temperature in both SI and U.S. customary units are given in Appendix C.

2–4 INORGANIC NONMETALLIC CONSTITUENTS

The chemical constituents of wastewater are typically classified as inorganic and organic. Inorganic chemical constituents of concern include nutrients, nonmetallic constituents, metals, and gases. Inorganic nonmetallic constituents considered in this section include pH, nitrogen, phosphorus, alkalinity, chlorides, sulfur, other inorganic constituents, gases, and odors. Metallic constituents are considered in Section 2–5.

Sources of Inorganic Nonmetallic Constituents

The sources of inorganic nonmetallic constituents in wastewater derive from the background levels in the water supply and from the additions resulting from domestic use, from the addition of highly mineralized water from private wells and groundwater, and from industrial use. Domestic and industrial water softeners also contribute significantly to the increase in mineral content and, in some areas, may represent the major source. Occasionally, water added from private wells and groundwater infiltration will (because of its high quality) serve to dilute the mineral concentration in the wastewater. Because concentrations of various inorganic constituents can greatly affect the beneficial uses made of the waters, the constituents in each wastewater must be considered separately.

pH

Because the concentration of the species of most chemical constituents is dependent on the hydrogen-ion concentration in solution, the hydrogen-ion concentration is an important quality parameter of both natural waters and wastewaters. The usual means of expressing the hydrogen ion concentration is as pH, which is defined as the negative logarithm of the hydrogen ion concentration:

$$\text{pH} = -\log_{10}[\text{H}^+] \tag{2–27}$$

The concentration range suitable for the existence of most biological life is quite narrow and critical (typically 6 to 9). Wastewater with an extreme concentration of the hydrogen ion is difficult to treat by biological means, and if the concentration is not altered before discharge, the wastewater effluent may alter the concentration in the natural receiving waters. For treated effluents discharged to the environment, the allowable pH range usually varies from 6.5 to 8.5.

The hydrogen ion concentration in water is connected closely with the extent to which water molecules dissociate. Water will dissociate into hydrogen and hydroxyl ions as follows:

$$H_2O \rightleftarrows H^+ + OH \tag{2–28}$$

Applying the law of mass action [Eq. (2–7)] to Eq. (2–28) yields

$$\frac{[H^+][OH^-]}{[H_2O]} = K \tag{2–29}$$

where the brackets indicate concentration of the constituents in moles per liter. Because the concentration of water in a dilute aqueous system is essentially constant, this concentration can be incorporated into the equilibrium constant K to give

$$[H^+][OH^-] = K_w \tag{2–30}$$

K_w is known as the ionization constant or ion product of water and is approximately equal to 1×10^{-14} at a temperature of 25°C. Equation (2–30) can be used to calculate the hydroxyl ion concentration when the hydrogen ion concentration is known and vice versa.

With pOH, which is defined as the negative logarithm of the hydroxyl ion concentration, it can be seen from Eq. (2–30) that, for water at 25°C,

$$pH + pOH = 14 \tag{2–31}$$

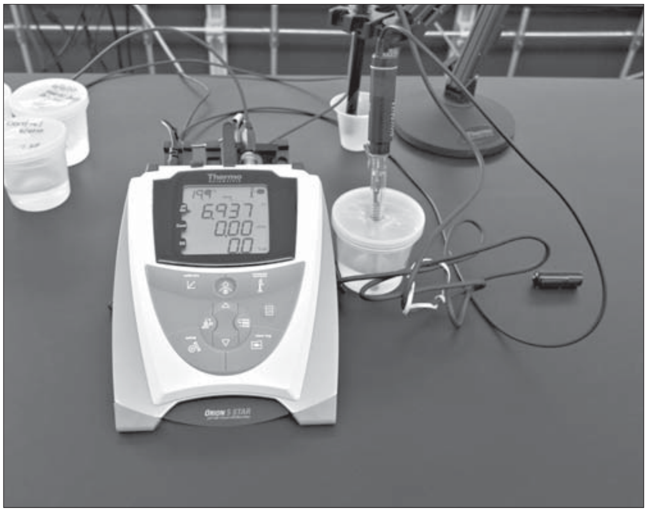
The pH of aqueous systems typically is measured with a pH meter (see Fig. 2–14). Various pH papers and indicator solutions that change color at definite pH values are also used. The pH is determined by comparing the color of the paper or solution to a series of color standards.

Chlorides

Chloride is a constituent of concern in wastewater as it can affect the final reuse applications of treated wastewater. Chlorides in natural water result from the leaching of chloride containing rocks and soils with which the water comes in contact, and in coastal areas from saltwater intrusion. In addition, agricultural, industrial, and domestic wastewaters discharged to surface waters are a source of chlorides.

Figure 2–14

Typical meter used for the measurement of pH and specific ion concentrations.



Human excreta, for example, contain about 6 g of chlorides per person per day. In areas where the hardness of water is high, home regeneration–type water softeners will also add large quantities of chlorides. Because conventional methods of waste treatment do not remove chloride to any significant extent, higher than usual chloride concentrations can be taken as an indication that a body of water is being used for waste disposal. Infiltration of groundwater into sewers adjacent to saltwater is also a potential source of high chlorides as well as sulfates.

Alkalinity

Alkalinity in wastewater results from the presence of the hydroxides $[OH^-]$, carbonates $[CO_3^{2-}]$, and bicarbonates $[HCO_3^-]$ of elements such as calcium, magnesium, sodium, potassium, and ammonia. Of these, calcium and magnesium bicarbonates are most common. Borates, silicates, phosphates, and similar compounds can also contribute to the alkalinity. The alkalinity in wastewater helps to resist changes in pH caused by the addition of acids. Wastewater is normally alkaline, receiving its alkalinity from the water supply, the groundwater, and the materials added during domestic use. The concentration of alkalinity in wastewater is important where chemical and biological treatment is to be used (see Chaps. 6 and 7, respectively), in biological nutrient removal (see Chap. 8), and where ammonia is to be removed by air stripping (see Chap. 11 and 15).

Alkalinity is determined by titrating against a standard acid; the results are expressed in terms of calcium carbonate, mg/L as $CaCO_3$. For most practical purposes alkalinity can be defined in terms of molar quantities, as

$$\text{Alk, eq/m}^3 = \text{meq/L} = [HCO_3^-] + 2[CO_3^{2-}] + [OH^-] - [H^+] \qquad (2\text{–}32)$$

The corresponding expression in terms of equivalents is

$$\text{Alk, eq/m}^3 = (HCO_3^-) + (CO_3^{2-}) + (OH^-) - (H^+) \qquad (2\text{–}33)$$

In practice, alkalinity is expressed in terms of calcium carbonate. To convert from meq/L to mg/L as $CaCO_3$ it is helpful to remember that

$$\begin{aligned}\text{Milliequivalent mass of } CaCO_3 &= \frac{(100\text{ mg/mmole})}{(2\text{ meq/mmole})} \qquad (2\text{–}34)\\ &= 50\text{ mg/meq}\end{aligned}$$

Thus 3 meq/L of alkalinity would be expressed as 150 mg/L as $CaCO_3$.

$$\begin{aligned}\text{Alkalinity, Alk as } CaCO_3 &= \frac{3.0\text{ meq}}{\text{L}} \times \frac{50\text{ mg } CaCO_3}{\text{meq } CaCO_3}\\ &= 150\text{ mg/L as } CaCO_3\end{aligned}$$

Nitrogen

The elements nitrogen and phosphorus, essential to the growth of microorganisms, plants, and animals, are known as nutrients or biostimulants. Trace quantities of other elements, such as iron, are also needed for biological growth, but nitrogen and phosphorus are, in most cases, the major nutrients of importance. Because nitrogen is an essential building block in the synthesis of protein, nitrogen data will be required to evaluate the treatability of wastewater by biological processes. Insufficient nitrogen can necessitate the addition of nitrogen to make the waste treatable. Nutrient requirements for biological waste treatment are discussed in Chaps. 7 and 8. Where control of algal growths in the receiving water is necessary, removal or reduction of nitrogen in wastewater prior to discharge may be desirable.

Sources of Nitrogen. The principal sources of nitrogen compounds are (1) the nitrogenous compounds of plant and animal origin, (2) sodium nitrate, and (3) atmospheric nitrogen. Ammonia derived from the distillation of bituminous coal is an example of nitrogen obtained from decayed plant material. Sodium nitrate ($NaNO_3$) is found principally in mineral deposits in Chile and in the manure found in sea birds' rookeries. The production of nitrogen from the atmosphere is termed *nitrogen fixation*. Because fixation is a biologically mediated process and because $NaNO_3$ deposits are relatively scarce, most sources of nitrogen in soil/groundwater are of biological origin.

Forms of Nitrogen. The chemistry of nitrogen is complex because of the several oxidation states that nitrogen can assume and the fact that changes in the oxidation state can be brought about by living organisms. To complicate matters further, the oxidation state changes brought about by bacteria can be either positive or negative depending upon whether aerobic or anaerobic conditions prevail. The oxidation states of nitrogen are summarized (Sawyer et al., 2003):

$$\begin{array}{ccccccc} -\mathrm{III} & 0 & \mathrm{I} & \mathrm{II} & \mathrm{III} & \mathrm{IV} & \mathrm{V} \\ NH_3 — & N_2 — & N_2O — & NO — & N_2O_3 — & NO_2 — & N_2O_5 \end{array} \tag{2–37}$$

The most common and important forms of nitrogen in wastewater and their corresponding oxidation state in the water/soil environment are ammonia (NH_3, $-$III), ammonium (NH_4^+, $-$III), nitrogen gas (N_2, 0), nitrite ion (NO_2^-, +III), and nitrate ion (NO_3^-, +V). The oxidation state of nitrogen in most organic compounds is $-$III.

Total nitrogen, as reported in Table 2–6, is composed of organic nitrogen, ammonia, nitrite, and nitrate. The organic fraction consists of a complex mixture of compounds including amino acids, amino sugars, and proteins (polymers of amino acids). The compounds that comprise the organic fraction can be soluble or particulate. The nitrogen in these compounds is readily converted to ammonium through the action of microorganisms in the aquatic or soil environment. Urea, readily converted to ammonium carbonate, is seldom found in untreated municipal wastewaters.

Organic nitrogen is determined analytically using the Kjeldahl method. The aqueous sample is first boiled to drive off the ammonia, and then it is digested. During digestion the organic nitrogen is converted to ammonium through the action of heat and acid. Total Kjeldahl nitrogen (TKN) is determined in the same manner as organic nitrogen, except that the ammonia is not driven off before the digestion step. Total Kjeldahl nitrogen is, therefore, the total of the organic and ammonia nitrogen. An alternative method is the perfulfate digestion procedure in which organic nitrogen is oxidized to nitrate nitrogen at high

Table 2–6

Definition of the various terms used to define various nitrogen species

Form of nitrogen	Abbrev.	Definition
Ammonia gas	NH_3	NH_3
Ammonium ion	NH_4^+	NH_4^+
Total ammonia nitrogen	TAN[a]	$NH_3 + NH_4^+$
Nitrite	NO_2^-	NO_2^-
Nitrate	NO_3^-	NO_3^-
Total inorganic nitrogen	TIN[a]	$NH_3 + NH_4^+ + NO_2^- + NO_3^-$
Total Kjeldahl nitrogen	TKN[a]	Organic N $+ NH_3 + NH_4^+$
Organic nitrogen	Organic N[a]	TKN $- (NH_3 + NH_4^+)$
Total nitrogen	TN[a]	Organic N $+ NH_3 + NH_4^+ + NO_2^- + NO_3^-$

[a] All species expressed as N.

temperatures (100–110°C in an autoclave for 1 hour) following the addition of potassium sulfate and sodium hydroxide. If the sample contains NH_4–N, NO_2–N, and NO_3–N, their concentrations are determined by other means so that these concentrations can be subtracted from the persulfate digestion total nitrogen concentration to determine the organic nitrogen concentration.

Ammonia nitrogen exists in as aqueous solution as either the ammonium ion (NH_4^+) or ammonia gas (NH_3), depending on the pH of the solution, in accordance with the following equilibrium reaction:

$$NH_4^+ \rightleftarrows NH_3 + H^+ \tag{2–38}$$

Applying the law of mass action [Eq. (2–7)] to Eq. (2–38) yields

$$\frac{[NH_3][H^+]}{[NH_4^+]} = K_a \tag{2–39}$$

where K_a = acid ionization (dissociation) constant at 25°C $= 10^{-9.25}$ or 5.62×10^{-10}

Because the distribution of the ammonia species is a function of the pH, the percentage ammonia can be determined using the following relationship:

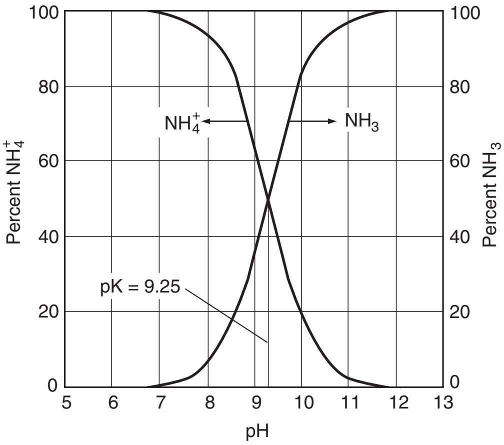
$$NH_3,\% = \frac{[NH_3] \times 100}{[NH_3] + [NH_4^+]} = \frac{100}{1 + [NH_4^+]/[NH_3]} = \frac{100}{1 + [H^+]/K_a} \tag{2–40}$$

Using Eq. (2–40) the distribution of the ammonia species as a function of pH is shown on Fig. 2–15. At pH levels above 7, the equilibrium is displaced to the left; at levels below pH 7, the ammonium ion is predominant. Ammonia is determined by raising the pH, distilling off the ammonia with the steam produced when the sample is boiled, and condensing the steam that absorbs the gaseous ammonia. The measurement is made colorimetrically, titrimetrically, or with specific-ion electrodes.

Nitrite nitrogen is relatively unstable and is easily oxidized to the nitrate form. It is an indicator of past pollution in the process of stabilization and seldom exceeds 1 mg/L in wastewater or 0.1 mg/L in surface waters or groundwaters. Although present in low concentrations, nitrite can be very important in wastewater or water pollution studies because it is extremely toxic to most fish and other aquatic species. Nitrites present in wastewater effluents are oxidized by chlorine and thus increase the chlorine dosage requirements and the cost of disinfection.

Figure 2–15

Distribution of ammonia (NH_3) and ammonium ion (NH_4^+) as a function of pH at 25°C.



Nitrate nitrogen is the most highly oxidized form of nitrogen found in wastewaters. Where secondary effluent is to be reclaimed for groundwater recharge, the nitrate concentration is important. The U.S. EPA drinking water standards (U.S. EPA, 1977) limit it to 10 mg/L as NO_3–N because of its serious and occasionally fatal effects on infants. Nitrates may vary in concentration from 0 to 20 mg/L as N in wastewater effluents. Assuming complete nitrification has taken place, the typical range found in treated effluents is from 15 to 25 mg/L as N.

Nitrous oxide (N_2O) is a greenhouse gas with an impact that is about 300 times that for CO_2 (U.S. EPA, 2008). Agriculture accounts for the main source of N_2O production due to human activity, although other anthropogenic sources include combustion of fossil fuels and manufacturing of nylon (Maier et al., 2009). Nitrous oxide can also be converted photolytically to nitric oxide (NO), which contributes to the depletion of the Earth's ozone layer. The production of N_2O and NO from biological wastewater treatment nitrification and denitrification processes is of interest and is addressed in Chapter 7. Nitrogen dioxide (NO_2) is a reddish-brown toxic gas that can be formed at high temperatures by oxidation of N_2 by oxygen, with internal combustion engines and power stations being major sources.

Nitrogen Fractions in Wastewater. As biological nutrient removal has become more common, information on the various organic nitrogen fractions has become important. The principal fractions are particulate and soluble. In biological treatment studies, the particulate and soluble fractions of organic nitrogen are fractionated further to assess wastewater treatability (see discussion in Sec. 8–2 in Chap. 8). Fractions that have been used include (1) ammonia (ionized and free), (2) biodegradable soluble organic nitrogen, (3) biodegradable particulate organic nitrogen, (4) nonbiodegradable soluble organic nitrogen, and (5) nonbiodegradable particulate organic nitrogen. The presence of nonbiodegradable forms of nitrogen often make meeting extremely low discharge limits for nitrogen difficult, if not impossible, without additional treatment steps. Unfortunately, there is little standardization on the definition of soluble versus particulate organic nitrogen (see discussion of under Solids in Sec. 2–3). Where filtration is the technique used to fractionate the sample, the relative distribution between soluble and particulate organic nitrogen will vary depending on the pore size of the filter used. In many cases, colloidal organic nitrogen has been classified as soluble or dissolved. The lack of standardized definition will also affect other aggregate constituents (i.e., chemical oxygen demand and total organic carbon).

In areas where surface water quality has been impaired due to eutrophication, more strict effluent permit requirements are imposed and an effluent total nitrogen (TN) concentration goal of less than 3.0 g/m^3 has been commonly applied. In such cases, the effluent soluble organic nitrogen (SON) concentration may account for over 40 percent of the effluent TN concentration. In addition the nondegradable SON persists in advanced biological nutrient removal processes, and has been shown to increase with increasing aeration times (Makinia et al., 2011). The inorganic forms of nitrogen are readily available for algal growth, but only recently has the SON impact on algal growth been investigated. From algal bioassay work, Haizhou et al. (2012) have shown that 20–40 percent of effluent SON is not readily available for algae growth and the nonavailable portion is likely related to more hydrophobic and possibly humic and higher molecular weight compounds

Nitrogen Pathways in Nature. The various forms of nitrogen that are present in nature and the pathways by which the forms are changed in an aquatic environment are depicted on Fig. 2–16. The nitrogen present in fresh wastewater is primarily combined in proteinaceous matter and urea. Decomposition by bacteria readily changes the organic form to ammonia. The age of wastewater is indicated by the relative amount of ammonia that is present. In an aerobic environment, bacteria can oxidize the ammonia nitrogen to nitrites and nitrates. The predominance of nitrate nitrogen in wastewater indicates that the waste has been stabilized with respect

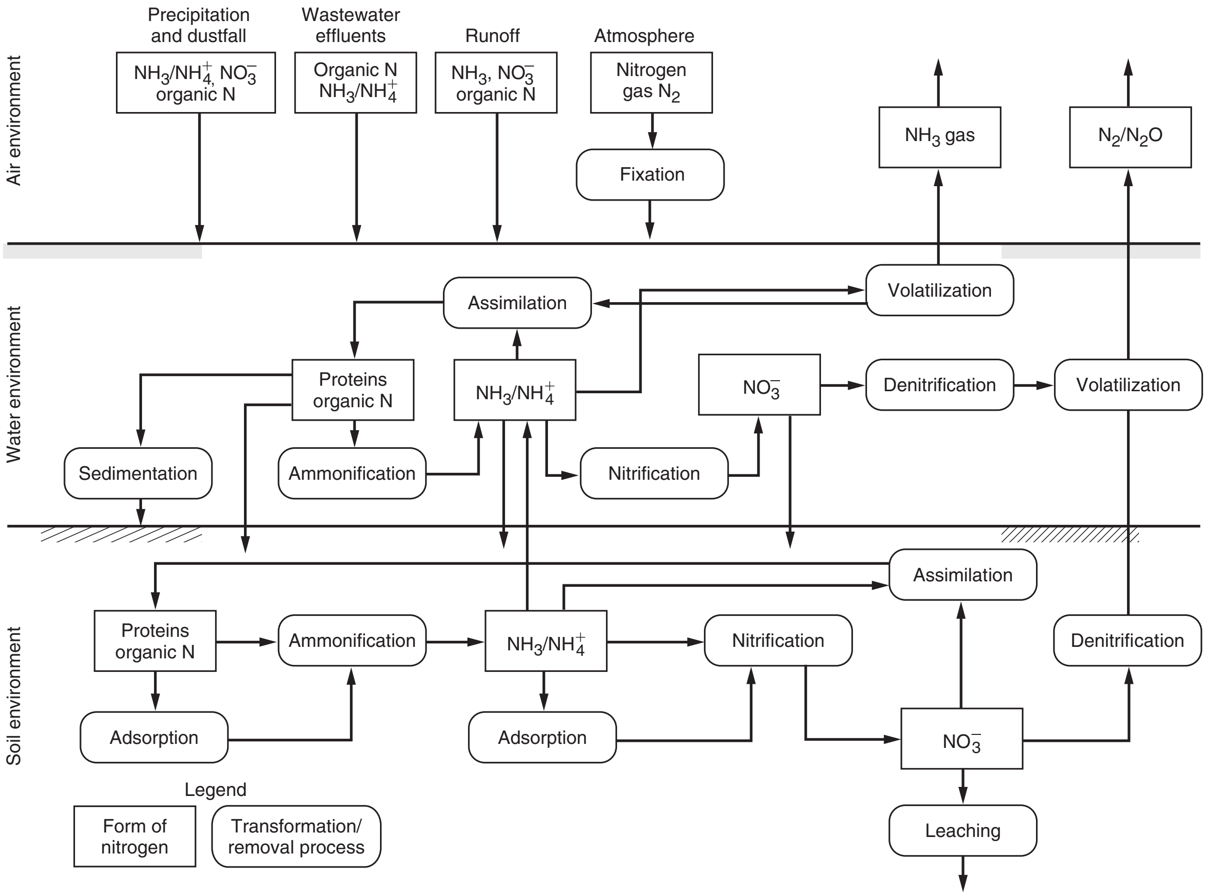


Figure 2–16

Generalized nitrogen cycle in the environment.

to oxygen demand. Nitrates, however, can be used by plants and animals to form protein. Death and decomposition of the plant and animal protein by bacteria again yield ammonia. Thus, if nitrogen in the form of nitrates can be reused to make protein by algae and other plants, it may be necessary to remove or reduce the nitrogen that is present to prevent these growths.

Phosphorus

Phosphorus is also essential to the growth of algae, agricultural crops, and other biological organisms. Unlike nitrogen, phosphorus does not have a gaseous form that can be discharged to the atmosphere. Because of noxious algal blooms that occur in surface waters, there is presently much interest in controlling the amount of phosphorus compounds that enter surface waters in domestic and industrial waste discharges and natural runoff. Further, the unsustainable nature of phosphorus mining and the potential future shortages is driving the development of processes for the recovery of phosphorus from wastewater as discussed in Chap. 15. Municipal wastewaters, for example, typically contain from 3.7 to 11 mg/L of phosphorus as P (see Table 3–18 in Chap. 3).

Phosphorus in Wastewater. Phosphorus in wastewater can be classified broadly into two fractions: particulate and dissolved. Each of these fractions can be differentiated further as reactive and nonreactive. Reactive phosphorus is defined as the forms of phosphorus that respond to colorimetric tests without preliminary hydrolysis or oxidative digestion. Note that these so-called reactive orthophosphates include both soluble forms as well as those that are loosely attached or adsorbed on to precipitates. The nonreactive forms include acid hydrolyzable and digestable forms, which may be an organic form of nonreactive phosphorus. The soluble forms of phosphorus that are found in aqueous solutions include the orthophosphates (reactive), polyphosphate (acid hydrolysable), and organic (digestable) phosphate. The orthophosphates (e.g., PO_4^{3-}, HPO_4^{2-}, $H_2PO_4^-$, H_3PO_4) are available for biological metabolism without further breakdown. The polyphosphates include those molecules with two or more phosphorus atoms, oxygen atoms, and in some cases, hydrogen atoms combined in a complex molecule. Polyphosphates undergo hydrolysis in aqueous solutions and revert to the orthophosphate forms; however, this hydrolysis is usually quite slow. The organically bound phosphorus is usually of minor importance in most domestic wastes, but it can be an important constituent of industrial wastes and wastewater sludges.

Soluble Nonreactive Forms. The soluble nonreactive forms of phosphorus are of great interest because they are not removed easily using current biological and chemical treatment processes. Effluent soluble nonreactive phosphorus concentrations in the range of 0.004 to 0.042 g/m^3 were found after high chemical dosing in tertiary filtration or membrane separation processes (Gu et al., 2011). Thus, the presence of nonreactive forms of phosphorus, as with nonbiodegradable nitrogen, can make meeting extremely low discharge permit limits difficult to achieve. As with SON, the impact of the remaining phosphorus after extensive chemical treatment and particulate removal on algal growth is also of interest. Based on algal bioassay testing of tertiary treated effluent samples with high alum dose it was found that not all of the effluent phosphorus measured was readily available to algal growth (Li and Brett, 2012). As the treatment level and alum addition increased, the percent bioavailable phosphorus for algal growth decreased from 60 percent for an effluent total phosphorus concentration of 0.50 g/m^3 to 15 percent at an effluent concentration of 0.02 g/m^3.

Sulfur

The sulfate ion occurs naturally in most water supplies and is present in wastewater as well. Sulfur is required in the synthesis of proteins and is released in their degradation. Sulfate is reduced biologically under anaerobic conditions to sulfide which, in turn, can combine with hydrogen to form hydrogen sulfide (H_2S). The following generalized reactions are typical:

$$\text{Organic matter} + SO_4^{2-} \xrightarrow{\text{bacteria}} S^{2-} + H_2O + CO_2 \tag{2–41}$$

$$S^{2-} + 2H^+ \rightarrow H_2S \tag{2–42}$$

If lactic acid is used as the precursor organic compound, the reduction of sulfate to sulfide occurs as follows:

$$\underset{\text{lactic acid}}{2CH_3CH(OH)COOH} + \underset{\text{sulfate}}{SO_4^{2-}} \xrightarrow{\text{bacteria}} \underset{\text{acetate}}{2CH_3COOH} + \underset{\substack{\text{sulfide}\\ \text{ion}}}{S^{2-}} + 2H_2O + 2CO_2 \tag{2–43}$$

Hydrogen sulfide gas, which will diffuse into the headspace above the wastewater in sewers that are not flowing full, tends to collect at the crown of the pipe. The accumulated H_2S can then be oxidized biologically to sulfuric acid, which is corrosive to concrete sewer pipes. This corrosive effect, known as "crown rot," can seriously threaten the structural integrity of the sewer pipe (ASCE, 1989; U.S. EPA, 1985a).

Sulfates are reduced to sulfides in sludge digesters and may upset the biological process if the sulfide concentration exceeds 200 mg/L. Fortunately, such concentrations are rare. The H_2S gas, which is evolved and mixed with the wastewater gas ($CH_4 + CO_2$), is corrosive to the gas piping, and if burned in gas engines, the products of combustion can damage the engine and severely corrode exhaust gas heat recovery equipment, especially if allowed to cool below the dew point.

Gases

Gases commonly found in untreated wastewater include nitrogen (N_2), oxygen (O_2), carbon dioxide (CO_2), hydrogen sulfide (H_2S), ammonia (NH_3), and methane (CH_4). The first three are common gases of the atmosphere and will be found in all waters exposed to air. The latter three are derived from the decomposition of the organic matter present in wastewater and are of concern with respect to worker health and safety. Although not found in untreated wastewater, other gases with which the environmental engineer must be familiar include chlorine (Cl_2) and ozone (O_3) (for disinfection and odor control), and the oxides of sulfur and nitrogen (in combustion processes). The following discussion is limited to those gases that are of interest in untreated wastewater. Under most circumstances, the ammonia in untreated wastewater will be present as the ammonium ion (see "Nitrogen"). However, before discussing the individual gases, it will be useful to review the ideal gas law, to consider the solubility of gases in water, and to review Henry's law as applied to the gases of interest.

Solubility of Gases in Water. The actual quantity of a gas that can be present in solution is governed by (1) the solubility of the gas as defined by Henry's law, (2) the partial pressure of the gas in the atmosphere, (3) the temperature, and (4) the concentration of the impurities in the water (e.g., salinity, suspended solids, etc.).

The Ideal Gas Law. The ideal gas law, derived from a consideration of Boyle's Law (volume of a gas is inversely proportional to pressure at constant temperature) and Charles' law (volume of a gas is directly proportional to temperature at constant pressure) is

$$PV = nRT \tag{2–44}$$

where P = absolute pressure, atm
V = volume occupied by the gas, L, m^3
n = moles of gas, mole
R = universal gas law constant, 0.082057 atm·L/mole·K
= 0.000082057 atm·m^3/mole·K
T = temperature, K (273.15 + °C)

Using the universal gas law it can be shown that the volume of gas occupied by one mole of a gas at standard temperature [0°C, (32°F)] and pressure (1.0 atm) is equal to 22.414 L.

$$V = \frac{nRT}{P}$$

$$V = \frac{(1\text{ mole})(0.082057\text{ atm}\cdot\text{L/mole}\cdot\text{K})[(273.15 + 0)\text{K}]}{1.0\text{ atm}} = 22.414\text{ L}$$

The following relationship, based on the ideal gas law, is used to convert between gas concentrations expressed in ppm_v and $\mu g/m^3$.

$$\mu\text{g/m}^3 = \frac{(\text{concentration, ppm}_v)(\text{mw, g/mole of gas})(10^6\,\mu\text{g/g})}{(22.414 \times 10^{-3}\,\text{m}^3/\text{mole of gas})} \tag{2–45}$$

The application of Eq. (2–45) is illustrated in Example 2–6.

EXAMPLE 2–6 **Conversion of Gas Concentration Units** The off gas from a wastewater force main (i.e., pressure sewer) was found to contain 9 ppm_v (by volume) of hydrogen sulfide (H_2S). Determine the concentration in $\mu g/m^3$ and in mg/L at standard conditions (0°C, 101.325 kPa).

Solution

1. Compute the concentration in $\mu g/L$ using Eq. (2–45).
   The molecular weight of $H_2S = [2(1.01) + 32.06] = 34.08$

$$9\ \text{ppm}_v = \left(\frac{9\ \text{m}^3}{10^6\ \text{m}^3}\right)\left[\frac{(34.08\ \text{g/mole H}_2\text{S})}{(22.4 \times 10^{-3}\ \text{m}^3\text{/mole of H}_2\text{S})}\right]\left(\frac{10^6\ \mu\text{g}}{g}\right) = 13{,}693\ \mu\text{g/m}^3$$

2. The concentration in mg/L is

$$13{,}693\ \mu\text{g/m}^3 = \left(\frac{13{,}693\ \mu\text{g}}{\text{m}^3}\right)\left(\frac{1\ \text{mg}}{10^3\ \mu\text{g}}\right)\left(\frac{1\ \text{m}^3}{10^3\ \text{L}}\right) = 0.0137\ \text{mg/L}$$

Comment If gas measurements, expressed in $\mu g/L$, are made at other than standard conditions, the concentration must be corrected to standard conditions, using the ideal gas law, before converting to ppm.

Henry's Law for Dissolved Gases. The equilibrium or saturation concentration of gas dissolved in a liquid is a function of the type of gas and the volume fraction or partial pressure of the gas in contact with the liquid. The relationship between the mole fraction of the gas in the atmosphere above the liquid and the mole fraction of the gas in the liquid is given by the following form of Henry's law:

$$p_g = \frac{H}{P_T} x_g \tag{2–46}$$

where p_g = mole fraction of gas in air, mole gas/mole of air

H = Henry's law constant, $\dfrac{\text{atm (mole gas/mole air)}}{\text{(mole gas/mole water)}}$

P_T = total pressure, usually 1.0 atm

x_g = mole fraction of gas in water, mole gas/mole water

$$= \frac{\text{mole gas } (n_g)}{\text{mole gas } (n_g) + \text{mole water } (n_w)}$$

In Eq. (2–46) it is helpful to remember that the mole fraction of a gas corresponds to the partial pressure or volume fraction of the gas. If the partial pressure of the gas is used, Eq. (2–46) is written as follows:

$$P_g = Hx_g \tag{2–47}$$

where P_g = partial pressure of gas, atm
other terms = as defined above

In practice and in the literature, the Henry's law constant is often reported as atm, with the mole fraction being implied. Henry's law constant is a function of the type of gas, temperature, and nature of the liquid. Values of Henry's law constants for various gases in water at 20°C are given in Table 2–7. It is important to note that reported values of the

Table 2–7

Henry's law constants at 20°C, unitless Henry's law constants at 20°C, and temperature dependent coefficients[a]

Parameter	Henry's constant, atm	Henry's constant, unitless	Temperature coefficients	
			A	*B*
Air	66,400	49.68	557.60	6.724
Ammonia	0.75	5.61×10^{-4}	1887.12	6.315
Carbon dioxide	1420	1.06	1012.40	6.606
Carbon monoxide	53,600	40.11	554.52	6.621
Chlorine	579	0.43	875.69	5.75
Chlorine dioxide	1500	1.12	1041.77	6.73
Hydrogen	68,300	51.10	187.04	5.473
Hyrogen sulfide	483	0.36	884.94	5.703
Methane	37,600	28.13	675.74	6.880
Nitrogen	80,400	60.16	537.62	6.7392
Oxygen	41,100	30.75	595.27	6.644
Ozone	5300	3.97	1268.24	8.05
Sulfur dioxide	36	2.69×10^{-2}	1207.85	5.68

[a] Adapted in part from Crittenden et al. (2012), Cornwell (1990), and Hand et al. (1998).

Henry's law constant found in the literature will vary depending on the date of the reference and the specific method used to estimate the constant. Use of the data in Table 2–7 is illustrated in Example 2–7.

The change in the Henry's law constant with temperature can be estimated using the following empirical equation, derived from a consideration of the van't Hoff-Arrhenius relationship:

$$\log_{10} H = \frac{-A}{T} + B \tag{2–48}$$

where H = Henry's law constant at temperature T, atm
A = empirical constant that takes into account the enthalpy change in water due to the dissolution of a component in water and the universal gas law constant
T = temperature, K = 273.15 + °C
B = empirical constant

Values of A and B for various gases of interest in wastewater treatment are presented in Table 2–7. It should be noted that the values given in Table 2–7 for A and B are approximate and will vary depending on the source and the method used to derive them.

Unitless Form of Henry's Law. In the literature, the unitless form of Henry's law is often used to compute the solubility of trace gases in water or wastewater. The unitless form is usually written as

$$\frac{C_g}{C_s} = H_u \tag{2–49}$$

where C_g = concentration of constituent in gas phase, $\mu g/m^3$, mg/L
C_s = saturation concentration of constituent in liquid, $\mu g/m^3$, mg/L
H_u = Henry's law constant, unitless

The unitless form is obtained by noting that at 1.0 atm pressure and 0°C, the volume occupied by 1.0 mole of air is 22.414 L. At other temperatures, 1.0 mole of air is equal to 0.082 T L of air, where T is temperature in Kelvin, K (273.15 + °C). Using these conversions, the unitless form of Henry's law is

$$H_u = \left[H\frac{\text{atm (mole gas/mole air)}}{\text{(mole gas/mole water)}}\right]\left(\frac{\text{mole air}}{0.082\,T\,\text{L}}\right)\left(\frac{\text{L}}{55.6\text{ mole water}}\right)$$

$$H_u = \left(\frac{H}{4.559\,T}\right) \tag{2–50}$$

For example, at 20°C, H_u equals

$$H_u = \left[\frac{H}{4.559(273.15 + 20)}\right] = H \times (7.49 \times 10^{-4}) \text{ at } 20°\text{C}$$

If atmospheric conditions prevail and Henry's constant is expressed in terms of atm·m³/mole (another form of Henry's law used commonly in the literature), the unitless form of Henry's law is obtained as follows:

$$H_u = \frac{H}{RT} \tag{2–51}$$

where H_u = Henry's law constant, unitless as used in Eq. (2–49)
H = Henry's law constant values expressed in atm·m³/mole
R = universal gas law constant, 0.00008205 atm·m³/mole·K
T = temperature, K = 273.15 + °C

EXAMPLE 2–7 **Saturation Concentration of Oxygen in Water** What is the saturation of oxygen in water in contact with dry air at 1 atm and 20°C?

Solution–Method 1 Using Eq. (2–46)

1. Dry air contains about 21 percent oxygen by volume (see Appendix B). Therefore, $p_g = 0.21$ mole O_2/mole air
2. Determine x_g.
   a. From Table 2–7, at 20°C, Henry's constant is

$$H = 4.11 \times 10^4 \frac{\text{atm (mole gas/mole air)}}{\text{(mole gas/mole water)}}$$

   b. Using Eq (2–46), the value of x_g is

$$\begin{aligned} x_g &= \frac{P_T}{H}p_g \\ &= \frac{1.0\text{ atm}}{4.11 \times 10^4 \dfrac{\text{atm (mole gas/mole air)}}{\text{(mole gas/mole water)}}}(0.21\text{ mole gas/mole air}) \\ &= 5.11 \times 10^{-6}\text{ mole gas/mole water} \end{aligned}$$

3. One liter of water contains 1000 g/(18 g/mole) = 55.6 mole, thus

$$\frac{n_g}{n_g + n_w} = 5.11 \times 10^{-6}$$
$$\frac{n_g}{n_g + 55.6} = 5.11 \times 10^{-6}$$

Because the number of moles of dissolved gas in a liter of water is much less than the number of moles of water,

$$n_g + 55.6 \approx 55.6$$

and $n_g \approx (55.6)5.11 \times 10^{-6}$

$$n_g \approx 2.84 \times 10^{-4} \text{ mole O}_2\text{/L}$$

4. Determine the saturation concentration of oxygen.

$$C_s \approx \frac{\left(\dfrac{2.84 \times 10^{-4} \text{ mole O}_2}{\text{L}}\right)\left(\dfrac{32 \text{ g}}{\text{mole O}_2}\right)\left(\dfrac{10^3 \text{ mg}}{1 \text{ g}}\right)}{(1 \text{ g}/10^3 \text{ mg})}$$
$$\approx 9.09 \text{ mg/L}$$

Solution–Method 2 Using Eq. (2–49)

1. The density of air at 20°C from Appendix B is 1204 kg/m^3.
2. The percent of oxygen in air from Appendix B is about 23.18 percent oxygen by weight.
3. Determine the saturation concentration of oxygen.
   a. From Table 2–7, at 20°C, the unitless form of Henry's constant is

   $$H_u = 30.75$$

   b. Using Eq (2–49), the value of C_s is

   $$C_s = \frac{C_g}{H_u}$$
   $$C_s = \frac{(1.204 \text{ kg/m}^3)(10^3 \text{ g/kg})(0.2318)}{30.75}$$
   $$= 9.08 \text{ g/m}^3 = 9.08 \text{ mg/L}$$

Comment The computed values (9.09 and 9.08 mg/L) are essentially the same as the value given in Appendix E (9.09 mg/L). It should be noted that the values for the Henry's law constant given in Table 2–7 will vary depending on the source and the method used to derive them. Also, the relationship at different temperatures is not linear.

Oxygen (O_2). Dissolved oxygen (DO) is required for the respiration of aerobic microorganisms as well as all other aerobic life forms. However, O_2 is only slightly soluble in water. The actual quantity of O_2 (and other gases too) that can be present in a solution is governed by (1) the solubility of the gas, (2) the partial pressure of the gas in the atmosphere, (3) the temperature, and (4) the concentration of the impurities in the water (e.g., salinity, suspended solids). The interrelationship of these variables is delineated in Chap. 6 and is illustrated in Appendix E, where the effect of temperature and salinity on DO concentration is presented.

Because the rate of biochemical reactions that use O_2 increases with increasing temperature, dissolved oxygen levels tend to be more critical in the summer months. The problem is compounded in summer months because stream flows are usually lower, and thus the total quantity of O_2 available is also lower. The presence of DO in wastewater is desirable because it prevents the formation of noxious odors. The role of O_2 in wastewater treatment is discussed in Chaps. 5, 7, 8, and 9.

Hydrogen Sulfide (H_2S). Hydrogen sulfide is formed, as mentioned previously, from the anaerobic decomposition of organic matter containing sulfur or from the reduction of mineral sulfites and sulfates. It is not formed in the presence of an abundant supply of oxygen. This gas is a colorless, inflammable compound having the characteristic odor of rotten eggs. Hydrogen sulfide is also toxic, and great care must be taken in its presence. High concentrations of H_2S can overwhelm olfactory glands, resulting in a loss of smell. This loss of smell can lead to a false sense of security that is very dangerous. The blackening of wastewater and sludge usually results from the formation of H_2S that has combined with the iron present to form ferrous sulfide (FeS). Various other metallic sulfides are also formed. Although H_2S is the most important gas formed from the standpoint of odors, other volatile compounds such as indol, skatol, and mercaptans, which may also be formed during anaerobic decomposition, may cause odors far more offensive than that of H_2S.

Methane (CH_4). The principal by-product from the anaerobic decomposition of the organic matter in wastewater is methane gas (see Chaps. 10 and 13). Methane is a colorless, odorless, combustible hydrocarbon of high fuel value. Normally, large quantities of CH_4 are not encountered in untreated wastewater because even small amounts of oxygen tend to be toxic to the organisms responsible for the production of CH_4. Occasionally, however, as a result of anaerobic decay in accumulated bottom deposits, CH_4 has been produced. Because methane is highly combustible and the explosion hazard is high, access ports (manholes) and sewer junctions or junction chambers where there is an opportunity for gas to collect should be ventilated with a portable blower during and before the time required for operating personnel to work in them for inspection, renewals, or repairs. In treatment plants, CH_4 is produced from the anaerobic treatment process used to stabilize wastewater sludges (see Chap 13). In treatment plants where CH_4 is produced, notices should be posted about the plant warning of explosion hazards, and plant employees should be instructed in safety measures to be maintained while working in and about the structures where CH_4 may be present. Methane is also a serious greenhouse gas with an impact of over 25 times that of CO_2 (U.S. EPA, 2008).

Odors

Odors in domestic wastewater are usually caused by gases produced by the decomposition of organic matter or by substances added to the wastewater. Fresh wastewater has a distinctive, somewhat disagreeable odor, which is less objectionable than the odor of wastewater which has undergone anaerobic (devoid of oxygen) decomposition. The most characteristic odor of stale or septic wastewater is that of hydrogen sulfide, which, as discussed previously, is produced by anaerobic microorganisms that reduce sulfate to sulfide. Industrial wastewater may contain either odorous compounds or compounds that produce odors during the process of wastewater treatment. The management of odors from wastewater treatment plants is considered in Chap. 16.

Public Concern. Odors have been rated as the foremost concern of the public relative to the implementation of wastewater treatment facilities. Within the past few years, the control of odors has become a major consideration in the design and operation of wastewater collection, treatment, and disposal facilities, especially with respect to the public acceptance of these facilities. In many areas, projects have been rejected because of the concern over the potential for odors. In view of the importance of odors in the field of wastewater management, it is appropriate to consider the effects they produce, how they are detected, and their characterization and measurement.

Effects of Odors. The importance of odors at low concentrations in human terms is related primarily to the psychological stress they produce rather than to the harm they do to the body. Offensive odors can cause poor appetite for food, lowered water consumption,

Table 2–8

Major odorous compounds and their corresponding odor thresholds associated with untreated wastewater[a]

Odorous compound	Chemical formula	Molecular weight	Odor threshold (typical), ppm_v[b]	Characteristic odor
Ammonia	NH_3	17.0	0.035–53 (1.5)	Ammoniacal, pungent
Chlorine	Cl_2	71.0	0.0095–4.7 (0.15)	Pungent, suffocating
Crotyl mercaptan	$CH_3—CH{=}CH—CH_2—SH$	90.19	0.00003	Skunk like
Dimethyl sulfide	$(CH_3)2^S$	62	0.0001–0.02 (0.002)	Decayed vegetables
Diphenyl sulfide	$(C_6H_5)2^S$	186	0.00005–0.005 (0.0004)	Unpleasant
Ethyl mercaptan	$CH_3(CH_2)SH$	62	0.000009–0.03 (0.0002)	Decayed cabbage
Hydrogen sulfide	H_2S	34	0.00007–1.4 (0.003)	Rotten eggs
Indole	C_8H_6NH	117	0.0001–0.0003 (0.0001)	Fecal, nauseating
Methyl amine	CH_3NH_2	31	0.02–8.7 (0.11)	Putrid, fishy
Methyl mercaptan	CH_3SH	48	0.00002–0.04 (0.0007)	Decayed cabbage
Skatole	C_9H_9N	131	0.00000007–0.05 (0.0002)	Fecal, nauseating
Sulfur dioxide	SO_2	64.07	0.009–5.0 (0.6)	Pungent, irritating
Thiocresol	$CH_3(C_6H_4)SH$	124	0.00006–0.001 (0.0002)	Skunk, rancid

[a] Adapted in part from Patterson et al. (1984) and U.S. EPA (1985a).

[b] Parts per million by volume.

impaired respiration, nausea and vomiting, and mental perturbation. In extreme situations, offensive odors can lead to the deterioration of personal and community pride, interfere with human relations, discourage capital investment, lower socioeconomic status, and deter growth. Also, some odorous compounds (e.g., H_2S) are toxic at elevated concentrations. These problems can result in a decline in market and rental property values, tax revenues, payrolls, and sales.

Detection of Odors by the Human Olfactory System. The odorous compounds responsible for producing psychological stress in humans are detected by the olfactory system, but the precise mechanism involved is at present not well understood. Since 1870, more than 30 theories have been proposed to explain olfaction. One of the difficulties in developing a universal theory has been the inadequate explanation of why compounds with similar structures may have different odors and why compounds with very different structures may have similar odors. At present, there appears to be some general agreement that the odor of a molecule must be related to the molecule as a whole as opposed to some functional group attached to a molecule. Over the years, many attempts have been made to classify odors in a systematic fashion. The major categories of offensive odors and the compounds involved are listed in Table 2–8 along with the corresponding threshold odor values. All of these compounds may be found or may develop in domestic wastewater, depending on local conditions.

Characterization of Odor. In the past, four independent factors were identified typically as being required for the complete characterization of an odor: character, detectability, hedonics, and intensity (see Table 2–9). More recently a fifth factor, persistence, has been added to the list (see Table 2–9). Detectability and persistence are considered further below.

Detectability. Odor, as shown on Fig. 2–17, can be measured (detected) by sensory methods and specific odorant concentrations can be measured by instrumental methods. The application of both methods of odor detection is considered below. In sensory methods, human subjects (often a panel of subjects) are exposed to odors that have been diluted with odor-free air, and the number of dilutions required to reduce an odor to its minimum

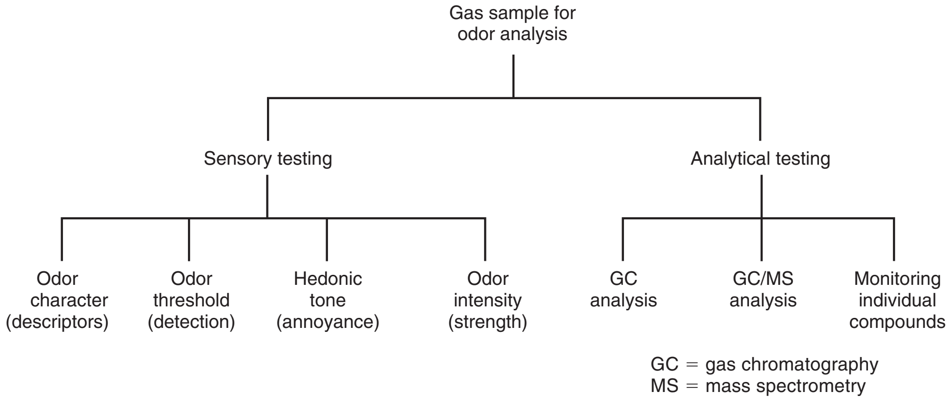
Table 2–9

Factors that must be considered for the complete characterization of an odor

Factor	Description
Character	Relates to the mental associations made by the subject in sensing the odor. Determination can be quite subjective. Typical odor descriptors are listed in the last column of Table 2–8.
Detectability (also threshold)	The number of dilutions required to reduce an odor to its minimum detectable threshold odor concentration (MDTOC).
Hedonics (tone)	The relative pleasantness or unpleasantness of the odor sensed by the subject.
Intensity	The preceived relative strength of the odor above the detection threshold. Usually measured by the butanol olfactometer or calculated from the D/T (dilutions to threshold ratio) when the relationship is established.
Persistence	The rate at which the odor intensity changes with concentration. Persistence can be represented as a dose response function.

Figure 2–17

Classification of methods used to detect odors.



detectable threshold odor concentration (MDTOC) are noted. The detectable odor concentration, reported as the dilutions to the MDTOC, commonly called *dilutions-to-threshold* (D/T) is given by the following expression:

$$\text{D/T} = \frac{\text{Volume of odor free air}}{\text{Volume of odorous air}} \tag{2–52}$$

Thus, if four volumes of odor free air must be added to one volume of odorous air to reduce the odorant to its MDTOC, the odor concentration would be reported as 4 D/T. It should be noted that a number of other measures are used to define the intensity of an odor including the ED_{50} value which represents the number of times an odorous air sample must be diluted before the average person (50 percentile) can barely detect an odor in the diluted sample. To date, detectability is the only factor that has been used in the development of statutory regulations for nuisance odors. The application of D/T values to assess odor impacts is considered in Sec. 16–3 in Chap. 16.

The threshold odor of a water or wastewater sample is determined by diluting the sample with odor-free water. Depending on the nature of the odorous constituents, the diluted sample can be heated to enhance the release of diluted odorous constituents. The "threshold odor number" (TON) corresponds to the greatest dilution of the sample with odor free water at which an odor is just perceptible. The recommended sample size is 200 mL. The numerical value of the TON is determined as follows:

$$\text{TON} = \frac{A + B}{A} \tag{2–53}$$

where TON = threshold odor number
A = mL of sample
B = mL of odor free water

The odor emanating from the liquid sample is determined as discussed above with human subjects (often a panel of subjects). Details for this procedure may be found in Standard Methods (2012).

Persistence. Persistence corresponds to the rate at which a perceived odor intensity decreases as the odor is diluted. Typically odor intensity is defined as

$$I = kC^n \tag{2–54}$$

where I = odor intensity, ppm$_v$ n-butanol
C = concentration of odor, number of dilutions
k, n = coefficients for each specific odor or combination of odors

Three odor intensity measurements at different dilutions are used to establish the dose response. When Eq. (2–54) is linearized and plotted, the slope of the line of best fit corresponds to n. Thus, as the slope of the line decreases, the odor is more persistent. The application of Eq. (2–54) is illustrated in Example 2–8.

Sensory Measurement of Odors. It has been shown that, under carefully controlled conditions, the sensory (organoleptic) measurement of odors by the human olfactory system can provide meaningful and reliable information. Therefore, the sensory method is often used to measure the odors emanating from wastewater-treatment facilities. The availability of a direct reading meter for hydrogen sulfide, described below, which can be used to detect concentrations accurately to 3 ppb and with less accuracy to 1 ppb, is a significant development.

Field Olfactometers. The sensory determination of the minimum threshold odor concentration is subject to a number of errors. Adaptation and cross adaptation, synergism, subjectivity, and sample modification (see Table 2–10) are the principal errors. To avoid errors in sample modification during storage in sample collection containers, direct reading field olfactometers have been developed to measure odors at their source without using sampling containers.

A field olfactometer is a hand-held device in which odorous air can be passed sequentially through a series of graduated orifices and mixed (diluted) with air that has been purified by passing through activated carbon. The orifices are typically sized to provide D/T values of 2, 4, 7, 15, 30, and so on. The dilution ratios are determined by the ratio of the size of the odorous to purified air inlets. Two commonly used field olfactometers, the Scentometer® (Barnebey-Cheney, 1987) and the Nasal Ranger® (St. Croix Sensory, 2006),

Table 2–10

Types of errors in the sensory detection of odors

Description	Type of error
Adaptation and cross adaptation	When exposed continually to a background concentration of an odor, the subject is unable to detect the presence of that odor at low concentrations. When removed from the background odor concentration, the subject's olfactory system will recover quickly. Ultimately, a subject with an adapted olfactory system will be unable to detect the presence of an odor to which his system has adapted.
Sample modification	Both the concentration and composition of odorous gases and vapors can be modified in sample collection containers and in odor detection devices. To minimize problems associated with sample modification, the period of odor containment should be minimized or eliminated, and minimum contact should be allowed with any reactive surfaces.
Subjectivity	When the subject has knowledge of the presence of an odor, random error can be introduced in sensory measurements. Often, knowledge of the odor may be inferred from other sensory signals such as sound, sight, or touch.
Synergism	When more than one odorant is present in a sample, it has been observed that it is possible for a subject to exhibit increased sensitivity to a given odor because of the presence of another odor.

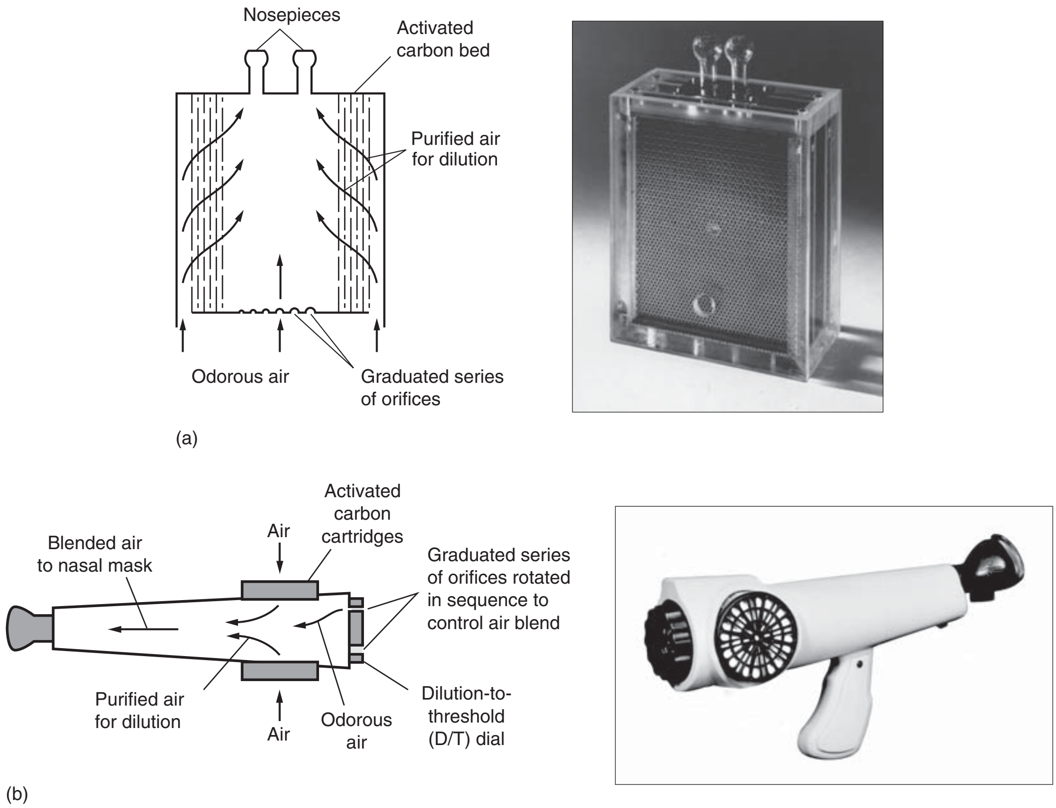


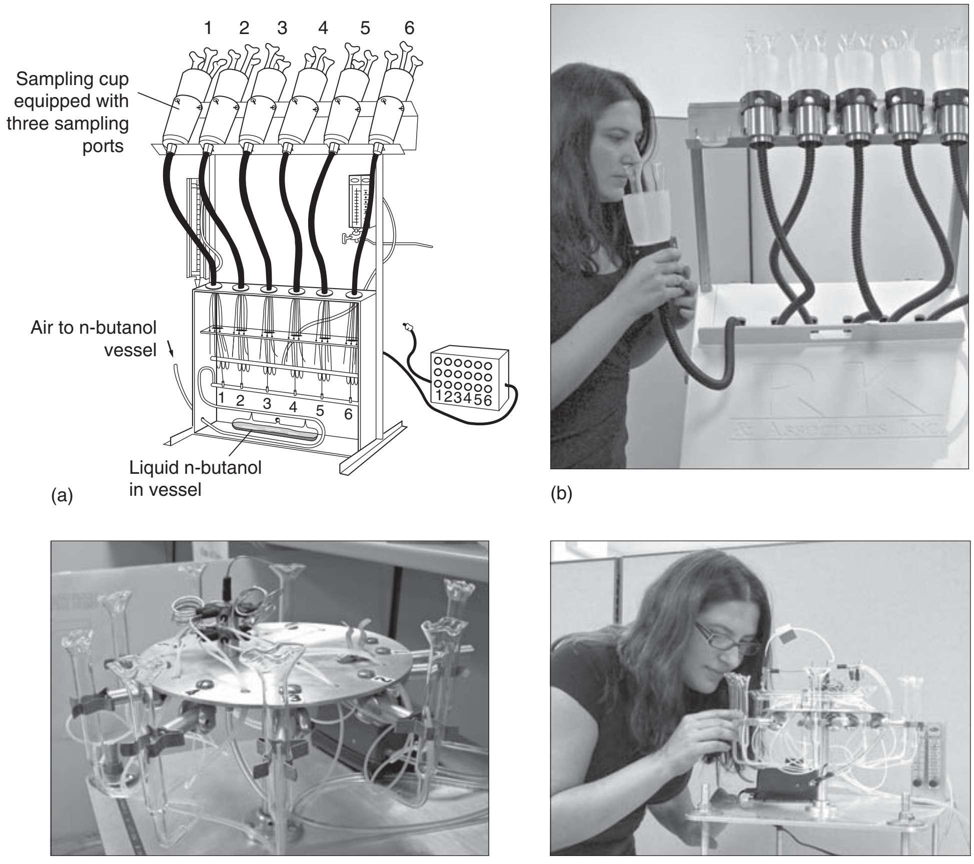
Figure 2–18

Examples of field hand-held olfactometers used for field odor studies: (a) Scentometer® schematic and front view looking at nose pieces (5 in × 6 in × 2.5 in, from Barnebey & Sutcliffe Corp.) and (b) Nasal Ranger® schematic and pictorial view (from St Croix Sensory Inc.).

are shown on Fig. 2–18. Field olfactometers are very useful for making odor determinations over a large area surrounding a treatment plant. Often a mobile odor laboratory, which contains several types of olfactory and analytical equipment in a single van type vehicle, is used for field sites.

Fixed Olfactometers. Equipment used in a laboratory setting to analyze odors includes (1) the triangle olfactometer, (2) the butanol wheel, and (3) a variety of other specialized olfactometers. The triangle olfactometer enables the operator to introduce the odorous air sample at different concentrations at five or six different cups each equipped with three sampling ports [see Fig. 2–19(a)]. At each cup, two ports contain purified air, and one port contains a diluted sample. Each odor panel member (usually six) then sniffs each of the three ports and *must* select the port he or she believes contains the sample [see Fig. 2–19(b)]. The procedure is repeated at the remaining four or five cups. The concentration of the odorous air is increased in successive cups, typically doubling in each successive cup (ASTM, 2004). The results are analyzed using a standardized statistical program based on signal detection theory (Green and Swets, 1966).

The Butanol Wheel is a device used to measure the intensity of an odor against various concentrations of n-butanol. The device comprises eight sampling ports located on a rotatable disk [see Fig. 2–19(c)]. Dilutions of n-butanol that increase by a factor of two are delivered to



(a) (b) (c) (d)

Figure 2–19

Examples of fixed olfactometers: (a) schematic of dynamic forced-choice triangle olfactometer, (b) panel member sniffing one of the three sample ports, (c) view of Butanol wheel, (d) panel member sniffing one of sample ports. [Figs. (b), (c) and (d) courtesy of RK & Associates, Inc.]

each successive port. Each odor panel member first sniffs the odorous sample being tested and then compares it to the various dilutions of n-butanol starting with Port 1 [see Fig. 2–19(d)]. The test is continued until the panelist identifies the n-butanol dilution which most closely matches the intensity of the odorous sample. The results are reported in ppmv n-butanol odor intensity. Application of Butonal Wheel test results is illustrated in Example 2–8.

EXAMPLE 2–8 **Determination of Relative Persistence** Intensity measurements were made at different dilutions for two odor samples. Using the data provided, determine which of the two odors is more persistent.

n-butanol odor intensity, ppm$_v$	Dilution-to-threshold, D/T	
	Sample A	Sample B
10,000	0	0
100	25	3.2
10	316	10
0	3160	32

Solution

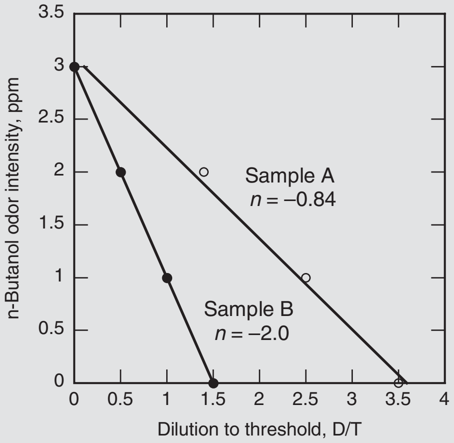
1. Linearize Eq. 2–54 and log transform the given data.
   a. The linearized form of Eq. 2–54 is

      $\log I = \log k + n \log C$

   b. The log transformed data are:

	log D/T	
Log *I*	Sample A	Sample B
3	0	0
2	1.4	0.5
1	2.5	1.0
0	3.5	1.5

3. Plot log *I* versus log *C* and determine the slope *n* to determine which sample is more persistent.
   a. The required plot is given below



   b. The slopes for the two samples are

      Sample A −0.84
      Sample B −2.0

   c. Based on the slopes, Sample A is more persistent than Sample B.

Many of the specialized laboratory olfactometers are designed to work in conjunction with instrumental methods of analysis. For example, the Gerstel ODP2® is used in conjunction with a GC or MS chromatograph for the detection of compounds that as they elute from the separation column. Thus, there is a simultaneous instrumental and olfactory characterization of an odorous compounds (Agus et al., 2011).

Instrumental Measurement of Odors. It is often desirable to know the specific compounds responsible for odor. Although gas chromatography has been used successfully for this purpose, it has not been used as successfully in the detection and quantification

Figure 2–20

Portable H_2S meter used for field odor studies. (From Arizona Instrument Corporation, Jerome Instrument Division.)

of odors derived from wastewater collection, treatment, and disposal facilities. Equipment developed and found useful in the chemical analysis of odors is the triple-stage quadrupole mass spectrometer. The spectrometer can be used as a conventional mass spectrometer to produce simple mass spectra or as a triple-stage quadrupole to produce collisionally activated disassociation spectra. The former operating mode provides the masses of molecular or parent ions present in samples, while the latter provides positive identification of compounds. Types of compounds that can be identified include ammonia, amino acids, and volatile organic compounds (Agus et al., 2011).

An instrumental method has been developed for the measurement of hydrogen sulfide with mid-range accuracy ranging from 0.003 to 25 ppm and with an overall range of 0.001 to 50 ppm. The portable AZI Jerome Model 631 shown on Fig. 2–20 utilizes an inline pump to pull air containing hydrogen sulfide over a gold film sensor for a fixed period of time. The sensor absorbs the hydrogen sulfide and the change in resistance of the gold film sensor is related to the mass concentration of hydrogen sulfide. The sensor is re-zeroed before the next sample cycle. Eventually, when the sensor becomes saturated, a heat cycle is initiated to remove the accumulated hydrogen sulfide from the sensor. In addition to the portable meter, the company also makes a variety of stationary units, which are designed to operate automatically.

2–5 METALLIC CONSTITUENTS

Trace quantities of many metals, such as cadmium (Cd), chromium (Cr), copper (Cu), iron (Fe), lead (Pb), manganese (Mn), and mercury (Hg), nickel (Ni), and zinc (Zn) are important constituents of most waters. Many of these metals are also classified as priority pollutants. However, most of these metals are necessary for growth of biological life, and absence of sufficient quantities of them could limit growth of algae, for example. The presence of any of these metals in excessive quantities will interfere with many beneficial uses of the water because of their toxicity; therefore, it is frequently desirable to measure and control the concentrations of these substances.

Sources of Metallic Constituents

The sources of trace metals in wastewater include the discharges from commercial and industrial activities, products used in residential applications such as cleaning agents and personal care products, and groundwater infiltration. Many of the sources of heavy metals are identified in Table 2–11. For example, cadmium, chromates, lead, and mercury are

Table 2–11

Typical metallic wastes produced by commercial, industrial, and agricultural activities that have been classified as priority pollutants

Name	Formula	Use	Concern
Arsenic	As	Alloying additive for metals, especially lead and copper, battery grids, cable sheaths, boiler tubes. High purity (semiconductor) grade	Carcinogen and mutagen. *Long term*—Sometimes can cause fatigue and loss of energy; dermatitis
Barium	Ba	Getter alloys in vacuum tubes, deoxidizer for copper, Frary's metal, lubricant for anode rotors in x-ray tubes, spark-plug alloys	Flammable at room temperature in powder form. *Long term*—Increased blood pressure and nerve block
Cadmium	Cd	Electrodeposited and dipped coatings on metals, bearing and low-melting alloys, brazing alloys, fire protection system, nickel-cadmium storage batteries power transmission wire, TV phosphors, basis of pigments used in ceramic glazes, machinery enamels, fungicide, photography and lithography, selenium rectifiers, electrodes for cadmium-vapor lamps and photoelectric cells	Flammable in powder form. Toxic by inhalation of dust or fume. A carcinogen. Soluble compounds of cadmium are highly toxic. *Long term*—Concentrates in the liver, kidneys, pancreas, and thyroid; hypertension suspected effect
Chromium	Cr	Alloying and plating element on metal and plastic substrates for corrosion resistance, chromium-containing and stainless steels, protective coating for automotive and equipment accessories, nuclear and high-temperature research, constituent of inorganic pigments	Hexavalent chromium compounds are carcinogenic and corrosive on tissue. *Long term*—Skin sensitization and kidney damage
Lead	Pb	Storage batteries, gasoline additive, cable covering, ammunition, piping, tank linings, solder and fusible alloys, vibration damping in heavy construction, foil, babbit and other bearing alloys	Toxic by ingestion or inhalation of dust or fumes. *Long term*—Brain and kidney damage; birth defects
Mercury	Hg	Amalgams, catalyst electrical apparatus, cathodes for production of chlorine and caustic soda, instruments, mercury vapor lamps, mirror coating, arc lamps, boilers	Highly toxic by skin absorption and inhalation of fume or vapor. *Long term*—toxic to central nervous system, may cause birth defects
Selenium	Se	Electronics, xerographic plates, TV cameras, photocells, magnetic computer cores, solar batteries, (rectifiers, relays), ceramics (colorant for glass) steel and copper, rubber accelerator, catalyst, trace element in animal feeds	*Long term*—Red staining of fingers, teeth, and hair; general weakness; depression; irritation of nose and mouth
Silver	Ag	Manufacture of silver nitrate, silver bromide, photo chemicals; lining vats and other equipment for chemical reaction vessels, water distillation, etc.; mirrors, electric conductors, silver plating electronic equipment; sterilant, water purification, surgical cements, hydration and oxidation catalyst, special batteries, solar cells, reflectors for solar towers, low temperatures brazing alloys, table cutlery, jewelry, dental medical and scientific equipment, electrical contacts, bearing metal, magnet windings, dental amalgams, colloidal silver used as a nucleating agent in photography and medicine, often combined with protein	Toxic metal. *Long term*—Permanent grey discoloration of skin, eyes, and mucus membranes

often present in industrial wastes. These are found particularly in metal-plating wastes and should be removed by pretreatment at the site of the industry rather than be mixed with the municipal wastewater. Fluoride, a toxic anion, is found commonly in wastewater from electronics manufacturing facilities.

Importance of Metals

Metals of importance in the treatment, reuse, and disposal of treated effluents and biosolids are summarized in Table 2–12. All living organisms require varying amounts (macro and micro) of metallic elements, such as iron, chromium, copper, zinc, and cobalt, for proper growth. Although macro and micro amounts of metals are required for proper growth, the same metals can be toxic when present in elevated concentrations. As more use is made of treated wastewater effluent for irrigation and landscape watering, the presence of a variety of

Table 2–12

Metals of importance in wastewater management[a]

		Nutrients necessary for biological growth				
Metal	**Symbol**	**Macro**	**Micro**[b]	**Concentration threshold of inhibitory effect on heterotrophic organisms, mg/L**	**Used to determine SAR**[a] **for land application of effluent**	**Used to determine if biosolids are suitable for land application**
Arsenic	As			0.05		✔
Cadmium	Cd			1.0		✔
Calcium	Ca	✔			✔	
Chromium	Cr		✔	10[c], 1[d]		
Cobalt	Co		✔			
Copper	Cu		✔	1.0		✔
Iron	Fe	✔				
Lead	Pb		✔	0.1		✔
Magnesium	Mg	✔	✔		✔	
Manganese	Mn		✔			
Mercury	Hg			0.1		✔
Molybdenum	Mo		✔			✔
Nickel	Ni		✔	1.0		✔
Potassium	K	✔				
Selenium	Se		✔			✔
Sodium	Na	✔			✔	
Tungsten	W		✔			
Vanadium	V		✔			
Zinc	Zn		✔	1.0		✔

[a] SAR = sodium adsorption ratio.

[b] Often identified as trace elements needed for biological growth.

[c] Total chromium.

[d] Hexavalent chromium.

metals must be determined to assess any adverse affects that may occur. Calcium, magnesium, and sodium are of importance in determining the sodium adsorption ratio (SAR) which is used to assess the suitability of treated effluent for agricultural use (see Asano, et al., 2007). Where composted sludge is applied in agricultural applications, the concentration of arsenic, cadmium, copper, lead, mercury, molybdenum, nickel, selenium, and zinc must be determined.

Sampling and Methods of Analysis

Methods for determining the concentrations of these substances vary in complexity according to the interfering substances that may be present (Standard Methods, 2012). Metals are determined typically by flame atomic absorption, electrothermal atomic absorption, inductively coupled plasma, or IPC/mass spectrometry. Various classes of metals are defined as (1) *dissolved metals,* those metals present in unacidified samples that pass through a 0.45 μm membrane filter; (2) *suspended metals,* those metals present in unacidified samples that are retained on a 0.45 μm membrane filter; (3) *total metals,* the total of the dissolved and suspended metals or the concentration of metals determined on an unfiltered sample after digestion; and (4) *acid extractable metals,* those metals in solution after an unfiltered sample is treated with a hot dilute mineral acid (Standard Methods, 2012).

Typical Effluent Discharge Limits for Metals

Increasingly, metallic constituents in effluent discharges and in biosolids are being regulated. Typical discharge requirements for metals and other toxic constituents are reported in Table 2–13. In addition to complying with existing U.S. EPA requirements, many states have adopted more restrictive standards to protect specific beneficial uses.

2–6 AGGREGATE ORGANIC CONSTITUENTS

Organic compounds are normally composed of a combination of carbon, hydrogen, and oxygen, together with nitrogen in some cases. The organic matter in wastewater typically consists of proteins (40 to 60 percent), carbohydrates (25 to 50 percent), and oils and fats (8 to 12 percent). Urea, the major constituent of urine, is another important organic compound contributing to fresh wastewater. Because urea decomposes rapidly, urea is seldom found in other than very fresh wastewater. Because of the complex nature of wastewater, the organic characteristics of interest in wastewater are classified as aggregate and individual. Aggregate organic constituents are comprised of a number of individual compounds that cannot be or are not distinguished separately as opposed to constituents that are determined individually.

Sources of Aggregate Organic Constituents

Along with the proteins, carbohydrates, fats and oils, and urea, derived from food and human wastes, wastewater typically contains small quantities of a very large number of different synthetic organic molecules, with structures ranging from simple to extremely complex. Sources of synthetic organic molecules include unused medicine, personal care products, and household cleaning and maintenance products.

Measurement of Organic Content

In general, the analyses used to measure aggregate organic material may be divided into those used to measure gross concentrations of organic matter greater than about 1.0 mg/L and those used to measure trace concentrations in the range of 10^{-12} to 1 mg/L. Laboratory methods commonly used today to measure gross amounts of organic matter (typically

Table 2–13

Typical discharge limits for toxic constituents found in secondary effluent

Constituent	Units	Average value[a] Daily	Monthly
Arsenic	μg/L	20	
Cadmium	μg/L	1.1	
Chromium	μg/L	11	
Copper	μg/L	4.9	
Lead[b]	μg/L	5.6	
Mercury	μg/L	2.1	0.012
Nickel[b]	μg/L	7.1	
Selenium[b]	μg/L	5.0	
Silver	μg/L	2.3	
Zinc[b]	μg/L	58	
Dieldrin[c]	μg/L	0.0019	0.00014
Lindane	μg/L	0.16	0.063
Tributyltin	μg/L	0.01	0.005
PAHs[d,e]	μg/L	0.049	

[a] Limits apply to the average concentration of all samples collected during the averaging period (daily–24-h period; monthly-calendar month).

[b] Effluent limitation may be met as a 4-d average. If compliance is to be determined based on a 4-d average, then concentrations of four 24-h composite samples must be reported as well as the average of four.

[c] Compliance will be based on the practical quantification level (PQL), 0.07 μg/L.

[d] PAHs = polynuclear aromatic hydrocarbons.

[e] Compliance will be based on the practical quantification level (PQL) for each PAH, 4 μg/L.

Source: Bay Area Regional Water Quality Control Board, Oakland, CA.

greater than 1 mg/L) in wastewater include (1) biochemical oxygen demand (BOD), (2) chemical oxygen demand (COD), and (3) total organic carbon (TOC). Complementing these laboratory tests is the theoretical oxygen demand (ThOD), which is determined from the chemical formula of the organic matter.

Biochemical Oxygen Demand (BOD)

The most widely used parameter of organic pollution applied to both wastewater and surface water is the 5-d BOD (BOD_5). This determination involves the measurement of the dissolved oxygen used by microorganisms in the biochemical oxidation of organic matter. Despite the widespread use of the BOD test, it has a number of limitations, as discussed later in this section. It is hoped that, through the continued efforts of workers in the field, one of the other measures of organic content, or perhaps a new measure, will ultimately be used in its place. Why, then, if the test suffers from serious limitations, is further space devoted to it in this text? The reason is that BOD test results are now used (1) to determine the approximate quantity of oxygen that will be required to biologically stabilize the organic matter present, (2) to determine the size of waste treatment facilities, (3) to measure the efficiency of some treatment processes, and (4) to determine compliance with wastewater discharge permits. Because it is likely that the BOD test will continue to be used for some time, it is important to know the details of the test and its limitations.

Basis for BOD Test. If sufficient oxygen is available, the aerobic biological decomposition of an organic waste will continue until all of the waste is consumed. Three more or less distinct activities occur. First, a portion of the waste is oxidized to end products to obtain energy for cell maintenance and the synthesis of new cell tissue. Simultaneously, some of the waste is converted into new cell tissue using part of the energy released during oxidation. Finally, when the organic matter is used up, the new cells begin to consume their own cell tissue to obtain energy for cell maintenance. This third process is called endogenous respiration. Using the term COHNS (which represents the elements carbon, oxygen, hydrogen, nitrogen, and sulfur) to represent the organic waste and the term $C_5H_7NO_2$ to represent cell tissue, the three processes are defined by the following generalized chemical reactions:

Energy reaction (oxidation)

$$\text{COHNS} + O_2 + \text{bacteria} \rightarrow CO_2 + H_2O + NH_3 + \text{other end products} + \text{energy} \qquad (2\text{–}55)$$

Synthesis reaction

$$\text{COHNS} + O_2 + \text{bacteria} + \text{energy} \rightarrow \underset{\text{New cell tissue}}{C_5H_7NO_2} \qquad (2\text{–}56)$$

Endogenous respiration

$$C_5H_7NO_2 + 5O_2 \rightarrow 5CO_2 + NH_3 + 2H_2O \qquad (2\text{–}57)$$

If only the oxidation of the organic carbon that is present in the waste is considered, the ultimate BOD is the oxygen required to complete the three reactions given above. This oxygen demand is known as the ultimate *carbonaceous* or first-stage BOD and is usually denoted as UBOD.

As will be discussed later, the ammonia produced in the energy reaction, Eq. (2–55), can be oxidized further to nitrite and nitrate. Thus, the BOD test only represents the amount of oxygen need to oxidize the carbonaceous material in a sample.

Description of BOD Test Procedure. The standard BOD test [see Fig. 2–21(a)] involves placing a small sample of the wastewater in a BOD bottle (volume = 300 mL). The bottle is then filled with dilution water saturated in oxygen and containing the nutrients required for biological growth. To ensure that meaningful results are obtained, the sample must be suitably diluted with a specially prepared dilution water so that adequate nutrients and oxygen will be available during the incubation period. Normally, several dilutions are prepared to cover the complete range of possible values. Before stoppering the bottle, the oxygen concentration in the bottle is measured (see Fig. 2–22). When testing wastewaters with low concentrations of microorganisms, a seeded BOD test is conducted [see Fig. 2–21(b)]. The organisms contained in the effluent from primary sedimentation facilities are used commonly as the seed for the BOD test. Seed organisms can also be obtained commercially. When the sample contains a large population of microorganisms (e.g., untreated wastewater), seeding is not necessary.

The standard incubation period is usually five days at 20°C, but other lengths of time and temperatures can be used. After incubating for a period of 5-d at 20°C, the dissolved oxygen concentration is measured again. The BOD of the sample is the difference in the

Figure 2–21

Procedure for setting up BOD test bottles: (a) with unseeded dilution water and (b) with seeded dilution water (Tchobanoglous and Schroeded, 1985).

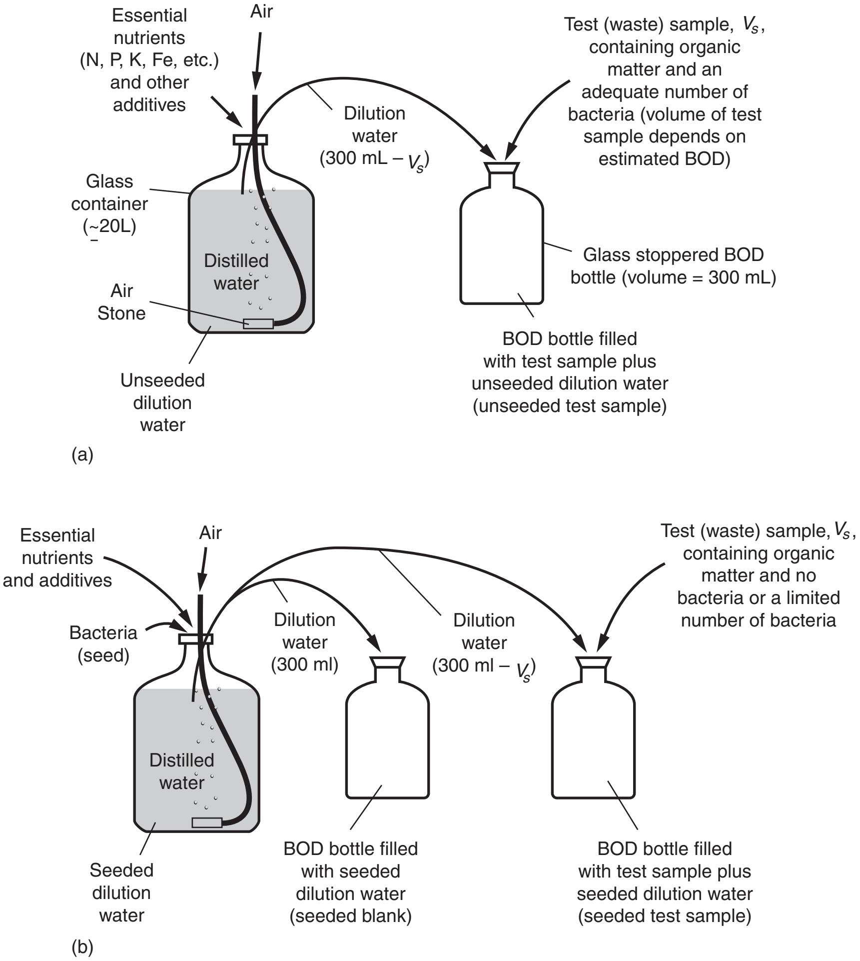
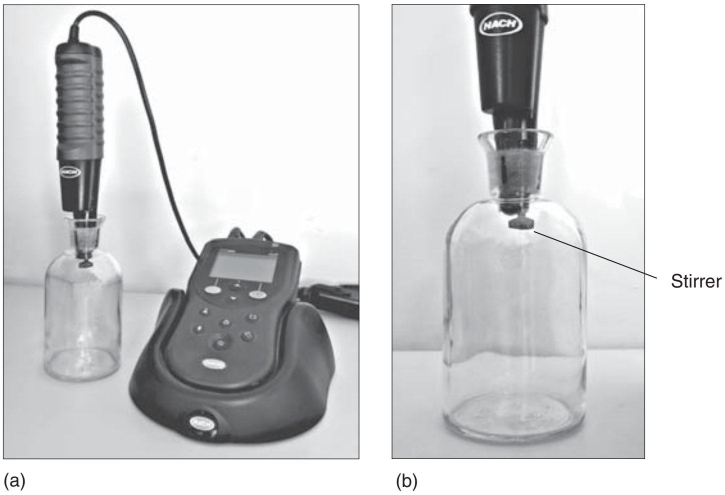


Figure 2–22

Measurement of oxygen in BOD bottle: (a) with a DO probe equipped with a stirring mechanism and (b) close up of stirrer.



dissolved oxygen concentration values, expressed in milligrams per liter, divided by the decimal fraction of sample used (Standard Methods, 2012). The computed BOD value is known as the 5-d, 20°C biochemical oxygen demand. The 5-d incubation period dates back to the use of the BOD test to assess river pollution in England the late 1800s. Because the maximum time of flow of any river in England from the headwaters to the ocean is 5 days, the 5-d period was, and is, utilized for the test.

Longer time periods (typically seven days), which correspond to work schedules, are often used, especially in small plants where the laboratory staff is not available on the weekends. The temperature, however, should be constant throughout the test. The 20°C temperature used is an average value for slow-moving streams in temperate climates and is easily duplicated in an incubator. Different results would be obtained at different temperatures because biochemical reaction rates are temperature dependent.

Modeling of BOD Reaction. The rate of BOD oxidation ("exertion") is modeled based on the assumption that the amount of organic material remaining at any time, t, is governed by a first order function (see Chap. 1).

$$\frac{d\text{BOD}_r}{dt} = k_1\text{BOD}_r \tag{2–58}$$

Integrating between the limits of UBOD and BOD_t and $t = 0$ and $t = t$ yields:

$$\text{BOD}_r = \text{UBOD}\,(e^{-k_1 t}) \tag{2–59}$$

where BOD_r = amount of waste remaining at time t (d) expressed in oxygen equivalents, mg/L
UBOD = the total or ultimate carbonaceous BOD, mg/L
k_1 = first-order reaction rate constant, 1/d
t = time, d

Thus the BOD exerted up to time t is given by

$$\text{BOD}_t = \text{UBOD} - \text{BOD}_r = \text{UBOD} - \text{UBOD}(e^{-k_1 t}) = \text{UBOD}(1 - e^{-k_1 t}) \tag{2–60}$$

Equation (2–60) is the standard expression used to define the BOD for wastewater. The basis for this equation is discussed in Sec. 1–5 in conjunction with the analysis of a batch reactor. It should be noted that in the literature dealing with the characterization of wastewater, the terms L and BOD_u are often used to denote ultimate carbonaceous BOD (UBOD).

Biochemical oxidation theoretically takes an infinite time to go to completion because the rate of oxidation is assumed to be proportional to the amount of organic matter remaining. Within a 20-d period, the oxidation of the carbonaceous organic matter is about 95 to 99 percent complete, and in the 5-d period used for the BOD test, oxidation is from 60 to 70 percent complete.

BOD Reaction Rate Coefficients. The value of k_1 for untreated wastewater is generally about 0.12 to 0.46 d^{-1} (base e), with a typical value of about 0.23 d^{-1}. The range of k_1 values for effluents from biological treatment processes is from 0.12 to 0.23 d^{-1}. For a given wastewater, the value of k_1 at 20°C can be determined experimentally by observing the variation with time of the dissolved oxygen in a series of incubated samples. If k_1 at 20°C is equal to 0.23 d^{-1}, the 5-d oxygen demand is about 68 percent of the ultimate first-stage demand. Occasionally, the first-order reaction rate constant will be expressed in log (base 10) units. The relationship between k_1 (base e) and K_1 (base 10) is as follows:

$$K_1(\text{base } 10) = \frac{k_1(\text{base } e)}{2.303} \tag{2–61}$$

As discussed above, the temperature at which the BOD of a wastewater sample is determined is usually 20°C. It is possible, however, to determine the reaction constant k at a temperature other than 20°C using the following relationship developed in the discussion on the effects of temperature in Chap. 1, Sec. 1–6:

$$\frac{k_2}{k_1} = \theta^{(T_2 - T_1)} \tag{1–44}$$

The value of the temperature coefficient θ has been found to vary from 1.056 in the temperature range between 20 and 30°C to 1.135 in the temperature range between 4 and 20°C (Schroepfer et al., 1964). A value of θ often quoted in the literature is 1.047 (Phelps, 1944), but it has been observed that this value does not apply at cold temperatures (e.g., below 20°C). Equation (2–60), along with Eq. (1–44), makes it possible to convert test results from different time periods and temperatures to the standard 5-d 20°C test, as illustrated in Example 2–9.

EXAMPLE 2–9 **Calculation of Different BOD Values** Determine the 1-d BOD and ultimate first-stage BOD for a wastewater whose 5-d, 20°C BOD is 200 mg/L. What would have been the 5-d BOD if the test had been conducted at 25°C? The reaction constant k (base e) = 0.23 d^{-1}, and $\theta = 1.047$.

Solution

1. Determine the ultimate carbonaceous BOD.

$$\text{BOD}_5 = \text{UBOD} - \text{BOD}_r = \text{UBOD}(1 - e^{-k_1 t})$$
$$200 = \text{UBOD}(1 - e^{-0.23\times 5}) = \text{UBOD}(1 - 0.317)$$
$$\text{UBOD} = 293 \text{ mg/L}$$

2. Determine the 1-d BOD.

$$\text{BOD}_t = \text{UBOD}(1 - e^{-k_1 t})$$
$$\text{BOD}_1 = 293(1 - e^{-0.23\times 1}) = 293(1 - 0.795) = 60.1 \text{ mg/L}$$

3. Determine the 5-d BOD at 25°C.

$$k_{1_T} = k_{1_{20}}(1.047)^{T-20}$$
$$k_{1_{25}} = 0.23(1.047)^{25-20} = 0.29 \text{ d}^{-1}$$
$$\text{BOD}_5 = \text{UBOD}(1 - e^{-k_1 t}) = 293(1 - e^{-0.29\times 5}) = 224 \text{ mg/L}$$

For polluted water and wastewater, a typical value of k_1 (base e at 20°C) is 0.23 d^{-1} (K_1, base 10, = 0.10 d^{-1}). The value of the reaction rate constant varies significantly, however, with the type of waste. The range may be from 0.05 to 0.3 d^{-1} (base e) or more. For the same ultimate BOD, the oxygen uptake will vary with time and with different reaction rate constant values (see Fig. 2–23).

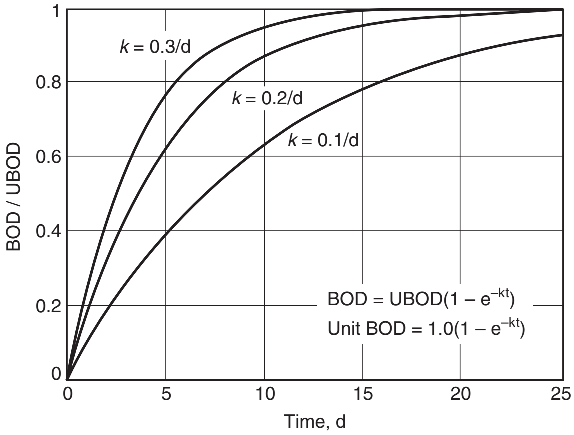
Nitrification in the BOD Test. Noncarbonaceous matter, such as ammonia, is produced during the hydrolysis of proteins. It is now known that a number of bacteria are capable of oxidizing ammonia to nitrite and subsequently to nitrate. The generalized reactions are as follows:

Conversion of ammonia to nitrite (as typified by *Nitrosomonas*):

$$NH_3 + 3/2O_2 \rightarrow HNO_2 + H_2O \tag{2–62}$$

Figure 2–23

Effect of the rate constant k_1 on BOD (for a unit UBOD value).



Conversion of nitrite to nitrate (as typified by *Nitrobacter*):

$$HNO_2 + 1/2O_2 \rightarrow HNO_3 \tag{2–63}$$

Overall conversion of ammonia to nitrate:

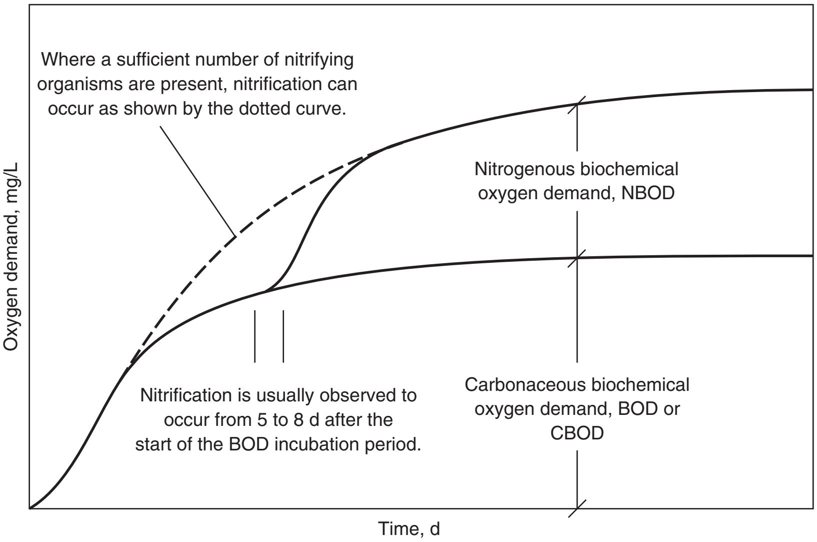
$$NH_3 + 2O_2 \rightarrow HNO_3 + H_2O \tag{2–64}$$

The oxygen demand associated with the oxidation of ammonia to nitrate is called the nitrogenous biochemical oxygen demand (NBOD). The normal exertion of the oxygen demand in a BOD test for a domestic wastewater is shown on Fig. 2–24. Because the reproductive rate of the nitrifying bacteria is slow, it normally takes from 6 to 10 d for them to reach significant numbers to exert a measurable oxygen demand. However, if a sufficient number of nitrifying bacteria is present initially, the interference caused by nitrification can be significant.

When nitrification occurs in the BOD test, erroneous interpretations of treatment operating data are possible. For example, assume the effluent BOD from a biological treatment process is 20 mg/L without nitrification and 40 mg/L with nitrification. If the influent BOD to the treatment process is 200 mg/L, then the corresponding BOD removal efficiency

Figure 2–24

Definition sketch for the exertion of the carbonaceous and nitrogenous biochemical oxygen demand in a waste sample.

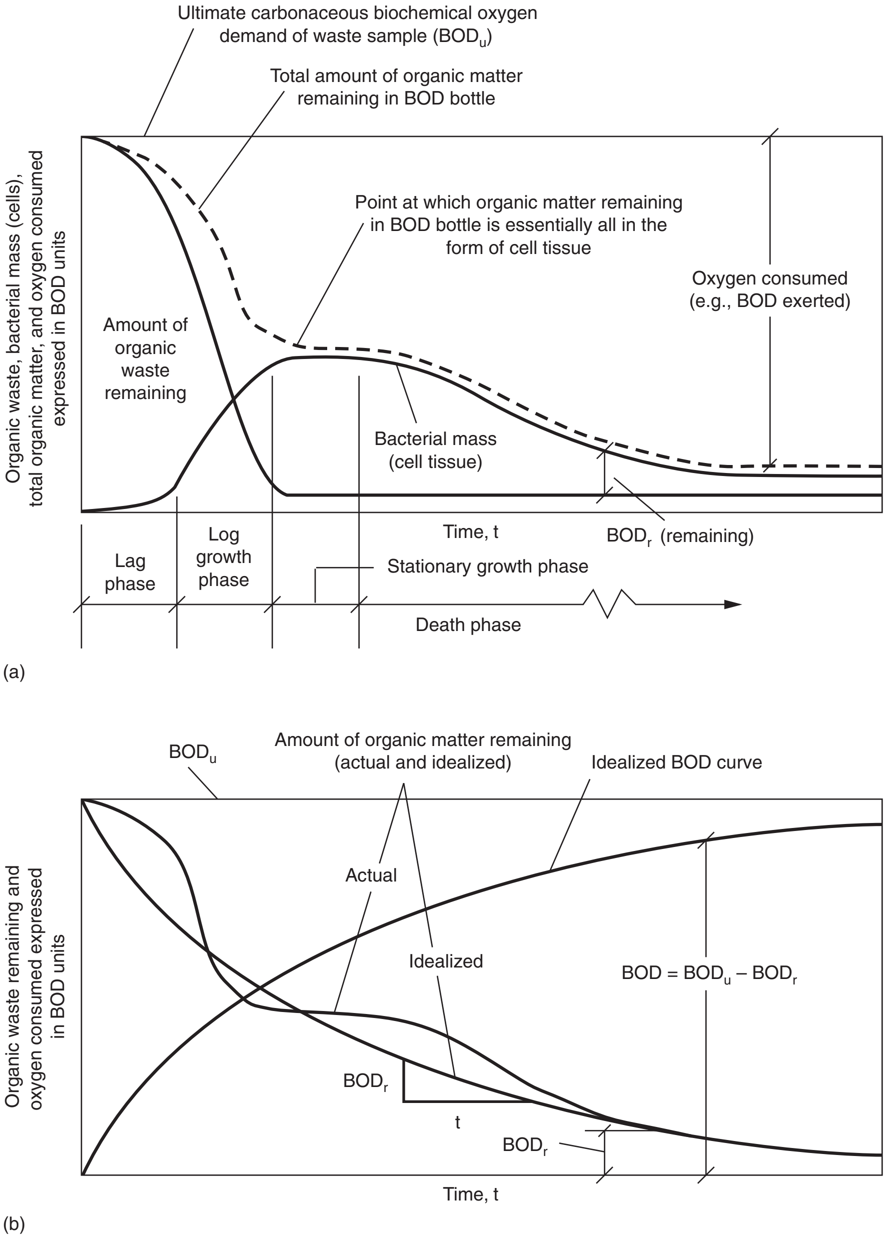


would be reported as 90 and 80 percent without and with nitrification, respectively. Thus, if nitrification is occurring but is not suspected, it might be concluded the treatment process is not performing well, when in actuality it is performing quite well.

Carbonaceous Biochemical Oxygen Demand. When nitrification occurs, the measured BOD value will be higher than the true value due to the oxidation of carbonaceous material (see Fig. 2–25). If a given percentage of carbonaceous biochemical oxygen demand (CBOD) removal must be achieved to meet regulatory permit limits, early nitrification can pose a serious problem. The effects of nitrification can be overcome either by using

Figure 2–25

Functional analysis of the BOD test: (a) interrelationship of organic waste, bacterial mass (cell tissue), total organic waste, and oxygen consumed in BOD test and (b) idealized representation of the BOD test (Tchobanoglous and Schroeder, 1985).



various chemicals to suppress the nitrification reactions, or by treating the sample to eliminate the nitrifying organisms (Young, 1973). Pasteurization and chlorination/dechlorination are two methods that have also been used to suppress the nitrifying organisms.

When the nitrification reaction is suppressed, the resulting BOD is known as the carbonaceous biochemical oxygen demand (CBOD). In effect, the CBOD is a measure of the oxygen demand exerted by the oxidizable carbon in the sample. The CBOD test, in which the nitrification reaction is suppressed chemically, should only be used on samples that contain small amounts of organic carbon (e.g., treated effluent). Errors have sometimes been observed in the measured BOD values when the CBOD test is used on wastewater containing significant amounts of organic matter such as untreated wastewater.

Analysis of BOD Data. The value of k is needed if the BOD_5 is to be used to obtain UBOD, the ultimate or 20-d BOD. The usual procedure followed when these values are unknown is to determine k_1 and UBOD from a series of BOD measurements. There are several ways of determining k_1 and UBOD from the results of a series of BOD measurements, including the method of least-squares, the method of moments (Moore et al., 1950), the daily-difference method (Tsivoglou, 1958), the rapid-ratio method (Sheehy, 1960), the Thomas method (Thomas, 1942, 1950) and the Fujimoto method (Fujimoto, 1961). The least-squares and Fujimoto method are illustrated in the 4th edition of this textbook (Tchobanoglous et al., 2003).

Effect of Particle Size on BOD Reaction Rates. If a separation and analysis technique, such as membrane filtration (see Figs. 2–4 and 2–8), is used to quantify the size distribution of the solids in the influent wastewater, the various size fractions can be correlated to observed oxygen (BOD) uptake rates, determined using a respirometer. As reported in Table 2–14, the observed BOD reaction rate coefficients are affected significantly by the size of the particles in wastewater. Based on the data given in Table 2–14, it is clear that the treatment of a wastewater can be affected by modifying the particle size distribution. Further, wastewaters with significantly different particle size distributions will respond differently, depending on the method of treatment (e.g., in constructed wetlands).

Limitations in the BOD Test. The limitations of the BOD test are as follows: (1) a high concentration of active, acclimated seed bacteria is required; (2) pretreatment is needed when dealing with toxic wastes, and the effects of nitrifying organisms must be reduced; (3) only the biodegradable organics are measured; (4) the test does not have stoichiometric validity after the soluble organic matter present in solution has been used (see Fig. 2–25); and (5) the relatively long period of time required to obtain test results. Of the above, perhaps the most serious limitation is that the 5-d period may or may not correspond to the point where the soluble organic matter that is present has been used. The lack of stoichiometric validity at all times reduces the usefulness of the test results.

Table 2–14

Effect of the size of the biodegradable particles found in wastewater on observed BOD reaction rates[a]

Fraction	Size range μm	k (base 10), d^{-1}
Settleable	>100	0.08
Supracolloidal	1–100	0.09
Colloidal	0.1–1.0	0.22
Soluble	<0.1	0.39

[a] Adapted from Balmat (1957).

Total and Soluble Chemical Oxygen Demand (COD and SCOD)

The COD test is used to measure the oxygen equivalent of the organic material in wastewater that can be oxidized chemically using dichromate in an acid solution, as illustrated in the following equation when the organic nitrogen is in the reduced state (oxidation number $= -3$) (Sawyer et al., 2003).

$$C_nH_aO_bN_c + dCr_2O_7^{2-} + (8d + c)H^+ \rightarrow nCO_2 + \frac{a + 8d - 3c}{2}H_2O + cNH_4^+ + 2dCr^{3+} \tag{2–65}$$

where $d = \frac{2n}{3} + \frac{a}{6} - \frac{b}{3} - \frac{c}{2}$

Although it would be expected that the value of the ultimate carbonaceous BOD would be as high as the COD, this is seldom the case. Some of the reasons for the observed differences are as follows: (1) many organic substances which are difficult to oxidize biologically, such as lignin, can be oxidized chemically, (2) inorganic substances that are oxidized by the dichromate increase the apparent organic content of the sample, (3) certain organic substances may be toxic to the microorganisms used in the BOD test, and (4) high COD values may occur because of the presence of inorganic substances with which the dichromate can react. From an operational standpoint, one of the main advantages of the COD test is that it can be completed in about 2.5 h, compared to 5 or more d for the BOD test. To reduce the time further, a rapid COD test that takes only about 15 min has been developed.

As new methods of biological treatment have been developed, especially with respect to biological nutrient removal, it has become more important to fractionate the COD. The principal fractions are particulate and soluble COD. In biological treatment studies, the particulate and soluble fractions are fractionated further to assess wastewater treatability (see discussion in Chap. 8, Sec. 8–2). Fractions that have been used include (a) readily biodegradable soluble COD, (b) slowly biodegradable colloidal and particulate (enmeshed) COD, (c) nonbiodegradable soluble COD, and (d) nonbiodegradable colloidal and particulate COD. The readily biodegradable soluble COD is often fractionated further into complex COD that can ferment to volatile fatty acids (VFAs) and short chain VFAs (see Fig. 8–4 in Chap. 8). Unfortunately, as noted previously, there is little standardization on the definition of soluble versus particulate COD. Where filtration is the technique used to fractionate the sample, the relative distribution between soluble and particulate COD will vary greatly depending on the pore size of the filter. An alternative method used to determine the soluble COD involves precipitation of the suspended solids and a portion of the colloidal material. The COD of the clarified liquid corresponds to the soluble COD.

Total and Dissolved Organic Carbon (TOC and DTOC)

The TOC test, done instrumentally, is used to determine the total organic carbon in an aqueous sample. The test methods for TOC utilize heat and oxygen, ultraviolet radiation, chemical oxidants, or some combination of these methods to convert organic carbon to carbon dioxide which is measured with an infrared analyzer or by other means. The TOC of a wastewater can be used as a measure of its pollution characteristics and in some cases it has been possible to relate TOC to BOD and COD values. The TOC test is also gaining in favor because it takes only 5 to 10 min to complete. If a valid relationship can be established between results obtained with the TOC test and the results of the BOD test for a given wastewater, use of the TOC test for process control is recommended.

A continuous online TOC analyzer has been developed, in conjunction with the space program, which can be used to detect TOC concentrations in the ppb (parts per billion) range.

Such instruments are used to detect the residual TOC in the treated effluent from microfiltration and reverse osmosis (RO) treatment units. Continuous TOC measurements may be used to monitor the performance of the full-scale RO units, to be used in conjunction with repurification projects in which repurified effluent is proposed to be blended with other waters.

Along with COD, it has also become more important to fractionate the TOC. The principal fractions are particulate TOC and dissolved (soluble) TOC (DTOC). As with COD, the particulate and soluble TOC fractions are fractionated further to assess treatability. It should be noted that the pore size of the filter paper recommended in Standard Methods (2012) for differentiating between dissolved and particulate TOC is 0.45 μm, in contrast to the pore size (2.0 μm or less) used to define TSS and TDS. Again, because of the pore size of the filter paper used, the colloidal material that passes through the filter will be classified as dissolved. Because of the interest in the chemical constituents that make up the DTOC, advanced methods of analysis have been developed to quantify the constituent groupings as illustrated on Fig. 2–26.

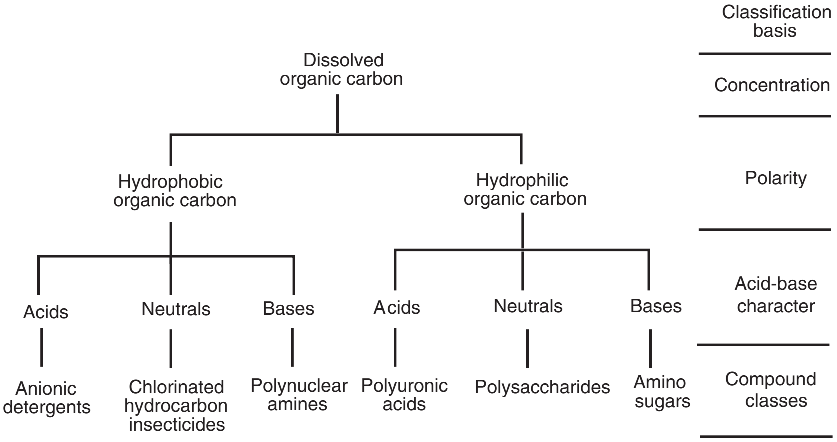
UV-Absorbing Organic Constituents

A number of organic compounds are found in wastewater including humic substances, lignin, tannin, and various aromatic compounds, that strongly absorb ultraviolet (UV) radiation. As a result, UV absorption has been used as a surrogate measure for the organic compounds cited above. The UV wavelengths at which adsorption is determined are typically in the range from 200 to 400 nm, with the value of 254 nm being reported most commonly. The results of UV absorption measurements are reported in units of cm^{-1}, along with the pH and the UV wavelength (e.g., $\text{UV}_\lambda^{\text{pH}}$ where λ is the UV wavelength). This method has proven useful in assessing the aggregate presence of UV absorbing compounds in wastewater, although interfering compounds can render the test invalid.

The results of UV absorption measurements at a wavelength of 254 nm are also correlated to the amount of dissolved organic carbon (DOC) present in a sample that has been filtered through a filter with a pore size of 0.45 μm. The results are reported as the *specific ultraviolet adsorption* (SUVA) per mg/L of DOC. It should be noted that although the UV measurement is correlated to the DOC, SUVA is in fact a measure of the nature of the carbon in the sample being analyzed, more specifically the extent to which the carbon is aromatic. Thus, the SUVA test is used most commonly to distinguish between different water samples. The SUVA test has also been used to assess the potential for the formation of trihalomethanes (THMs) (see Sec. 2–7).

Figure 2–26

Procedure for the characterization of the organic fractions that comprise the TOC.



Theoretical Oxygen Demand (ThOD)

Organic matter of animal or vegetable origin in wastewater is generally a combination of carbon, hydrogen, oxygen, and nitrogen. The principal groups of these elements present in wastewater are, as previously noted, carbohydrates, proteins, oils and grease, and products of their decomposition. The biological decomposition of these substances is discussed in Chap. 7. If the chemical formula of the organic matter is known, the ThOD may be computed as illustrated in Example 2–10.

EXAMPLE 2–10 **Calculation of ThOD** Determine the ThOD for glycine [$CH_2(NH_2)COOH$] using the following assumptions:

1. In the first step, the organic carbon and nitrogen are converted to carbon dioxide (CO_2) and ammonia (NH_3), respectively.
2. In the second and third steps, the ammonia is oxidized sequentially to nitrite and nitrate.
3. The ThOD is the sum of the oxygen required for all three steps.

Solution

1. Write balanced reaction for the carbonaceous oxygen demand.

$$CH_2(NH_2)COOH + 3/2O_2 \rightarrow NH_3 + 2CO_2 + H_2O$$

2. Write balanced reactions for the nitrogenous oxygen demand.

$$\begin{array}{ll} \text{(a)} & NH_3 + 3/2O_2 \rightarrow HNO_2 + H_2O \\ \text{(b)} & \underline{HNO_2 + 1/2O_2 \rightarrow HNO_3 \qquad\quad} \\ & NH_3 + 2O_2 \quad \rightarrow HNO_3 + H_2O \end{array}$$

3. Determine the ThOD.

$$\begin{aligned} \text{ThOD} &= (3/2 + 4/2) \text{ mole } O_2/\text{mole glycine} \\ &= 7/2 \text{ mole } O_2/\text{mole glycine} \times 32 \text{ g/mole } O_2 \\ &= 112 \text{ g } O_2/\text{mole glycine} \end{aligned}$$

Interrelationships between BOD, COD, and TOC

Typical values for the ratio of BOD/COD for untreated municipal wastewater are in the range from 0.3 to 0.8 (see Table 2–15). If the BOD/COD ratio for untreated wastewater is 0.5 or greater, the waste is considered to be easily treatable by biological means. If the ratio is below about 0.3, either the waste may have some toxic components or acclimated microorganisms may be required in its stabilization. The corresponding BOD/TOC ratio for untreated wastewater varies from 1.2 to 2.0. In using these ratios it is important to remember

Table 2–15

Comparison of ratios of various parameters used to characterize wastewater

Type of wastewater	BOD/COD	BOD/TOC
Untreated	0.3–0.8	1.2–2.0
After primary settling	0.4–0.6	0.8–1.2
Final effluent	0.1–0.3[a]	0.2–0.5[b]

[a] CBOD/COD.

[b] CBOD/TOC.

that they will change significantly with the degree of treatment the waste has undergone, as reported in Table 2–15. The theoretical basis for these ratios is explored in Example 2–11.

EXAMPLE 2–11 **Determination of BOD/COD, BOD/TOC, and TOC/COD ratios** Determine the theoretical BOD/COD, BOD/TOC, and TOC/BOD ratios for the compound $C_5H_7NO_2$. Assume the value of the BOD first-order reaction rate constant is 0.23/d (base *e*) (0.10/*d* base 10).

Solution

1. Determine the COD of the compound using Eq. (2–57).

$$C_5H_7NO_2 + 5O_2 \rightarrow 5CO_2 + NH_3 + 2H_2O$$

$$\text{mw } C_5H_7NO_2 = 113, \text{ mw } 5O_2 = 160$$

$$\text{COD} = 160/113 = 1.42 \text{ mg } O_2/\text{mg } C_5H_7NO_2$$

2. Determine the BOD of the compound.

$$\frac{\text{BOD}}{\text{UBOD}} = (1 - e^{-k_1 t}) = (1 - e^{-0.23 \times 5}) = 1 - 0.32 = 0.68$$

$$\text{BOD} = 0.68 \times 1.42 \text{ mg } O_2/\text{mg } C_5H_7NO_2 = 0.97 \text{ mg BOD/mg } C_5H_7NO_2$$

3. Determine the TOC of the compound.

$$\text{TOC} = (5 \times 12)/113 = 0.53 \text{ mg TOC/mg } C_5H_7NO_2$$

4. Determine BOD/COD, BOD/TOC, and TOC/BOD ratios.

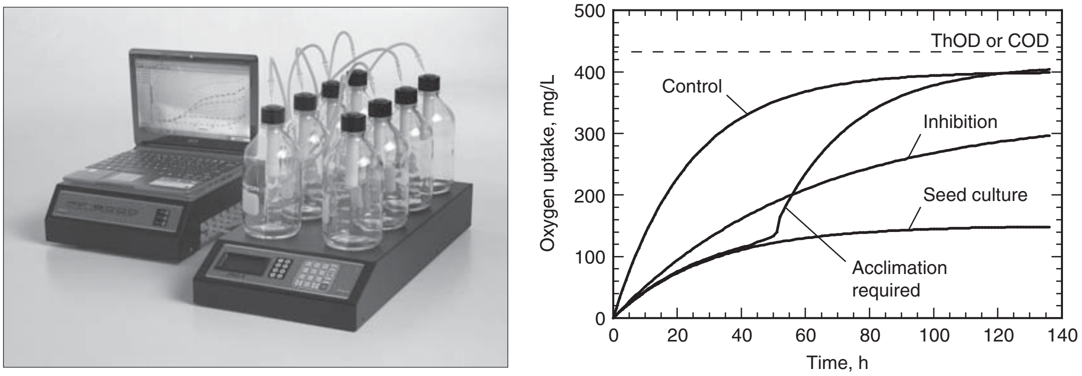
$$\frac{\text{BOD}}{\text{COD}} = \frac{0.68 \times 1.42}{1.42} = 0.68$$

$$\frac{\text{BOD}}{\text{TOC}} = \frac{0.68 \times 1.42}{0.53} = 1.82$$

$$\frac{\text{TOC}}{\text{COD}} = \frac{0.53}{1.42} = 0.37$$

Respirometric Characterization of Aggregate Organic Constituents. Determination of the BOD value and the corresponding rate constant k_1 can be accomplished more effectively using a respirometer as compared to using the bottle technique as described above (Young and Baumann, 1976a, 1976b; Young, et al., 2003). Respirometers are devices that are used to measure the rate of respiration of living microorganisms in aerobic, anoxic, and anaerobic environments.

Description. Modern headspace-gas respirometers work by maintaining a constant oxygen pressure over a sample containing microorganisms that are in the process of metabolizing an organic substrate by replacing the oxygen as it is consumed by the microorganisms. Oxygen replacement is accomplished by means of an electrolysis cell, a bubble-type flow cell, or by transducer-controlled pneumatic injection. An example of a



(a) (b)

Figure 2–27

View of respirometer and response curves: (a) commercial headspace-gas respirometer (courtesy of Respirometer Systems and Applications, LLC) and (b) typical oxygen uptake curves for wastewater samples having different biodegradation characteristics (Young and Cowan, 2004).

typical commercially available respirometer is shown on Fig. 2–27(a). The principal advantages of modern headspace-gas respirometers over manometric respirometers, such as Gilson or Warburg respirometers (Tchobanoglous and Burton, 1991), are (1) the use of a large (1-L) sample that minimizes the errors of grab sampling and pipetting in dilutions and (2) oxygen consumption is measured continuously, thereby providing much more detail about the progress of the biological reaction.

Respirometric Applications. Used initially for the determination of the BOD and rate constants, respirometry is now used in a number of different applications in the field of wastewater treatment, including (1) monitoring oxygen uptake rates in activated sludge mixed liquors, (2) assessing biodegradability and treatability of industrial wastewaters, (3) assessing toxicity of industrial chemicals to wastewater treatment processes, and (4) assessing nutrient deficiencies (Young and Cowan, 2004). Biodegradation characteristics can vary among chemical types and wastewater sources, as illustrated on Fig. 2–27(b). The curve labeled "control" represents oxygen uptake of readily biodegradable substances. The curve labeled "inhibition" is characteristic of the oxygen uptake for chemicals that may be toxic or have low rates of biodegradation. When acclimation is required, a delay in oxygen uptake will occur, but the initial rates of oxygen uptake will be similar to that of the seed culture. Other patterns can occur depending on the type of stresses imposed on the seed culture.

Oil and Grease

The term *oil and grease,* as commonly used, includes the fats, oils, waxes, and other related constituents found in wastewater. The term *fats, oil, and grease* (FOG) used previously in the literature has been replaced by the term oil and grease. The oil and grease content of a wastewater can be determined by several methods based on liquid-liquid extraction and solid phase adsorption followed by liquid extraction (Standard Methods, 2012). Following the extraction step, the solvent used in the extraction is evaporated and the residual oil and grease content is determined gravimetrically. Other extractable substances include mineral oils, such as kerosene and lubricating and road oils. Oil and grease are quite similar chemically; they are compounds (esters) of alcohol or glycerol (glycerin) with fatty acids. The glycerides of fatty acids that are liquid at ordinary temperatures are called oils, and those that are solids are called grease (or fats).

If grease is not removed before discharge of treated wastewater, it can interfere with the biological life in the surface waters and create unsightly films. The thickness of oil required to form a translucent film on the surface of a water body is about 0.0003048 mm (0.0000120 in.), as given in the following table.

	Film thickness		**Quantity spread**	
Appearance	**in.**	**mm**	**gal/mi²**	**L/ha**
Barely visible	0.0000015	0.0000381	25	0.365
Silvery sheen	0.0000030	0.0000762	50	0.731
First trace of color	0.0000060	0.0001524	100	1.461
Bright bands of color	0.0000120	0.0003048	200	2.922
Colors begin to dull	0.0000400	0.0010160	666	9.731
Colors are much darker	0.0000800	0.0020320	1332	19.463

Source: Eldridge (1942).

Fats and oils are contributed to domestic wastewater in butter, lard, margarine, and vegetable fats and oils. Fats are also commonly found in meats, in the germinal area of cereals, in seeds, in nuts, and in certain fruits. The low solubility of fats and oils reduces their rate of microbial degradation. Mineral acids attack them, however, resulting in the formation of glycerin and fatty acid. In the presence of alkalies, such as sodium hydroxide, glycerin is liberated, and alkali salts of the fatty acids are formed. These alkali salts are known as soaps. Common soaps are made by saponification of fats with sodium hydroxide. They are soluble in water, but in the presence of hardness constituents, the sodium salts are changed to calcium and magnesium salts of the fatty acids, or so-called mineral soaps. These are insoluble and are precipitated.

Kerosene, lubricating, and road oils are derived from petroleum and coal tar and contain essentially carbon and hydrogen. These oils sometimes reach the collection system in considerable volume from shops, garages, and streets. For the most part, they float on the wastewater, although a portion is carried into the sludge on settling solids. To an even greater extent than fats, oils, and soaps, the mineral oils tend to coat surfaces. The particles interfere with biological action and cause maintenance problems.

Surfactants

Surfactants, or surface-active agents, are large organic molecules that are slightly soluble in water and cause foaming in wastewater treatment plants and in the surface waters into which the waste effluent is discharged. Surfactants are composed most commonly of a strongly hydrophobic group combined with a strongly hydrophilic group. Typically, the hydrophobic group is a hydrocarbon radical (R) made up of 10 to 20 carbon atoms. Two types of hydrophobic groups are used: those that will and those that will not ionize in water. Anionic surfactants are negatively charged [e.g., $(RSO_3N)^-Na^+$], whereas cationic surfactants are positively charged [e.g., $(RMe_3N)^+Cl^-$]. Nonionizing (nonionic) surfactants commonly contain a polyoxyethylene hydrophilic group ($ROCH_2CH_2OCH_2CH_2 \ldots OCH_2CH_2OH$, often abbreviated RE_n, where n is the average number of $—OCH_2CH_2—$ units in the hydrophilic group). Hybrids of these types also exist. In the United States, ionic surfactants amount to about two-thirds of the total surfactants used and nonionics to about one-third (Standard Methods, 2012).

Surfactants tend to collect at the air-water interface with the hydrophilic group in the water and the hydrophobic group in the air. During aeration of wastewater, these compounds collect on the surface of the air bubbles and thus create a very stable foam. Before 1965, the type of surfactant present in synthetic detergents, called alkyl-benzene-sulfonate (ABS), was especially troublesome because it resisted breakdown by biological means. As a result of legislation in 1965, ABS has been replaced in detergents by linear-alkyl-sulfonate (LAS), which is biodegradable. Because surfactants come primarily from synthetic detergents, the foaming problem has been greatly reduced. It should be noted that so called “hard” synthetic detergents are still used extensively in many foreign countries.

Two tests are now used to determine the presence of surfactants in water and wastewater. The MBAS (methylene blue active substances) test is used for anionic surfactants. The determination of surfactants is accomplished by measuring the color change in a standard solution of methylene blue dye. Nonionic surfactants are measured using the CTAS (cobalt thiocyanate active substances) test. Nonionic surfactants will react with the CTAS to produce a cobalt containing product which can be extracted into an organic liquid and then measured. It should be noted that the CTAS method requires sublimation to remove nonsurfactants and ion exchange to remove the cationic and anionic surfactants (Standard Methods, 2012).

Chemical Energy in Wastewater and Biosolids

The chemical energy content of the organic constituents in untreated wastewater, primary sludge, and biosolids can be determined by (1) using a full-scale boiler as a calorimeter, (2) using a laboratory bomb calorimeter, and (3) by calculation, if the elemental composition is known. Because of the difficulty in instrumenting a full-scale boiler, most of the experimental data on the energy content of the organic constituents of wastewater, sludge, and biosolids are based on the results of bomb calorimeter tests (Shizas and Bagley, 2004; Zanoni and Mueller, 1982).

The energy content of wastewater can be estimated from an elemental analysis of the constituents in organic compounds using the following expression, which is a modified form of the DuLong formula developed by Channiwala (1992), also Channiwala and Parikh (2002).

$$\text{HHV (MJ/kg)} = 34.91\,C + 117.83\,H - 10.34\,O - 1.51\,N + 10.05\,S - 2.11A \tag{2–66}$$

Where HHV is the high heating value and C is the weight fraction of carbon; H of hydrogen; O of oxygen; N of nitrogen; S of sulfur; and A of ash as derived from an ultimate analysis or from the chemical formula, if known. When the HHV is used, it is assumed that the water component is in liquid state at the end of combustion. Another estimate of the heating value of a combustible material is the lower LHV (lower heating value), in which it is assumed that latent heat of vaporization is not recovered. In general, the LHV is about 6 to 8 percent lower than the corresponding HHV. For stationary combustion units with exhaust heat recovered, use of the HHV is the most appropriate. Where exhaust heat is not recovered, use of the LHV is most appropriate. Also, in most of the European literature, LHVs are reported, whereas HHVs are reported in the American literature. The application of Eq. 2–66 is illustrated in Example 2–12.

EXAMPLE 2–12 **Estimate the Chemical Energy Content of Untreated Wastewater and Biosolids** Estimate the chemical energy content, on a COD basis, of (1) the organic fraction of untreated wastewater, composed of 50, 40, and 10 percent proteins, carbohydrates, and fat, respectively and (2) biosolids comprised of bacterial cell biomass. Assume the chemical composition of untreated wastewater is $C_{7.9}H_{13}O_{3.7}NS_{0.04}$ with an ash content of 3 percent. The composition of the cell biomass is $C_5H_7O_2N$ (Hoover and Porges, 1952) with an ash content of 3 percent. Express results on the basis of MJ/kg organic fraction or biosolids COD.

Solution-Part 1 Untreated Wastewater

1. Determine the energy content of the wastewater using Eq. 2–66.
   a. Determine the weight fractions of the elements and ash comprising the wastewater.

Component	Coefficient	Molecular weight	Molecular mass	Weight fraction
Carbon	7.9	12	94.8	0.50[a]
Hydrogen	13	1	13	0.07
Oxygen	3.7	16	59.2	0.31
Nitrogen	1	14	14	0.08
Sulfur	0.04	32	1.28	0.01
Ash	0			0.03
			182.28	1.00

[a] (94.8/182.28) × 0.97 = 0.50.

   b. The energy content of the organic fraction using Eq. 2–66 is:

$$\text{HHV (MJ/kg organic fraction)} = 34.91\,(0.50) + 117.83\,(0.07) - 10.34\,(0.31) - 1.51\,(0.08) + 10.05(0.01) - 2.11\,(0.03)$$

$$\text{HHV (MJ/kg organic fraction)} = 17.45 + 8.25 - 3.21 - 0.12 + 0.10 - 0.06 = 22.41$$

2. Determine the COD of the organic fraction.
   a. Write a balanced reaction for the chemical oxidation of the biomass neglecting sulfur.

$$\underset{182.28}{C_{7.9}H_{13}NO_{3.7}} + \underset{8.55(32)}{8.55O_2} \rightarrow 7.9CO_2 + NH_3 + 5H_2O$$

   b. The COD of the organic fraction is

$$\text{COD} = 8.55(32\text{ g }O_2\text{/mole})/(182.28\text{ g organic fraction/mole})$$

$$= 1.50\text{ g }O_2\text{/g organic fraction}$$

3. Determine the energy content of the biomass in terms of MJ/kg biosolids COD

$$\text{HHV (MJ/kg organic fraction COD)} = \frac{(22.77\text{ MJ/kg of organic fraction})}{(1.50\text{ kg }O_2\text{/kg of organic fraction})}$$

$$= 15.1\text{ MJ/kg of organic fraction COD}$$

Solution-Part 2 Biosolids

1. Determine the energy content of the biosolids using Eq. 2–66.
   a. Determine the weight fractions of the elements and ash comprising the biosolids.

Component	Coefficient	Molecular weight	Molecular mass	Weight fraction
Carbon	5	12	60	0.52[a]
Hydrogen	7	1	7	0.06
Oxygen	2	16	32	0.27
Nitrogen	1	14	14	0.12
Sulfur	0	32	0	0
Ash	0			0.03
			113	1.00

[a] (60/113) × 0.97 = 0.52.

b. The energy content of the biosolids using Eq. 2–66 is

$$\text{HHV (MJ/kg biosolids)} = 34.91\,(0.52) + 117.83\,(0.06) - 10.34\,(0.27) - 1.51(0.12) - 2.11\,(0.03)$$

$$\text{HHV (MJ/kg biosolids)} = 18.15 + 7.07 - 2.79 - 0.18 - 0.06 = 22.19$$

2. Determine the COD of the biosolids.

a. Write a balanced reaction for the chemical oxidation of the biomass.

$$\underset{113}{C_5H_7NO_2} + \underset{5(32)}{5O_2} \rightarrow 5CO_2 + NH_3 + 2H_2O$$

b. The COD of the biosolids is

$$\begin{aligned}\text{COD} &= 5(32\text{ g }O_2\text{/mole})/(113\text{ g/mole biosolids})\\ &= 1.42\text{ g }O_2\text{/g biosolids}\end{aligned}$$

3. Determine the energy content of the biomass in terms of MJ/kg biosolids COD.

$$\begin{aligned}\text{HHV (MJ/kg biosolids COD)} &= \frac{(22.19\text{ MJ/kg of biosolids})}{(1.42\text{ kg }O_2\text{/kg of biosolids})}\\ &= 15.63\text{ MJ/kg of biosolids COD}\end{aligned}$$

Comment In practice, because the HHV is difficult to recover, the LHV is often used in place of the HHV. If the difference between the HHV and LHV is assumed to be eight percent, the corresponding LHV of the organic fraction and biosolids is 13.74 MJ/kg organic fraction and 14.38 MJ/ kg biosolids COD.

2–7 INDIVIDUAL ORGANIC COMPOUNDS

Individual organic compounds are determined to assess the presence of priority pollutants identified by the U.S. Environmental Protection Agency (U.S. EPA) and a number of new emerging compounds of concern (for which regulations have not been written). Priority pollutants (both inorganic and organic) have been and are continuing to be selected on the basis of their known or suspected carcinogenicity, mutagenicity, teratogenicity, or high acute toxicity. As the techniques used to identify specific compounds continue to improve, several other organic compounds have been detected in public water supplies and in treated wastewater effluents.

Sources of Individual Organic Compounds

The individual compounds of concern can be grouped into several categories: (1) priority pollutants, (2) volatile organic compounds (VOCs), (3) disinfection byproducts, (4) pesticides and agricultural chemicals, and (5) unregulated trace organic compounds. The sources of priority pollutants are primarily from commercial and industrial discharges and, to a very limited extent, from products used in domestic applications. Volatile organic compounds are derived primarily from commercial and industrial sources. Pesticides and agricultural chemicals found in wastewater are primarily from surface runoff from agricultural, vacant, and park lands. Unregulated trace organic compounds in question are derived, in large part, from (1) human and veterinary antibiotics, (2) human prescription and nonprescription drugs, (3) industrial and household wastewater products, and (4) sex and steroidal hormones.

Priority Pollutants

The U.S. EPA initially identified 129 priority pollutants in 65 classes to be regulated by categorical discharge standards; three were removed in 1981 (Federal Register, 1982). Priority pollutants (both inorganic and organic) were selected on the basis of their known or suspected carcinogenicity, mutagenicity, teratogenicity, or high acute toxicity. Many of the organic priority pollutants are also classified as volatile organic compounds (VOCs). Two types of standards are used to control pollutant discharges to public-owned treatment works (POTWs). The first, "prohibited discharge standards," applies to all commercial and industrial establishments which discharge to POTWs. Prohibited standards restrict the discharge of pollutants that may create a fire or explosion hazard in collection systems or treatment works, are corrosive ($pH < 5.0$), obstruct flow, upset treatment processes, or increase the temperature of the wastewater entering the plant to above 40°C. "Categorical Standards" apply to industrial and commercial discharges in 25 industrial categories ("categorical industries") and are intended to restrict the discharge of priority pollutants. It is anticipated that this list will continue to be updated in future.

Volatile Organic Compounds (VOCs)

Organic compounds that have a boiling point less than or equal to 100°C or a vapor pressure greater than 1 mm Hg at 25°C are generally considered to be volatile organic compounds (VOCs). For example, vinyl chloride, which has a boiling point of -13.9°C and a vapor pressure of 2548 mm Hg at 20°C, is an example of an extremely volatile organic compound. Volatile organic compounds are of great concern because (1) once such compounds are in the vapor state they are much more mobile and, therefore, more likely to be released to the environment; (2) the presence of some of these compounds in the atmosphere may pose a significant public health risk; and (3) they contribute to a general increase in reactive hydrocarbons in the atmosphere, which can lead to the formation of photochemical oxidants. The release of these compounds in collection systems and at treatment plants, especially at the headworks, is of particular concern with respect to the health of the collection system and treatment plant workers. The physical phenomena involved in the release and control of VOCs is considered in more detail in Chap. 16.

Disinfection Byproducts

It has been found that when chlorine is added to water containing organic matter several organic compounds containing chlorine are formed. Collectively, these compounds, along with others, are known as disinfection byproducts (DBPs). Although generally present in low concentrations, they are of concern because many of them are known or suspected potential human carcinogens. Typical classes of compounds include trihalomethanes (THMs), haloacetic acids (HAAs), trichlorophenol, and aldehydes.

In the last decade N-Nitrosodimethylamine (NDMA) has been found in the effluent from wastewater treatment plants. The reason for concern over this compound is because as a group of compounds, nitrosamines are among the most powerful carcinogens known (Snyder, 1995). These compounds are also known to be carcinogenic to various fish species at low concentrations. The U.S. EPA action limit for NDMA is two parts per trillion. Based on the results of recent studies, NDMA appears to be formed during the chlorination process. In treated effluent, the nitrite ion can react with hydrochloric acid, present as result of the use of chlorine for disinfection, to form nitrous acid. In turn, nitrous acid can react with dimethylamine to form NDMA (Hill, 1988). The compound dimethylamine is common in wastewater and surface waters, being found in urine, feces, algae, and plant tissues. Dimethylamine is also part of some polymers used for water treatment (such as polydiallyl dimethylamine) and for ion exchange resins. The formation of NDMA under basic and alkaline conditions has been reported by Wainwright (1986).

Because of the concern over the formation of DBPs and NDMA, considerable attention has been focused on the use of ultraviolet (UV) disinfection as a possible replacement for chlorine. In addition, considerable attention has been focused on the modifications to conventional treatment processes to improve the treatment of these compounds and to advanced treatment processes for the removal of these substances. The use of UV radiation for disinfection and the destruction of NDMA is considered in Chap. 12.

Pesticides and Agricultural Chemicals

Pesticides, herbicides, and other agricultural chemicals, are toxic to many organisms and, therefore, can be significant contaminants of surface waters. Concentrations of these chemicals can result in fish kills, in contamination of the flesh of fish that decreases their value as a source of food, and in impairment of water supplies. These chemicals are not common constituents of domestic wastewater.

Unregulated Trace Organic Compounds

In addition to the compounds discussed above for which requirements have been established, varieties of new unregulated (often referred to as emerging, trace or micropollutant compounds) compounds have been identified in many of the nation's water supplies and in treated wastewater effluents at low ng/L or low ug/L concentrations. The compounds in question are derived, in large part, from (1) human and veterinary antibiotics, (2) human prescription and nonprescription drugs, (3) industrial and household wastewater products, and (4) sex and steroidal hormones. Typical examples of types of compounds involved are reported in Table 2–16. As more becomes known about the health impacts of these compounds, it is anticipated that discharge limits may be developed for a number of these compounds. Given that over 30 million organic compounds are known to exist, it is clear that the list of emerging compounds will continue to grow as analytical techniques continue to improve.

Analysis of Individual Organic Compounds

The analytical methods used to determine individual organic compounds require the use of sophisticated instrumentation capable of measuring trace concentrations in the range of 10^{-12} to 10^{-3} mg/L. Gas chromatographic (GC) and high-performance liquid chromatographic (HPLC) methods are used most commonly to detect individual organic compounds. Different types of detectors are used with each method, depending on the nature of the compound being analyzed. Typical detectors used in conjunction with gas chromatography include electrolytic conductivity, electron capture (ECD), flame ionization (FID),

Table 2–16

Representative organic constituents found in wastewater and in streams[a]

Veterinary and human antibiotics

Carbadox
Chlortetracycline
Ciprofloxacin
Dolcycline
Enrofloxacin
Erythromycin
Erythromycin-H_2O
Ivermectin
Lincomycin
Norfioxacin
Oxytetracycline
Roxarsone
Roxithromycin
Sarafloxacin
Spectinomycin
Sulfachlorpyridazine
Sulfadimethoxine
Sulfamethazine
Sulfamethazine
Sulfamethiazole
Sulfathiazole
Sulfamethoxazole
Tetracycline
Trimethoprim
Tylosin
Virginiamycin

Sex and steroidal hormones

Cis-androsternone
3-β-coprostanol
Cholesterol
Equilenin
Equilin
Estrone
Estriol
17α-estradiol
17β-estradiol
17α-ethynylestradiol
Mestranol
19-norethisterone
Progesterone
Testosterone

Human prescription and non-prescription drugs (general use)

Acetominophen (antipyretic)
Albuterol (antiasthmatic)
Amoxicillin (antibiotic)
Caffeine (stimulant)
Carbamazepine (anticonvulsant)
Cimetidine (antacid)
Codeine (analgesic)
Cotinine (nicotine metabolite)
Dehydronifedipine (antianginal)
Digoxigenin (digoxin metabolite)
Diltiazem (antihypertensive)
Diphenhydramine (antihistamine)
Enalaprillat (antihypertensive)
Fluoxetine (antidepressant)
Furosemide (diuretic)
Gemfibrozil (lipotropic agent)
Ibuprofen (anti-inflammatory)
Mefformin (antidiabetic agent)
Paroxetine (paxil metabolite)
Paraxanthine (caffeine metabolite)
Ranitidine (antacid)
Salbutamol (antiasthmatic)
Sulfamethoxazole (antibiotic)
Trimethoprim (antibiotic)
Warfarin (anticoagulant)

Industrial and Household Wastewater Products (general use)

Acetophenone (fragrance)
Anthracene (PAH)[b]
Benzo(a)pyrene (PAH)
Benzophenone (used in plastics)
2,6-di-tert-para-benzoquinone (antioxidant)
5methyl 1 H benzotriazole (antioxidant)
Lindane (pesticide)
Methyl parathion (pesticide)
Napthalene (PAH)
NPEO1 – total (detergent metabolite)
NPEO2 – total (detergent metabolite)
OPEO1 (detergent metabolite)

(*continued*)

Table 2–16 (*Continued*)

Industrial and household wastewater products (continued)	
Bisphenol A (used in polymers)	OPEO2 (detergent metabolite)
Bis(2-ethylhexyl)phthalate (plasticizer)	Pentachlorophenol (wood preservative)
2,6-di-te-butylphenol (antioxidant)	Phenanthrene (PAH)
Butylated hydroxyanisole (antioxidant)	Phenol (disinfectant)
Butylated hydroxytoluene (antioxidant)	Para–nonylphenol-total (detergent metabolite)
Caffeine (stimulant)	Phthalic anhydride (used in plastics)
Cholesterol (plant/animal steroid)	Pyrene (PAH)
Codeine (analgesic)	Stigmastanol (plant sterol)
Cotinine (nicotine metabolite)	Tetrachloroethylene (solvent)
3b-coprostanol (carnivore fecal indicator)	Tributyl phosphate (fire retardant)
Para-cresol (wood preservative)	Triclosan (antimicrobial disinfectant)
Diethylphthalate (plasticizer)	Tri(2-butoxyethyl) phosphate (plasticizer)
1,4-dichorobenzene (fumigant)	Tri(2-chloroethyl) phosphate (fire retardant)
Ethanol, 2-butoxy-, phosphate (plasticizer)	Tri(dichlorisipropyl) phosphate (fire retardant)
Fluoranthene (PAH)	Triphenyl phosphate (plasticizer)
Fungicide, herbicides, insecticides, and pesticides (general use)	
Bromacil (herbicide)	Diazinon (insecticide)
Carbazole (insecticide)	Dieldrin (pesticide)
Carbraryl (insecticide)	Metolachlor (herbicide)
Chlorpyrifos (insecticide)	N,N-diethyltoluamide (DEET) (insecticide)
Chlorpyrifos (pesticide)	Prometon (herbicide)
Cis-chlordane (pesticide)	Thiabendazole (fungicide)
Other compounds (general use)	
Anthraquinone (aromatic organic compound used in manufacturing)	1,3,4,6,7,8-hexahydro-4,6,6,7,8,8,-hexamethylcyclopenta-γ-2-benzopyran (fragrances)
β-sitostrol (plant steroid)	
β-stigmastanol (plant steroid)	

[a] Adapted in part from USGS (2000).

[b] PAH = polynuclear aromatic hydrocarbon.

photoionization (PID) and mass spectrometer (GCMS). Typical detectors for high-performance liquid chromatography include photodiode array (PDAD) and post column reactor (PCR). It should also be noted that many of the individual organic constituents can be determined by two or more of the above methods (Standard Methods, 2012).

Over 180 individual organic compounds the can be determined using one or more of the methods cited above. The principal categories containing the individual organic compounds are reported in Table 2–17. As instrumental methods of analysis have improved, the detection limits for these compounds have become increasingly small, typically below 10 ng/L. The specific organic compounds that are analyzed for will depend on the application. For example, for indirect reuse applications, scans of disinfection by-products may be required where chlorine is used for disinfection.

Table 2–17

Typical classes of organic compounds whose members are identified as individual compounds

Name	Occurrence/source	Concern
Volatile organic compounds	Found in ground and surface waters	Potential for tetratogenesis or carcinogenesis in humans
1,2-Dibromoethane (EDB) and 1,2-Dibromo-3-Chloropropane (DBCP)	Found in groundwater supplies, especially where these compounds have been used as fumigants	Detrimental effects on human health
Trihalomethanes (THMs)	Found in most chlorinated water supplies	Disinfection byproduct. Potential human carcinogen
Chlorinated organic solvents	Found in raw water supplies resulting from industrial contamination	Potential human carcinogen
Halocetic acids (HAAs)	Form from the chlorination of natural organic matter (humic and fulvic acids)	Disinfection byproduct. Potential human carcinogen
Trichlorophenol	Form from the chlorination of natural organic matter (humic and fulvic acids)	Disinfection byproduct. Dichloacetic acid and trichloroacetic acid are animal carcinogens
Aldehydes	Formed from the application of ozone to water containing organic matter	Disinfection byproduct
Extractable base/neutral and acids	Many semivolitile compounds including polynuclear aromatic hydrocarbons, phthalates, phenolics, organochlorine pesticides, and PCBs	Many of the listed compounds are toxic or carcinogenic
Phenols	Generally traceable to industrial discharges or landfills	Impart a taste to water at low levels. May have detrimental impact on human health at higher levels
Polychlorinated biphenyls (PCB)	Found in water supplies contaminated by transformer oils	These compounds are toxic, bioaccumulative, and extremely stable in water
Polynuclear aromatic hydrocarbons (PAHs)	Byproducts of petroleum processing or combustion	Many compounds in this group are highly carcinogenic at relatively low levels
Carbamate pesticides	Found in water supplies contaminated by pesticides	
Organochlorine pesticides	Found in water supplies contaminated by pesticides	Many compounds in this group are bioaccumulative, and relatively stable, as well as toxic or carcinogenic
Acidic herbicide compounds	Used for weed control these compounds are found in aquatic systems	
Glyphosphate herbicide	Broad spectrum nonselective post emergence herbicide. Water supplies can become contaminated through runoff and spray drift	

2–8 RADIONUCLIDES IN WASTEWATER

Radionuclides are unstable atoms that are transformed through the process of radioactive decay. Radioactive decay is the spontaneous disintegration of an element, resulting in greater atomic stability through change of electron orbits or release of radioactive particles or radiation (Crittenden et al., 2012). Radionuclides are of interest because of the health effects resulting from exposure to radioactive particles and their occurrence in natural waters. Radionuclides are being detected in wastewater and are of concern when wastewater containing radionuclides is reused or discharged to the environment. Wastewater treatment plant processes also have the potential to concentrate radioactive materials in biosolids.

Sources of Radionuclide

Radioactive materials may be present in wastewater from natural occurring sources, medical facilities, and regulated users of Atomic Energy Act radioactive materials. Natural radionuclides are the most common sources of radioactivity in the environment and are formed from the dissolution of rock formations containing uranium ore and gases released from deep in the earth's crust. The principal source of natural radionuclides is uranium ore (U_3O_8), and its abundance in the earth's crust is 1 part per 10^{12} parts. Natural occurring radionuclides can enter the collection system from (1) community water supplies, principally those with predominant amounts of groundwater; (2) water infiltrating into the sewer system; and (3) residuals from water purification systems. Runoff into storm sewers can also occur from global radioactive fallout such as that resulting from the nuclear power plant disaster in Japan in 2011.

The anthropogenic radionuclides come from sources such as a nuclear power plants used to supply electrical energy, medical facilities that provide nuclear medicines and x-ray services, academic and research facilities using nuclear materials for research, commercial products such as televisions and smoke detectors, and nuclear weapons (Crittenden et al., 2012). The National Research Council (NRC) estimates that more than 9000 out of 22,000 regulated users have the potential to release radioactive materials to domestic collection systems (Bastian, 2011).

Units of Expression

The life span of a radionuclide in water and wastewater is measured by its half-life. A half-life is defined in terms of probability. It is the time when the expected value of a number of entities that have decayed is equal to half of the original number. The units of expression for radionuclide concentrations in water are picocuries per liter (pCi/L).

Description of Isotopes Found in Wastewater and Sludge

The primary forms of radioactive decay are (1) alpha (particle) radiation, (2) beta (particle) radiation, and (3) gamma (ray) radiation. The release of alpha and beta particles transforms an isotope into a different element, while the release of gamma radiation reduces the energy of the element. Alpha, beta, and gamma radiation are known as ionizing radiation because of their ability to free electrons from their orbit in adjacent atoms (Crittenden et al., 2012). In an environmental radiation program conducted by the State of Washington, for example, several isotopes as well as total uranium were detected in wastewater and wastewater sludge samples from six wastewater treatment plants. The decay mode in most cases was found to be either beta or gamma. The isotopes identified in the sampling study, their half-life, and possible sources are given in Table 2–18 (WDOH 1997).

Recently, radioactive tritium, which occurs naturally in lakes, rivers, and public water supplies, has been detected near nuclear testing facilities and nuclear power plants. Tritium at very low concentrations has been detected in tertiary effluents. In a report by the U.S. EPA, the levels of radioactive materials found in biosolids and ash samples from most wastewater treatment plants indicated that radiation exposure to workers and the general public is very low and is not likely of concern (Bastian, 2011).

Treatment Technologies for the Removal of Radionuclides

No technologies have been reported in the literature for the removal of radionuclides in wastewater. However, technologies that have been investigated for the removal of radionuclides from drinking water that can be considered for wastewater are presented in Table 2–19. Laboratory analyses and bench- and pilot-plant testing are required to determine feasible methods of treatment.

Table 2–18

Isotopes Found in Wastewater and Sludge in the State of Washington

Isotope	Half-life	Decay mode	Possible sources
Beryllium-7 (Be-7)	53 d	Gamma	Be-7 is a naturally occurring radioisotope produced by the interactions of cosmic radiation and the upper atmosphere. Be-7 could enter the wastewater system through the water supply, from runoff in a combined system, or in washwater which has picked up dust
Cesium-137 (Cs-137)	30 y	Beta, Gamma	Cs-137 is a fission product produced in nuclear reactors and explosions. It is encountered virtually everywhere in the world as the result of atmospheric testing of nuclear weapons. Cs-137 can be found in wastewater systems from the water supply (if not supplied by wells), runoff in combined systems, and people via the food chain
Cobalt-57 (Co-57)	270 d	Gamma	Co-57 is produced by accelerators. The source of Co-57 detected in sludge is not apparent although Co-57 is used in some medical procedures
Cobalt-58 (Co-58)	71 d	Gamma	See Co-57
Cobalt-60 (Co-60)	5.2 y	Beta, Gamma	Co-60 is produced usually in nuclear reactors when neutrons are by Co-59 atoms in the steel. It is not readily apparent where Co-60 in collection systems comes from
Gross Beta	NA	Beta	Gross Beta is a screening tool to determine if further analyses of Beta emitting isotopes are needed. Subsequent testing of isotopes is expensive and time consuming
Iodine-131 (I-131)	8 d	Beta, Gamma	I-131 can be created as a fission product of a nuclear reactor or weapon, or by irradiating material in a reactor accelerator. It is also used in a number of medical procedures. I-131 found in sludge is likely from medical procedures
Manganese-54 (Mn-54)	312 d		Mn-54 is an activation product produced in nuclear reactors
Potassium-40 (K-40)	1.2×10^9 y	Beta, Gamma	K-40 is an naturally occurring radioisotope; approximately 0.01% of all potassium is K-40. The K-40 in sludge is attributed principally to excrement although it could be from runoff of fertilizer used in landscape and agricultural applications
Strontium-89 (Sr-89)	51 d	Beta	Sr-89 is a fission fragment produced in nuclear reactors or weapon explosions. It is sometimes used in medical procedures
Strontium-90 (Sr-90)	230 y	Beta	Sr-90 is a fission fragment produced in nuclear reactors or weapon explosions. Sr-90 is found in the environment from weapons test fallout and can travel up the food chain
Technetium-99 (Tc-99)	213,000 y	Beta	Tc-99 is created as a fission fragment when a uranium atom is split. It also has several medical uses
Thallium-201 (Tl-201)	3 d	Gamma	Th-201 is produced in accelerators and used in examining heart functions
Total Uranium (Total U)			Total U can occur is a natural form or from processing in a nuclear reactor. Uranium in the wastewater system can arrive from the water supply (especially wells) or from runoff in combined systems and washwater from uranium processing facilities
Zinc-65 (Zn-65)	243.9 d		Zn-65 is a neutron activation product, usually associated with nuclear reactors

[a] Adapted from WDOH (1997).

Table 2–19

Possible treatment technologies considered for the removal of radionuclides[a]

Treatment technology	Application
Activated alumina	Activated alumina has been used to remove fluoride and arsenic in full-scale drinking water applications and may be suitable for the removal of uranium. Removal rates for uranium are reported to be about 90 percent.
Aeration	Radon-222 gas has a very high Henry's law constant and is very amenable to aeration. Aeration methods such as diffused air and packed towers are reported to be very effective in removing radon-222. Removal rates for radon are reported to be up to 99 percent.
Coagulation-filtration	Conventional coagulation-filtration using alum or iron salts can be effective in removing uranium from drinking water. Removal rates are reported to be in the range of 80 to 98 percent. Enhanced coagulation-filtration with alum or ferric can be used in large applications for the removal of uranium.
Ion exchange	Strong-acid cation (SAC) resins in the sodium form and weak acid cation (WAC) resins can be used to remove radium-226 and radium-228 from aqueous systems. The advantages of WAC resins are that they are easier to regenerate and require less regenerate then SAC resins; the disadvantages are the resin tends to swell and corrosion resistant materials are required if HCl is the regenerant solution. Reported removal rates are from 81 to 100 percent for radium and 90 to 100 percent for uranium.
Reverse osmosis	Reverse osmosis has been found to be an excellent treatment technology for the removal of radionuclides from drinking water. Several types of membranes have been effective in removing natural uranium from groundwater in Florida. Reported removal rates are 90 to 95+ percent for radium and 90 to 99 percent for uranium.

[a] Adapted from Crittenden et al. (2012) and Malcolm Pirnie Inc. (2008).

2–9 BIOLOGICAL CONSTITUENTS

The biological characteristics of wastewater are of fundamental importance in the control of diseases caused by pathogenic organisms of human origin and, because of the extensive and fundamental role played by bacteria and other microorganisms in the decomposition and stabilization of organic matter and the transformation of inorganic compounds, both in nature and in wastewater treatment plants. To develop a general understanding of the microbial constituents in wastewater, the material presented in this section has been organized into the following subsections: (1) the sources of the microorganisms found in wastewater, (2) the general classification of the microorganisms found in wastewater, (3) the enumeration and detection microorganisms, (4) the pathogenic microorganisms associated with human disease, (5) the use of indicator organisms, and (6) a brief discussion of evolving pathogenic microorganisms. Following the general introduction presented in subsections (1) and (2), the remainder of the section is devoted to consideration of the pathogenic microorganisms found in wastewater. The growth, metabolic, and environmental requirements of microorganisms responsible for the treatment of wastewater are considered in Chap. 7. The use of microorganisms for the treatment of wastewater is considered in Chaps. 8, 9, and 10. The biological conversion of waste sludge to biosolids is considered in Chap. 13.

Sources of Microorganisms in Wastewater

Microorganisms found in untreated wastewater include bacteria, fungi, algae, protozoa, helminthes, viruses, and other microscopic plants and animals. The principal source of these organisms is from human waste. Other sources include commercial and industrial activities and surface water infiltration and inflow. While a large number of microorganisms are always present in wastewater, others may only be present during disease outbreaks resulting from a specific disease or specific events or occurrences. The concentration of microorganisms resulting from a disease outbreak will depend on the number of persons shedding microorganisms (in feces) and the duration of the shedding.

General Classification. In 1990, a three domain system for classifying microoganisms was proposed (Woese et al., 1990). The three domains proposed are *Bacteria, Archaea,* and *Eucarya (*or *Eukarya*). The relationship between these three domains along with the principal phyla within each domain are illustrated on Fig. 7–8 in Chap. 7. The Bacteria and Archaea are single-cellular organisms and classified as prokaryotes, which do not contain their chromosome inside a nucleus. The Eucarya can be single or multicellular organisms and have their chromosome inside a nucleus. The Archaea are separated from Bacteria due to their deoxyribonucleic acid (DNA) composition and unique cellular chemistry, such as differences in the cell wall and ribosome structure. Many Archaea can grow under extreme conditions of temperature and salinity and also include methanogenic-producing microorganisms within the phylum Euryarchaeota, important in anaerobic treatment processes, and ammonia-oxidizing populations within the phylum Crenarchaetota. There are no known pathogenic microorganisms within the Archaea. Some of the key differences between bacteria and Archaea (prokaryotes) and eukaryote microorganisms are summarized in Table 2–20, and a schematic of their cell structures is shown on Fig. 2–28. Clearly, the eukaryotes are much more complex and encompass both multicellular and single-cell plants and animals including algae, protozoa, and fungi. Also, the eukaryote organisms are generally much larger than the prokaryotes. Additional information on prokaryote cell structure and composition and the role and importance of DNA and ribonucleic nucleic acid (RNA) are discussed in Chap. 7.

Viruses are obligate *intracellular infectious* agents that are able to direct the metabolic machinery of a host cell to duplicate their molecular structure. Although viruses contain the genetic information (either DNA or RNA) needed to replicate themselves, they are unable to reproduce outside of a host cell. Viruses are composed of a nucleic acid core (DNA or RNA) surrounded by an outer coat of protein and glycoprotein. In the past, viruses were often classified according to the host infected. Bacteriophage, as the name implies, are viruses that infect bacteria and are common microbial constituents found in wastewater.

General Description. A general description of the microorganisms found in wastewater is given in Table 2–21 using the terminology introduced in the previous paragraphs. Data on the shape, resistant form, and size of the microorganisms of interest are presented in Table 2–22. Information on the size of the microorganisms, especially the resistant form, is needed to determine the type of treatment that will be required to treat or remove them.

An important feature of some microorganisms is their ability to generate resistant cell forms. For example, selected species of bacteria can form endospores (formed within the cell), the structure of which is extremely complex. The endospore contains all of the information necessary for reproduction and is coated with several layers of proteins. Endospores are extremely resistant to heat, dessication, and disinfecting chemicals. It has been speculated that endospores may remain dormant for decades, and perhaps even centuries. A spore can become viable in a suitable environment in a three step process: activation,

Table 2–20

Comparison of prokaryote and eucaryote cells[a]

Cell characteristic	Prokaryote	Eukaryote (Eukarya)
Phylogenetic group	Bacteria, blue-green algae (cyanobacteria), archaea	Single cell: algae, fungi, protozoan Multicell: plants, animals
Size[b]	Small, 0.2 to 3.0 μm	2–100 μm for single cell organisms
Cell wall	Composed of petidoglycan (bacteria), other polysaccharides, protein, glycoprotein (archaea)	Absent in animals and most protozoan, present in plants, algae, fungi: usually polysaccharide
Nuclear structure		
Nuclear membrane	Absent	Present
DNA	Single molecular, plasmids	Several chromosomes
Internal membranes	Simple, limited	Complex, endoplasmic reticulum, golgi, mitochondria several present
Membrane organelles	Absent	Several present
Photosynthetic pigments	In internal membranes chloroplasts absent	In chloroplasts
Respiratory system	Part of cytoplasmic membrane	Mitochondria

[a] Adapted from Ingraham and Ingraham (1995), Madigan et al. (2009), and Stanier et al. (1986).

[b] For additional size information see Table 2–20.

Figure 2–28

Typical structure of microorganism cells: (a) prokaryotic and (b) eukaryotic.

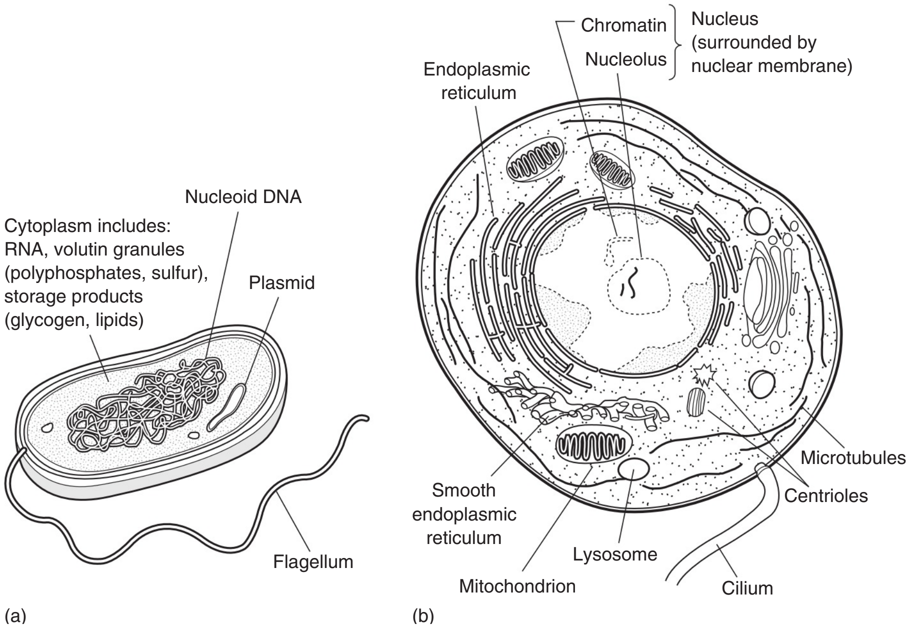


Table 2–21

Typical descriptions of the microorganisms found in natural waters, wastewater, and wastewater treatment processes

Organism	Description
Bacteria	Bacteria are single-cell prokaryotic organisms. The interior of the cell contains a colloidal suspension of proteins, carbohydrates, and other complex organic compounds, called the cytoplasm. The cytoplasmic area contains ribonucleic acid (RNA), whose major role is in the synthesis of proteins. Also within the cytoplasm is deoxyribonucleic acid (DNA). DNA contains all the information necessary for the reproduction of all the cell components and may be considered to be the blueprint of the cell. Their usual mode of reproduction is by binary fission or by budding.
Archaea	Similar to bacteria in size and basic cell components. Their cell wall, cell material, and RNA composition is different. Important in anaerobic processes and also found under extreme conditions of temperature and chemical composition.
Fungi/ yeast	Fungi are multicellular, non-photosynthetic, heterotrophic eukaryotes. Most fungi are either strict or facultative aerobes which reproduce sexually or asexually, by fission, budding, or spore formation. Molds or "true fungi" produce microscopic units (hyphae), which collectively form a filamentous mass called the mycelium. Yeasts are fungi that cannot form a mycelium and are therefore unicellular. Fungi have the ability to grow under low-moisture, low nitrogen conditions and can tolerate an environment with a relatively low pH. The ability of the fungi to survive under low pH and nitrogen-limiting conditions, coupled with their ability to degrade cellulose, makes them very important in the composting of sludge.
Protozoa	Protozoa are motile, microscopic eukaryotes that are usually single cells. The majority of protozoa are aerobic heterotrophs, some are aerotolerant anaerobes and a few are anaerobic. Protozoa are generally an order of magnitude larger than bacteria and often consume bacteria as an energy source. In effect, the protozoa act as polishers of the effluents from biological waste treatment processes by consuming bacteria and particulate organic matter.
Helminths	Helminth is a general term used to describe worms collectively. Classified as invertebrates, helminths are usually elongated, flat, or round. The three stages of helminth development are eggs, larval, adult. Worldwide, worms are one of the principal causative agents of human disease.
Rotifers	Rotifers are aerobic heterotrophic animal eukaryotes. The name is derived from the fact that they have two sets of rotating cilia on their head which are used for motility and capturing food. Rotifers are very effective in consuming dispersed and flocculated bacteria and small particles of organic matter. Their presence in an effluent indicates a highly efficient aerobic biological purification process.
Algae	Algae are unicellular or multicellular, autotrophic, photosynthetic eukaryotes or prokaryotes. They are of importance in biological treatment processes. In wastewater treatment lagoons, the ability of algae to produce oxygen by photosynthesis is vital to the ecology of the water environment. The blue-green algae cyanobacteria is a prokaryotic organism.
Viruses	Viruses are composed of a nucleic acid core (either DNA or RNA) surrounded by an outer shell of protein called a capsid. Viruses are infectious agents that only multiply within a host cell, where they redirect the cell's biochemical system to reproduce themselves. Viruses can also exist in an extracellular state in which the virus particle (known as a viron) is metabolically inert. Bacteriaphages are viruses that infect bacteria as the host; they have not been implicated in human infections.

Table 2–22

Typical data on the shape, size, and resistant forms of classes of microorganisms and selected species found in wastewater

Microorganism	Shape	Size, μm[a]	Environmentally resistant form
Bacteria			
Bacilli	rod	0.3–1.5 D × 1–10 L	endospores or dormant cells
Bacillus (*E. coli*)	rod	0.6–1.2 D × 2–3 L	dormant cells
Cocci	spherical	0.5–4	dormant cells
Spirilla	spiral	0.6–2 D × 20–50 L	dormant cells
Vibrio	rod, curved	0.4–2 D × 1–10 L	dormant cells
Protozoa			
Cryptosporidium[b]			
Oocysts	spherical	3–6	oocysts
Sporozite	tear drop	1–3 W × 6–8 L	
Entamoeba histolytica			
Cysts	spherical	10–15 D	cysts
Trophozite	semi-spherical	10–20	
Giardia lamblia[c]			
Cysts	ovid	6–8 W × 8–14 L	cysts
Trophozite	pear or kite	6–8 W × 12–16 L	
Helminths			
Ancylostoma duodenale (hookworm) eggs	elliptical or egg	36–40 W × 55–70 L	filariform larva
Ascaris lumbricoides (roundworm) eggs	lemon or egg	35–50 W × 45–70 L	embryonated egg
Trichuris trichiura (whipworm) eggs	elliptical or egg	20–24 W × 50–55 L	embryonated egg
Viruses			
MS2	spherical	0.022–0.026	virion
Enterovirus	spherical	0.020–0.030	virion
Norwalk	spherical	0.020–0.035	virion
Polio	spherical	0.025–0.030	virion
Rotavirus	spherical	0.070–0.080	virion

[a] D = diameter, L = length, and W = width.

[b] Member of the phylum Apicomplexa.

[c] Member of the phylum Sarcomastigophora, order Diplomonadida.

germination, and outgrowth (Madigan et al., 2009). The resistant forms in protozoans are known as cysts or oocycts. Resistant forms in helminths are eggs and oocysts.

The term *parasite* is used to describe an organism that lives at the expense of another. Parasites that live on the surface of a host organism are *ectoparasites*. Parasites that live internally within the host are known as *endoparasites* (Roberts and Janovy, 1996).

Enumeration and Identification of Microorganisms

In terms of wastewater treatment, the most dramatic change that has occurred is in the identification of the microorganisms found in water and wastewater. The methods now used include a combination of traditional and newer molecular-based methods. The principal methods used for microorganism identification include (1) observational, (2) culture methods, (3) physiological methods, (4) immunological methods, and (5) nucleic-acid based methods. To describe all of the individual tests used within each method would encompass several books, and that is not the purpose here. The principal methods used to enumerate various microorganisms in wastewater treatment are summarized in Table 2–23. These methods are considered briefly in the following discussion. An appreciation of the methods used to both enumerate and identify microorganisms is important in assessing the significance of the data reported in the literature. Additional details on the methods listed in Table 2–23 may be found in Standard Methods (2012) and Maier et al. (2009). Also, because the methods are evolving so rapidly, especially the newer molecular-based methods, the current literature should be consulted for the latest methods.

Observational Methods. The direct observation of microorganisms dates back to the mid-1600s when the microscope was developed. Leeuwenhoek, a Dutch merchant, is credited with providing the first accurate descriptions of bacteria, protozoa, and fungi between 1670 and 1723. Observational methods are still very much in use for algae, fungi, and protozoa as well as other observable microorganisms, such as the filamentous microorganisms common in activated sludge treatment systems (see Chap. 8).

Culture Methods. The traditional culture-based methods of analysis that have been used to enumerate bacteria, as reported in Table 2–23, include (1) pour and spread plate method, (2) membrane filtration, (3) multiple tube fermentation, (4) enzyme substrate coliform tests, (5) heterotrophic plate count, and (6) presence absence test. Because of the widespread use of these tests, they are described further below. Details on these methods, and many others, may be found in Standard Methods (2012).

Pour and Spread Plate Method. Robert Koch, a German scientist regarded as the father of clinical bacteriology, first reported the use of a gel (solid medium) for the culture of microorganisms in 1873. Use of a Petri dish with a solid medium was first reported in 1877, by Petri who worked in Koch's laboratory (Madigan et al., 2009). The pour and spread plate methods are illustrated on Fig. 2–29. The separate distinct bacterial colonies formed on the petri dish after incubation are counted, and the results reported as colony forming units (cfu) per unit volume of sample (typically cfu/mL). In the past, it was assumed that each colony developed from a single bacterium, but use of the term cfu does not assume that one bacteria formed each colony. The use of the pour plate method for the enumeration of bacteriophage is illustrated on Fig. 2–30.

Membrane Filtration. The membrane filtration (MF) method (see Fig. 2–31) evolved from the pour and spread plate methods. Bacteria are retained on the filter because they are larger than the size of the pores of the membrane filter (typically 0.45 μm). After incubation, the colonies formed on the surface of the filter can be counted and the concentration in the original water sample determined. The membrane filter technique has the advantage of being faster than the MPN procedure and of giving a direct count of the number of organisms (e.g., coliform organisms). The results are reported as colony forming units per 100 milliliters (CFU/100 mL).

Multiple Tube Fermentation. The multiple tube method (see Fig. 2–32) for enumerating the number of microorganisms was first proposed by Theobald Smith in the United States

Table 2–23

Representative methods used to enumerate and identify microorganisms and to assess performance of various treatment processes[a]

Test	Description	Typical application
Observational		
Conventional light microscopy	Used for observation of cell morphology and size. Various dyes and stains can be used to enhance visualization and identification. Direct counts can be obtained using a Petroff-Hauser counting chamber.	Cell counts, characterization of filamentous bacteria
Flow cytometry	Light scattering and fluorescence detection with flow of single particles through detector allows for rapid cell counts. Number and size of cells related to change in electric conductivity. Common for algae enumeration.	Cell counts
Electron microscopy	Electron microscopes include transmission and scanning types and are capable of magnification around 10,000,000x. However, the process of obtaining images with electron microscopes can be time consuming and expensive.	Visualization of microorganisms
Culture methods		
Pour and spread plate method	Diluted sample is mixed with agar and poured into culture dish. Agar is allowed to solidify and the dish incubated. After incubation, colonies formed on the agar are counted. Results are reported as colony forming units per milliliter (cfu/mL). In the spread plate method, diluted sample is spread on a surface culture dish containing a suitable culture medium.	Bacterial counts
Membrane filter technique	Sample is passed through a membrane filter and the filter placed right side up in contact with an agar or other solid media. After incubation, colonies formed on the surface of the filter are counted.	Bacterial counts
Multiple-tube fermentation	Sample is diluted serially and added to fermentation tubes and incubated. Positive tubes (cloudy) are counted. Based on the principle of dilution to extinction, as illustrated on Fig. 2–33, *the most probable number per 100 mL* (MPN/100 mL) is computed using the Poisson distribution for extreme values.	Bacterial counts
Enzyme substrate coliform test	Enzyme based methods used to simultaneously determine total coliform bacteria and *E. coli*. Bacterial enzymes present in total coliform group hydrolyze an added substrate resulting in a color change (yellow). *E. coli* cleave a fluorogenic substrate, resuling in the release of fluorogen, which fluoresces under ultraviolet light.	Total coliform bacteria and *E. coli*
Heterotrophic plate count (HPC)	Pour plate, spread plate, or membrane filter method, as described above, can be used to determine HPC. Colonies of bacteria, derived from pairs, chains, clusters, or single cells, are measured. The results are reported as colony forming units per milliliter (CFU/mL).	Bacterial counts

(*continued*)

Table 2–23 (*Continued*)

Test	Description	Typical application
Culture methods (continued)		
Presence-absence (P-A) test	A single 100 mL sample is tested for the P-A of coliform organisms using a selective media. The P-A test is used for highly treated samples such as effluent from a water treatment plant.	Presence of bacteria
Agar overlay method	Sample is mixed with agar and *E. coli*. Solution poured onto solid agar plate and incubated. If colifphage are present, the bacterial cells will lyse resulting in the presence of clear spots. Clear spots, called plaques, are reported as plaque forming units (e.g., pfu/100 mL).	Coliphage counts[b]
Tissue culture (agar overlay method)	Virus assays are performed in the laboratory by inoculating sample concentrate onto monolayers of cultured cells (hence the name tissue culture). Buffalo Green Monkey Kidney (GBMK) is most common cell line for enteroviruses. Virus destroy the infected cells. The destroyed cells appear as a hole or plaque in the cell monolayer. Each plaque (plaque forming unit or PFU) is the result of the presence of a single or a clump of viruses.	Virus counts
Physiological methods		
Respiration gases	Measurement of the rate of gas consumption or production, i.e., oxygen consumption, carbon dioxide evolution, and methane evolution.	Microbial activity and substrate conversion
Microelectrodes	Ulta-fine probes are inserted into a microbial sample followed by continuous measurement of various cell activities, including oxygen uptake and nitrate reduction.	Microbial activity
Labeled constituents	Introduction of a radiolabeled constituents into a microbial sample, including (a) labeled substrate followed by measurement of labeled carbon within the cell, liquid, and evolved as carbon dioxide and (b) labeled thymidine followed by the measurement of the rate of incorporation into DNA.	Microbial activity and substrate conversion
Cell products	Methods include measurement of (a) proteins expressed under variable conditions, (b) enzyme activity through the production of fluorescent products generated by the hydrolysis of fluorescein diacetate, (c) dehydrogenase activity through the reduction of tetrazolium salts, and (d) metabolically active biomass through the adenylate energy charge, or ratio of ATP to total adenylates.	Microbial activity
Immunological methods		
Fluorescent immunolabeling	An antibody is tagged with a fluorescent dye. Once tagged, an antibody becomes attached to an antigen associated with a microorganism; the sample can then be examined using fluorescence microscopy. Fluorescein isothiocyanate (FITC) is the most commonly used fluorescent dye.	Spatial distribution of antigen, detection of bacteria, virus, protozoa, helminths
Enzyme-linked immunosorbent assay (ELISA)	An enzyme-antibody probe is added to a sample containing an antigen. After attachment, the substrate for the enzyme is added to the sample, resulting in a color change.	Quantification of biomass in biofilms, various assays

(*continued*)

Table 2–23 (*Continued*)

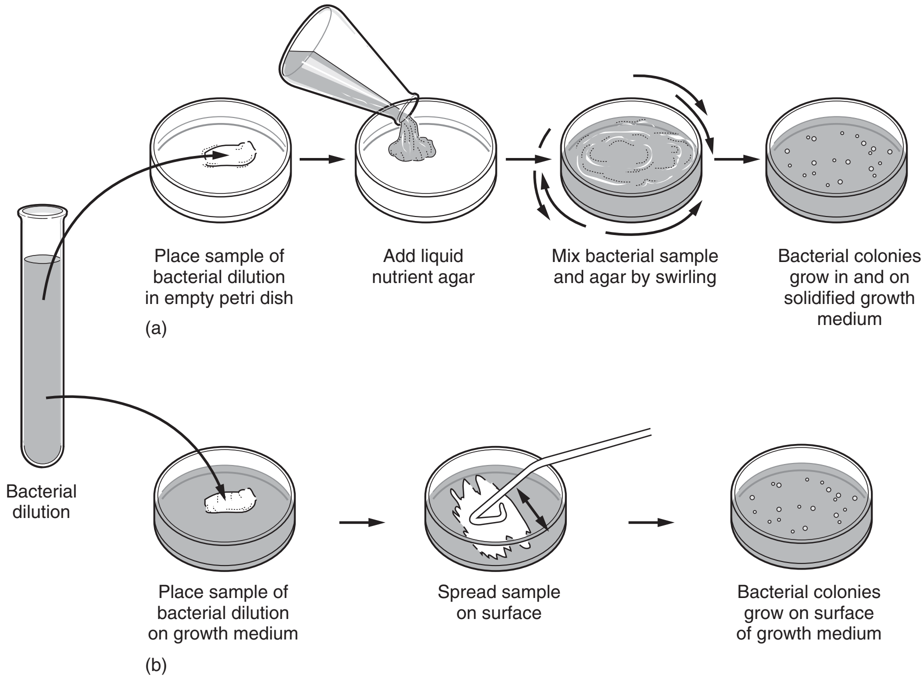
Test	Description	Typical application
Nucleic-acid methods		
Cloning	The process of cloning consists of the insertion of an isolated fragment of DNA of interest into a host cell, typically *E. coli*. The host cell, or clone, then generates identical replicates of the DNA fragment. The DNA fragments are typically analyzed by sequencing.	Replication of genetic material
Nucleic acid probes	A nucleic acid probe is a molecule having a strong interaction with a known complementary genomic sequence unique to the targeted organism(s) and possessing a means for detection. Typical methods include (a) fluorescent in-situ hybridization (FISH), (b) detection of DNA or RNA on gels following electrophoresis, and (c) screening gene expression using an array of gene probes known as microarrays.	Identification of specific microorganisms, including spatial distribution in flocs and biofilms
Polymerase chain reaction (PCR)	Amplification of the DNA of the genome of the microorganisms being tested by using complementary DNA fragments known as primers to bind to target DNA of virus The primer triggers a reaction which results in the production of many millions of copies of the microorganism DNA. Examples include reverse transcription PCR (RT-PCR), nested PCR, multiplex PCR, integrated cell culture PCR (ICC-PCR), and real-time quantitative PCR (qPCR).	Amplification of genetic material
Sequencing	The coding of genetic material can be determined through the process of sequencing, which usually takes place at commercial laboratories. The DNA sequence can then be compared to databases to determine the relationship of the genetic material to other organisms that have had their DNA sequenced. The 16S rRNA region of the sequence has been found to be the most useful for determining the identity of isolated microorganisms.	Identification of isolated microorganisms
Restriction fragment length polymorphism (RFLP)	A method that uses enzymes to cut purified DNA or PCR products into small fragments at specific segments of the genome. The fragments are then analyzed by gel or capillary electrophoresis to obtain a microbial community fingerprint.	Microbial community fingerprint
Gel gradient electrophoresis	A method that subjects PCR fragments to an increasing concentration of denaturant (DGGE) or temperature (TGGE) to allow for visualization of diversity in the genetic material resulting from differential melting or denaturing of the PCR fragments.	Microbial community diversity
Metagenomics	The analysis of the collective genetic material recovered from an environment sample.	Microbial community diversity and metabolism

[a] Adapted from Ingraham and Ingraham (1995), Madigan et al. (2009), Maier et al. (2009),and Stanier et al. (1986).

[b] A bacteriophage is a virus that infects and replicates within bacteria. A coliphage is a type of bacteriophage that infects *E. coli*.

Figure 2–29

Schematic of plate culture methods used for the enumeration of bacteria: (a) pour plate and (b) spread plate.

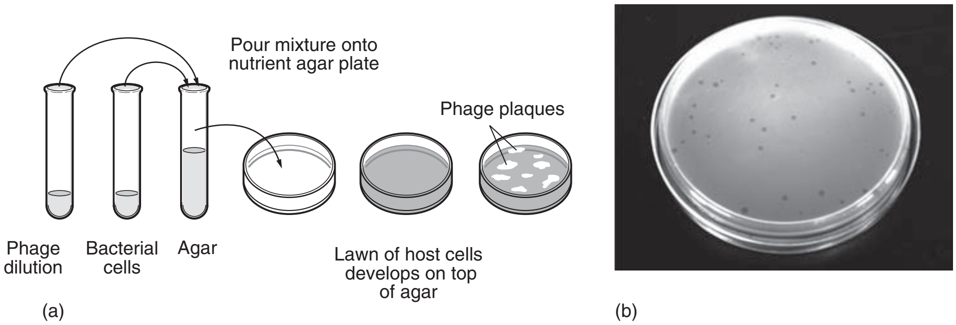


in 1893 (Smith, 1893). The method is based on the principle of dilution to extinction as illustrated on Fig. 2–32. Initially, the results obtained with the multiple tube method were identified as the *indicated number*. The name was changed to multiple tube method in the 1930s. Concentrations of total coliform bacteria are typically reported as the *most probable number per 100 mL* (MPN/100mL). The MPN is based on the application of the Poisson distribution for extreme values to the analysis of the number of positive and negative results obtained when testing multiple portions of equal volume and in portions constituting a geometric series. It is emphasized that the MPN is not the absolute concentration of organisms that are present, but only a statistical estimate of that concentration.

Enzyme Substrate Coliform Tests. In addition to the modified MPN test, several commercial enzymatic assays have been developed that can be used for the simultaneous detection of both total coliform bacteria and *E. coli*. In the enzymatic assays, powdered ingredients

Figure 2–30

Schematic of the technique used for the enumeration of coliphage: (a) schematic, (b) development of coliphage on *E coli* lawn. The clear spots correspond to coliphage colonies (see footnote b, Table 2–23).



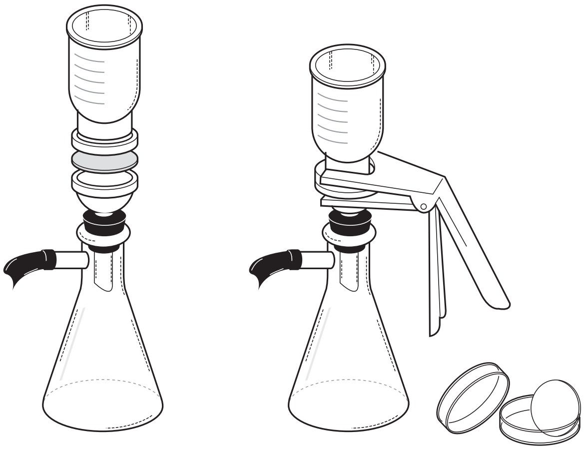


Figure 2–31

Membrane filter apparatus used to test for bacteria in relatively clean waters. After centering the membrane filter on the filter support, the funnel top is attached and the water sample to be tested is poured into the funnel. To aid in the filtration process, a vacuum line is attached to the base of the filter apparatus. After the sample has been filtered, the membrane filter is placed in a petri dish containing a culture medium for incubation and subsequent bacterial enumeration.

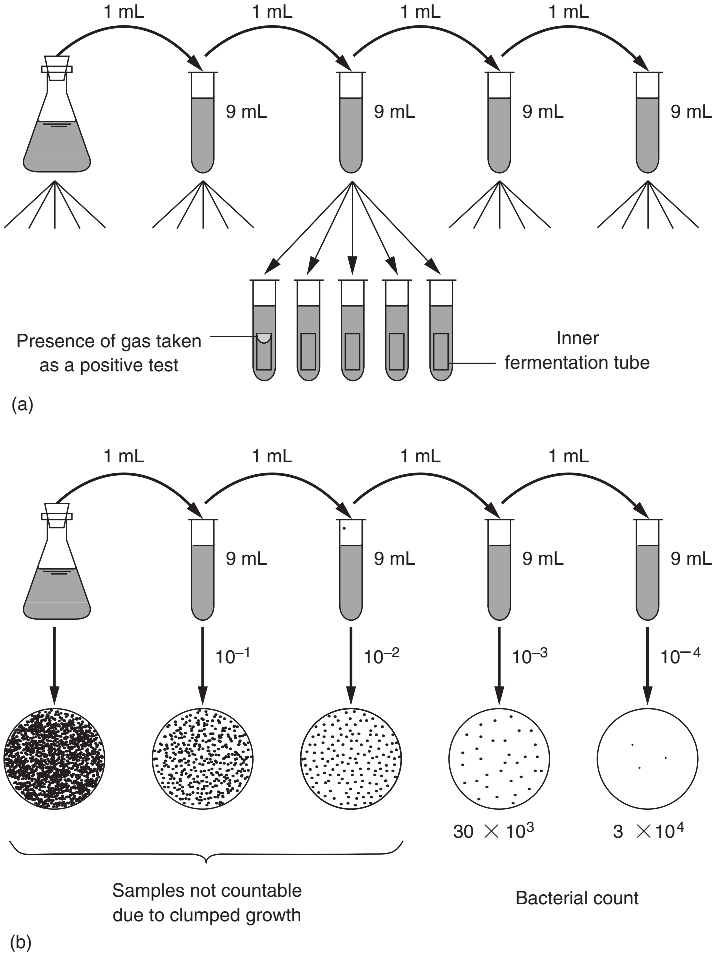


Figure 2–32

Schematic illustration of the methods used to obtain bacterial counts: (a) multiple tube fermentation technique using a liquid medium and (b) plate count method using a solid medium.

comprised of salts and specific enzyme substrates that serve as the sole carbon source are added to various wastewater samples. When metabolized by total coliform and *E coli*, the specific enzyme substrates produce a yellow color and or fluoresce. After incubation, samples containing coliform organisms turn yellow, and samples containing *E. coli* will fluoresce when exposed to long-wave UV illumination [see Fig. 2–33(a)]. The enzymatic test can be used in two different modes: presence/absence and quantification. In the presence/absence mode the chemical ingredients are added to 100 mL bottles containing the sample to be analyzed. The quantification mode can be carried out using the multiple-tube method or specialized apparatus such as the Colilert-18/Quanti-Tray method [see Fig. 2–33(b)]. The results are reported as present or absent in a 100 mL sample and as MPN/100mL in the quantification test.

Heterotrophic Plate Count. The heterotrophic plate count (HPC) is a procedure for estimating the number of live heterotrophic bacteria in wastewater samples. The HPC method

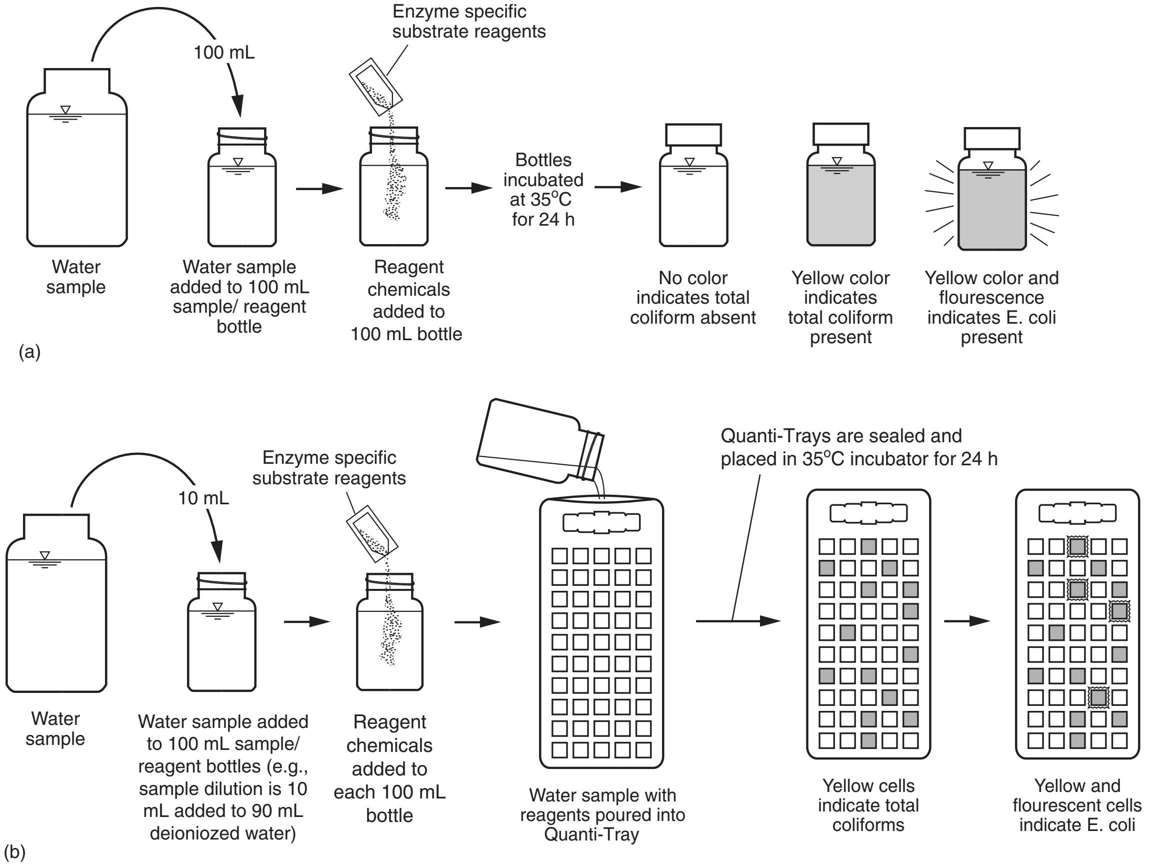


Figure 2–33

Schematic of the enzyme specific substrate coliform test for total coliform and *E coli:* (a) presence absence test using 100 mL bottles and (b) quantification test using Quanti-Tray apparatus. Note that sample dilutions can be prepared as necessary for a given sample (dilution shown is for example only). When using the Quanti-Tray, the number of total coliform or *E. coli* organisms present in the sample can be determined by counting the positive wells and then using IDEXX most probable number (MPN) tables.

evolved from first plate count method which was included in the first edition of Standard Methods (Standard Methods, 1905). The HPC can be determined using the (1) pour plate method, (2) spread plate method, or (3) membrane filter method as described above. In the HPC test, colonies of bacteria, which may be derived from pairs, chains, clusters, or single cells, are measured. The results are reported as colony forming units per milliliter (CFU/mL).

Presence Absence Test. The presence-absence (P-A) test for coliform organisms is a modification of the multiple tube fermentation technique described above. The test is used intended for use for samples collected from water distribution systems or water treatment plants. Rather than using multiple dilutions, a single 100 mL sample is tested for the P-A of coliform organisms using lauryl sulfate tryptose lactose broth as used in the MPN test. Coliform organisms are present if a distinct yellow color forms, indicating that lactate fermentation has occurred in the sample. The test is based on the rationale that no organisms should be present in 100 mL. It has been used in wastewater for highly treated samples. Enzymatic methods are also used in the P-A test.

Physiological Methods. Physiological methods are used to address the metabolic processes carried out by a microbial community, Measurement of microbial activity can be used to estimate the amount of active biomass, the response of a microbial system to a disturbance, and for determination of the status of an engineered biological process, such as biological wastewater treatment or composting of organic wastes. Common examples of physiological methods include measurement of the rate and type of substrate utilization, the rate of oxygen uptake, or the formation of respiration products.

Immunological Methods. Analytical methods that utilize an antibody for the detection or quantification of a target antigen are known as immunoassays. A key element of immunoassays is the visualization of the antibody-antigen interaction. The visualization step is accomplished typically through the use of signal molecules, which can be designed to allow for detection by various mechanisms, including color change, fluorescence, and radioactivity. Signal molecules can be attached to the antibody directly, known as direct labeling, or the signal molecule may be added after the antibody-antigen attachment has taken place, known as indirect labeling. Indirect labeling uses a secondary antibody to attach with the primary antibody, which is already attached to the target antigen. While nonspecific attachment is a key disadvantage for many immunoassays, the ability to observe the spatial arrangement of target microorganisms is a powerful advantage.

Nucleic-Acid-Based Methods. Molecular methods are based on the use of specific (DNA or RNA) sequences to identify microorganisms, and the use of DNA or RNA amplification procedures (e.g., polymerase chain reaction, PCR) to detect extremely low concentrations of nucleic acid. The PCR technique was developed by Dr. Kary B. Mullis in 1983 while he worked as chemist at the Cetus Corporation. He won the Nobel Prize in 1993 for his invention. Since 1983, the original procedure has been improved and modified to include a number of procedures such as those reported in Table 2–21. Additional details may be found in Madigan et al. (2009) and Maier et al. (2009). Application of some of these procedures to identify and follow specific microbial populations and activity are presented in Chap. 7.

Pathogenic Organisms and Prions

Pathogenic organisms and agents found in wastewater may be excreted by human beings and animals who are infected with disease or who are carriers of a particular infectious disease. The pathogenic organisms found in wastewater can be classified into four broad categories: bacteria, protozoa, helminths, and viruses. The principal pathogenic organisms found in untreated wastewater are reported in Table 2–24, along with the diseases and

Table 2–24

Infectious agents potentially present in untreated domestic wastewater[a]

Organism	Disease	Remarks/symptoms
Bacteria		
Campylobacter jejuni	Gastroenteritis	Diarrhea
Escherichia coli (Enteropathogenic)	Gastroenteritis	Diarrhea
Legionella pneumophila	Legionnaires disease	Malaise, myalgia, fever, headache, respiratory illness
Leptospira (spp.)	Leptospirosis	Jaundice, fever (Weil's disease)
Salmonella typhi	Typhoid fever	High fever, diarrhea, ulceration of small intestine
Salmonella ($\approx$2,100 serotypes)	Salmonellosis	Food poisoning
Shigella (4 spp.)	Shigellosis	Bacillary dysentery
Vibrio cholera	Cholera	Extremely heavy diarrhea, dehydration
Yersinia enterolitica	Yersinosis	Diarrhea
Protozoa		
Balantidium coli	Balantidiasis	Diarrhea, dysentry
Cryptosporidium parvum	Cryptosporidiosis	Diarrhea
Cyclospora cayetanensis	Cyclosporasis	Severe diarrhea, stomach cramps, nausea, and vomiting lasting for extended periods
Entamoeba histolytica	Amebiasis (Amoebic dysentery)	Prolonged diarrhea with bleeding, abscesses of the liver and small intestine
Giardia lamblia	Giardiasis	Mild to severe diarrhea, nausea, indigestion
Helminths[b]		
Ascaris lumbricoides	Ascariasis	Roundworm infestation
Enterobius vermicularis	Enterobiasis	Pinworm
Fasciola hepatica	Fascioliasis	Sheep liver fluke
Hymenolepis nana	Hymenolepiasis	Dwarf tapeworm
Taenia saginata	Taeniasis	Beef tapeworm
T. solium	Taeniasis	Pork tapeworm
Trichuris trichiura	Trichuriasis	Whipworm
Viruses		
Adenovirus (31 types)	Respiratory disease and gastrointestinal illness	
Enteroviruses (more than 100 types, e.g., polio, echo, and coxsakie viruses)	Gastroenteritis, heart anomalies, meningitis	
Hepatitis A virus	Infectious hepatitis	Jaundice, fever
Norovirus	Gastroenteritis	Vomiting
Pavrovirus (2 types)	Gastroenteritis	
Rotavirus	Gastroenteritis	

[a] Adapted from Feachem et al. (1983), Madigan et al. (2009), and Crook (1998).

[b] The helminths listed are those with a worldwide distribution.

Table 2–25

Microorganism concentration found in untreated wastewater and septic tank effluent and the corresponding infectious dose[a]

Organism	Concentration in raw wastewater[b] MPN/100 mL	Infectious dose, number of organisms[c]
Bacteria		
Bacteroides	10^7–10^{10}	
Colifom, total	10^7–10^9	
Colifom, fecal	10^6–10^8	10^6–10^{10}
Coliform, *E coli.*	10^5–10^7	
Clostridium perfringens	10^3–10^5	1–10^{10}
Enterococci	10^4–10^5	
Fecal streptococci	10^4–10^7	
Pseudomonas aeruginosa	10^3–10^6	
Shigella	10^0–10^3	10–20
Salmonella	10^2–10^4	
Protozoa		
Cryptosporidium parvum oocysts	10^1–10^3	1–10
Entamoeba histolytica cysts	10^{-1}–10^1	10–20
Giardia lamblia cysts	10^3–10^4	<20
Helminth		
Ova	10^1–10^3	
Ascaris lumbricoides	10^{-2}–10^0	1–10
Viruses		
Enteric virus	10^3–10^4	1–10
Coliphage	10^3–10^4	

[a] Adapted in part from Crook (1998) and Feacham et al. (1983).
[b] Value will vary with portion of population shedding at any given time.
[c] Infections dose will vary with serotype or strain of organism, and the individual's general health.

disease symptoms associated with each pathogen. Bacterial pathogenic organisms of human origin typically cause diseases of the gastrointestinal tract, such as typhoid and paratyphoid fever, dysentery, diarrhea, and cholera. Because these organisms are highly infectious, they are responsible for many thousands of deaths each year in areas with poor sanitation, especially in the tropics. It has been estimated that that up to 4.5 billion people are or have been infected with some parasite (Madigan et al., 2009). Typical data on the quantity of selected pathogenic organisms found in wastewater and the corresponding concentration needed for an infectious dose are reported in Table 2–25.

Bacteria. Many types of harmless bacteria colonize the human intestinal tract and are routinely shed in the feces. Because pathogenic bacteria are present in the feces of infected individuals, domestic wastewater contains a wide variety and concentration range of nonpathogenic and pathogenic bacteria. One of the most common bacterial pathogens found in domestic wastewater is the genus *Salmonella.* The *Salmonella* group contains a wide variety of species that can cause disease in humans and animals. Typhoid fever, caused by *Salmonella typhi*, which is specific to humans, is the most severe and serious. The most common disease associated with *Salmonella* is food poisoning identified as salmonellosis. *Shigella*, a less common genus of bacteria, is responsible for an intestinal disease known as bacillary dysentery or shigellosis. Waterborne outbreaks of shigellosis have been reported from recreational swimming areas and where wastewater has contaminated wells used for drinking water (Crook, 1998; Maier et al., 2009).

Other bacteria isolated from raw wastewater include *Vibrio, Mycobacterium, Clostridium, Leptospira* and *Yersinia* species. *Vibrio cholerae* is the disease agent for cholera, which is not common in the United States but is still prevalent in other parts of the world. Humans are the only known hosts, and the most frequent mode of transmission is through water. *Mycobacterium tuberculosis* has been found in municipal wastewater, and outbreaks have been reported among persons swimming in water contaminated with wastewater (Crook, 1998; Maier et al., 2009).

Waterborne gastroenteritis of unknown cause is frequently reported, with the suspected agent being bacterial. One potential source of this disease is certain gram-negative bacteria normally considered to be nonpathogenic. These include the enteropathogenic *Escherichia coli* and certain strains of *Pseudomonas,* which may affect the newborn and have been implicated in gastrointestinal disease outbreaks. *Campylobacter jejuni* has been identified as the cause of a form of bacterial diarrhea in humans. While it has been well established that this organism causes disease in animals, it has also been implicated as the etiologic agent in human waterborne disease outbreaks (Crook, 1998).

Protozoa. Of the disease causing organisms reported in Table 2–24, the protozoans *Cryptosporidium parvum, Cyclospora, Entamoeba histolytica,* and *Giardia lamblia* (see Fig. 2–34) are of great concern because of their significant impact on individuals with compromised immune systems, including very young children, the elderly, persons undergoing treatment for cancer, and individuals with acquired immune deficiency syndrome (AIDS). The life cycle of *Cryptosporidium parvum* and *Giardia lamblia* is illustrated on Fig. 2–35. As shown, infection is caused by the ingestion of water contaminated with oocycts and cysts. It is also important to note that numerous nonhuman sources of *Cryptosporidium parvum* and *Giardia lamblia* are present in the environment. Further, not all of the oocycsts and cysts that are present are viable in terms of their ability to cause disease. To determine the potential risk from these microorganisms, infectivity studies must be conducted.

Pathogenic protozoan disease outbreaks have been significant, highlighted by the 1993 outbreak of cryptosporidiosis in Milwaukee in which it is estimated that more than 400,000 persons became ill and that there were outbreaks of cyclosporiasis in ten other

Figure 2–34

Definition sketch for (a) *Giardia lamblia* cyst and trophozite and (b) *Cryptosporidium parvum* oocyst and sporozite.

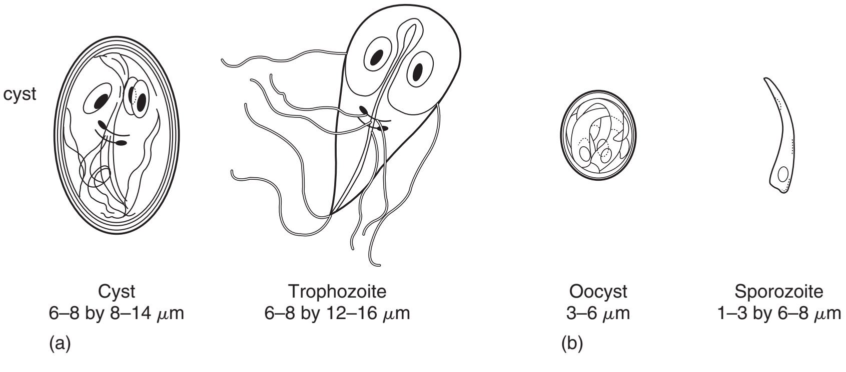
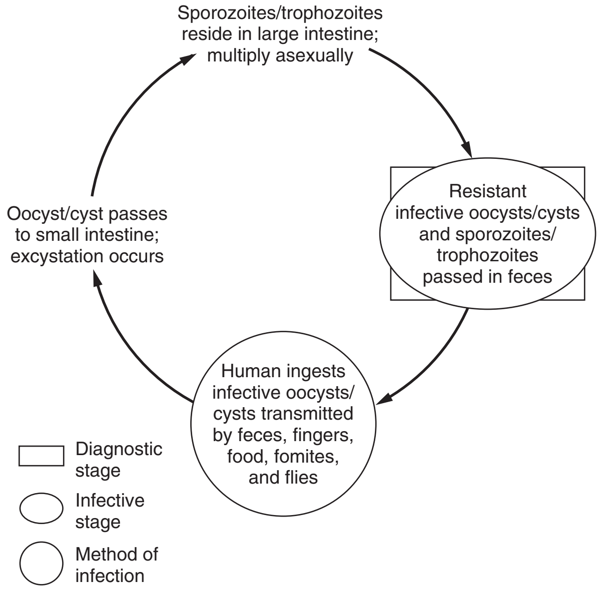


Figure 2–35

Life cycle of *Cryptosporidium parvum* and *Giardia lamblia.*



states. As noted in Table 2–22, these protozoan organisms may cause symptoms which can include severe diarrhea, stomach cramps, nausea, and vomiting lasting for extended periods. Despite intensive trials in humans and animals, no effective treatment has been found for cryptosporidiosis (Roberts and Janovy, 1996). The oocysts of *Cryptosporidium parvum* and the cysts of *Giardia lamblia* are the most resistant forms (see Table 2–22). These organisms are of particular concern because they are found in almost all wastewaters and because conventional disinfection techniques using chlorine have not proven to be effective in their inactivation or destruction. However, it has been found that UV disinfection is extremely effective in the inactivation of oocysts of *Cryptosporidium parvum* and the cysts of *Giardia lamblia*.

Helminths. The term *helminths* is used to describe parasitic worms collectively. In the United States, as result of the improvements in the provision of sanitation and wastewater treatment facilities and in food handling practices, the prevalence of helminth infections has decreased dramatically over the last century. Nevertheless, due to increased levels of immigration to the United States of persons from countries where worms are endemic, the transmission of helminths by wastewater and particularly by biosolids remains a concern. In fact, the eggs of worms are found in wastewater throughout the United States. In particular, small, nonparasitic nematodes are universally present, even in finished drinking water at the tap (Cooper, 2012). As noted in Table 2–21, worldwide, worms are one of the principal causative agents of human disease. It is estimated that the number of human infections caused by helminths collectively is on the order of 4.5 billion (Roberts and Janovy, 1996).

Most of the helminths fall into three major phyla: Nematoda (roundworms), Platyhelminths (flatworms), and Annelida (segmented worms). Most human infections are associated with nematodes and flat worms, while the segmented worms are primarily ectoparisitic, such as leaches. The phylum Nematoda collectively represents one of the most abundant animal groups on earth, most of which are harmless to humans. Included among its pathogenic members are the large round worm *Ascaris lumbricoides,* the whipworm *Trichuris trichiura,* the hookworms *Necator americanus* and *Ancylostoma duodenale,* and the threadworm *Strongyloides stercoralis*. *Ascaris lumbricoides* is considered to be the most prevalent parasitic infection worldwide with over one and a half billion persons infected (Crompton, 1999; Maier et al., 2009; Roberts and Janovy, 1996). It has been estimated that there are on the order of 4 million cases in the United States (Khuroo, 1996).

The phylum platyhelminths includes the tapeworms *Taenia saginata* (beef tapeworm), *Taenia solium* (pork tapeworm), and *Schistosoma* species. *Taenia saginata*, transmitted primarily by the ingestion of infected beef products, is the most common tapeworm found in humans. The trematodes *Schistosoma mansoni*, *S. haematobium, and S. japonicum,* also known as blood flukes, are medically important members of the trematoda class. More than 200 million infections are ascribed to these worms worldwide. It is estimated that more than 400,000 infected individuals, most of whom were infected outside of the United States, live in the United States (West and Olds, 1992).

The human infective stage of helminths varies; in some species it is either the adult organism or larvae, while in other species it is the eggs, but it is primarily the eggs that are present in wastewater. Helminth eggs, which range in size from about 10 μm to more than 100 μm, can be removed by many commonly used wastewater treatment processes such as sedimentation, filtration, and stabilization ponds. However, some helminth eggs are extremely resistant to environmental stresses and may survive usual wastewater and sludge disinfection procedures. Chlorine disinfection and mesophilic anaerobic digestion, for example, are not effective at inactivating many helminth eggs. It has been found that the eggs of *Ascaris* can survive for up to ten years in the sediments of oxidation ponds (Nelson, 2011). The long survival times of *Ascaris* and other worm eggs are of particular importance in the management of biosolids.

Viruses. More than 100 different types of enteric viruses capable of producing infection or disease are excreted by humans. Enteric viruses multiply in the intestinal tract and are released in the fecal matter of infected persons. From the standpoint of health, the most important human enteric viruses are the enteroviruses (polio, echo, and coxsackie), caliciviruses including noroviruses (formally known as Norwalk agents), rotaviruses, reoviruses, adenoviruses, and hepatitis A virus. Of the viruses that cause diarrheal disease, only the Norwalk virus and rotavirus have been shown to be major waterborne pathogens. The reoviruses and adenoviruses, known to cause respiratory illness, gastroenteritis, and eye infections, have been isolated from wastewater. There is no evidence that the human immunodeficiency virus (HIV), the pathogen that causes the acquired immunodeficiency syndrome (AIDS), can be transmitted via the waterborne route (Crook, 1998; Madigan et al., 2009; Maier et. al., 2009; Rose and Gerba, 1991). The biology of viruses is delineated in Voyles (1993).

Prions. Prions consist of small protein molecules that have a distinct extracellular form but do not contain either DNA or RNA. Interest in prions stems from the fact that they can cause disease in animals, such as mad cow disease and scrapie in sheep, and can also infect humans. Prions can exist in two forms: healthy or pathogenic. The healthy form is found in most animals. The pathogenic form enters the host, which contains healthy prions, and replicates by converting the healthy prions present in the host to the pathogenic form. To date, no model has been put forth to explain how the pathogenic prion subverts the healthy prion to become pathogenic. A public health assessment of prion diseases in humans may be found in Belay and Schonberger (2005).

One form of prion disease has been observed in humans that have consumed contaminated meat from cattle infected with pathogenic prions (Johnson et al., 2011). Pathogenic prions may also be transferred in animal manures, thus there is potential for prions to be present in the waste from animal facilities such as slaughterhouses where infected tissues may be present. In wastewater treatment, concern centers on the fact that prions are

Table 2–26

Typical pathogen survival times at 20–30°C in various environments[a]

Pathogen	Survival time, d Fresh water and wastewater	 Crops	 Soil
Bacteria			
Fecal coliforms[b]	<60 but usually <30	<30 but usually <15	<120 but usually <50
Salmonella spp[b]	<60 but usually <30	<30 but usually <15	<120 but usually <50
Shigella[b]	<30 but usually <10	<10 but usually <5	<120 but usually <50
Vibrio cholerae[c]	<30 but usually <10	<5 but usually <2	<120 but usually <50
Protozoa			
E. histolytica cysts	<30 but usually <15	<10 but usually <2	<20 but usually <10
Helminths			
A. lumbricoides eggs	Many months	<60 but usually <30	<Many months
Viruses[b]			
Enteroviruses[d]	<120 but usually <50	<60 but usually <15	<100 but usually <20

[a] Adapted from Feacham et al. (1983).

[b] In seawater, viral survival is less, and bacterial survival is very much less than in fresh water.

[c] *V. cholerae* survival in aqueous environments is a subject of current uncertainty.

[d] Includes polio, echo, and coxsackie viruses.

difficult to inactivate by conventional means. Temperatures exceeding 100°C have been found to be an effective treatment strategy (Kirchmayer et al., 2006). Because they have been found in digested biosolids, there is also concern about their potential impacts on farm animals as a result of land application.

Survival of Pathogenic Organisms. Of great concern in the management of disease-causing organisms is the survival of these organisms in the environment. Typical data on the survival of microorganisms in the environment are presented in Table 2–26. Although the data given in Table 2–26 can be used as a rough guide, numerous exceptions have been reported in the literature.

Use of Indicator Organisms. Because the numbers of pathogenic organisms present in wastes and polluted waters are usually few and difficult to isolate and identify, microorganisms, which are more numerous and more easily tested for, are commonly used as surrogate (i.e., indicator) organisms for the target pathogen(s). The general features of an ideal indicator organism and the use of bacterial and other indicators is considered briefly in the following discussion.

Characteristics of an Ideal Indicator Organism. An ideal organism should have the following characteristics (adapted from Cooper 2012; Maier et al., 2009):

1. The indicator organism must be present when fecal contamination is present.
2. The numbers of indicator organisms present should be equal to or greater than those of the target pathogenic microorganism.
3. The indicator organism must exhibit the same or greater survival characteristics in the environment as the target pathogen organism for which it is a surrogate.

4. The isolation and quantification procedure for the indicator organism must be faster, simpler, and less expensive than that of the target pathogen.
5. The organism should be a member of the intestinal microflora of warm-blooded animals.

Some authors have stated the first characteristic as "the indicator organism must be present when the target pathogen is present." Unfortunately, the target pathogen(s) may not be present during the entire year because the shedding rate of pathogenic organisms is not uniform throughout the year. Thus, it is important that the indicator organism be present when fecal contamination is present if public health is to be protected. To date, no ideal indicator organism has yet been found. Microorganisms that have been proposed for use as indicators of fecal contamination are summarized in Table 2–27. Indicator organisms that have been used to establish performance criteria for various water uses are reported in Table 2–28.

Table 2–27

Specific organisms that have been used or proposed for use as indicators of fecal contamination

Indicator organism	Characteristics
Total coliform bacteria	Species of gram-negative rods which ferment lactose with gas production (or produce a distinctive colony within 24 ± 2 h to 48 ± 3 h incubation on a suitable medium) at 35 ± 0.5°C. There are strains which do not conform to the definition. The total coliform group includes four genera in the *Enterobacteriaceae* family: *Escherichia*, *Citrobactor*, *Enterobacter*, and *Klebisella*. Of the group, the *Escherichia* genus (*E. coli* species) appears to be most representative of fecal contamination.
Fecal coliform bacteria	A fecal coliform bacteria group was established based on the ability to produce gas (or colonies) at an elevated incubation temperature (44.5 ± 0.2°C for 24 ± 2 h).
Klebisella	The total coliform population includes the genera *Klebisella*. The thermotolerant *Klebisella* are also included in the fecal coliform group. This group is cultured at 35 ± 0.5°C for 24 ± 2 h.
E. coli	The *E. coli* is one of the coliform bacteria population and is more representative of fecal sources than other coliform genera.
Bacteroides	Bacteroides, an anaerobic organism, has been proposed as a human specific indicator.
Fecal Streptococci	This group had been used in conjunction with fecal coliforms to determine the source of recent fecal contamination (man or farm animals). Several strains appear to be ubiquitous and cannot be distinguished from the true fecal streptococci under usual analytical procedures, which detract from their use as an indicator organism.
Enterococci	Two strains of fecal streptococci, S. *faecalis* and S. *faecium*, are the most human specific members of the fecal streptococcus group. By eliminating the other strains through the analytical procedures, the two strains known as enterocci can be isolated and enumerated. The enterocci are generally found in lower numbers than other indicator organisms, however, they exhibit better survival in seawater.
Clostridium perfringens	This organism is a spore-forming anaerobic persistent bacteria, and the characteristics make it a desirable indicator where disinfection is employed, where pollution may have occurred in the past, or where the interval before analysis is protracted.
P. aeruginosa and A. hydrophila	These organisms may be present in sewage in large numbers. Both can be considered aquatic organisms and can be recovered in water in the absence of immediate sources of fecal pollution.

Table 2–28

Indicator organisms used in establishing performance criteria for various water uses

Water use	Indicator organism
Drinking water	Total coliform Fecal coliform *E. coli*
Freshwater recreation	Fecal coliform *E. coli* Enterococci
Saltwater recreation	Fecal coliform Total coliform Enterococci
Shellfish growing areas	Total coliform Fecal coliform
Agricultural irrigation (For reclaimed water)	Total coliform
Wastewater effluent	Total coliform
Disinfection	Total coliform Fecal coliform *E. coli* MS2 coliphage

Use of Coliform Organisms as Indicators. The intestinal tract of humans contains a large population of rod-shaped bacteria known collectively as coliform bacteria. Each person discharges from 100 to 400 billion coliform bacteria per day, in addition to other kinds of bacteria. Thus, the presence of coliform bacteria in environmental samples has, over the years, been taken as an indication that pathogenic organisms associated with feces (e.g., viruses) may also be present. The absence of coliform bacteria is taken as an indication that the water is free from disease producing organisms.

The total coliform group includes four genera in the *Enterobacteriaceae* family: *Escherichia, Citrobactor, Enterobacter*, and *Klebisella*. The coliform bacteria have common biochemical and morphological attributes. Typically, these organisms are gram negative, non-spore forming, rod-shaped organisms (see Fig. 2–36) that ferment lactose in 24 to 48 hours at 35 ± 0.5°C (Standard Methods, 2012). The term *gram negative* refers to a staining procedure used to differentiate groups of organisms. Three groups of coliform organisms that have been used as bacterial indicators are total coliform, fecal coliform, and *E coli*.

Total Coliform. The coliform group of bacteria is commonly found in the environment in soil or vegetation and in the intestines and feces of humans and warm-blooded animals. In general, coliform bacteria are harmless. When total coliform are detected in drinking water, their presence is taken as an indication that disease-causing pathogens could also potentially be present. If total coliforms are detected in drinking water, repeat samples are taken to verify their presence, and an investigation is undertaken to determine whether the water treatment and distribution system has been breached and to locate the source of contamination, if one exists.

Fecal Coliform. Fecal coliform, sub-group of the total coliform group, are found in large quantities in the intestines and feces of humans and warm-blooded animals. The presence of fecal coliform in drinking water is taken as an indication that there is a greater risk,

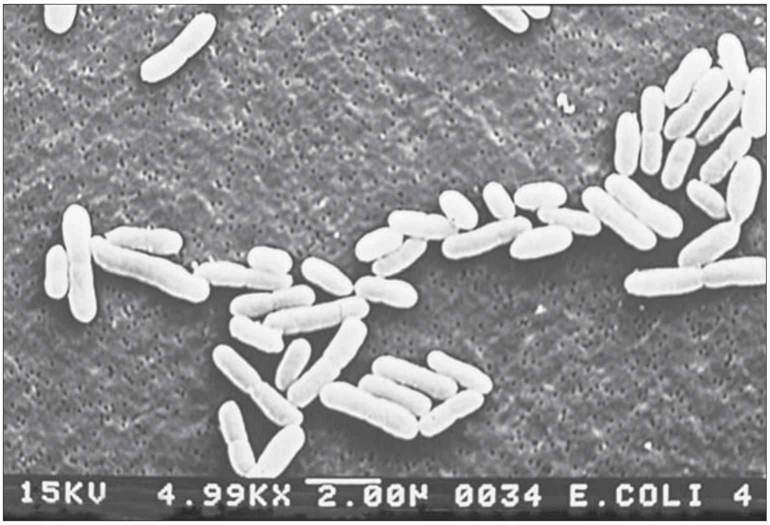


Figure 2–36

Micrograph of a pure culture of *E coli*.

as compared to the presence of total coliform, that disease-causing pathogens could also be present. If fecal coliform are detected, the same procedure as discussed for total coliform is followed to identify and eliminate the source of contamination, if one exists.

E. coli. The organism *Escherichia coli* (*E. coli*), specific to the intestines and feces of humans and warm blooded animals, has historically been the target organism tested for with the *total coliform test*. Although most *E. coli* strains are harmless to humans, serotype O157:H7 can cause food poisoning and, in some cases, become life threatening. Early on, it was found that the coliform test was not specific for fecal coliform or *E. coli* and that a variety of other coliform organisms were included in the test results. In recent years, tests have been developed that distinguish among total coliforms, fecal coliforms, and *E. coli*, and all three are currently being reported in the literature. The presence of *E. coli* in drinking water is taken as an indication that there is a greater risk, as compared to the presence of fecal or total coliform, that disease-causing pathogens could also be present. If *E. coli* are detected, the same procedure, as discussed above for total coliform, is followed to identify and eliminate the source of contamination, if one exists.

As a historical note, *Escherichia coli* (*E. coli*) is named after the German pediatrician Theodor Escherich (1857–1911), who first discovered this species in the feces of healthy infants and reported his findings in 1885. Initially, it was called *Bacterium coli commune* because it was found in the colon. Following a revision of the domain *Bacteria* in the late 1890s, it was reclassified in 1895 as *Bacillus coli*. In 1919, it was named *Escherichia coli* after its original discoverer (Castllani and Chalmers, 1919). The name *Escherichia coli* was adopted for use in Standard Methods in 1958. One of the most recent and important developments in the evolution of the coliform test is the ability to identify and quantify *E. coli* through the use of elevated temperatures and a specific growth medium 4-methylumbelliferyl-β-D-glucuronide (MUG). When *E. coli* are present, they are able to cleave the fluorogenic substrate MUG from the growth medium because they possess a specific enzyme (β-glucuronidase). The presence of a bright blue flourescnce is taken as a positive response for *E coli*.

Other Indicator Organisms. While total and fecal coliform organisms and *E. coli* may be present, it has not been demonstrated that they are indicators of the presence of enteric viruses and protozoa. Further, concerns for newly emerging pathogenic organisms which may arise from nonhuman reservoirs (e.g., pathogenic *E. coli, Cryptosporidium parvum,* and *Giardia lamblia*) have led to the questioning of the use of indicators that arise primarily from fecal inputs. In a recent study, it was concluded that coliform bacteria are adequate indicators for the potential presence of pathogenic bacteria and viruses but are inadequate as an indicator of the presence of waterborne protozoa. Waterborne disease outbreaks have also occurred in drinking water systems that have not violated their microbial water quality standards (Craun et al., 1997).

Given the limitations in using coliform organisms as indicators of potential contamination by wastewater, attention has now focused on the use of bacteriophages as an indicator organism and more specifically as indicators of enteric viruses. Bacteriophages are viruses that can infect prokaryotic cells. There are six major families of bacteriophages, five of which are DNA-based and one of which is RNA-based. Of the five DNA-based bacteriophages, three are double stranded and two are single stranded. Bacteriophages that infect *E. coli* are known as coliphages. Coliphages that attach directly to the cell wall are known as *somatic*. Coliphages that infect only male strains of *E. coli* (possess pilli) are known as male-specific (F^+) coliphages. The source of male-specific phages is thought only to be feces.

Within the male specific family there are four serotypes. Groups II and III are primarily of human origin whereas groups I and IV are of animal origin, with the exception of pigs, which may harbor groups II and III. Interest in using coliphages as an indicator of enterovirses is based on the fact that the phages are approximately the same size as most enteric viruses of interest (e.g., polio), are of fecal origin, and are always present in raw municipal wastewater. Coliphages have been used extensively in disinfection studies (see Sec. 12–9 in Chap. 12).

Evolving Pathogenic Microorganisms

In recent years there has been a disturbing increase in the number of disease outbreaks in the United States and in many other parts of the world, especially in light of the fact that it was thought that a number of endemic contagious diseases had been controlled or eliminated (only smallpox to date) (Levins et al., 1994). The bacteria *Legionella pneumophila,* the causative agent in Legionnaire's disease, is ubiquitous and is found in drinking water, wastewater, and reclaimed wastewater, is an example of a disease causing organism that has been identified relatively recently (Levins et al., 1994). The significance of the identification of new disease organisms, disease outbreaks, and the reemergence of old diseases is that the concern for public health must remain the primary objective of wastewater management.

2–10 TOXICITY

Wastewater can contain a variety of constituents, many of which can cause adverse impacts if discharged to the environment. *Toxicity* is a measure of the degree to which single or multiple constituents that may be present in untreated and treated wastewater can cause adverse impacts (damage) to human and animal health, sensitive aquatic biota, and ecosystems. To provide a general introduction to the subject of toxicity and toxicity testing, it will be useful to consider the following topics: (1) sources of toxicity in untreated and treated wastewater, (2) the evolution and application of toxicity testing, (3) toxicity testing procedures, (4) the analysis of toxicity test results, (5) the application of toxicity test results, and (6) methods that can be used to identify specific toxicity constituents.

Sources of Toxicity

The sources of toxicity in untreated and treated wastewater are derived from the constituents added during usage, treatment, and disinfection with chemical agents.

Constituents Added During Usage. Where a separate wastewater collection system is used, constituents added during usage can include (1) physical properties such as elevated temperature and TDS; (2) inorganic nonmetallic constituents such as ammonia and hydrogen sulfide; (3) metallic constituents such as chromium, mercury, and silver; (4) aggregate organic constituents such as cleaning and personal care products; and (5) individual organic compounds such as identified in Table 2–16. Where a combined wastewater collection system is used, an additional source of toxicity is from runoff. Constituents in runoff that can cause toxicity include pesticides and nutrients from yards, city landscaping, and agricultural lands, and heavy metals and organic and inorganic (e.g., salt) constituents from streets and highways.

Constituents Added During Treatment. Constituents added during treatment that can result in toxicity issues can include flocculent aids that contain contaminants, chemicals to precipitate phosphorus that contain contaminants, chemicals added to control foaming and frothing, and chemicals added to control algae growths.

Constituents Added During Disinfection. One of the most important sources of toxicity is from the disinfection byproducts formed during the disinfection of treated effluent with chemicals such as chlorine, chlorine dioxide, and ozone. The formation and control of chlorine, chlorine dioxide, and ozone disinfection byproducts is considered in Secs. 12–3, 12–4, and 12–6 in Chap. 12, respectively.

Evolution and Application of Toxicity Testing

Until the latter part of the twentieth century, pollution control measures were focused primarily on conventional pollutants (such as oxygen-demanding materials, suspended solids, etc.) which were identified as causing water quality degradation. During the past 30 y, increased attention has been focused on the control of toxic substances, especially those contained in wastewater treatment plant discharges. The national policy prohibiting the discharge of toxic pollutants in toxic amounts is documented in Section 101(a) (3) of the federal Clean Water Act. Because it is not economically feasible to determine the specific toxicity of each of the thousands of potentially toxic substances in complex effluents, whole effluent toxicity testing using aquatic organisms is a direct, cost-effective means of determining effluent toxicity. Whole effluent toxicity testing involves the introduction of appropriate bioassay organisms into test aquariums (see Fig. 2–37) containing various concentrations of the effluent in question and observing their responses. The whole effluent test procedure is used to determine the aggregate toxicity of unaltered effluent discharged into receiving waters. Toxicity is the only parameter measured.

Even though the focus of this section is on effluent toxicity, it should be noted that toxicity testing has number of other applications, including

1. Assess the suitability of environmental conditions for aquatic life.
2. Establish acceptable receiving water concentrations for conventional parameters (such as dissolved oxygen, pH, temperature, salinity, or turbidity).
3. Study the effects of water quality parameters on wastewater toxicity.
4. Assess the toxicity of wastewater to one or more freshwater, estuarine, or marine test organisms.
5. Establish relative sensitivity of a group of standard aquatic organisms to effluent as well as standard toxicants.
6. Assess the degree of wastewater treatment needed to meet water quality requirements.

Figure 2–37

Typical setup used to conduct of whole effluent toxicity tests where mortality is the test end point.

7. Determine the effectiveness of wastewater treatment methods.
8. Establish permissible effluent discharge rates.
9. Determine compliance with federal and state water quality standards and water quality criteria associated with NPDES permits (Standard Methods, 2012).

Such tests provide results that are useful in protecting human health, aquatic biota, and the environment from impacts caused by the release of constituents found in wastewater into surface waters. Toxicity identification, in which the constituents or compounds responsible for the observed toxicity are delineated, is another important aspect of toxicity assessment.

Toxicity Testing

Terms commonly encountered when considering the conduct of toxicity tests and the analysis, interpretation, and application of test results are summarized in Table 2–29. Because the terms reported in Table 2–29 are subject to change as new and improved methods of toxicity testing are developed, it is imperative that the latest version of Standard Methods and related U.S. EPA protocols be reviewed before undertaking any toxicity testing.

Toxicity tests are classified according to (1) duration (short-term, intermediate-term, or long-term); (2) method of adding test solutions (static, recirculation, renewal, or flow-through); (3) type of test (*in vitro,* tests in petri dishes or test tubes, or *in vivo*, toxicity tests using the whole organism); and (4) purpose (NPDES permit requirements, mixing zone determinations, etc). In vitro toxicity testing has been validated widely in recent years.

Even though organisms vary in sensitivity to effluent toxicity, the U.S. EPA has documented that (1) toxicity of effluents correlate well with toxicity measurements in the receiving waters when effluent dilution was measured, and that (2) predictions of impacts from both effluent and receiving water toxicity tests compare favorably with ecological community responses in the receiving waters. The U.S. EPA has conducted nationwide tests with freshwater, estuarine and marine ecosystems. Methods include both acute as

Table 2–29

Terms used in evaluating the effects of contaminants on living organisms[a, b]

Term	Description
Acute toxicity	Exposure that will result in significant response shortly after exposure (typically a response is observed within 48 or 96 h)
Chronic toxicity	Exposure that will result in sublethal response over a long-term, often 1/10 of the life span or more
Chronic value (ChV)	Geometric mean of the NOEC and LOEC from partial and full cycle tests and early-life-stages tests
Cumulative toxicity	Effects on an organism caused by successive exposures
Dose	Amount of a constituent that enters the test organism
Effective concentration (EC)	Constituent concentration estimated to cause a specified effect in a specified time period (e.g., 96-h EC_{50})
Exposure time	Time period during which a test organism is exposed to a test constituent
Inhibiting concentration (IC)	Constituent concentration estimated to cause a specified percentage inhibition or impairment in a qualitative function
In vitro	Tests conducted in glass petri dishes or test tubes
In vivo (in life)	Toxicity tests conducted using the whole organism
Lethal concentration (LC)	Constituent concentration estimated to produce death in a specified number of test organisms in a specified time period (e.g., 96-h LC_{50})
Lowest-observed-effect concentration (LOEC)	Lowest constituent concentration in which the measured values are statistically different than the control
Maximum allowable toxicant concentration (MATC)	Constituent concentration that may be present in receiving water without causing significant harm to productivity or other uses
Median tolerance limit (TLm)	An older term used to denote the constituent concentration at which at least 50 percent of the test organisms survive for a specified period of time. Use of the term "median tolerance limit" has been superseded by the terms median lethal concentration (LC_{50}) and median effective concentration (EC_{50})
No observed-effect concentration (NOEC)	Highest constituent concentration at which the measured effects are no different from the control
Sublethal toxicity	Exposure that will damage organism, but not cause death
Toxicity	Potential for a test constituent to cause adverse effects on living organisms
Whole effluent toxicity (WET)	The total (or aggregate) toxicity effect of treated effluent measured directly in a toxicity test

[a] Adapted from Hughes (1996) and Standard Methods (2012).

[b] It should be noted that the terms given in this table apply only to aquatic organisms and are, for the most part, distinct from the terms used for animals and humans.

Table 2–30

Typical examples of short-term chronic toxicity test methods using various freshwater and marine/estuarine aquatic species[a]

Species/common name	Test duration	Test endpoints
Freshwater species		
Cladoceran *Ceriodaphnia dubia*	Approximately 7 d (until 60 percent of control have 3 broods)	Survival, reproduction
Fathead minnow *Pimephales promelas*	7 d 9 d	Larval growth, survival Embryo-larval survival, percent hatch, percent abnormality
Freshwater algae *Selenastrum capricomutum*	4 d	Growth
Marine/estuarine species		
Sea urchin *Arbacia punctulata*	1.5 h	Fertilization
Red macroalgae *Champia parvula*	7–9 d	Cystocarp production (fertilization)
Mysid *Mysidopsis bahia*	7 d	Growth, survival, fecundity
Sheepshead minnow *Caprinodon variegatus*	7 d 7–9 d	Larval growth, survival Embryo-larval survival, percent hatch, percent abnormality
Inland silverside *Menidia beryijina*	7 d	Larval growth, survival

[a] Adapted from U.S. EPA (1988, 1989).

well as chronic exposures. Typical short-term chronic toxicity test methods are reported in Table 2–30. Detailed contemporary testing and analysis protocols are summarized in Standard Methods (2012) and in U.S. EPA publications (U.S. EPA, 1985b, c, d, e).

Analysis of Toxicity Test Results

Methods used to analyze both short-term (acute) and long-term (chronic) toxicity data are considered in the following discussion.

Acute Toxicity Data. The median lethal concentration (LC_{50}) when mortality is the test end point, or median effective concentration (EC_{50}) when a sublethal effect (e.g., immobilization, fatigue in swimming, "avoidance") is the end point, is typically used to define acute toxicity (Stephen, 1982). A typical bioassay setup using fish where mortality is the test endpoint is shown on Fig. 2–37. A fish swimming chamber is used to assess sublethal effects. A fish is placed in a chamber where the flow-through velocity can be increased until the fish is swept out of the chamber. The washout velocity for fish exposed to a specific compound can be compared to the washout velocity for the control fish.

Because the LC_{50} value is the median value, it is important to provide some information on the variability of the test population. The LC_{50} values can be determined graphically or analytically using the Spearman Karber, moving average, binomial, and probit methods. The 95 percent confidence limits are usually specified. Most standard statistical packages available for desktop computers include a probit analysis program. Determination of LC_{50} values, both graphically and by means of probit analysis, is illustrated in Example 2–13. Typically, LC_{50} values are computed based on survival at both 48-and 96-hour exposures.

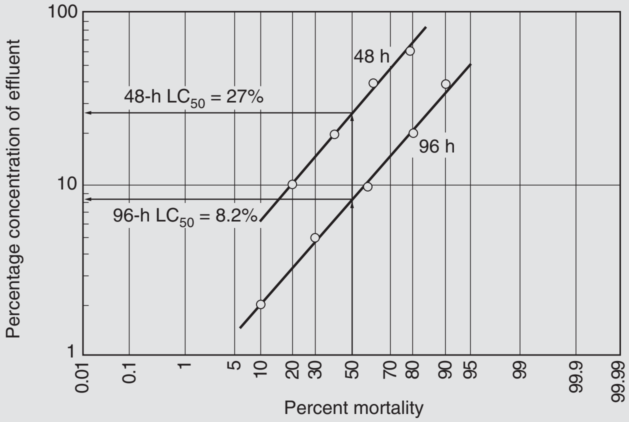
EXAMPLE 2–13 **Analysis of Acute Toxicity Data** Determine graphically and by probit analysis the 48-and 96-h LC_{50} values in percent by volume for the following toxicity test data obtained using flathead minnows.

Concentration of waste, % by volume	No. of test animals	No of test animals dead after[a] 48 h	96 h
60	20	16 (80)	20 (100)
40	20	12 (60)	18 (90)
20	20	8 (40)	16 (80)
10	20	4(20)	12(60)
5	20	0 (0)	6 (30)
2	20	0(0)	2 (10)

[a] Percentage values are given in parentheses.

Solution

1. Plot the concentration of wastewater in percent by volume (log scale) against the test animals that have died in percent (probability scale). The required plot is given below.



2. Fit a line to the data points by eye giving most consideration to the points lying between 16 and 84 percent mortality, which corresponds to approximately one standard deviation.
3. Find the wastewater concentration causing 50 percent mortality. The estimated LC_{50} values are
   a. 48-h $LC_{50} = 27.0\%$
   b. 96-h $LC_{50} = 8.2\%$

4. Compare the results obtained with a probit analysis to the values determined in Step 3. The probit analysis is conducted using a standard statistical analysis. The results of the probit analysis are as follows:
   a. 48-h $LC_{50} = 27.6\%$, 95% confidence limits 21.0 and 37.8%
   b. 96-h $LC_{50} = 8.1\%$, 95% confidence limits 5.8 and 10.9%

Comment Although the LC_{50} values obtained using the graphical analysis approach are approximate, they are quite close to the values obtained using the probit analysis approach and serve as a good check. To obtain confidence limits, a probit or similar analysis must be performed.

Chronic Toxicity Data. Results of chronic toxicity tests often are analyzed statistically to determine the lowest observed effect concentration (LOEC), the no observed effect concentration (NOEC), and the chronic value (ChV). Generally statistical significance is assumed to be at the $p = 0.05$ level. The chronic value (ChV) is calculated as the geometric mean of the LOEC and the NOEC.

Chronic toxicity limits may be specified with either NOEC or ChV as the end point. The term maximum acceptable toxicant concentration (MATC) often is used interchangeably with the chronic value. Similar to acute toxicity data, lethal concentration (LC) or effective concentration (EC) values can be used with chronic toxicity data to describe chronic toxicity tolerance levels. Recently, the concept of the inhibiting concentration (IC) has been introduced to characterize effects in chronic tests. A variety of nonparametric and parametric statistical methods are available to determine NOECs and LOECs and LCs, ECs, and ICs (Standard Methods, 2012).

Application of Toxicity Test Results

In applying acute and chronic toxicity test results, the toxic units (TU) approach has been adopted by a number of federal and state agencies. In the toxic units approach (U.S. EPA, 1985b), a TU concentration is established for the protection of aquatic life.

Toxic Unit Acute (TU_a). TU_a is defined as the reciprocal of the wastewater concentration that caused the acute effect by the end of the exposure period.

$$TU_a = 100/LC_{50} \tag{2–67}$$

Toxic Unit Chronic (TU_c). The TU_c is defined as the reciprocal of the effluent concentration at which the measured effects, by the end of the chronic exposure period, are no different from the control.

$$TU_c = 100/NOEC \tag{2–68}$$

where NOEC = no observed effect concentration.

Depending on the use to be made of the toxicity test results, a variety of different numerical values have been used for TU_a and TU_c as a basis for assessing the suitability of a given effluent for discharge to the environment. For example, to protect against acute toxicity it has been suggested that the MATC should be less than 0.3 × TU_c. Because the limiting values vary from location to location, current regulatory standards must reviewed in applying toxicity results. The application of toxicity test results is illustrated in Example 2–14.

EXAMPLE 2–14 **Application of Toxicity Test Results** A critical initial dilution of 100:1 is achieved for a treated effluent discharged to marine receiving waters. Toxicity tests were conducted with the wastewater treatment plant effluent using three marine species. Based on the toxicity test results, it was found that *Champia parvula* was the most sensitive species acute end point (2.59 percent effluent) as measured by the EC_{50}, and also the most sensitive species chronic end point (1.0 percent) as measured by the NOEC. For protection of the aquatic environment, the acute and chronic toxicity requirements have been set at 10 TU_a and 1.0 Tu_c, respectively.

Results of Acute Toxicity Tests

Species	Control exposure, h	Survival, %	Percent effluent	
			LC_{50} or EC_{50}[a]	NOael
Mysidopis bahia	96	100	18.66	10.00
Cyprinodon variegatus	96	100	>100	50.00
Champia parvula	48/168	100	2.59	12.25

[a] EC_{50} results based on reduction of cystocarp production.

Results of Chronic Toxicity Tests

Species	Control exposure, d	Survival, %	Percent effluent	
			NOEC	LOEC
Mysidopis bahia	7	82.0	6.0	10.00
Cyprinodon variegatus	7	98.8	15.0	15.00>
Champia parvula	7	100.0	1.0	2.25

Solution

1. Check compliance with acute toxicity requirements.
   a. Based on data for the most sensitive species tested, the number of acute toxic units (TU_a), based on Eq. (2–67), is

   $$TU_a = 100/LC_{50} = 100/2.59 = 38.6$$

   b. Following an initial dilution of 100, the TU_a value is

   $$TU_a /100 = 38.6/100 = 3.86\ TU_{ad} \text{ (after dilution)}$$

   Because the TU_a value after dilution (3.86 TU_{ad}) is less than 10 TU_a the acute toxicity requirement has been met.

2. Check compliance chronic toxicity requirements.
   a. Based on data for the most sensitive species tested, the number of chronic toxic units (TU_c), based on Eq. (2–68), is

      $TU_c = 100/\text{NOEC} = 100/1.0 = 100$

   b. Following an initial dilution of 100, the TU_c value is

      $TU_c = 100/100 = 1.0\ Tu_{cd}$ (after dilution)

      Because the TU_c value after dilution (1.0 TU_{cd}) is equal to 1.0 TU_c, the chronic toxicity requirement has been met.

In summary, there are a number of advantages to the use of whole effluent toxicity testing. In this approach, the bioavailability of the toxics is measured, and the effects of any synergistic interactions are also considered. Because the aggregate toxicity of all components of the wastewater effluent is determined, the toxic effect can be limited by limiting only one parameter, the effluent toxicity. Because contemporary receiving water management strategies are based on site-specific water quality criteria, toxicity testing facilitates comparison of effluent toxicity with site-specific water quality criteria designed to protect representative, sensitive species and yet allow for establishment of discharge limitations that will protect aquatic environments.

Identification of Toxicity Components

Toxicity testing can also be used to determine the source of toxicity, an especially important test for industrial discharges. For example, is the toxicity caused by the suspended solids, the colloidal solids, the long- or short-chain dissolved organic constituents, or the dissolved inorganic constituents? To determine the source(s) of the toxicity, each of the potential sources must be isolated from the other constituents in the sample and tested for toxicity. The various methods that can be used to fractionate a wastewater sample are illustrated graphically on Fig. 2–38 and described in Table 2–31.

Figure 2–38

Separation techniques that can be used to fractionate a wastewater sample. (Adapted from Eckenfelder, 1999.)

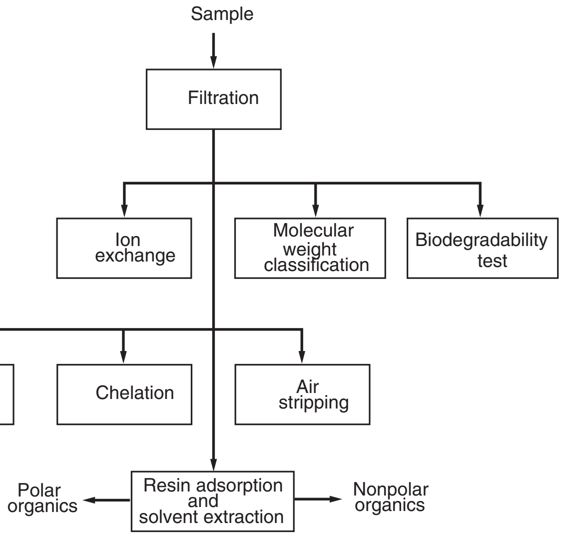


Table 2–31

Description of separation techniques that can be used to fractionate a wastewater sample[a]

Separation technique	Description
Filtration for suspended solids	Filtration is generally performed first to determine whether the toxicity is related to the soluble or insoluble phase of the sample. Typically, 1 μm glass fiber filters that have been prewashed with ultrapure water are used. The insoluble phase should be resuspended in control water to insure that filtration, not adsorption on the filter medium, removed the toxicity.
Filtration for colloidal solids	A 0.1-μm filter should be used to determine if the colloidal fraction is responsible for the toxicity.
Ion exchange	Inorganic toxicity can be studied by using cationic and anionic exchange resins to remove potentially toxic inorganic compounds or ions.
Molecular weight classification	Evaluating the molecular weight distribution of the influent, and the toxicity of each molecular weight range, can often narrow the list of suspected contaminants.
Biodegradability test	Controlled biological treatment of effluent samples in the lab can result in almost compete oxidation of the biodegradable portion of organics. Bioassay analysis can then quantify the toxicity associated with the nonbiodegradable components, as well as the reduction in toxicity attainable by biological treatment.
Oxidant reduction	Residual chemical oxidants carried over from a process (e.g., chlorine and chloramines used for disinfection, or ozone and hydrogen peroxide used in sludge conditioning) can be toxic to most organisms. A simple batch reduction of these oxidants at various concentrations, using an agent such as sodium thiosulfate, can be used to assess the toxicity of any remaining oxidants.
Metal chelation	The toxicity of the sum of all cationic metals (with the exception of mercury) can be determined by chelation of samples, using varying concentrations of ethylenediaminetetraacetic acid (EDTA) and evaluating the change in toxicity.
Air stripping	Batch air stripping at acid, neutral, and basic pH can remove essentially all volatile organics. At a basic pH, ammonia is removed as well. Thus, if both volatile organics and ammonia are suspected toxicants, an alternative ammonia removal technique, such as a zeolite exchange, should be used. (Note that ammonia is toxic in the nonionized form, so ammonia toxicity is very pH dependent.)
Resin adsorption and solvent extraction	Specific nonpolar organics can sometimes be identified as toxics, using a resin-adsorption/solvent-extraction process. A sample is adsorbed on a long-chain organic resin, the organics are reextracted from the resin with a solvent (e.g., methanol), and the toxicity of the sample is determined using a bioassay test procedure.

[a] Adapter from Eckenfelder (2000).

PROBLEMS AND DISCUSSION TOPICS

2–1 The following test results were obtained for three treated wastewater effluent samples. Check the accuracy of the analytical measurements for one of the waters (to be selected by instructor). Do you suspect that a constituent of significance has been neglected? If so, is it a cation or anion?

	Concentration, mg/L				Concentration, mg/L		
	Wastewater sample no.				Wastewater sample no.		
Cation	1	2	3	Anion	1	2	3
Ca^{2+}	121.3	76.0	190.2	HCO_3^-	280.0	128.2	260.0
Mg^{2+}	36.2	27.2	84.1	SO_4^{2-}	116.0	240.0	64.0
Na^+	8.1	22.9	75.2	Cl^-	61.0	37.2	440.4
K^+	12.0	18.7	5.1	NO_3^-	15.6	2.0	35.1
Fe^{2+}	—	2.1	0.2	CO_3^{2-}	—	—	30.0

2–2 Determine the mole fraction of Ca^{2+}, Mg^{2+}, and SO_4^{2-} for one of the following treated wastewater effluents (to be selected by instructor).

	Concentration, mg/L				Concentration, mg/L		
	Wastewater sample no.				Wastewater sample no.		
Cation	1	2	3	Anion	1	2	3
Ca^{2+}	206.6	161.4	226.1	HCO_3^-	525.4	438.7	476.6
Mg^{2+}	95.3	47.5	62.1	SO_4^{2-}	219.0	153.2	483.2
Na^+	82.3	71.4	46.2	Cl^-	303.8	163.8	20.6
K^+	5.9	2.2	3.5	NO_3^-	19.2	8.1	9.3
Fe^{2+}			3.1	CO_3^{2-}			

2–3 Determine the ionic strength and activity coefficients and activity for the constituents for one of the wastewater effluents in Problem 2–1 or 2–2 (to be selected by instructor).

2–4 Estimate the TDS for one of the water samples in Problem 2–1 or 2–2 (to be selected by instructor) using Eq. (2–11) and by summing the mass of the individual ionic species. How do the computed values compare?

2–5 Using the following formula, modified appropriately, determine the concentration of total and volatile solids, expressed as mg/L, for one of the wastewater samples (to be selected by instructor).

$$\mathrm{TS} = \frac{\left[\begin{pmatrix}\text{mass of evaporating}\\ \text{dish plus residue, g}\end{pmatrix} - \begin{pmatrix}\text{mass of evaporating}\\ \text{dish, g}\end{pmatrix}\right]\left(\dfrac{10^3\ \text{mg}}{\text{g}}\right)}{\text{sample size, L}}$$

		Weight, g			
		Sample number			
Item	Unit	1	2	3	4
Sample size	mL	90	100	120	200
Tare mass of evaporating dish	g	22.6435	22.6445	22.6550	22.6445
Mass of evaporating dish plus residue after evaporation at 105°C	g	22.6783	22.6832	22.6995	22.6667
Mass of evaporating dish plus residue after ignition at 550°C	g	22.6768	22.6795	22.6832	22.6433

2–6 The following test results were obtained for a wastewater sample taken at the headworks to a wastewater treatment plant. All of the tests were performed using a sample size of 50 mL. Determine the concentration of total solids, total volatile solids, total suspended solids, volatile suspended solids, and dissolved solids for one of the samples (to be selected by instructor) (refer to the formula given in Problem 2–5).

	Weight, g			
	Sample number			
Item	1	2	3	4
Tare mass of evaporating dish	53.5435	53.5434	53.5436	53.5433
Mass of evaporating dish plus residue after evaporation at 105°C	53.5765	53.5693	53.5725	53.5793
Mass of evaporating dish plus residue after ignition at 550°C	53.5515	53.5489	53.5495	53.5523
Tare mass of Whatman GF/C filter	1.5433	1.5435	1.5436	1.5434
Mass of Whatman GF/C filter plus residue after drying at 105°C	1.5533	1.5521	1.5635	1.5541
Mass of Whatman GF/C filter plus residue after ignition at 550°C	1.5457	1.5455	1.5456	1.5457

2–7 The following data were obtained from a serial filtration test of a settled effluent after biological treatment. Prepare a plot of one of the samples (to be selected by instructor). How great an error in the measurement of the total suspended and colloidal solids would occur if a filter paper with a nominal pore size of 1.2 μm was used to determine the total suspended solids, as compared to using a filter with a nominal pore size of 0.1 μm filter?

Nominal pore size, μm	Suspended solids, mg/L			
	Sample number			
	1	2	3	4
12	20.2	29.4	22.5	25.1
8	8.8	11.5	8.0	15.1
5	4.1	3.5	4.9	2.2
3	7.5	5.1	11.6	8.9
1	15.1	13.5	21.2	25.0
0.1	9.9	15.1	24.9	17.5

2–8 Determine the coefficients A and β in Eq. (2–16) for the following particle size data obtained using a particle counter with arithmetic channel settings.

	Number of particles			
	Sample number			
Channel size, μm	1	2	3	4
1–2	27,000	3980	25,119	1000
2–5	9029	1690	4979	599
5–10	4050	450	561	199
10–15	1418	100	123	100
15–20	405	60	45	45
20–30	203	40	26	40
30–40	36	20	8	20
40–60	16	9		20
60–80	8	5		10
80–100	5	3		6
100–140	4	2		

2–9 If the average intensity of the UV irradiation to which a sample was exposed is 5, 10, 15 mW/cm^2 (to be selected by instructor), determine the UV intensity measured at the water surface in a Petri dish. The depth of water in the Petri dish is 8, 10 or 12 mm (to be selected by instructor). Assume the absorptivity, $k(\lambda = 254\ \text{nm})$, is equal to 1.25 cm^{-1}.

2–10 Estimate the alkalinity, expressed as mg/L as $CaCO_3$, for one of the water samples in Problem 2–1 (to be selected by instructor).

2–11 What is the molar mass of a gas at 20°C assuming the gas has a density of 0.68 g/L at standard temperature and pressure (STP)?

2–12 At what pH would 95 percent of the ammonia be present as gas? Apply Eq. (2–40) and give the acid ionization (dissociation) constant at 25°C?

2–13 Compare the saturation concentrations of O_2, N_2, and CO_2 between San Francisco (sea level) and Taos, NM (elevation 2150 m), Denver, CO (elevation 1600 m) or La Paz, Bolivia (elevation 4270 m) (city to be selected by instructor).

2–14 Using Henry's law, determine the saturation concentration of O_2, N_2, or CO_2 (gas to be selected by instructor) in water at 0, 10, 20, 30, 40, and 50°C. Construct a plot of O_2, N_2, or CO_2 saturation concentration as a function of temperature.

2–15 Determine the concentration of oxygen in the liquid phase of a covered pure-oxygen-activated sludge plant subject to 2, 2.5, or 3 atm of pressure (pressure to be selected by instructor). For the purposes of this problem, neglect any reactions occurring in the system. The composition of the gas in the head space above the wastewater is 80 percent oxygen, 15 percent nitrogen, and 5 percent carbon dioxide by volume.

2–16 Bottled soda water is produced by increasing the gas pressure and carbon dioxide content. For a bottle having a total gas pressure of 2 atm, determine the pH of the soda water at 25°C. The composition of the gas in the head space above the soda water is 95 percent carbon dioxide by volume.

2–17 Intensity measurements were made at different dilutions for three odor samples. Using the data provided, determine which of the three odors is more persistent.

n-butanol odor intensity, ppm$_v$	Dilution-to-threshold, D/T Sample 1	 2	 3
10,000	0	0.0	0
100	25	3.2	11
10	316	10.0	56
0	3160	32.0	265

2–18 A 175, 200, 225 mL volume (volume to be selected by instructor) of distilled odor-free water was required to reduce the odor of a 25 mL sample of treated wastewater to a level that is just perceptible. What is the threshold odor number (TON)? What is the corresponding value expressed in D/T (dilutions-to–threshold)? Assuming the odorous compound is hydrogen sulfide, find the concentration of hydrogen sulfide in the original sample of treated wastewater. Assume the sample is at equilibrium with the atmosphere. Assume the data in Table 2–8 apply.

2–19 The local Air Pollution Control District has threatened to fine and penalize the local wastewater management agency, your client, because of frequently recurring odor complaints from residents who live downwind of the plant. The plant manager, a full-time employee at the treatment plant, claims that no problem exists. He proves his point by consistently finding less than 5 dilutions to MDTOC at the plant boundary using a hand-held sniff dilution olfactometer as employed by the local Air Pollution Control District. You, however, live downwind of the plant and have frequently detected odors from it. Why do these differences exist? How would you resolve them objectively?

2–20 You have been asked to review an odor-control system that has apparently failed to control odors from a sludge-dewatering building adequately. The wastewater management agency, your client, claims the system has failed to perform according to specifications. The engineering contractor who installed the system claims that the specifications were not adequate.

In your investigation you find that the agency employed a reputable odor consultant to develop the odor-control-system specifications. The consultant used the ASTM Panel Method for odor measurement, using evacuated glass cylinders for sample collection. Several measurements were made, and the maximum observed value was doubled to develop the control-system specifications. In this way a 90 percent odor-removal requirement was established to meet the desired final odor-emission limit of 2.8×10^4 odor units per minute (the product of airflow in m^3/min and number of dilutions to MDTOC).

Using a direct-reading olfactometer, you find that the control system removes 99 percent of the odor, and that at a rate of 10^6 odor units per minute the final odor emission is 10^6 odor units per minute. What reasons might explain your findings? How would you resolve the problem?

2–21 In a BOD determination, 6 mL of wastewater are mixed with 294 mL (total volume of BOD bottle is 300 mL) of diluting water containing 9.1 mg/L of dissolved oxygen. After a 5-d incubation at 20°C, the dissolved oxygen content of the mixture is 2.8 mg/L. Calculate the BOD of the wastewater. Assume that the initial dissolved oxygen concentration in the wastewater is zero and that the following equation applies:

$$\text{BOD, mg/L} = \frac{D_1 - D_2}{P}$$

where D_1 = dissolved oxygen of diluted sample immediately after preparation, mg/L
D_2 = dissolved oxygen of diluted sample after 5-d incubation at 20°C, mg/L
P = fraction of wastewater sample volume to total combined volume

2–22 Solve Problem 2–21 for one of the following conditions (to be selected by instructor). In all cases, the total volume of the BOD bottle is 300 mL.

a. sample size = 8 mL, oxygen in dilution water = 9.0 mg/L, oxygen in mixture after 7-d incubation at 20°C = 1.8 mg/L
b. sample size = 6 mL, oxygen in dilution water = 9.2 mg/L, oxygen in mixture after 6-d incubation at 20°C = 1.65 mg/L
c. sample size = 6 mL, oxygen in dilution water = 8.9 mg/L, oxygen in mixture after 4-d incubation at 20°C = 1.5 mg/L
d. sample size = 10 mL, oxygen in dilution water = 9.15 mg/L, oxygen in mixture after 5-d incubation at 20°C = 1.42 mg/L

2–23 Determine the UBOD and BOD_5 (in mg/L) of a mixture of 150 mg/L glucose ($C_6H_{12}O_6$) and 150 mg/L glutamic acid ($C_5H_{10}N_2O_3$). Assume the value of the BOD_5 first-order reaction rate constant is $0.23\ d^{-1}$ (base e), at 20°C.

2–24 If the 5-d 20°C BOD of a wastewater is 185, 200, or 220 mg/L (to be selected by instructor), what will be the ultimate BOD (UBOD)? What will be the 10-d demand? If the bottle had been incubated at 15°C and the first order reaction rate constant, k_1, is $0.23\ d^{-1}$ (base e), what would the 5-d BOD have been?

2–25 If the BOD value at 20°C of a wastewater measured at 2 and 8 d, 1 and 9 d, 2 and 7 d, and 3 and 10 d (day to be selected by instructor) was found to be 125 and 225 mg/L, respectively, determine the 5-d value using the first-order rate model.

2–26 The following results for a wastewater sample were determined at 20°C. For a sample (to be selected by instructor) determine the ultimate carbonaceous oxygen demand, the ultimate nitrogenous oxygen demand (NOD), the carbonaceous BOD reaction-rate constant (k), and the nitrogenous NOD reaction-rate constant (k_n). Determine $k(\theta = 1.05)$ and $k_n(\theta = 1.08)$ at 25°C.

	BOD, mg/L			
	Sample number			
Time, d	1	2	3	4
0	0	0	0	0
2	18	30	45	36
4	26	43	75	58
6	30	52	95	70
8	33	58	114	80
10	56	60	135	90
12	69	90	144	98
14	77	104	149	102
16	82	114	151	145
18	84	120	152	170
20	87	125	152	182
25	90	135	170	210
30	91.5	142	239	222
35	92.5	147	260	233
40	93	148	268	239
45	94	149	271	240
50	94.5	150	272	241

2–27 If the observed first-order carbonaceous BOD reaction rate coefficients are 0.23 and 0.28 d^{-1} at 20 and 30°C, 0.22 and 0.15 d^{-1} at 20 and 12°C, or 0.15 and 0.30 d^{-1} at 10 and 20°C (to be selected by instructor), estimate the activation energy for the reaction. Hint: see Example 1–1.

2–28 Compute the carbonaceous and nitrogenous oxygen demand of a wastewater represented by the formula $C_6N_2H_4O$, $C_9N_2H_6O_2$, $C_{10}N_2H_8O_2$ or $C_{12}N_4H_6O_2$, (formula to be selected by instructor). Assume N is converted to NH_3 in the first step.

2–29 The following data have been obtained for five different wastewater samples. Estimate the total quantity of oxygen in mg/L that must be furnished to stabilize completely one of the wastewaters (to be selected by instructor). What are the corresponding COD and the ThOD for the wastewater?

		Wastewater sample				
Item	Unit	1	2	3	4	5
BOD	mg/L	400	375	225	185	325
k (base *e*)	d^{-1}	0.29	0.23	0.027	0.025	0.023
NH_3	mg/L	80	65	75	67	83

2–30 An industrial wastewater is known to contain only glycine ($C_2H_5O_2N$), glucose ($C_6H_{12}O_6$), and stearic acid ($C_{18}H_{36}O_2$). The results of a laboratory analysis for four different samples are as follows. For one of these samples (to be selected by instructor), determine the concentration of each of the three constituents in mg/L. Base your answer on 10^5 L so that the mole numbers are greater than one.

a. organic nitrogen = 11 mg/L, organic carbon = 130 mg/L, and COD = 425 mg/L. Determine the concentration of each of the three constituents in mg/L.

b. organic nitrogen = 13 mg/L, organic carbon = 109 mg/L, and COD = 440 mg/L. Determine the concentration of each of the three constituents in mg/L.

c. organic nitrogen = 9 mg/L, organic carbon = 123 mg/L, and COD = 625 mg/L. Determine the concentration of each of the three constituents in mg/L.

d. organic nitrogen = 12 mg/L, organic carbon = 143 mg/L, and COD = 425 mg/L. Determine the concentration of each of the three constituents in mg/L.

2–31 How many mg/L of $Cr_2O_7^{-2}$ are consumed if the COD of a wastewater sample is found to be 300, 375, 450, or 525 mg/L (value to be selected by instructor)?

2–32 If the chemical composition of food waste is $C_{21.53}H_{34.21}O_{12.66}N_{1.00}S_{0.07}$, estimate the chemical energy content on a COD basis. Express results on the basis of MJ/kg COD computed as HHV on dry weight basis. How does the computed energy content for food waste compare to that of wastewater as determined in Example 2–12? How much energy has been removed from the food waste?

2–33 Bacteria have equivalent diameters of 2×10^{-6} μm and densities of approximately 1 kg/L. Under optimal conditions, bacteria can divide every 30 min. Determine the mass of bacteria that would accumulate in 72 h under continuing optimal growth conditions. Can this occur? Explain.

2–34 If the bacteria found in feces have an average volume of 2.0 μm^3, determine the concentration of suspended solids that would be represented by a bacterial density equal to 10^8 organisms/mL. Assume the density of the bacteria is 1.005 kg/L.

2–35 Using the following joint probability equation, based on the Poisson distribution, derive an expression that can be used to compute the MPN based on a single sample comprised of 5 fermentation tubes. What is the MPN if the sample size is 0.1 mL and 3 of 5 fermentation tubes are positive?

$$y = 1/a[(1 - e^{-n\lambda})^p(e^{-n\lambda})^q]$$

where y = probability of occurrence of a given result
a = constant for a given set of conditions
n = sample size, mL
λ = coliform density, number/mL
p = number of positive tubes
q = number of negative tubes

2–36 Seven effluent samples have been analyzed for total coliform using the standard confirmed test. Using the standard MPN tables, determine the coliform density, expressed as MPN, for three of the seven samples (to be selected by instructor). Refer to the most recent edition of Standard Methods for MPN tables.

Size of portion, mL	Positive tubes/total tubes						
	Sample number						
	1	2	3	4	5	6	7
100.0			5/5	5/5	5/5	5/5	5/5
10.0		4/5	4/5	5/5	5/5	5/5	5/5
1.0	4/5	5/5	5/5	5/5	5/5	5/5	5/5
0.1	3/5	3/5	3/5	2/5	1/5	2/5	5/5
0.01	1/5	2/5	2/5	3/5	2/5	2/5	5/5
0.001					0/5	1/5	1/5

2–37 Discuss the advantages and disadvantages of using the fecal coliform and fecal streptococci tests to indicate bacteriological contamination. Cite a minimum of three references from the literature.

2–38 A coliphage virus test using MS-2 was performed to determine the titer of the undiluted sample. During the test, one milliliter of each dilution was added to each plate. The results of the plaque counting are shown in the table given below. Based on the results, calculate the titer of the culture in the undiluted sample.

	Count/plate				
	Sample				
Dilution	Original	Dup. 1	Dup. 2	Dup. 3	Dup. 4
10^{-7}	FP[a]	FP	FP	FP	FP
10^{-8}	TNTC[b]	TNTC	TNTC	TNTC	TNTC
10^{-9}	120	110	116	TNTC	123
10^{-10}	60	51	38	43	56
10^{-11}	1	2	0	1	1

[a] FP = full plaque on plate.
[b] TNTC = too numerous to count.

2–39 Using the following toxicity test data, determine the 48 and 96 h LC_{50} values in percent by volume.

Concentration of waste, % by volume	Number of test animals	Number of test animals dead after	
		48 h	96 h
80	20	17	20
60	20	13	20
40	20	10	15
20	20	6	13
10	20	3	9
5	20	1	4
2	20	0	2

2–40 Using the following toxicity test data, determine the 48 and 96 h LC_{50} values in percent by volume.

Concentration of waste, % by volume	Number of test animals	Number of test animals surviving after		
		24 h	48 h	96 h
12	20	0	2	8
10	20	0	5	10
8	20	1	8	13
6	20	3	11	16
4	20	6	16	20
2	20	14	20	20

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3

Wastewater Flowrates and Constituent Loadings

WORKING TERMINOLOGY

Term	Definition
Combined sewer overflow (CSO)	A hydraulic relief structure within a combined collection system to allow excess wet-weather flow to be discharged ahead of the interceptor, pumping station or wastewater treatment plant.
Combined sewer system	A collection system which conveys both sanitary wastewater at all times and stormwater runoff during wet weather, as well as snowmelt.
Delayed inflow	Stormwater that may require several hours or days or more to drain through the collection system.
Direct inflow	Those types of inflow that have a direct stormwater runoff connection to the sanitary collection system and cause an almost immediate increase in wastewater flowrates.
Domestic wastewater	Wastewater discharged from residences and from commercial, institutional, and similar facilities including infiltration.
Exfiltration	Flow out of a collection system through breaks in the pipe wall, defective pipe joints or connections, or breaks in access port (manhole) walls.
Flow equalization	The dampening of flowrate variations to obtain a constant or nearly constant flowrate, usually by means of a storage (equalization) basin.
Industrial wastewater	Wastewater from nondomestic sources in which industrial wastes predominate.
Infiltration	Water entering a collection system from a variety of entry points including service connections and from the ground through defective pipe joints, connections, or breaks in access port (manhole) walls.
Inflow	Extraneous water that is discharged to the collection system such as from roof leaders, area drains, access port covers, cross connections from storm drains and catch basins, and combined systems.
Instantaneous peak flowrate	Highest recorded flowrate occurring for a period consistent with the recording equipment. In many situations the recorded peak flow may be considerably below the actual peak flow because of metering and recording equipment limitations.
Mass loading rate	The product of flowrate times constituent concentration.
Peaking factor	The ratio of the peak flowrate to the average flowrate.
Sanitary sewer overflow (SSO)	The release of wastewater from the sanitary sewer system caused by backups, clogging or hydraulic overloading.
Sanitary sewer system	A collection system in which primarily domestic wastewater is conveyed.
Steady inflow	Water discharged from cellar and foundation drains, cooling water discharges, and drains from springs and swampy areas.
Stormwater	Runoff resulting from rainfall and snowmelt.
Sustained flowrates	Flowrates that are equalled or exceeded for a specified number of consecutive days based on annual operating data.
Sustained mass loadings	The mass loading rate value sustained or exceeded for a given period of time (e.g., 1 h, 1 d, or 1 mo).

Determining wastewater flowrates and constituent mass loadings is a fundamental step in initiating the conceptual process design for upgrading existing and developing new wastewater treatment facilities. Reliable data for existing and projected flowrates are essential for determining the hydraulic characteristics, sizing, and operational considerations of the treatment system components. Constituent mass loading rate, the product of constituent concentration and flowrate, is necessary to determine the capacity and operational characteristics of the treatment facilities and ancillary equipment to ensure that treatment objectives are met.

Important factors and issues, typical to most planning and design projects, addressed in this chapter include (1) wastewater sources and flowrates, (2) impact of the collection system on wastewater flowrates, (3) analysis of wastewater flowrate data, (4) analysis of wastewater constituents, (5) analysis of constituent concentrations and mass loading rate data, (6) selection of design flowrates and mass loading rates, and (7) flow equalization.

3–1 WASTEWATER SOURCES AND FLOWRATES

The components that make up the wastewater flow from a community, identified previously in Chap. 1, and repeated here for convenience, are

Domestic (also called sanitary) wastewater. Wastewater discharged from residences and from commercial, institutional, and public facilities.

Industrial wastewater. Wastewater in which industrial process wastes predominate.

Infiltration/inflow (I/I). Water that enters the collection system through indirect and direct means. Infiltration is extraneous water that enters the collection system through leaking joints, cracks and breaks, or porous walls. Inflow is stormwater that enters the collection system from storm drain connections (catch basins), roof leaders, foundation and basement drains, or through access port (manhole) covers or breaks in the access port the walls.

Stormwater. Runoff resulting from rainfall and snowmelt.

Data that can be used to estimate average wastewater flowrates from various domestic, commercial, institutional, and industrial sources are presented in this section. The contributions associated with the collection system are considered in the following section.

Municipal Uses of Water

To understand the sources of wastewater it is helpful to review briefly the municipal use of water. Municipal uses of water may be divided into various categories as reported in Table 3–1. Domestic use includes water used indoors in private residences, apartment houses, etc., for drinking, cooking, hand washing, bathing, laundry, toilet flushing, and other uses, and outdoors for landscape irrigation, car washing, and other outdoor purposes.

Commercial and industrial use includes water used by commercial establishments and industries. In small residential communities the commercial and industrial use may be as low as 40 L/capita·d (10 gal/capita·d), but in industrial cities it may as high as 400 L/capita·d (100 gal/capita·d). Public use includes the water required for use in parks, civic buildings, schools, hospitals, churches, street washing, etc. Water that leaks from the system, meter slippage, unauthorized connections, and all other unaccounted-for water is classified as loss and waste. The loss and waste category is often estimated at about 75 L/capita·d (20 gal/capita·d), but with proper construction and careful maintenance it can be reduced to less than 20 L/capita·d (5 gal/capita·d).

Table 3–1

Municipal uses of water and typical quantities in the United States[a]

Use	Flowrate, gal/capita·d		Flowrate, L/capita·d	
	Range	Typical	Range	Typical
Domestic				
Indoor use	40–80	65[b]	150–300	250
Outdoor use	16–90	35[c]	60–340	132
Commercial	10–75	40	40–300	150
Public	15–25	20	60–100	75
Loss and waste	15–25	20	60–100	75
Total	96–255	170	370–990	682

[a] Data developed from numerous sources and the authors experience.

[b] Based on current (mid-2013) level of water conservation.

[c] In some parts of the country, outdoor water use is significantly higher than indoor use, depending on the season of the year.

Domestic Wastewater Sources and Flowrates

The principal sources of domestic wastewater in a community are the residential areas and commercial districts. Other important sources include public and private institutional facilities and public recreational facilities. Knowledge of the wastewater flowrates is of fundamental importance in the design and operation of wastewater treatment systems. Flowrates are measured both within the collection system and wastewater treatment plants.

Flowrate Measurements in the Collection System. For areas served with collection systems, wastewater flowrates are determined commonly from existing records or by direct field measurements. Flowrate measurements can be made within gravity sewers by installing flow meters in access ports. In the past, measuring flumes (e.g., Palmer Bowlus) or weirs were installed with level measuring equipment calibrated to output direct flowrate readings. Although flumes and weirs are still used occasionally, most collection system flowrate measurements are now made using area-velocity meters. Area-velocity devices are used to measure simultaneously the depth of flow and the velocity in the collection system without restricting the flow path. Both ultrasonic and radar based devices are used [see Figs. 3–1(a) and (b)]. Unmetered pump stations in the collection system can also be used for gathering flowrate information by measuring the wet well volume and determining the time between pump starting and stopping.

Flowrate Measurements at Treatment Plants. Flowrate measurements at wastewater treatment plants are made with a variety of flowmeters. In the past, influent flowrate measurements were made with Venturi meters for measurements in force mains and Parshall flumes for open channel measurements. Today, the preference is for magnetic meters for force mains because they have a shorter laying length, no flow constrictions, and minimal headloss [see Figs. 3–1(c), (d) and (e)]. Ultrasonic meters are also used. Parshall flumes are the least complicated and flow can be measured manually in case the metering equipment is out of calibration or inoperative [see Figs. 3–1(f) and (g)]. The only problem with the Parshall flume is that to operate properly it requires a free discharge, and thus, the headloss is higher than other measuring devices.

Flowrate Estimates from Available Data. For new developments or newly sewered areas, wastewater flowrates are derived from an analysis of population data and estimates of per capita wastewater flowrates from similar communities. These records are

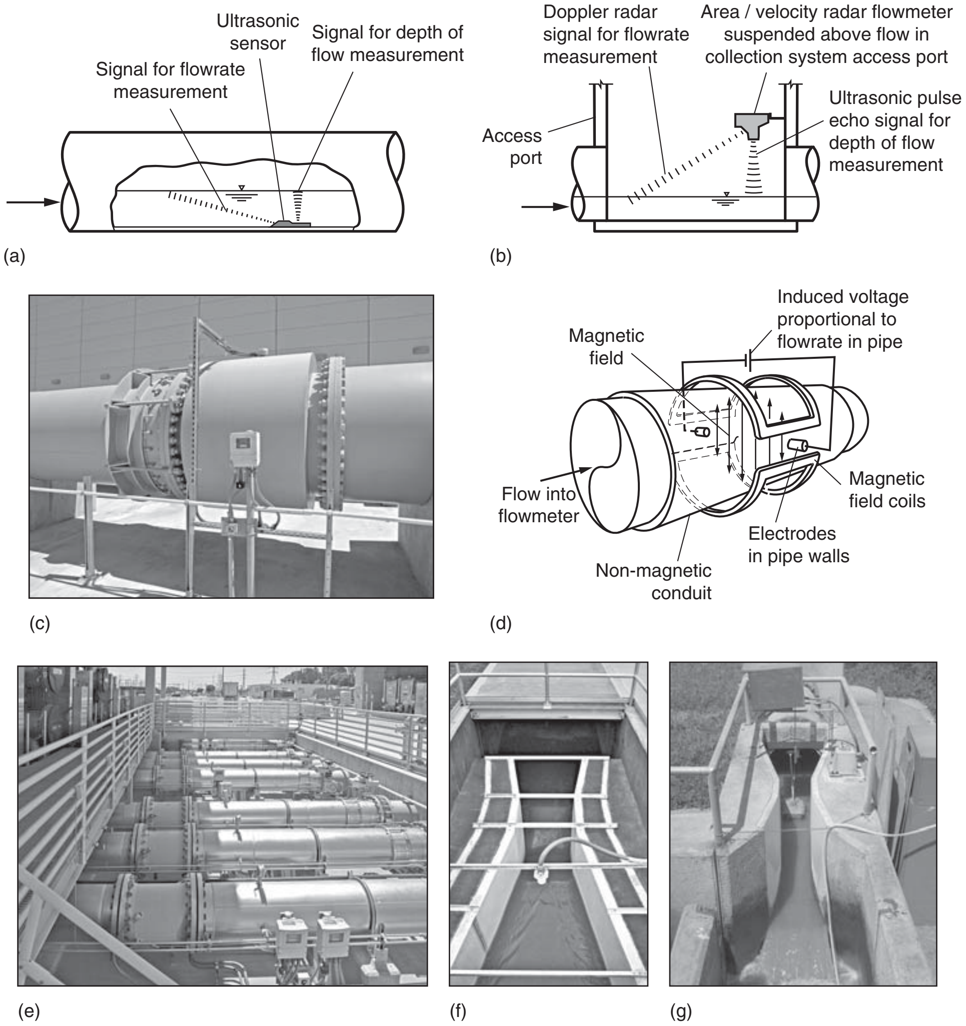


Figure 3–1

Examples of devices used for measurement of wastewater flowrates: (a) schematic of area/velocity ultrasonic flowrate meter sensor, (b) schematic of area/velocity radar flow meter sensor. Velocity is measured using Doppler Radar and ultrasonic echo plus is used to sense depth of flow, (c) view of magnetic flow meter and (d) schematic of magnetic flow meter. When a conductive fluid moves through a magnetic field a voltage is generated; the magnitude of the voltage is proportional to the flowrate, (e) view of multiple magnetic flow meters used to monitor flowrates to advanced oxidation facilities, and (f) Parshall flume equipped with a sonic water level indicator to determine the depth of flow that is correlated to the flowrate and (g) Parshall flume equipped with a float can also be used to determine depth. (Note: Parshall flumes are used commonly at smaller wastewater treatment plants.)

especially useful in other parts of the world where water use for landscape irrigation is limited and 90 percent or more of the water used becomes wastewater. In the United States, on average about 50 to as high as 90 percent (e.g., from high-rise apartments) of the per capita water consumption becomes wastewater. The higher percentages apply to the northern states during cold weather; the lower percentages are applicable to the semi-arid region of southwestern United States where landscape irrigation is used extensively. When water consumption records are used for estimating wastewater flowrates, the amount of water consumed for purposes such as landscape irrigation (that is not discharged to the collection system), leakage from water mains and service pipes, or product water that is used by manufacturing establishments must be evaluated carefully.

Residential Areas. For many residential areas, wastewater flowrates are commonly determined based on population and the average per capita contribution of wastewater. For residential areas where large residential development is planned, it is often advisable to develop flowrates on the basis of land-use areas and anticipated population densities. Where possible, these rates should be based on actual flow data from selected similar communities, preferably in the same locale. In the past, the preparation of population projections for use in estimating wastewater flowrates was often the responsibility of the engineer, but today population projection data are usually available from local, regional, and state planning agencies.

Wastewater flowrates can vary depending on the quantity and quality of the water supply, the potable water billing rate structure, the extent of conservation measures, geographic location, rates of infiltration/inflow, and other economic and social characteristics of the community. In a report published by the U.S. Geological Survey (Kenny et al., 2009), the average domestic water use per capita by state varied from a high of 715 L/capita·d (189 gal/capita·d) to a low of 193 L/capita·d (51 gal/capita·d). The average for the country was 375 L/capita·d (99 gal/capita·d), which is consistent with the data reported in Table 3–1. If it is assumed that 50 to 90 percent of the water supply becomes wastewater, the average wastewater flowrates would vary from 188 to 338 L/capita·d (50 to 89 gal/capita·d). The reason that an average range is given is that no one has an accurate estimate of the extent of water conservation measures that have been implemented, which varies from community to community. The typical average range reflects an estimate of the current (2013) extent to which conservation measures have been implemented in different parts of the country.

Data on the typical flowrate values for residential sources in the United States as a function of the number of residents with the current (2013) level of conservation and with significant water conservation are given in Table 3–2. The data from Table 3–2 are plotted on Fig. 3–2. As shown on Fig. 3–2, as the number of persons per residence increases, the average wastewater flowrate per capita decreases. In the case with significant conservation measures, the per capita value approaches a value of about 150 L/capita·d (39.6 gal/capita·d). At the present time this asymptotic value represents a reasonable estimate of what can be achieved with extensive water conservation measures and fixtures and appliances. Assuming an occupancy rate of 3.3 persons per residence, the range of anticipated wastewater flowrates will vary from about 250 to 175 L/capita·d (66 to 46 gal/capita·d). The value of 250 L/capita·d (66 gal/capita·d) is within the range of typical per capita values reported above. With the passage of time, it is anticipated that the right-hand curve on Fig. 3–2 will move to the left. It is estimated that reaching significant conservation may take more than 20 y. Reduced household water use not only changes the quantity of wastewater generated, but, as discussed later in this chapter, the characteristics of wastewater as well.

Commercial Districts. Depending on the function and activity, unit flowrates for commercial facilities can vary widely. Because of the wide variations that have been observed,

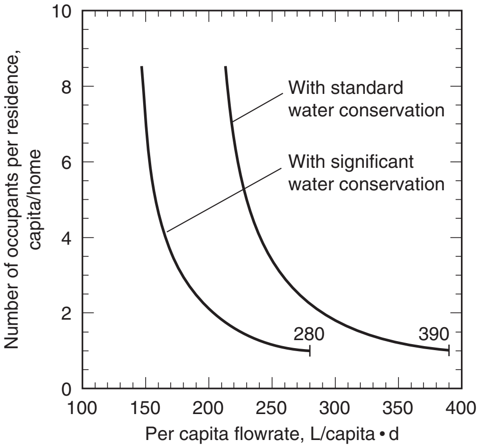
Table 3–2

Typical wastewater flowrates from urban residential sources in the United States

Household size, no. of persons	Flowrate, gal/capita·d: With current level of conservation	Flowrate, gal/capita·d: With extensive conservation	Flowrate, L/capita·d: With current level of conservation	Flowrate, L/capita·d: With extensive conservation
1	103	74	390	280
2	77	54	290	205
3	68	48	257	180
4	63	44	240	168
5	61	42	230	160
6	59	41	223	155
7	58	40	218	151
8	57	39	215	149

Figure 3–2

Per capita flowrate from individual residences as a function of the number of residents.



every effort should be made to obtain records from actual or similar facilities. If no other records are available, estimates for selected commercial sources, based on function or persons served, may be made using the data presented in Table 3–3. In the past, commercial wastewater flowrates were often based on existing or anticipated future development or comparative data. Flowrates were generally expressed in terms of quantity of flow per unit area [i.e., m^3/ha·d (gal/ac·d)]. Typical unit area flowrate allowances for commercial developments normally range from 7.5 to 14 m^3/ha·d (800 to 1500 gal/ac·d). The latter approach can be used to check the values obtained from existing records or estimates made using Table 3–3.

Institutional Facilities. Typical flowrates from some institutional facilities are shown in Table 3–4. Again, it is stressed that flowrates vary with the region, climate, and type of facility. The actual records of institutions are the best sources of flow data for design purposes.

Recreational Facilities. Wastewater flowrates from many recreational facilities are subject to seasonal variations. Typical data on wastewater flowrates from recreational facilities are presented in Table 3–5.

Strategies for Reducing Interior Water Use and Wastewater Flowrates

Because of the importance of conserving both resources and energy, various means for reducing wastewater flowrates are available. The reduction of wastewater flowrates from domestic sources results directly from the reduction in interior water use. Therefore, the terms *interior water use* and *domestic wastewater flowrates* are sometimes used interchangeably. Representative water use rates for various devices and appliances are reported in Table 3–6. Information on the relative distribution of water use within a residence is reported in Table 3–7. Devices and appliances that can be used to reduce interior domestic water use and wastewater flows are described in Table 3–8.

Another method of achieving flow reduction that has been adopted by a number of communities is to restrict the water usage of certain appliances, such as automatic dishwashers and kitchen food-waste grinders (i.e., garbage disposal units), that tend to increase water consumption. The use of one or more of the flow-reduction devices is specified for all new residential dwellings in many communities; in others, the use of waste-food grinders has been limited in new housing developments. Further, many individuals concerned

Table 3–3

Typical wastewater flowrates from commercial sources in the United States[a]

		Flowrate, gal/unit·d		Flowrate, L/unit·d	
Source	**Unit**	**Range**	**Typical**	**Range**	**Typical**
Airport	Passenger	2.4–3.8	3	9–14	11
Apartment	Person	32–45	38	120–170	145
Automobile service station	Vehicle served	6–11	8	23–42	30
	Employee	7–11	10	26–42	38
Bar/cocktail lounge	Seat	8–15	11	30–57	43
	Employee	8–12	10	30–45	37
Boarding house	Person	20–45	30	76–170	115
Conference center	Person	5–8	6	20–30	24
Department store	Toilet room	280–450	300	1000–1700	1100
	Employee	6–11	8	23–42	30
Hotel	Guest	52–56	53	200–215	200
	Employee	6–11	8	23–42	30
Industrial building (sanitary waste only)	Employee	12–26	15	45–98	60
Laundry (self-service)	Machine	320–413	338	1210–1560	1280
	Load	36–41	38	136–155	145
Mobile home park	Unit	100–113	105	380–430	400
Motel (with kitchen)	Guest	36–60	38	135–230	145
Motel (without kitchen)	Guest	32–53	34	120–200	130
Office	Employee	6–12	10	23–45	38
Public lavatory	User	2.4–3.8	3	9–14	12
Restaurant:					
Conventional	Customer	6–8	6	23–30	24
With bar/cocktail lounge	Customer	6–9	7	23–34	26
Shopping center	Employee	6–10	8	23–38	30
	Parking space	0.8–2.3	1.5	3–9	6
Theater (Indoor)	Seat	1.6–3	2.3	6–11	9

[a] Adapted in part from Tchobanoglous et al. (2003).

about conservation have installed such devices on their own as a means of reducing water consumption. New designs in front-loading clothes washers also offer significant reductions in water use, on the order of 50 to 75 percent of older models. A comparison of residential interior water use (and resulting per capita wastewater flows) is given in Table 3–9 for homes with the current levels of conservation and with extensive water conserving appliances and fixtures. The potential savings of employing selected water-efficient devices is illustrated in Example 3–1.

Table 3–4

Typical wastewater flowrates from institutional sources in the United States[a]

Source	Unit	Flowrate, gal/unit·d		Flowrate, L/unit·d	
		Range	Typical	Range	Typical
Assembly hall	Guest	1.6–3	2.3	6–11	9
Church	Seat	1.6–3	2.3	6–11	9
Hospital	Bed	128–240	150	480–900	570
	Employee	4–11	7.5	15–42	30
Institutions other than hospitals	Bed	60–94	75	230–360	285
	Employee	4–11	7.5	15–42	28
Prison	Inmate	60–110	90	240–430	340
	Employee	4–11	7.5	15–42	28
School, day					
With cafeteria, gym, and showers	Student	12–23	19	45–90	70
With cafeteria only	Student	8–15	11	30–60	42
School, boarding	Student	32–60	38	120–230	140

[a] Adapted in part from Tchobanoglous et al. (2003).

Table 3–5

Typical wastewater flowrates from recreational facilities in the United States[a]

Facility	Unit	Flowrate, gal/unit·d		Flowrate, L/unit·d	
		Range	Typical	Range	Typical
Apartment, resort	Person	40–53	45	150–200	170
Cabin, resort	Person	6.4–38	30	24–145	115
Cafeteria	Customer	1.6–3	2.3	6–11	9
	Employee	6.4–9	7.5	24–34	28
Camp					
With toilets only	Person	12–23	18.8	45–87	70
With central toilet and bath facilities	Person	28–38	33.8	106–144	128
Day	Person	12–15	11.3	45–57	43
Cottages, (seasonal with private bath)	Person	32–45	37.5	120–170	142
Country club	Member present	16–30	18.8	60–115	70
	Employee	8–11	9.8	30–42	37
Dining hall	Meal served	3–7.5	5.3	11–28	20

(*continued*)

Table 3–5 (*Continued*)

Facility	Unit	Flowrate, gal/unit·d Range	Flowrate, gal/unit·d Typical	Flowrate, L/unit·d Range	Flowrate, L/unit·d Typical
Dormitory, bunkhouse	Person	16–38	30	120–200	115
Fairground	Visitor	0.8–2.3	1.5	3–9	6
Picnic park with flush toilets	Visitor	4–7.5	3.8	15–28	14
Recreational vehicle park					
With individual connection	Vehicle	60–113	75	230–430	284
With comfort station	Vehicle	32–38	33.8	120–145	128
Roadside rest areas	Person	2.4–4	2.5	9–15	11
Swimming pool	Customer	4–9	6.8	15–34	26
	Employee	6.4–9	7.5	24–34	28
Vacation home	Person	20–45	37.5	76–170	142
Visitor center	Visitor	2.4–4	2.5	9–15	11

[a] Adapted in part from Tchobanoglous et al. (2003).

Table 3–6

Typical rates of water use for various devices and appliances in the United States

Device or appliance	US Customary Units Units	US Customary Units Range	US Customary Units Typical	SI Units Units	SI Units Range	SI Units Typical
Bathtub	gal/use	25–35	30	L/use	95–130	114
Dishwasher	gal/load	5–15	10	L/load	19–57	38
Faucet, typical	gal/min·use	0.5–4	2.5	L/min·use	1.9–15	9
Kitchen food waste grinder	gal/d	1–2	1.5	L/d	4–8	6
Shower, standard	gal/min·use	4–7	5	L/min·use	15–26	19
Shower, low-flow	gal/min·use	2–2.5	2.5	L/min·use	8–9.5	9
Toilet, pre 1980s	gal/use	4–7	6	L/use	15–26	23
Toilet, 1980-1992-3	gal/use	3–4	3.5	L/use	11–15	13
Toilet, tank, low-flow	gal/use	0.9[a]–1.6[b]	1.6	L/use	3.4–6	6
Washbasin	gal/min·use	1–3	2	L/min·use	8–11	8
Washing machine						
Top loading, standard	gal/load	40–50	45	L/load	150–190	170
Front loading, low-flow	gal/load	12–25	20	L/load	45–95	76

[a] Dual flush type.

[b] Currently, some states have adopted regulations mandating the use of 1.28 gal/flush toilets. In the future, it is anticipated that the 1.28 gal/flush toilet may become a national standard and that in the future the allowable water usage per flush may be reduced further to 1.0 gal/flush.

Table 3–7

Typical distribution of residential indoor water use in the United States[a]

Use	Percent of total		Typical water usage[b]	
	Range	Typical	gal/capita·d	L/capita·d
Bath	1.5–2	1.8	1.2	4.4
Clothes washing (Laundry)	20–24	23	15.0	56.6
Dishwashing	1–1.5	1.4	0.9	3.4
Faucet	15–18	16	10.4	39.4
Shower	16–20	18	11.7	44.3
Toilet flushing	24–30	28	18.2	68.9
Other domestic[c]	2–3	2.2	1.4	5.4
Leakage	8–12	9.6	6.2	23.6
Total		100	65.0	246.0

[a] Typical range of indoor water use in the United States ranges from 150 to 300 L/capita·d (40 to 80 gal/capita·d).

[b] Based on an indoor water usage rate of 246 L/capita·d (65 gal/capita·d).

[c] Houseplant watering, water for pets, etc.

Table 3–8

Flow reduction devices and appliances

Device/appliance	Description and/or application
Faucet aerators	Increases the rinsing power of water by adding air and concentrating flow, thus reducing the amount of washwater used
Flow-limiting shower heads	Restricts and concentrates water passage by means of orifices that limit and divert shower flow for optimum use by the bather
Low-flush toilets	Reduces the amount of water per flush
Pressure reducing valve	Reduces home water pressure below that of the water distribution system, decreases the probability of leaks and dripping faucets
Pressurized shower	Water and compressed air are mixed together. Impact provides the sensation of conventional shower
Retrofit kits for bathroom	Kits may consist of shower flow restrictors, toilet dams or fixture displacement bags, and toilet leak detector tablets
Toilet dam	A partition in the toilet tank that reduces the amount of water per flush
Toilet leak detectors	Tablets that dissolve in the toilet tank and release dye to indicate leakage of the flush valve
Vacuum toilet	A vacuum along with a small amount of water is used to remove solids from toilet
Water efficient dishwasher	Reduces the amount of water used to wash dishes
Water efficient clothes washer	Reduces the amount of water used to wash clothes. New front-loading machines have been developed that not only use less water but are more energy efficient

EXAMPLE 3–1 **Determine Water Savings by Employing Water Efficient Appliances** A new subdivision of 2000 homes is planned, and a condition of the building permit is to determine the potential savings in water consumption (and wastewater flows) if the following water-efficient appliances are used: front-loading washing machines, ultra-low flush toilets, and ultra-low flow shower heads. Use 3.5 residents per home and values for devices and appliances from Table 3–9 to determine the potential savings.

The estimated water use and percentage savings are illustrated in the following table.

		Unit water use, L/capita·d		Water Use, L/d	
Appliance/device	No. of Residents	With current level of conservation	With extensive conservation	With current level of conservation	With extensive conservation
Clothes washing	7000	56.6	36.0	396,200	252,000
Toilets	7000	68.9	31.0	482,300	217,000
Showers	7000	44.3	26.1	310,100	182,700
Total				1,188,600	651,700
Savings, %					45

Comment Three of the largest water using appliances and devices utilized in the home are those described in this example. Interior water use and the generation of wastewater can be reduced significantly with the community-wide installation of water-efficient appliances and devices, thus reducing the flows that will have to be handled by the collection system and treatment plant. Where high infiltration rates occur within the collection system, it is difficult or impossible to assess the beneficial effects of using water conservation devices.

Water Use in Other Parts of the World

The typical flowrates and use patterns presented in Tables 3–1 through 3–7 and Table 3–9 are based on water use and wastewater flowrate data from communities and facilities in the United States. Many developed countries (e.g., Canada) have flowrates in similar ranges. Water use, and consequently the per capita wastewater generation rate in less developed countries, is significantly lower. In some cases, the water supply is only available for limited periods of the day. Water use data from other parts of the world are given in Table 3–10. The striking aspect of the data presented in Table 3–10 is the variability that exists between countries. In general, information such as presented in Table 3–10 is not very useful, other than for gross comparisons, as it is not presented in the context of the country dynamics and the basis for the collection and reporting of data is not uniform.

Sources and Rates of Industrial (Nondomestic) Wastewater Flows

Nondomestic wastewater flowrates from industrial sources vary with the type and size of the facility, the degree of water recycling within the facility, and the presence of any onsite wastewater pretreatment or final treatment methods. Extremely high peak flowrates may be reduced by the use of onsite detention tanks and equalization basins. Typical design values for estimating the flows from industrial areas that have no or little

Table 3–9

Typical comparisons of indoor water use with current level of conservation and with extensive conservation practices and devices in the United States

	Flow, gal/capita·d		Flow, L/capita·d	
Use	**With current level of conservation[a,b,c]**	**With extensive conservation[d]**	**With current level of conservation**	**With extensive conservation**
Bath	1.2 (30)	1.2 (30)	4.4	4.5
Clothes washing (Laundry)	15.0 (30)	9.5 (20)	56.6	36.0
Dishwashing	0.9 (10)	0.7 (8)	3.4	2.6
Faucet	10.4 (3)	6.9 (2)	39.4	26.1
Shower	11.7 (4)	6.9 (2.5)	44.3	26.1
Toilet flushing	18.2 (3.3)[e]	8.2 (1.6)	68.9	31.0
Other domestic	1.4 (3)	1.4 (3)	5.4	5.3
Leakage	6.2	6.0	23.6	22.7
Total	65.0	40.8	246.0	154.4

[a] Rates of indoor water use based on values given in Table 3–7.

[b] Current level of conservation assumed to reflect mid-2013 values.

[c] Number in parenthesis is the assumed current water usage rate per use for various uses and devices as given in Table 3–6. For example, the number in parenthesis for the bath is 30 gal/use.

[d] Number in parenthesis is assumed to correspond to the water usage rate with extensive conservation based on the values given in Table 3–6.

[e] Estimated average of currently installed toilets.

wet-process type industries are 7.5 to 14 m^3/ha·d (1000 to 1500 gal/ac·d) for light industrial developments and 14 to 28 m^3/ha·d (1500 to 3000 gal/ac·d) for medium industrial developments. For industries without internal water recycling or reuse programs, it can be assumed that about 85 to 95 percent of the water used in the various operations and processes will become wastewater. For large industries with internal water-recycling programs, separate estimates based on actual water consumption records must be made. Average domestic (sanitary) wastewater contributed from industrial facilities per employee may vary from 30 to 95 L/capita·d (8 to 25 gal/capita·d).

Variations in Wastewater Flowrates

Wastewater flowrates vary during the time of day, day of the week, season of the year, and year to year depending upon the nature of the discharges to the collection system and sources and rates of infiltration/inflow. Short-term, seasonal, multiyear, and industrial variations in wastewater flowrates are briefly discussed here.

Short-Term Variations. The typical current diurnal pattern of wastewater flows observed at treatment plants, such as shown on Fig. 3–3(a) for an intermediate size community, began in the latter half of the 20th century. In the early part of the twentieth century, the daily variation in wastewater flowrates was characterized by a single peak in the morning as illustrated on Fig. 3–3(a). The shift from a single to a dual peak reflects the changes that have occurred in the workplace, perhaps the most notable event being the employment of women in factories during the early 1940s as part of the war effort.

Table 3–10

Water consumption in various countries and the United States in 2000[a]

Country	Per capita water consumption gal/d	 L/d
Argentina	93	350
Austria	113	430
Canada	196	742
Chile	63	238
Germany	41	156
Greece	93	350
Hungary	139	526
India	34	129
Kuwait	53	200
Libya	74	279
Nepal	8	30
Mexico	92	348
Mozambique	3	11
Norway	29	110
Russian Federation	72	274
Saudi Arabia	50	189
South Africa	59	224
United States	100	380

[a] Adapted from United Nations (2005).

Referring to Fig. 3–3(a), minimum flows occur during the early morning hours when water consumption is lowest and when the base flow consists of infiltration and small quantities of sanitary wastewater. The first flowrate peak generally occurs in the late morning when wastewater from the peak morning water use reaches the treatment plant. A second flowrate peak generally occurs in the early evening between 7 and 9 p.m. In some bedroom communities, the amplitude of the second peak will exceed the morning peak. It should also be noted that a shift occurs on the weekends with respect to the morning peak [see Fig. 3–3(a)] as people tend to get up a bit later.

The time of occurrence and the amplitude of the flowrate peaks vary with the size of the community and the length and storage capacity of the collection system [see Fig. 3–3(b)]. In the curves shown on Fig, 3–3(b), there is essentially no late afternoon peak. The reason for this occurrence is that the collection system has excess storage capacity and there is a significant travel time to reach the treatment plant. The same phenomena will be observed where a centralized treatment plant for a larger community also serves a number of smaller communities located some distance from the plant. Because of the travel time, the flows from the outlying communities will arrive later than the peak flow from the main community. Arriving later, these flows tend to dampen the second peak that would have been observed had the treatment plant only received wastewater from the single community. Further, as a community increases in size, the diurnal variations tend to be reduced as shown on Fig. 3–3(c). The masking effect caused by high infiltration rates during a storm event is illustrated on Fig. 3–3(d), in

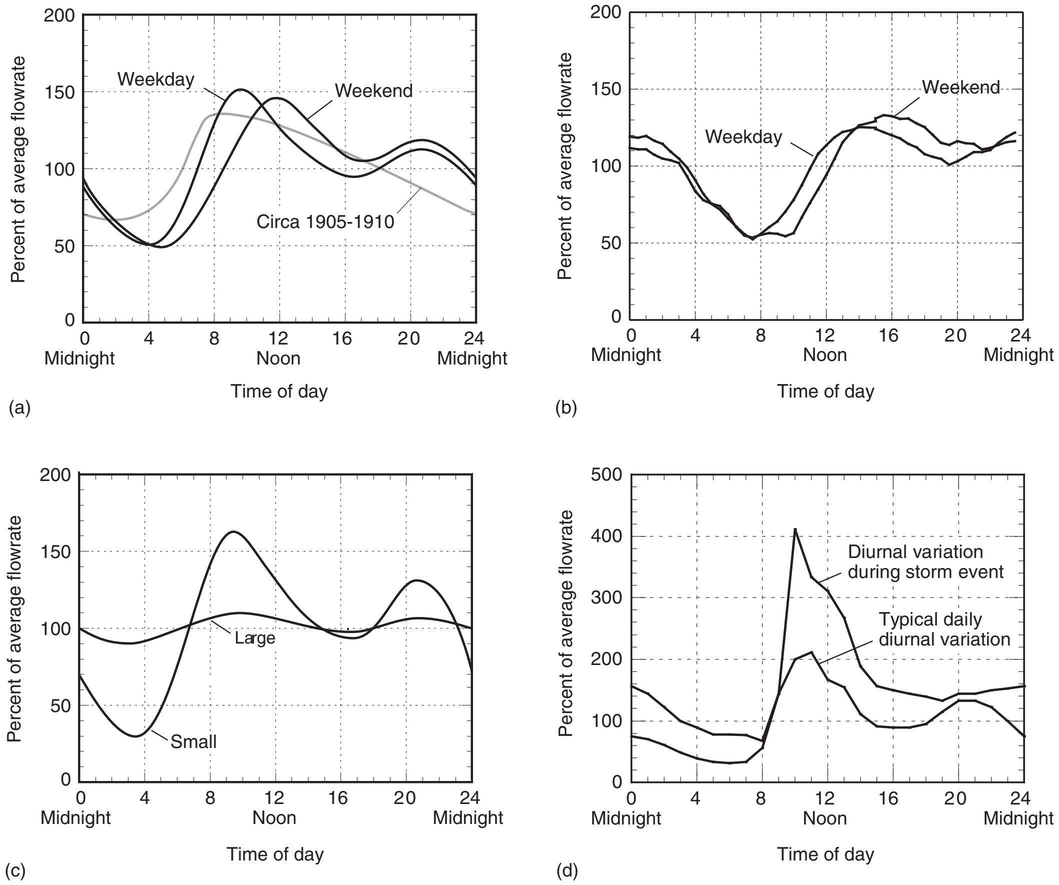


Figure 3–3

Typical variations in normalized influent flowrate data for domestic wastewater systems: (a) generalized diurnal pattern for weekdays and weekends in the latter part of the 20th century for an intermediate sized community (40,000–400,000 m^3/d), and the corresponding curve observed in the early part of the 20th century (ca. 1905–1910); (b) observed flowrate variations for the City of Davis (population 65,000), based on half hour flowrate measurements (courtesy of West Yost and Associates); (c) generalized flowrate variations for small (4000–40,000 m^3/d) and large (>400,000 m^3/d) communities; and (d) masking effect of excess stormwater infiltration, based on hourly flowrate measurements (note change in vertical scale).

which the late afternoon peak is masked by the continued flow resulting from the storm event.

When extraneous flows are minimal, wastewater discharge curves resemble water consumption curves, but with a lag of several hours. As the community size increases, the variations between the high and low flows decrease due to (1) the increased storage in the collection system of large communities that tends to equalize flowrates and (2) changes in the economic and social makeup of the community.

Industrial Variations. Industrial wastewater discharges are difficult to predict. Many manufacturing facilities generate relatively constant flowrates during production, but the flowrates change markedly during cleanup and shutdown. While internal process changes may lead to reduced discharge rates, plant expansion and increased production may lead to increased wastewater generation. Where joint treatment facilities are to be constructed, special attention should be given to industrial flowrate projections, whether they are prepared by the industry or jointly with the city's staff or engineering consultant. Industrial discharges are most troublesome in smaller wastewater treatment plants where there is limited capacity to absorb shock loadings.

Seasonal Variations. Seasonal variations depend on location and the nature of the community. In the eastern part of the United States, where it tends to rain throughout the year, there is less seasonal variation in observed flowrates as compared to the western United States where there are distinct wet (November through April) and dry (May through October) periods. Snowmelt is a significant factor in the areas with high rates of infiltration observed each spring, due to seasonably high groundwater levels in the Northeast and other snow-belt states. The difference in observed flowrates is illustrated on Fig. 3–4. It is important to note that the general patterns and the magnitude of the flowrates shown on Fig. 3–4 will vary considerably with increased or decreased rainfall patterns resulting from global climate change.

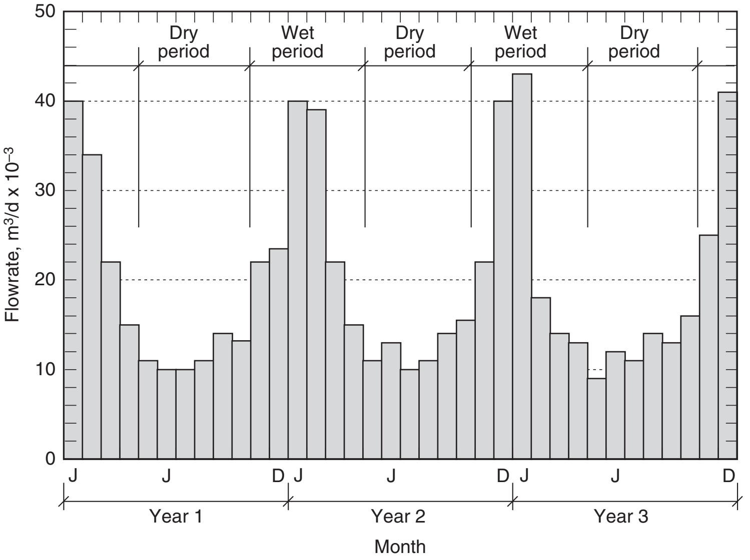
Seasonal variations in domestic wastewater flows are commonly observed at resort areas, in small communities with college campuses, and in communities that have seasonal commercial and industrial activities. The magnitude of the variations to be expected depends on both the size of the community and the seasonal activity.

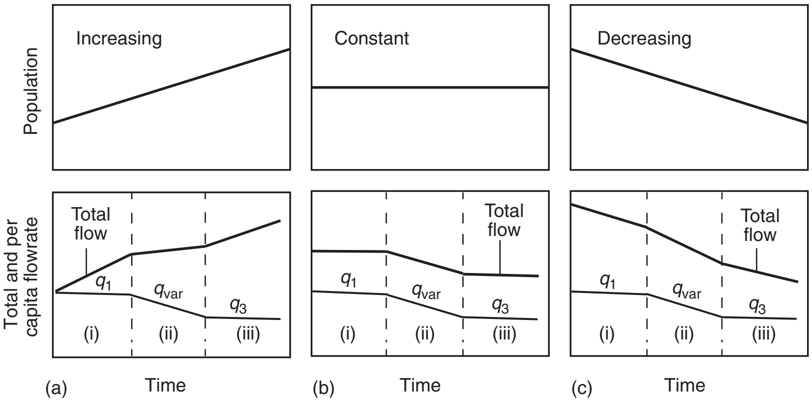
Long-Term Multiyear Variations Due to Conservation

In addition to the daily and seasonal changes described above, some significant trends have been observed in the long-term flowrates measured at wastewater treatment plants at many large cities in the United States. In general, three major trends in flowrates are observed based on cities with (1) increasing population, (2) relatively constant population, and (3) decreasing population as illustrated on Fig. 3–5. In all cases, the recent and ongoing implementation of low-flow appliances and fixtures and practices that result in water conservation will result in an increase in wastewater constituent concentrations.

Figure 3–4

Seasonal flowrate variations observed in the Western United States in the dry period (May through October) and wet period (November through April). In some locations, the distinction between dry and wet periods is becoming blurred due to global climate change.





q = per capita wastewater flowrate

(i) Pre-1992

(ii) Improved water conservation, period end point unknown

(iii) Maximum water conservation

Figure 3–5

Flowrate variations observed in cities in the United States with (a) increasing population, (b) constant population, and (c) decreasing population. The three time periods depicted in these plots are as follows: (i) time period up to about 1990 with relatively limited implementation of water conservation measures, (ii) period following 1990 when the per capita flowrate is decreasing as a result of greater implementation of conservation devices and public awareness, and (iii) period after the time that maximum water conservation has been implemented. The time when maximum conservation has been implemented will vary with each community.

Increasing Population. In cities where the population is increasing, the per capita wastewater flowrates were relatively constant up to the early nineties and the total wastewater flowrate increased in proportion to the population [see Fig. 3–5(a), period (i)]. After the early nineties, the per capita flowrate started to decrease reflecting changes in the plumbing code and the installation of water conserving appliances (e.g., low flush toilets). As more water conservation measures are adopted and water conservation devices and appliances become the norm [Note: the per capita flowrate, q_{var}, is variable in period (ii)], the total flowrate will continue to increase, but at a slower rate of increase [see Fig. 3–5(a), period (ii)]. At some point in the future, water conservation practices and low-flow devices and appliances will be installed in essentially all homes. At that point, the total flow will continue to increase with population, but at a stable and reduced per capita flowrate [see Fig. 3–5(a), period (iii)]. The result of these practices will be that while the constituent mass loading to the wastewater treatment plant will be expected to increase in proportion to the population increase, the concentration of the constituents will be higher than before.

Relatively Constant Population. In cities where the population has been relatively constant, the per capita flowrate was also more or less constant up to the early nineties [see Fig. 3–5(b), period (i)]. After the nineties, the total per capita flowrate started to decrease reflecting the installation of water conserving appliances (e.g., low flush toilets) [see Fig. 3–5(b), period (ii)]. The total flowrate will continue to decrease as more water conservation measures are adopted and low-flow devices and appliances are installed. As discussed above, at some point in the future water conservation devices and appliances will be installed in essentially all of the homes. At that point, the total per capita flowrate will again become more-or-less constant, but at a reduced per capita flowrate [see Fig. 3–5(b), period (iii)]. The constituent mass loading to the wastewater treatment plant is expected to remain relatively constant, but, as above, the concentration of the constituents will increase to reflect the reduced per capita flowrate.

Decreasing Population. In cities with a decreasing population, the per capita wastewater flowrate was relatively constant up to the nineties and the total flowrate decreased in proportion to the population [see Fig. 3–5(c), period (i)]. With the adoption of water conservation measures, total flowrate has decreased along with the the per capita wastewater flowrate [see Fig. 3–5(c), period (ii)]. When water conservation devices and appliances will be installed in essentially all of the homes, the total flow will again decrease with decreasing population, but at a reduced per capita flowrate [see Fig. 3–5(c), period (iii)]. The constituent mass loadings to the wastewater treatment plant are expected to decrease in proportion to the population decrease, while the concentration of the constituents will increase to reflect the reduced per capita flowrate.

Impact of Water Conservation on Future Planning

Because of the variations in flowrates now observed, great care must be taken in the selection of future flowrates for planning purposes. An important aspect in the selection of future flowrates will be to try to estimate the degree to which conservation measures will have already been implemented, the projected maximum level of water conservation that is expected, and the possible future use of satellite water reuse systems. Satellite water reuse systems, including onsite greywater systems, divert wastewater flow to onsite or local reuse applications and effectively reduce the flow of wastewater to the centralized wastewater treatment plant (Tchobanoglous and Leverenz, 2013).

3–2 IMPACT OF COLLECTION SYSTEM ON WASTEWATER FLOWRATES

In addition to the domestic and industrial sources of wastewater, as discussed above, other significant variables in estimating wastewater flowrates are the contributions from infiltration/inflow and from stormwater runoff and snowmelt, where combined collection systems are used. Exfiltration from collection systems can also affect the total quantity of wastewater. The flowrate contributions from these sources are considered in this section.

Infiltration/Inflow

Extraneous flows in collection systems, described as infiltration and inflow, are illustrated on Fig. 3–6 and are defined as follows:

Infiltration. Water entering a collection system from a variety of entry points, including service connections and from the ground through such means as defective pipes, pipe joints, connections, or access port (manhole) walls and joints.

Steady inflow. Water discharged from cellar and foundation drains, cooling-water discharges, and direct connections from springs and swampy areas. This type of inflow is steady and is identified and measured along with infiltration.

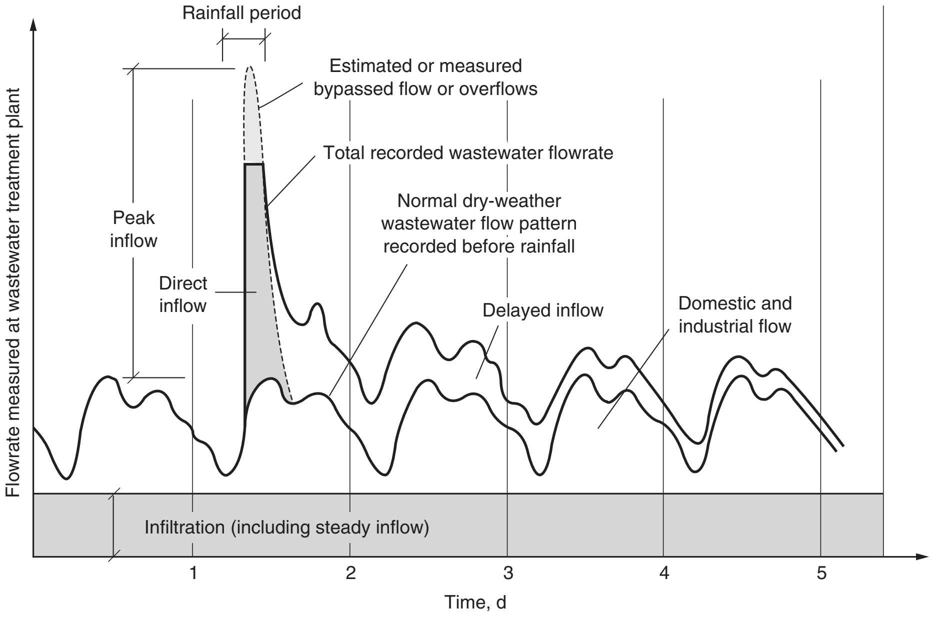
Direct inflow. Inflow sources that result from direct stormwater runoff connections to the sanitary or combined collection system and cause an almost immediate increase in wastewater flowrates. Possible sources are roof leaders, yard and areaway drains, access port covers, cross connections from storm drains and although typically confined to combined systems, catch basins. River or tidal intrusion from faulty tide or backwater gates is another source of direct inflow.

Total inflow. The sum of the direct inflow at any point in the system plus any flow leaving from the system upstream through sanitary sewer overflow (SSO) and combined sewer overflow (CSO) discharges, pumping station bypasses, and the like.

Delayed inflow. Delayed inflow, which is not caused by the direct connections listed above, can include foundation drains, the discharge of sump pumps from cellars, and the slowed entry of surface water through access ports (manholes) in ponded areas.

Figure 3–6

Graphic identification of infiltration/inflow.



The initial impetus in the United States for defining and identifying infiltration/inflow was the Federal Water Pollution Control Act Amendments of 1972. As a condition of receiving a federal grant for the design and construction of wastewater treatment facilities, grant applicants had to demonstrate that their wastewater collection systems were not subject to excessive infiltration/inflow. The several benefits to the community that are realized by "tightening" the collection system and reducing the rates of infiltration/inflow include (1) reducing wastewater backups and SSOs in separate sanitary collection systems and reducing CSOs in combined collection systems (see Fig. 1–2 in Chap. 1), (2) increasing the efficiency of operation of wastewater treatment plants, and (3) improving the utilization of collection system hydraulic capacity for wastewater requiring treatment instead of for infiltration/inflow. Because an understanding of the effects of infiltration/inflow is important in determining treatment plant flowrates, a discussion of excessive infiltration/inflow is included in this section.

Infiltration into Collection Systems. Rainfall runoff can be divided into three components for a given area: one portion runs quickly into the stormwater systems or other drainage channels, another portion evaporates or is absorbed by vegetation, and the remainder percolates into the ground, becoming groundwater. The proportion of the rainfall that percolates into the ground depends on the character of the surface and soil formation and on the rate and distribution of the precipitation. Any reduction in permeability, such as that due to buildings, pavements, or frost, decreases the opportunity for precipitation to become groundwater and increases the surface runoff correspondingly. The amount of groundwater flowing from a given area may vary from a negligible amount for a highly impervious district or a district with dense subsoil to 25 or 30 percent of the rainfall for a semi-pervious district with sandy subsoil permitting rapid passage of water. The percolation of water through the ground from rivers or other bodies of water sometimes has considerable effect on the groundwater table, which rises and falls continually.

Impact of High Groundwater. The presence of high groundwater levels results in leakage into the collection systems and in an increase in the quantity of wastewater and the expense of its conveyance and treatment. This occurrence is of particular importance in the Northeast when spring snowmelt occurs. The amount of flow that can enter a collection system from groundwater infiltration, may range from 0.01 to 1.0 m^3/d·mm-km (100 to 10,000 gal/d·in.-mi) or more. The number of millimeter-kilometers (inch-miles) in a wastewater collection system is the sum of the products of sewer diameters, in millimeters (inches), times the lengths, in kilometers (miles), of sewers of corresponding diameters.

Estimating Infiltration. Infiltration may also be estimated based on the area served by the collection system and may range from 0.2 to 28 m^3/ha·d (20 to 3000 gal/ac·d) (Metcalf & Eddy, 1981). The variation in the amount of infiltration encompasses a wide range of values because the lot sizes may vary in area, which in turn affects the length and extent of the collection system network. During heavy rains, when there may be leakage through access port covers, or inflow as well as infiltration, the rate may exceed 500 m^3/ha·d (50,000 gal/ac·d).

Factors Affecting Infiltration. Infiltration/inflow is a variable part of the wastewater flowrate and depends on the quality of the material and workmanship in constructing the collection systems and building connections, the collection system age, the collection system maintenance program, and the elevation of the groundwater compared with that of the collection system. The rate and quantity of infiltration depend on the length of the collection system, the area served, the soil and topographic conditions, and, to a certain extent, the population density (which affects the number and total length of house connections). Although the elevation of the water table varies with the quantity of rain and melting snow percolating into the ground, the leakage through defective joints, porous concrete, and cracks has been, in some cases, large enough to lower the groundwater table to the wastewater level in the collection system.

The pipe joints in most of the collection systems built during the first half of the twentieth century were constructed with cement mortar or hot-poured bituminous compounds. Access ports were almost always constructed of brick masonry. Deterioration of pipe joints, pipe-to-access port joints, and the waterproofing of brickwork has resulted in a high potential for infiltration into these old sewers. The use of high-quality pipe with dense walls, precast access port sections, and joints sealed with rubber or synthetic gaskets is standard practice in modern collection system design. The use of these improved materials has greatly reduced rates of both infiltration and exfiltration from newly constructed collection systems. However, as these newer systems age, infiltration rates will rise over time, though at a much lower rate than with older sewers.

As noted above, a key factor in the rate of infiltration of both older and newer collection systems is the quality of the actual installation. Even some newer systems, however, exhibit higher than expected rates of infiltration, and in some cases inflow as well, due to poor construction practices or lack of construction supervision and inspection.

Inflow into Collection Systems

As described previously, the type of inflow that causes a "steady flow" cannot be identified separately and so is included in the measured infiltration. The direct inflow can cause an almost immediate increase in flowrates in sanitary systems. The effects of inflow on peak flowrates that must be handled by a wastewater treatment plant are shown in Example 3–2.

EXAMPLE 3–2 **Determine Infiltration/Inflow from Wastewater Flow Records** A large city has measured high flowrates during the wet season of the year. The flow during the dry period of the year, when rainfall is rare and groundwater infiltration is negligible, averages 120,000 m^3/d (31.7 Mgal/d). During the wet period when groundwater levels are elevated, the flowrate averaged 230,000 m^3/d (60.8 Mgal/d) excluding those days during and following any significant rainfall events. During a recent storm, hourly flowrates were recorded during the peak flow period, as well as several days following the storm. The flow plots are shown on the accompanying figure. Compute the infiltration and inflow and determine if the infiltration is excessive. Excessive infiltration is defined by the local regulatory agency as rates over 0.75 $m^3/d\cdot mm\text{-}km$ (8000 gal/d·in.-mi) of collection system. The composite diameter-length of the collection system is 270,000 mm-km (6600 in.-mi).

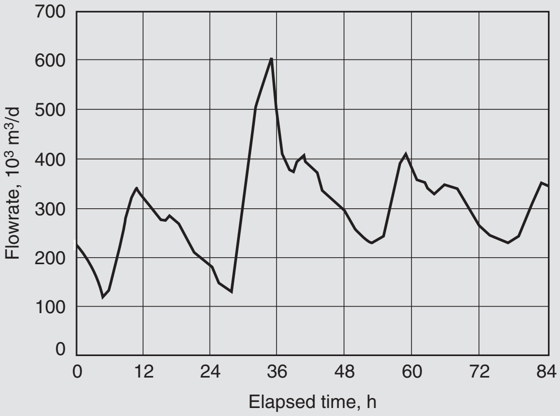
Solution

1. Determine the infiltration and inflow components during the wet season.
   a. Because infiltration is low during dry periods, high groundwater infiltration is

$$\begin{aligned}\text{Infiltration} &= (230{,}000 - 120{,}000)\ m^3/d \\ &= 110{,}000\ m^3/d\end{aligned}$$

   b. The maximum hourly inflow is determined graphically from the accompanying figure as the difference between the peak hourly wet-weather flow during the storm and the comparable flowrate on the preceding day. The maximum flowrate at hour 35 is 606,000 m^3/d, and the flow at hour 11 is 340,000 m^3/d. In this case, the maximum inflow is

$$\begin{aligned}\text{Inflow} &= (606{,}000 - 340{,}000)\ m^3/d \\ &= 266{,}000\ m^3/d\end{aligned}$$



2. Determine if the infiltration is excessive.
   a. Calculate the infiltration by dividing the calculated flowrate by the composite diameter-length of the collection system.

$$\frac{(110{,}000\ m^3\cdot d)}{(270{,}000\ mm\cdot km)} = 0.407\ m^3/d\cdot mm\text{-}km$$

   b. Using the regulatory agency criterion of 0.75 $m^3/d\cdot mm\text{-}km$, the infiltration is not excessive.

Comment In this example, the peak flow during the storm period was 4.7 times the average dry weather flow. As discussed later in this chapter, the peak flow factor is high for a system of this size. Because inflow represents over 50 percent of the peak flow and requires oversizing of the hydraulic capacity of the treatment plant, methods of inflow reduction should be investigated to decrease the hydraulic load on the collection system and treatment facilities.

Exfiltration from Collection Systems

Collection systems that exhibit high rates of infiltration may also exhibit high rates of exfiltration during low groundwater conditions. When exfiltration occurs, untreated wastewater leaks out of pipe joints and service connections. If the piping and joints are in poor condition, significant quantities of wastewater may seep into the ground, travel through the gravel bedding of the piping system, or even surface in extreme cases. Seepage of untreated wastewater into the ground near shallow wells can result in pollution of the water supply. Well contamination has occurred in urban areas such as Los Angeles, CA, where collection systems are within 300 m (1000 ft) of water wells. Exfiltration in collection systems near surface water bodies can also contribute to ongoing high coliform counts in those water bodies that may be difficult to correct. Reduction of inflow/filtration in collection systems may serve to limit exfiltration and remove potential threats to water supplies and public health. The potential adverse effects of exfiltration on surface water quality are illustrated in Example 3–3.

EXAMPLE 3–3 **Determine the Pollution Contributing Effects of Exfiltration on Nearby Water Body** Untreated wastewater from a damaged collection system pipeline leaked into a small nearby lake. The leakage is estimated to be 10,000 L/d. What increase in concentration of organisms can be expected in the lake if the initial coliform count in the wastewater is 10^7 organisms/100 mL, the wastewater is diluted by a ratio of 1000 to 1, and the leak lasted for one day? Assume the number of coliform organisms in the dilution water is zero.

Solution

1. To solve this problem, a concentration balance must be prepared (see Chap. 1 for additional details for mass and concentration balances). The required concentration mass balance is:

Total number of organisms in mixture	=	total number of organisms in leakage	+	total number of organisms in dilution water
$Q_M C_M$	=	$Q_L C_L$	+	$Q_{DW} C_{DW}$

Where Q_M = the volume of the mixture $(Q_L + Q_{DW})$
C_M = number of organisms/100 mL in the mixture
Q_L = volume of leakage
C_L = number of organisms/100 mL in the leakage
Q_{DW} = volume of the dilution water
C_{DW} = number of organisms/100 mL in the dilution water

2. Substitute for the various quantities and solve for the number of organisms in the mixture.

$$[(10^4\text{ L/d})(1\text{ d}) + (10^3)(10^4\text{ L})]\left(\frac{C_M}{100\text{ mL}}\right)$$

$$= (10^4\text{ L/d})(1\text{ d})\left(\frac{10^7}{100\text{ mL}}\right) + (10^3)(10^4\text{ L})\left(\frac{0}{100\text{ mL}}\right)$$

$$(10^4\text{ L} + 10^7\text{ L})\left(\frac{C_M}{100\text{ mL}}\right) = (10^4\text{ L})\left(\frac{10^7}{100\text{ mL}}\right)$$

$$C_M \approx \frac{10^4}{100\text{ mL}}$$

Comment Because untreated wastewater contains a high concentration of coliform organisms, the receiving water, even under the assumed condition that the exfiltration is well mixed, can contain high concentration levels. Such high coliform counts would be taken as indicators that a potential health hazard may exist. Therefore, in this example, because the exfiltration from the collection system into the lake is significant, receiving water quality objectives will be difficult to maintain.

Combined Collection System Flowrates

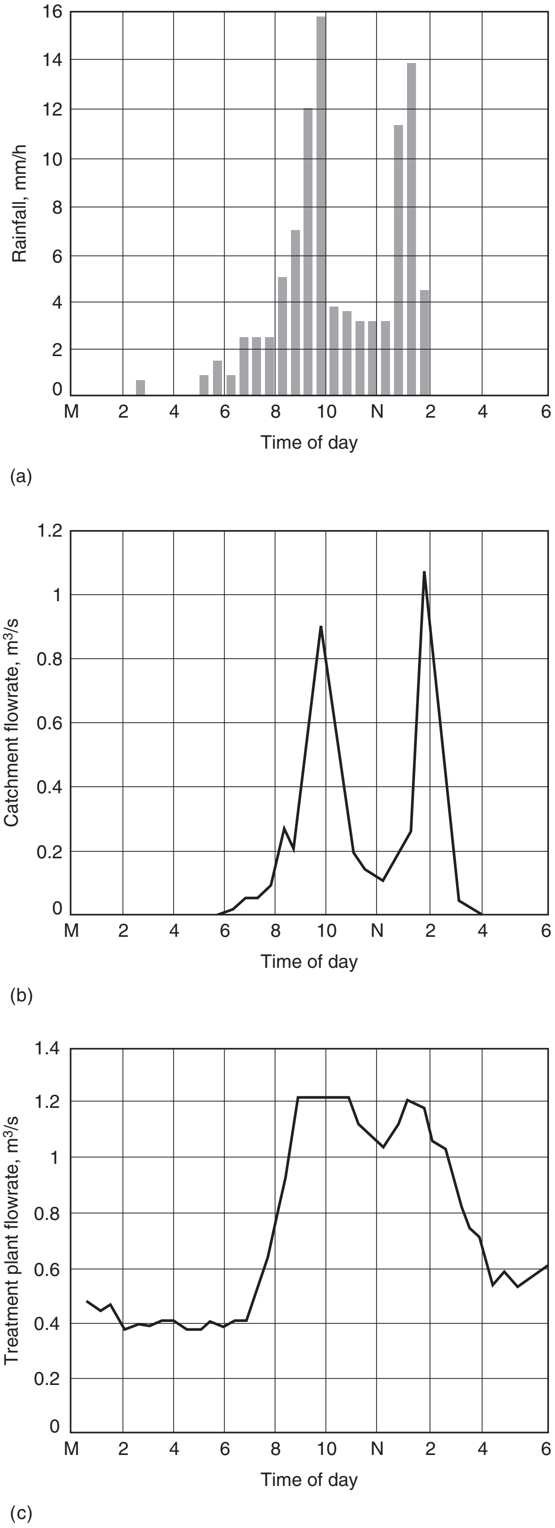
Flow in a combined collection system is composed of the same flow components found in a sanitary sewer system, but with the addition of two major sources of inflow: stormwater runoff during wet-weather periods and snowmelt, typically during late winter and early spring (see Fig. 1–1 in Chap. 1). The dry-weather flow may include domestic, commercial, and industrial wastewater plus infiltration and inflow. During a rainfall event, the amount of flow directly attributable to the storm—the inflow—is significantly larger than the dry-weather wastewater flow and can completely mask the typically observed dry-weather flow patterns. Further, some recent short-duration, high-intensity rainfall events, attributed to global climate change, have resulted in flowrates in combined collection system far in excess of the wet-weather flowrates observed normally.

Past Design Constraints. While combined systems were designed to convey both sanitary and stormwater flows, the downstream interceptors, pumping stations, and wastewater treatment plants were typically designed, at a much later date, to only convey and treat a fraction of the total wet-weather flow generated within the system. Hydraulic relief points, consisting of regulator or diversion structures and CSO outfalls, were strategically located to limit the flow to both the interceptor sewers and treatment plant. While CSO discharges are legal, communities with combined systems are required to develop plans to eliminate or greatly reduce the frequency and magnitude of these wet-weather discharges so that the water quality standards of the receiving waters are met. In some cases where the combined system is undersized, flooding or surcharging (surcharging results when the pipeline capacity is exceeded) may occur at various upstream locations within the system, which create potential threats to public health.

Effects of Combined System Flowrates. The effects of combined system flowrates are illustrated on Fig. 3–7. The catchment hydrograph (flow versus time), as illustrated on Fig. 3–7(b), closely resembles that of the variations in the rainfall hyetograph [see Fig. 3–7(a)]. The short response time between the rainfall event and the increase in the flowrate can be taken as an indication of a short travel time for flow from all points in the upstream combined system. In contrast, the hydrograph at the treatment plant [see Fig. 3–7(c)] shows less distinct flow peaks and a lag time for flows to return to normal dry weather levels following rainfall cessation. The higher flows at this location are due to the larger contributing combined system, and the smoothed peaks result from loss of flow through CSO outfalls and hydraulic routing effects. The peak flowrates and accompanying mass loadings, however, must be accounted for in the hydraulic design of the treatment plant or dedicated CSO control facilities and in the selection of appropriate unit operations and processes.

Figure 3–7

Flow variations in a combined collection system during wet weather; (a) rainfall hyetograph, (b) typical catchment flowrate, and (c) observed treatment plant flowrate.



Direct Measurement of Combined Sewer Flowrates and Wastewater Characteristics

Combined wastewater flows and characteristics can be monitored at various points within the collection system, including in the combined sewers, at regulator structures, CSO outfalls, dedicated CSO control facilities, and the wastewater treatment plant. When performing receiving water impact studies, monitoring of the flow and characteristics in the CSO outfalls is desired. Monitoring points in the collection or interceptor system may be needed for a number of other reasons, including defining the flow to be controlled, diverted, or treated.

Temporary Flow Monitoring. In performing studies of combined wastewater collection systems, temporary monitoring facilities are installed and left in place for several storm events. Flow metering in such installations is typically performed using portable, battery-operated, depth-and-velocity–sensing instrumentation [see Figs. 3–1(a) and (b)]. Similarly, wastewater samples are taken using portable, battery-operated, programmable sampling devices. These samplers are preset for desired sampling time intervals and are level or flow actuated.

Permanent Flow Monitoring. Permanent flow monitoring installations are used in some systems to allow continuous flow records to be obtained at critical points. Also, these installations can allow centralized control of transport system facilities to maximize storage of combined wastewater in the system or to control flow to the downstream treatment plant. The flow data recorded at the site may be periodically recovered manually, or the data may be telemetered to a central location for data analysis or as inputs into real time control systems.

Need to Collect Rainfall Data. To understand how combined wastewater flowrate and characteristics respond to a storm event, rainfall data must be obtained. Therefore, it is often necessary to install temporary rainfall monitoring equipment in close proximity to the drainage area tributary to the monitoring location. Continuous recording rain gauges capable of monitoring the rainfall depth over time should be used. The tipping bucket rain gauge is one type that can be used to record rainfall continuously in 0.25 mm (0.01 in.) increments on a clock-driven recorder chart. If the combined system being monitored encompasses a large area, installation of several rain gauges may be necessary to record spatial variation of rainfall characteristics across the entire area. Sophisticated radar-rainfall systems are now used typically for large regional systems where SSO or CSO monitoring is essential to the mitigation planning.

Calculation of Combined Sewer Flowrates

Calculation of flowrates in a combined system is a complicated and challenging undertaking. The variability in weather patterns and changing nature of rainfall events (e.g., short-duration high-intensity rainfall events) that have been observed in recent years, especially on the Midwest and East Coast, has further complicated the task. The first step in the process involves quantifying wastewater, rainfall-runoff, and other sources of flow such as groundwater infiltration. These sources of flow are then combined and routed through the various components of the system. Finally, the volumes of flow exiting the system through CSO outlets, entering the downstream treatment facility, or being transported to other points in the system are determined.

Computer Modeling. Due to the complexity of combined systems, it is normally necessary to use computer models to simulate the complete combined system including dry weather wastewater flows, hydraulic routing through the piping system, discharges through the outlets, and the flow through the interceptor and to the treatment plant. New systems are typically separated, but existing combined systems may require evaluation and design relative to increased sanitary wastewater inputs and, mostly, relative to CSO control. The models used for combined sewer systems need to be transient to simulate flow variations during storms. There are a number of models available, with ever-increasing capabilities such as graphical user interfaces, GIS interaction (for model development), multi-user functionality, semi-automatic calibration, graphical results displays, animations, and two-dimensional above-ground flow simulation.

The USEPA Stormwater Management Model (SWMM) is a public domain model that has seen widespread use and upgrades since its development in the 1970s. Several proprietary models or user interfaces are based on SWMM, including PC-SWMM, XP-SWMM, and InfoSWMM. A number of independent proprietary models have also been used widely, including Mike Urban, InfoWorks, and SewerGEMS. These models are also applicable to separate systems where infiltration and inflow (I/I) is similar to runoff, except that it can only be simulated empirically based on flow monitoring data. Software capabilities are now such that large, complex systems can be simulated for extended periods of time. But, for any application, model calibration with flow monitoring data is essential to ensure reliability of the results, which are often the basis of large capital investments.

Model Calibration and Verification. The process of calculating flows in a combined system using any of the computer models normally involves comparison of measured versus predicted flows at selected locations in the system. During model calibration, the model is run with rainfall data collected from one storm, and the calculated results are compared with the observed field results. Estimated input parameters are then adjusted within reasonable bounds to obtain best fit between predictions and measurements. During verification, data sets from other storms are used, and no adjustments of parameters are allowed. This calibration and verification process is essential in assessing the predictive capability of the model.

3–3 ANALYSIS OF WASTEWATER FLOWRATE DATA

Because the hydraulic design of both collection and treatment facilities is affected by variations in wastewater flowrates, the flowrate characteristics have to be analyzed carefully from existing records. In cases where only flowrate data in the collection system are available, it must be recognized that the flowrates may differ somewhat from the flowrate entering the treatment plant because of the flow dampening effect of the wastewater collection system. The statistical analysis of flowrate data and the determination of design parameters are considered in this section.

Statistical Analysis of Flowrate Data

The statistical analysis of wastewater flowrate and constituent concentration data involves the determination of statistical parameters used to quantify a series of measurements. The statistical parameters and graphical techniques used for the analysis of wastewater management data are summarized in Appendix D.

Type of Probability Distribution. The two most common types of probability distributions encountered in the statistical analysis of flowrate data are (1) the standard normal distribution and (2) the log-normal distribution, in which log of the values is distributed normally. If the data are distributed normally, the statistical measure used to define the distribution include the mean, variance, skewness, and coefficient of kurtosis. Skewness and coefficient of kurtosis are needed to quantify the nature of a given distribution. If a distribution is highly skewed, as determined by the coefficient of skewness, normal statistics cannot be used. Another term used to assess the relative variability in the data is *coefficient of* *variation* (see Appendix D). If the data are distributed log-normally, the statistical measure used to define the distribution include the geometric mean and the geometric standard deviation (see Appendix D). In general, most flowrate data are distributed log-normally because of the wide range of wet-weather flowrates as compared to the average dry-weather flowrates. The statistical analysis of flowrate data is illustrated in Example 3–4.

EXAMPLE 3–4 **Statistical Analysis of Wastewater Flowrate Data** Using the following weekly flowrate data obtained, from an industrial discharger, for the dry (May–October) and wet (November–April) periods, determine the statistical characteristics and estimate the maximum weekly flowrate that will occur during each period.

	Flowrate, m^3/wk			Flowrate, m^3/wk	
Week	Dry period	Wet period	Week	Dry period	Wet period
1	13,500[a]	20,000[b]	14	37,000	51,600
2	25,900	16,250	15	30,100	41,250
3	28,750	40,350	16	21,250	35,000
4	10,750	18,600	17	23,500	30,750
5	12,500	18,300	18	16,750	23,900
6	9850	18,750	19	8350	16,350
7	13,900	21,800	20	18,100	30,200
8	15,100	20,200	21	9250	21,100
9	23,400	23,750	22	9900	21,750
10	21,900	42,500	23	8750	20,800
11	23,700	32,000	24	15,500	24,500
12	18,000	28,300	25	7600	14,400
13	26,400	28,300	26	8700	15,200

[a] First wk of May.
[b] First wk of November.

Solution

1. Determine the nature of the distribution by plotting the data on arithmetic and log-probability paper.
   a. Set up a data analysis table with four columns as described below.
      i. In column 1, enter the rank serial number starting with number 1
      ii. In column 2, enter the probability plotting position (see Appendix D)
      iii. In columns 3 and 4, arrange the weekly flowrate data in ascending order

		Flowrate, m^3/wk				Flowrate, m^3/wk	
No.	Plotting position, %[a]	Dry period	Wet period	No.	Plotting position, %[a]	Dry period	Wet period
1	3.7	7600	14,400	14	51.9	16,750	23,750
2	7.4	8350	15,200	15	55.6	18,000	23,900
3	11.1	8700	16,250	16	59.3	18,100	24,500
4	14.8	8750	16,350	17	63.0	21,250	28,300

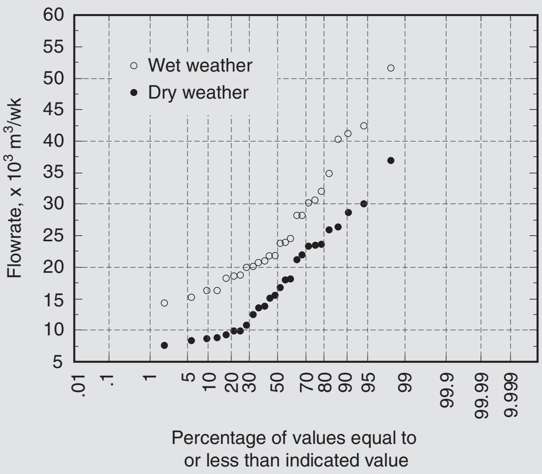
(*continued*)

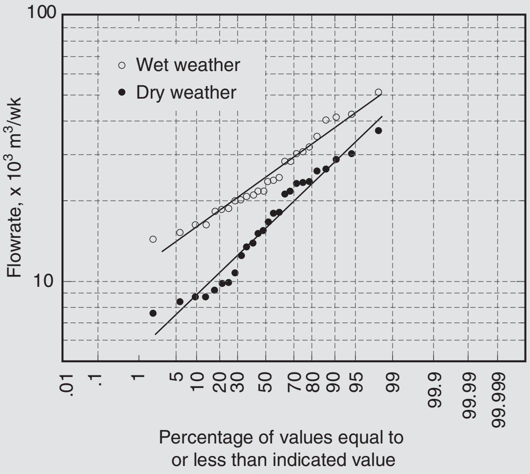
(*Continued*)

No.	Plotting position, %[a]	Flowrate, m^3/wk		No.	Plotting position, %[a]	Flowrate, m^3/wk	
		Dry period	Wet period			Dry period	Wet period
5	18.5	9250	18,300	18	66.7	21,900	28,300
6	22.2	9850	18,600	19	70.4	23,400	30,200
7	25.9	9900	18,750	20	74.1	23,500	30,750
8	29.6	10,750	20,000	21	77.8	23,700	32,000
9	33.3	12,500	20,200	22	81.5	24,600	35,000
10	37.0	13,500	20,800	23	85.2	25,900	40,350
11	40.7	13,900	21,100	24	88.8	28,750	41,250
12	44.4	15,100	21,750	25	92.6	30,100	42,500
13	48.2	15,500	21,800	26	96.3	37,000	51,600

[a] Plotting position, % = $\left(\frac{m}{n+1}\right) \times 100$, where $n = 26$ [Equation (D–10), Appendix D].

b. Plot the weekly flowrates expressed in m^3/wk versus the plotting position. The resulting plots are presented below. Because the data fall on a straight line on log-probability paper, they are log-normally distributed.





2. Determine the geometric mean for the dry and wet periods and the corresponding geometric standard deviation using Eq. (D–9) from Appendix D.

$$s_g = \frac{P_{84.1}}{M_g} = \frac{M_g}{P_{15.9}}$$

Period	M_g	$P_{84.1}$	s_g
Dry	15,948	25,198	1.58
Wet	24,504	34,391	1.40

Based on the value of the geometric standard deviation, it can be concluded that the observed variation is large.

3. Estimate the probable annual maximum weekly flowrate during the dry and wet periods.
   a. Determine the probability factor

$$\text{Peak week} = \left(\frac{m}{n+1}\right) \times 100 = \left(\frac{26}{26+1}\right) \times 100 = 96.3$$

   b. Determine the flowrate from the figure given in Step 1b at the 96.3 percentile
      Peak dry week flowrate = 35,948 m^3/wk
      Peak wet week flowrate = 44,900 m^3/wk

Comment The statistical analysis of data is important in establishing the design conditions for wastewater treatment plants. The application of statistical analysis to the selection of design flowrates and mass loadings is considered in the following section.

Developing Design Parameters from Flowrate Data

Quantifying the variations in flowrates is important in the design and operation of wastewater treatment plants. Before considering the quantification of flowrate variations, it will be helpful to define some of the parameters that are used commonly to quantify the observed variations.

Flowrate Parameters. The principal parameters used to quantify the observed variations in flowrate are defined in Table 3–11. As will be discussed in Sec. 3–7, these terms are also of importance in the selection and sizing of individual unit treatment processes and operations. As reported in Table 3–11 the parameters used to describe flowrate variations can be grouped into five categories: (1) average values including wet- and dry-weather and annual flows, (2) various peak values, (3) maximum values, (4) minimum values, and (5) sustained values. Using a 1-d record for the purposes of illustration, the following parameters are identified on Fig. 3–8: average daily flowrate, instantaneous peak, peak hour, minimum hour, minimum, and sustained 13.5-h flowrate. Obviously, a two- or three-year record is needed to establish the parameters shown on Fig. 3–8 and the long-term parameters identified in Table 3–11.

Flowrate Ratios. Because it is difficult to compare numerical peak flow values from different wastewater treatment plants, peak flowrate values are normalized by dividing by the long-term average flowrate. The resultant ratio, known as a peaking factor, is defined as follows.

$$\text{Sustained peaking factor, PF} = \frac{\text{peak flowrate (e.g., hourly, daily, monthly)}}{\text{long-term average flowrate}} \tag{3–1}$$

Peaking factors are applied most frequently to determine the peak hourly flowrate. For example, the sustained hourly peaking factor relative to the average flowrate for the daily flowrate record given on Fig. 3–8 is 1.72 [(183 m^3/d)/(106 m^3/d)]. Where flowrate records are available, at least three years of data should be analyzed to define the peak to average day peaking factor. The peaking factor is particularly useful in estimating the maximum hydraulic conditions that might occur and have to be accommodated. Peaking factors can also be applied to mass loadings.

Table 3–11

Parameters used to quantify observed variations in flowrate and constituent concentrations[a]

Parameter	Description[b]
Average dry weather flow (ADWF)	The average of the daily flow data for dry weather periods[c]
Average wet weather flow (AWWF)	The average of the daily flow for wet weather periods[c]
Average daily flow	The average flow occurring over a 24-h period based on total annual flowrate data
Instantaneous peak flowrate	Highest recorded flowrate occurring for a period consistent with the reporting period for the metering equipment. In many situations the recorded peak flow may be considerably below the actual peak flow because of metering and recording equipment limitations
Peak hourly flowrate	The peak sustained hourly flowrate occurring during a 24-h period based on annual operating data (see also Instantaneous peak flow)
Maximum day flow	The maximum flow occurring over a 24-h period based on annual operating data
Maximum monthly flowrate	The maximum daily flow sustained for a period of 1 mo in the record examined. In practice, the maximum month is typically the maximum 30-d value that occurs in the reporting period per the NPDES permit. Although not statistically correct, it is the value that is reported
Minimum hourly flowrate	The minimum sustained hourly flowrate occurring over a 24-h period based on annual operating data
Minimum daily flow	The minimum flow that occurs over a 24-h period based on annual operating data
Minimum monthly flow	The minimum daily flow that occurs over a period of 1 mo based on annual operating data (see Maximum month above)
Sustained flow or load	The flow or mass loading rate sustained or exceeded for a specified period of time (e.g., 1 h, 1 d, or 1 mo) based on annual operating data

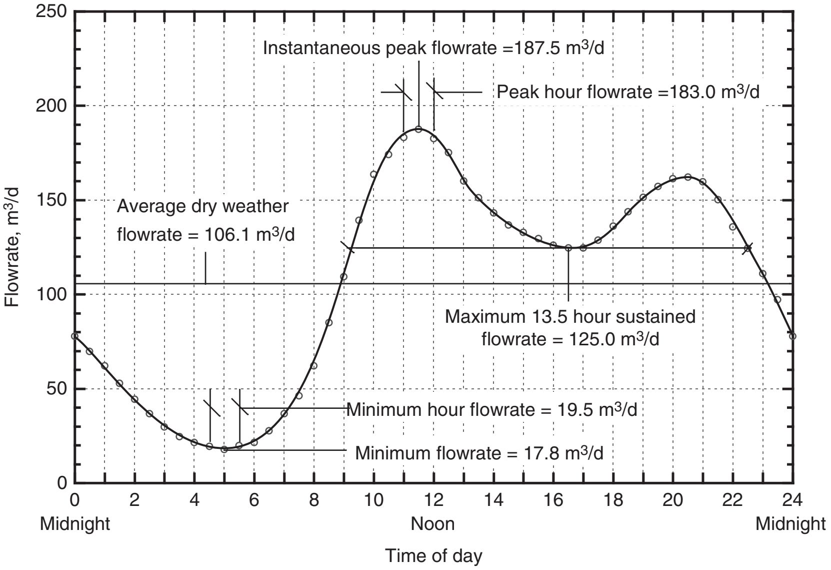
[a] Adapted in part from Crites and Tchobanoglous (1998).
[b] The definitions can also be applied to constituent mass loadings.
[c] In some parts of the country, the distinction between wet and dry periods is becoming blurred due to climate change effects.

Observed Variability in Influent Flowrates

The influent flowrate to a treatment facility, as noted previously, is dependent on factors such as the time of day, season, size and characteristics of the contributing population, and infiltration/inflow to and exfiltration from the collection system. In large cities the wastewater flow is distributed more evenly because there is a greater diversity of lifestyles and a high amount of activity at night. In contrast, wastewater treatment plants used for small residential communities are more likely to experience higher peak flow relative to mean flow values. Peaking factors, as discussed above, are used to estimate the maximum values that would be expected. One method used to characterize the variability of wastewater parameters and treatment processes is the use of the geometric standard deviation, s_g, as computed in Example 3–5. The value of s_g can be used to approximate an entire distribution of all expected values if a mean value is known or can be estimated. As discussed in

Figure 3–8

Identification of flowrate parameters based on variations observed on a typical weekday.



Appendix D, the greater the numerical value of s_g, the greater the observed range in the measured values.

Peaking factors are also related to s_g by specification of a frequency value. The peaking factor is calculated as the value at a given frequency divided by the mean value. For example, the peak day value, which corresponds to one event per year, is the value that occurs at a frequency of 99.7 percent [(364/365) × 100]. The typical ranges of observed values for s_g for influent flowrates for small, medium, and large capacity wastewater treatment plants are given in Table 3–12. The relationship between s_g values and the peaking factors for peak day, week, and month can be determined using the curves given on Fig. 3–9. An example of the use of the s_g value and the curves given on Fig. 3–9 is illustrated subsequently in Example 3–9.

Figure 3–9

Relationship of s_g values to peaking factors for peak day, week and month: (a) diagram for facilities with large peak to mean variations and (b) expanded portion of diagram (a) for facilities with small peak to mean variations.

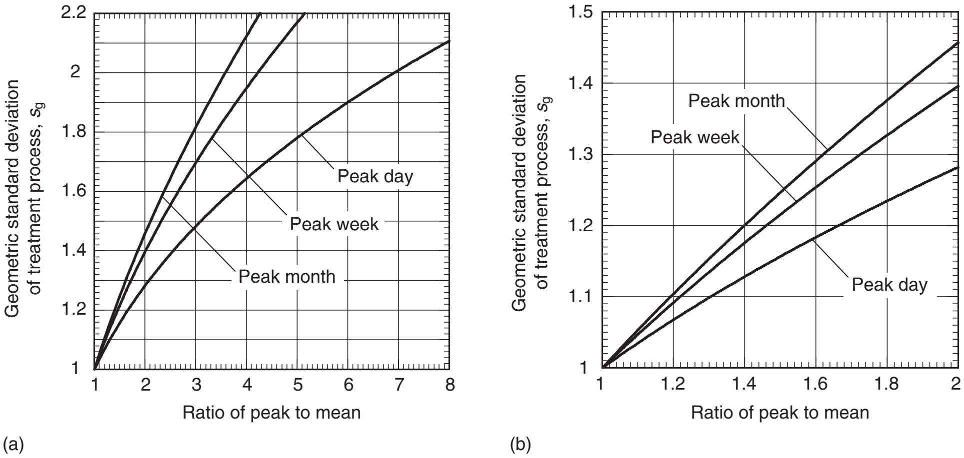


Table 3–12

Ranges of geometric standard deviations (s_g) for influent wastewater flowrates and selected constituents observed at small, intermediate, and large wastewater treatment facilities

	Ranges of s_g values for typical wastewater treatment facilities[a]					
	Small[b]		Medium[c]		Large[d]	
Parameter	Range	Typical	Range	Typical	Range	Typical
Flowrate	1.4–2.0	1.6	1.1–1.5	1.25	1.1–1.2	1.15
BOD	1.4–2.1	1.6	1.3–1.6	1.3	1.1–1.3	1.27
COD	1.5–2.2	1.7	1.4–1.8	1.4	1.1–1.5	1.30
TSS	1.4–2.1	1.6	1.3–1.6	1.3	1.1–1.3	1.27

[a] Excluding systems with large amounts of infiltration in the collection system.

[b] Flowrate of 4000–40,000 m^3/d (1–10 Mgal/d).

[c] Flowrate of 40,000–400,000 m^3/d (10–100 Mgal/d).

[d] Flowrate >400,000 m^3/d (>100 Mgal/d).

3–4 ANALYSIS OF WASTEWATER CONSTITUENTS

The complete analysis of wastewater data involves the determination of the flowrate, constituent concentrations, and constituent mass loading and their variations. The analysis may involve determining the concentrations of specific constituents, mass loadings (flowrate times concentration), or sustained mass loadings (loadings that occur over a defined period of time). From the standpoint of treatment processes, one of the most serious deficiencies results when the design of a treatment plant is based on average flowrates and average BOD and TSS loadings, with little or no recognition of peak conditions and other factors that may affect the average values.

In many communities, peak influent flowrates and BOD and TSS loading rates can reach two or more times average values. It must also be emphasized that, in nearly all cases, peak flowrates and BOD and TSS loading rates do not occur at the same time. Analysis of current records is the best method of arriving at appropriate peak and sustained mass loadings. The principal factors responsible for loading variations are (1) the established habits of community residents, which cause short-term (hourly, daily, and weekly) variations; (2) seasonal conditions, which usually cause longer-term variations; and (3) industrial activities, which cause both long- and short-term variations.

Wastewater Constituents Discharged By Individuals

The physical, chemical, and biological characteristics of wastewater vary throughout the day. An adequate determination of the waste characteristics will result only if the sample tested is representative. Typically, composite samples made up of portions of samples collected at regular intervals during a day are used (see Fig. 3–10, see also Fig. 2–2 in Chap 2). The amount of liquid collected for each sample is proportional to the flowrate at the time the sample was collected. Adequate characterization of mass of the various constituents discharged by individuals is of fundamental importance in the design of treatment and disposal processes.

In the past, it was quite common practice to report measured constituent concentrations, most commonly as mg/L. In the twenty-first century this practice is less useful because, as noted previously in the discussion of flowrates, the per capita flowrates are decreasing as a result of conservation and the use of water conserving appliances and fixtures. Use of current measured concentration values for future projections can be misleading because future concentration values will be understated, due to the impacts of

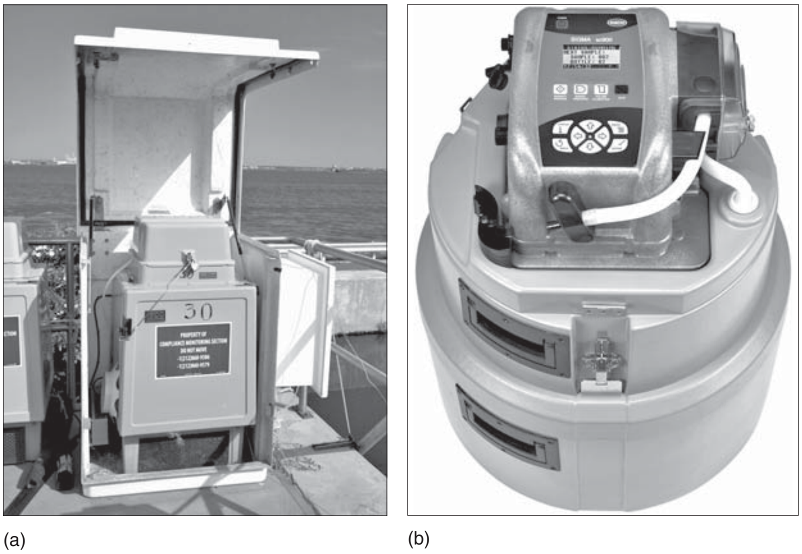


Figure 3–10

Samplers used to collect wastewater samples: (a) refrigerated, locked, and enclosed sampler used for the collection of treatment plant effluent compliance samples (see Fig. 2–2 for view of internal workings of sampler) and (b) portable sampler used to collect individual hourly samples throughout a day to assess process performance and to study potential benefits of process modifications (courtesy of Hach Company, www.hach.com). Portable samplers are also used to conduct collection system evaluations.

(a) (b)

continued water conservation. A more rational approach is to consider the constituent mass discharge rate on a per capita basis. With the availability of such data, the corresponding wastewater concentrations can be determined easily. In what follows typical constituent mass data are presented for the United States and for foreign countries.

Per Capita Mass Constituent Discharges in the United States. Typical data on the total mass of waste discharged per person per day (dry weight basis) from individual residences are reported in Table 3–13. The data presented in Table 3–13 have been gathered from numerous sources. The values reported in Table 3–13 represent the combined total for feces and urine. The reason that the range of values for the individual constituents is much greater than that observed in other countries (see Table 3–14) is due to the fact that about 37.5 percent of all adults in the United States are classified as obese (Ogden et al., 2012). The total number of pathogenic organisms discharged will depend on whether an individual is ill and is shedding pathogens. If one or more members of a family are ill and shedding pathogens, the number of measured organisms can increase by several orders of magnitude. The data given in Table 3–13 will be used subsequently to determine constituent concentrations based on various volumes of water. In the future, it is anticipated that constituent mass per capita data, as given in Table 3–13, will be used as a basis for determining expected constituent concentrations for the design of treatment plant upgrades as well as for new plants.

Per Capita Mass Constituent Discharges in Countries Outside of the United States. The amounts of waste discharged by individuals in other countries can vary significantly due to cultural and socioeconomic differences. Constituent data for 12 other countries as compared to the United States are reported in Table 3–14. The difference in the range of constituent values for other countries as compared to the United States is striking and, as described above, is largely cultural. Water use by individuals also differs significantly in other countries; in most cases the quantities used are significantly less. Consequently, the strength of the wastewater may be much higher than found in the United States and can affect significantly wastewater treatability. In some cases, the wastewater composition may be high in organic content but low in alkalinity and thus cannot be nitrified fully. Concentrations of TSS and BOD for another culture compared to the United States are illustrated in Example 3–5.

Table 3–13

Quantity of waste discharged by individuals on a dry weight basis in the United States[a]

	Value, lb/capita·d			Value, g/capita·d		
Constituent (1)	**Range (2)**	**Typical without ground up kitchen waste (3)**	**Typical with ground up kitchen waste (4)**	**Range (5)**	**Typical without ground up kitchen waste (6)**	**Typical with ground up kitchen waste (7)**
BOD_5	0.11–0.26	0.15	0.20	50–120	70	93
COD	0.30–0.65	0.40	0.50	110–295	180	230
TSS	0.13–0.33	0.15	0.19	60–150	70	87
NH_3 as N	0.011–0.026	0.017	0.017	5–12	7.6	7.9
Organic N as N	0.009–0.022	0.012	0.013	4–10	5.4	6.0
TKN[b,c] as N	0.020–0.040	0.029	0.031	9–18	13	13.9
Organic P as P	0.002–0.004	0.0026	0.0029	0.9–1.8	1.2	1.3
Inorganic P as P	0.001–0.006	0.0020	0.0020	0.50–2.7	0.90	0.90
Total P[c] as P	0.003–0.010	0.0046	0.0048	1.5–4.5	2.1	2.2
Potassium, K[c]	0.009–0.015	0.013	0.014	4–7	6.0	6.2
Oil and grease	0.022–0.077	0.062	0.070	10–35	28.0	32

[a] Adapted in part from Tchobanoglous et al. (2003).

[b] TKN is total Kjeldahl nitrogen.

[c] The percentage distribution of the total nitrogen, phosphorus, and potassium in feces and urine is 14 to 17, 28 to 35, and 12 to 18 percent, respectively.

Table 3–14

Typical wastewater constituent data for various countries[a]

Country/ constituent	BOD, g/capita·d	TSS, g/capita·d	TKN, g/capita·d	NH_3-N, g/capita·d	Total P, g/capita·d
Brazil	55–68	55–68	8–14	ND	0.6–1
Denmark	55–68	82–96	14–19	ND	1.5–2
Egypt	27–41	41–68	8–14	ND	0.4–0.6
Germany	55–68	82–96	11–16	ND	1.2–1.6
Greece	55–60	ND	ND	8–10	1.2–1.5
India	27–41	ND	ND	ND	ND
Italy	49–60	55–82	8–14	ND	0.6–1
Japan	40–45	ND	1–3	ND	0.15–0.4
Palestine[b]	32–68	52–72	4–7	3–5	0.4–0.7
Sweden	68–82	82–96	11–16	ND	0.8–1.2
Turkey	27–50	41–68	8–14	9–11	0.4–2
Uganda	55–68	41–55	8–14	ND	0.4–0.6
United States[c]	50–120	60–150	9–18	5–12	1.5–4.5

[a] Adapted from Tchobanoglous et al. (2003).

[b] West Bank and Gaza Strip.

[c] From Table 3–13.

Characteristics of Urine. Because there is increased interest in urine separation and its implications for wastewater treatment and recovery for use as a fertilizer, typical composition data for urine are reported in Table 3–15. The values in the table are identified as approximate because the composition of urine will vary for each individual depending on the amount of water ingested, the amount and composition of the food ingested, the time of day, general health, blood pressure, and the temperature. As noted earlier the bulk of the nutrients (i.e., nitrogen, phosphorus, and potassium) in wastewater are present in urine.

Table 3–15

Approximate properties and overall composition of urine[a,b]

		Value, mg/L[c]	
Constituent	**Formula**	**Range**	**Typical**
Organic molecules			
Urea	CON_2H_4	9000–23,000	20,000
Creatinine	$C_4H_9N_3O_2$	900–1200	1000
Uric acid	$C_5H_4N_4O_3$	200–400	300
Trace organics			
Inorganic elements			
Ammonium	NH_4^+	400–600	500
Bicarbonate	HCO_3^-	20–600	300
Calcium	Ca^{2+}	100–300	150
Chloride	Cl^-	1600–8000	1900
Magnesium	Mg^{2+}	80–120	100
Potassium	K^+	1200–1700	1500
Sodium	Na^+	5000–7000	6000
Inorganic compounds			
Sulphates	SO_4^{2-}	1600–2000	1800
Phosphates	$H_2PO_4^-$, HPO_4^{2-} PO_4^{3-}	1000–1500	1200
Overall composition			
Urea	CON_2H_4	16,000–24,000	20,000
Organic compounds	various	4000–8000	6000
Organic ammonium salts	various	4000–6000	5000
Inorganic salts	various	12,000–16,000	15,000
Other	various	2500–6000	4000
Solids	%	4–7	5
Water	%	93–96	95
Density	g/mL	1.002–1.030	1.010

[a] Adapted in part from Putnam (1971); Ryan (1966); Gotaas (1956); and numerous other sources.

[b] Values in the table are identified as approximate because the composition of urine will vary for each individual depending on the amount of water ingested, the amount and composition of the food ingested, the time of day, general health, blood pressure, and the temperature.

[c] Per capita amount of urine discharged per day will vary from 0.8 to 1.3 L/capita·d.

Constituent Concentrations Based on Individual Mass Discharges

The corresponding constituent concentrations for the United States, assuming the per capita mass quantities of waste constituents given in Table 3–13, are reported in Table 3–16. In preparing Table 3–16, it was assumed that (1) 25 percent of the homes were equipped with kitchen food waste grinders and (2) the constituent mass quantities were diluted in 190 and 460 L (50 and 120 gal) of water. Two different dilutions are used to illustrate the impact that the dilution has on the resulting constituent concentrations.

The method used to determine the constituent concentration values is illustrated in Example 3–5.

EXAMPLE 3–5 **Estimate Waste Constituent Concentration** Using data from Table 3–14, determine the BOD, TSS, and ammonia nitrogen concentrations for the West Bank and the Gaza Strip assuming the water supply is intermittent and the wastewater flowrate is 60 L/capita·d.

Solution

1. From Table 3–15, use the following average constituent contributions:
   a. BOD = 50 g/capita·d
   b. TSS = 62 g/capita·d
   c. NH_3-N = 4 g/capita·d
2. Compute BOD concentration.

$$\text{BOD} = \left[\frac{(50\ \text{g/capita·d})}{(60\ \text{L/capita·d})}\right]\left(\frac{10^3\ \text{L}}{1\ \text{m}^3}\right) = 833\ \text{g/m}^3$$

3. Compute TSS concentration.

$$\text{TSS} = \left[\frac{(62\ \text{g/capita·d})}{(60\ \text{L/capita·d})}\right]\left(\frac{10^3\ \text{L}}{1\ \text{m}^3}\right) = 1033\ \text{g/m}^3$$

4. Compute the NH_3-N concentration.

$$\text{NH}_3\text{-N} = \left[\frac{(4\ \text{g/capita·d})}{(60\ \text{L/capita·d})}\right]\left(\frac{10^3\ \text{L}}{1\ \text{m}^3}\right) = 66.7\ \text{g/m}^3$$

Comment In many parts of the world where water usage is low, constituent concentrations for BOD and TSS may range up to 1000 g/m^3 (mg/L). In the above example, the concentrations of BOD and TSS are nearly 2 to 4 times the BOD and TSS concentrations found typically in the United States (see Table 3–16). Ammonia nitrogen (NH_3-N) concentrations may also be higher, perhaps on the order of two times or more of those in the United States.

Mineral Increase Resulting from Water Use

Data on the increase in the mineral content of wastewater resulting from water use and the variation of the increase within a collection system are especially important in evaluating the reuse potential of wastewater. Typical data on the incremental increase in mineral content that can be expected in municipal wastewater resulting from domestic use are reported in Table 3–17. Increases in the mineral content of wastewater may be due in part from addition of highly mineralized water from private wells and groundwater and from

Table 3–16

Typical unit loading factors and expected wastewater constituent concentrations from individual residences in the United States[a]

Constituent	Typical value[b], g/capita·d	Concentration, mg/L Volume, L/capita·d (gal/capita·d) 190(50)	380(100)
BOD_5	76.0	399.0	199.0
COD	193.0	1013.0	507.0
TSS	74.0	391.0	195.0
NH_3 as N	7.7	40.0	20.0
Org. N as N	5.5	29.0	14.0
TKN as N	13.2	70.0	35.0
Org. P as P	1.2	6.4	3.2
Inorg. P as P	0.9	4.7	2.4
Total P as P	2.1	11.0	5.6
Potassium	6.1	32.0	16.0
Oil and grease	29.0	153.0	76.0

[a] Adapted from Tchobanoglous et al. (2003).
[b] Data from Table 3–13, Columns 6 and 7, assuming 25 percent of the homes have kitchen waste food grinders. For example, $BOD_5 = [70 + (93 - 70)(0.25)]$ mg/L = 76 mg/L.

industrial use. Domestic and industrial water softeners also contribute significantly to the increase in mineral content and, in some areas, may represent the major source. Occasionally, water added from private wells and groundwater infiltration will (because of its high quality) serve to dilute the mineral concentration in wastewater.

Composition of Wastewater in Collection Systems

Typical data on the composition of untreated domestic wastewater as found in wastewater collection systems (in the United States) are reported in Table 3–18. The data presented in Table 3–18 for medium strength wastewater are based on an average flow of 460 L/capita·d (120 gal/capita·d) and include constituents added by commercial, institutional, and industrial sources. Typical concentrations for low-strength and high strength wastewater, which reflect different amounts of infiltration, are also given. Because there is no "typical" wastewater, it must be emphasized that the typical data presented in Table 3–18 *should only be used as a guide*.

Variations in Constituent Concentrations

Several types of constituent concentrations can occur depending upon the characteristics of the contributors to the wastewater collection system.

Short-Term Variation in Constituent Concentrations. Constituent concentrations variations change significantly during the course of a day. An example of typical

Table 3–17

Typical mineral increase from domestic water use

Constituent	Typical increment range[a, b] Mass, g/capita·d	Concentration, mg/L[c]
Anions:		
Bicarbonate (HCO_3)	23–46	60–121
Carbonate (CO_3)	0–5	0–13
Chloride (Cl)	9–23	24–60
Sulfate (SO_4)	7–14	18–37
Cations:		
Calcium (Ca)	3–7	8–18
Magnesium (Mg)	2–5	4–13
Sodium (Na)	18–32	47–84
Other constituents		
Aluminum (Al)	0.04–0.09	0.11–0.24
Boron (B)	0.04–0.09	0.11–0.24
Fluoride (F)	0.09–0.2	0.24–0.53
Manganese (Mn)	0.09–0.2	0.24–0.53
Silica (SiO_2)	0.9–5	2.4–13
Total alkalinity (as $CaCO_3$)	28–55	74–145
Total dissolved solids (TDS)	69–175	182–460

[a] Values do not include commercial and industrial additions.
[b] Excluding the addition from domestic water softeners.
[c] Based on 380 L/capita·d (100 gal/capita·d) as used in Tables 3–16 and 3–18.

Note: mg/L = g/m^3.

variations in BOD and TSS in domestic wastewater are shown on Fig. 3–11. In general, the variations in BOD and TSS follow the flowrate variation. The peak BOD (organic matter) concentration typically occurs in the evening, whereas the peak TSS concentration occurs in the morning. The amplitude of the BOD and TSS variations depends on the size and characteristics of the community. Also, the time that the peak value occurs will depend on whether it is a weekday or the weekend [see Fig. 3–3(b)].

Seasonal Variation in Constituent Concentrations. For domestic flow only, and neglecting the effects of infiltration, the unit (per capita) loadings and the strength of the wastewater from most seasonal sources, such as resorts, will remain about the same on a daily basis throughout the year even though the total flowrate varies. The total mass of BOD and TSS of the wastewater, however, will increase directly with the population served.

Infiltration/inflow, as discussed earlier in this chapter, is another source of water flow into the collection system. In most cases, the presence of this extraneous water tends to decrease the concentrations of BOD and TSS, depending on the characteristics of the water entering the sewer. In some cases, concentrations of some inorganic constituents may actually increase where the groundwater contains high levels of dissolved constituents.

Table 3–18

Typical composition of untreated domestic wastewater

		Concentration[b]		
Constituent	Unit	Low strength	Medium strength	High strength
Solids, total (TS)	mg/L	537	806	1612
Dissolved, total (TDS)	mg/L	374	560	1121
Fixed	mg/L	224	336	672
Volatile	mg/L	150	225	449
Suspended solids, total (TSS)	mg/L	130	195	389
Fixed	mg/L	29	43	86
Volatile	mg/L	101	152	304
Settleable solids	mL/L	8	12	23
Biochemical oxygen demand, 5-d, 20°C BOD	mg/L	133	200	400
Total organic carbon (TOC)	mg/L	109	164	328
Chemical oxygen demand (COD)	mg/L	339	508	1016
Nitrogen (total as N)	mg/L	23	35	69
Organic	mg/L	10	14	29
Free ammonia	mg/L	14	20	41
Nitrites	mg/L	0	0	0
Nitrates	mg/L	0	0	0
Phosphorus (total as P)	mg/L	3.7	5.6	11.0
Organic	mg/L	2.1	3.2	6.3
Inorganic	mg/L	1.6	2.4	4.7
Potassium	mg/L	11	16	32
Chlorides[c]	mg/L	39	59	118
Sulfate[c]	mg/L	24	36	72
Oil and grease	mg/L	51	76	153
Volatile organic compounds (VOCs)	μg/L	<100	100–400	>400
Total coliform	No./100 mL	10^6–10^8	10^7–10^9	10^7–10^{10}
Fecal coliform	No./100 mL	10^3–10^5	10^4–10^6	10^5–10^8
Cryptosporidium oocysts	No./100 mL	10^{-1}–10^1	10^{-1}–10^2	10^{-1}–10^3
Giardia lamblia cysts	No./100 mL	10^{-1}–10^2	10^{-1}–10^3	10^{-1}–10^4

[a] Adapted from Tchobanoglous et al. (2003).

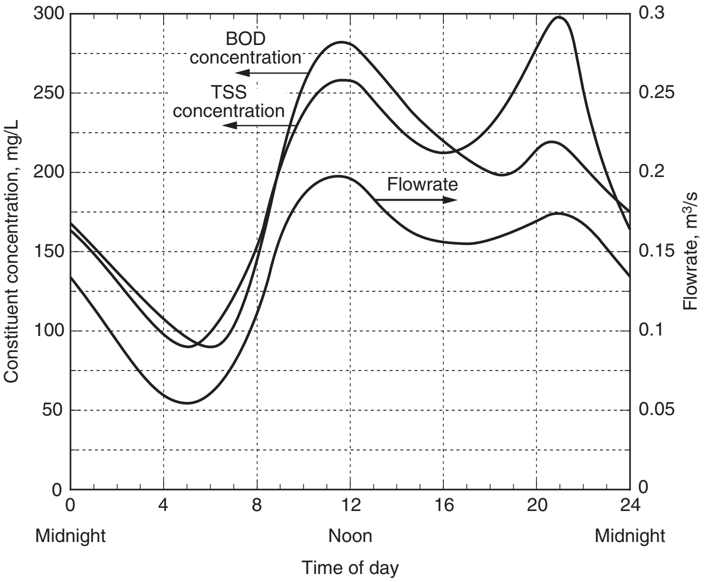
[b] Low strength is based on an approximate wastewater flowrate of 570 L/capita·d (150 gal/capita·d). Medium strength is based on an approximate wastewater flowrate of 380 L/capita·d (100 gal/capita·d). High strength is based on an approximate wastewater flowrate of 190 L/capita·d (50 gal/capita·d).

[c] Values should be increased by amount of constituent present in domestic water supply.

Note: mg/L = g/m^3.

Figure 3–11

Typical hourly variations observed in flow and BOD and TSS concentrations in domestic wastewater.



Variations in Industrial Wastewater. The composition of wastewater from industrial operations varies widely depending on the function and activity of the particular industry. Examples of the constituent concentration variability are illustrated in Table 3–19. From these examples, it can be observed that flow values and water quality measurements may vary by several orders of magnitude over a period of a year. Because of this variation, it is often difficult to define "typical operating conditions" for industrial activities.

The concentrations of both BOD and TSS, as well as a wide range of other constituents, in industrial wastewater can vary significantly throughout the day. For example, the BOD and TSS concentrations contributed from vegetable-processing facilities during the noon wash-up period may far exceed those contributed during working hours. Problems with high short-term loadings most commonly occur in small treatment plants that have limited reserve capacity to handle these so-called "shock loadings." The seasonal impact of industrial wastes such as canneries can cause both the flow and BOD loading to increase from two to five times average conditions.

If industrial wastes are to be discharged to the collection system for treatment in a municipal wastewater facility, it will be necessary to characterize the wastes adequately to identify the ranges in constituent concentrations and mass loadings. Such characterization is also needed to determine if pretreatment is required before the waste is permitted to be discharged into the collection system. If pretreatment is needed, the effluent from the pretreatment facilities must also be characterized. Further, any proposed future process changes should also be assessed to determine what effects they might have on the wastes to be discharged. Where data are not available, every effort should be made to obtain information from similar facilities. With sufficient characterization of the wastewater from industrial discharges, suitable pretreatment facilities can be provided and plant upsets can be avoided.

Variations in Constituent Concentrations in Combined Collection Systems. Flowrates, constituent concentrations, and mass loads emanating from combined collection systems can vary widely from community to community, season to season,

Table 3–19

Typical examples of the range of effluent flowrates and constituent concentrations for two industrial activities[a]

		Wool textile mill		Tomato cannery	
Constituent	Unit	Annual average	Daily maximum	Peak season[b]	Off season[c]
Flowrate	m^3/d	—	—	4164–22,300	1140–6400
pH	—	5.92[d]	—	7.2–8.0	7.2–8.0
BOD	mg/L	90.7	169	460–1100	29–56
COD	mg/L	529	1240	—	—
SS	mL/L	—	—	6–80	0.5–2.2
TSS	mg/L	93.4	860	270–760	69–120
TDS	mg/L	—	—	480–640	360–520
Nitrate-N	mg/L	—	—	0.4–5.6	0.1–2.2
Ammonia-N	mg/L	8.1	54	—	—
Phosphorus	mg/L	—	—	1.5–7.4	0.3–3.9
Sulfate	mg/L	—	—	15–23	7.1–9.9
DO	mg/L	—	—	0.9–3.8	1.6–9.8
Oil and grease	mg/L	27.4	45.2	—	—
Temperature	°C	—	—	18–23	13–19

[a] Adapted from Tchobanoglous et al. (2003).

[b] Peak season is from early July to late September, when fresh-harvest tomatoes are canned. Treatment consists of screening and sedimentation.

[c] Off-season is from November to June, when canned tomatoes are remanufactured into tomato paste, tomato sauce, and other tomato products (e.g., salsa, ketchup, spaghetti sauce). Treatment typically consists of screening, aeration, and sedimentation.

[d] Median value.

Note: $m^3/d \times 0.264 \times 10^{-3} = Mgal/d$.

storm to storm and within a storm. Typical factors influencing the characteristics of wastewater from combined collection systems are shown in Table 3–20. Example variations of BOD, TSS, and fecal coliform measured in a combined system are shown on Fig. 3–12, during and after a storm event. As shown, the BOD and fecal coliform bacteria concentrations are low during the storm when runoff flows are high. After the storm, when runoff subsides and the flow consists primarily of wastewater, concentrations rise significantly. When this rise occurs, it can be concluded that the BOD and fecal coliform concentrations in the stormwater are significantly lower than in the wastewater component. The characteristics of combined wastewater and stormwater as compared to municipal wastewater are shown in Table 3–21.

Unlike BOD and fecal coliform bacteria, TSS concentrations rise slightly during the storm and remain unchanged after the storm, indicating that TSS concentrations from stormwater runoff and wastewater are similar. The slight rise in the TSS concentration during the peak flow may be due to a phenomenon common to many combined sewer systems known as the “first flush.” The first flush has often been observed following the initial phase of a rainfall event in which much of the accumulated surface contaminants are washed into the combined system. In combined collection systems, the increased flows may be capable of resuspending material deposited previously during low flow periods.

Table 3–20

Typical factors influencing the characteristics of combined wastewater

Parameter	Quantity-related factors	Quality-related factors
Precipitation	Rainfall depth (volume) Storm intensity Storm duration	Antecedent dry weather Atmospheric quality
Wastewater sources	Peak flowrate and variability Land use	Contributing sources (domestic, industrial, commercial, etc.)
Drainage basin characteristics	Size, shape, slope, time of concentration Land use Impervious area Soil characteristics Stormwater management practices	Effectiveness of best management practices/minimum controls with respect to street cleaning and litter control Pollutant buildup and wash-off Stormwater management practices
Sewer system, interceptor design and condition	Pipe size, slope and shape Quantity of base infiltration Number and location of inflow sources (catch basins, roof leaders, etc.) Surcharging or backwater conditions Type of flow regulation or diversion structures including pumping station overflows Capacity reduction from sediment build-up River or tidal inflow	Effectiveness of best management practices/minimum controls with respect to catch basin and sewer cleaning Pollutant buildup and resuspension Chemical and biological transformations

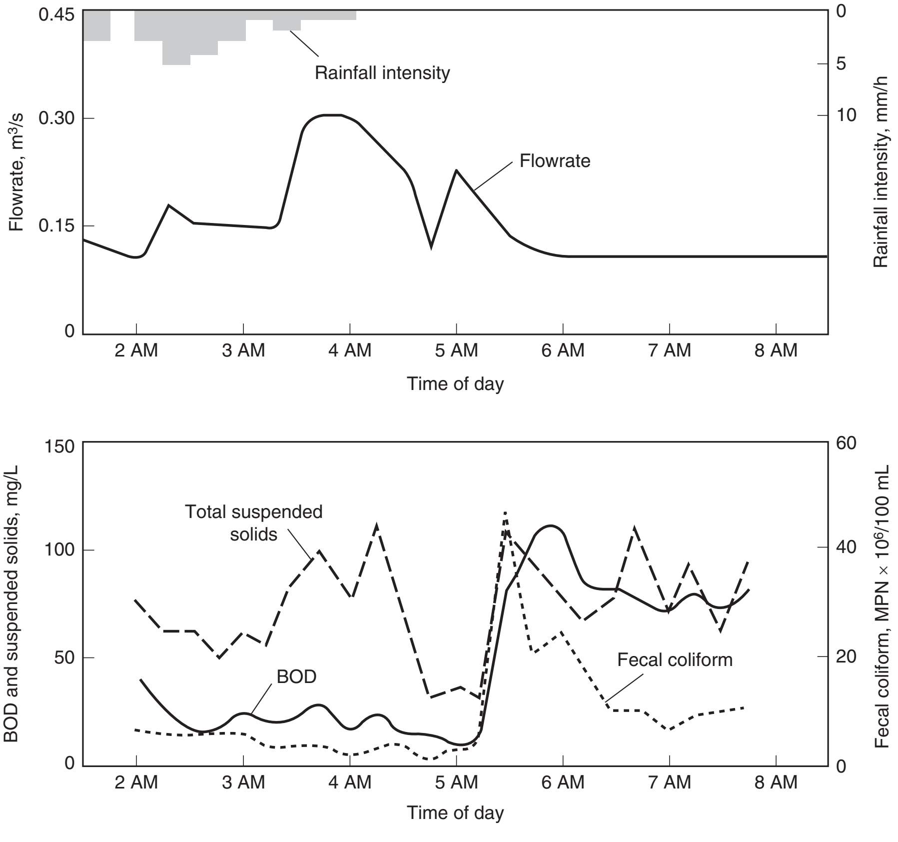


Figure 3–12

Typical variations of flowrate, BOD, TSS, and fecal coliform in a combined collection system during a storm event.

Table 3–21

Comparison of characteristics of combined wastewater with other sources[a]

Parameter	Unit	Rainfall	Stormwater runoff	Combined wastewater	Municipal wastewater
Total suspended solids, TSS	mg/L	<1	67–101	270–550	120–400
Biochemical oxygen demand, BOD	mg/L	1–13	8–10	60–220	110–350
Chemical oxygen demand, COD	mg/L	9–16	40–73	260–480	250–800
Nitrogen (as N)					
Total Kjeldahl nitrogen, TKN	mg/L		0.40–1.00	4–17	20–70
Nitrate, NO_3	mg/L		0.05–1.0	0.48–0.91	0
Phosphorus (total as P)	mg/L	0.02–0.15	0.67–1.66	1.2–2.8	4–12
Metals					
Copper, Cu	μg/L		27–33		
Lead, Pb	μg/L	30–70	30–144	140–600	
Zinc, Zn	μg/L		135–226		
Fecal coliform bacteria	MPN/100 mL		10^3–10^4	10^5–10^6	10^5–10^8

[a] Adapted from Tchobanoglous et al. (2003).

Together, the resuspended material and contaminants washed off surfaces result in high contaminant concentrations. Factors known to contribute to the magnitude and frequency of the first flush effect include combined sewer slopes, street and catch basin cleaning frequency and design, rainfall intensity, duration, and surface buildup of debris and contaminants.

Wastewater from combined collection systems usually contains more inorganic matter than wastewater from sanitary collection systems because of the larger quantities of stormwater runoff that enter the combined sewer system. This observation is particularly true in snowbelt states where sand is used to treat the streets during snow and ice conditions.

Statistical Analysis of Constituent Concentrations

The statistical analysis of constituent concentration data is essentially the same as described previously for flowrate data in Sec. 3–3 and illustrated in Example 3–4. The most common type of probability distribution encountered in the statistical analysis of constituent concentration data is the log-normal distribution, in which log of the constituent concentration is normally distributed. In general, wastewater constituents that can be altered biologically are distributed log-normally. Inorganic constituents, such as chloride and sulfate, can be distributed both normally or log-normally.

Observed Variability in Influent Constituent Concentrations

The influent constituent concentrations to a treatment facility, as noted above, and illustrated on Fig. 3–12, are also quite variable depending on the time of day, season, size and characteristics of the contributing population, and infiltration to and exfiltration from the collection system. The variability observed in constituent concentrations can be quantified using the geometric standard deviation, s_g, as computed in Example 3–4. The value of s_g can be used to approximate an entire distribution of all expected values if a mean value is known or can be estimated. For example, the peak day value, which corresponds to one event per year, is the value that occurs at a frequency of 99.7 percent [(364/365) × 100]. The typical range of observed values for s_g for influent BOD, TSS, and COD concentrations for small-, medium-, and large-capacity wastewater treatment plants are given in Table 3–12, presented previously. The application of the s_g values given in Table 3–12 is illustrated subsequently in Example 3–9.

3–5 ANALYSIS OF CONSTITUENT MASS LOADING DATA

The analysis of wastewater constituent data involves the determination of simple average or flow-weighted average concentrations, mass loadings rates, and sustained peak mass loading rates.

Simple Average

The simple or arithmetic average of a number of individual measurements is given by

$$\bar{x} = \frac{1}{n}\sum_{i=1}^{n} x_i \tag{3–2}$$

Where $\bar{x}$ = arithmetic average of individual measurements
n = number of measurements
x_i = average value of measurement during the ith time period

To analyze the BOD and TSS data given on Fig. 3–6, for example, the usual procedure is to divide the day's record into 24 1-h increments, sum the 24 individual average hourly values, and divide by 24. Although arithmetic averages are still used, they are of little value because the magnitude of the flow at the time of the measurement is not taken into account. If the flowrate remains constant, the use of a simple average is acceptable.

Flow-Weighted Average

Flow-weighted constituent concentrations are obtained by multiplying the flow (typically hourly values over a 24-h period) times the corresponding constituent concentration, summing the results, and dividing by the summation of the flows as given by Eq. 3–3.

$$C_W = \frac{\sum_{i=1}^{n} q_i C_i}{\sum_{i=1}^{n} q_i} \tag{3–3}$$

where C_w = flow-weighted average constituent concentration
n = number of observations
q_i = average flowrate during ith time period
C_i = average concentration of the constituent during ith time period

Whenever possible, flow-weighted constituent concentrations should be used because they are a more accurate representation of the actual wastewater strength that must be treated.

Determination of the simple arithmetic average and flow-weighted constituent concentrations is illustrated in Example 3–6.

EXAMPLE 3–6 **Calculation of Flow-Weighted BOD and TSS Concentrations** Compute the flow-weighted BOD and TSS values using the data provided below for a community of about 5000 persons. Compare the flow-weighted values to the simple arithmetic averages. What is the significance of the difference?

Time	Flowrate, m^3/s	BOD, g/m^3	TSS, g/m^3	Time	Flowrate, m^3/s	BOD, g/m^3	TSS, g/m^3
Mid	0.120	165	175	Noon	0.195	255	280
1 a.m.	0.115	150	155	1 p.m.	0.180	242	265
2	0.095	130	135	2	0.170	229	265
3	0.075	110	120	3	0.164	230	235
4	0.060	100	110	4	0.160	212	222
5	0.055	90	100	5	0.158	217	210
6	0.060	100	90	6	0.159	234	200
7	0.085	120	110	7	0.163	250	200
8	0.115	150	150	8	0.169	270	212
9	0.160	190	210	9	1.174	295	216
10	0.187	238	253	10	0.164	250	203
11	0.195	256	273	11	0.155	199	189
12	0.195	255	280	12	0.135	165	175

Solution

1. Create a spreadsheet for calculating the flow-weighted values with columns for time, flowrate, BOD, TSS, q × BOD, and q × TSS. Enter the time intervals (e.g., Noon–1 p.m.) in column 1. The required table is presented.
2. For each time period, calculate the average flowrate value during the interval and enter the value in column 2. For example, the average flowrate value during the first interval (12–1 a.m.) is

   Value at beginning of interval = 0.120 m^3/s
   Value at end of interval = 0.115 m^3/s

   $$\text{Average flowrate} = \frac{(0.120\ \text{m}^3/\text{s} + 0.115\ \text{m}^3/\text{s})}{2} = 0.118\ \text{m}^3/\text{s}$$

   The average flowrate values for each successive time interval are entered in column 2.
3. Enter the average values for BOD and TSS in columns 3 and 4, respectively.
4. For each time period, multiply the average flowrate value (column 2) times the average BOD (column 3), and enter the results in column 5.
5. For each time period, multiply the average flowrate (column 2) value by the average TSS (column 4), and enter the results in column 6.
6. Calculate the sum and simple arithmetic average for columns 2 through 6.

7. Divide the sum of the values in columns 5 and 6 (q × BOD and q × TSS, respectively) by the sum of the values in column 2 (flowrate) to obtain the flow-weighted average for BOD and TSS. The resulting values are given in the last two lines of the spreadsheet.

Time interval (1)	Flowrate, q m^3/s (2)	BOD, g/m^3 (3)	TSS, g/m^3 (4)	q × BOD, kg/d (5) = (2) × (3)	q × TSS, kg/d (6) = (2) × (4)
12–1 a.m.	0.118	157.5	165.0	18.59	19.47
1–2	0.105	140.0	145.0	14.70	15.23
2–3	0.085	120.0	127.5	10.20	10.84
3–4	0.068	105.0	115.0	7.14	7.82
4–5	0.058	95.0	105.0	5.51	6.09
5–6	0.058	95.0	95.0	5.51	5.51
6–7	0.073	110.0	100.0	8.03	7.30
7–8	0.100	135.0	130.0	13.50	13.00
8–9	0.138	170.0	180.0	23.46	24.84
9–10	0.174	214.0	231.5	37.24	40.28
10–11	0.191	247.0	263.0	47.18	50.23
11–12	0.195	255.5	276.5	49.82	53.92
12–1 p.m.	0.188	248.5	272.5	46.72	51.23
1–2	0.175	235.5	265.0	41.21	46.38
2–3	0.167	229.5	250.0	38.33	41.75
3–4	0.162	221.0	228.5	35.80	37.02
4–5	0.159	214.5	216.0	34.11	34.34
5–6	0.159	225.5	205.0	35.85	32.60
6–7	0.161	242.0	200.0	38.96	32.20
7–8	0.166	260.0	206.0	43.16	34.20
8–9	0.172	282.5	214.0	48.59	36.81
9–10	0.169	272.5	209.5	46.05	35.41
10–11	0.160	224.5	196.0	35.92	31.36
11–12	0.145	182.0	182.0	26.39	26.39
Totals	3.346	4682	4578	711.96	694.19
Average	0.139	195.1	190.8		
Flow-weighted concentration values				212.8[a]	207.5

[a] $C_W = \sum_{i=1}^{n} q_i C_i / \sum_{i=1}^{n} q_i = 711.96/3.346 = 212.8.$

Comment When comparing the computation of a simple average to a flow-weighted value, the differences can be significant. In this example, if simple averages were used, the BOD loading would have been understated by 17.7 mg/L (8.3 percent), and the TSS loading by 16.7 mg/L (8.1 percent). If simple averages had been used in establishing process loading values, the treatment facilities could be under designed by about 8 percent in this case.

Mass Loading Rates

Constituent mass loading rates are usually expressed in kilograms per day (kg/d)[pounds per day (lb/d)] and may be computed using Eq. (3–4a) when the flowrate is expressed in cubic meters per day (m^3/d), or Eq. (3–4b) when the flowrate is expressed in million gallons per day (Mgal/d). Note that in the SI system of units, the concentration expressed in milligrams per liter (mg/L) is equivalent to grams per cubic meter (g/m^3).

$$\text{Mass loading rates, kg/d} = \frac{(\text{concentration, g/m}^3)\ (\text{flowrate, m}^3\text{/d})}{(10^3\ \text{g/1 kg})} \tag{3–4a}$$

$$\text{Mass loading rates, lb/d} = (\text{concentration, mg/L})(\text{flowrate, Mgal/d})\left[\frac{8.34\ \text{lb}}{\text{Mgal}\cdot(\text{mg/L})}\right] \tag{3–4b}$$

To design treatment processes to function properly under varying loading conditions, data must be available for the sustained peak mass loading rates of constituents that are to be expected. In the past, such information has seldom been available. When the data are not available, curves similar to those shown on Fig. 3–13 can be used. The curves for BOD, TSS, TKN (total Kjeldahl nitrogen), NH_3 (ammonia), and phosphorus were derived from an analysis of the records of over 50 treatment plants throughout the country. It should be noted that significant variations will be observed from plant to plant, depending on the size of the system, the percentage of combined wastewater, the size and slope of the interceptors, and the types of wastewater contributors.

The procedure used to develop the mass loading rate curves shown on Fig. 3–13 is as follows. First, the average mass loading rate is determined for the period of record. Second, the records are reviewed for the highest and lowest sustained one-day mass loading rate. These values are divided by the average mass loading rate and the numbers are plotted. Third, the same procedures are followed for two consecutive days, three consecutive days, etc., sustained loadings until ratio values are found for the period of interest (usually 10 to 30 d).

The daily mass loading rates for the various plants were developed using hourly data and the following expression:

$$\text{Daily mass loading rate, kg/d} = \sum_{i=1}^{24} \frac{(\text{concentration, g/m}^3)(\text{flowrate, m}^3\text{/d})}{(10^3\ \text{g/1 kg})} \tag{3–5a}$$

$$\text{Daily mass loading rate, lb/d} = \sum_{i=1}^{24} (\text{concentration, mg/L})(\text{flowrate, Mgal/d})\left[\frac{8.34\ \text{lb}}{\text{Mgal}\cdot(\text{mg/L})}\right] \tag{3–5b}$$

The development of a sustained peak mass loading curve is illustrated in Example 3–7.

EXAMPLE 3–7 **Development of BOD Sustained Mass Loading Rate Values** Develop a sustained BOD peak mass loading curve for a treatment plant with a design flowrate of 1 m^3/s (22.8 Mgal/d) and using sustained peak loading rate factors from Fig. 3–13(a). Assume that the long-term daily average BOD concentration is 200 g/m^3.

Solution

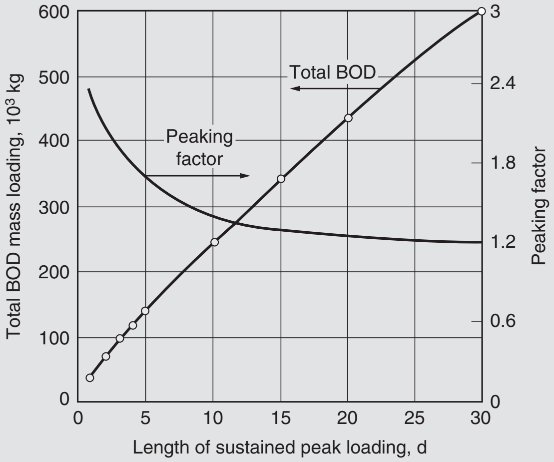
1. Compute the daily mass loading rate value for BOD.

$$\text{Daily BOD mass loading rate, kg/d} = \frac{(200\ \text{g/m}^3)(1\ \text{m}^3\text{/s})(86{,}400\ \text{s/d})}{(10^3\ \text{g/1 kg})} = 17{,}280\ \text{kg/d}$$

2. Set up a computation table for the development of the necessary information for the peak sustained BOD mass loading curve (see following table).
3. Obtain the sustained peak BOD loading rate factors from Fig. 3–13(a), and determine the sustained mass loading rates for various time periods (see table, cols. (1), (2), and (3)).
4. Develop data for the sustained mass loading curve and prepare a plot of the resulting data (see following figure).

Length of sustained peak, d (1)	Peaking factor[a] (2)	Peak BOD mass loading rate, kg/d (3)	Total mass loading, kg[b] (4)
1	2.4	41,472[c]	41,472
2	2.1	36,288	72,576
3	1.9	32,832	98,496
4	1.8	31,104	124,416
5	1.7	29,376	146,880
10	1.4	24,192	241,920
15	1.3	22,464	336,960
20	1.25	21,600	432,000
30	1.21	19,872	596,160
365	1.0	17,280	

[a] From Fig. 3–13a.
[b] Col. 1 × col. 3 = col. 4.
[c] 41,472 = 17,280 × 2.4.



Comment The interpretation of the curve plotted for this example is as follows. If the sustained peak loading period were to last for 10 d, the total amount of BOD that would be received at a treatment facility during the 10-d period would be 241,695 kg. The corresponding amounts for sustained peak periods of 1 and 2 d would be 41,401 and 72,451 kg, respectively. Computations for an example of this type are normally done using a spreadsheet program.

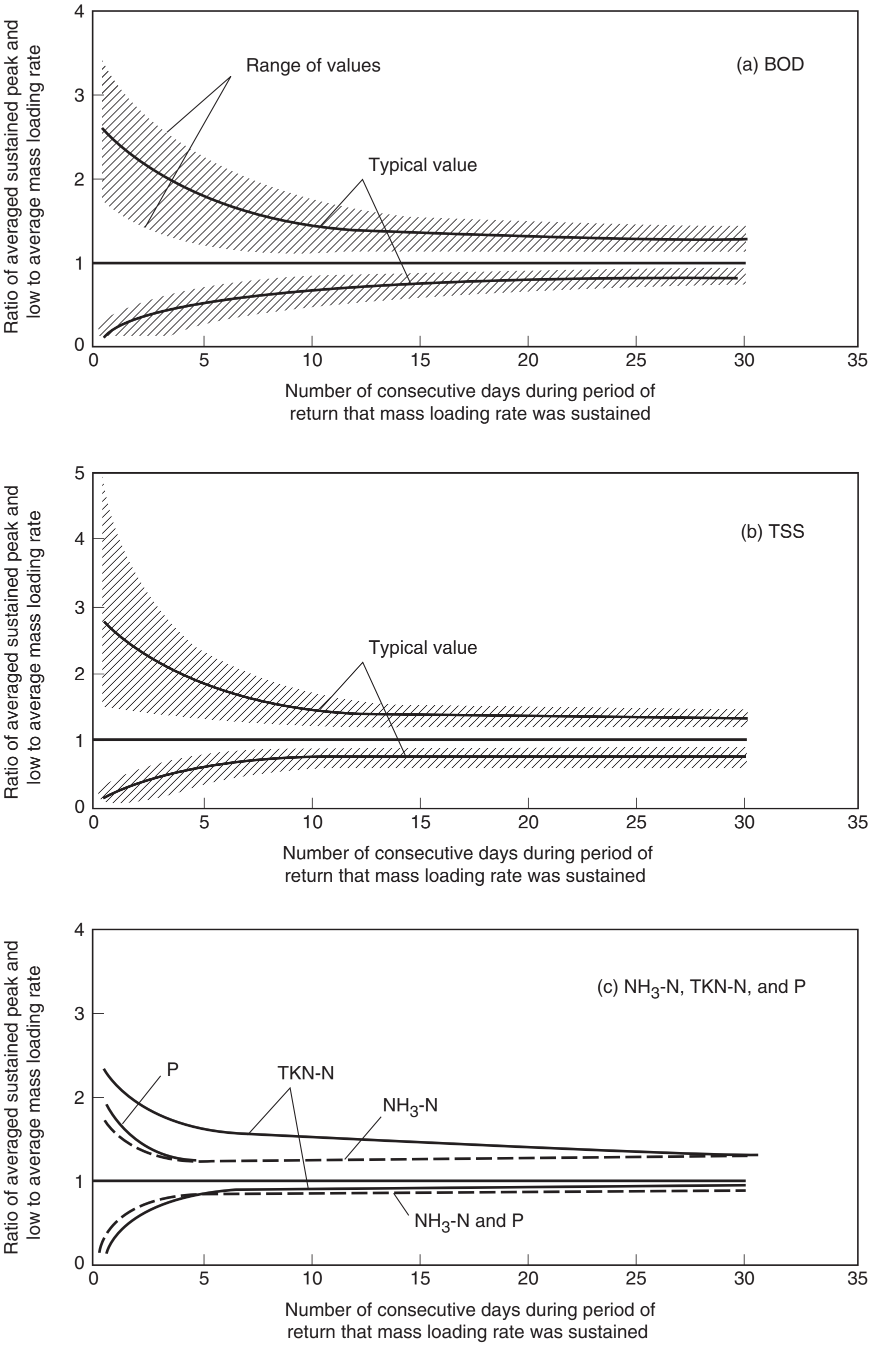


Figure 3–13

Typical information on the ratio of averaged peak and low-to average mass loading rates for: (a) biochemical oxygen demand, BOD; (b) total suspended solids,TSS; and (c) ammonia nitrogen, NH_3-N, total Kjeldahl nitrogen, TKN-N, and phosphorus, P.

Effect of Mass Loading Variability on Treatment Plant Performance

During the course of a day, the mass loading rate that is received by the treatment plant can vary widely as illustrated on Fig. 3–14. The variations are more pronounced in small collection systems where the collection system storage capacity does not provide a significant dampening effect. The impact of these load variations is seen most dramatically in the effects on

Figure 3–14

Illustration of typical hourly variations observed in flow and BOD and TSS concentrations and the corresponding BOD and TSS and mass loading rates. (Typical hourly variations are from Fig. 3–11.)

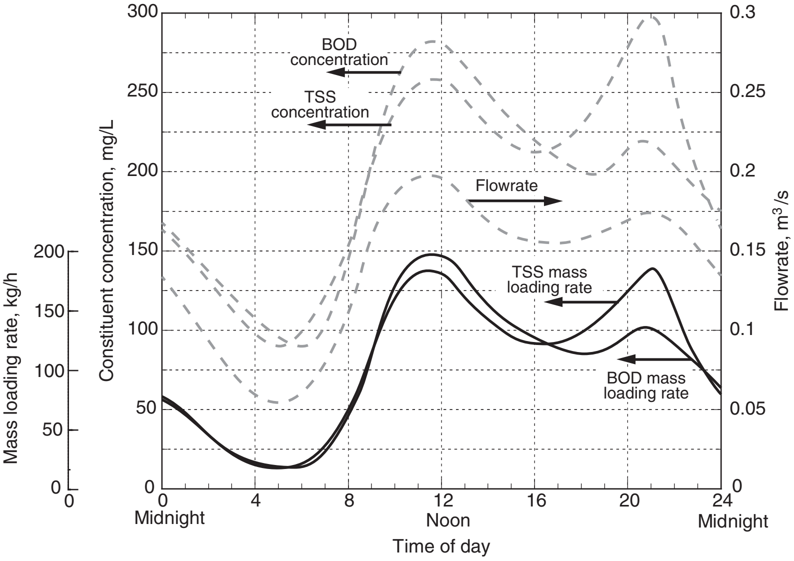
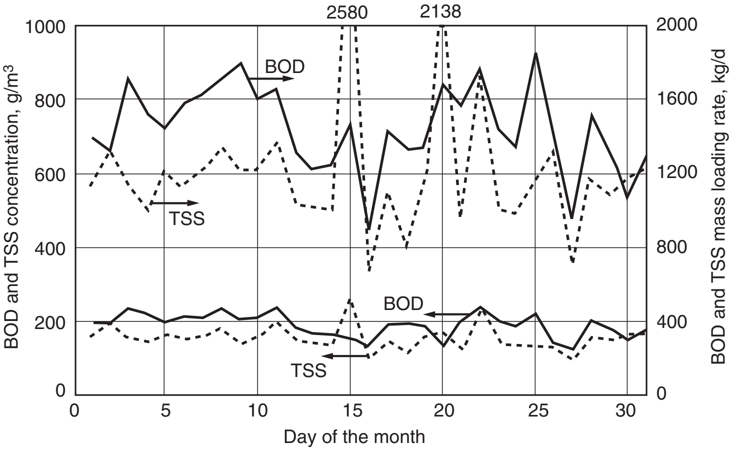


Figure 3–15

Example variations of TSS and BOD concentrations and mass loading rates over a monthly period.



biological treatment operating conditions. The maximum hourly BOD loading rate may vary as much as 3 to 4 times the minimum hourly BOD loading rate in a 24-h period. Over longer periods of time, the mass loading rates can also vary widely (see Fig. 3–15). These types of variations have to be accounted for in the design of the biological treatment system. In extreme cases, flow or load equalization may be required.

3–6 SELECTION OF DESIGN FLOWRATES AND MASS LOADINGS

The rated capacity of wastewater treatment plants is normally based on the average annual daily flowrate at the design year plus an allowance for future growth. As a practical matter, however, wastewater treatment plants have to be designed to meet a number of conditions that are influenced by flowrates, wastewater characteristics and constituent concentrations, and a combination of both flowrate and concentration (mass loading). Conditions that must be considered include peak and minimum hydraulic flowrates and the maximum, minimum, and sustained process constituent mass loading rates. The importance of wastewater flowrates in process design and operation is considered in this section; mass loadings are discussed in a following section.

Additionally, periods of initial operation and low flows and loads must be taken into consideration in design. Typical flowrate and mass loading factors that are important in the design and operation of wastewater treatment facilities are described in Table 3–22. The overall objective of wastewater treatment is to provide a wastewater treatment system that

Table 3–22

Typical flowrate and mass loading factors used for the design and operation of wastewater treatment facilities[a]

Flowrate factor	Purpose for design and operation
Flowrate	
Average daily flow	Development of flowrate ratios and for estimating pumping and chemical costs. Also used to estimate operator staffing requirements
Peak hour	Sizing of pumping facilities and conduits; sizing of physical unit processes: grit chambers, sedimentation tanks, and filters; sizing chlorine contact tanks. Getting the peak hourly flowrate through the treatment plant without flooding channels and weirs is a major design challenge. Also important in developing process control strategies.
Maximum day	Sizing of equalization basins, sedimentation tank and clarifier effluent troughs and weirs, chlorine contact tanks, and sludge pumping system. Also important in developing process control strategies for managing high flows
Maximum month	Record keeping and reporting; selection of maximum number of operating units required during high flow periods; sustained high flow operational strategies; and sizing of chemical storage facilities
Minimum hour	Sizing turndown of pumping facilities and chemical feed systems and determining low range of plant flowmeter(s)
Minimum day	Sizing of influent channels to control solids deposition; sizing effluent recycle requirements for trickling filters
Minimum month	Selection of minimum number of operating units required during low flow periods; scheduling shutdown for maintenance
Mass loading	
Minimum month	Process turndown requirements
Minimum day	Sizing of trickling filter recycle rates
Maximum day	Sizing of selected process units
Maximum month	Sizing of sludge storage facilities; sizing of composting requirements
Maximum 15-d	Sizing anaerobic and aerobic digesters
Sustained loading	Sizing of selected process units and ancillary process equipment

[a] Adapted from Tchobanoglous et al. (2003).

is capable of coping with a wide range of probable wastewater conditions while complying with the overall performance requirements.

Design Flowrates

The development and forecasting of flowrates is necessary to determine the design capacity as well as the hydraulic requirements of the treatment system. Flowrates need to be developed both for the initial period of operation and for the future (design) period. Consideration of the flowrates during the early years of operation is often overlooked, and oversizing of equipment and inefficient operation can result. The focus of the following discussion is on the development of various design flowrates.

Rationale for the Selection of Flowrates. The rationale for selecting flowrates is based on hydraulic and process considerations. As stated, the process units and hydraulic conduits must be sized to accommodate the anticipated peak flowrates that will pass through the treatment plant. Provisions have to be made to ensure bypassing of wastewater does not occur either in the collection system or at the treatment plant. Many of the process units are designed based on detention time or overflow rate (flowrate per unit of surface area) to achieve the desired removal rates of BOD and TSS. Because the performance of these units can be affected significantly by varying flowrate conditions and mass loadings, minimum and peak flowrates must be considered in design.

Forecasting Flowrates. In determining the design flowrate, elements to be considered are (1) the existing base flows; (2) estimated future flows for residential, commercial, institutional, and industrial sources; and (3) non-excessive infiltration/inflow. Existing base flows equal actual metered flowrates minus excessive infiltration/inflow (defined as infiltration/inflow that can be controlled by cost-effective improvements to the collection system).

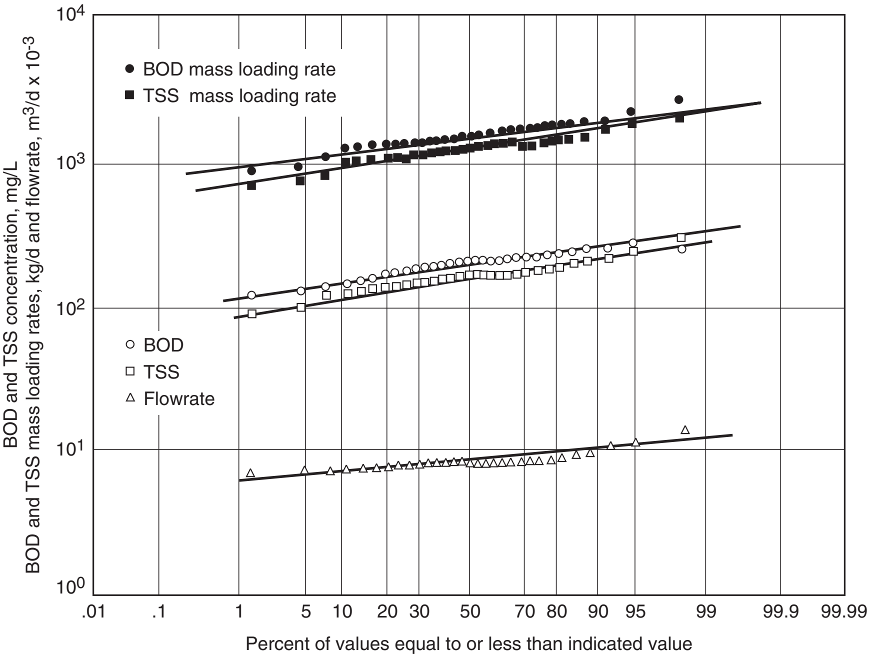
A yardstick by which total dry-weather base flow can be measured where infiltration is not excessive is 420 L/capita·d (110 gal/capita·d). This base flow includes 230 L/capita·d (60 gal/capita·d) for domestic flows, 40 L/capita·d (10 gal/capita·d) for commercial and small industrial flows, and 150 L/capita·d (40 gal/capita·d) for infiltration.

A useful technique in forecasting flowrates is probability analysis, discussed earlier in this chapter. Where flowrate data are available, preferably for at least two years, future flowrates for design can be predicted with a reasonable certainty. An example of a probability analysis of flowrates, as well as BOD and TSS concentrations and mass loadings, is shown on Fig. 3–16. The probability analysis can be used to estimate occurrences of peak flows and loads, and to establish a basis for selecting design flows and loads. For example, a maximum one-day occurrence can be determined based on a 99.7 percent probability; the value will not be equaled or exceeded in the time period analyzed. A probability value, such as the 95th percentile, can also be established for forecasting the design loadings to meet permit requirements.

Minimum Flowrates. As noted in Table 3–22, low flowrates are also of concern in treatment plant design, particularly during the initial years of operation when the plant is operating well below the design capacity, and in designing pumping stations. In cases where very low nighttime flow is expected, provisions for recycling treated effluent may have to be included to sustain the process, (e.g., biological treatment processes such as trickling filters, and to maintain optimal flowrates through ultraviolet disinfection systems). In absence of measured flowrate data, minimum daily flowrates may be assumed to range from 30 to 90 percent of average flowrates for small to large size communities, respectively [see also Fig. 3–3(c)] (WEF, 2010).

Figure 3–16

Typical probability plots for flowrate, BOD, and TSS.



Sustained Flowrates. Sustained flowrates are those that are equaled or exceeded for a specified number of consecutive days based on annual operating data. Data for sustained flowrates may be used in sizing equalization basins and other plant hydraulic components. An example plot of sustained and low flowrates is shown on Fig. 3–17. When developing plots similar to Fig. 3–17, the longest available period should be used.

Peak Flowrate Factors. The flowrate peaking factors (the ratio of peak flowrate to average flowrate) most frequently used in design are those for peak hour and maximum day (see Table 3–22). Peak hourly flowrates are used to size the hydraulic conveyance system and other facilities such as sedimentation tanks and chlorine contact tanks where little volume is available for flow dampening. Other peaking factors such as maximum week or maximum month may be used for treatment facilities such as pond systems that have long detention times or for sizing solids and biosolids processing facilities that also

Figure 3–17

Typical ratios of averaged sustained peak and low daily flowrates to average annual daily flowrates for time periods up to 30 d.

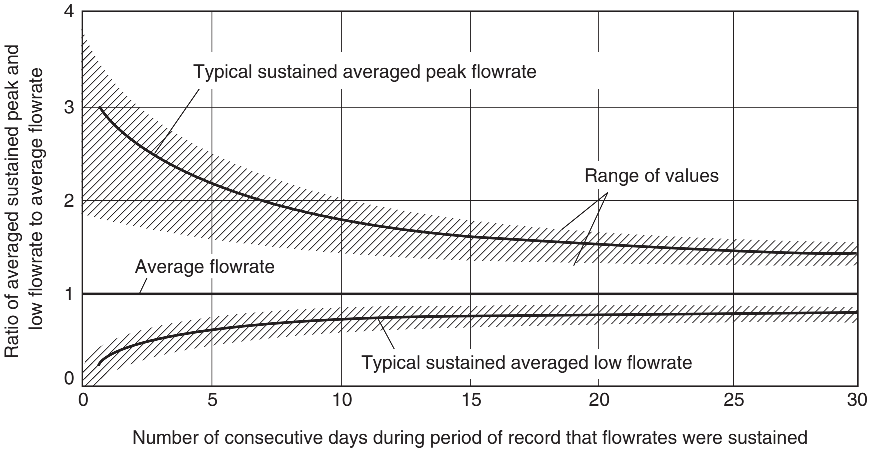
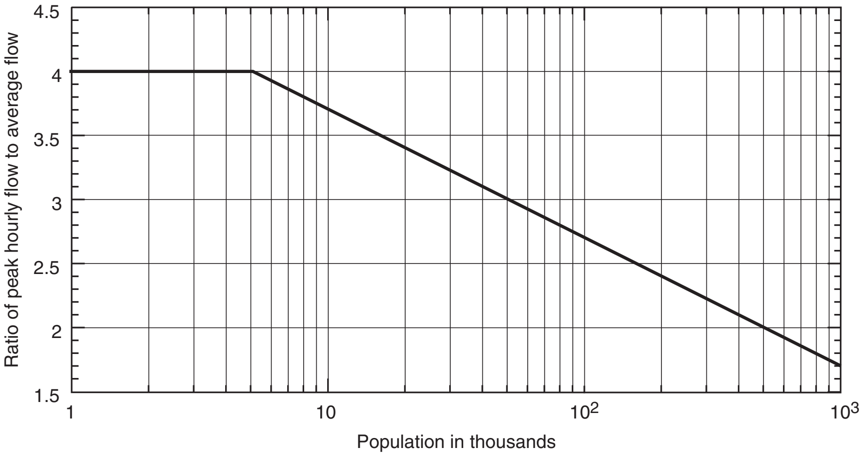


Figure 3–18

Peaking factor curve (ratio of peak hourly to average daily flow).



have long detention times or ample storage. Peaking factors may be developed from flowrate records or based on published curves or data from similar communities.

If flow measurement records are inadequate to establish peaking factors, published information may be used. Many sources for peaking factor data are available including state agencies, cities, and special districts that provide wastewater collection and treatment services, and professional publications from national organizations such as the Water Environment Federation and the American Society of Civil Engineers. (An example peaking factor curve is given on Fig. 3–18, and it may be used for estimating peak hourly flowrates from domestic sources.) The curve given on Fig. 3–18 was developed from analysis of the records of numerous communities throughout the United States. The curve is based on average residential flowrates, exclusive of infiltration/inflow, and includes small amounts of commercial flows and industrial wastes.

In developing factors for peak hourly flowrates, the characteristics of the collection system serving the wastewater treatment plant must be considered carefully. Improvements to or rehabilitation of the collection system may also increase or decrease the peaking factors. For pumped flows where reliable metering data are not available, factors to be considered include

- Interviews with operators regarding observations of operating conditions
- Review of pumping records (historical data on number of pumps in service and running time, if available)
- Operating speed of pumps
- Condition of pumps from maintenance records (unit output will be lower if impellers are worn)

Field testing at pumping stations can also be performed to measure the combined output of a simulated historical high flow event. Assistance in performing pump tests is often available from the local energy service provider.

Where flow to the treatment plant is by gravity, the peak flowrate can be estimated based on the following:

- Capacity of the influent sewers
- Investigation of upstream access ports (i.e., manholes) to determine if a high water mark is visible
- Interviews with operating staff and review of any documented field records

Forecasting design flowrates, including the use of peaking factors, is illustrated in Example 3–8.

EXAMPLE 3–8 **Forecasting Design Flowrates** A residential community with a current population of 15,000 is planning to expand its wastewater treatment plant. In 20 y, the population is estimated to increase to 25,000. Given the following information, estimate the future average, peak, and minimum design flowrate at plus 10 and 20 y.

Current situation:

1. Population is 15,000 persons
2. Average daily wastewater flowrate is 7500 m^3/d.
3. Infiltration/inflow has been determined to be non-excessive. Infiltration is estimated to be 100 L/capita·d at average flowrate and 150 L/capita·d at peak flowrate.
4. Municipal use is estimated to be 40 L/capita·d at average flowrate and 60 L/capita·d at peak flowrate.
5. An industry contributes an average flowrate of 1000 m^3/d with a peak flowrate of 1500 m^3/d.

Future situation:

1. Population is estimated to increase linearly to 25,000 in 20 y.
2. Residential water use in new homes is expected to be 20 percent less than the existing residences because of the installation of water saving appliances and fixtures as new homes are built and old homes are renovated.
3. The per capita wastewater from existing homes will decrease by 20 percent linearly over 20 y.

Other assumptions:

1. Current wastewater peaking factor is 3.0, which is expected to decrease linearly to 2.0 in 20 y.
2. Current ratio of minimum to average flowrate is 0.35, which is expected to increase to 0.45 in 20 y.

Solution

1. Compute the present and future per capita wastewater flowrates.
   a. For current conditions, compute the average domestic flowrate excluding infiltration and municipal use.
      i. Compute infiltration

$$\text{Infiltration} = (15{,}000 \text{ persons})(100 \text{ L/capita·d})\left(\frac{1 \text{ m}^3}{10^3 \text{ L}}\right) = 1500 \text{ m}^3/\text{d}$$

      ii. Compute average municipal flowrate

$$\text{Municipal flow} = (15{,}000 \text{ persons})(40 \text{ L/capita·d})\left(\frac{1 \text{ m}^3}{10^3 \text{ L}}\right) = 600 \text{ m}^3/\text{d}$$

      iii. Compute average domestic flowrate

$$\begin{aligned}\text{Domestic flow, m}^3/\text{d} &= \text{total} - \text{infiltration} - \text{municipal} - \text{industrial}\\ &= 7500 - 1500 - 600 - 1000 = 4400 \text{ m}^3/\text{d}\end{aligned}$$

b. Compute present per capita flowrate by dividing the existing domestic flowrate by the present population.

$$\text{Per capita flowrate} = \frac{(4400\ \text{m}^3/\text{d})}{15{,}000\ \text{persons}} = 0.29\ \text{m}^3/\text{capita}\cdot\text{d}$$

c. For future conditions, reduce existing per capita flowrate by 20%.

$$\text{Future per capita flowrate} = 0.29 \times 0.8 = 0.232\ \text{m}^3/\text{capita}\cdot\text{d}$$

2. Compute future average flowrate.

	Flowrate, m^3/d		
Flowrate	Current	Plus 10	Plus 20
From current residents	4400	3960	3520
From future residents (5000 × 0.232 m^3/capita·d)		1160	
From future residents (10,000 × 0.232 m^3/capita·d)			2320
Subtotal residential	4400	5120	5840
Industrial flowrate	1000	1000	1000
Infiltration (15,000)(100 L/capita·d)(1 $m^3/10^3$ L)	1500		
Infiltration (20,000)(100 L/capita·d)(1 $m^3/10^3$ L)		2000	
Infiltration (25,000)(100 L/capita·d)(1 $m^3/10^3$ L)			2500
Municipal (15,000)(40 L/capita·d)(1 $m^3/10^3$ L)	600		
Municipal (20,000)(40 L/capita·d)(1 $m^3/10^3$ L)		800	
Municipal (25,000)(40 L/capita·d)(1 $m^3/10^3$ L)			1000
Total average flowrate =	7500	8920	10,340
Average per capita flowrate	0.50	0.45	0.41

3. Compute future peak flowrate.

	Flowrate, m^3/d		
Flowrate	Current	Plus 10	Plus 20
Residential peak flowrate (4400 × 3.0)	13,230		
Residential peak flowrate (5120 × 2.5)		12,800	
Residential peak flowrate (5840 × 2.0)			11,680
Industrial peak flowrate	1500	1500	1500
Infiltration (15,000)(150 L/capita·d)(1 $m^3/10^3$ L)	2250		
Infiltration (20,000)(150 L/capita·d)(1 $m^3/10^3$ L)		3000	
Infiltration (25,000)(150 L/capita·d)(1 $m^3/10^3$ L)			3750
Municipal (15,000)(60 L/capita·d)(1 $m^3/10^3$ L)	900		
Municipal (20,000)(60 L/capita·d)(1 $m^3/10^3$ L)		1200	
Municipal (25,000)(60 L/capita·d)(1 $m^3/10^3$ L)			1500
Total peak flowrate	17,880	18,500	18,430
Peak per capita flowrate	1.19	0.93	0.74

4. Compute the minimum flowrate

Flowrate	Flowrate, m^3/d Current	Plus 10	Plus 20
Residential minimum flowrate (4400 × 0.35)	1323		
Residential minimum flowrate (5120 × 0.40)		2048	
Residential minimum flowrate (5840 × 0.45)			2628
Industrial (facilities are shut down at night)	0	0	0
Infiltration (15,000)(100 L/capita·d)(1 $m^3/10^3$ L)	1500		
Infiltration (20,000)(100 L/capita·d)(1 $m^3/10^3$ L)		2000	
Infiltration (25,000)(100 L/capita·d)(1 $m^3/10^3$ L)			2500
Municipal (600 × 0.35)	210		
Municipal (800 × 0.40)		320	
Municipal (1000 × 0.45)			450
Total minimum flowrate	3033	4368	5578
Peak per capita minimum flowrate	0.2	0.22	0.22

Comment In this example, infiltration/inflow contributes about 50 percent of the minimum flowrate and 20 percent of the average flowrate, an illustration of the influence of extraneous flows on treatment plant design. If wastewater flow records are not adequate or are unavailable, future average daily flow may be calculated based on the future population and unit wastewater flowrates, given in Sec. 3–1. Appropriate adjustments should be made in the calculations to account for any special conditions such as flow reduction, infiltration/inflow allowances, and industrial flows. When peak flowrates for more than one flow component are calculated, some adjustment in the total peak flowrate should be made if the peaks from the components do not occur simultaneously.

Upstream Control of Peak Flowrates. Planning wastewater plants to handle peak flowrates may involve other considerations, including (1) improvements to the collection system to reduce peak flow related to infiltration/inflow (I/I) and (2) installation of flow equalization facilities to provide storage either in the collection system or at the treatment plant. Other alternatives for peak flowrate control at the treatment plant, namely provision for flow splitting and bypass facilities, are discussed in Sec. 3–7.

Improvement to the collection system may involve a lengthy and costly process and may not have an immediate effect on significantly reducing peak flowrates. In some cases, the amount of flow reduction resulting from collection system rehabilitation has been less than anticipated, particularly if infiltration is a significant component of the total flow stream. In unusual circumstances, the flowrates have actually increased after completion of the collection system improvement program. Therefore, safety factors should be considered when estimating possible peak flowrate reduction resulting from collection system improvements.

Flow equalization can be an effective measure in reducing peak flows. Benefits derived by upstream flow equalization include (1) reduced hydraulic loading on already overtaxed collection facilities, (2) reduced collection system overflows (and reduced public health threats), and (3) reduced peak hydraulic loading of the treatment plant.

Equalization depends on available volume and may be of limited value in extreme peak flow conditions. Siting of equalization facilities in the collection system is often difficult because of limited available space at locations that are compatible with the system hydraulics. Operation and maintenance may also be difficult to manage, particularly in remote areas. Ease of operation, maintenance, and control and environmental factors are major reasons that many equalization facilities are located at treatment plants. The analysis for sizing flow equalization facilities is presented in Sec. 3–7.

Design Mass Loading Rates

The importance of mass loading rates in the design of wastewater treatment plants is identified in Table 3–22. For example, the sizing of aeration facilities and the amounts of solids and biosolids produced are related directly to the mass of BOD that must be processed. Further, the performance of the preliminary and primary treatment facilities have to be taken into account as ineffective operation of these facilities can result in the transfer of greater organic loads to the biological treatment system. Peak process loading rates are also important in sizing the process units and their support systems so that treatment plant performance objectives can be achieved consistently and reliably. The performance of primary sedimentation tanks is discussed in Chap. 5.

EXAMPLE 3–9 Estimation of Variability of Influent Wastewater Design Parameters

Compute the expected maximum values for the influent parameters flowrate, BOD, COD, and TSS for the case of a small and large size wastewater treatment facility. Assume the following mean design values apply:

		Mean design values	
Parameter	Unit	Small	Large
Flowrate	m^3/d	10,000	500,000
BOD	mg/L	250	250
COD	mg/L	600	600
TSS	mg/L	200	200

Determine the maximum value of the influent parameters for maximum day and maximum month. Comment on the importance of the results.

Solution

1. Select s_g values from Table 3–12 that correspond to the wastewater parameters of interest. In the absence of site and regionally specific information, use the typical values given in Table 3–12 as follows:

	Parameter			
Size of facility	Flowrate	BOD	COD	TSS
Small	1.6	1.6	1.7	1.6
Large	1.15	1.27	1.30	1.27

2. Locate the selected s_g value on Fig. 3–9 for a given frequency and determine the corresponding peaking factor.

Using the s_g values determined in Step 1, the corresponding peaking factors for peak day and peak month can be found on Fig. 3–9(a) for a small facility and Fig. 3–9(b) for a large facility. The peaking factors are summarized in the following table.

	Small facility			Large facility		
		Peaking factor			Peaking factor	
Parameter	s_g	d	mo	s_g	d	mo
Flowrate	1.6	3.70	2.35	1.15	1.48	1.29
BOD	1.6	3.70	2.35	1.27	1.95	1.55
COD	1.7	4.40	2.65	1.30	2.20	1.62
TSS	1.6	3.70	2.35	1.27	1.95	1.55

3. To obtain the maximum values for a specified frequency, multiply the peaking factors determined in step 2 by the mean value from the table given in the problem statement.
   a. For the peak day flowrate for the small facility the peaking factor is 3.70 and the mean design value is 10,000 m^3/d:

      $(3.70)(10{,}000\ m^3/d) = 37{,}000\ m^3/d$

   b. The design values for the two facilities are summarized in the following table.

		Design values					
		Small facility			Large facility		
Parameter	Unit	Mean	Peak d	Peak mo	Mean	Peak d	Peak mo
Flowrate	m^3/d	10,000	37,000	23,500	500,000	740,000	645,000
BOD	mg/L	250	925	587.5	250	487.5	387.5
COD	mg/L	600	2640	1590	600	1320	972
TSS	mg/L	200	740	470	200	390	310

Comment As shown in the summary table presented in Step 3, the smaller facility must be designed to accommodate a larger range in influent wastewater parameters relative to the large facility.

3–7 FLOW AND CONSTITUENT LOAD EQUALIZATION

The variations of influent wastewater flowrate and characteristics at wastewater treatment facilities were discussed in Secs. 3–4 and 3–6. Flow equalization is a method used to overcome the operational problems caused by flowrate variations, to improve the performance of the downstream processes, and to reduce the size and cost of downstream treatment facilities. Similarly, load equalization is a method used to reduce capital and operating cost of downstream treatment facilities. A variation of flow equalization involves the use of large offline storage basins or tunnels that are built specifically to capture and temporally store excess stormwater flow that would otherwise be discharged through CSO outfalls (see Fig. 3–19). The discussion below, however, focuses primarily on flow equalization as it pertains to wastewater treatment plant design.

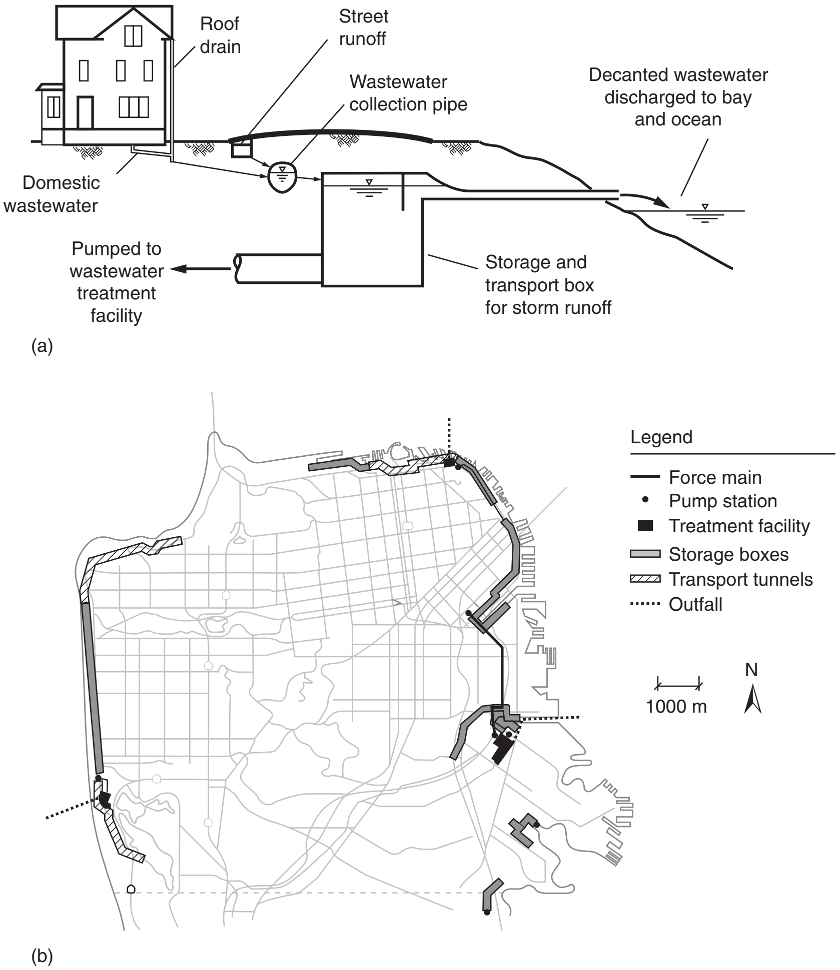


Figure 3–19

Details of combined collection system in the City of San Francisco, CA. (a) domestic wastewater, water from roof drains, and street runoff are collected in a combined collection system. Excess stormwater is discharged to large stormwater transport/storage boxes and tunnels. Once the stormwater flow has receded and treatment capacity becomes available, the wastewater from the storage boxes is treated before being discharged to San Francisco Bay or the Pacific Ocean, and (b) location of transport/storage boxes and tunnels around the periphery of the city (courtesy of City of San Francisco, CA).

Description/Application of Flow Equalization

Flow equalization simply is the dampening of flowrate variations to achieve a constant or nearly constant flowrate and can be applied in a number of different situations, depending on the characteristics of the collection system. The principal applications are for the equalization of (1) dry-weather flows to reduce peak flows and loads, (2) wet-weather flows in sanitary collection systems experiencing inflow/infiltration, or (3) combined stormwater and sanitary system flows as briefly noted above.

The application of flow equalization in wastewater treatment is illustrated in the two flow diagrams given on Fig. 3–20. In the inline arrangement [see Fig. 3–20(a)], all of the flow passes through the equalization basin. This arrangement can be used to achieve a considerable amount of constituent concentration and flowrate dampening. In the offline arrangement [see Fig. 3–20(b)], only the flow above some predetermined flow limit is diverted into the equalization basin. Although pumping requirements are minimized in this arrangement, the amount of constituent concentration dampening is considerably reduced. Offline equalization is sometimes used to capture the "first flush" from combined collection systems.

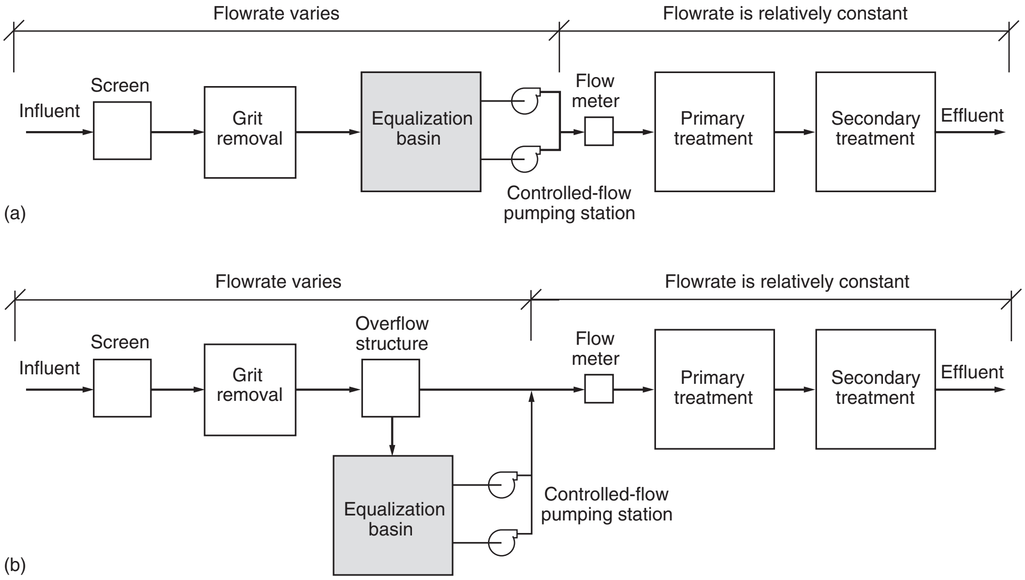


Figure 3–20

Typical wastewater treatment plant flow diagram incorporating flow equalization (a) inline equalization and (b) offline equalization.

The Benefits of Flow Equalization

The principal benefits that are cited as deriving from application of flow equalization are (1) biological treatment is enhanced, because shock loadings are eliminated or can be minimized, inhibiting substances can be diluted, and pH can be stabilized; (2) the effluent quality and thickening performance of secondary sedimentation tanks following biological treatment is improved through improved consistency in solids loading and the elimination of flow surges; (3) effluent filtration, or other tertiary treatment systems, surface area or volumetric requirements are reduced, performance is improved, and, for filtration systems, more uniform filter-backwash cycles are possible by lower hydraulic loading; and (4) in chemical treatment, dampening of mass loading improves chemical feed control and process reliability. Apart from improving the performance of most treatment operations and processes, flow equalization is an attractive option for upgrading the performance of overloaded treatment plants. Disadvantages of flow equalization include (1) relatively large land areas or sites are needed, (2) additional operation and maintenance is required, (3) potential for odors, and (4) increased capital cost. The use of, benefits, advantages, and disadvantages of flow equalization in municipal wastewater treatment were reviewed in an early U.S. EPA report (Ongerth, 1979).

Design Considerations

The design of flow equalization facilities is concerned with the following questions:

1. Where in the treatment process flowsheet should the equalization facilities be located?
2. What type of equalization flowsheet should be used, inline or offline?
3. What is the required equalization volume?
4. What are the features that should be incorporated into design?
5. How can the deposition of solids and potential odors be controlled?

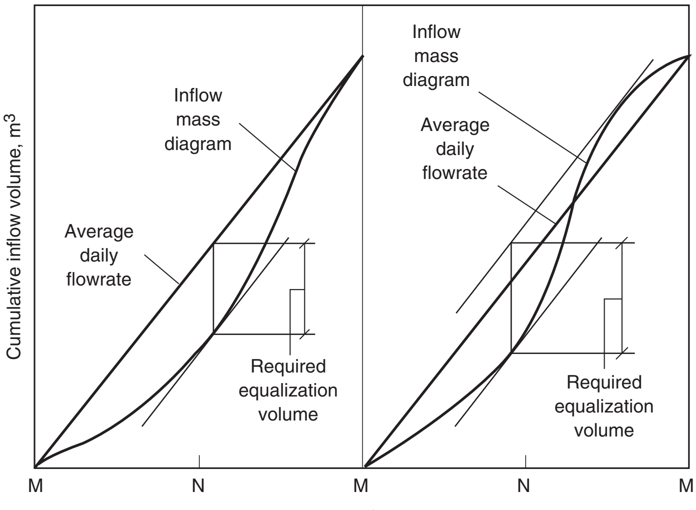
Location of Flow Equalization Facilities. The optimal location for siting equalization facilities must be determined for each system. Because the location will vary depending upon the characteristics of the collection system and the wastewater to be handled, land requirements and availability, and the type of treatment required, detailed studies should be performed for several locations throughout the system. Where equalization facilities are considered for location adjacent to the wastewater treatment plant, it is necessary to evaluate how they could be integrated into the treatment process flowsheet. In some cases, equalization after primary treatment and before biological treatment may be appropriate. Equalization after primary treatment causes fewer problems with solids deposits and scum accumulation. If flow equalization systems are to be located ahead of primary settling and biological systems, the design must provide for sufficient mixing to prevent solids deposition and concentration variations, and aeration to prevent odor problems.

Determination of Volume Requirements for Equalization Basin. The volume required for flowrate equalization is determined by using an inflow cumulative volume diagram in which the cumulative inflow volume is plotted versus the time of day. The average daily flowrate, also plotted on the same diagram, is the straight line drawn from the origin to the end point of the diagram. Diagrams for two typical flowrate patterns are shown on Fig. 3–21.

To determine the required volume, a line parallel to the coordinate axis, defined by the average daily flowrate, is drawn tangent to the mass inflow curve. The required volume is then equal to the vertical distance from the point of tangency to the straight line representing the average flowrate [see Fig. 3–21(a)]. If the inflow mass curve goes above the line representing the average flowrate [see Fig. 3–21(b)], the inflow mass diagram must be bounded with two lines that are parallel to the average flowrate line and tangent to extremities of the inflow mass diagram. The required volume is then equal to the vertical distance between the two lines. The determination of the required volume for equalization is also illustrated in Example 3–10. The procedure is exactly the same as if the average hourly volume were subtracted from the volume flow occurring each hour, and the resulting

Figure 3–21

Schematic mass diagrams for the determination of the required equalization basin storage volume for two typical flowrate patterns.



(a) Flowrate pattern A (b) Flowrate pattern B cumulative volumes were plotted. In this case, the low and high points of the curve would be determined using a horizontal line.

The physical interpretation of the diagrams shown on Fig. 3–21 is as follows. At the low point of tangency (flowrate pattern A) the storage basin is empty. Beyond this point, the basin begins to fill because the slope of the inflow mass diagram is greater than that of the average daily flowrate. The basin continues to fill until it becomes full at midnight. For flowrate pattern B, the basin is filled at the upper point of tangency.

In practice, the volume of the equalization basin will be larger than that theoretically determined to account for the following factors:

1. Continuous operation of aeration and mixing equipment will not allow complete drawdown, although special structures can be built.
2. Volume must be provided to accommodate the concentrated plant recycle streams that are expected, if such flows are returned to the equalization basin (a practice that is not recommended because of the potential to create odors).
3. Some contingency should be provided for unforeseen changes in diurnal flow.

Although no fixed value can be given, the additional volume will vary from 10 to 20 percent of the theoretical value, depending on the specific conditions.

EXAMPLE 3–10 **Determination of Flowrate Equalization Volume Requirements and Effects on BOD Mass Loading** For the flowrate and BOD concentration data given in following table, determine (1) the inline storage volume required to equalize the flowrate graphically (Note: the analytical spreadsheet solution is left to the reader) and (2) the effect of flow equalization on the BOD mass loading rate.

	Given data		Derived data	
Time period	Average flowrate during time period, m^3/s	Average BOD concentration during time period, mg/L	Cumulative volume of flow at end of time period, m^3	BOD mass loading during time period, kg/h
M–1	0.275	150	990	149
1–2	0.220	115	1782	91
2–3	0.165	75	2376	45
3–4	0.130	50	2844	23
4–5	0.105	45	3222	17
5–6	0.100	60	3582	22
6–7	0.120	90	4014	39
7–8	0.205	130	4752	96
8–9	0.355	175	6030	223
9–10	0.410	200	7506	295
10–11	0.425	215	9036	329
11–N	0.430	220	10,584	341
N–1	0.425	220	12,114	337

(*continued*)

(*Continued*)

	Given data		Derived data	
Time period	Average flowrate during time period, m^3/s	Average BOD concentration during time period, mg/L	Cumulative volume of flow at end of time period, m^3	BOD mass loading during time period, kg/h
1–2	0.405	210	13,572	306
2–3	0.385	200	14,958	277
3–4	0.350	190	16,218	239
4–5	0.325	180	17,388	211
5–6	0.325	170	18,558	199
6–7	0.330	175	19,746	208
7–8	0.365	210	21,060	276
8–9	0.400	280	22,500	403
9–10	0.400	305	23,940	439
10–11	0.380	245	25,308	335
11–M	0.345	180	26,550	224
Average	0.307			213

Note: $m^3/s \times 35.3147 = ft^3/s$
$m^3 \times 35.3147 = ft^3$
$mg/L = g/m^3$.

Solution

1. Determine the volume of the inline basin required for the flow equalization.
   a. The first step is to develop a cumulative volume curve of the wastewater flowrate expressed in cubic meters. The cumulative volume curve is obtained by converting the average flowrate (q_i) during each hourly period to cubic meters, using the following expression, and then cumulatively by summing the hourly values to obtain the cumulative flow volume.

$$\text{Volume, m}^3 = (q_i, \text{ m}^3/\text{s})(3600 \text{ s/h})(1.0 \text{ h})$$

For example, for the first three time periods shown in the data table, the corresponding hourly volumes are as follows:
For the time period M–1:

$$V_{M-1} = (0.275 \text{ m}^3/\text{s})(3600 \text{ s/h})(1.0 \text{ h})$$
$$= 990 \text{ m}^3$$

For the time period 1–2:

$$V_{1-2} = (0.220 \text{ m}^3/\text{s})(3600 \text{ s/h})(1.0 \text{ h})$$
$$= 792 \text{ m}^3$$

The cumulative flow, expressed in m^3, at the end of each time period is determined as follows:
At the end of the first time period M–1:

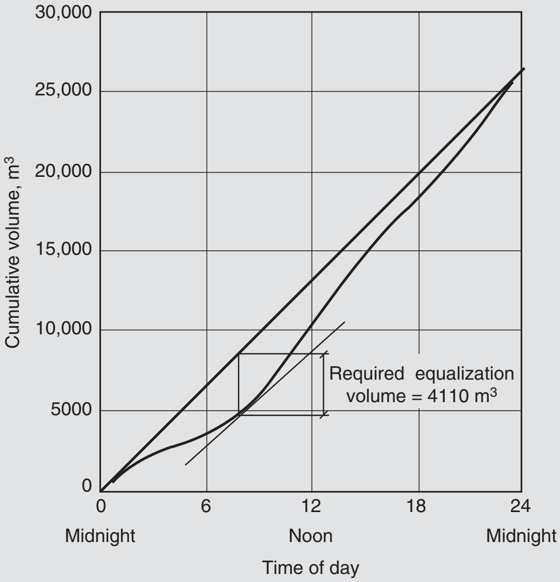
$$V_1 = 990 \text{ m}^3$$

At the end of the second time period 1–2:

$$V_2 = 990 + 792 = 1782 \text{ m}^3$$

The cumulative flows for all the hourly time periods are computed in a similar manner (see derived data in data table)

b. The second step is to prepare a plot of the cumulative flow volume, as shown in the following diagram. As will be noted, the slope of the line drawn from the origin to the end point of the inflow mass diagram represents the average flowrate for the day, which in this case is equal to 0.307 m^3/s.



c. The third step is to determine the required inline storage volume. The required storage volume is determined by drawing a line parallel to the average flowrate tangent to the low point of the inflow mass diagram. The required volume is represented by the vertical distance from the point of tangency to the straight line representing the average flowrate. Thus, the required volume is equal to

Volume of equalization basin, $V \sim 4100\ m^3\ (144{,}790\ ft^3)$

2. Determine the effect of the equalization basin on the BOD mass loading rate. Although there are alternative computation methods, perhaps the simplest way is to perform the necessary computations starting with the time period when the equalization basin is empty. Because the equalization basin is empty at about 8:00 a.m., the necessary computations will be performed starting with the 8–9 time period.

a. The first step is to compute the liquid volume in the equalization basin at the end of each time period. The volume required is obtained by subtracting the equalized hourly flowrate expressed as a volume from the inflow flowrate also expressed as a volume. The volume corresponding to the equalized flowrate for a period of 1 h is $0.307\ m^3/s \times 3600\ s/h = 1106\ m^3$. Using this value, the volume in storage is computed using the following expression:

$$V_{sc} = V_{sp} + V_{ic} - V_{oc}$$

where V_{sc} = volume in the equalization basin at the end of current time period
V_{sp} = volume in the equalization basin at the end of previous time period
V_{ic} = volume of inflow during the current time period
V_{oc} = volume of outflow during the current time period

Thus, using the values in the original data table, the volume in the equalization basin for the time period 8–9 is as follows:

$$V_{sc} = 0 + 1278 \text{ m}^3 - 1106 \text{ m}^3 = 172 \text{ m}^3$$

For time period 9–10:

$$V_{sc} = 172 \text{ m}^3 + 1476 \text{ m}^3 - 1106 \text{ m}^3 = 542 \text{ m}^3$$

The volume in storage at the end of each time period has been computed in a similar way (see following computation table)

Time period	Volume of flow during time period, m^3	Volume in storage at end of time period, m^3	Average BOD concentration during time period, mg/L	Equalized BOD concentration during time period, mg/L	Equalized BOD mass loading during time period, kg/h
8–9	1278	172	175	175	193
9–10	1476	542	200	197	218
10–11	1530	966	215	210	232
11–N	1548	1408	220	216	239
N–1	1530	1832	220	218	241
1–2	1458	2184	210	214	237
2–3	1386	2464	200	209	231
3–4	1260	2618	190	203	224
4–5	1170	2680	180	196	217
5–6	1170	2746	170	188	208
6–7	1188	2828	175	184	203
7–8	1314	3036	210	192	212
8–9	1440	3370	280	220	243
9–10	1440	3704	305	245	271
10–11	1368	3966	245	245	271
11–M	1242	4102	180	230	254
M–1	990	3986	150	214	237
1–2	792	3972	115	196	217
2–3	594	3160	75	179	198
3–4	468	2522	50	162	179
4–5	378	1794	45	147	162
5–6	360	1048	60	132	146
6–7	432	374	90	119	132
7–8	738	0	130	126	139
Average					213

Note: $\text{m}^3 \times 35.3147 = \text{ft}^3$
kg × 2.2046 = lb
$\text{g/ m}^3 = \text{mg/L}$.

b. The second step is to compute the average concentration leaving the storage basin. Using the following expression, which is based on the assumption that the contents of the equalization basin are mixed completely, the average concentration leaving the storage basin is

$$C_{oc} = \frac{(V_{ic})(C_{ic}) + (V_{sp})(C_{sp})}{V_{ic} + V_{sp}}$$

where C_{oc} = average concentration of BOD in the outflow from the storage basin during the current time period, g/m^3 (mg/L)
V_{ic} = volume of wastewater inflow during the current period, m^3
C_{ic} = average concentration of BOD in the inflow wastewater volume, g/m^3
V_{sp} = volume of wastewater in storage basin at the end of the previous time period, m^3
C_{sp} = concentration of BOD in wastewater in storage basin at the end of the previous time period, g/m^3

Using the data given in column 2 of the above computation table, the effluent concentration is computed as follows:
For the time period 8–9:

$$C_{oc} = \frac{(1278 \text{ m}^3)(175 \text{ g/m}^3) + (0)(0)}{1278 \text{ m}^3}$$
$$= 175 \text{ g/m}^3 \text{ (mg/L)}$$

For the time period 9–10:

$$C_{oc} = \frac{(1476 \text{ m}^3)(200 \text{ g/m}^3) + (172 \text{ m}^3)(175 \text{ g/m}^3)}{(1476 + 172) \text{ m}^3}$$
$$= 197 \text{ mg/L}$$

All the concentration values computed in a similar manner are reported in the above computation table.

c. The third step is to compute the hourly mass loading rate using the following expression:

$$\text{Mass loading rate, kg/h} = \frac{(C_{oc}, \text{ g/m}^3)(q_i, \text{ m}^3\text{/s})(3600 \text{ s/h})}{(1000 \text{ g/kg})}$$

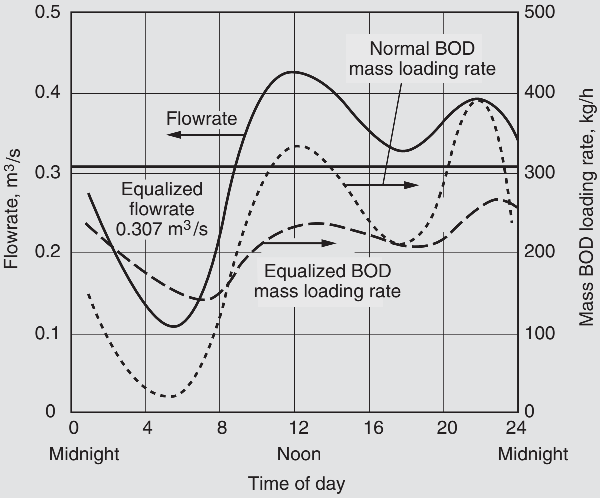
For example, for the time period 8–9, the mass loading rate is

$$\frac{(175 \text{ g/m}^3)(0.307 \text{ m}^3\text{/s})(3600 \text{ s/h})}{(1000 \text{ g/kg})} = 193 \text{ kg/h (426 lb/h)}$$

All hourly values are summarized in the computation table. The corresponding values without flow equalization are reported in the original data table.

d. The effect of flow equalization can be shown best graphically by plotting the hourly unequalized and equalized BOD mass loading (see the following plot). The following flowrate ratios, derived from the data presented in the table given in the problem statement and the computation table prepared in Step 2a, are also helpful in assessing the benefits derived from flow equalization:

	BOD mass loading	
Ratio	Unequalized	Equalized
$\frac{\text{Peak}}{\text{Average}}$	$\frac{439}{213} = 2.06$	$\frac{271}{213} = 1.27$
$\frac{\text{Minimum}}{\text{Average}}$	$\frac{17}{213} = 0.08$	$\frac{132}{213} = 0.62$
$\frac{\text{Peak}}{\text{Minimum}}$	$\frac{439}{17} = 25.82$	$\frac{271}{132} = 2.05$

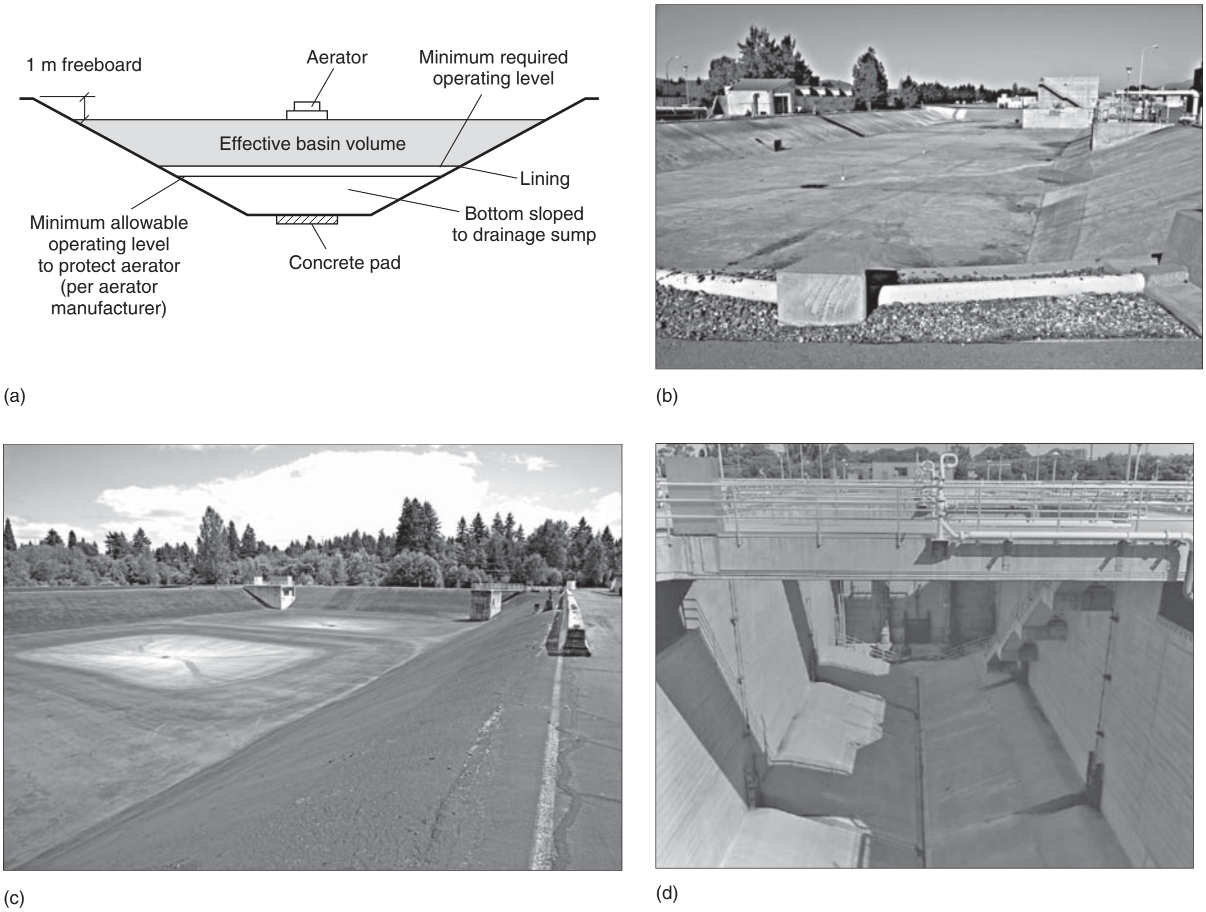


Comment Where inline flow equalization basins are used, additional dampening of the BOD mass loading rate can be obtained by increasing the volume of the basins. Alternatively, offline storage can be used to further reduce the variability of the BOD mass loading rate to the biological treatment process. Although the flow to a treatment plant was equalized in this example, flow equalization would be used, more realistically, in locations with high infiltration/inflow or peak stormwater flows.

Basin Configuration and Construction. In equalization basin design, the principal factors that must be considered are (1) basin geometry; (2) basin construction including cleaning, access, and safety; (3) mixing and air requirements; (4) operational appurtenances; and (5) pump and pump control systems.

Basin Geometry. The importance of basin geometry varies somewhat, depending on whether inline or offline flowrate equalization is used. If inline equalization is used to dampen both the flow and the mass loadings, it is important to use a geometry that allows the basin to function as a continuous-flow stirred-tank reactor. Therefore, elongated designs should be avoided, and the inlet and outlet configurations should be arranged to minimize short circuiting. Discharging the influent near the mixing equipment usually minimizes short circuiting. If the geometry of the basins is controlled by the available land area and an elongated geometry must be used, it may be necessary to use multiple inlets and outlets. Provisions should be included in the basin design for access by cleaning equipment such as front end loaders. Multiple compartments are also desirable to reduce cleaning costs and for odor control.

Basin Construction. New basins may be of earthen, concrete, or steel construction; earthen basins are generally the least expensive (see Fig. 3–22). Depending on local conditions, the interior side slopes may vary between 3:1 and 2:1. A section through a typical earthen basin is shown on Fig. 3–22(a). In most installations, a liner is required to prevent groundwater contamination [see Figs. 3–22(b) and (c)]. Basin depths will vary depending on land availability, groundwater level, and topography. If a liner is used in areas of high groundwater, the effects of hydraulic uplift on the liner have to be considered. The freeboard required depends on the surface area of the basin and local wind conditions. If a floating aerator is used to provide mixing and prevent septicity and odor formation, a minimum operating level is needed to protect the aerator. Typically, the minimum water depth can vary from 1.5 to 2 m (5 to 6 ft). With floating aerators, a concrete pad should be provided below the aerators to minimize erosion. To prevent wind-induced erosion in the



(a)

(b)

(c)

(d)

Figure 3–22

Typical flow equalization basins: (a) typical cross-section shallow flow equalization basins, (b) and (c) shallow lined earthen basins, and (d) deep concrete basin.

upper portions of the basin, it may be necessary to protect the slopes with riprap, soil cement, or a partial concrete layer. Fencing should also be provided to prevent public access to the basins.

In areas of high groundwater, drainage facilities should be provided to prevent embankment failure. To further ensure a stable embankment, the tops of the dikes should be of adequate width. The use of an adequate dike width will facilitate the use of mechanical equipment for maintenance and will also reduce construction costs, especially where mechanical compaction equipment is used.

Mixing and Air Requirements. The proper operation of both inline and offline equalization basins generally requires proper mixing and aeration. Mixing equipment should be sized to blend the contents of the tank and to prevent deposition of solids in the basin. To minimize mixing requirements, grit removal facilities should precede equalization basins where possible. Mixing requirements for blending a medium strength municipal wastewater (see Table 3–18), having a suspended solids concentration of approximately 210 mg/L, range from 0.004 to 0.008 kW/m^3 (0.02 to 0.04 $hp/10^3$ gal) of storage. Aeration is required to prevent the wastewater from becoming septic and odorous. To maintain aerobic conditions, air should be supplied at a rate of 0.01 to 0.015 $m^3/m^3 \cdot min$ (1.25 to 2.0 $ft^3/10^3$ $gal \cdot min$). In equalization basins that follow primary sedimentation and have short detention times (less than two hours), aeration may not be required.

Where mechanical aerators are used, baffling may be necessary to ensure proper mixing, particularly with a circular tank configuration. To protect the aerators in the event of excessive level drawdown, low-level shutoff controls should be provided. Because it may be necessary to dewater the equalization basins periodically, the aerators should be equipped with legs or draft tubes that allow them to come to rest on the bottom of the basin without damage. Various types of diffused air systems may also be used for mixing and aeration including static tube, jet, and aspirating aerators (see Sec. 5–12).

Operational Appurtenances. Among the appurtenances that should be included in the design of equalization basins are (1) facilities for flushing any solids and grease that may tend to accumulate on the basin walls, (2) an emergency overflow in case of pump failure, (3) a high water takeoff for the removal of floating material and foam, and (4) water sprays to prevent the accumulation of foam on the sides of the basin and to aid in scum removal. Solids removed from equalization basins should be returned to the head of the plant for processing.

Pumps and Pump Control. Because flow equalization imposes an additional head requirement within the treatment plant, pumping facilities are frequently required. Pumping may precede or follow equalization, but pumping into the basin is generally preferred for reliability of treatment operation. In some cases, pumping of both basin influent and equalized flows will be required. For offline flow equalization basins designed for the control of wet-weather peak flows, pump-in configurations may have very high and costly pumping requirements. If the hydraulic grade line allows, a gravity-in, pump out configuration is the most cost-effective. The pumps used to empty an offline, peak-flow equalization basin are usually much smaller than the pumps that would be required to pump flow into the basin.

An automatically controlled flow-regulating device will be required where gravity discharge from the basin is used. Where basin effluent pumps are used, instrumentation should be provided to control the preselected equalization rate. Regardless of the discharge method used, a flow measuring device should be provided on the outlet of the basin to monitor the equalized flow.

Equalization of Constituent Mass Loading Rates

While the focus of this section has been on flow equalization, it is important to note that equalization of constituent (e.g., BOD and TSS) mass loading rates is another approach that can be used to enhance biological treatment. The benefits of constituent load equalization include (1) improved biological treatment, because shock constituent loadings are eliminated or can be minimized, (2) improved utilization of aeration equipment, (3) reduced peak power requirements, and (4) reduced capacity of aeration blowers and related equipment.

Process Description. Constituent mass loading equalization is accomplished most effectively with offline storage facilities. Functionally, a portion of the flow is diverted during the period when the constituent concentrations are high (i.e., typically in the mid to late morning and early evening hours). The diverted flow and constituents are fed back into the system during periods when the aeration facilities are not used fully, typically in the late evening and early morning hours. Operationally, information must be available on the typical variation in the flowrate and constituent concentrations. Flowrate metering is routine. Also, new online continuous suspended solids meters are now available. The quantity of flow that must be diverted would be based on a control strategy utilizing flowrate and TSS measurements in conjunction with an appropriate algorithm.

Recent Development. Another approach to load equalization is to utilize primary effluent filtration, an old idea that has resurfaced because of the availability of new filter technologies. Operationally, the primary effluent filtration would only be used during periods when the concentration of the constituents in the influent wastewater is increasing such as in the mid to late morning hours or in the early evening hours. The mostly organic solids in the filter backwash water would be diverted to a relatively small storage facility. The organic solids from the storage facility would then be returned to the treatment process during periods when the aeration equipment is underutilized as described above. The diverted organic solids could also be sent to a fermenter for the production of volatile fatty acids needed for enhanced phosphorus removal.

Equalization of Sludge and Biosolids Processing Return Flows

Return flows from solids processing operations including sludge thickening, digester supernatant, and centrate and/or filtrate from biosolids dewatering historically have been returned to the headworks of the biological treatment process for reprocessing. However, as more restrictive wastewater treatment standards are being implemented, the practice of returning these flows to the headworks has made it difficult to achieve low discharge limits, especially for nitrogen and phosphorus. The impact of return flows is even greater because most of the return flows are reintroduced during the daytime hours when the biosolids dewatering facilities are normally operated.

The need to improve plant performance has lead to the implementation of flow equalization and/or separate treatment facilities for return flows. Flow equalization facilities are used most commonly at smaller treatment plants, whereas separate treatment and flow equalization facilities are more common at larger treatment plants. The sizing for flow equalization facilities can done using the principles outlined in this section. Flow equalization facilities for return flows at smaller treatment plants are designed to capture all of the return flows during the day so that the stored return flows can be added to the plant inflow over a 12-h period starting in the evening hours or when the incoming loadings to the plant are reduced. If space is limited, it may only be possible to reduce the peak of the return flows. Use of inactive treatment process tankage has proven to be effective. The separate treatment of return flows is considered in Chap. 15.

PROBLEMS AND DISCUSSION TOPICS

3–1 A community is rapidly approaching the design capacity of its wastewater treatment plant. As an alternative to expanding the plant, a water conservation program has been proposed. If the current average residential unit flowrate is 320 L/capita·d and the proposed water conservation rate reduction is 25, 35, 40 percent (to be selected by instructor), is the proposal reasonable? State your justification.

3–2 A new commercial development is being considered, and four developers have submitted competing proposals. An Environmental Assessment Report is being prepared to evaluate the wastewater flows from each of the competing proposals. Estimate the average daily and peak wastewater flowrates from one of the proposed developments (to be selected by instructor). The proposals consist of the following elements:

Type of facility	Units	Developer			
		1	2	3	4
Hotel	Number of guest rooms	120	80	60	250
	Number of employees	25	16	14	40
Department stores	Number of toilet rooms	8	12	16	
	Number of employees	40	60	80	
Self-service laundry	Number of machines		20	16	18
Restaurant, no bar	Number of seats	125	100	100	50
Restaurant, with bar	Number of seats	100	125	75	80
Theater (indoor)	Number of seats	500	400		350

3–3 Estimate the average and maximum flowrates from a recreational area (to be selected by instructor) with the following facilities. State all of your assumptions clearly.

Type of facility	Units	Recreational area			
		1	2	3	4
Visitor center	No. of visitors	250	300	400	500
Motel with kitchen	No. of guests		60	100	60
Resort cabins	No. of guests		100	40	
Cottages	No. of guests	60		60	120
Campground (toilets only)	No. of persons	140	120		200
Recreational vehicle park	Number of individual connections	40	50	20	50
Self-service laundry	Number of machines	8	10	6	10
Shopping center	Number of employees	10	15	15	20
	Number of parking spaces	30	40	40	60
Automobile service station	Number of vehicles	80	120	160	200
Restaurant with bar	Number of customers	200	300	400	500

3–4 A college dormitory complex, with four separate dormitories, is planning to institute a water conservation program to reduce its wastewater flows. Using the data given below, develop a water conservation program that will accomplish the flowrate reduction goal for one of the dormitories (to be selected by instructor).

		Dormitory			
Characteristics	Unit	1	2	3	4
Wastewater flowrate	m^3/d	125	105	140	160
Number of beds	No.	300	250	300	350
Existing fixture flowrates					
Toilets,	L/use	9	10	11	11
Showerheads,	L/min·use	18	20	23	23
Faucets,	L/capita·d	10	8	9	11
Flowrate reduction required	%	15	20	25	25

3–5 Compute the flow-weighted BOD and TSS concentrations from one of the following flowrate regimes (to be selected by instructor).

			Flowrate, m^3/d		
			Flowrate regime		
Time	BOD, mg/L	TSS, mg/L	1	2	3
02:00	130	150	8000	7200	10,000
04:00	110	135	6000	6400	8400
06:00	160	150	9400	9800	13,600
08:00	220	205	12,800	13,500	19,200
10:00	230	210	13,000	13,800	19,500
12:00	245	220	14,400	14,500	21,800
14:00	225	210	12,000	12,500	18,500
16:00	220	200	9600	10,000	14,800
18:00	210	205	11,000	10,500	15,000
20:00	200	210	8000	8500	11,500
22:00	180	185	9000	8200	12,600
24:00	160	175	8400	7700	11,600

3–6 The data in the table below consist of population values for a city, and average monthly influent flow values at the city wastewater treatment plant, from 2007 through 2010. The city is located in an area of high groundwater. Using these data answer the following questions.

a. What is the nature of the distribution of the monthly flowrate values? Use the plotting position method used in Example 3–4 [see Eq. (D–10) in Appendix D] to plot the monthly influent values versus the corresponding probabilities for each year on arithmetic- and log-probability paper, and check for linearity.

b. What is the average annual flow, average dry-weather flow (ADWF), and average wet-weather flow (AWWF) for each year? If the data are arithmetically distributed, use the arithmetic mean; if the data are log-normally distributed, use the geometric mean (see Table D–1 in Appendix D). Assume the dry season occurs from June to October and the wet season occurs from November to May.

c. What is the per capita flow contribution from commercial and light industrial activities? Assume that the residential contribution to the dry weather flow is 260 L/capita·d, and that commercial and light industrial activities make up the remaining flow.

d. What is the per capita flow contribution from infiltration and inflow for each year? Assume that the difference between the wet and dry weather flows is due to infiltration and inflow.

Community population and average monthly influent flowrate data for the period from 2007 through 2010

Year	2007	2008	2009	2010
Population	8690	9400	11,030	12,280
Month	Influent flowrate, m^3/d			
January	8800	13,900	8300	10,000
February	6200	9900	11,800	18,400
March	6800	8100	9400	13,000
April	4000	4200	6500	5000
May	4000	5700	5300	7600
June	3600	3600	4800	4600
July	2400	2600	3300	3800
August	2000	1500	3800	3100
September	2800	2000	2800	2200
October	3200	4800	4400	4400
November	4800	3200	6000	6500
December	5200	6700	7300	8600

3–7 Consider the city described in Problem 3–6. Assume that the build-out population for the community is 16,000 persons, and that the residential wastewater flowrate will be 300 L/capita·d. The commercial flowrate in 2010 (1000 m^3/d) is 80 percent of what it will be at build-out. Due to high infiltration and inflow (I/I) rates, a sewer repair program will be implemented. The I/I contribution will be either 500, 400, 300, or 200 L/cap·d (to be selected by instructor), depending on the degree of repair achieved. Estimate the ADWF, AWWF, and the average annual flow that will be received at the community treatment plant at build out. Justify the use of the AWWF as the nominal design capacity for the treatment plant.

3–8 Consider the treatment plant for the city described in Problem 3–6. Use the plotting position method illustrated in Example 3–4 [see Eq. (D–10) in Appendix D] to plot the monthly influent values versus the corresponding probabilities for each year. Determine the nature of the distribution by plotting the values on arithmetic- and log-probability paper and checking for linearity. If the average annual flow to the treatment facility at build-out is estimated to be 8000 m^3/d, what will the peak monthly flow be? (Hint: the average annual flow occurs at the 50 percent line on the probability graph. Use the slope of the wettest year to pass a line through the 50 percent line at 8000 m^3/d, and read the flow value for the highest month from the graph.)

3–9 Use the following influent data for a wastewater treatment plant to conduct the statistical analyses described in Problem 3–6, steps a, b, c, and d. If the data in the table below are not distributed either arithmetic-normally or log-normally, suggest a method for determining the required parameters.

Community population and average monthly influent flowrate data for the period from 2007 through 2010

Year	2007	2008	2009	2010
Population	17,040	17,210	17,380	17,630
Month	Influent flowrate, m^3/d			
January	8800	7760	9360	7600
February	9440	7280	7920	7840
March	8640	7200	8800	7680
April	7840	6960	8080	7440
May	7440	6800	7680	7280
June	7200	6880	7520	7360
July	7120	6960	7280	7200
August	7040	6720	7200	7280
September	6880	6880	7040	7200
October	6960	6800	7280	7440
November	7120	7120	7360	7680
December	7360	7600	7680	8000

3–10 A treatment plant is planned to be installed in a recreational resort that contains a developed campground for 200 persons, lodges and cabins for 100 persons, and resort apartments for 150 persons. Assume that persons staying in lodges use the dining hall for 3 meals per day and that a 50-seat cafeteria with 4 employees and an estimated 200 customers per day has been constructed. Daily attendance at the visitor center is 500 persons. Other facilities include a 10-machine laundromat, a 20-seat cocktail lounge, and three gas stations (1100 L/d per station). Determine the average wastewater flowrate in L/d using the unit flows assuming the housing facilities are at maximum capacity.

3–11 Obtain an annual report or one year of flow and BOD data from your local wastewater treatment facility. From these records, prepare probability plots for the flowrate and mass loadings. Determine the values at 50 and 95 percent probability.

3–12 From the flowrate data obtained for Problem 3–11, determine the mean and standard deviation.

3–13 The wastewater treatment plant has been experiencing high wastewater flowrates during the wet-weather months. The average monthly flows for four flow regimes are reported below (flow regime to be selected by instructor). The rapid increase in flows during the winter months is due mainly to increased infiltration/inflow. Infiltration is estimated to be 67 percent of the excess flow. The lengths of the collection system pipelines that need to be repaired are also listed below. The average repair cost is \$200,000/km and the repair will be effective in reducing the infiltration by 30 percent. How many years from now will it take to pay back the cost of the sewer repair program based on the annual savings in treatment cost, assuming the future annual flowrates are equal to those in the table below? The current cost of treatment is \$1.50/m^3 and the future cost of treatment is estimated to escalate at 6 percent per year. Assume the sewer repair will be complete in three years for Areas 1 and 2, and four years for Areas 3 and 4.

	Average monthly flowrate, m^3/d			
	Flowrate regime			
Month	1	2	3	4
January	293,000	410,000	460,000	470,000
February	328,000	459,000	440,000	485,000
March	279,000	391,000	515,000	560,000
April	212,000	296,000	333,000	400,000
May	146,000	204,000	230,000	300,000
June	108,000	151,000	170,000	225,000
July	95,000	133,000	150,000	200,000
August	89,000	125,000	141,000	188,000
September	93,000	130,000	140,000	165,000
October	111,000	155,000	167,000	192,000
November	132,000	185,000	200,000	240,000
December	154,000	215,000	225,000	215,000
Length of collection system, km	300	400	450	600

3–14 Nine months of flow records have been collected for a new wastewater treatment plant. Inspection of the records indicates that weekend flowrates tend to be higher than weekday flowrates. The weekday and weekend flowrates have been averaged and arranged in ascending order as given below. Develop arithmetic- and log-probability plots for both the weekday and weekend data for either flowrate regime 1 or 2 (to be selected by instructor) and comment on the skewness of the data. Determine the mean and 95th percentile values for each set of flow data, and determine the probable one day maximum flowrate. Discuss the significance of the data analysis.

	Flowrate regime 1		Flowrate regime 2	
Number	Weekend average flowrate, m^3/d × 10^3	Weekend average flowrate, m^3/d × 10^3	Weekend average flowrate, m^3/d × 10^3	Weekend average flowrate, m^3/d × 10^3
1	39.7	42.8	55.7	56.4
2	40.5	43.1	56.1	57.5
3	40.9	43.5	56.6	58.1
4	41.3	43.9	57.2	58.6
5	42.0	44.3	57.7	59.5
6	42.1	44.7	58.2	60.6
7	42.2	45.0	58.5	60.8
8	42.4	45.4	59.1	61.1
9	42.9	45.8	59.6	61.8

(*continued*)

(*Continued*)

	Flowrate regime 1		Flowrate regime 2	
Number	Weekend average flowrate, $m^3/d \times 10^3$	Weekend average flowrate, $m^3/d \times 10^3$	Weekend average flowrate, $m^3/d \times 10^3$	Weekend average flowrate, $m^3/d \times 10^3$
10	43.5	46.2	60.1	62.6
11	43.9	46.6	60.7	63.2
12	44.3	46.7	60.8	63.8
13	44.7	46.9	61.0	64.4
14	45.0	47.7	62.1	64.8
15	45.4	47.9	62.3	65.4
16	45.6	48.8	63.5	65.8
17	45.7	49.2	64.0	66.1
18	46.0	50.0	65.1	66.5
19	46.4	50.3	65.5	66.8
20	46.9	51.1	66.5	67.9
21	47.7	51.5	67.0	69.6
22	48.4	53.0	69.0	70.7
23	48.8	53.4	69.5	71.2
24	49.0	53.7	69.9	71.5
25	49.2	54.9	71.5	72.2
26	49.6	55.3	72.0	72.4
27	50.5	56.0	72.9	73.7
28	51.1	56.8	73.9	74.6
29	52.2	57.2	74.5	76.2
30	53.0	58.3	75.9	77.5
31	53.2	59.1	76.9	78.6
32	54.3	60.6	78.9	80.7
33	55.3	60.9	79.2	82.8
34	56.0	61.7	80.3	85.0
35	60.6	62.1	80.8	88.4
36	62.5	63.6	82.8	91.1

3–15 Land use in a new development is given in the following first table. A new school being built will have 1500 students. The average flowrate is 75 L/student, and the peaking factor (ratio of peak flowrate to average flowrate) is 4.0. Average flowrate allowances and the peaking factors for the other developments are shown in the second table. Determine the peak wastewater flowrate from one of the areas (to be selected by instructor).

	Area, ha			
Type of development	1	2	3	4
Residential	125	150	150	160
Commercial	11	10	15	16
School	4	4	4	4
Industrial	6	8	20	10

Type of development	Average flowrate, $m^3/ha \cdot d$	Peaking factor
Residential	40	3.0
Commercial	20	2.0
Industrial	30	2.5

3–16 Using the flowrate data presented below (city to be selected by instructor), estimate the size of an offline equalization tank needed to reduce the flowrate variation. The maximum hourly flowrate to the treatment process should not exceed 1.25, 1.5, and 1.75 times the average daily flowrate (ratio to be selected by instructor).

Time period	Average flowrate during time period, m^3/s		Time period	Average flowrate during time period, m^3/s	
	City 1	City 2		City 1	City 2
M–1	0.300	0.250	N–1	0.460	0.330
1–2	0.220	0.190	1–2	0.420	0.310
2–3	0.180	0.165	2–3	0.390	0.305
3–4	0.160	0.160	3–4	0.355	0.310
4–5	0.160	0.165	4–5	0.331	0.330
5–6	0.185	0.175	5–6	0.315	0.370
6–7	0.240	0.210	6–7	0.320	0.400
7–8	0.300	0.270	7–8	0.346	0.420
8–9	0.385	0.340	8–9	0.362	0.440
9–10	0.440	0.370	9–10	0.392	0.420
10–11	0.480	0.375	10–11	0.360	0.390
11–N	0.480	0.355	12–M	0.300	0.300

3–17 Using the information given in the data table presented in Example 3–10 determine (a) the offline storage volume needed to equalize the flowrate and (b) the effect of flow equalization on the BOD mass loading rate. How does the BOD mass loading rate curve determined in this problem compare with the curve shown in step 2d of Example 3–10? In your estimation, does the difference in the mass loading rate justify the cost of the larger basin required for inline storage?

3–18 Using the information given in the data table presented in Example 3–10 estimate the inline volume required to reduce the variation in the BOD mass loading rate between the peak and minimum from the existing ratio of 25.8.1:1 to a peak value of 5:1.

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4

Wastewater Treatment Process Selection, Design, and Implementation

WORKING TERMINOLOGY

Term	**Definition**
Bench-scale tests	Tests conducted in the laboratory with small quantities of samples to answer specific treatment questions.
Biosolids	Sludge from wastewater treatment processes that has been stabilized to meet the criteria in the U.S. EPA's 40 CFR 503 regulations and, therefore, can be used beneficially.
CCA	Critical component analysis. A method used to determine which mechanical components in the wastewater treatment plant will have the most immediate impact on effluent quality should failure occur.
Class A Biosolids	Biosolids that contain less than 1000 most probable number (MPN)/g of fecal coliforms and less than 3 MPN/4g of Salmonella bacteria and meet one of six stabilization alternatives given in 40 CFR 503. The material also must meet the pollutant limits and vector attraction reduction requirements set forth in 40 CFR 503.
ENRCCI	Engineering News-Record Construction Cost Index used to adjust construction cost information.
Flow diagram	The graphical representation of a particular combination of unit processes used to achieve specific treatment objectives.
Hydraulic profile	A graphical representation of the elevation of the free surface and hydraulic grade line of wastewater as it flows through the various treatment units.
NPDES	National Pollution Elimination Discharge System (NPDES), established based on uniform technological minimums with which each point source discharger has to comply.
Peaking factor	A factor that is applied customarily to average long term (typically flowrate and mass loading rate values) to determine maximum values.
Plant layout	The spatial arrangement of the physical facilities of the treatment plant identified in the flow diagram.
Pilot plant studies	Studies conducted at test beds at a scale larger than bench-scale, to establish the suitability of a process in the treatment of a specific wastewater under specific environmental conditions and to obtain data that can be used for full-scale design.
POTW	Publicly Owned Treatment Works.
Priority pollutants	Organic and inorganic compounds defined under Section 307 of Clean Water Act and listed in the Code of Federal Regulations at 40 CFR 401.15. Priority pollutants are selected on the basis of their known or suspected carcinogenicity, mutagenicity, teratogenicity, or high acute toxicity.
Privatization	Private sector ownership and operation of facilities and services used by government entities in performing their public function.
Process design criteria	The criteria used as the bases for sizing individual unit processes and their support systems.
Reliability, treatment process	Probability of adequate performance for a specified period of time under specified conditions or the percent of the time that effluent concentrations meet specified permit requirements.

Solids balance	The identification of the quantities of solids entering and leaving each unit process.
Value engineering	An intensive review of a project to determine best value, or value improvement, which may or may not result in cost reduction. Typically conducted at the 20 to 30 percent design stage of the project.
Variability, inherent	Based on the laws of chance, all physical, chemical, and biological treatment processes exhibit some measure of variability with respect to the performance that can be achieved. Variability is inherent in biological treatment processes.

Since the early 1900s, when the field of environmental engineering was in its infancy in the United States, there has been a steady evolution and development in the methods used for wastewater treatment. Descriptions of the many methods and variations that have been tried to date would fill several large volumes. The approach followed in this text is to identify and discuss basic principles and their application to wastewater treatment. The purpose of this chapter is to provide perspective on how the principles presented in Chaps. 1 through 3 and 5 through 18, along with other subjects not discussed in detail in this textbook, fit into the overall design, construction, operation and maintenance, and implementation of new and existing wastewater management projects. The following topics are covered: (1) planning considerations for wastewater treatment plants, (2) process selection, (3) treatment process reliability and selection of design values, (4) elements of process design, (5) implementation of wastewater management programs, and (6) financing.

4–1 PLANNING FOR NEW AND UPGRADING EXISTING WASTEWATER TREATMENT PLANTS

Most of the wastewater plants implemented using federal grant funding were designed using empirical design guidelines and were sized to accommodate long-term population growth (a 20 to 30 y timeline was the norm). Plants were located typically in remote areas close to water bodies that provided natural buffer zones between the treatment facilities and the communities they served. As sanitary sewer collection systems use gravity flow to collect wastewater, treatment plants were often located in lowlands adjacent to the receiving waters. In many cases, municipal growth has encroached on the natural buffer zones so that treatment facilities that once were considered remote are now in close proximity to the community. Going forward, new wastewater treatment plants will be required (1) to support population growth; (2) to deal with changing demographics; (3) to deal with changing wastewater characteristics, especially increasing wastewater concentrations as discussed in Chap. 3; (4) to meet new and more stringent effluent discharge standards; (5) to meet water reuse needs, including potable reuse; (6) to meet new stormwater management objectives; and (7) to replace existing aging infrastructure.

Need to Upgrade Existing Wastewater Treatment Plants

Over the next decade, many existing treatment plants, built during the 1970s and 1980s, will need to be upgraded to maintain treatment efficiency, to provide higher levels of treatment when discharging to environmentally sensitive receiving waters, to meet new water reuse opportunities, and to meet new stormwater management objectives. As more wastewater reuse is being implemented, additional treatment processes, including advanced oxidation, may have to be implemented. Treatment plants must also be upgraded aesthetically to minimize odors, noise, and visual impacts to the neighboring community.

Climate change considerations are also of concern, especially with recurring stormwater flooding due to short-duration high-intensity rainfall events. Sea level rise, along with stormwater surges due to increased stream channelization, currently threatens wastewater facilities in coastal communities constructed in low-lying areas. Issues related to sea level rise will become more serious in the future. Clearly the range of issues that must be considered in the upgrading of existing facilities goes beyond the specifics of process design.

Planning for New Wastewater Treatment Plants

During the late 1980s and continuing onward to the present time (2012), the focus on wastewater engineering has changed from the goals of meeting secondary standards for the removal of solids and organic matter to the specific requirements for constituent removal for health and environmental protection. During the next decade as additional demands are placed on the quality and quantity of the nation's water supplies necessitating water reuse, greater emphasis will be placed on maintaining high standards of wastewater treatment plant performance consistently and reliably. Without federal grant financing, local governments will expect that new facilities be implemented considering not only the initial capital investment but with greater emphasis on the long-term financial impact on the community's resources.

In planning for new wastewater treatment plants (WWTPs), consideration must be given not only to the physical facilities required to produce a high quality effluent, as discussed above, but also to opportunities to minimize operating costs associated with labor, energy (both electrical and for heating), and byproduct stabilization and disposal/reuse. Recognizing that municipal wastewater is inherently a resource, the recovery of energy, nutrients, and potable water from wastewater will become a major focus of the design of future WWTPs. Further, as scientific research continues in defining wastewater constituents that may cause adverse effects, greater levels of treatment will be needed, and, in some cases, will require the use of new technologies. It is anticipated that many of these new technologies currently under development will revolutionize the treatment of wastewater, including many of the approaches discussed in this textbook.

Treatment Process Design Considerations

The design of wastewater treatment facilities is an iterative process that involves consideration of all viable process alternatives that meet the treatment objectives and regulations, as discussed above. Design considerations that must be factored into process selection decisions to ensure that treatment facilities meet their treatment objectives, are financially sustainable, and are aesthetically integrated into the communities they serve, as discussed in this section. Examples of treatment processes that can be considered to reliably and cost-effectively meet these objectives are shown on Fig. 4–1. Important issues that must be considered in the planning and design of new treatment facilities are summarized in Table 4–1 and highlighted in the following discussion.

Liquid Streams. One of the most important issues that the designer of new plants in the United States will have to face is the exceedance criterion for effluent discharges described in Sec. 4–3. The focus of treatment plant design will be on the nature and variability of the wastewater constituents and the processes that will be required to meet the treatment objectives. In some cases, it may be necessary to develop source control programs to limit the effects of both flow and constituent concentrations for discharge to the collection system. The source control program may also include curbing excessive extraneous flows such as infiltration/inflow and wet weather discharges to the collection system.

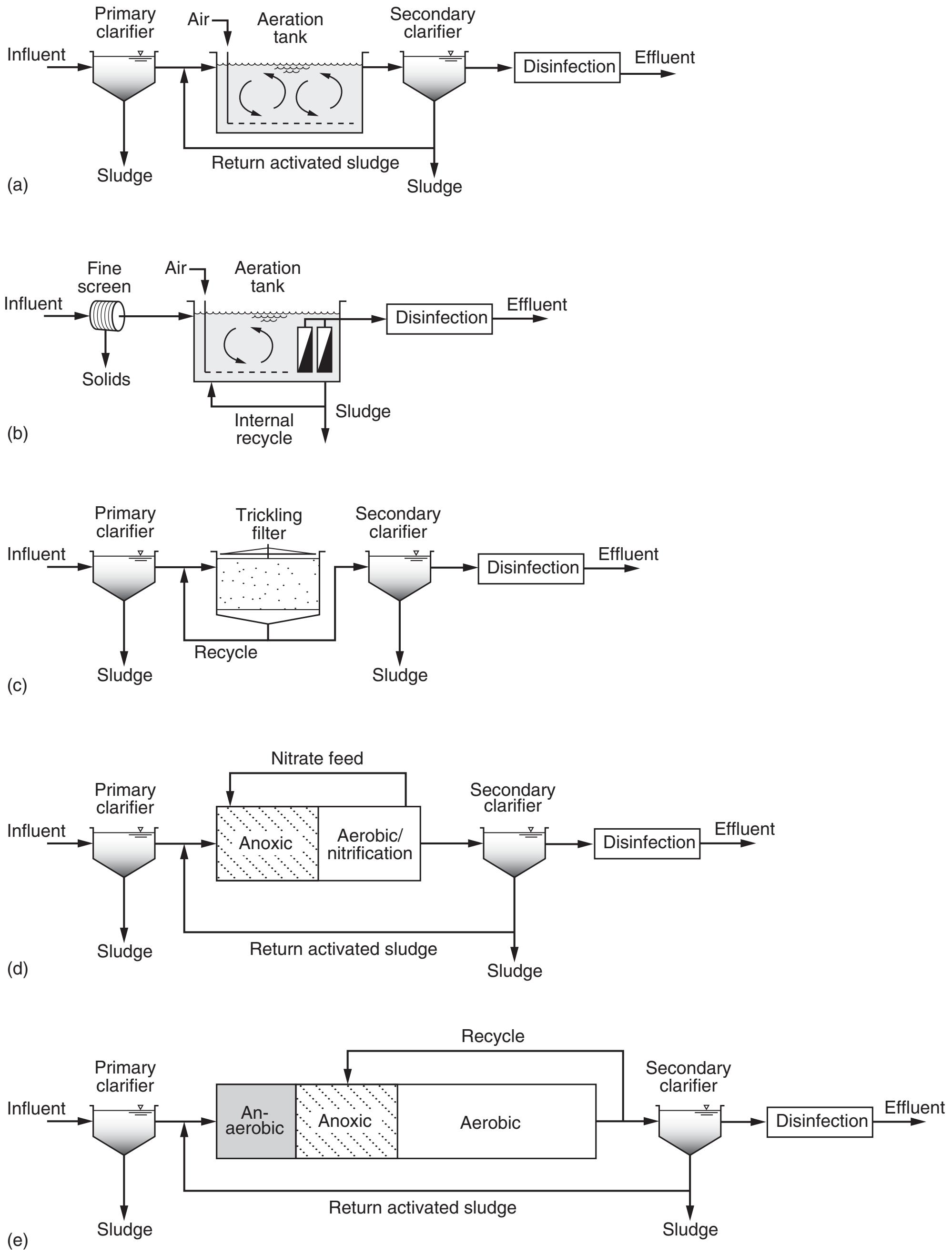


Figure 4–1

Generalized process flow diagrams for typical treatment processes all of which include disinfection for pathogen control: (a) activated sludge for TSS and BOD removal and nitrification; (b) membrane bioreactor for TSS and BOD removal and nitrification; (c) trickling filter for TSS and BOD removal; (d) suspended growth biological treatment for TSS, BOD, and nitrogen removal; and (e) suspended growth biological treatment for TSS, BOD, nitrogen, and phosphorus removal.

Table 4–1

Important considerations in the upgrading and design of new wastewater treatment plants

Issue	Description
Liquid stream processing	
Constituent and treatment process variability	New treatment processes should be designed to meet not-to-exceed permit limits taking into account constituent and treatment process variability
Flow equalization	Improved performance by eliminating flow surges through treatment facilities, especially wet-weather flows, typically required for reliably producing reclaimed water
Organic load equalization	Improved performance by equalizing organic loading rate to treatment processes throughout the day
Automatic process control	Provide provisions and facilities for automatic control of dissolved oxygen and solids retention time (SRT)
Enhanced disinfection for reuse	Improved and alternative technologies for disinfection and the control of disinfection byproducts
Advanced treatment processes	Processes for the removal of residual constituents and constituents not removed by conventional treatment
Conventional and advanced oxidation processes	Removal of specific constituents may require the use of advanced oxidation processes
Combined processes for specific constituents	To meet stringent not-to-exceed permit limits, two or more processes may have to be used in a series arrangement
Water reuse	Issues related to risk assessment will have to addressed
Treatment of wet-weather flows	Less costly to treat at WWTP as compared to individual treatment facilities at individual overflow locations
Energy management	Implementation of enough physical facilities, such as flow equalization, to allow for improved utilization of off-peak power for wastewater treatment
Solids processing	
Improved screening	Improved screening to remove extraneous materials that end up in the biosolids or foul treatment equipment such as aeration diffusers
Grit removal	Removal of grit that can settle in primary sedimentation tanks and digesters
Enhanced pathogen control	Enhanced pathogen control is needed to produce class A biosolids
Enhanced vector control	Enhanced vector control is needed to produce class A biosolids
Separate treatment of return flows	To improve the performance of liquid waste stream, especially with respect to the removal of nitrogen
Odor control	
Odor control in collection system	Implementation of source control program to minimize odors generated in collection system. Use of pure oxygen (see Fig. 16–3 in Chap. 16)
Odor formation in treatment facilities	Careful attention to hydraulic design to avoid dead zones with respect to flow, handling of return flows from solids processing facilities
Odor containment	Cover facilities to eliminate odors
Odor treatment	Separate or combined treatment of odorous gases from odor containment facilities
Process control	
Computer simulation models	Development of improved operational strategies through the use of mathematical simulation models
Pilot-plant testing	Ongoing program of testing new technologies

Upstream flow equalization may also be required. Facilities to handle in-plant variations of flows and loads including return flows will also have to be considered. Water reuse will also impose additional constraints on effluent management.

Solids Processing. In selecting the appropriate methods of solids processing, reuse, and disposal, consideration must be given to the appropriate regulations. In the United States, pollutant numerical limits and management practices for the reuse and disposal of solids generated from the processing of municipal wastewater are defined in the 40 CFR Part 503 regulations. The regulations are designed to protect public health and the environment from any reasonably anticipated adverse effects of pollutants contained in the biosolids. As many communities investigate options for solids processing, reuse, and disposal, greater emphasis is being placed on producing a cleaner product and meeting the Class A biosolids requirements. Important issues that have to be considered in solids processing are discussed in Chaps. 13 and 14. Control of pathogens and vectors is of special importance for the protection of public health.

Treatment of Return Flows and Recovery of Nutrients. As discharge standards for nitrogen and phosphorus become more stringent, it is clear that the practice of introducing return flows into the liquid processing facilities, typically before biological treatment, must be eliminated or modified if stringent nitrogen and phosphorus limits are to be achieved. Although flow equalization can be used to reduce the impact of return flows on treatment process performance, nutrient recovery can be a more effective solution. The subject of return flow treatment and nutrient recovery is considered in detail in Chap. 15.

Treatment of Wet-Weather (Stormwater) Flows. When new plants are being planned and existing plants are being upgraded, an important component of the analysis will be the management and treatment of wet-weather flows to minimize or eliminate untreated overflows. One of the approaches to controlling discharge of untreated overflows from the collection system is to maximize conveyance of wet-weather flows to the WWTP. Compared to other options, treatment of excess wet-weather flow at a new or existing WWTP can be an attractive option. Maximizing the use of unit processes at the WWTP can avoid or reduce the cost and impacts associated with controlling overflows at or near the point where excess flow is diverted in the collection system. While there are limitations on peaking factors that can be handled by WWTPs, particularly the biological treatment processes, options for WWTPs to handle increased quantities of wet-weather flow include (1) providing offline flow equalization storage to hold excess wet weather flow during periods of high flow, with the excess flow blended back into the treatment processes when flows drop and capacity is available; (2) increasing or expanding the capacities of conventional, continuous-flow unit processes, and either processing all flow through all unit processes or employing flow blending during peak flow periods; and (3) providing intermittent-use wet-weather treatment units that can process excess wet-weather flow which is blended with effluent processed through conventional, dry weather processes at the WWTP. Intermittent treatment facilities for wet-weather flows are considered in Sec. 18–5 in Chap. 18.

Control of Emissions. Gaining public acceptance for siting wastewater treatment plants will depend, to a large extent, on meeting community concerns over odors. The prevention, control, and treatment of odors are now mandatory parts of any treatment plant design (see Fig. 4–2). With proper planning for odor management, the need to correct odor

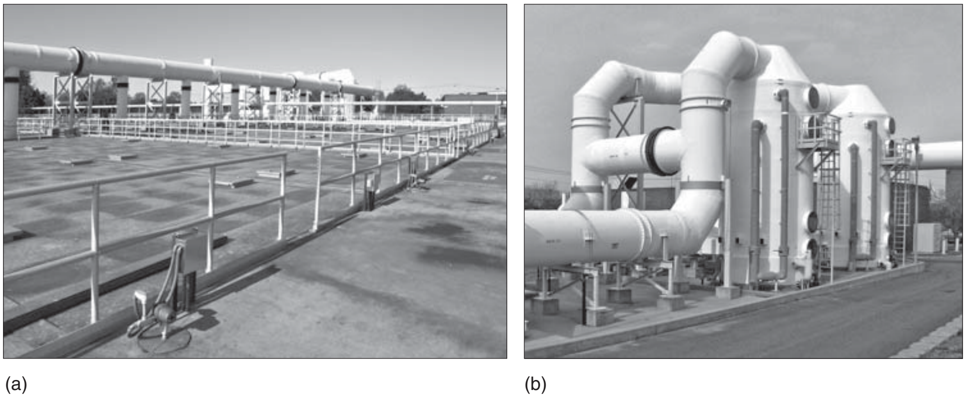

(a) (b)

Figure 4–2

Typical odor control facilities for new and existing wastewater treatment plants: (a) covered primary sedimentation facilities (b) odor control air scrubbers to treat odors from covered primary sedimentation tanks.

related issues after the fact and restore public confidence in facilities operation can be avoided. Important considerations in the control of emissions are discussed in Chap. 16.

Owner Needs

A factor often overlooked in the selection of a treatment process is the needs of the owner of the facilities. Owner needs may take the form of limitations of cost and the ability to pay for the project, operating capabilities where existing staff will be utilized, process preferences based upon personal experience, concerns about using proven processes or equipment that are not experimental, and concern about possible environmental impacts. Owner needs are especially important in small communities where there is a limited history of construction and operation of new treatment systems. For projects both large and small, it is important for the design engineer and the owner to reach an understanding about their mutual goals and objectives so that the needs of the owner are satisfied and the selected treatment process improvements meet the basic purposes for its selection, that is, meeting waste discharge regulations or effluent reuse in the most cost-effective manner and mitigating adverse environmental impacts.

Asset management is a management process to minimize the total cost of owning and operating the infrastructure assets while delivering the service levels needed. While it is not required to implement asset management in the United States, the long-term benefits have been recognized, and many organizations have been promoting the use of asset management. Asset management is discussed further in Chap. 18.

Environmental Considerations

The environmental impacts of a proposed wastewater treatment facility are as important as cost considerations if not more so. In addition to the potential impact of discharged effluent on the aquatic environment in the receiving waters, increasing efforts are being made to address the emission of greenhouse gas (GHG) from the treatment facility. These considerations are included in the triple bottom line (TBL) analysis which evaluates (1) economic, (2) environmental, and (3) social aspects of the project as part of a decision making process. While detailed environmental review procedures are not covered in this text, GHG emissions are considered in Chap. 16 and TBL is discussed briefly in Chap. 18.

National Environmental Policy Act. The protocol for evaluation of environmental impacts is set forth in the National Environmental Policy Act (NEPA) of 1969 (42 USC 4321–4347 as amended). Environmental evaluations should focus on social, technical, ecological, economic, political, legal, and institutional (STEEPLI) criteria.

Application of the NEPA regulations requires that an Environmental Impact Statement (EIS) be prepared for any proposed federal action which is determined to have a significant impact on the quality of the human environment. The development of an EIS is controlled by the Council on Environmental Quality (CEQ) Regulations for Implementing the Procedural Provisions of the National Environmental Policy Act (40 CFR 1500-1508).

The NEPA regulations ensure that the probable environmental effects are identified, that a reasonable number of alternative actions and their environmental impacts are considered, that the environmental information is available for public understanding and scrutiny, and that the public and governmental agencies participate as a part of the decision process. All pertinent regulations and the inherent protection afforded must be disclosed in the EIS. The NEPA neither prohibits nor permits any action, but requires full disclosure of environmental information and public participation in the decision-making process.

Environmental Information Document. The procedures and requirements for implementing the NEPA regulations for Municipal Wastewater Treatment Construction Grants Program under the Clean Water Act are set forth in Subpart E of the EPA regulations. The basic elements of the process include the Environmental Information Document (EID) generated by the grantee (owner) as an integral part of a facilities plan, consistent with Section 201 of the Clean Water Act. The EID is the basis for agency review of the environmental impacts of the facilities plan and preparation of an Environmental Assessment (EA). The EA must be of sufficient detail so as to be an adequate basis for EPA's independent review and decision to issue a Finding of No Significant Impact (FNSI) or to issue a notice of intent for an EIS and subsequent Record of Decision. If an EIS is required, then following development of a draft EIS and input based on public hearings, a final EIS is prepared. In the resultant Record of Decision the findings and the recommended actions selected are summarized.

Compatibility with Existing Facilities

An important consideration not to be overlooked in the expansion and upgrading of existing wastewater treatment facilities is the compatibility with existing process units including the impact of the new process on plant hydraulics (both at peak and low flow), impact on other unit processes, structural impacts, and impact on existing instrumentation and control system. The introduction of a new process into an existing facility represents new operating requirements and additional hiring and training of personnel for the proper operation and maintenance of a new process unit. Construction of the new facilities while operating the existing wastewater treatment plant to meet the discharge/reuse standards is another challenge. Often, equipment furnished by the same manufacturer as the existing installation permit fewer spare parts to be kept on hand, provided the equipment has a good record of service.

Energy and Resource Requirements

Concern over the rate of consumption of natural resources and energy has increased as shortages have occurred and worldwide demands have been increased. Because the operation of wastewater management facilities depends on energy resources to a large extent, it is important to appraise the requirements realistically. The operation of facilities is the main consumer of energy at treatment plants. Because energy consumption of different unit processes varies greatly and because there are innumerable combinations possible, data must be available for each prospective treatment process considered. Additionally, many opportunities exist for generating heat and power at the treatment plant that can be utilized to meet all or a part of the plant's demands. The recovery of heat energy from wastewater is also being incorporated both in collection systems (especially in Germany) and in new wastewater treatment plant designs. Considerations for energy management are further discussed in Chap. 17.

Energy Sources. The main energy sources are (1) electric power, (2) natural gas or propane, (3) diesel fuel or gasoline, and (4) recovered heat and power (see Chap. 17). Electric power is used mainly for running the electric motors for the process equipment and for providing lighting and power for various ancillary support systems. Natural gas or propane is used for building and digester heating and is used as a fuel source for standby engine-generators. Diesel fuel or gasoline is used similarly for standby engine-generators and for vehicle fuel. Recovered heat and power can meet all or part of the demands. Particular attention needs to be paid to the electrical energy costs because of the complex pricing structure used by utilities. In northern climates, where unit processes need to be covered, significant capital and operating costs can be associated with ventilation and heating.

Electric Energy Costs. Electric energy charges are commonly assessed based upon energy use, power factor charges, and demand charges. Power factor charges are concerns for plants having large electric motor–driven equipment. The demand charges are assessed by the utility companies when they commit sufficient power generating capacity to meet the entire demands of the treatment system. Peak power use for as little as 15 min may establish a demand charge for up to 12 mo. Demand charges can be reduced in some instances by providing power generating capability at the treatment plant. The recovery and use of digester gas for meeting energy needs and reducing demand is one example how both user charges and demand charges can be reduced with resulting cost savings to the treatment plant. Digester gas use is discussed in more detail in Chap. 13. As part of an energy cost evaluation, a sensitivity analysis should be considered to assess the impacts of future changes in energy costs on the overall cost of operation for the treatment facilities.

Cost Considerations

Of major significance in the selection and design of alternative wastewater treatment facilities, especially to the client, is the question of costs—not only initial construction costs but also annual operation and maintenance costs. Although cost estimating is not covered in this text, a few comments about the preparation of cost estimates are appropriate. Ordinarily, cost estimates are divided into three levels of detail: (1) order of magnitude estimates that are used for conceptual planning and are derived from cost curves and selected publications; (2) budget estimates (prepared during the preliminary design stage) derived from published or historical bid information, manufacturers' quotations, or limited quantity takeoffs; and (3) definitive estimates derived from detailed quantity takeoffs of completed plans and specifications. The accuracy of the estimates vary according to the level of detail; therefore, contingencies of varying percentages are added to the estimates to account for undefined items and for unforeseen conditions.

Construction Cost Estimates. When preparing estimates of construction cost particularly during times of high inflation, the same basis of comparison should be used to evaluate all the alternatives and for projecting future costs. Methods commonly used for projecting costs are (1) escalation based on an assumed rate of inflation or (2) a published cost index. The Engineering News-Record Construction Cost Index (ENRCCI), published in the magazine ENR (a McGraw-Hill publication) is used most commonly in the field of wastewater engineering.

Data in engineering reports and in the literature can be adjusted to a common basis for purposes of comparison by using the following relationship:

$$\text{Current cost} = \frac{\text{Current value of index}}{\text{Value of index at time of estimate}} \times \text{Estimated cost} \tag{4–1}$$

When possible, index values should also be adjusted to reflect current local costs. Both the ENRCCI and EPA indexes include costs for various geographical locations. The ENR publishes cost indexes for 20 cities. When using the ENRCCI, if the month of the year that the facilities were built is not given, it is common practice to use the June end-of-the-month index value. To project costs into the future, the following relationship can be used. The future value of the index is often projected to the one-third or mid-point of the construction period.

$$\text{Future cost} = \frac{\text{Projected future value of index}}{\text{Current value of index}} \times \text{Current cost} \tag{4–2}$$

It should be noted, however, that updating or projecting costs for periods of more than 3 to 5 y can result in gross inaccuracies, especially if the index has increased or decreased significantly.

Operations and Maintenance Cost Estimates. The annual costs for operations and maintenance (O&M) are important factors in the evaluation of alternative treatment processes. The principal elements of O&M costs are labor, energy, chemicals, and materials and supplies. Where possible, each of these elements should be estimated separately because costs of each may escalate at different rates. Energy costs should be estimated based on the estimated energy consumption by the process equipment and the appropriate energy rate obtained from the utility furnishing the energy. Chemical costs should be computed similarly based on the estimated amounts consumed and the appropriate unit price. Materials and supplies are estimated on predicted usage and should be included.

Cost Comparisons. In the evaluation of alternative treatment process flow diagrams, costs may be compared using present worth, total annual costs, or life cycle costs. In a present worth analysis, all future expenditures are converted to a present worth cost at the beginning of the planning period. A discount rate is used in the analysis and represents the time value of money (the ability of money to earn interest). In a total annual cost comparison, the capital costs are amortized based on probable interest rates for bonds and the duration of the bond issue. The annual fixed (amortized) cost is added to the annual operating and maintenance costs to determine the total annual cost. Life cycle costs are used to determine the total cost of a facility over its total useful life (up to 50 y for structures) and include the capital cost and the operating and maintenance costs. Life cycle costs are particularly useful in comparing the costs of a rehabilitated existing facility as compared to a new facility. However, when comparing costs between treatment alternatives, consideration should also be given to the level of accuracy of the estimate. For example, if the cost estimate accuracy carries a contingency of 40 percent, and all options are within 5 to 8 percent of each other, then the options could be considered equal.

Other Design Considerations

Additional important design considerations include (1) equipment availability, and (2) personnel requirements. Although a detailed discussion of equipment availability and personnel requirements is beyond the scope of this textbook, these subjects are introduced briefly below to provide a perspective on how they fit into the overall picture.

Equipment Availability. The availability of equipment plays an important part in process selection because of (1) the need to provide redundant systems when there are long delivery times for spare parts and replacement units and (2) when equipment delivery is critical to the construction schedule. Most of the equipment used in wastewater treatment is custom manufactured, other than for items such as small pumps, motors, and valves. Some items of equipment may be manufactured from alloy materials such as stainless steel that require special manufacturing techniques or are proprietary and only available from limited sources, perhaps even from overseas suppliers. Therefore, the design engineer should consider carefully the equipment components that make up the process or system to determine their potential effects upon the design, construction, and operation and maintenance of the facilities.

Personnel Requirements. In the selection of treatment processes, consideration must be given both to the numbers of operating and maintenance personnel needed and to the skills that may be required. The simpler and less complex the process, the fewer the number of highly skilled people are needed. Where facilities are being added to an existing treatment plant, capabilities of the existing personnel should be evaluated so that the new facilities can be added without causing major staffing problems and the need for extensive retraining.

Some of the more complex processes require high levels of automatic controls utilizing electronic instruments and devices (see Chap. 18). Proper instrumentation and controls can save labor and even allow some of the small plants to operate unattended. However, complex instrumentation and control systems may require the on-staff services of highly skilled instrumentation technicians. Instrumentation specialists may be difficult to recruit and maintain on staff because of the high demand for well-qualified technicians. The extent and complexity of the control systems and the staffing levels required have to be evaluated carefully.

4–2 CONSIDERATIONS IN PROCESS SELECTION

Process selection involves the detailed evaluation of the various factors that must be considered when evaluating unit processes and other treatment methods to meet current and future treatment objectives. The purpose of process analysis is to select the most suitable unit processes and the optimum operational criteria. The purpose of this section is to introduce the important factors that must be considered in process selection and to consider the basis for process design. The impact of treatment plant reliability on the selection of specific treatment process design criteria is examined in the following section.

Important Factors in Process Selection

The most important factors that must be evaluated in process analysis and selection are identified in Table 4–2. Each factor is important in its own right, but some factors require additional attention and explanation. The first factor, "process applicability," stands out above all others and reflects directly upon the skill and experience of the design engineer. Many resources are available to the designer to determine applicability, including past experience in similar type projects. Available resources include performance data from operating installations, published information in technical journals, manuals of practice published by the Water Environment Federation (WEF, 2010b), process design manuals published by WEF and U.S. EPA, and the results of pilot plant studies.

Table 4–2

Important factors that must be considered when evaluating and selecting unit processes

Factor	Comment
1. Process applicability	The applicability of a process is evaluated on the basis of past experience, data from full-scale plants, published data, and from pilot plant studies. If new or unusual conditions are encountered, pilot plant studies are essential.
2. Applicable flow range	The process should be matched to the expected range of flowrates. For example, stabilization ponds are not suitable for extremely large flowrates, in highly populated areas.
3. Applicable flow variation	Most unit operations and processes have to be designed to operate over a wide range of flowrates. Most processes work best at a relatively constant flowrate. If the flow variation is too great, flow equalization may be necessary.
4. Influent wastewater characteristics	The characteristics of the influent wastewater affect the types of processes to be used (e.g., chemical or biological) and the requirements for their proper operation.
5. Inhibiting and unaffected constituents	What constituents are present and may be inhibitory to the treatment processes? What constituents are not affected during treatment?
6. Climatic constraints	Temperature affects the rate of reaction of most chemical and biological processes. Temperature may also affect the physical operation of the facilities. Warm temperatures may accelerate odor generation and also limit atmospheric dispersion.
7. Process sizing based on reaction kinetics or process loading criteria	Reactor sizing is based on the governing reaction kinetics and kinetic coefficients. If kinetic expressions are not available process loading criteria are used. Data for kinetic expressions and process loading criteria usually are derived from experience, published literature, and the results of pilot plant studies.
8. Process sizing based on mass transfer rates or process loading criteria	Reactor sizing is based on mass transfer coefficients. If mass transfer rates are not available process loading criteria are used. Data for mass transfer coeficients and process loading criteria usually are derived from experience, published literature, and the results of pilot plant studies.
9. Performance	Performance is usually measured in terms of effluent quality and its variability, which must be consistent with the effluent discharge requirements.
10. Treatment residuals	The types and amounts of solid, liquid, and gaseous residuals produced must be known or estimated. Often, pilot plant studies are used to identify and quantify residuals.
11. Sludge processing	Are there any constraints that would make sludge processing and disposal infeasible or expensive? How might recycle loads from sludge processing affect the liquid unit operations or processes? The selection of the sludge processing system should go hand-in-hand with the selection of the liquid treatment system.
12. Environmental constraints	Environmental factors, such as prevailing winds and wind directions and proximity to residential areas, may restrict or affect the use of certain processes, especially where odors may be produced. Noise and traffic may affect selection of a plant site. Receiving waters may have special limitations, requiring the removal of specific constituents such as nutrients.
13. Chemical requirements	What resources and what amounts must be committed for a long period of time for the successful operation of the unit operation or process? What effects might the addition of chemicals have on the characteristics of the treatment residuals and the cost of treatment?
14. Energy requirements	The energy requirements, as well as probable future energy cost, must be known if cost-effective treatment systems are to be designed.
15. Other resource requirements	What, if any, additional resources must be committed to the successful implementation of the proposed treatment system using the unit operation or process being considered?
16. Personnel requirements	How many people and what levels of skills are needed to operate the unit operation or process? Are these skills readily available? How much training will be required?

(*continued*)

Table 4–2 (*Continued*)

Factor	Comment
17. Operating and maintenance requirements	What special operating or maintenance requirements will need to be provided? What spare parts will be required and what will be their availability and cost?
18. Ancillary processes	What support processes are required? How do they affect the effluent quality, especially when they become inoperative?
19. Reliability	What is the long-term reliability of the unit operation or process being considered? Is the operation or process easily upset? Can it stand periodic shock loadings? If so, how do such occurrences affect the quality of the effluent?
20. Complexity	How complex is the process to operate under routine or emergency conditions? What levels of training must the operators have to operate the process?
21. Compatibility	Can the unit operation or process be used successfully with existing facilities? Can plant expansion be accomplished easily?
22. Adaptability	Can the process be modified to meet future treatment requirements?
23. Economic life-cycle analysis	Cost evaluation must consider initail capital cost and long-term operating and maintenance costs. The plant with lowest Initail capital cost may not be the most effective with respect to operating and maintenance costs. The nature of the available funding will also affect the choice of process
24. Land availability	Is there sufficient space to accommodate not only the facilities currently being considered but possible future expansion? How much of a buffer zone is available to provide landscaping to minimize visual and other impacts?

Where the applicability of a process to a given situation is unknown or uncertain, pilot plant studies must be conducted to determine performance capabilities and to obtain design data upon which a full-scale design can be based. The following discussion will deal briefly with process design based on reaction kinetics, mass transfer, and the use of loading criteria. As part of the discussion, the conduct of bench and pilot plant studies is considered along with process variability. The other factors in Table 4–2 are discussed throughout the remainder of the book. They are identified here to indicate the diverse nature of the information that must be available to make a proper evaluation of unit processes used for the treatment of wastewater.

Process Selection Based on Reaction Kinetics

In process selection and sizing based on reaction kinetics, particular emphasis is placed on defining the nature of the reactions occurring within the process, the appropriate values of the kinetic coefficients, and the selection of the reactor type.

Selection of Appropriate Kinetic Rate Expression(s) and Coefficients. The nature of the reactions occurring within a process must be known to apply the reaction kinetics approach to design. For example, it is of critical importance to know if the reaction is zero, first, retarded first, or second order, or if the reaction is a saturation type. Reaction rates are discussed in Sec. 1–10 in Chap. 1. Selection of appropriate kinetic rate coefficients for the process that is to be designed is also based on (1) information obtained from the literature, (2) experience with the design and operation of similar systems, or (3) data derived from pilot plant studies. In cases where significantly different wastewater characteristics occur or new applications of existing technology or new processes are being considered, pilot plant testing is recommended. The various rate expressions that have been developed for biological treatment are considered in Chaps. 7 through 10. In addition, there are a number of commercially available models of the activated sludge process that take into account myriad of variables. Although comprehensive, the key issue with these models is to understand what the model output means. For example, does the output seem reasonable, real, or defensible? The subject of model output assessment is considered further in Chaps. 8 and 9.

Selection of Reactor Types. Operational factors that must be considered in the type of reactor or reactors to be used in the treatment process include (1) the nature of the wastewater to be treated, (2) the nature of the reaction kinetics governing the treatment process, (3) special process requirements, and (4) local environmental conditions. As noted previously, for biological treatment with the activated sludge process, for zero-order kinetics, there is no difference in the size of the reactor required (i.e., $V_{\text{complete-mix}} = V_{\text{plug-flow}}$). For example, a complete-mix reactor might be selected over a plug-flow reactor, because of its dilution capacity, if the influent wastewater is known to contain toxic constituents that cannot be removed by pretreatment. Alternatively, a plug-flow or multistage reactor might be selected over a complete-mix reactor to control the growth of filamentous microorganisms. In practice, the construction costs and operation and maintenance costs also affect reactor selection.

Process Selection Based On Mass Transfer

In addition to process selection based on reaction kinetics and loading criteria, a number of treatment processes will be based on mass transfer considerations, as introduced in Chap. 1. The principal operations in wastewater treatment involving mass transfer are aeration, especially the addition of oxygen to water; the drying of biosolids and sludge; the removal of volatile organics from wastewater; the stripping of dissolved constituents such as ammonia from digester supernatant; and the exchange of dissolved constituents as in ion exchange. Fortunately, there is a considerable body of literature on these subjects as well as a vast amount of practical experience. Additional details on these subjects are presented in the subsequent chapters.

Process Design Based on Loading Criteria

If appropriate reaction rate expressions and mass transfer coefficients cannot be developed, generalized loading criteria are used frequently. Early design loading criteria for activated sludge biological treatment systems were based on aeration tank capacity [e.g., kg of BOD/m^3 (lb BOD/10^3 ft^3)]. For example, if a process that is loaded at 10 kg/m^3 produces an acceptable effluent and one loaded at 20 kg/m^3 does not, the successful experience tends to be repeated. Unfortunately, records often are not well maintained, and the limits of such loading criteria are seldom defined. Examples of loading criteria are presented in the design chapters for unit processes. It should be noted that with the new activated sludge biological treatment process variations and new aeration equipment, the use of loading factors should be avoided.

Bench-Scale Tests and Test-Bed Pilot-Scale Studies

Where the applicability of a process for a given situation is unknown, but the potential benefits of using the process are significant, bench-scale or pilot-scale tests must be conducted. Bench-scale tests are conducted in the laboratory with small quantities of the wastewater in question. Pilot-scale tests are conducted typically with flows that can vary from 0.1 to 5 percent of the design flows. It is important to note that the term *pilot-scale*

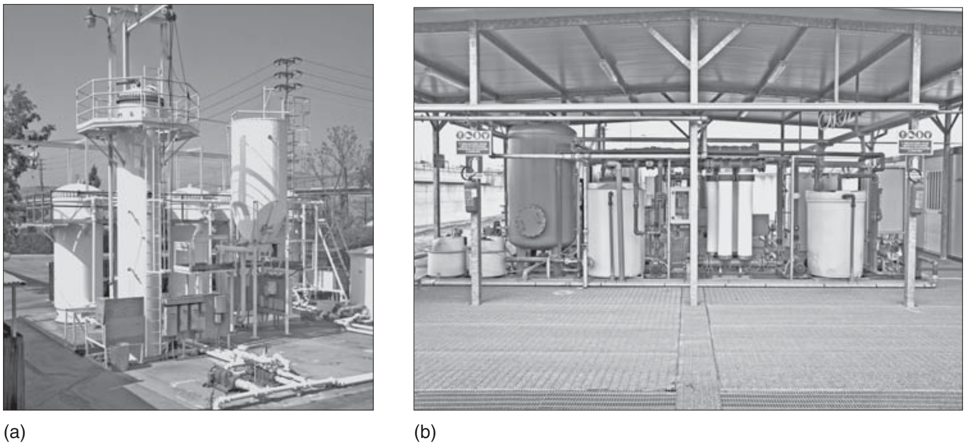

(a) (b)

Figure 4–3

Typical example of a test-bed facility where a number of different technologies can be tested at pilot-scale: (a) test bed (a.k.a. pilot plant) for Pomona Virus Study conducted by the County Sanitation Districts of Los Angeles County (circa 1975–1977) and (b) test bed facility, Toranto, Italy.

is used typically to denote hydraulic capacity of the facility and not necessarily the scale of the physical facilities being tested. For example, a full-scale membrane filtration system may be comprised of 500 microfiltration units, whereas a pilot-scale facility might have 10 units or less. Where scale-up issues and computational methods are too complex, the individual units being tested are the same as those that will be used in the full-scale installation. The term *test-bed* is used to describe a physical facility (also a geographic location, urban area, or city) where technologies and concepts can be tested and evaluated. An example, a pilot-scale (i.e., reduced flow) test-bed facility for the evaluation of advanced treatment technologies, is illustrated on Fig. 4–3.

The purpose of conducting pilot-plant studies is to establish the suitability of the process in the treatment of a specific wastewater under specific environmental conditions and to obtain the necessary data on which to base a full scale design. Factors that should be considered in planning pilot-plant studies for wastewater treatment are presented in Table 4–3. The relative importance of the factors presented in Table 4–3 will depend on the specific application and the reasons for conducting the testing program. For example, testing of UV disinfection systems is typically done (1) to verify manufacturers performance claims, (2) to quantify effects of effluent water quality constituents on UV performance, (3) to assess the effect(s) of system and reactor hydraulics on UV performance, (4) to assess the effect(s) of effluent filtration on UV performance, and (5) to investigate photoreactivation and impacts.

Wastewater Discharge Permit Requirements

In most wastewater discharge permits, effluent constituent requirements are based on 7-d and 30-d average concentrations. Because wastewater treatment effluent quality is variable for a number of reasons (varying organic loads, changing environmental conditions, etc.), it is necessary to ensure that the treatment system is designed to produce effluent concentrations equal to or less than the permit limits. Two approaches in process selection and design are (1) the use of arbitrary safety factors, and (2) statistical analysis of treatment plant performance to determine a functional relationship between effluent quality and the probable frequency of occurrence. The latter approach, termed the "reliability concept," is preferred because it can be used to provide a consistent basis for analysis of uncertainty and a rational basis for the analysis of performance and reliability. Treatment process reliability is considered in the following section.

Table 4–3

Considerations in setting up pilot-plant testing programs

Item	Consideration
Reasons for conducting pilot testing	Test new process Simulation of another process Predict process performance Document process performance Optimize system design Satisfy regulatory agency requirements Satisfy legal requirements
Pilot plant size	Bench or laboratory-scale model Pilot-scale tests Full- (prototype) scale tests
Nonphysical design factors	Available time, money, and labor Degree of innovation and motivation involved Quality of water or wastewater Location of facilities Complexity of process Similar testing experience Dependent and independent variables
Physical design factors	Scale-up factors Size of prototype Flow variations expected Facilities and equipment required and setup Materials of construction
Design of pilot testing program	Dependent variables including ranges Independent variables including ranges Time required Test facilities Test protocols Statistical design of data acquisition program Phased approach to adjust protocol as data are collected and analyzed

4–3 TREATMENT PROCESS RELIABILITY AND SELECTION OF DESIGN VALUES

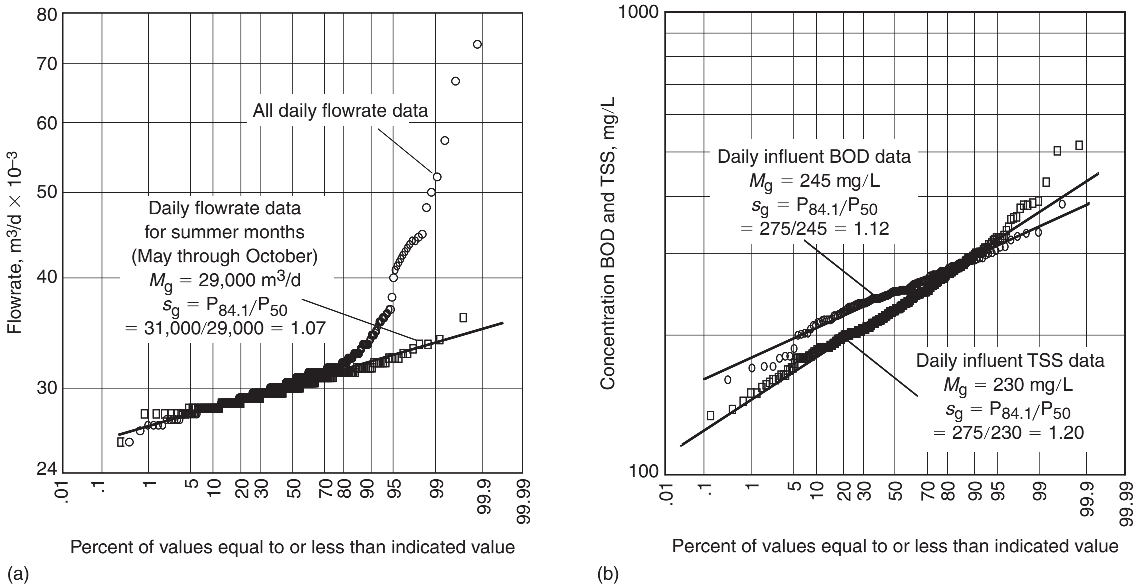
Important factors in process selection and design are treatment plant performance and reliability in meeting permit requirements. Reliability of a treatment plant or a treatment process may be defined as the probability of adequate performance for a specified period of time under specified conditions or, in terms of treatment plant performance, the percent of the time that effluent concentrations meet specified permit requirements. For example, a treatment process with a reliability of 99 percent is expected to meet the performance requirements 99 percent of the time. For one percent of the time, or three to four times per year, the not-to-exceed permit limits are expected to be exceeded. Such a level of performance may or may not be acceptable, depending on the permit requirements. For each specific case where the reliability concept is to be employed, the levels of reliability must be evaluated, including the cost of the facilities required to achieve the specified levels of reliability and the associated operating and maintenance costs. Thus, the purpose of this section is to examine how treatment process variability is assessed and how the performance of combined processes can be evaluated. The specific topics to be discussed are (1) variability in wastewater treatment, (2) selection of process design parameters, (3) the performance of combined processes, and (4) the development of input-output relationships.

Variability in Wastewater Treatment

Three categories of variability that can affect the design, performance, and reliability of a wastewater treatment plant are (1) variability of the influent wastewater flowrate and characteristics, (2) inherent variability in wastewater treatment processes, and (3) variability caused by mechanical breakdown, design deficiencies, and operational failures. Following a brief discussion of the characterization of variability in wastewater treatment, each of the above categories of variability is considered in the following discussion. Variability of the influent wastewater flowrate and characteristics was discussed in Chap. 3, and is reviewed here briefly for completeness.

Characterization of Parameter and Process Variability. One common method used to characterize the variability of wastewater parameters and treatment processes is the use of the geometric standard deviation, s_g (see Appendix D). The value of s_g can be used to approximate an entire distribution of all expected values if a mean value is known or can be estimated. As discussed in Appendix D, the greater the numerical value of s_g, the greater the observed range in the measured values.

Variability of Influent Wastewater Flowrates. As discussed in Sec. 3–3 in Chap. 3, the influent flowrate to a treatment facility is dependent on factors such as the time of day, season, size and characteristics of the contributing population, and infiltration to and exfiltration from the collection system. A typical example of the variability that can be observed in influent wastewater flowrates is illustrated on Fig 4–4(a).



(a) (b)

Figure 4–4

Probability distributions for daily influent wastewater characteristics collected over one year: (a) flowrate and (b) biochemical oxygen demand (BOD) and total suspended solids (TSS).

Table 4–4

Range of geometric standard deviations (s_g) for influent parameters observed at small, intermediate, and large wastewater treatment facilities

	Range of s_g for typical wastewater treatment facilities[a]					
	Small[b]		Intermediate[c]		Large[d]	
Parameter	Range	Typical	Range	Typical	Range	Typical
Flowrate	1.4–2.0	1.6	1.1–1.5	1.25	1.1–1.2	1.15
BOD	1.4–2.1	1.6	1.3–1.6	1.3	1.1–1.3	1.27
COD	1.5–2.2	1.7	1.4–1.8	1.4	1.1–1.5	1.30
TSS	1.4–2.1	1.6	1.3–1.6	1.3	1.1–1.3	1.27

[a] Excluding systems with large amounts of infiltration in the collection system.

[b] Flowrate of 4000–40,000 m^3/d.

[c] Flowrate of 40,000–400,000 m^3/d.

[d] Flowrate $>$ 400,000 m^3/d.

As shown on Fig. 4–4(a), the summer flowrates are very stable and follow a log-normal distribution, whereas the entire daily set of flowrate data is influenced significantly by the high winter flowrates and is, therefore, extremely variable. In fact, the complete daily flowrate data set cannot be modeled with either an arithmetic or a log-normal distribution. As will be discussed subsequently, such variability, which is not uncommon, is of concern where stringent discharge requirements must be met. In some cases, it may be necessary to reduce the amount of infiltration/inflow in the collection system and/or install flow equalization facilities (discussed in Chap. 3) to improve treatment process performance. The typical range of observed values for s_g for influent flowrates for small, medium, and large capacity wastewater treatment plants are given in Table 4–4.

Variability in Constituent Concentrations. The variability of the constituents in wastewater, discussed previously in Sec. 3–3, must be considered carefully in the design of biological treatment processes, especially with respect to the design of the aeration facilities [see Fig 4–4(b)]. Geometric standard deviation values for the variability observed in influent BOD, COD, and TSS constituent concentrations are given in Table 4–4 in terms of the s_g value. The range of s_g values given in Table 4–4 corresponds to the range of values reported in the literature and in the authors' experience.

Inherent Variability In Wastewater Treatment Processes. All physical, chemical, and biological treatment processes exhibit some measure of variability with respect to the performance that can be achieved. Typical ranges of mean effluent constituent values that can be achieved with various biological treatment processes are reported in Table 4–5. The range of variability observed in the effluent BOD and TSS performance of various activated sludge processes is illustrated graphically on Fig. 4–5. Further, as illustrated on Fig. 4–5 and discussed in Chap. 8, the physical characteristics of the secondary sedimentation facilities can have a significant impact on the observed performance of the activated sludge process. The variability in terms of s_g values for BOD, TSS, and turbidity for the three processes considered in Table 4–5 are presented in Table 4–6. The range of s_g values is representative of the values reported in the literature. Use of the data in Table 4–6 is illustrated in Example 4–1.

Table 4–5

Typical range of effluent quality after secondary treatment[a]

			Range of effluent quality after indicated treatment		
Constituent	**Unit**	**Untreated wastewater**	**Conventional activated sludge[b]**	**Activated sludge with BNR[c]**	**Membrane bioreactor**
Total suspended solids (TSS)	mg/L	120–400	5–25	5–20	≤1
Biochemical oxygen demand (BOD)	mg/L	110–350	10–30	5–15	<3
Chemical oxygen demand (COD)	mg/L	250–800	40–80	20–40	15–30
Total organic carbon (TOC)	mg/L	80–260	20–40	10–20	5–10
Ammonia nitrogen	mg N/L	12–45	1–10	0.7–3.0	0.7–3.0
Nitrate nitrogen	mg N/L	0–trace	10–30	2–10	2–10[d]
Nitrite nitrogen	mg N/L	0–trace	0–trace	0–trace	0–trace
Total nitrogen	mg N/L	20–70	15–35	5–10	3–10[d]
Total phosphorus	mg P/L	4–12	4–10	0.5–2.0	0.5–2.0[d]
Turbidity	NTU		2–15	2–8	≤1
Volatile organic compounds (VOCs)	mg/L	<100–>400	10–40	10–20	10–20
Metals	mg/L	1.5–2.5	1–1.5	1–1.5	trace
Surfactants	mg/L	4–10	0.5–2	0.1–1	0.1–0.5
Totals dissolved solids (TDS)	mg/L	270–860	500–700	500–700	500–700
Trace constituents[f]	mg/L	10–50	5–40	5–30	0.5–20
Total coliform	No./100 mL	10^6–10^9	10^4–10^5	10^4–10^5	<100
Protozoan cysts and oocysts	No./100 mL	10^1–10^4	10^1–10^2	0–10	0–1
Viruses	PFU/100 mL[e]	10^1–10^4	10^1–10^3	10^1–10^3	10^0–<10^3

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[a] From Chap. 3, Table 3–12 and 3–14.

[b] Conventional secondary is defined as activated sludge treatment with nitrification.

[c] BNR is defined as biological nutrient removal for the removal of nitrogen and phosphorus.

[d] With BNR process.

[e] Plaque forming units.

[f] For example, prescription and non-prescription drugs.

For constituents that are not modified significantly by biological treatment, such as inorganic TDS, it has been found that both arithmetic and log-normal distributions can be used to model process performance. Also, where the variability in performance is not great, both the arithmetic and log-normal distributions can be used to model the observed performance. The Weibull distribution (Kokoska and Zwillinger, 2000) has also proven useful in the analysis of the performance of advanced wastewater treatment processes (WCPH, 1996, 1997).

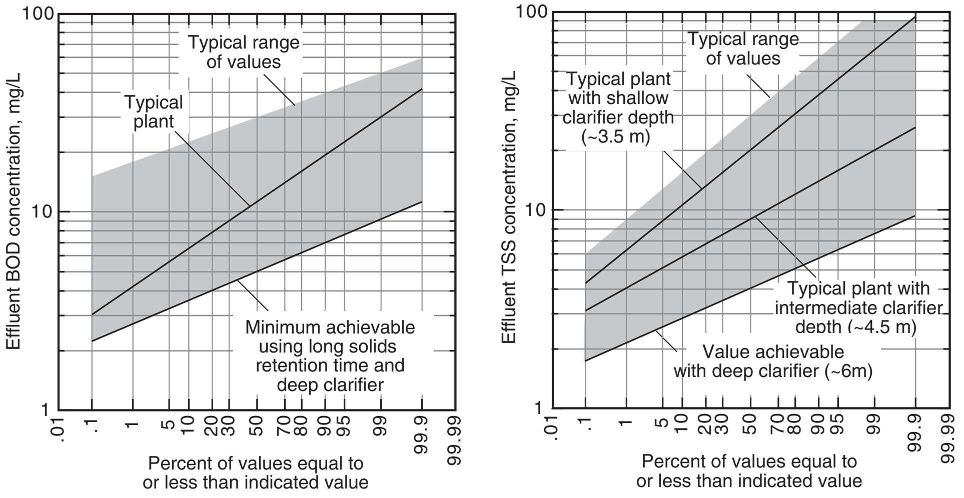


Figure 4–5

Variability of effluent constituents observed in the performance of the activated sludge process: (a) BOD and (b) TSS.

Table 4–6

Typical range of effluent quality variability observed from secondary treatment processes[a]

Biological treatment process	Unit	Range of effluent values	Geometric standard deviation, s_g[b]	
			Range	Typical
Conventional activated sludge				
BOD	mg/L	5–25	1.3–2.0	1.5
TSS	mg/L	5–25	1.2–1.8	1.4
Turbidity	NTU	5–15[c]	1.2–1.6	1.4
Activated sludge with BNR				
BOD	mg/L	5–15	1.3–2.0	1.5
TSS	mg/L	5–20	1.2–1.8	1.4
Turbidity	NTU	2–8	1.2–1.6	1.4
Membrane bioreactor				
BOD	mg/L	<3	1.3–1.6	1.4
TSS	mg/L	≤ 1	1.3–1.9	1.5
Turbidity	NTU	≤ 1	1.1–1.4	1.3

[a] All of the reported distributions are log normal, Mg = geometric mean, s_g = geometric standard deviation.

[b] $s_g = P_{84.1}/P_{50}$.

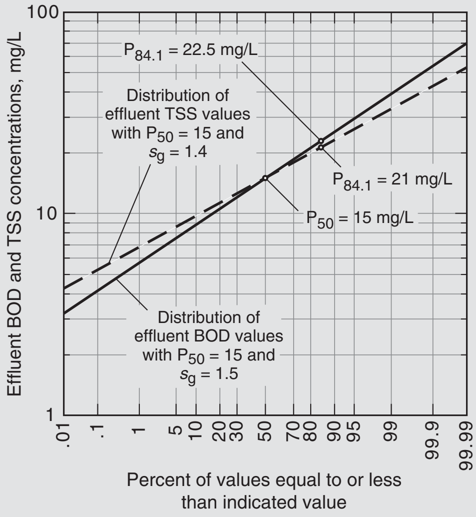
[c] Turbidity values of less than 2 NTU have been observed in plants with deep clarifiers (e.g., sidewater depths of 5.5 to 6 m). Corresponding BOD and TSS values are in the range from 3 to 6 mg/L.

EXAMPLE 4–1 **Evaluation of Activated Sludge Process Reliability** A conventional activated sludge process has been designed to have a mean effluent BOD and TSS value of 15 mg/L. Determine the maximum BOD and TSS values that are expected to occur with a frequency of (a) once per year and (b) once every three years. If the effluent limit for both BOD and TSS is 30 mg/L, estimate how often the effluent limits will be exceeded annually.

Solution

1. Select s_g values for BOD and TSS from Table 4–4 that correspond to the effluent BOD and TSS for a conventional activated sludge process. From Table 4–4, use the typical s_g values of 1.5 and 1.4 for BOD and TSS, respectively.
2. Determine the probability distribution of the effluent BOD and TSS values.
   a. Using the s_g values, compute the BOD and TSS values corresponding to the plotting position on $P_{84.1}$ (see footnote b from Table 4–6).
      i. For BOD
      $P_{84.1} = s_g \times P_{50} = 1.5 \times 15 \text{ mg/L} = 22.5 \text{ mg/L}$
      ii. For TSS
      $P_{84.1} = s_g \times P_{50} = 1.4 \times 15 \text{ mg/L} = 21 \text{ mg/L}$

   b. Estimate the distribution of effluent BOD and TSS values by plotting the $P_{84.1}$ and P_{50} values. As the effluent BOD and TSS values are expected to follow a log normal distribution, a straight line can be drawn through the $P_{84.1}$ and P_{50} values, as shown on the following plot.



3. Compute the effluent BOD and TSS values expected to occur with the frequency of interest.
   a. The probability of occurrence of a given event with a frequency of once per year is (1/365) × 100 = 0.3 %. Therefore, the percent of events occurring less than once per year is 100 − 0.3 = 99.7 %. Using the plot developed in step 2, the effluent BOD and TSS values corresponding to 99.7 % are
      i. For BOD
      $P_{99.7} = 45.8 \text{ mg/L}$
      ii. For TSS
      $P_{99.7} = 37.8 \text{ mg/L}$

b. Similarly, the probability of occurrence of a given event with a frequency of once in three years (i.e., 99.9%) is
   i. For BOD
      $P_{99.9} = 52.6$ mg/L
   ii. For TSS
      $P_{99.9} = 42.5$ mg/L

4. Estimate how often the annual effluent BOD and TSS values will exceed the effluent standard of 30 mg/L.
   a. From the plot presented in step 2, the effluent BOD will exceed 30 mg/L approximately 4.5% of the time (~16 d/y).
   b. From the plot presented in step 2, the effluent TSS will exceed 30 mg/L approximately 2.0% of the time (~7 d/y).

Comment As found in step 4, the effluent BOD and TSS values will exceed the discharge limit of 30 mg/L about 4.5 and 2.0 percent of the time, respectively. If the BOD and TSS are not to exceed the effluent limits, then either the process will have to be designed for a lower mean value or some form of effluent filtration must be added to meet the discharge limits reliably. The impact of adding some form of filtration is considered in Chap. 11.

Mechanical Process Reliability. In addition to the variability in the influent wastewater flowrate and characteristics and the inherent variability in the response of wastewater treatment processes, the variability associated with the mechanical equipment used at wastewater treatment facilities must also be considered in analyzing what design values and how much standby equipment must be available to meet stringent standards at some specified reliability value (e.g., 99 or 99.9 percent). A number of approaches are available for analyzing mechanical reliability of a treatment plant and include (WCPH, 1996):

1. Critical component analysis (CCA)
2. Failure modes and effects analysis
3. Event tree analysis
4. Fault tree analysis

All four of these approaches are cited frequently in the literature and are used by a variety of industries. The critical component analysis (CCA) approach was developed by the U.S. EPA to determine the in-service reliability, maintainability, and operational availability of selected critical wastewater treatment components (U.S. EPA, 1982). The objective of the CCA is to determine which mechanical components in the wastewater treatment plant will have the most immediate impact on effluent quality should failure occur. The statistical parameters used most commonly in applying the CCA method, as defined in Table 4–7, are *mean time before failure* (MTBF), *expected time before failure* (ETBF), *inherent availability* (AVI), and *operating availability* (AVO).

A complete process reliability analysis was performed for a treatment system designed to produce 3800 m^3/d (1.0 Mgal/d) of reclaimed water. The treatment facility included preliminary treatment (coarse screening and grit removal), primary treatment (rotary drum and disk screens), secondary treatment (water hyacinth ponds), and tertiary treatment (a package plant consisting of coagulation, softening, sedimentation, and filtration). Advanced water treatment consists of ultraviolet disinfection, reverse osmosis, air stripping, and granular activated carbon adsorption. Sodium hypochlorite is used for plant

Table 4–7

Statistical measures to assess equipment reliability[a]

Statistical measure	Description
Mean time before failure (MTBF)	A measure of the mechanical reliability of equipment, determined by the number of failures. The usual approach is to divide the operating hours by the number of failures
Expected time before failure (ETBF)	Similar to the MTBF, but the actual elapsed time is used as the total time in service
Inherent availability (AVI)	Fraction of calendar time that the component or unit was operating
Operating availability (AVO)	Fraction of time the component or unit can be expected to be operational excluding preventive maintenance

[a] Adapted from U.S. EPA (1982), WCPH (1996,1997).

effluent disinfection, with the required contact time taking place within the distribution system. The results of the process reliability analysis are presented in Table 4–8. As shown, the preliminary treatment process has the lowest MTBF. Typical problems experienced with the preliminary treatment facilities included: tripped breaker, packing leak, and gear box failure. With the exception of three treatment processes, the AVO for the remaining processes was greater than 0.99. The AVI was greater than 0.99 for all of the treatment process. Information such as presented in Table 4–8 can be used to determine maintenance schedules and the requirements for standby parts and backup components (WCPH, 1996, 1997).

Selection of Process Design Parameters to Meet Discharge Permit Limits

Because of the variations in effluent quality, treatment plants must now be designed to produce an average effluent concentration below the permit requirements. The question is what mean value should be used for process design to be assured that constituent concentrations in

Table 4–8

Summary statistics on the mechanical reliability for Aqua III[a,b]

	Statistical measure[c]			
Item	MTBF, y	90% CL MTBF, y	AVO	AVI
Preliminary (headworks)	0.35	0.57	0.9953	0.9998
Primary	0.82	0.65	0.9967	0.9981
Secondary	2.12	1 75	0.9757	0.9953
Package plant	2.24	1 78	0.9994	0.9995
UV disinfection	0.58	0.25	0.9991	0.9984
Reverse osmosis	1.22	0.99	0.9900	0.9903
Aeration tower	1.16	0.50	0.7835	0.9995
Carbon tower	1.86	1.02	0.9963	0.9999
Product water	0.56	0.45	0.9771	0.9964

[a] Adapted from WCPH (1997).

[b] Aqua III data 10/9/94 through 9/30/95.

[c] See Table 4–7 for definitions of statistical measures.

the effluent will be equal to or less than a specified limit with a specified degree of reliability? Two approaches can be used to estimate the design mean value needed to meet prescribed standards: (1) a statistical approach involving the coefficient of reliability and (2) a graphical approach. Both of these approaches are described and illustrated in the following discussion. Whether the mean value arrived at can be designed for is addressed subsequently.

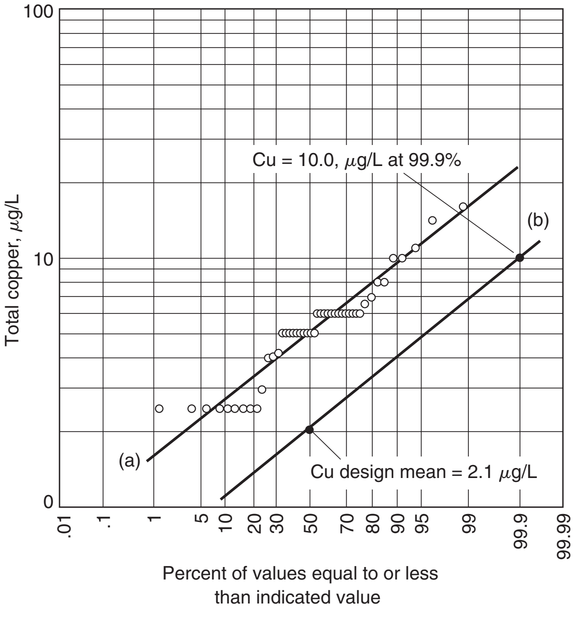
Statistical Approach to Selection of Mean Design Value. One approach that can be used to determine the mean design value involves the use of the coefficient of reliability (COR) approach developed by Niku et al. (1979, 1981). In the COR method, the mean constituent values (design values) are related to the standards that must be achieved on a probability basis. This method of analysis is documented in the 4th edition of this textbook.

Graphical Approach to Selection of Mean Design Value. Another method of determining appropriate mean design values to meet the specified effluent standards is the graphical probability method. If it is assumed that the geometric standard deviation can be used as a measure of reliability and that the value remains approximately constant at other design values, then the required effluent value can be set at the specified reliability level (e.g., 10 mg/L at 99 percent) and a line is passed through the value with the same geometric standard deviation as the measured data. The value at a probability value of 50 percent is the new mean design value.

The graphical approach is illustrated on Fig. 4–6. The plotted data correspond to the monthly total copper in the effluent from a wastewater treatment plant. If the permit limit is to be 10 μg/L at the 99.9 percent reliability value, that value is plotted and a line with the same geometric standard deviation is drawn through the point. The value at a probability of 50 percent on the line drawn through permit limit corresponds to the required mean design value, in this case 2.1 μg/L. For many constituents, it will be found that the required mean design value cannot be met with an existing process. Where the required mean value cannot be met with a single process, it may be necessary to use two or more treatment processes in series. The typical geometric standard deviation values given in Table 4–6 can also be used for the design of new treatment plants if the discharge permit limits are known.

Figure 4–6

Probability distributions:
(a) monthly efffluent total copper (Cu) samples collected over a period of two years and
(b) corresponding distribution to achieve 99.9 percent compliance with a copper limit of 10 mg/L, drawn with the same geometric standard deviation as the original distribution.



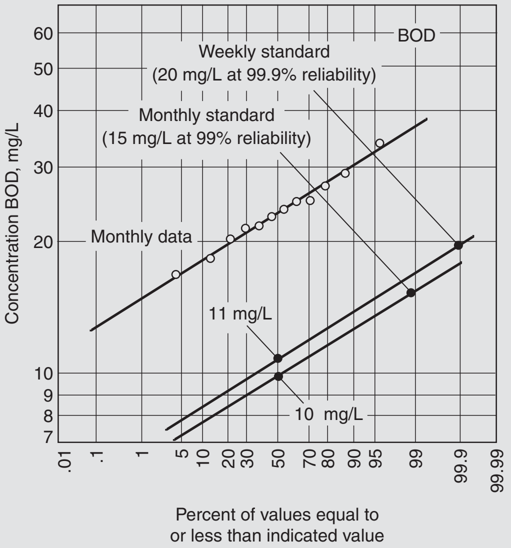
EXAMPLE 4–2 **Estimating Effluent Design BOD and TSS Concentrations Based on Reliability Considerations** An existing activated sludge plant is required to be expanded and upgraded to meet new permit requirements. The new effluent requirements are as given below. Determine the mean design effluent BOD and TSS concentrations required to meet the 99 percent reliability level for the monthly standard and the 99.9 percent reliability level for the weekly standard using the log-probability graphical method. Average monthly effluent BOD and TSS data for the existing facility for a period of one year are also given below.

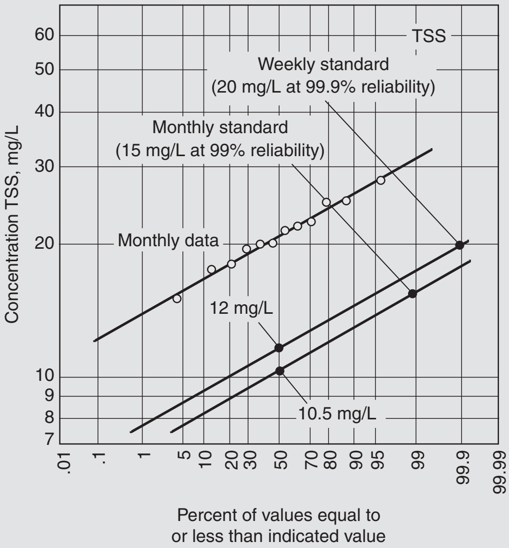
Constituent	Monthly mean	Weekly mean
BOD, mg/L	15	20
TSS, mg/L	15	20

Month	BOD, mg/L	TSS, mg/L
Jan	34.0	15.0
Feb	27.1	18.0
Mar	29.0	17.5
Apr	25.0	22.5
May	25.1	22.0
Jun	22.0	24.9
Jul	21.7	28.0
Aug	20.5	25.1
Sep	17.0	19.5
Oct	18.5	20.0
Nov	23.1	20.1
Dec	24.0	21.5

Solution

1. Plot the monthly data for BOD and TSS on log-probability paper. The required plots for BOD and TSS are shown below.





2. Estimate the design effluent concentrations for BOD and TSS for (a) 99 percent reliability for the monthly standard and (b) 99.9 percent reliability for the weekly standard.
   a. Determine the design effluent concentrations for BOD. The BOD concentrations are determined by passing lines with the same slope as the measured data through the points at 15 mg/L and 99% and 20 mg/L and 99.9% and noting the corresponding values at 50%. The values so determined are

$$\text{BOD}_{\text{design}} \text{ at } 15 \text{ mg/L and } 99\% = 10.0 \text{ mg/L}$$
$$\text{BOD}_{\text{design}} \text{ at } 20 \text{ mg/L and } 99.9\% = 11.0 \text{ mg/L}$$

b. Determine the design effluent concentrations for TSS. The TSS concentrations are determined by passing lines with the same slope as the measured data through the points at 15 mg/L and 99% and 20 mg/L and 99.9% and noting the corresponding values at 50%. The values so determined are
   $\text{TSS}_{\text{design}}$ at 15 mg/L and 99% = 10.5 mg/L
   $\text{TSS}_{\text{design}}$ at 20 mg/L and 99.9% = 12 mg/L

Comment When the concept of reliability is used, the mean effluent values selected for design will typically be significantly lower than permit requirements. Loge et al. (2001) demonstrated how the coefficient of variation was reduced (i.e. process variability) when processes are used in series with respect to the disinfection of treated effluent with ultraviolet radiation. Based on numerous past designs in England and Europe, it has been found that to achieve the prescribed limit at the 95 percent level the average design value should be about 50 percent of the prescribed limit.

Performance of Combined Processes

Applying either the statistical or the graphical procedures as illustrated in Example 4–2, it will often be found that the resulting mean design value for a given process is well below the range where any factual knowledge exists on how to design the process. For example, assume that to meet NPDES permit requirements, an activated sludge process must be designed to meet an average effluent suspended solids concentration of 4 mg/L. The fact is that it is difficult, if not impossible, to design a secondary clarifier to meet a specific design value. What is typically assumed is that with good design and effective operation of the secondary process, an average value of 4 or 5 mg/L can be achieved. Unfortunately, such assumptions are not acceptable when not-to-exceed permit limits must be met. In such a situation, it will usually be necessary to add an additional process, such as depth or surface filtration, to meet the permit requirements consistently. In the following example, the basis for determining the performance of the combined treatment processes is addressed.

EXAMPLE 4–3 **Estimating the Performance of Combined Treatment Processes Based on Reliability Considerations** Estimate the combined performance at the 98.3, 99.2, and 99.9 percent levels for an activated sludge process followed by a granular medium depth filter with respect to the removal of TSS and turbidity. Assume the following data apply to the activated sludge process. Also, assume that no chemicals are to be used with the depth filter.

1. Distribution for effluent TSS is log-normal
2. Geometric mean for effluent TSS, $M_g = 15$ mg/L
3. Geometric standard deviation for TSS, $s_g = 1.25$
4. $s_g = P_{84.1}/P_{50}$

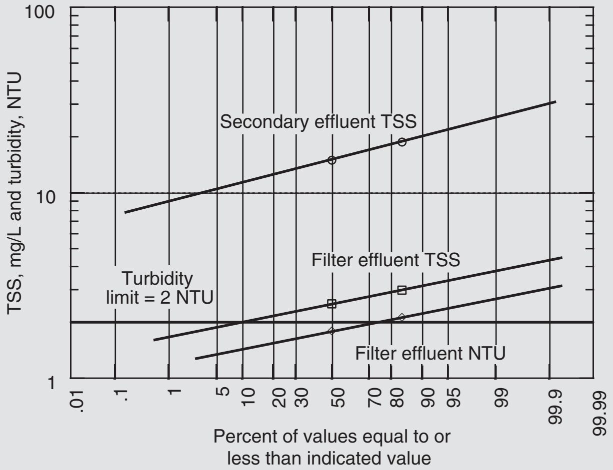
Also assume the following relationship can be used to describe the performance of the filter.

Filter effluent turbidity, NTU = 0.5 NTU + 0.2 (filter influent turbidity, NTU)

Solution

1. To determine the TSS values after filtration, use the following relationships as given in Chap. 2.
   a. Secondary effluent TSS, mg/L = (2.3 mg/L/NTU) (effluent turbidity, NTU) [Eq. (2–18)]
   b. Filter effluent TSS = (1.4 mg/L/NTU) (filter effluent turbidity, NTU) [Eq. (2–18)]

2. Using the above relationships, determine the values corresponding to 50 percent and 84.1 percent (one standard deviation) after filtration
   a. At 50 percent
      i. Secondary effluent turbidity = (15 mg/L)/(2.3 mg/L/NTU) = 6.52 NTU
      ii. Filter effluent turbidity = 0.5 NTU + 0.2(6.52 NTU) = 1.8 NTU
      iii. Filter effluent TSS = (1.4 mg/L/NTU) (1.8 NTU) = 2.52 mg/L
   b. At 84.1 percent [use Eq. (D-9) from Appendix D]
      i. $P_{84.1} = P_{50} \times s_g = 15 \text{ mg/L} \times 1.25 = 18.75 \text{ mg/L}$
      ii. Secondary effluent turbidity = (18.75 mg/L)/(2.3 mg/L/NTU) = 8.15 NTU
      iii. Filter effluent turbidity = 0.5 NTU + 0.2(8.15 NTU) = 2.13 NTU
      iv. Filter effluent TSS = (1.4 mg/L/NTU) (2.13 NTU) = 2.98 mg/L
3. Plot the secondary effluent TSS and the filter effluent turbidity and TSS and prepare a summary compliance table.
   a. Log-probability plot



   b. Summary table

Item	Unit	Value at indicated probability		
		98.3%	99.2%	99.9%
TSS	mg/L	2.6	2.9	4.2
Turbidity	NTU	2.5	2.8	3.0

Comment As shown in the above summary table, for six-TSS exceedance events per year (which corresponds to 98.3 percent probability) the TSS concentration would be equal to or greater than 2.6 mg/L, for three TSS exceedance events per year (99.2 percent probability) the TSS concentration would be equal to or greater than 2.9 mg/L and for one-TSS exceedance event in three years (99.9 percent probability) the TSS concentration would be equal to or greater than 4.2 mg/L. If a turbidity level of 2 NTU had to be met for reuse applications, the 2-NTU limit would be exceeded 25 percent of the time, without the use of additional treatment.

4–4 ELEMENTS OF PROCESS DESIGN

Treatment process design is undertaken simultaneously with the selection of the treatment process and appropriate design coefficients based on the wastewater flowrate and characteristics and treatment objectives and goals. The principal elements of process design are (1) establishing the design period for facilities, (2) development of the process flow diagram, (3) establishing process design criteria, (4) preliminary sizing of treatment units, (5) preparation of solids balances, (6) site layout considerations, (7) evaluation of plant hydraulics (hydraulic profile), and (8) energy management. Each of these elements are introduced and discussed in this section.

Design Period

The design period corresponds to the target date when the design capacity of the facilities is reached. Design periods may vary for individual components, depending upon the ease or difficulty of expansion. Typical design periods for various types of facilities are given in Table 4–9. Longer periods are preferred for structures and hydraulic conduit systems, which cannot be expanded easily. The selection of the design period depends upon growth characteristics, environmental considerations, and the availability and source of construction funds.

Treatment Process Flow Diagrams

Treatment process flow diagrams are graphical representations of particular combinations of unit operations and processes. Depending on the constituents that must be removed, an almost limitless number of different flow diagrams can be developed by combining various unit processes. Apart from the analysis of the suitability of the types of individual treatment units, the exact configuration of process units selected will also depend on factors such as (1) the designer's past experience, (2) design and regulatory agency policies on the application of specific treatment methods, (3) the availability of suppliers of equipment for specific treatment methods, (4) the maximum use that can be made of existing facilities, (5) initial construction costs, and (6) future operation and maintenance costs. A typical process flow diagram for the treatment of wastewater to meet secondary treatment standards, as defined by the U.S. EPA is shown on Fig. 4–7(a).

Process Design Criteria

After one or more preliminary process flow diagrams have been developed, the next step is to determine the process design criteria for the selected treatment processes so that the size of the physical facilities can be determined. The hydraulic detention time would be an

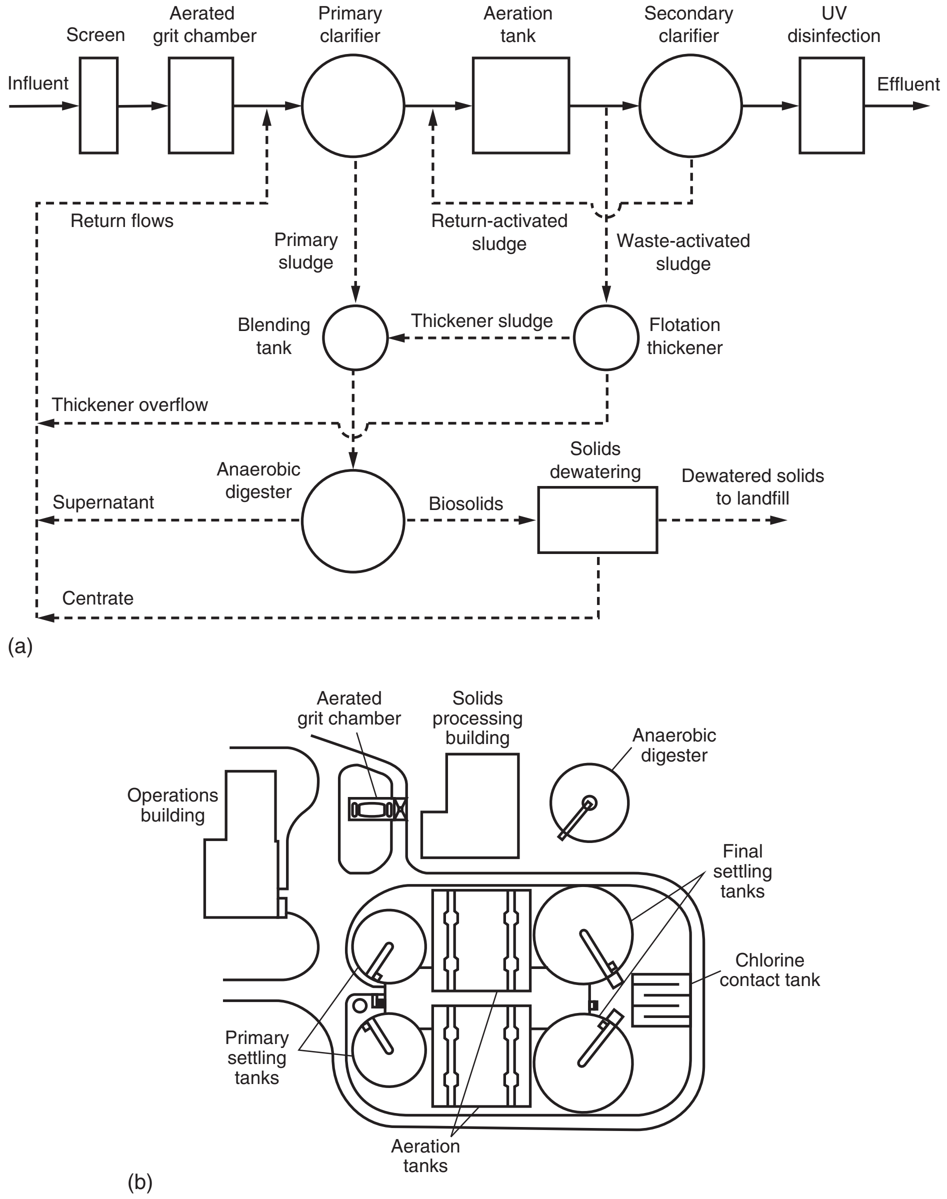
Table 4–9

Typical design periods for wastewater treatment facilities

Facility	Planning period range, y
Collection Systems	20–40
Pump stations	
Structures	20–40
Pumping equipment	10–20
Treatment plants	
Process structures	20–40
Process equipment	10–20
Hydraulic conduits	20–40

Figure 4–7

Wastewater treatment plant designed to meet US EPA secondary wastewater treatment standards (see Table 1–2) (a) schematic flow diagram and (b) schematic layout.



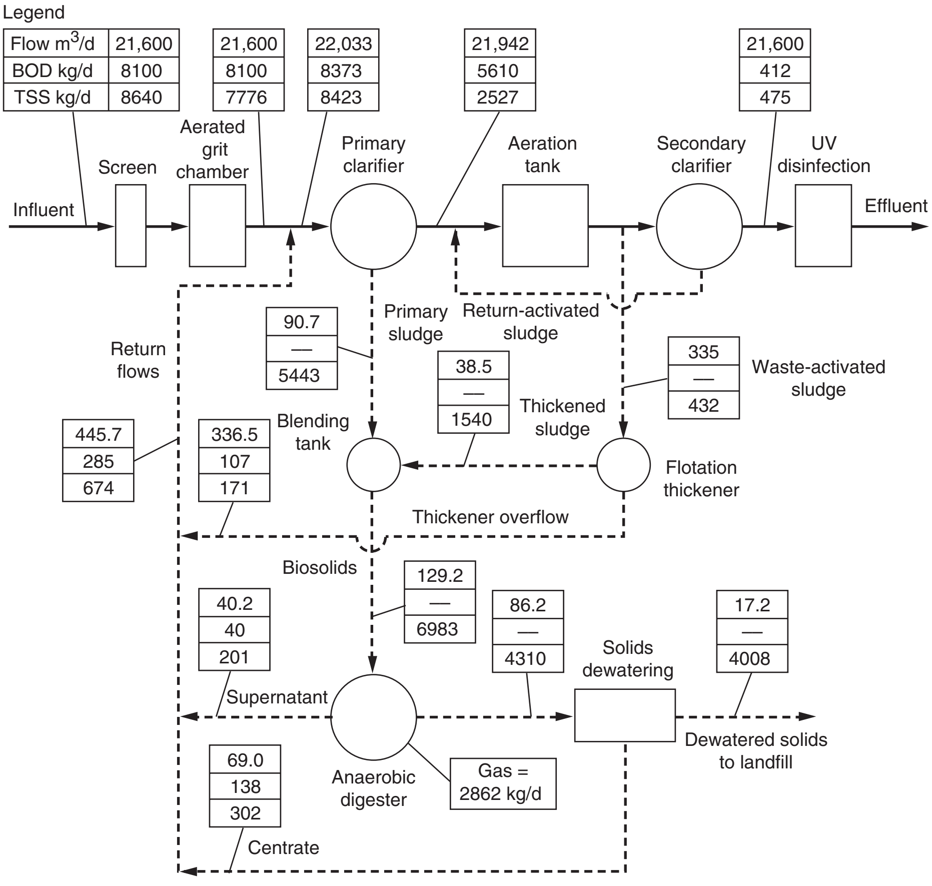
example of the process design criteria for the grit chamber. Similar procedures are followed for each unit process. All the key design criteria should be listed in a summary table. Because most treatment plants are designed to be effective for some time in the future (up to 40 y), design criteria are given generally for the time when the facilities will first be put into operation, and for the end of the design period. The latter will be influenced by projections of the population to be served and the economic studies of cost effectiveness for various design periods.

Preliminary Sizing

After the design criteria have been established, the next step is to determine the number and size of the physical facilities needed. For example, if the hydraulic detention time in the aerated grit chamber shown on Fig. 4–7(a) is to be 3.5 min at a peak flowrate, the corresponding grit chamber volume required would be calculated. In considering sizing, physical site constraints need to be considered: for example, will the site accommodate the use of round sedimentation tanks or will rectangular tanks have to be used? Operational considerations such as flow splitting and load balancing will have to be evaluated, particularly in process trains that combine different numbers of unit processes—for example, two primary clarifiers and three aeration tanks. Maintenance factors have to be considered in selecting the number of units so that provisions are included for taking a unit out of service for maintenance and repair. In small plants where a single unit is being considered, maintenance of that unit may be a particular problem unless special provisions, such as temporary storage, are included.

Solids Balance

After the design criteria are established and the preliminary sizing is completed, solids balances should be prepared for each process flow diagram. They should be prepared for the average load with appropriate peaking factors applied for maximum loads. Such information must be available to size (1) sludge thickening and storage facilities, (2) sludge digesters, (3) sludge dewatering facilities, (4) thermal reduction systems, (4) composting facilities, and (5) sludge piping and pumping equipment and other appurtenant facilities. The solids balance for the flow diagram shown on Fig. 4–7(a) is presented on Fig. 4–8. The details involved in the preparation of the solids balance are illustrated in Chap. 14.



Solids balance for second iteration

Figure 4–8

Results of solid mass balance for the treatment process flow diagram shown on Fig. 4–7. Computational details are presented in Sec. 14–7 in Chap. 14.

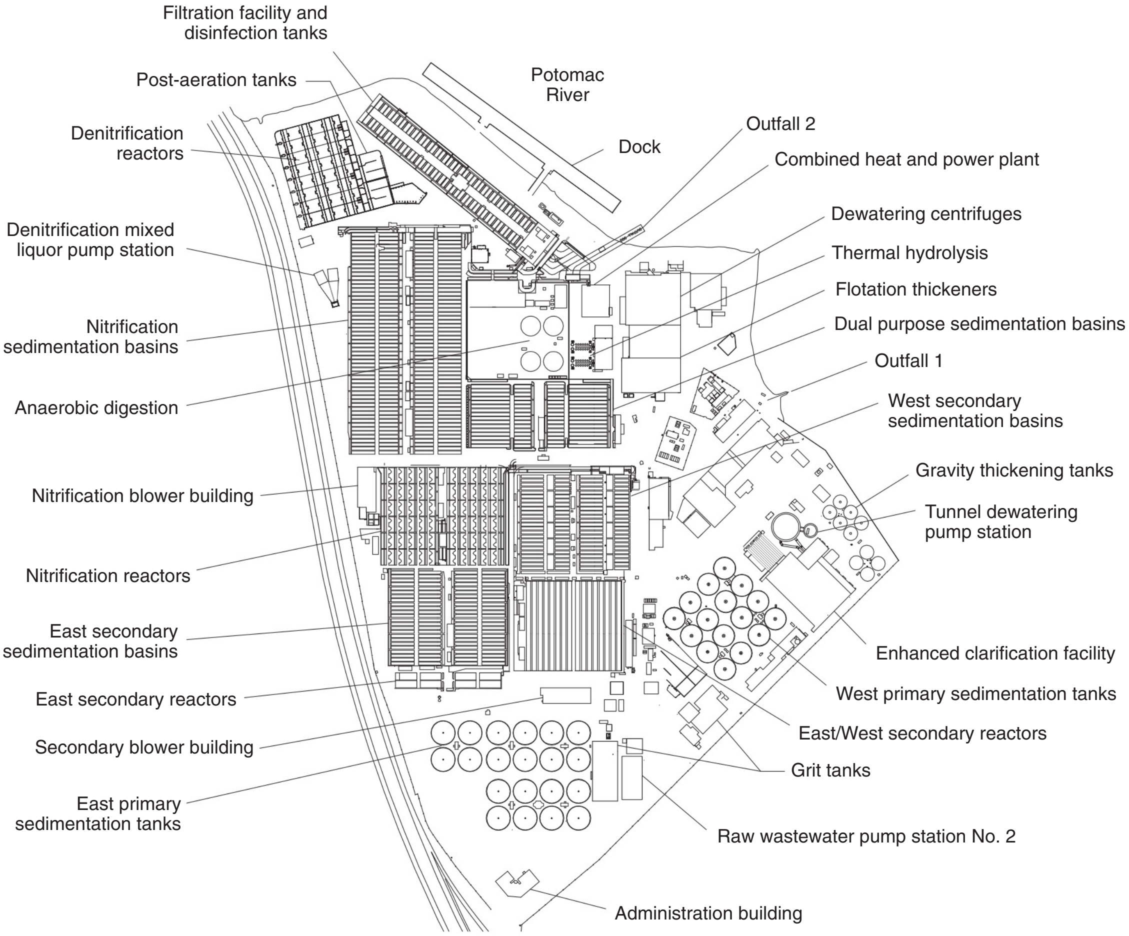


Figure 4–9

Layout of the Blue Plains Advanced Wastewater Treatment Plant with a capacity 16.2 m^3/s (370 Mgal/d) serving the Washington, DC area and environs. (Coordinates: 38.8178 N, 77.0220 W, view at altitude 4 km, also shown on the cover of this book.)

Plant Layout

Plant layout refers to the spatial arrangement of the physical facilities required to achieve a given treatment objective. The overall plant layout includes the location of the control and administrative buildings and any other necessary structures. Several different layouts, using computer generated overlays, are normally evaluated before a final selection is made. Among the factors that must be considered when laying out a treatment plant are the following: (1) geometry of the available treatment plant sites, (2) topography, (3) soil and foundation conditions, (4) location of the influent sewer, (5) location of the point of discharge, (6) plant hydraulics, preferably with straight flow paths between units to minimize headloss and provide symmetry for flow splits, (7) types of processes involved, (8) process performance and efficiency, (9) transportation access, (10) accessibility to operating personnel, (11) reliability and economy of operation, (12) aesthetics, (13) environmental control, and (14) provisions for future plant expansion including additional area. The physical layouts for a small and large wastewater treatment plant are shown on Figs. 4–7(b) and 4–9, respectively. The layout shown on Fig. 4–7(b) corresponds to the flow diagram shown on Fig. 4–7(a). The plant layout shown on Fig. 4–9 is of the large Blue Plains Wastewater Treatment Plant, Washington, DC, which includes a number of advanced treatment processes.

Plant Hydraulics

After the process flow diagram has been selected and the size of the corresponding physical facilities is determined, hydraulic computations and profiles are prepared for both average and peak flowrates. Hydraulic computations are made to size the interconnecting conduits and channels and to compute the headlosses through the plant. Typical ranges of headlosses through treatment units are given in Table 4–10. In designing the plant hydraulic system, consideration needs to be given to (1) equalizing the flow splitting between the treatment units, (2) making provisions for bypassing secondary treatment units at extreme peak flows to prevent loss of biomass, (3) provision for removing treatment facilities during periods of sustained low flow, and (4) minimizing the number of changes in direction of wastewater flow in conduits and channels.

Hydraulic profiles are prepared for three reasons: (1) to ensure that the hydraulic gradient is adequate for the wastewater to flow through the treatment facilities by gravity, (2) to establish the head requirement for the pumps where pumping will be needed, and (3) to ensure that the plant facilities will not be flooded or backed up during periods of peak flow. The hydraulic profile for the flow diagram given on Fig. 4–7 is shown on Fig. 4–10. In preparing a hydraulic profile, distorted vertical and horizontal scales are commonly used to depict the physical facilities.

Hydraulic profile computations involve the determination of the headloss as the wastewater flows through each of the physical facilities in the process flow diagram. Specific computational procedures may vary depending on local conditions. For example, if a downstream discharge condition is the control point, some designers prepare the hydraulic profile by working backward from the control point. Other designers prefer to

Table 4–10

Typical headlosses across various treatment units[a]

	Headloss range	
Treatment unit	**ft**	**m**
Bar Screen	0.5–1.0	0.2–0.3
Grit chambers		
Aerated	1.5–4.0	0.1–1.2
Velocity controlled	1.5–3.0	0.5–0.9
Primary sedimentation	1.5–3.0	0.5–0.9
Aeration tank	0.7–2.0	0.2–0.6
Trickling filter		
Low-rate	10.0–20.0	3.0–6.1
High-rate, rock media	6.0–16.0	1.8–4.9
High-rate, plastic media	16.0–40.0	4.9–12.2
Secondary sedimentation	1.5–3.0	0.5–0.9
Filtration	10.0–16.0	3.0–4.9
Carbon adsorption	10.0–20.0	3.0–6.1
Chlorine-contact tank	0.7–6.0	0.2–1.8

[a] The reported values do not reflect designs optimized for minimum energy usage.

work from the head end of the plant. Still others work from the center in each direction, adjusting the elevations at the end of the computations. The use of mathematical models and digital computers allow many possible hydraulic conditions to be analyzed.

Energy Management

Water and wastewater utilities in the United States consume about 2 to 4 percent of the total amount of electricity produced (WEF 2010a). Typically, 30 percent of the operating cost of a wastewater treatment plant is budgeted for energy use. During the next 20 to 30 y

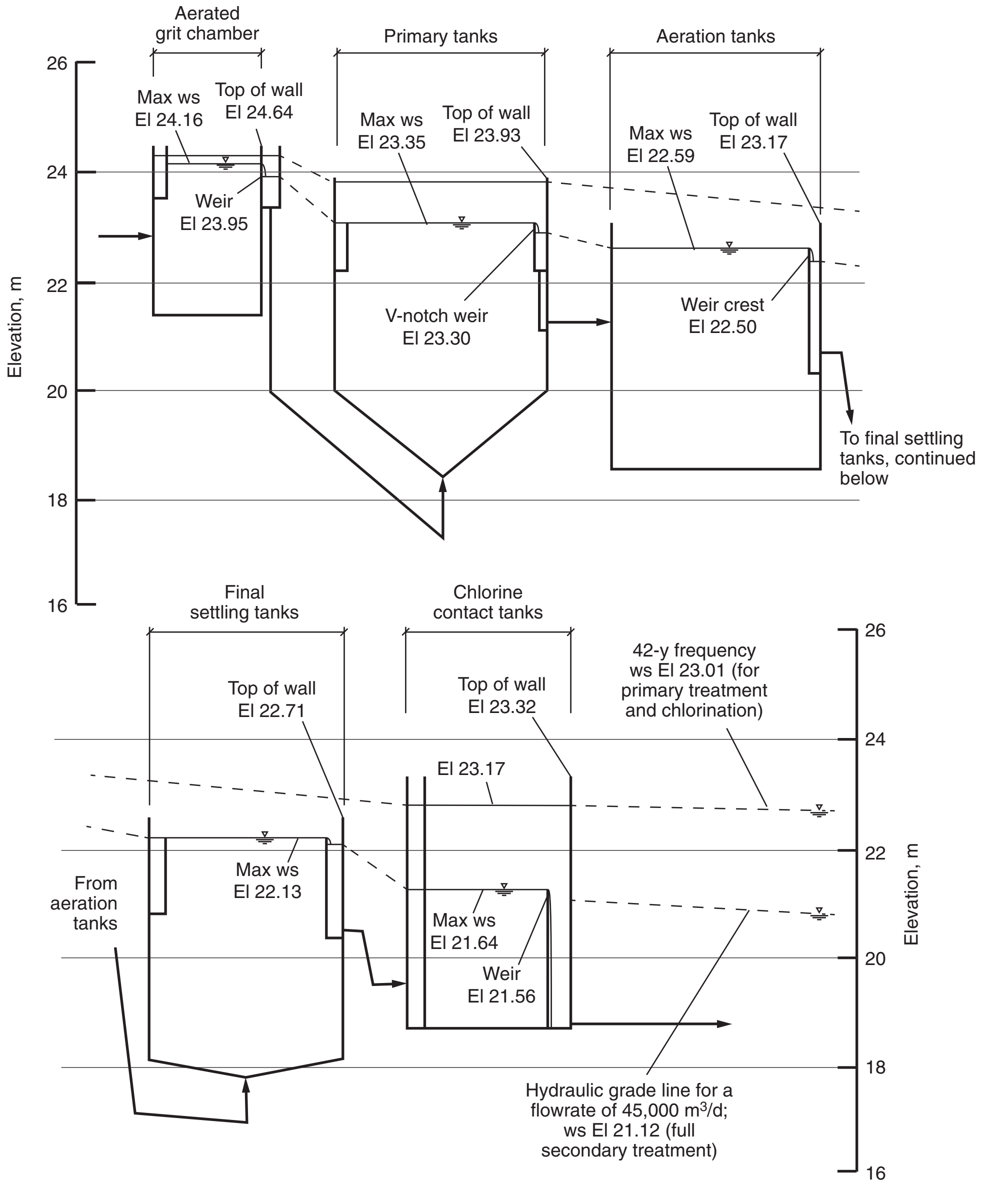


Figure 4–10

Hydraulic profile for wastewater treatment plant shown schematically on Fig. 4–7.
Note: ws = water surface, El = elevation.

the electricity requirements for wastewater treatment in the United States are expected to increase an additional 30 to 40 percent. In an era where there are concerns about the adequacy of fuel supplies, cost of energy, and the increasingly higher levels of treatment that result in increased energy consumption, the design and operation of wastewater treatment plants is focused increasingly on improving the efficiency of electric energy use and reducing the cost of treatment. Thus, the importance of energy usage and procurement in wastewater treatment cannot be overstated. Reference to energy usage and management is a constant theme throughout this book. A comprehensive analysis of the use of energy in wastewater treatment and the measures that can be employed to improve energy efficiency is presented in Chap. 17.

4–5 IMPLEMENTATION OF WASTEWATER MANAGEMENT PROGRAMS

Many of the major considerations in the overall design process and wastewater management programs are detailed in WEF (2010b). The principal elements of a wastewater management program, highlighted in this section, include (1) facilities planning, (2) design, (3) value engineering, (4) construction, and (5) startup and operation. Most major projects having a construction cost over \$10 to 20 million typically follow all steps. Smaller projects (less than \$10 million) may not include the value engineering step, although some simplified form of value engineering is highly desirable.

Facilities Planning

A facilities plan is a document established to analyze systematically the technical, economic, environmental, and financial factors necessary to select a cost-effective wastewater management plan. While the facilities plan itself may include an environmental impact assessment on major projects, the environmental assessment is usually a separate document. The scope of the facilities plan includes (1) defining the problem; (2) identifying life expectancies of major elements (usually 20–25 y for equipment and 50 y for structures); (3) defining, developing and analyzing alternative treatment and disposal systems; (4) selecting a plan; and (5) outlining an implementation plan including financial arrangements and a schedule for design and construction. The ultimate objective of a facilities plan is a well-defined, cost-effective, and environmentally sound project capable of being implemented and being acceptable to taxpayers and regulatory authorities. For more details about preparation of a facilities plan, U.S. EPA (1985) may be consulted.

Design

Following facilities planning, the approach used generally for designing a facility consists of conceptual design, preliminary design, special studies, and final design. The conceptual design is used to finalize the preliminary design criteria used in the facilities plan, to establish preliminary facilities layouts, and to define the necessary field investigations required such as surveys and geotechnical studies. The preliminary design is an expansion of the conceptual design and defines fully the facilities to be included in the project so that final design can proceed. Special studies may include field studies or testing necessary for the development of design criteria. The final design involves the production of the detailed contract plans and specifications used to bid and build the project. Mitigation measures may also be included in the design to reduce or lessen unavoidable environmental impacts. Because the design approach varies with the type and size of the project, only a general outline of the design process is provided in this text.

Value Engineering

Value engineering (VE) is an intensive review of a project to determine best value, or value improvement, which may or may not result in cost reduction. The purpose of the VE analysis is to obtain the best project at the least cost without sacrificing quality or reliability. Depending on the size and complexity of the project, the VE effort may vary from one team and one review session to multiple teams and multiple reviews. For large projects, two review sessions are usually held, each lasting about one week: one at approximately the 20 to 30 percent stage of design completion, and a second at the 65 to 75 percent stage. The VE team members are senior professionals who are not involved with the design of the project. For detailed information about the VE process, U.S. EPA (1985) may be consulted.

Construction

The quality of the design plans and specifications are often measured by (1) ease of integration of new facilities into existing sites, (2) clarity of presentation that allows contractors to submit bids with small allowances for undefined or unforeseen conditions, (3) specification of high quality materials of construction to ensure a long useful life of the facilities, (4) timely completion of the work, and (5) minimization of changes required during construction. Some of the construction considerations and management techniques for construction are discussed below.

Construction Considerations. In the preparation of the final plans and specifications, the design engineer must consider many of the details of construction. Some of the principal considerations are (1) how will the plant be built, (2) how will it interface with existing facilities, and (3) what will be the materials of construction. The buildability of a set of plans will be reflected in the bid price and the number of changes that must be made during construction. Numerous changes can result in costly change orders. Integrating a new facility with an existing one may present problems in (1) maintaining operations during construction, (2) continuing treatment at a level that will not violate discharge permit requirements, and (3) avoiding safety hazards to personnel. The construction contract must define clearly how these issues are addressed.

In selecting materials of construction, three principles are fundamental to the engineering design of process oriented facilities: (1) durability—the life of the equipment is expected to last at least 20 y, and up to 50 y for structures; (2) reliability—good quality materials and equipment to minimize maintenance and replacement; and (3) environmental suitability—realizing that wastewater and its attendant chemicals may be corrosive. For these reasons, most process structures are constructed of reinforced concrete and other materials of construction are selected based upon their corrosion-resistant properties. For information about materials of construction for wastewater treatment plants, WEF (2010a) may be consulted.

Construction and Program Management. Techniques used to ensure timely construction of the project in accordance with the plans and specifications include construction management and program management. Construction management usually provides review of the contract plans and specifications and management oversight of the construction contractor's operations. The purposes of construction management are to (1) verify the technical adequacy, operability, and constructability of the plans and specifications before construction begins; (2) establish construction schedules consistent with the program objectives and to optimize cash resources; (3) review the contractor's operation to ensure conformance with the plans and specifications; and (4) control change orders and possible construction claims. Program management differs from construction management in that it provides a single source of responsibility and authority (accountable to the owner) for the management, planning, engineering, permitting, financing, construction, and startup operations of the total wastewater management program. Program management is often used in very large projects or projects that are privatized (see Sec. 4–6).

Facilities Startup and Operation

Some of the principal concerns in wastewater engineering relate to the startup, operation, and maintenance of treatment plants. The challenges facing the design engineer and the treatment plant operator are (1) providing, operating, and maintaining a treatment plant that consistently meets its performance requirements; (2) managing operation and maintenance costs within the required performance levels required; (3) maintaining equipment to ensure proper operation and service; and (4) training operating personnel. Therefore, the design has to be done with the operations in mind, and the plant has to be operated in accordance with the design concept. One of the principal tools used for plant startup, operation, and maintenance is the operations and maintenance (O&M) manual. The purpose of an O&M manual is to provide treatment system personnel with the proper understanding, recommended operating techniques and procedures, and references necessary to efficiently operate and maintain their facilities. The design engineer usually has the lead responsibility in preparing the O&M manual.

4–6 FINANCING

As discussed earlier in this chapter, the traditional funding sources for wastewater treatment plants have changed. The U.S. government has provided grants for construction of treatment facilities for over 40 y. The 1987 Water Quality Act included a transition program that phased out the construction grants program and phased in a state revolving loan fund program. The revolving loan program pays only a portion of the costs; the wastewater agencies have to provide the balance. Therefore, cities, towns, and small communities have to investigate their funding options carefully to determine what is the most economical financing method for them. Alternative financing methods that are used commonly include (1) long-term municipal debt financing (with or without federal or state grants or loans), (2) non-debt financing, (3) leasing, and (4) private financing (privatization). Asset management, as described in Chap. 18, may be adopted to maximize the long-term benefits of the investment while maintaining the level of services provided. Because financing is becoming more integrally involved with wastewater treatment design, construction, and operation, a brief discussion of the financing methods is provided in this section.

Long-Term Municipal Debt Financing

For projects with major capital expenditures, public agencies often use long-term debt to spread the cost of the project over a number of years. Long-term financing mechanisms include general obligation bonds, limited or special obligation bonds, revenue bonds, special assessment bonds, industrial development bonds, locally issued bonds, and small denomination bonds called "mini-bonds." Of these options, general obligation and revenue bonds are used most frequently. General obligation bonds are debt instruments backed by the full faith and credit of the issuing agency. The bonds are secured by an unconditional pledge of the issuing agency to levy unlimited taxes to meet the bond obligations. Revenue bonds are used to finance projects that generate revenue and are expected to be self-sustaining. Principal and interest charges are paid from the revenues; no taxes are levied. Tax exempt bonds result in lower interest rates as the earnings are not subject to federal or local taxes. The 1984 and 1986 tax acts substantially limit the ability of agencies to issue debt that is fully tax exempt by restricting the use of bond proceeds. To increase the marketability of bonds and revenue, several features or variations may be added to the bond structure. Also, to reduce risk during periods of uncertain economic conditions, municipal bond insurance and letters of credit may be used to enhance the credit-worthiness of the bonds.

Non-Debt Financing

Non-debt financing is a method of generating revenues from system charges and is sometimes called "pay-as-you-go" financing. The funds generated annually by rates or charges that are not used for operations and maintenance or for debt payments can be used to finance new construction. Techniques used in non-debt financing may be connection charges, special assessments, system development charges, and increasing rates in advance of construction. This method of financing may be the limited to smaller projects depending upon the amount of funds that can be generated by these techniques.

Leasing

Leasing is an alternative form of facility financing that has limited application for wastewater treatment facilities. Leasing is complex, involving tax benefits to the lessor and tax implications to the lessee. The tax acts of 1984 and 1986 substantially reduced the benefits of tax-oriented leasing. Therefore, the legal and tax consequences have to be investigated carefully before undertaking a lease. In some cases, leases may be attractive for municipal agencies as a means of acquiring needed facilities and equipment where debt limitations restrict direct purchase and ownership. Many leases include an option to buy at the end of the contract period as an ultimate ownership feature.

Privatization

Privatization refers to private sector ownership and operation of facilities and services used by government entities in performing their public function (SERC, 2004). The term privatization came into vogue after the federal income tax amendments of 1981. The tax act focused attention on private sector tax benefits that could be shared with the public sector, thereby lowering the cost of facilities for the public sector and reducing user fees. In addition to cost savings, privatization may offer advantages in construction and operating efficiencies and in meeting effluent standards. Construction efficiencies may be realized by reduced construction time, greater flexibility in flow-matching the sizing to meet current needs, and the increased use of modular designs.

The overall result of privatization is, in most cases, a reduction in life cycle cost. Reductions as high as 20 to 30 percent have been achieved as compared to conventionally financed, constructed, and operated projects. Operating efficiencies may result under private operation by centralized administration, bulk ordering of chemicals and supplies, and sharing of key personnel among multiple facilities. Assurances in meeting effluent standards may be provided by the resources available from the private operator such as required management skills and trained operating personnel.

PROBLEMS AND DISCUSSION TOPICS

4–1 Prepare a brief summary of the history of wastewater treatment in your community. Identify major events that helped to bring about change, or improvements. If any of the events were related to crisis situations, try to assess whether the same result would have been achieved with proper planning.

4–2 If an EIR was prepared for the construction of your local wastewater treatment plant, obtain a copy and review it specifically with respect to the recommendations made concerning odor and energy management. Are the findings or recommendations made in light of today's concerns?

4–3 The current construction cost for a small wastewater treatment plant is estimated to be $\$5 \times 10^6$. If the construction of the plant is to be delayed 5 or 10 y (to be selected by instructor), estimate the cost when the plant will be constructed in the future. Use end-of-year ENRCCI values in making your projection.

4–4 Determine the year when your local wastewater treatment plant was constructed or expanded and its construction costs. What would the cost be to construct or expand the plant today? What has been the average rate of inflation from the time your plant was constructed to the present?

4–5 If the energy usage of an advanced secondary treatment plant, incorporating nutrient removal, is 2400 mJ/10^3 m^3 (2500 kWh/Mgal), estimate the annual cost for energy for a 4000 m^3/d wastewater treatment plant using your local electrical cost.

4–6 Search the literature and find 3 probability plots of effluent characteristics. How do the reported reliabilities compare with the values given in Table 4–6?

4–7 A superintendent of a wastewater treatment plant has requested that his staff collect and analyze 6 effluent samples of a specific constituent for submittal to the regulatory agency. If the regulatory agency, in turn, sets the not-to-be-exceeded value (i.e., 99.9 percent) in the effluent discharge permit at the maximum value measured in the 6 samples, has the superintendent potentially shot himself in the foot? Explain and illustrate your analysis.

4–8 Using the data in Table 4–5, estimate the total logs of removal for coliphage with a treatment process composed of activated sludge followed by microfiltration and reverse osmosis for influent coliphage concentrations of 20,000, 40,000, 28,000, 50,000 pfu (plaque forming units)/100 mL (value to be selected by instructor). Assume no removal of coliphage through the microfilter and that the logs of removal achieved with the particular reverse osmosis membrane used are 2, 3.2, 3.0, or 3.7 (log removal value to be selected by instructor). What is the removal achieved at 99 and 99.9 percent reliability?

4–9 The following monthly effluent constituent concentrations have been obtained from four different existing activated sludge treatment plants. Each of these plants is to be replaced with a new plant to meet new and more stringent discharge requirements. For one of these plants (to be selected by instructor), determine the mean design value assuming the following maximum monthly permit limits must be met: (a) BOD and TSS = 15 mg/L at 99 and 99.9 percent reliability, (b) BOD and TSS = 10 mg/L at 99 and 99.9 percent reliability, (c) BOD = 5 mg/L and TSS = 8 mg/L at 99 and 99.9 percent. On average, what percentage improvement would be required to meet the new discharge permit requirements. In your opinion, is the percentage improvement reasonable?

	Constituent concentration, mg/L							
	Treatment plant number							
	1		2		3		4	
Month	BOD	TSS	BOD	TSS	BOD	TSS	BOD	TSS
Jan	11.0	14.0	4.0	5.0	10.0	40.0	8.4	6.5
Feb	14.0	11.0	5.0	5.0	15.0	39.0	10.2	4.2
Mar	7.0	10.0	6.0	6.0	17.0	23.0	17.9	5.9
Apr	6.0	6.0	7.5	7.0	20.0	30.0	10.3	10.3

(*continued*)

(*Continued*)

	Constituent concentration, mg/L							
	Treatment plant number							
	1		2		3		4	
Month	BOD	TSS	BOD	TSS	BOD	TSS	BOD	TSS
May	11.0	13.0	9.0	8.0	25.0	33.0	13.2	10.9
Jun	8.0	12.0	13.0	10.0	29.0	10.0	9.3	8.10
Jul	9.0	8.0	16.0	14.0	30.0	18.0	8.6	6.9
Aug	10.0	8.0	18.0	17.0	25.0	50.0	12.0	8.2
Sep	16.0	10.0	6.5	15.0	35.5	60.0	13.7	9.1
Oct	10.2	10.0	10.0	1.5	25.0	70.0	13.8	14.0
Nov	7.5	10.0	12.0	10.0	40.5	77.0	16.3	14.0
Dec	4.8	10.0	5.5	9.0	50.0	82.0	17.0	18.2

4–10 Average monthly performance data have been collected over a period of a year for an activated sludge process treating primary settled effluent followed by depth filters operated in parallel. Using these data, estimate what overall average monthly removal would be expected with the activated sludge process followed by one of the depth filters (to be selected by instructor) assuming the influent constituent concentration is 150, 200, 275, or 300 mg/L (to be selected by instructor).

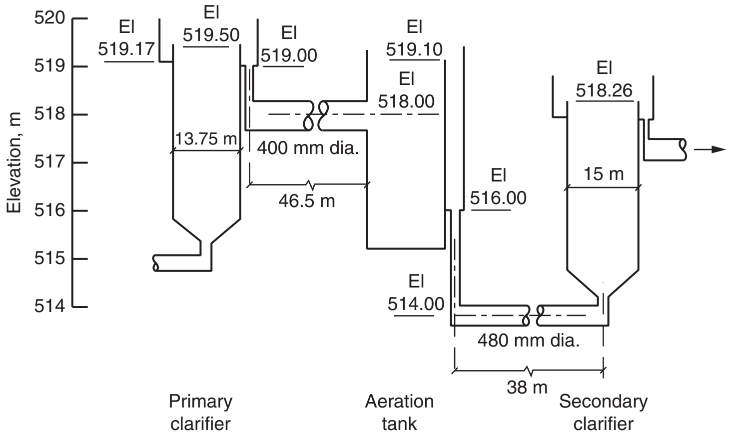
Removal, %			
Biological treatment	Depth filter number		
	1	2	3
80	65	45	41
98	65	50	44
80	65	49	49
84	65	55	45
90	65	58	47
85	65	68	45
78	65	70	43
93	65	40	46
88	65	45	45
92	65	57	43
94	65	61	48
89	65	54	42

4–11 An existing conventional activated sludge wastewater treatment plant having a design capacity of 15,000 m^3/d is considering upgrading their existing solids processing facilities to include either aerobic digestion or anaerobic digestion. List the factors the design engineer will have to consider in selecting the process and comment on the energy implications associated with the process selection.

4–12 Develop the hydraulic profile for average and peak flow conditions for the schematic portion of the wastewater treatment plant shown below. Assume that the recycle sludge is returned directly to the aeration tank, that 90° v-notch weirs are used around the periphery of the primary and secondary clarifiers and that the overflow weir in the aeration tank is a Francis type with two end contractions. Other pertinent data and information are as follows:

Q_{avg} = 4000 m^3/d plus 100 percent sludge recycle
Q_{peak} = 8000 m^3/d plus 50 percent sludge recycle

Q_{low} = 2000 m^3/d plus 100 percent sludge recycle
Spacing of v-notch weirs = 600 mm
Width of aeration-tank effluent weir = 1400 mm



4–13 Referring to the portion of the wastewater treatment plant given in Problem 4–12, develop the hydraulic profile for peak and low flow conditions for the following conditions:

Q_{avg} = 7500 m^3/d plus 100 percent sludge recycle
Q_{peak} = 15,000 m^3/d plus 50 percent sludge recycle
Q_{low} = 2500 m^3/d plus 100 percent sludge recycle
Number of primary and secondary clarifiers = 2 each
Diameter of line from aeration tank to each clarifier = 400 mm
Spacing of v-notch weirs = 600 mm
Width of aeration-tank effluent weir = 1400 mm

4–14 Based on a tour of your local wastewater treatment plant, what consideration was given to energy conservation in the design of your local wastewater treatment plant? What, if anything, is being done currently to bring about a reduction in the use of energy?

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5

Physical Unit Processes

WORKING TERMINOLOGY

Term	**Definition**
Absorption	The process by which atoms, ions, molecules, and other constituents are transferred from one phase and are distributed uniformly in another phase (see also adsorption).
Adsorption	The process by which atoms, ions, molecules, and other constituents are transferred from one phase and accumulate on the surface of another phase (see also absorption).
Air stripping	The removal of volatile and semi-volatile contaminants from a liquid by passing air and liquid counter-currently through a packed tower.
Ballasted flocculation	A process in which a flocculation aid and a ballasting agent, typically a silica microsand, are used to form dense microfloc particles, which settle rapidly.

Term	Definition
Blowers	Low-pressure air compressors that are used to deliver large quantities of air to aeration tanks to sustain biological activity.
Comminutors	In-stream grinders used for size reduction of coarse solids in wastewater.
Computational fluid dynamics (CFD)	A series of algorithms and computations used to predict or validate the behavior of liquids or gases flowing over or through constructed surfaces.
Desorption	The removal of gas from a liquid.
Diffusers	Devices used for dispersing air in a liquid.
Dissolved-air flotation (DAF)	The removal of particulate matter by attaching the particles to a blanket of rising air bubbles in a specially designed flotation tank.
Dispersed-air flotation (sometimes called induced air flotation)	The removal of oil and grease, principally in industrial wastewater, by inducing air into the wastewater by a downward pumping action; oil and solids attach to fine bubbles, which rise to the surface and are skimmed off.
Flocculation	A process by which small particles in a suspension increase in size resulting from particle collisions; flocculation is enhanced usually by gentle stirring.
Grinders	Devices used for the size reduction of coarse solids after the solids have been removed from the wastewater stream, usually by a bar screen.
Grit	Sand, gravel, cinders, other heavy inorganic materials and also organic matter such as eggshells, bone chips, seeds, and coffee grounds.
Grit chambers	Tanks designed for the removal of grit in wastewater by gravity sedimentation, helical flow, or centrifugal force.
Grit classifier (sometimes called a grit washer)	A mechanical device that uses an inclined screw or reciprocating rake to wash putrescible organics from grit.
Headloss	Energy loss for a fluid moving through a conduit, structure, or screening device caused by friction and/or turbulence.
High-rate clarification	A physical/chemical treatment process that employs special flocculation and sedimentation systems to achieve rapid settling; ballasted flocculation and tube or inclined-plate settlers are often used.
Lamella-plate clarification	A counter-current settling process in which a series of inclined plates enhance the separation of solids from clear liquid (supernatant).
Mass transfer	The transfer of material from one homogeneous phase to another; aeration, gas stripping, and adsorption are examples of mass transfer.
Mechanical aerators	Devices used to agitate water to promote mixing with atmospheric air or high purity oxygen in the head space of high purity oxygen activated sludge tank.
Mixing	The agitation of a liquid-solids suspension for the purpose of blending the mixture and keeping solids in suspension, entraining gases, or for accelerating a chemical reaction.
Preliminary treatment	Treatment steps (e.g., comminution, screening, grit removal, and preaeration) that prepare wastewater influent for further treatment.
Primary sedimentation	Customarily the first step in the wastewater treatment process in which significant amounts of suspended solids are removed by gravity sedimentation.
Screenings	Coarse material greater than 6 mm (0.25 in.) comprised of debris such as rocks, branches, pieces of lumber, leaves, paper, tree roots, plastics, bottles, cans, and rags.
Screens	Devices used to retain coarse materials (screenings) contained in wastewater; types of screens include coarse screens, fine screens, and microscreens.

Term	Definition
Sedimentation basins (also called clarifiers)	Circular or rectangular tanks that provide quiescent conditions to allow gravity settling of solid particles.
Short circuiting	Deviations from ideal flow patterns in tanks that results in shortened residence times and increased solids carryover.

Processes used for the treatment of wastewater in which change is brought about by means of or through the application of physical forces are known as *physical unit processes*. Because physical unit processes were derived originally from observations of the physical world, they were the first treatment methods to be used. Today, the physical unit processes shown on Fig. 5–1 are a major part of most wastewater-treatment systems. The physical unit processes most commonly used in preliminary and primary treatment of wastewater include some or all of the following (1) screening, (2) coarse solids reduction (comminution, maceration, and screenings grinding), (3) mixing and flocculation, (4) gravity separation, (5) grit removal, (6) primary sedimentation, (7) high-rate clarification, and (8) flotation. These topics are considered in the first 8 sections of this chapter.

New approaches for primary treatment involving physical unit processes are introduced in Sec. 5–9. The fundamentals of gas transfer discussed in Sec. 5–10 and aeration

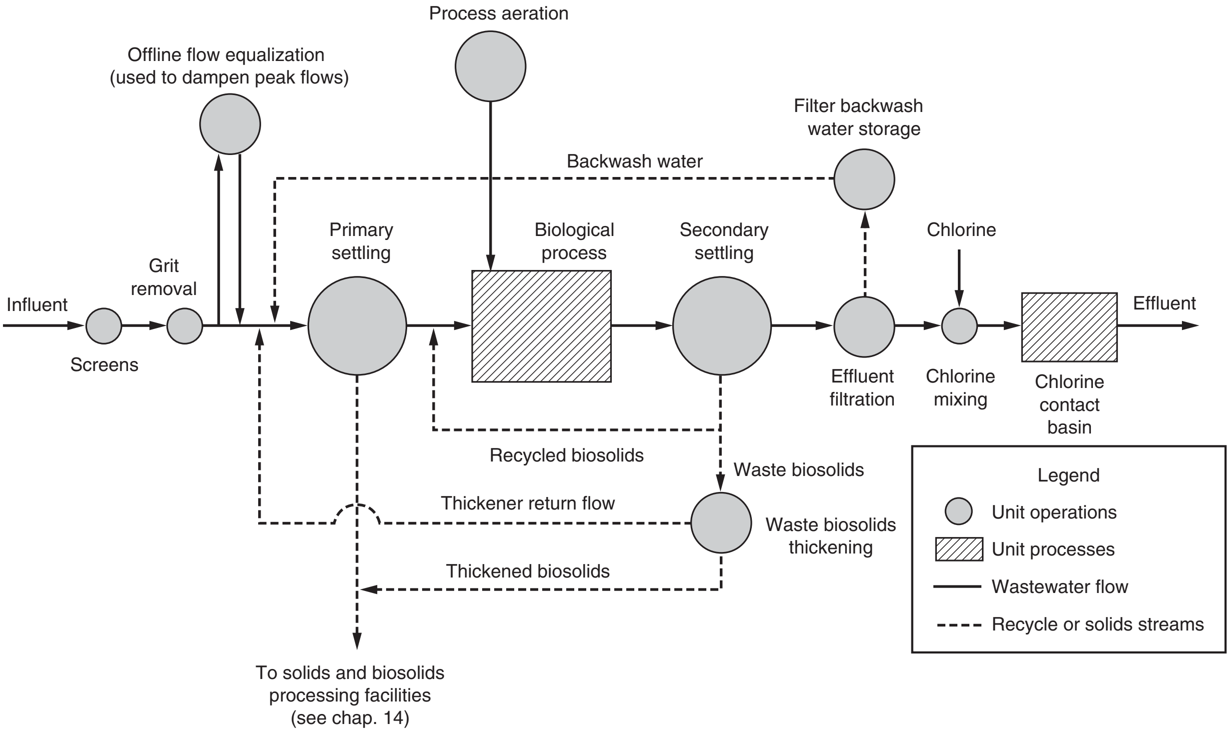


Figure 5–1

Location of physical unit operations in a conventional wastewater treatment plant flow diagram.

systems discussed in Sec. 5–11 are of critical importance in the biological treatment of wastewater discussed in Chaps. 7, 8 and 10, which follow. Physical unit processes that apply to advanced wastewater treatment systems such as depth and surface filtration, membrane separation systems, and ammonia stripping are discussed separately in Chap. 11. Unit processes associated with the processing of solids and biosolids (sludge) are covered in Chap. 13. The principal applications of these processes and treatment devices used in their implementation are summarized in Table 5–1.

Table 5–1

Typical physical unit processes used for wastewater treatment

Operation	Application	Device	See Section
Screening, coarse	Removal of coarse solids such as sticks, rags and other debris in untreated wastewater by interception (surface straining)	Bar Rack	5–1
Screening, fine	Removal of small particles	Fine Screen	5–1
Screening, micro	Removal of fine solids, floatable matter, and algae	Microscreen	5–1, 11–5
Comminution	In-stream grinding of coarse solids to reduce size	Comminutor	5–2
Grinding/ maceration	Grinding of solids removed by bar racks. Sidestream grinding of coarse solids	Screenings grinder Macerator	5–2 5–2
Mixing	Blending chemicals with wastewater and for homogenizing and maintaining solids in suspension	Rapid mixer	5–3
Flocculation	Promoting the aggregation of small particles into larger particles to enhance their removal by gravity sedimentation	Flocculator	5–3
Accelerated sedimentation	Removal of grit	Grit chamber	5–5
	Removal of grit and coarse solids	Vortex separator	5–5
Sedimentation	Removal of settleable solids	Primary clarifier	5–6
		High-rate clarifier	5–7
	Thickening of solids and biosolids	Gravity thickener	14–6
Flotation	Removal of finely divided suspended solids and particles with densities close to that of water; also thickens biosolids	Dissolved-air flotation (DAF)	5–8 14–6
	Removal of oil and grease	Induced-air flotation	5–8
Aeration	Addition of oxygen to biological process	Diffused-air aeration	5–11
		Mechanical aerator	5–11
	Post-aeration of treated effluent	Cascade aerator	5–11
VOC control	Removal of volatile and semi-volatile organic compounds from wastewaters	Gas stripper	16–4
Depth filtration	Suspended solids removal	Depth filters	11–4
Surface filtration	Suspended solids removal	Surface filters	11–5
Membrane filtration	Removal of suspended and colloidal solids and dissolved organic and inorganic matter	Reverse osmosis and other membrane systems	11–6
Air stripping	Removal of ammonia, hydrogen sulfide, and other gases from wastewater and digester supernatant	Packed tower	11–8

5–1 SCREENING

The first unit process generally encountered in wastewater treatment plants is screening. A screen is a device with openings, generally of uniform size, that is used to retain large solids found in the influent wastewater to the treatment plant or in combined wastewater collection systems subject to overflows, especially from stormwater. The principal role of screening is to remove coarse materials from the flow stream that could (1) damage or clog subsequent process equipment, (2) reduce overall treatment process reliability and effectiveness, or (3) contaminate waterways. Fine screens are sometimes used in place of or following coarse screens where greater removals of solids are required to (1) protect equipment which may be more sensitive to solids such as membrane bioreactors or (2) eliminate materials that may inhibit the beneficial reuse of biosolids.

All aspects of screenings removal, transport, and disposal must be considered in the application of screening devices, including (1) the degree of screenings removal required because of potential effects on downstream treatment processes and equipment, (2) health and safety of the operators as screenings contain pathogenic organisms and attract insects, (3) odor potential, and (4) requirements for handling, transport, prior to disposal (i.e., removal of organics by washing and reduced water content by pressing), and (5) disposal options. Thus, an integrated approach is required to achieve effective screenings management.

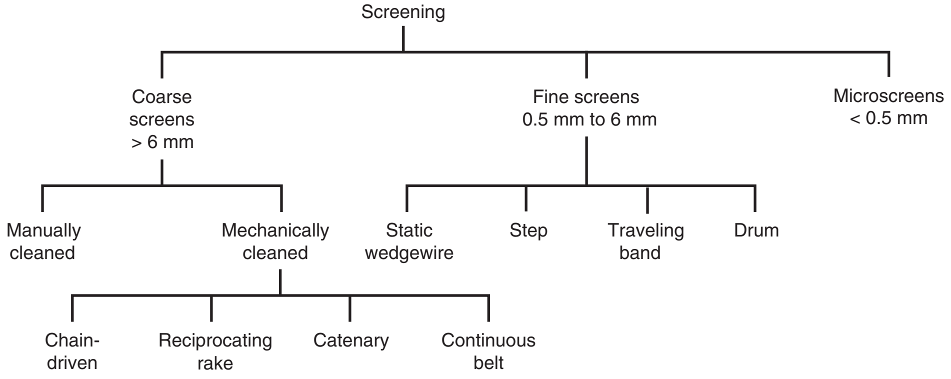
Classification of Screens

Types of screens commonly used in wastewater treatment are identified on Fig. 5–2. Two general types of screens, coarse and fine screens, are used in preliminary treatment. Fine screens may also be used as an alternative primary treatment process or for removal of additional organic solids from sludge streams prior to sludge processing. Microscreens are used principally for removing residual solids from treated effluents.

Screening elements may consist of parallel bars, rods or wires, grating, wire mesh, or perforated plate, and the openings may be of any shape, but generally are circular or rectangular slots. A screen composed of parallel bars or rods is often called a "bar rack" or a coarse screen and is used for the removal of coarse solids. Fine screens are devices consisting of perforated plates, wedge wire elements, and wire cloth that have smaller openings. The materials removed by these devices are known as screenings.

Figure 5–2

General classification for the types of screens used in wastewater treatment.



Screenings Characteristics and Quantities

Screenings are the material retained on bar racks and screens. The smaller the screen opening, the greater will be the quantity of collected screenings. While no precise definition of screenable material exists, and no recognized method of measuring quantities of screenings is available, screenings exhibit some common properties.

Screenings Retained on Coarse Screens. Coarse screenings greater than 6 mm (0.25 in.) consist of debris such as rocks, branches, pieces of lumber, leaves, paper, tree roots, plastics, bottles, cans, and rags. Organic matter associated with screenings are also removed with organic content increasing as the spacing decreases. The accumulation of oil and grease can be a serious problem, especially in cold climates. The quantity and characteristics of screenings collected for disposal vary, depending on the type of bar screen, the size of the bar screen opening, the type of sewer system, and the geographic location. Typical data on the characteristics and quantities of coarse screenings to be expected at wastewater treatment plants served by conventional gravity sewers are reported in Table 5–2.

The characteristics of the collection system will also impact the volume of screenings received. Where the collection system has many lift stations with their own screening equipment, the volume of screenings received at the treatment plant will be less. Combined sewer systems typically produce more screenings than separate sanitary sewers and may produce volumes of screenings at up to 20 times the average rate during the "first flush" of peak wet weather flow event (WEF, 2009).

Screenings Retained on Fine Screens. Fine screenings consist of materials retained on screens with openings of 0.5 mm (0.02 in.) to 6 mm (0.25 in.). The materials retained on fine screens include small rags, paper, plastic materials of various types, razor blades, grit, undecomposed food waste, feces, etc. Typical data on characteristics and quantities of screenings removed from various types of screens are reported in Table 5–3. Compared to coarse screenings, the specific weight of the fine screenings is lower and the moisture content is higher than for coarse screens. Fine screenings contain putrescible matter (including fecal material), substantial amounts of grease and scum and may be or become extremely odorous. Consequently, manual handling of collected fine screenings should be minimized and screening should be conveyed using enclosed conveyors and include grit washing and compacting equipment.

Table 5–2

Typical information on the characteristics and quantities of screenings removed from wastewater with coarse screens

Size of opening between bars, mm	Moisture content, %	Specific weight, kg/m³	Volume of screenings			
			ft³/Mgal		L/1000 m³	
			Range	Typical	Range	Typical
6.0[a]	60–90	700–1100	7–13.5	9.5	51–100	67
12.5	60–90	700–1100	5–10	7.0	37–74	50
25.0	50–80	600–1000	2–5	3.0	15–37	22
37.5	50–80	600–1000	1–2	1.5	7–15	11
50.0	50–80	600–1000	0.5–1.5	0.8	4–11	6

[a] Information on fine screen added for the purpose of comparison.

Note: mm × 0.3937 = in.
kg/m³ × 8.3492 = lb/1000 gal.

Table 5–3

Typical information on the characteristics and quantities of screenings removed from wastewater with various types of screens

Operation	Size of opening, mm	Moisture content, %	Specific weight, kg/m^3	Volume of screenings, $ft^3/Mgal$, Range	Volume of screenings, $ft^3/Mgal$, Typical	Volume of screenings, $L/1000\ m^3$, Range	Volume of screenings, $L/1000\ m^3$, Typical
Fine bar screens	12.5	80–90	900–1100	6–15	10	44–110	75
Static wedge wire	9.0	80–90	900–1100	5–12	8	37–85	60
Rotary drum[a]	6.0	80–90	900–1100	4–8	6	30–60	45

[a] Following coarse screening.

Note: mm × 0.3937 = in.
$kg/m^3 \times 8.3492$ = lb/1000 gal.

Coarse Screens (Bar Racks)

In wastewater treatment, coarse screens are used to protect pumps, valves, pipelines, and other appurtenances from damage or clogging by rags and large objects. Typically coarse screens precede fine screens to protect them from damage from large debris. Industrial waste treatment plants may not need coarse screens, depending on the character of the wastes. According to the method used to clean them, coarse screens are designated as either manually cleaned or mechanically cleaned. Screenings, handling, processing, and disposal are discussed following the discussion of the types of screens.

Manually Cleaned Coarse Screens. Manually cleaned coarse screens are used frequently ahead of pumps in small wastewater pumping stations and sometimes at the headworks of small- to medium-sized wastewater treatment plants. Often they are used for standby screening in bypass channels for service during high-flow periods, when mechanically cleaned screens are being repaired, or in the event of a power failure. Normally, mechanically cleaned screens are provided in lieu of manually cleaned screens to minimize labor required to clean the screens and to reduce flooding resulting from clogging of infrequently cleaned screens.

Where used, the length of the manually cleaned bar rack should not exceed the distance that can be conveniently raked, approximately 3 m (10 ft). The screen bars are welded to spacing bars located at the rear face, out of the way of the tines of the rake. A perforated drainage plate should be provided at the top of the rack where the rakings may be stored temporarily for drainage.

The screen channel should be designed to prevent the accumulation of grit and other heavy materials in the channel ahead of the screen and following it. The channel floor should be level or should slope downward through the screen without pockets to trap solids. Fillets may be desirable at the base of the sidewalls. The channel preferably should have a straight approach, perpendicular to the bar screen, to promote uniform distribution of screenable solids throughout the flow and on the screen. Typical design information for manually cleaned bar screens is provided in Table 5–4.

Mechanically Cleaned Bar Screens. The design of mechanically cleaned bar screens has evolved over the years to reduce the operating and maintenance problems and

Table 5–4
Typical design information for manually and mechanically cleaned bar racks

	U.S. customary units			SI units		
		Cleaning method			Cleaning method	
Parameter	**Unit**	**Manual**	**Mechanical**	**Unit**	**Manual**	**Mechanical**
Bar size						
Width	in.	0.2–0.6	0.2–0.6	mm	5–15	5–15
Depth	in.	1.0–1.5	1.0–1.5	mm	25–38	25–38
Clear spacing between bars	in.	1.0–2.0	0.6–3.0	mm	25–50	15–75
Slope from vertical	deg	30–45	0–30	deg	30–45	0–30
Approach velocity						
Maximum	ft/s	1.0–2.0	2.0–3.25	m/s	0.3–0.6	0.6–1.0
Minimum	ft/s		1.0–1.6	m/s		0.3–0.5
Allowable headloss	in.	6	6–24	mm	150	150–600

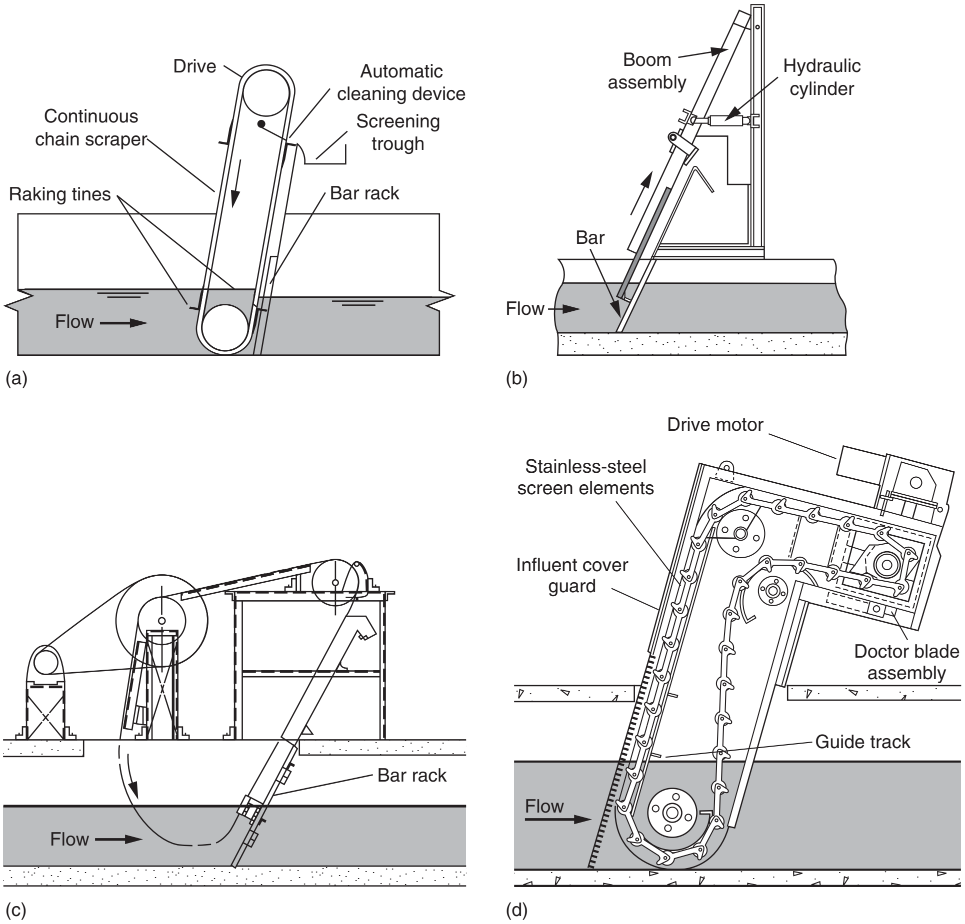
to improve the screenings removal capabilities. Many of the newer designs include extensive use of corrosion-resistant materials including stainless steel and plastics. Mechanically cleaned bar screens are divided into four principal types: (1) chain-driven, (2) reciprocating rake, (3) catenary, and (4) continuous belt. Cable-driven bar screens were used extensively in the past but have been replaced in most wastewater applications by the other types of screens. However, cable-driven raking mechanisms are still used in specific applications such as screens ahead of tunnel dewatering pump stations where deep tunnels are used for storage or conveyance of wet weather flow from combined sewer systems. Typical design information for mechanically cleaned screens is also included in Table 5–4. Examples of the different types of mechanically cleaned bar screens are shown on Fig. 5–3 and the advantages and disadvantages of each type are presented in Table 5–5.

Chain-Driven Screens. Chain-driven mechanically cleaned bar screens can be divided into categories based on whether the screen is raked to clean from the front (upstream) side or the back (downstream) side and whether the rakes return to the bottom of the bar screen from the front or back. Each type has its advantages and disadvantages, although the general mode of operation is similar. In general, front-cleaned, front-return screens [see Fig. 5–3(a)] are more efficient in terms of retaining captured solids, but they are less rugged and are susceptible to jamming by solids that collect at the base of the rake. In front-cleaned, back-return screens, the cleaning rakes return to the bottom of the bar screen on the downstream side of the screen, pass under the bottom of the screen, and clean the bar screen as the rake rises. The potential for jamming is minimized, but a hinged plate, which is also subject to jamming, is required to seal the pocket under the screen.

In back-cleaned screens, the bars protect the rake from damage by the debris. However, a back-cleaned screen is more susceptible to solids carryover to the downstream side, particularly as rake wipers wear out. The bar rack of the back-cleaned, back-return screens is less rugged than the other types because the top of the rack is unsupported so the rake

Figure 5–3

Typical mechanically-cleaned coarse screens: (a) front-cleaned, front-return chain-driven, (b) reciprocating rake, (c) catenary, and (d) continuous belt.



tines can pass through. Most of the chain-operated screens share the disadvantage of submerged sprockets that require frequent operator attention and are difficult to maintain. Additional disadvantages include the adjustment and repair of the heavy chains, and the need to dewater the channels for inspection and repair of submerged parts.

Reciprocating Rake (Climber) Screen. The reciprocating-rake-type bar screen [see Fig. 5–3(b)] imitates the movements of a person raking the screen. The rake moves to the base of the screen, engages the bars, and pulls the screenings to the top of the screen, where they are removed. Most screen designs utilize a cogwheel drive mechanism for the rake. The drive motors are either submersible electric or hydraulic type. A major advantage is that all parts requiring maintenance are above the waterline and can be easily inspected and maintained without dewatering the channel. The front-cleaned, front-return feature minimizes solids carryover. The screen uses only one rake instead of multiple rakes that are used with other types of screens. As a result, the reciprocating rake screen may have limited capacity in handling heavy screening loads, particularly in deep channels where a long "reach" is necessary. For example, heavy leaf loadings in the autumn on screens serving combined sewer systems can lead to clogging of the screens due to the long cycle time between raking. The high overhead clearance required to accommodate the rake mechanism can limit its use in retrofit applications.

Table 5–5

Advantages and disadvantages of various types of bar screens

Type of screen	Advantages	Disadvantages
Chain–driven screens		
Front clean/ back return	Multiple cleaning elements (short cleaning cycle)	Unit has submerged moving parts that require channel dewatering for maintenance
	Used for heavy duty applications	Less efficient screenings removal, i.e., carryover of residual screenings to screened wastewater channel
Front clean/ front return	Multiple cleaning elements (short cleaning cycle)	Unit has submerged moving parts that require channel dewatering for maintenance
	Very little screenings carryover	Submerged moving parts (chains, sprockets, and shafts) are subject to fouling
		Heavy objects may cause rake to jam
Back clean/ back return	Multiple cleaning elements (short cleaning cycle)	Unit has submerged moving parts that require channel dewatering for maintenance
	Submerged moving parts (chains, sprockets, and shafts) are protected by bar rack	Long rake teeth are susceptible to breakage
		Some susceptibility to screenings carryover
Reciprocating rake	No submerged moving parts; maintenance and repairs can be done above operating floor	Requires more headroom than other screens
	Can handle large objects (bricks, tires, etc.)	Long cycle time; raking capacity may be limiting
	Effective raking of screenings and efficient discharge of screenings	Grit accumulation in front of bar may impede rake movement
	Relatively low operating and maintenance costs	Relatively high cost due to stainless steel construction
	Stainless steel construction reduces corrosion	
Catenary	Sprockets are not submerged; most maintenance can be done above the operating floor	Because design relies on weight of chain for engagement of rakes with bars, chains are very heavy and difficult to handle
	Required headroom is relatively low	Because of the angle of inclination of the screen (45 to 75-deg), screen has a large footprint
	Multiple cleaning elements (short cleaning cycle)	Misalignment and warpage can occur when rakes are jammed
	Can handle large objects	May emit odors because of open design
	Very little screenings carryover	
Continuous belt	Most maintenance can be done above operating floor	Overhaul or replacement of the screening elements is a time-consuming and expensive operation
	Unit is difficult to jam	

Catenary Screen. A catenary screen is a type of front-cleaned, front-return, chain-driven screen, but it has no submerged sprockets. In the catenary screen [see Fig. 5–3(c)], the rake is held against the rack by the weight of the chain. If heavy objects become jammed in the bars, the rakes pass over them instead of jamming. The screen, however, has a relatively large "footprint" and requires greater space for installation.

Continuous Belt Screen. The continuous belt screen is a continuous, self-cleaning screening belt that can be used to remove both fine and coarse solids [see Fig. 5–3(d)]. A large number of screening elements (rakes) are attached to the drive chains; the number of screening elements depends on the depth of the screen channel. Because the screen openings can range from 0.5 to 30 mm (0.02 to 1.18 in.), it can be used as either a fine or a coarse screen. When used as a fine screen, this type of screen is better able to handle coarse solids and upstream protective devices may not be required. Hooks protruding from the belt elements are provided to capture large solids such as cans, sticks, and rags. The screen has no submerged sprocket. Newer designs of internally fed, continuous-belt, rotary screens using wedge wire instead of screen fabric are structurally more rugged.

Design of Coarse Screen Installations. Considerations in the design of screening installations include (1) location; (2) approach velocity; (3) clear openings between bars or mesh size; (4) headloss through the screens; (5) screenings handling, processing, and disposal; and (6) controls. Because the purpose of coarse screens is to remove large objects that may damage or clog downstream equipment, in nearly all cases, they should be installed ahead of the grit chambers. If grit chambers are placed before screens, rags and other stringy material could foul the grit chamber collector mechanisms, wrap around air piping, and settle with the grit. If grit is pumped, further fouling or clogging of the pumps will likely occur.

In manually cleaned installations, it is essential that the velocity of approach be limited to approximately 0.45 m/s (1.5 ft/s) at average flow to provide adequate screen area for accumulation of screenings between raking operations. Additional area to limit the velocity may be obtained by widening the channel at the screen and by placing the screen at a flatter angle to increase the submerged area. As screenings accumulate, partially plugging the screen, the upstream head will increase, submerging new areas for the flow to pass through. The structural design of the screen should be adequate to prevent collapse if it becomes plugged completely.

For most mechanically cleaned coarse screen installations, two or more units should be installed so that one unit may be taken out of service for maintenance. Slide gates or recesses in the channel walls for the insertion of stop logs should be provided ahead of, and behind, each screen so that the unit can be dewatered for screen maintenance and repair. If only one unit is installed, it is essential that a bypass channel with a manually cleaned bar screen be provided for emergency use. If the mechanically cleaned screen should become inoperative, especially during unattended hours, an overflow should be provided to direct flow to the manually cleaned bar screen. Flow through the bypass channel should be prevented by a closed slide or sluice gate. The screen channel should be designed to prevent the settling and accumulation of grit and other heavy materials. An approach velocity of at least 0.4 m/s (1.25 ft/s) is recommended to minimize solids deposition in the channel. To prevent the pass-through of debris at peak flowrates, the velocity through the bar screen should not exceed 0.9 m/s (3 ft/s).

The velocity through the bar screen can be controlled by installation of a down stream head control device such as a Parshall flume, or, for screens located upstream of a pumping station, by controlling the wetwell operating levels. If the channel velocities are controlled by wetwell levels, lower velocities can be tolerated provided flushing velocities occur during normal operating conditions.

Headloss through mechanically cleaned coarse screens is typically limited to about 150 mm (6 in.) by operational controls. The raking mechanisms are operated normally based on differential headloss through the screen or by a time clock. For time clock operation, a cycle length of approximately 15 min is recommended; however, either a high-water or high-differential contact should be provided that will place the screen in continuous operation when needed. Some mechanically cleaned screens are designed to operate at a higher speed at higher flowrates or higher differential flow conditions.

Hydraulic losses through bar screens are a function of approach velocity and the velocity through the bars. The headloss through coarse screens can be estimated using the following equation:

$$h_L = \frac{1}{C}\left(\frac{v_s^2 - v^2}{2g}\right) \tag{5–1}$$

where h_L = headloss, m (ft)
C = an empirical discharge coefficient to account for turbulence and eddy losses, typically 0.7 for a clean screen and 0.6 for a clogged screen
v_s = velocity of flow through the openings of the bar screen, m/s (ft/s)
v = approach velocity in upstream channel, m/s (ft/s)
g = acceleration due to gravity, 9.81 m/s^2 (32.2 ft/s^2)

The headloss calculated using Eq. (5–1) applies only when the bars are clean. Headloss increases with the degree of clogging. The buildup of headloss can be estimated by assuming that a part of the open space in the upper portion of the bars in the flow path is clogged. The use of Eq. (5–1) is illustrated in Example 5–1.

EXAMPLE 5–1 **Headloss Buildup in Coarse Screens** Determine the buildup of headloss through a bar screen when fifty percent of the flow area is blocked off due to the accumulation of coarse solids. Assume the following conditions apply:

Approach velocity = 0.6 m/s
Velocity through clean bar screen = 0.9 m/s
Open area for flow through clean bar screen = 0.19 m^2
Headloss coefficient for a clean bar screen = 0.7

Solution

1. Compute the clean water headloss through bar screen using Eq. (5–1).

$$h_L = \frac{1}{C}\left(\frac{v_s^2 - v^2}{2g}\right)$$

$$h_L = \frac{1}{0.7}\left[\frac{(0.9\,\text{m/s})^2 - (0.6\,\text{m/s})^2}{2(9.81\,\text{m/s}^2)}\right] = 0.033\,\text{m}$$

2. Estimate the headloss through the clogged bar screen (reducing the screen area by 50 percent results in a doubling of the velocity).

   The velocity through the clogged bar screen is

$$v_s = 0.9 \text{ m/s} \times 2 = 1.8 \text{ m/s}$$

Assuming the flow coefficient for the clogged bar screen is approximately 0.6, the estimated headloss is

$$h_L = \frac{1}{0.6}\left[\frac{(1.8\,\text{m/s})^2 - (0.6\,\text{m/s})^2}{2(9.81\,\text{m/s}^2)}\right] = 0.24\,\text{m}$$

Comment Where mechanically cleaned coarse screens are used, the cleaning mechanism typically is actuated by the buildup of headloss. Headloss is determined by measuring the water level before and after the screen. In some cases, the screen is cleaned at predetermined time intervals, as well as at a maximum head differential.

Although most screens use rectangular bars, optional shapes, i.e., "teardrop" and trapezoidal, are available. For the optional shapes, the wider width dimension is located on the upstream side of the bar rack to make it easier to dislodge materials trapped between the bars. The alternative shapes also reduce headloss through the rack.

Fine Screens

The applications for fine screens range over a broad spectrum and include preliminary treatment (following coarse bar screens), primary treatment (as a substitute for primary clarifiers), and treatment of combined sewer overflows. Fine screens can also be used to remove solids from primary effluent that could cause clogging problems in trickling filters or the membranes of membrane bioreactors. They may also be used for removal of solids from sludge flow streams prior to further treatment of the sludge (biosolids).

Screens for Preliminary and Primary Treatment. Fine screens used for preliminary treatment are (1) static wedgewire (fixed), (2) rotary drum (3) travelling belt, and (4) step type screens. Typically, the openings vary from 0.2 to 6 mm (0.01 to 0.25 in.). Examples of fine screens are illustrated on Fig. 5–4, descriptive information is provided in Table 5–6, and additional information is given below.

Fine screens may be used to replace primary treatment at small wastewater treatment plants, up to 0.13 to m^3/s (3 Mgal/d) in design capacity. Typical removal rates of BOD and TSS are reported in Table 5–7. Stainless-steel mesh or special wedge-shaped bars are used as the screening medium. Provision is made for the continuous removal of the collected solids, supplemented by water sprays to keep the screening medium clean. Headloss through the screens may range from about 0.8 to 1.4 m (2.5 to 4.5 ft).

Static Wedge Wire Screens. Static wedge wire screens [see Fig. 5–4(a)] customarily have 0.2 to 1.2 mm (0.01 to 0.06 in.) clear openings and are designed for flowrates of about 400 to 1200 $L/m^2 \cdot min$ (10 to 30 $gal/ft^2 \cdot min$) of screen area. Headloss ranges from 1.2 to 2 m (4 to 7 ft). The wedge wire medium consists of small, stainless steel, wedge-shaped bars with the flat part of the wedge facing the flow. Appreciable floor area is required for installation, and the screens must be cleaned once or twice daily with high-pressure hot water, steam, or degreaser to remove grease buildup. Static wedge wire screens are generally applicable to smaller plants or for industrial installations.

Drum Wedge Wire Screens. For the drum-type wedge wire screen [see Fig. 5–4(b)] and (c)), the screening or straining medium is mounted on a cylinder that rotates in a flow channel. The construction varies, principally with regard to the direction of flow through the screening medium. In the most common arrangements, wastewater flows either into

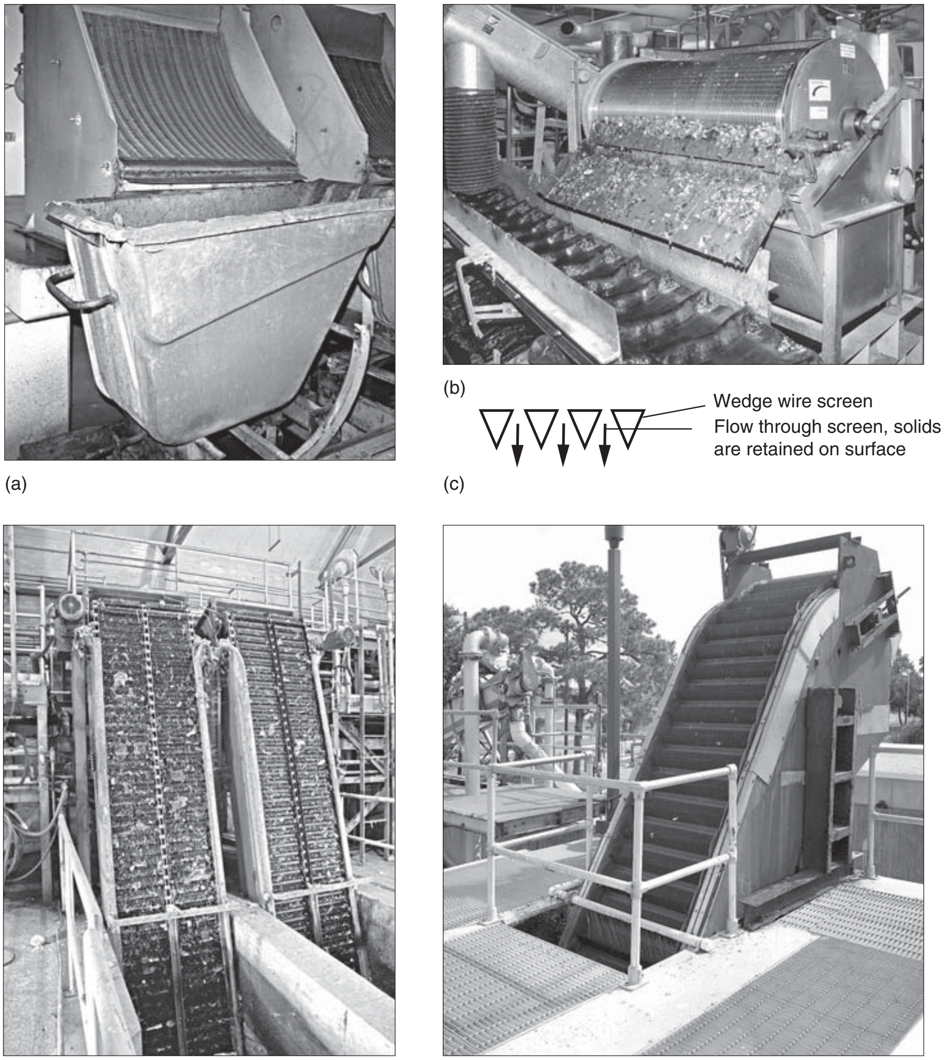


Figure 5–4

Typical fine screens: (a) static wedge wire, (b) wedge-wire drum screen, (c) section through wedge wire screen, (d) traveling band screen, and (e) step screen.

one end (single entry) of the drum and outward through the screen or from both sides (double entry) inward through the screens. In both cases, the screenings collect on the inner surface. As the screen rotates, the debris collected on the inside of the screen is lifted above the operating floor above debris hoppers located inside the screen structure and the debris is washed by a series of washwater jets. The collected debris and washwater then passes to the solids handling systems. In another arrangement, wastewater discharges into the top of the unit and passes through to the interior with solids collection on the exterior. However, this arrangement should only be used on low flowrates.

Traveling Band Screens. Traveling band screens [see Fig. 5–4(d)] are used for fine screening of influent wastewater at headworks or for secondary effluent ahead of tertiary membrane systems. In the typical application, the water being screened will pass through an opening in the front of the screen and then from the inside of the screen, through screen panels to the outside of the screen totally separating the influent and effluent. As the band of screen panels rotate, the debris collected on the inside of the screen is lifted above deck

Table 5–6

Description of screening devices used in wastewater treatment

Type of screening device	Screening surface				Application	See Fig.
	Size classification	Size range		Screen medium		
		in.	mm[a]			
Inclined (fixed)	Medium	0.01–0.1	0.25–2.5	Stainless steel wedge-wire screen	Primary treatment	5–4a
Drum (rotary)	Coarse	0.1–0.2	2.5–5	Stainless steel wedge-wire screen	Preliminary treatment	5–4b
	Medium	0.01–0.1	0.25–2.5	Stainless steel wedge-wire screen	Primary treatment	
	Fine	0.00024–0.0014	6–35 μm	Stainless steel and polyester screen cloths	Removal of residual secondary suspended solids	5–6
Horizontal reciprocating	Medium	0.06–0.17	1.6–4	Stainless steel bars	Combined sewer overflows/stormwater	5–5a
Tangential	Fine	0.0475	1200 μm	Stainless steel mesh	Combined sewer overflows	5–5b

[a] Unless otherwise noted.

level by lifting elevators, where it is washed into a debris trough within the head section of the screen. The debris trough will discharge by gravity to the screenings handling equipment. The screens may be controlled either automatically or locally by hand.

Step Screens. Step screens consist of two step-shaped sets of thin vertical plates, one fixed and one movable [see Fig. 5–4(e)]. The fixed and movable step plates alternate across the width of an open channel and together form a single screen face. The movable plates rotate in a vertical motion. Through this motion, solids captured on the screen face are automatically lifted up to the next fixed step landing, and are eventually transported to the top of the screen where they are discharged to a collection hopper. The circular pattern of

Table 5–7

Typical data on the removal of BOD and TSS with fine screens used to replace primary sedimentation[a]

Type of screen	Size of openings		Percent removal	
	in.	mm	BOD	TSS
Fixed parabolic	0.0625	1.6	5–20	5–30
Rotary drum	0.01	0.25	25–50	25–45

[a] The actual removal achieved will depend on the nature of the wastewater collection system and the wastewater travel time.

the moving plates provides a self-cleaning feature for each step. Normal ranges of openings between the screen plates are 3 to 6 mm (0.12 or 0.24 in.); however, openings as small as 1 mm (0.04 in.) are available. Solids trapped on the screen also create a "filter mat" that enhances solids removal performance. In addition to wastewater screening, step screens can be used for removal of solids from septage, primary sludge, or digested biosolids.

Fine Screens for Combined Sewer Overflows. Screens have also been developed specifically for the removal of floatable and other solids from combined sewer overflows. Basic types include horizontal reciprocating screens, tangential flow screens, and drum screens. The horizontal reciprocating screen is a rigid, weir-mounted screen configured of narrow stainless-steel bars that run the length of the device [see Fig. 5–5(a)]. The screening bars are parallel to the normal direction of flow and are designed in continuous runs with no intermediate supports to collect solids. As the water in the screen channel rises, wastewater begins to pass through the openings in the screen bars. Solids are trapped on the screen, and as the level continues to rise in the channel due to the entrapped solids on the screen, a hydraulically driven rake assembly is automatically activated to remove the accumulated solids from the screen. The rake carriage travels back and forth across the screen, combing the entrapped solids. The combing tines of the rake assembly carry the solids to one end of the screen for disposal either into the wastewater channel that carries flow to the wastewater treatment plant or to a solids-collection pit.

In the tangential flow screen, the technology relies on the natural motion of water to screen and trap solids. The separation process is effected using a fine-mesh cylindrical screen and requires no moving parts [see Fig. 5–5(b)]. As the wastewater flows into the separation chamber, a circular motion is generated that is designed to allow water to pass through the cylindrical screen while forcing solids to swirl toward the center of the chamber. The swirling water is regulated so that the tangential flow around the chamber is greater than the radial force attempting to push the solids outward. Thus, the accumulation of solids on the screen is minimized. The solids settle into a central sump where they can be removed. Floatables are trapped in the separation chamber until the flow stops, when they can be removed.

More recently, drum screens as described above have also been developed for this service. Single entry drum screens with the capability of handling very high stormwater flowrates are now available.

Design of Fine Screen Installations. Fine screens are usually preceded by mechanically cleaned coarse screens. An installation should have a minimum of two parallel screens, each with the capability of handling peak flowrates. Flushing water should be provided nearby so that the buildup of grease and other solids on the screen can be removed periodically. In colder climates, hot water or steam is more effective for grease removal.

The calculation of headloss through fine screens differs from that of coarse screens. The clear water headloss through fine screens may be obtained from manufacturers' rating tables, or calculated using Eq. (5–2):

$$h_L = \frac{1}{2g}\left(\frac{Q}{CA}\right)^2 \tag{5–2}$$

where h_L = headloss, m (ft)

C = coefficient of discharge for the screen (a typical value for a clean screen is 0.60)

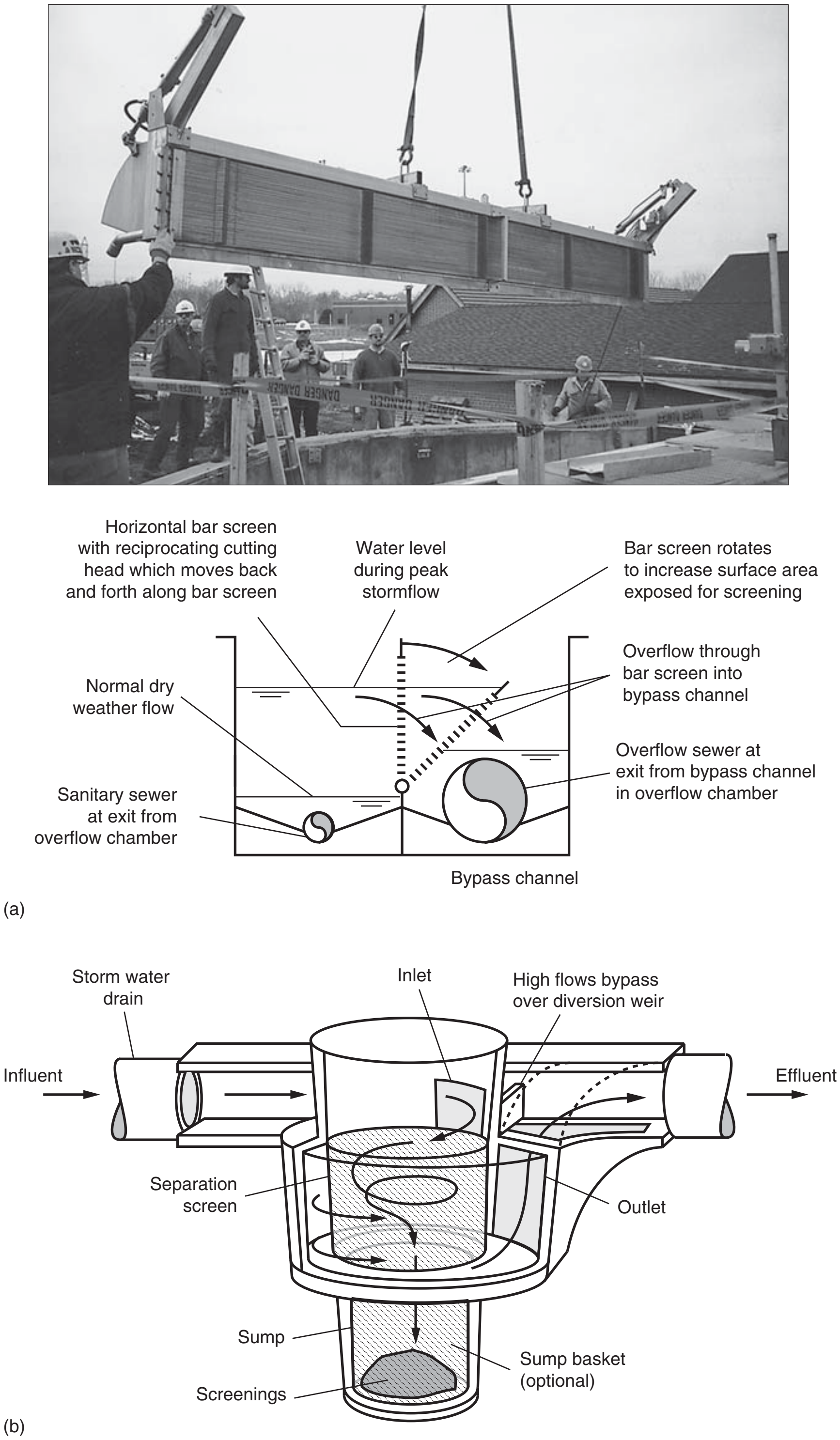
g = acceleration due to gravity, 9.81 m/s^2 (32.2 ft/s^2)

Q = discharge through screen, m^3/s (ft^3/s)

A = effective open area of submerged screen, m^2 (ft^2)

Figure 5–5

Devices used for the screening of combined sewer overflows: (a) view of horizontal screen during installation. The cutting mechanism travels back and forth horizontally along the bars. The screen can be rotated to increase the screening area exposed to the flow and (b) tangential flow device with separation screen.



Values of *C* and *A* depend on screen design factors, such as the size and milling of slots, the wire diameter and weave, and particularly the percent of open area, and must be obtained from the screen manufacturer or determined experimentally. The important determination is the headloss during operation; headloss depends on the size and amount of solids in the wastewater, the size of the apertures, and the method and frequency of cleaning.

Microscreens

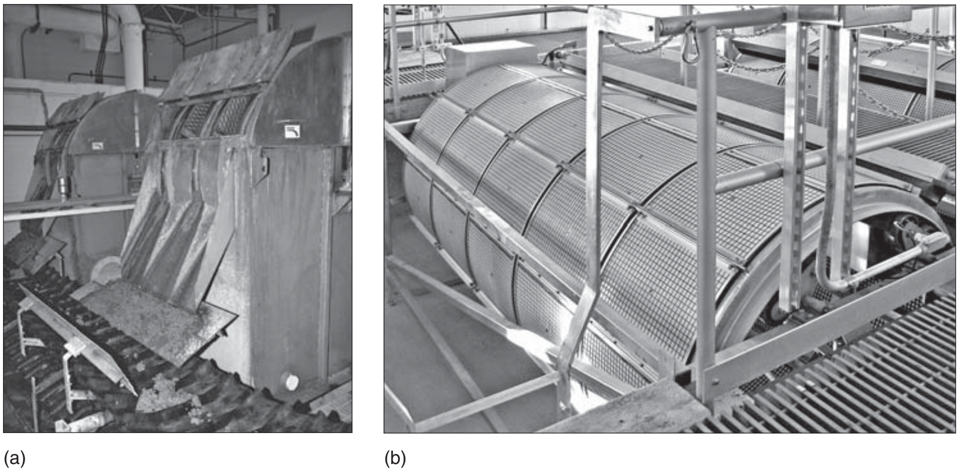
Microscreening involves the use of variable low-speed (up to 4 rev/min), continuously backwashed, rotating drum screens operating under gravity-flow conditions (see Fig. 5–6). The filtering fabrics have openings of 10 to 35 microns (micrometers) and are fitted on the drum periphery. The wastewater enters the open end of the drum and flows outward through the rotating-drum screening cloth. The collected solids are backwashed by high-pressure jets into a trough located within the drum (see also Sec. 11–5 for cloth medium surface filters used in advanced wastewater treatment). The principal applications for microscreens are to remove suspended solids from secondary effluent and from stabilization pond effluent.

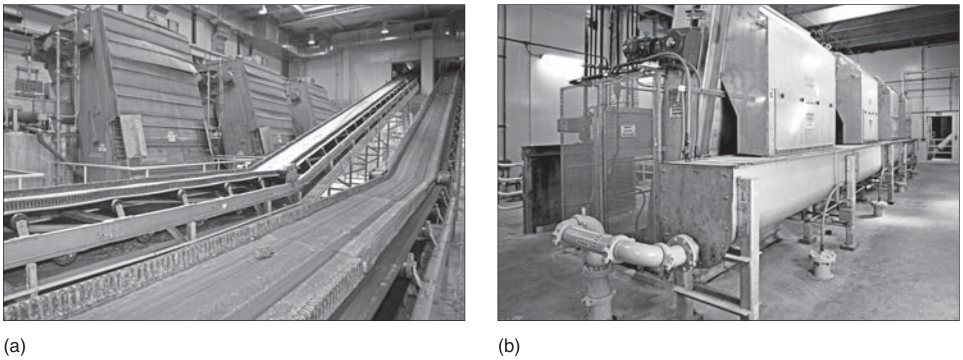
Typical suspended-solids removal achieved with microscreens ranges from 10 to 80 percent, with an average of 55 percent. Problems encountered with microscreens include incomplete solids removal and inability to handle solids fluctuations. Reducing the rotating speed of the drum and less frequent flushing of the screen have resulted in increased removal efficiencies but reduced capacity.

The functional design of a microscreen involves (1) characterizing the suspended solids with respect to the concentration and degree of flocculation, (2) selecting design parameters that will not only assure sufficient capacity to meet maximum hydraulic loadings rate with critical solids characteristics but also meet operating performance requirements over the expected range of hydraulic and solids loadings rates, and (3) providing backwash and cleaning facilities to maintain the capacity of the screen. Because of the variable performance of microscreens, pilot-plant studies are recommended, especially if the units are to be used to remove solids from stabilization-pond effluent, which may contain significant amounts of algae.

Figure 5–6

Microscreens used in wastewater treatment as a replacement for primary treatment: (a) disk type with stainless steel fabric, (b) continuous backwash drum type with stainless fabric. The size of the openings on the disk type screen is 250 μm (0.010 in.). (Courtesy of Xylem.)


(a)
(b)



(a) (b)

Figure 5–7

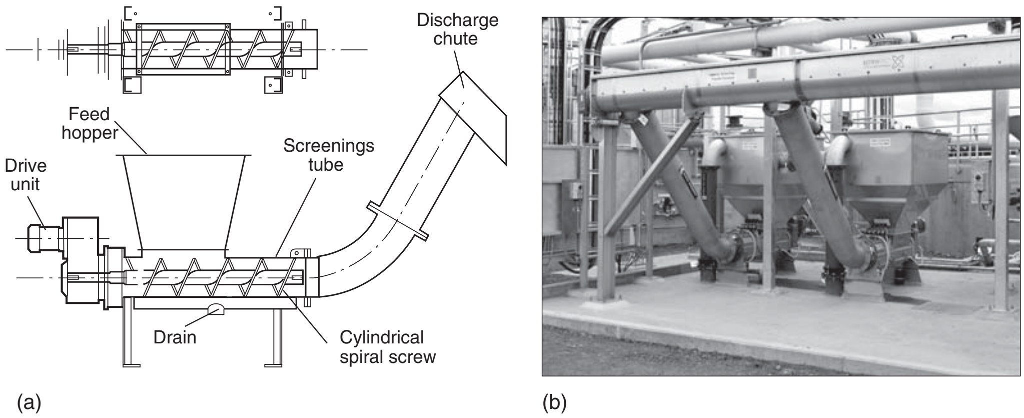
Conveyance system for coarse solids: (a) belt conveyor and (b) hydraulic transport.

Screenings Handling, Processing, and Disposal

In mechanically cleaned screen installations with multiple units, screenings are typically discharged from the screening unit onto a traveling belt conveyor or a hydraulic transport system (see Fig. 5–7) or pneumatic ejector system. Screenings can also be discharged directly into a screenings grinder, a pneumatic ejector, or a container for disposal. Belt conveyors and pneumatic ejectors are generally the primary means of mechanically transporting screenings. Belt conveyors offer the advantages of simplicity of operation, low maintenance, freedom from clogging, and low cost. Belt conveyors give off odors and may have to be provided with covers. Fluted conveyor belts can be problematic in operation due to screenings such as paper and latex products sticking to the flutes. Pneumatic ejectors are less odorous and typically require less space; however, they are subject to clogging if large objects are present in the screenings.

Screenings compactors can be used to dewater and reduce the volume of screenings (see Fig. 5–8). Such devices, including hydraulic ram and screw compactors, receive screenings directly from the bar screens and are capable of transporting the compacted screenings to a receiving hopper. Compactors can reduce the water content of the screenings by up to 50 percent and the volume by up to 75 percent. As with pneumatic ejectors, large objects can cause jamming, but automatic controls can sense jams, automatically reverse the mechanism, and actuate alarms and shut down equipment.

Macerators can be mounted above the pneumatic ejectors to grind up material and prevent clogging, however, these units must be designed to handle the characteristics of



(a) (b)

Figure 5–8

Typical device used for compacting screenings: (a) schematic and (b) view two screening compactors with overhead discharge trough.

the material to be collected to avoid bridging from excessive fats, oil, and grease (FOG) materials. The distance from screening units to the bin or truck loading area should be considered when determining height of screen and type of screening conveyance. In some cases it might be effective to confine the screening unit to the size required to remove the material from the flow path and use the conveying system to convey the material to upper levels where the storage bin or truck is located.

Means of disposal of screenings include (1) removal by hauling to disposal areas (landfill) including codisposal with municipal solid wastes, (2) disposal by burial on the plant site (small installations only), (3) incineration either alone or in combination with sludge and grit (large installations only), and (4) discharge to grinders or macerators where they are ground and returned to the wastewater. The first method of disposal is most commonly used. In some states, screenings are required to be lime stabilized for the control of pathogenic organisms before disposal in landfills. Grinding the screenings and returning them to the wastewater flow shares many of the disadvantages cited under comminution, as discussed in the following section.

5–2 COARSE SOLIDS REDUCTION

As an alternative to coarse bar screens or fine screens, comminutors and macerators can be used to intercept coarse solids and grind or shred them in the screen channel. High-speed grinders are used in conjunction with mechanically cleaned screens to grind and shred screenings that are removed from the wastewater. The solids are cut up into a smaller, more uniform size for return to the flow stream for subsequent removal by downstream treatment operations and processes. Comminutors, macerators, and grinders can theoretically eliminate the messy and offensive task of screenings handling and disposal. The use of comminutors and macerators is particularly advantageous in a pumping station to protect the pumps against clogging by rags and large objects and to eliminate the need to handle and dispose of screenings. They are particularly useful in cold climates where collected screenings are subject to freezing.

There is a wide divergence of views, however, on the suitability of using devices that grind and shred screenings at wastewater treatment plants. One school of thought maintains that once coarse solids have been removed from wastewater, they should not be returned, regardless of the form. The other school of thought maintains that once cut up, the solids are more easily handled in the downstream processes. Shredded solids often present downstream problems, particularly with rags and plastic bags, as they tend to form ropelike strands. Rag and plastic strands can have a number of adverse impacts, such as clogging pump impellers, sludge pipelines, and heat exchangers, and accumulating on air diffusers and clarifier mechanisms. Plastics and other nonbiodegradable material may also adversely affect the quality of biosolids that are to be beneficially reused.

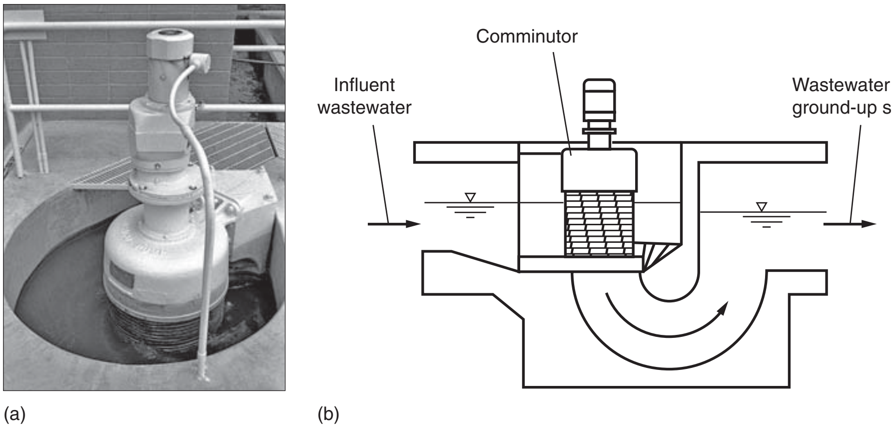
Approaches to using comminutors, macerators, and grinders are applicable in many retrofit situations. Examples of retrofit applications include plants where a spare channel has been provided for the future installation of a duplicate unit or in very deep influent pumping stations where the removal of screenings may be too difficult or costly to achieve. Alternative approaches may also be possible, such as using chopper pumps at pumping stations (see Sec. 13–4) or installing grinders ahead of sludge pumps.

Comminutors

Comminutors are used most commonly in small wastewater treatment plants, less than 0.2 m^3/s (5 Mgal/d). Comminutors are installed in a wastewater flow channel to screen and shred material to sizes from 6 to 20 mm (0.25 to 0.77 in.) without removing the shredded

Figure 5–9

Typical comminutor used for particle size reduction of solids in small plants: (a) view of installed unit and (b) schematic of typical installation.



solids from the flow stream. A typical comminutor uses a stationary horizontal screen to intercept the flow (see Fig. 5–9) and a rotating or oscillating arm that contains cutting teeth to mesh with the screen. The cutting teeth and the shear bars cut coarse material. The small sheared particles pass through the screen and into the down stream channel. Comminutors may create a string of material, namely rags, that can collect on downstream treatment equipment. Because of operating problems and high maintenance with comminutors, newer installations generally use a screen or a macerator described below.

Macerators

Macerators are slow-speed grinders that typically consist of two sets of counterrotating assemblies with blades [see Fig. 5–10(a) and (b)]. The assemblies are mounted vertically

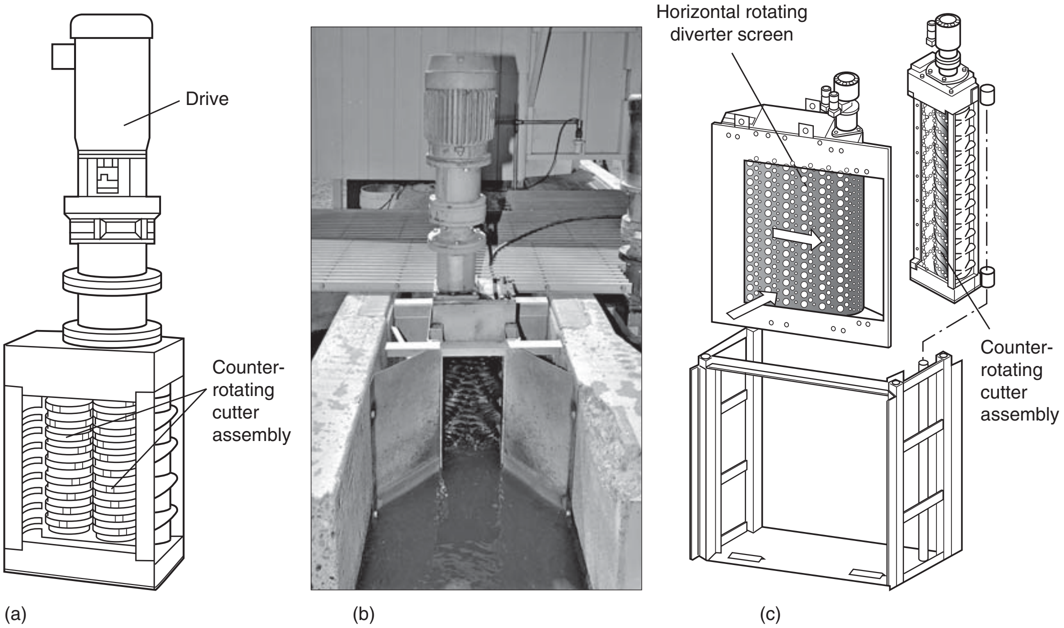


Figure 5–10

Typical macerators: (a) schematic of in-channel slow-speed grinder/macerator, (b) view of in-channel macerator, and (c) schematic of linked screen macerator.

in the flow channel. The blades or teeth on the rotating assemblies have a close tolerance that effectively chops material as it passes through the unit. The chopping action reduces the potential for producing ropes of rags or plastic that can collect on downstream equipment. Macerators can be used in pipeline installations to shred solids, particularly ahead of wastewater and sludge pumps, or in channels at smaller wastewater treatment plants. Sizes for pipeline applications typically range from 100 to 400 mm (4 to 16 in.) in diameter. Grinders for solids (sludge) applications are discussed in Chap. 13.

Another type of macerator used in channel applications is a moving, linked screen that allows wastewater to pass through the screen while diverting screenings to a grinder located at one side of the channel [see Fig. 5–10(c)]. Standard sizes of this device are available for use in large channels ranging from widths of 750 to 1800 mm (30 to 72 in.) and depths of 750 to 2500 mm (30 to 100 in.). The headloss is lower than that of the units with counterrotating blades shown on Fig. 5–10(a).

Grinders

High-speed grinders, typically referred to as hammermills, receive screened materials from bar screens. The materials are pulverized by a high-speed rotating assembly that cuts the materials passing through the unit. The cutting or knife blades force screenings through a stationary grid or louver that encloses the rotating assembly. Washwater is typically used to keep the unit clean and to help transport materials back to the wastewater stream. Discharge from the grinder can be located either upstream or downstream of the bar screen.

Design Considerations

Comminuting and macerating devices may be preceded by grit chambers to prolong the life of the equipment and to reduce the wear on the cutting surfaces. Comminutors should be constructed with a bypass arrangement so that a manual bar screen is used in case flowrates exceed the capacity of the comminutor or when there is a power or mechanical failure. Stop gates and provisions for dewatering the channel should also be included to facilitate maintenance. Headloss through a comminutor usually ranges from 0.1 to 0.3 m (4 to 12 in.) and can approach 0.9 m (3 ft) in large units at maximum flowrates. In cases where a comminutor or macerator precedes grit chambers, the cutting teeth are subject to high wear and require frequent sharpening or replacement. Units that use cutting mechanisms ahead of the screen grid should be provided with rock traps in the channel upstream of the comminutor to collect material that could jam the cutting blade.

Because these units are complete in themselves, no detailed design is necessary. Manufacturers' data and rating tables for these units should be consulted for recommended channel dimensions, capacity ranges, headloss, upstream and downstream submergence, and power requirements. Because manufacturers' capacity ratings are usually based on clean water, the ratings should be decreased by approximately 80 percent to account for partial clogging of the screen.

5–3 MIXING AND FLOCCULATION

Mixing is an important unit operation in many phases of wastewater treatment including (1) mixing of one substance completely with another, (2) blending of miscible liquids, (3) flocculation of wastewater particles, (4) continuous mixing of liquid suspensions, and (5) heat transfer. Most mixing operations in wastewater can be classified as continuous-rapid (less than 30 s) or continuous (i.e., ongoing). Continuous rapid mixing and continuous mixing are considered in this section. Each unit operation is described separately

Table 5–8

Typical mixing and flocculation devices, mixing times, and applications for mixing devices used in wastewater treatment facilities

Mixing device	Typical mixing times, s	Applications/remarks
Mixing and blending devices		
Static inline mixers	<1	Used for chemicals requiring instantaneous mixing such as alum (Al^{3+}), ferric chloride (Fe^{3+}), cationic polymer, chlorine (Cl_2)
Inline mixers	<1	Used for chemicals requiring instantaneous mixing such as alum (Al^{3+}), ferric chloride (Fe^{3+}), cationic polymer, chlorine (Cl_2)
High speed induction mixer	<1	Used for chemicals requiring instantaneous mixing such as alum (Al^{3+}), ferric chloride (Fe^{3+}), cationic polymer, chlorine (Cl_2)
Pressurized water jets	<1	Used in water treatment practice and for reclaimed water applications
Turbine and propeller mixers	2–20	Used in back mix reactors for the mixing of alum in sweep floc applications. Actual time depends on the configuration of the vessel in which mixing is taking place. Mixing of chemicals in solution feed tanks
Pumps	<1	Chemicals to be mixed are introduced in the suction intake of the pump
Other hydraulic mixing devices	1–10	Hydraulic jumps, weirs, Parshall flumes, etc.
Flocculation devices		
Static mixer	600–1800	Used for flocculation of coagulated colloidal particles
Paddle mixers	600–1800	Used for flocculation of coagulated colloidal particles
Turbine mixer	600–1800	Used for flocculation of coagulated colloidal particles
Continuous mixing		
Mechanical aerators	Continuous	Used to provide oxygen and to maintain mixed liquor suspended solids in suspension in suspended growth biological treatment processes
Hyperboloid mixers	Continuous	Used to maintain mixed liquor suspended solids in suspension in suspended-growth biological treatment processes
Pneumatic mixing	Continuous	Used to provide oxygen and to maintain mixed liquor suspended solids in suspension in suspended growth biological treatment processes

below followed by a description of the types of mixers and flocculation devices and an analysis of the energy requirements for these operations. Typical examples of the types of mixing devices used for mixing and blending, flocculation, and continuous mixing are given in Table 5–8.

Continuous Rapid Mixing in Wastewater Treatment

Continuous rapid mixing is used, most often, where one substance is to be mixed with another. The principal applications of continuous rapid mixing are in (1) the blending of chemicals with wastewater (e.g., the addition of alum or iron salts prior to flocculation and settling or for dispersing chlorine and hypochlorite into wastewater for disinfection), (2) the blending of miscible liquids, and (3) the addition of chemicals to sludge and biosolids to improve their dewatering characteristics.

Continuous Mixing in Wastewater Treatment

Continuous mixing is used where the contents of a reactor or holding tank or basin must be kept in suspension, such as in equalization basins, flocculation basins, suspended-growth biological treatment processes, aerated lagoons, and aerobic digesters. The two applications considered in the following discussion are flocculation and the maintenance of material in suspension.

Flocculation in Wastewater Treatment. The purpose of wastewater flocculation is to form aggregates or flocs from finely divided particles and from chemically destabilized particles. Flocculation is a transport step that brings about the collisions between the destabilized particles needed to form larger particles that can be removed readily by settling or filtration. Although not used routinely, flocculation of wastewater by mechanical or air agitation may be considered for (1) increasing removal of suspended solids and BOD in primary settling facilities, (2) conditioning wastewater containing certain industrial wastes, (3) improving performance of secondary settling tanks following the activated-sludge process, and (4) as a pretreatment step for the filtration of secondary effluent. When used, flocculation can be accomplished in separate tanks or basins specifically designed for the purpose, in inline facilities such as in the conduits and pipes connecting the treatment units, or in combination with flocculator-clarifiers.

Flocculation typically follows rapid mixing where chemicals have been added to destabilize the particles. The destabilization of particles resulting from the addition of chemicals is defined as "coagulation" and is considered in Chap. 6. There are two types of flocculation: (1) microflocculation and (2) macroflocculation. The distinction between these two types of flocculation is based on the particle sizes involved.

Microflocculation (also known as perikinetic flocculation) is the term used to refer to the aggregation of particles brought about by the random thermal motion of fluid molecules. The random thermal motion of fluid molecules is also known as *Brownian motion* or *movement* [see Fig. 5–11(a)]. Microflocculation is significant for particles that are in the size range from 0.001 to about 1 μm. *Macroflocculation* (also known as orthokinetic flocculation) is the term used to refer to the aggregation of particles greater than 1 or 2 μm. Macroflocculation can be brought about by (1) induced velocity gradients and (2) differential settling. Particles can be brought together (i.e., flocculated) by inducing velocity gradients in a fluid containing the particles to be flocculated. As illustrated on Fig. 5–11(b), faster-moving particles will overtake slower-moving particles in a velocity field. If the particles that collide stick together, a larger particle will be formed that will be easier to remove by gravity separation.

In macroflocculation by differential settling [see Fig. 5–11(b)], large particles overtake smaller particles during gravity settling. When the two particles collide and stick

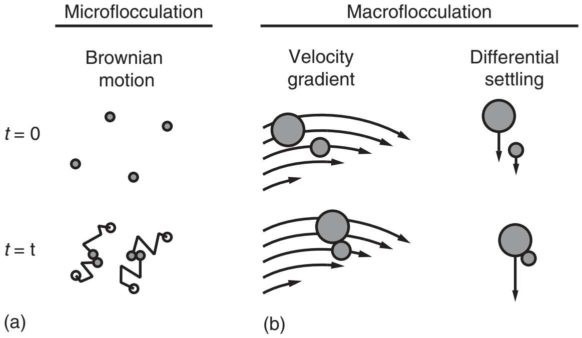


Figure 5–11

Schematic illustration of the two types of flocculation: (a) microflocculation (due to Brownian motion, also known as perikinetic flocculation) and (b) macroflocculation (also known as orthokinetic flocculation) due to (i) fluid shear and (ii) differential settling. (Adapted from Pankow, 1991; Logan, 2012.)

together, a larger particle is formed that settles at a rate that is greater than that of the larger particle before the two particles collided. Flocculent settling in the absence of induced velocity gradients is considered in the following section.

It should be noted that flocculation brought about by induced velocity gradients is ineffectual until the colloidal particles reach a size of 1 or 2 μm through contacts produced by Brownian motion. For example, macroflocculation cannot be used to aggregate viruses, which are 0.1 μm in size or smaller, until they are microflocculated or adsorbed or enmeshed in larger flocs or particles.

Maintaining Material in Suspension. Continuous mixing operations are used in biological treatment processes such as the activated-sludge process to maintain the mixed liquor suspended solids in suspension. In biological treatment systems the mixing device, in most cases, is also used to provide the oxygen transfer needed for the process. Thus, the aeration equipment must be able to provide the oxygen needed for the process and must be able to deliver the energy needed to maintain mixed conditions within the reactor. Both mechanical aerators and dissolved aeration devices are used. In both aerobic and anaerobic digestion, mixing is used to homogenize the contents of the digester to accelerate the biological conversion process, and to distribute uniformly the heat generated from biological conversion reactions or from external heating sources such as hot water circulation.

Energy Dissipation in Mixing and Flocculation

Mixing with an impeller in a reactor or mixing chamber causes two actions to occur: circulation and shearing of the fluid. The power input per unit volume of liquid can be used as a rough measure of mixing effectiveness, based on the reasoning that more input power creates greater turbulence, and greater turbulence leads to better mixing. Camp and Stein (1943) studied the establishment and effect of velocity gradients in coagulation tanks of various types and developed the following equations for use in the design and operation of systems with mechanical mixing devices (e.g., paddles).

$$G = \sqrt{\frac{P}{\mu V}} \tag{5–3}$$

where G = average velocity gradient, T^{-1}, 1/s
P = power requirement, W
μ = dynamic viscosity, $\mathrm{N{\cdot}s/m^2}$
V = flocculator volume, $\mathrm{m^3}$

In Eq. (5–3), it is important to note that the velocity gradient G is a measure of the average velocity gradient in the fluid. High G values will be observed near the blades of the mechanical mixing device, while significantly lower values will be observed at some distance from the blades of the mixing device.

As given by Eq. (5–3), the value of G depends on the power input, the viscosity of the fluid, and the volume of the basin. Multiplying both sides of Eq. (5–3) by the detention time $\tau = V/Q$ yields

$$G\tau = \frac{V}{Q}\sqrt{\frac{P}{\mu V}} = \frac{1}{Q}\sqrt{\frac{PV}{\mu}} \tag{5–4}$$

where τ = detention time, s
Q = flowrate, $\mathrm{m^3/s}$

Table 5–9

Typical detention time and velocity gradient, *G*, values for mixing and flocculation in wastewater[a]

Process	Range of values: Detention time	Range of values: *G* value, s^{-1}
Mixing		
Typical rapid mixing operations in wastewater treatment	5–30 s	500–1500
Rapid mixing for effective initial contact and dispersion of chemicals	<1 s	1500–6000
Rapid mixing of chemicals in contact filtration processes	<1 s	2500–7500
Flocculation		
Typical flocculation processes used in wastewater treatment	30–60 min	50–100
Flocculation in direct filtration processes	2–10 min	25–150
Flocculation in contact filtration processes	2–5 min	25–200

[a] The limitations associated with the use of the velocity gradient concept, as discussed in the text, must be considered in applying the *G* values given in this table.

Typical values that have been used for G for various mixing operations are reported in Table 5–9. The power required for various types of mixers is considered in the following discussion. The use of Eq. (5–3) is illustrated in Example 5–2.

EXAMPLE 5–2 **Power Requirement to Develop Velocity Gradients** Determine the theoretical power requirement to achieve a G value of 100/s in a tank with a volume of 2800 m^3 ($\sim 10^5$ ft^3). Assume that the water temperature is 15°C (59°F). What is the corresponding value when the water temperature is 5°C (41°F)?

Solution

1. Determine the theoretical power requirement at 15°C using Eq. (5–3) rearranged as follows.
   $P = G^2 \mu V$
   μ at 15°C $= 1.139 \times 10^{-3}$ N·s/m^2 (see Appendix C)
   $P = (100/s)^2(1.139 \times 10^{-3}\ \text{N·s/m}^2)(2800\ \text{m}^3)$
   $= 31{,}892$ W
   $= 31.9$ kW (23,524 ft·lb_f/s)

2. Determine the theoretical power requirement at 5°C
   μ at 5°C $= 1.518 \times 10^{-3}$ N·s/m^2 (see Appendix C)
   $P = (100/s)^2\ (1.518 \times 10^{-3}\ \text{N·s/m}^2)(2800\ \text{m}^3)$
   $= 42{,}504$ W
   $= 42.5$ kW (31,351 ft·lb_f/s)

While the use of the velocity gradient G has been popular in the water and wastewater field, it should be noted that use of the velocity gradient concept does not apply to microflocculation and, as will be discussed later, cannot be used as design parameter for some types of mixing devices. Insight into the reason that the velocity gradient G is not an effective measure for microflocculation can be gained by considering the following relationship developed by Kolmogoroff (1941) to describe the size of eddies formed as result of power input to a fluid (Davies, 1972).

$$l_K = \left(\frac{\nu^3}{P_M}\right)^{1/4} \tag{5–5}$$

where l_K = Kolmogoroff microscale length, m
ν = kinematic viscosity, m^2/s
P_M = Power per unit mass, W/kg, $[(kg{\cdot}m^2/s^3)/kg]$
$= G^2\nu$

Substituting $G^2\nu$ in Eq. (5–5) yields:

$$l_K = \left(\frac{\nu^2}{G^2}\right)^{1/4} \tag{5–6}$$

Equation (5–6) can now be used to estimate the smallest eddy that can be produced for a given average G value. For example, if $G = 1000/s$, and $\nu = 1.003 \times 10^{-6}\ m^2/s$ at 20°C, the corresponding value of the microscale length is 31.7 μm. Thus, particles smaller than 31.7 μm will not be affected. In fact, if the G value were increased to 10,000/s, the corresponding microscale length is 10.0 μm. Based on this analysis, it is clear that if particles smaller than 1 to 10 μm are to be removed they must first be destabilized and allowed to undergo microflocculation caused by Brownian motion. It can be argued that effective mixing is critical in keeping the particles in suspension so that collisions can occur by Brownian motion (Han and Lawler, 1992). Further, because the G value is an average value, the effectiveness of the mixing, which depends on the pumping characteristics of the mixer and the geometry of the mixing basin, must be evaluated carefully.

Time Scale in Mixing

The time scale for mixing is a consideration in the design of mixing facilities and operations. For example, if the reaction rate between the substance being mixed into a liquid and the liquid is rapid, the time of mixing is extremely important. For slowly reacting substances, the time of mixing is not as critical. The rationale for mixing times for various chemicals is considered in Chap. 6. Typical mixing times may be found in Table 6–24 in Chap. 6. As reported in Table 6–24, recommended mixing times for coagulants such as alum or iron salts and for dispersing chlorine and hypochlorite into solution are less than 1 s. Typical mixing times achievable for various mixing devices are reported in Table 5–8. It should be noted that achieving extremely short mixing times becomes increasingly difficult as the flowrate increases. In some applications, it may be preferable to use multiple mixing devices to achieve optimal mixing times.

Types of Mixers Used for Rapid Mixing in Wastewater Treatment

Many types of mixing devices are available, as summarized previously in Table 5–8, depending on the application and the time scale required for mixing. The principal devices used for rapid mixing in wastewater treatment applications include static inline mixers, inline mixers, high-speed induction mixers, pressurized water jets, and propeller and turbine mixers. Mixing can also be accomplished in pumps and with the aid of hydraulic devices such as hydraulic jumps, Parshall flumes, or weirs. Although hydraulic mixing can sometimes be highly efficient, the principal problem is that the energy input varies with the flowrate, and incomplete and ineffective mixing can occur at low flowrates.

Static Mixers. Static inline mixers contain internal vanes or orifice plates that bring about sudden changes in the velocity patterns as well as momentum reversals. Static mixers are principally identified by their lack of moving parts. Typical examples include inline static mixers that contain elements that bring about sudden changes in the velocity patterns as well as momentum reversals [see Fig. 5–12(a)] and mixers that contain orifice plates and nozzles [see Fig. 5–12(b)]. Static inline mixers are used most commonly for mixing of chemicals with wastewater. Inline mixers are available in sizes varying from

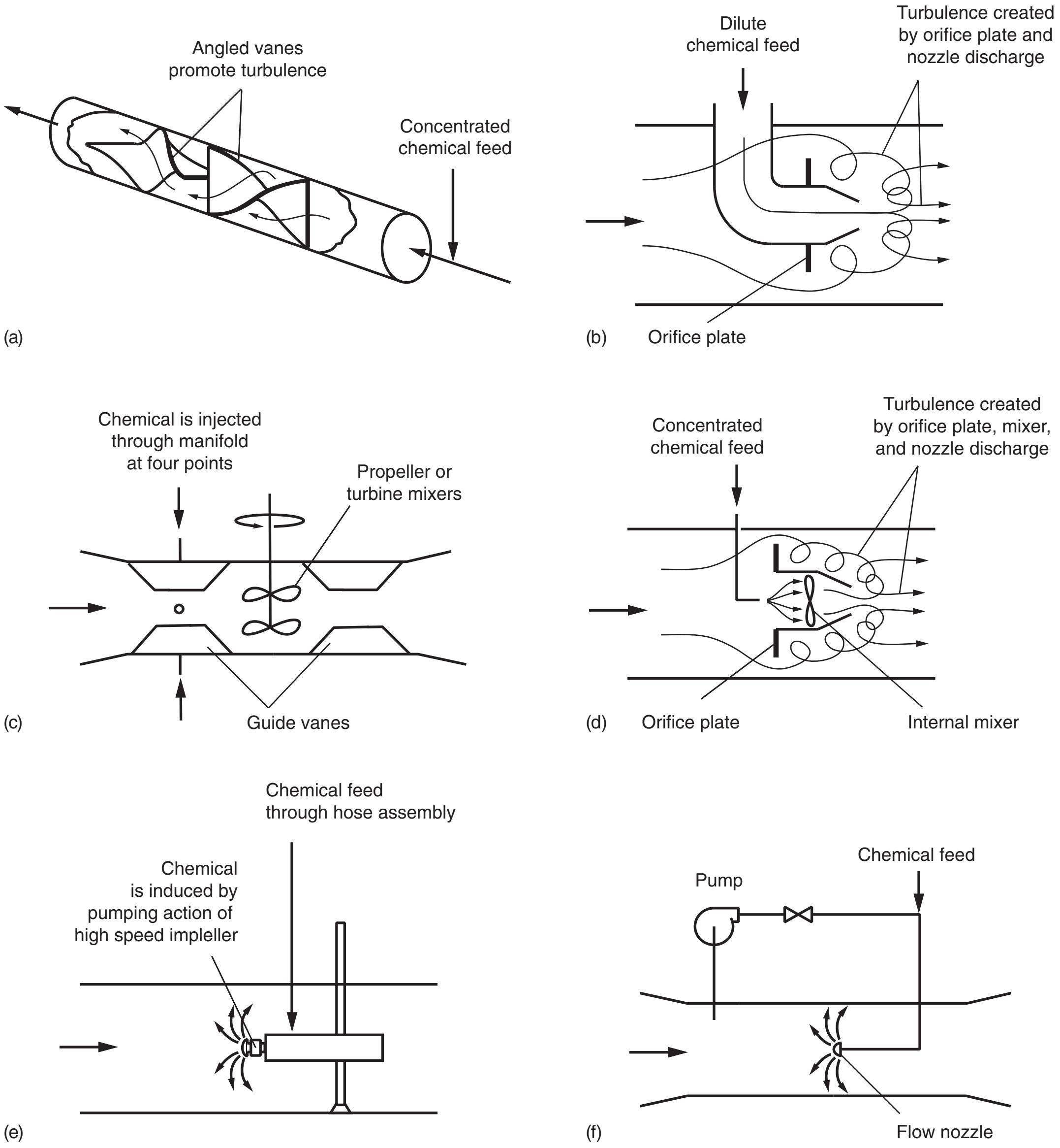


Figure 5–12

Typical mixers used in wastewater treatment for rapid mixing: (a) inline static mixer with internal vanes, (b) inline static mixer with orifice used to create internal turbulence for mixing dilute chemicals, (c) inline propeller or turbine mixer, (d) inline mixer with orifice and internal propeller mixer , (e) high speed induction mixer, and (f) pressurized water jet mixer.

about 12 mm to 3 m × 3 m open channels. Low-pressure-drop round, square, and rectangular inline static mixers have been developed for chlorine mixing in open channels and tunnels for flowrates varying from 0.22 to over 8.76 m^3/s (5 to over 200 Mgal/d) (Carlson, 2000).

For static inline mixers with vanes, the longer the mixing elements, the better the mixing; however, the pressure loss increases. It should also be noted that the shear rate and the scale (i.e., size) of the turbulent eddies formed in static mixers with vanes are more limited in range as compared to the wide range of values obtained with mechanical mixers. Mixing also occurs in a plug-flow regime in static inline mixers. Mixing times in static mixers are quite short, typically less than 1 s. The actual mixing time will vary with the length of the mixer, which depends on the number of mixing elements used, and the internal volume occupied by the mixing element. Thus, because the nature of the mixing that occurs in static mixers is quite different from that of mechanical mixers, use of the velocity gradient concept [see Eq. (5–3)] is inappropriate for static mixers.

For static mixers, the degree of mixing is related to the headloss (i.e., pressure drop) through the mixer. The headloss through the mixer can be estimated using the following expression:

$$h \approx k\left(\frac{v^2}{2g}\right) \approx K_{SM}v^2 \tag{5–7}$$

where h = headloss dissipated as liquid passes through mixing device, m
k = empirical coefficient characteristic of the mixing
v = approach velocity, m/s
K_{SM} = overall coefficient for mixing device, s^2/m
g = acceleration due to gravity, 9.81 m/s^2

Typical values for K_{SM} vary from about 1.0 to 4.0, with a value of 2.5 being typical for the mixers used in wastewater. However, because the specific geometry of the internal mixing vanes used in such mixers varies with each manufacturer, the headloss or pressure-drop curves provided by the manufacturer must be used for estimation purposes. The power dissipated by static mixing devices can be computed using the following equation:

$$P = \gamma Qh \tag{5–8}$$

where P = power dissipated, kW
γ = specific weight of water, kN/m^3
Q = flowrate, m^3/s

Inline Mixers. Inline mixers are similar to static mixers but contain a rotating mixing element to enhance the mixing process. Typical examples of inline mixers are illustrated on Fig. 5–12(c) and (d). In the inline mixer shown on Fig. 5–12(c), the power required for mixing is supplied by an external source. For the mixer shown on Fig. 5–12(d), the power for mixing is supplied by the energy dissipation caused by the orifice plate and by the power input to the propeller mixer.

High-Speed Induction Mixer. The high-speed induction mixer is an efficient mixing device for a variety of chemicals. A proprietary device, shown on Fig. 5–12(e) for chlorine mixing, consists of a motor-driven open propeller that creates a vacuum in the chamber directly above the propeller. The vacuum created by the impeller induces the chemical to be mixed directly from the storage container without the need for dilution water. The high operating speed of the impeller (3450 rev/min) provides a thorough mixing of the chemical that is being added to the water by the high velocity of the fluid leaving the impeller of the mixing device.

Pressurized Water Jets. Pressurized water jet mixers, such as illustrated on Fig. 5–12(f), can also be used to mix chemicals. An important design feature of pressurized water jet mixers is that the velocity of the jet containing the chemical to be mixed must be sufficient to achieve mixing in all parts of the pipeline (Chao and Stone, 1979; Pratte and Baines, 1967). As shown on Fig. 5–12(f), a reactor tube has been added to achieve effective mixing. With pressurized water jet mixers, the power for mixing is provided by an external source (i.e., the solution feed pump).

Types of Mixers Used for Maintaining Solids in Suspension in Wastewater Treatment and Chemical Mixing

A variety of different impeller types are used for maintaining solids in suspension and mixing chemicals. Information on the types of mixing impellers used in wastewater treatment is presented in Table 5–10. Turbine and propeller and hyperboloid mixers are discussed further below.

Turbine and Propeller Mixers. Turbine and propeller mixers are used commonly in wastewater treatment processes for mixing and blending of chemicals, for keeping material in suspension, and for aeration. Turbine or propeller mixers are usually constructed with a vertical shaft driven by a speed reducer and electric motor. Two types of impellers are used for mixing: (1) radial-flow impellers and (2) axial-flow impellers (see Fig. 5–13). Radial-flow impellers generally have flat or curved blades located parallel to the axis of the shaft. The vertical flat-blade turbine impeller is a typical example of a radial-flow impeller. Axial-flow impellers make an angle of less than 90° with the drive shaft. Axial-flow impellers are further classified as variable pitch–constant angle of attack

Table 5–10

Typical types of mixing impellers used in wastewater treatment[a]

Type of impeller	Flow	Shear	Pumping capacity	Applications
Vertical flat blade turbine (VFBT)	Radial	High	Low	Vertical flow flash mixing, suspension of solids, gas dispersion
Disk turbine	Radial	High	Low	Mixing, gas dispersion
Hyperboloid	Radial	Very low	High	Suspension of solids, flocculation, mixing, gas dispersion
Surface impeller	Radial	High	Moderate	Gas transfer
Pitched blade turbine (45 or 32° PBT)	Axial	Moderate	Moderate	Horizontal flash mixing, suspension of solids
Low shear hydrofoil (LS)	Axial	Low	High	Horizontal flow flash mixing, suspension of solids, blending, flocculation
Propeller	Axial	Very low	High	Horizontal flow flash mixing, suspension of solids, blending, flocculation

[a] Adapted, in part, from Philadelphia Mixer Catalog.

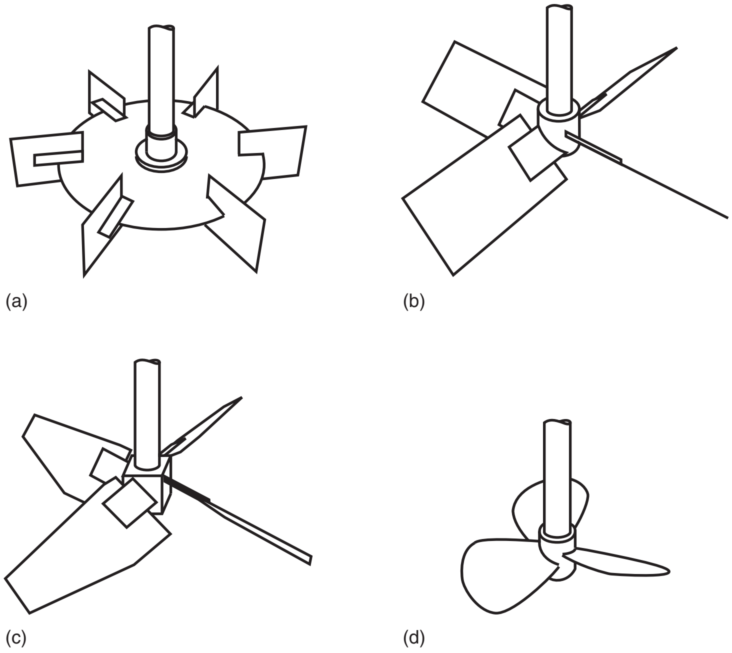


Figure 5–13

Typical impellers used for mixing: (a) disk-type radial-flow impeller, (b) axial-flow pitched (typically 45°) blade impeller, (c) axial-flow hydrofoil-type impeller, and (d) propeller mixer.

and constant pitch-variable angle of attack. Propellers and hydrofoils are typical examples. Propeller mixers may be provided with more than one set of propeller blades on a shaft.

Rapid mixing in wastewater-treatment processes usually occurs in the regime of turbulent flow in which inertial forces predominate. As a general rule, the higher the velocity and the greater the turbulence, the more efficient the mixing. On the basis of inertial and viscous forces, the following mathematical relationships can be used to estimate the power requirements for mixing and the pumping capacity of the mixer.

Power for mixing

$$P = N_P \rho n^3 D^5 \tag{5–9}$$

Pumping capacity

$$Q_i = N_Q n D^3 \tag{5–10}$$

where P = power input, W, (kg·m^2/s^3)
N_P = power number for impeller, unitless
ρ = density, kg/m^3
n = revolutions per second, r/s
D = diameter of impeller, m
Q_i = pump discharge, m^3/s
N_Q = flow number for impeller, unitless

Typical values for N_P and N_Q are presented in Table 5–11 for various types of impellers. It is important to consult manufacturers' catalogs for the appropriate values for N_P and N_Q for a specific piece of equipment. The values of N_P and N_Q must be adjusted for viscosity, blade characteristics, and the number of impellers on a single shaft.

Based on Eqs. (5–9) and (5–10), for a constant input of power as the impeller size is increased, more power is expended on flow and less on turbulence or shear. Thus, a small impeller operating at a high rotational speed will produce greater fluid shear and less pumping, whereas a large impeller operating at a slow speed will result in a high pumping

Table 5–11

Typical power and flow numbers for various impellers

Type of impeller	Power number, N_P	Flow number, N_Q	Pumping capacity
Vertical flat blade turbine(VFBT)	3.5–4.0	0.84–0.86	Low
Disk turbine			Low
Pitched blade turbine (45° PBT)	1.6	0.84–0.86	
Pitched blade turbine (32° PBT)	1.1	0.84–0.86	Moderate
Low shear hydrofoil (LS, 3-blade)	0.30	0.50	High
Low shear hydrofoil (LS, 4-blade)	0.60	0.55	High
Propeller			High

[a] Adapted, in part, from Philadelphia Mixer Catalog.

capacity and low fluid shear. Mixers with small impellers operating at high speeds are best for dispersing gases or small amounts of chemicals in wastewater. Mixers with slow-moving impellers are best for blending two fluid streams, or for flocculation.

Typically, Eq. (5–9) applies if the Reynolds number is in the turbulent range (greater than 10,000). For intermediate values of the Reynolds number, manufacturers' catalogs should be consulted. The Reynolds number is given by

$$N_R = \frac{D^2 n \rho}{\mu} \tag{5–11}$$

where D = diameter of impeller, m
n = rotational speed, rev/s
ρ = mass density of fluid, kg/m^3
μ = dynamic viscosity of fluid, N·s/m^2
Note: N = kg·m/s^2

Mixers are selected on the basis of laboratory or pilot-plant tests or similar data provided by manufacturers. No satisfactory method exists for scaling up from an agitator of one design to a unit of a different design. Geometrical similarity should be preserved, and the power input per unit volume should be kept the same.

Where propeller or turbine mixers are used, it is imperative that vortexing or mass swirling of the liquid be eliminated. Vortexing, in which the liquid to be mixed rotates with the impeller, causes a reduction in the difference between the fluid velocity and the impeller velocity and thus decreases the effectiveness of mixing. If the mixing vessel is fairly small, vortexing can be prevented by mounting the impellers off-center or at an angle with the vertical, or by having them enter the side of the basin at an angle. In circular and rectangular tanks the usual method used to limit vortexing is to install four or more vertical baffles extending approximately one-tenth the diameter out from the wall [see Fig. 5–14, see also Fig. 6–33(b)]. These baffles effectively break up the mass rotary motion and promote vertical mixing.

Typical design parameters for mixing operations, as given in Table 5–12, include (1) the velocity gradient G, subject to the caveats discussed previously, (2) the rotational

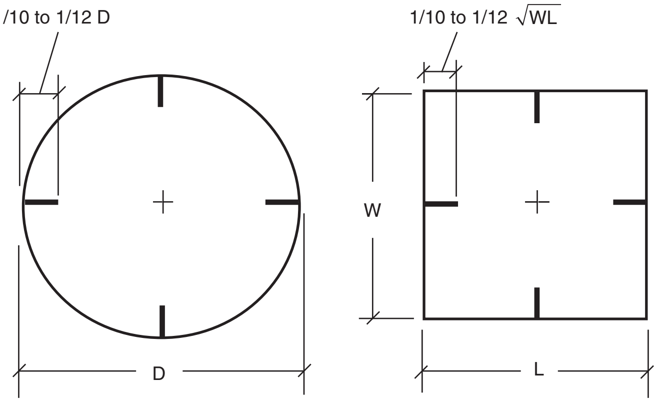


Figure 5–14

Definition sketch for the placement of baffles to limit vortexing in mixing tanks and reactors.

speed, and (3) the ratio of the impeller diameter to the equivalent tank diameter. The rotation speed will vary considerably, depending on whether the flow through the mixer is horizontal or vertical (see Fig. 5–15).

Hyperboloid Mixers. Hyperboloid mixers are used in water and wastewater treatment processes in a variety of applications including mixing and blending of chemicals, for flocculation, suspension of biosolids in anaerobic and anoxic basins, and for aeration in activated sludge and in sludge basins (see Fig. 5–16). Typically hyperboloid mixers are designed with a vertical shaft driven by a speed reducer and electric motor. The size of the hyperboloid impeller and the rotational speed depends on the basin size and geometry. The hyperboloid mixers have the advantage of being mounted close to the bottom of the tank which allows for the energy introduction where it is needed to suspend sludge floc.

Types of Mixers Used for Flocculation in Wastewater Treatment

The principal types of mixers used for flocculation can be classified as (1) static mixers, (2) paddle mixers, and (3) turbine, propeller, and hyperboloid mixers. Each of these types of mixers is considered briefly in the following discussion.

Table 5–12

Typical design parameters for mixing operations[a]

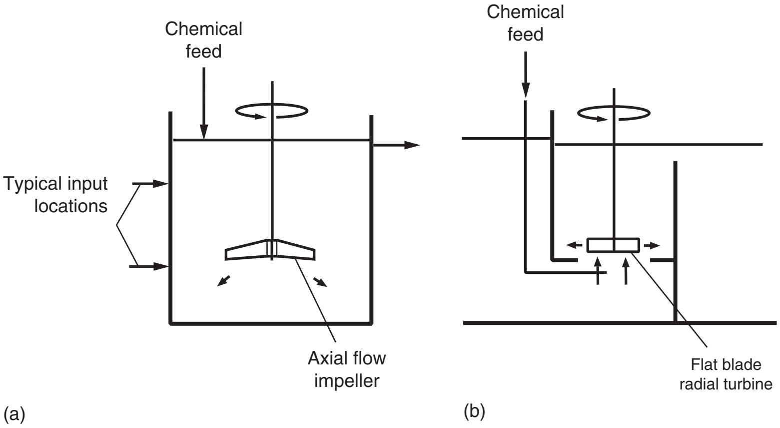
Parameter	Symbol	Unit	Value
Horizontal flow mixing			
Velocity gradient	G	1/s	500–2500
Rotational speed	n	rev/min	40–125
Ratio impeller diameter to equivalent tank diameter[b]	D/T_e	unitless	0.25–0.40
Vertical flow mixing			
Velocity gradient	G	1/s	500–2500
Rotational speed	n	rev/min	25–45
Ratio impeller diameter to equivalent tank diameter	D/T_e	unitless	0.40–0.60

[a] Adapted, in part, from Philadelphia Mixer Catalog.

[b] $T_e \approx 1.13\sqrt{L \times W}$ where L = length and W = width.

Figure 5–15

Definition sketch for rapid mixing in a tank or channel (typically mixing times are 10 to 30 s): (a) horizontal flow through a mixing tank equipped with an axial flow impeller and (b) vertical flow through a mixing basin equipped with radial flow impeller.

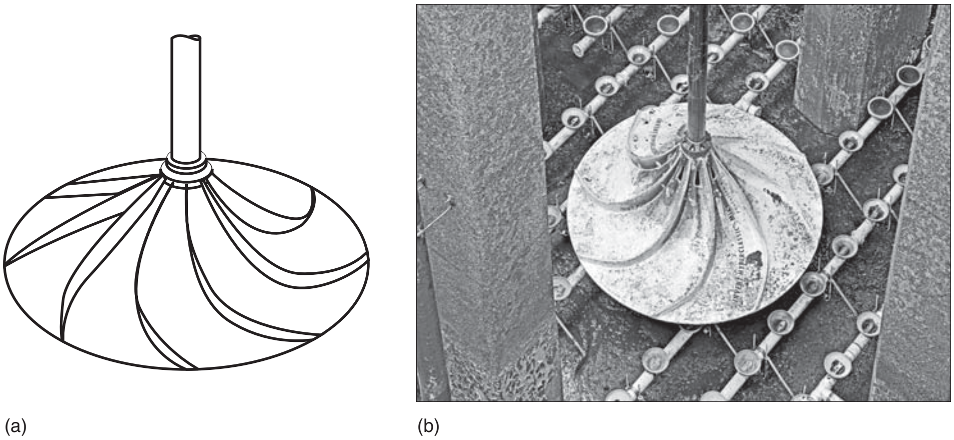


Baffled Static Mixers. In the most common type of baffled static mixer, the liquid to be treated is subjected to a series of flow reversals in which the direction of flow is changed. Baffled static mixers can be composed of over and under narrow flow channels, such as shown on Fig. 5–17(a), or the narrow flow channels can be laid out horizontally. The headloss caused by frictional resistance offered by the flow channels and the flow reversals provides the energy for flocculation. In some designs, the channel spacing is varied to provide a decreasing energy gradient so that the large floc particles formed toward the end of the flocculation basin will not be broken apart.

Paddle Mixers. Paddle mixers are used as flocculation devices when coagulants, such as aluminum or ferric sulfate, and coagulant aids, such as polyelectrolytes and lime, are added to wastewater or solids (sludge). Paddle flocculators consist of a series of appropriately spaced paddles mounted on either a horizontal or vertical shaft. Flocculation is promoted by gentle mixing brought about by the slow-moving paddles, which, as shown on Fig. 5–17(b), rotate the liquid and promote mixing. Increased particle contact promotes floc growth, but, if the mixing is too vigorous, the increased shear forces will break up the floc into smaller particles. Agitation should be controlled carefully so that the floc particles will be of suitable size and will settle readily. Variable-speed drives are often used to regulate the paddle speed. There has been a movement away from paddle flocculators to the

Figure 5–16

Hyperboloid mixer: (a) schematic and (b) view of mixer (air diffusers below the mixer have been blanketed off for the mixer to work properly).



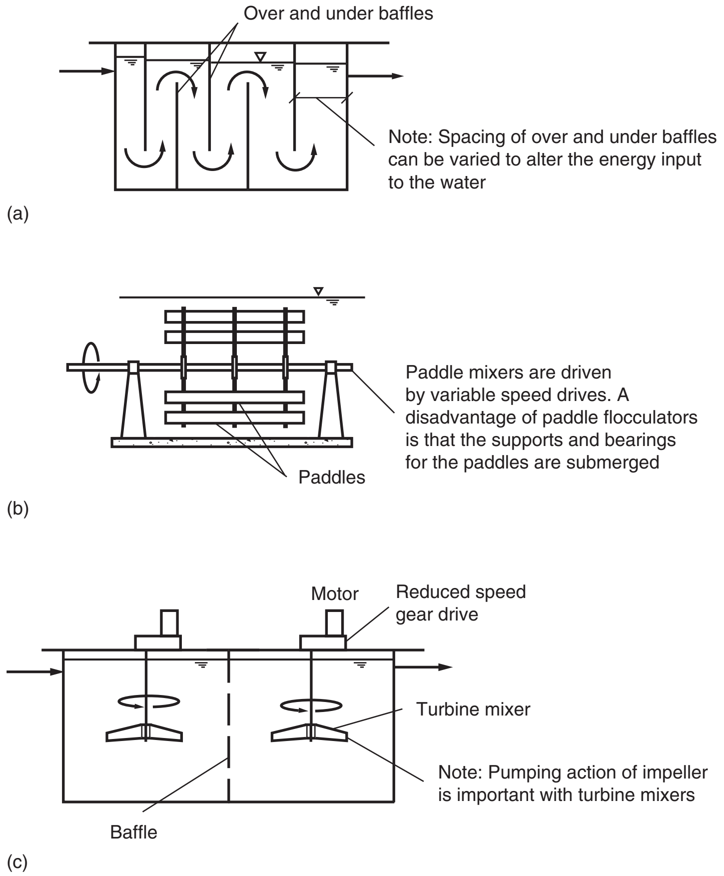


Figure 5–17

Typical mixers used for flocculation in wastewater treatment facilities: (a) over and under baffled reactor, (b) paddle mixer in baffled tank, and (c) turbine mixer in a baffled tank.

use of turbine flocculators because of the maintenance problems associated with paddle flocculators.

Power in a mechanical paddle system can be related to the drag force on the paddles as follows.

$$F_D = \frac{C_D A \rho v_p^2}{2} \tag{5–12}$$

$$P = F_D v_p = \frac{C_D A \rho v_p^3}{2} \tag{5–13}$$

where F_D = drag force, N
C_D = coefficient of drag of paddle moving perpendicular to fluid
A = cross-sectional area of paddles, m^2
ρ = mass density of fluid, kg/m^3
v_p = relative velocity of paddles with respect to the fluid, m/s, usually assumed to be 0.6 to 0.75 times the paddle-tip speed
P = power requirement, W ($kg \cdot m^2/s^3$)

The application of Eq. (5–13) is illustrated in Example 5–3.

EXAMPLE 5–3 **Power Requirements and Paddle Area for a Wastewater Flocculator**

Determine the theoretical power requirement and the paddle area required to achieve a G value of 50/s in a tank with a volume of 3000 m^3. Assume that the water temperature is 15°C, the coefficient of drag C_D for rectangular paddles is 1.8, the paddle-tip velocity υ is 0.6 m/s, and the relative velocity of the paddles υ_p is 0.75 υ.

Solution

1. Determine the theoretical power requirement using Eq. (5–3), rearranged as shown below.

$$\mu \text{ at } 15^\circ\text{C} = 1.139 \times 10^{-3} \text{ N·s/m}^2 \text{ (see Appendix C)}$$

$$\begin{aligned} P &= G^2 \mu V \\ &= (50/\text{s})^2 (1.139 \times 10^{-3} \text{ N·s/m}^2)(3000 \text{ m}^3) \\ &= 8543 \text{ (kg·m}^2/\text{s}^3) = 8543 \text{ W} \\ &= 8.54 \text{ kW} \end{aligned}$$

2. Determine the required paddle area using Eq. (5–13).

$$\rho = 999.1 \text{ kg/m}^3 \text{ (see Appendix C)}$$

$$\begin{aligned} A &= \frac{2P}{C_D \rho \upsilon_p^3} \\ &= \frac{2(8543 \text{ kg/m}^2 \cdot \text{s}^3)}{1.8(999.1 \text{ kg/m}^3)(0.75 \times 0.6 \text{ m/s})^3} \\ &= 104.3 \text{ m}^2 \end{aligned}$$

Turbine, Propeller, and Hyperboloid Flocculators. The rotating element of turbine- and propeller-type flocculators consists of three or four blades attached to a vertical shaft [see Fig. 5–17(c)]. The hyperboloid-type flocculator consists of a hyperboloid shaped mixer body that has 8 integrated so-called motion fins. Hyperboloid-type mixer was shown previously in Fig. 5–16. The flocculator is driven with an external gear reduction system powered with a variable-speed drive. The blades of the propeller may be rectangular in shape or have the shape of a hydrofoil. Blades shaped as hydrofoils or hyperboloid impellers are used to limit the amount of floc shearing while at the same time providing the velocity gradients and pumping capacity needed for mixing. Typical design parameters for flocculation facilities are presented in Table 5–13. In sizing turbine, propeller, or hyperboloid flocculators, both the power and the pumping requirements must be considered. In addition, the tip speed and the superficial velocity must also be considered. The required power and the pumping capacity can be estimated using Eqs. (5–9) and (5–10), presented previously. In wastewater flocculation, if variable-speed drives are provided for the flocculators (considering the minimum G required to keep particles in suspension), the mixer speed can be adjusted to optimize flocculation and energy use.

Types of Mixers Used for Continuous Mixing in Wastewater Treatment

Continuous mixing operations are used in biological treatment processes such as the activated-sludge process to maintain the mixed liquor suspended solids uniformly mixed state. In biological treatment systems the mixing device is also used to provide the oxygen needed for the process. Thus, the aeration equipment must be able to provide the oxygen needed for the process and the energy needed to maintain mixed conditions within the reactor.

Table 5–13

Typical design parameters for flocculation facilities[a]

Parameter	Symbol	Unit	Value
Velocity gradient	G	1/s	100–500
Rotational speed	n	rev/min	10–30
Ratio length to width	L/W	unitless	$1 \leq L/W \leq 1.25$
Ratio impeller diameter to equivalent tank diameter[b]	D/T_e	unitless	0.35–0.45
Ratio height to equivalent tank diameter[b]	H/T_e	unitless	0.9–1.1
Tip speed			
Flat blade turbine	TS	m/s	0.6–1.5
Pitch blade turbine (45 or 32°)	TS	m/s	1.8–2.4
Low shear propeller (3 or 4 blade)	TS	m/s	2–2.7
Superficial velocity[c]	SV	m/min	1–2
Hyperboloid impeller (8 motion fins)	TS	m/s	1.8–3.0

[a] Adapted, in part, from Philadelphia Mixer Catalog.

[b] $T_e = 1.13\sqrt{L \times W}$ where L = length and W = width in m.

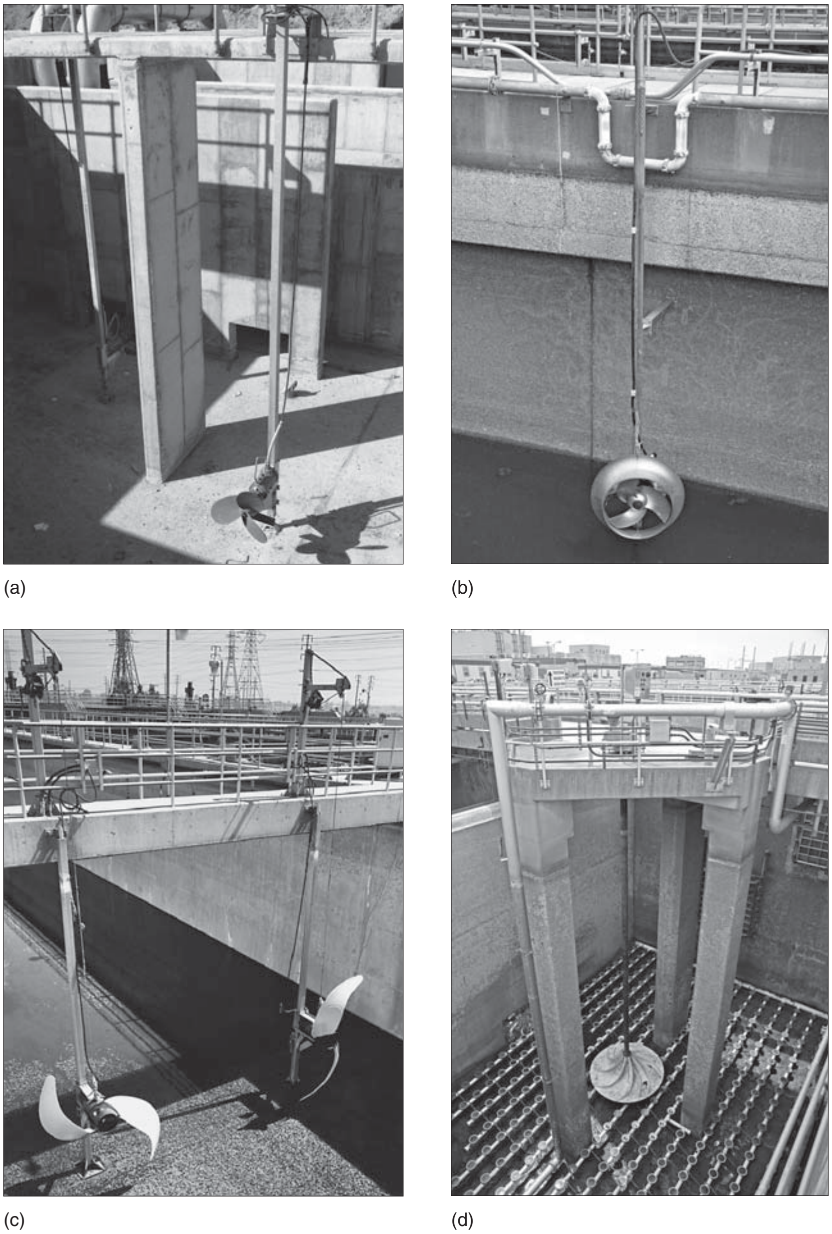
[c] $SV = Q/A$ where Q is the pumping rate and A is the basin cross-sectional area.

Both mechanical aerators and dissolved aeration devices are used. Diffused air is often used to fulfill both the mixing and oxygen requirements for aeration tanks and for maintaining solids in suspension in distribution channels with low velocities carrying raw, unsettled wastewater or mixed liquor. Alternatively, mechanical turbine-aerator mixers may be used. Hyperboloid mixer/aerators can be used in situations where a large variety in oxygen levels can be expected.

Where biological bioreactors are being used for biological nutrient removal, it is essential that the feed channels to the bioreactors and the anaerobic or anoxic zones within the bioreactor be non-aerated. Horizontal, submersible propeller mixers provide directional mixing and can be used to maintain channel velocities in oxidation ditches operating at low dissolved oxygen levels or to mix the contents of anoxic or anaerobic reactor zones, (see Fig. 5–18). Hyperboloid mixers can be used for non-aerated mixing of influent feed channels and are particularly applicable for non-directional mixing of anaerobic or anoxic zones with low depth to width ratios [see Fig. 5–16(b)]. In general, hyperboloid mixers will be more efficient than other types of mixing equipment.

Pneumatic Mixing. In pneumatic mixing, a gas (usually air or oxygen) is injected into the bottom of mixing or activated-sludge tanks, and the turbulence caused by the rising gas bubbles serves to mix the fluid contents of the tank. In aeration, soft bubbles are formed with an average diameter of 5 mm while the air flow is about 10 percent of the liquid flow. The velocity gradients due to bubble formation range from a $G_{\text{avg}} < 200\ \text{s}^{-1}$ to $G_{\max} = 8200\ \text{s}^{-1}$ (Masschelein, 1992).

Where air flocculation is employed, the air supply system should be adjustable so that the flocculation energy level can be varied throughout the tank. When air is injected in



(a) (b) (c) (d)

Figure 5–18

Mixers used to mix the contents of an anoxic biological reactor: (a) horizontal, submersible propeller, (b) submersible propeller with shroud, (c) airfoil mixer, and (d) hyperbolic mixer. (Air diffusers below the mixer have been blanked off. This tank has been designed to either operate with mixing or with diffused air.)

mixing or flocculation tanks or channels, the power dissipated by the rising air bubbles can be estimated with the following equation:

$$P = p_a V_a \ln \frac{p_c}{p_a} \tag{5–14}$$

where P = power dissipated, kW
p_a = atmospheric pressure, kN/m^2
V_a = volume of air at atmospheric pressure, m^3/s
p_c = air pressure at the point of discharge, kN/m^2

Equation (5–14) is derived from a consideration of the work done when the volume of air released under compressed conditions expands isothermally. If the flow of air at atmospheric pressure is expressed in terms of m^3/min (ft^3/min) and the pressure is expressed in term of meters (feet) of water, Eq. (5–14) can be written as follows:

$$P = KQ_a \ln\left(\frac{h + 10.33}{10.33}\right) \qquad \text{S.I. units} \tag{5–15a}$$

$$P = KQ_a \ln\left(\frac{h + 33.9}{33.9}\right) \qquad \text{U.S. customary units} \tag{5–15b}$$

where K = constant 1.689 (35.28 in U.S. customary units)
Q_a = air flowrate at atmospheric pressure, m^3/min (ft^3/min)
h = air pressure at the point of discharge expressed in meters of water, m (ft)

The velocity gradient G achieved in pneumatic mixing is obtained by substituting P from Eq. (5–15) into Eq. (5–3).

Mechanical Aerators and Mixers. The principal types of mechanical aerators used for continuous mixing are high- or low-speed surface aerators and hyperboloid mixer/aerators. These devices are discussed in Sec. 5–11, which deals with aeration, and in Chap 8. Typical power requirements for mixing with surface aerators range from 20 to 40 kW/10^3 m^3 (0.75 to 1.50 hp/10^3 ft^3), depending on the type of mixer and the geometry of the tank, lagoon, or basin. For hyperboloid mixer/aerators the power requirements for mixing are significantly lower because the mixing and the aeration functions are separated. The typical range depending on geometry of the tank, lagoon, or basin is 2 to 4 kW/10^3 m^3 (0.08 to 0.15 hp/10^3 ft^3).

New Developments in Mixing Technology

New analytical tools that are now being applied to the analysis of and design of mixing devices include (1) computational fluid dynamics (CFD), (2) digital particle image velocimetry (DPIV), (3) laser doppler anemometry (LDA), and (4) laser-induced fluorescence (LIF). Computational fluid dynamics is used to model the fluid flow patterns in mixing devices and for scale-up analysis. In respect to fluid flow, both two- and three-dimensional models are now available. Digital particle image velocimetry is used to understand fluid movement in mixing devices. The movement of neutrally buoyant fluorescent particles is photographed using laser beam illumination. Laser doppler anemometry is used to study turbulence and to obtain data on the mean velocity at a given location in the mixing chamber. To evaluate the mean velocity, two laser beams are focused so that the beams intersect. As a particle passes through the intersection of the beams, light is reflected. The wavelength of the reflected light is a function of the particle velocity. Laser-induced fluorescence is used to measure the degree of mixing of solutions. Dyes such as rhodimine and other materials will fluoresce when struck by laser light of a given wavelength. The scattering of light is measured to assess the degree of mixing. This technique is being used to study the diffusion and mixing of a substance by assessing the coefficient of variation of the mixed solution and to evaluate blending times.

5–4 GRAVITY SEPARATION THEORY

The removal of suspended and colloidal materials from wastewater by gravity separation is one of the most widely used physical unit processes in wastewater treatment. A summary of gravitational phenomena utilized in wastewater treatment is presented in Table 5–14.

Table 5–14

Types of gravitational phenomena utilized in wastewater treatment

Type of separation phenomenon	Description	Application/occurrence
Discrete particle settling	Refers to the settling of particles in a suspension of low solids concentration by gravity in a constant acceleration field. Particles settle as individual entities, and there is no significant interaction with neighboring particles	Removal of grit and sand particles from wastewater
Flocculant settling	Refers to a rather dilute suspension of particles that coalesce, or flocculate, during the settling operation. By coalescing, the particles increase in mass and settle at a faster rate	Removal of a portion of the TSS in untreated wastewater in primary settling facilities, and in upper portions of secondary settling facilities. Also removes chemical floc in settling tanks
Ballasted flocculant settling	Refers to the addition of an inert ballasting agent and a polymer to a partially flocculated suspension to promote rapid settling and improved solids reduction. A portion of the recovered ballasting agent is recycled to the process	Removal of a portion of the TSS in untreated wastewater, wastewater from combined systems, and industrial wastewater. Also reduces BOD and phosphorus
Hindered settling (also called zone settling)	Refers to suspensions of intermediate concentration, in which interparticle forces are sufficient to hinder the settling of neighboring particles. The particles tend to remain in fixed positions with respect to each other, and the mass of particles settles as a unit. A solids-liquid interface develops at the top of the settling mass	Occurs in secondary settling facilities used in conjunction with biological treatment facilities
Compression settling	Refers to settling in which the particles are of such concentration that a structure is formed, and further settling can occur only by compression of the structure. Compression takes place from the weight of the particles, which are constantly being added to the structure by sedimentation from the supernatant liquid	Usually occurs in the lower layers of a deep solids or biosolids mass, such as in the bottom of deep secondary settling facilities and in solids-thickening facilities
Accelerated gravity settling	Removal of particles in suspension by gravity settling in an acceleration field	Removal of grit and sand particles from wastewater
Flotation	Removal of particles in suspension that are lighter than water by air or gas flotation	Removal of greases and oils, light material that floats, thickening of solids suspensions

Sedimentation is the term applied to the separation of suspended particles that are heavier than water, by gravitational settling. The terms sedimentation and settling are used interchangeably. A sedimentation basin may also be referred to as a sedimentation tank, clarifier, settling basin, or settling tank. Accelerated gravity settling involves the removal of particles in suspension by gravity settling in an accelerated flow field. The fundamentals of gravity separation are introduced in this section. The design of facilities for the removal of grit and TSS are considered in Secs. 5–6 and 5–7, respectively.

Description

Sedimentation is used for the removal of grit, TSS in primary settling basins, biological floc removal in the activated-sludge settling basin, and chemical floc removal when the chemical coagulation process is used. Sedimentation is also used for solids concentration in sludge thickeners. In most cases, the primary purpose is to produce a clarified effluent, but it is also necessary to produce sludge with a solids concentration that can be handled and treated easily.

On the basis of the concentration and the tendency of particles to interact, four types of gravitational settling can occur: (1) discrete particle, (2) flocculent, (3) hindered (also called zone), and (4) compression. Because of the fundamental importance of the separation processes in the treatment of wastewater, the analysis of each type of separation process is discussed separately. In addition, tube settlers, used to enhance the performance of sedimentation facilities, are also described. Other gravitational separation processes include high-rate clarification, accelerated gravity settling, and flotation and are discussed subsequently.

Particle Settling Theory

The settling of discrete, nonflocculating particles can be analyzed by means of the classic laws of sedimentation formed by Newton and Stokes. Newton's law yields the terminal particle velocity by equating the gravitational force of the particle to the frictional resistance, or drag. The gravitational force is given by

$$F_G = (\rho_p - \rho_w) g V_p \tag{5–16}$$

where F_G = gravitational force, MLT^{-2} ($kg{\cdot}m/s^2$)
ρ_p = density of particle, ML^{-3} (kg/m^3)
ρ_w = density of water, ML^{-3} (kg/m^3)
g = acceleration due to gravity, LT^{-2} (9.81 m/s^2)
V_p = volume of particle, L^3 (m^3)

The frictional drag force depends on the particle velocity, fluid density, fluid viscosity, particle diameter, and the drag coefficient C_d (dimensionless), and is given by Eq. (5–17).

$$F_d = \frac{C_d A_p \rho_w \upsilon_p^2}{2} \tag{5–17}$$

where F_d = frictional drag force, MLT^{-2} ($kg{\cdot}m/s^2$)
C_d = drag coefficient (unitless)
A_p = cross-sectional or projected area of particles in direction of flow, L^2 (m^2)
υ_p = particle settling velocity, LT^{-1} (m/s)

Equating the gravitational force to the frictional drag force for spherical particles yields Newton's law:

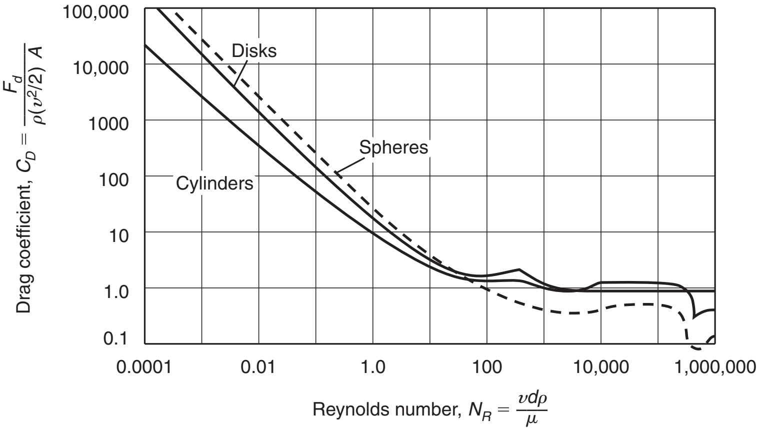
$$\upsilon_{p(t)} = \sqrt{\frac{4g}{3C_d}\left(\frac{\rho_p - \rho_w}{\rho_w}\right) d_p} \approx \sqrt{\frac{4g}{3C_d}(sg_p - 1) d_p} \tag{5–18}$$

where $\upsilon_{p(t)}$ = terminal velocity of particle, LT^{-1} (m/s)
d_p = diameter of particle, L (m)
sg_p = specific gravity of the particle

The coefficient of drag C_d takes on different values depending on whether the flow regime surrounding the particle is laminar or turbulent. The drag coefficient for various particles is shown on Fig. 5–19 as a function of the Reynolds number. As shown on Fig. 5–19, there are three more or less distinct regions, depending on the Reynolds number: laminar ($N_R < 1$), transitional (N_R = 1 to 2000), and turbulent ($N_R > 2000$).

Figure 5–19

Coefficient of drag as a function of Reynolds number.



Although particle shape affects the value of the drag coefficient, for particles that are approximately spherical, the curve on Fig. 5–19 is approximated by the following equation (upper limit of $N_R = 10^4$):

$$C_d = \frac{24}{N_R} + \frac{3}{\sqrt{N_R}} + 0.34 \tag{5–19}$$

The Reynolds number, N_R, for the settling of spherical particles is defined as

$$N_R = \frac{v_p d_p \rho_w}{\mu} = \frac{v_p d_p}{\nu} \tag{5–20}$$

where ν = kinematic viscosity, L^2T^{-1} (m^2/s)
μ = dynamic viscosity, MTL^{-2} ($N{\cdot}s/m^2$)

Other terms are as defined above.

Non Spherical Particles. Because most of the particles in wastewater are nonspherical, a sphericity factor, ψ, is introduced in the Reynolds number as shown in Eq. (5–21).

$$N_R = \frac{v_p d_p \rho_w \psi}{\mu} = \frac{v_p d_p \psi}{\nu} \tag{5–21}$$

where ψ = sphericity, dimensionless
Other terms are as defined above.

The sphericity factor is the ratio of the surface area of a sphere with the same volume as a given particle to the surface area of the particle [see Eq. (11–11) in Chap.11]. Sphericity factors vary from 1.0 for spheres to about 0.70 for crushed sand. An alternative approach to account for non spherical particles is to multiply the coefficient of drag by a shape factor, ϕ (Degremont, 2007). Typical values of ϕ for sand and anthracite are 2 and 2.25, respectively. Values for flocculent particles can range as high as 15 to 25 and higher. Because ψ, as defined above, is essentially equal to $1/\phi$, this latter approach is more conservative than the use of Eq. (5–21), and is used mostly for flocculent particles (see Example 5–10).

Settling in the Laminar Region. For Reynolds numbers less than about 1.0, viscosity is the predominant force governing the settling process, and the first term in

Eq. (5–19) predominates. Assuming spherical particles, substitution of the first term of the drag coefficient equation [Eq. (5–19)] into Eq. (5–18) yields Stokes' law:

$$v_p = \frac{g(\rho_p - \rho_w)d_p^2}{18\mu} \approx \frac{g(sg_p - 1)d_p^2}{18\nu} \tag{5–22}$$

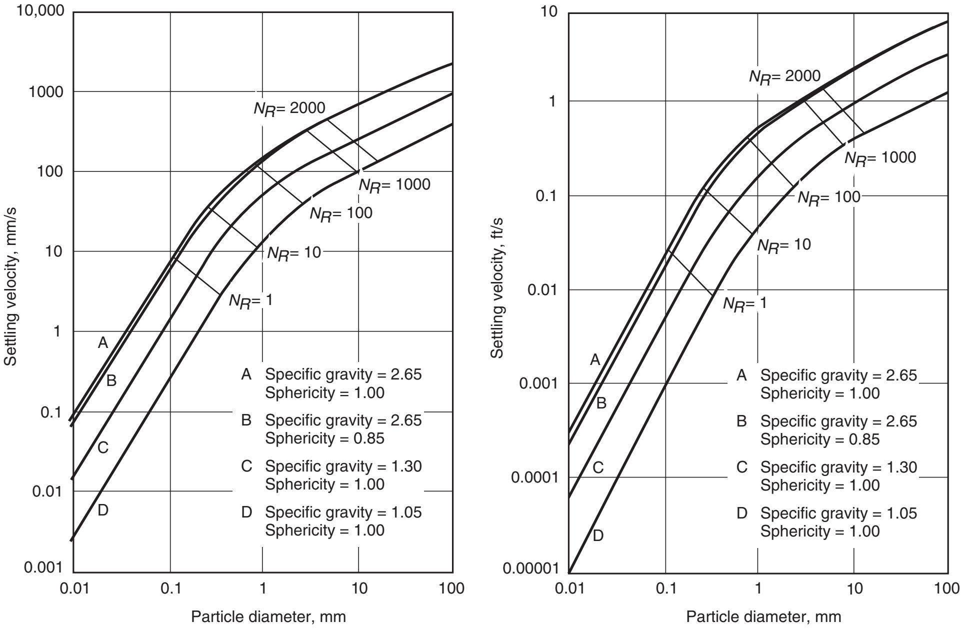
Terms are as defined previously. For laminar-flow conditions, Stokes found the drag force to be

$$F_D = 3\pi\mu v_p d_p \tag{5–23}$$

Stokes' law [Eq. (5–22)] can also be derived by equating the drag force found by Stokes to the effective weight of the particle [Eq. (5–16)].

Settling in the Transition Region. In the transition region, the complete form of the drag equation [Eq. (5–19)] must be used to determine the settling velocity, as illustrated in Example 5–4. Because of the nature of the drag equation, finding the settling velocity is an iterative process. As an aid in visualizing settling in the transition region, Fig. 5–20 has been prepared, which covers the laminar and the transition region for particle sizes of interest in environmental engineering.

Settling in the Turbulent Region. In the turbulent region, inertial forces are predominant, and the effect of the first two terms in the drag coefficient equation [Eq. (5–19)] is reduced. For settling in the turbulent region, a value of 0.4 is used for the



(a) (b)

Figure 5–20

Settling velocities for various particle sizes under varying conditions: (a) settling velocity in mm/s versus particle size in mm, and (b) settling velocity in ft/s versus particle size in mm (Crites and Tchobanoglous, 1998).

coefficient of drag. If a value of 0.4 is substituted into Eq. (5–21) for C_d, the resulting equation is

$$v_p = \sqrt{3.33\,g\left(\frac{\rho_p - \rho_w}{\rho_w}\right)d_p} \approx \sqrt{3.33\,g(sg_p - 1)\,d_p} \tag{5–24}$$

The use of Eqs. (5–18) through (5–22) is illustrated in Example 5–4.

EXAMPLE 5–4 **Determination of Particle Terminal Settling Velocity** Determine the terminal settling velocity for a sand particle with an average diameter of 0.5 mm (0.00164 ft), a sphericity factor of 0.85, and a specific gravity of 2.65, settling in water at 20°C (68°F). At this temperature, the kinematic viscosity value given in Appendix C is 1.003×10^{-6} m²/s (1.091×10^{-5} ft²/s).

Solution

1. Determine the terminal settling velocity for the particle using Stokes' Law (Eq. 5–22).

$$\begin{aligned} v_p &= \frac{g(sg_p - 1)d_p^2}{18\nu} \\ &= \frac{(9.81\ \text{m/s}^2)(2.65 - 1)(0.5 \times 10^{-3}\ \text{m})^2}{18(1.003 \times 10^{-6}\ \text{m}^2/\text{s})} = 0.224\ \text{m/s} \end{aligned}$$

2. Check the Reynolds number with Eq. (5–21) which includes the sphericity factor, ψ.

$$N_R = \frac{v_p d_p \psi}{\nu} = \frac{(0.224\ \text{m/s})(0.5 \times 10^{-3}\ \text{m})(0.85)}{(1.003 \times 10^{-6}\ \text{m}^2/\text{s})} = 94.9$$

The use of Stokes' Law is not appropriate for Reynolds number > 1.0. Therefore, Newton's Law (Eq. 5–18) must be used to determine the settling velocity in the transition region (see Fig. 5–20). The drag coefficient term in Newton's equation is dependent on the Reynolds number, which is a function of the settling velocity. Because the settling velocity is not known, an initial settling velocity must be assumed. The assumed velocity is used to compute the Reynolds number, which is used to determine the drag coefficient, which is used in the Newton equation to calculate the settling velocity. A solution is achieved when the initial assumed settling velocity is approximately equal to the settling velocity resulting from Newton's equation. The solution process is iterative, as illustrated below.

3. For the first assumed settling velocity, use the Stokes' Law settling velocity calculated above. Using the resulting Reynolds number, also determined previously, compute the drag coefficient.

$$C_d = \frac{24}{N_R} + \frac{3}{\sqrt{N_R}} + 0.34 = \frac{24}{94.9} + \frac{3}{\sqrt{94.9}} + 0.34 = 0.901$$

4. Use the drag coefficient in Newton's equation to determine the particle settling velocity.

$$v_p = \sqrt{\frac{4g(sg - 1)d}{3C_d}} = \sqrt{\frac{4(9.81\ \text{m/s}^2)(2.65 - 1)(0.5 \times 10^{-3}\ \text{m})}{3 \times 0.901}} = 0.109\ \text{m/s}$$

Because the initial assumed settling velocity (0.224 m/s) does not equal the settling velocity calculated by Newton's equation (0.109 m/s), a second iteration is necessary.

5. For the second iteration, assume a settling velocity value of 0.09 m/s, and calculate the Reynolds number. Use the Reynolds number to determine the drag coefficient, and use the drag coefficient in Newton's equation to find the settling velocity.

$$N_R = \frac{(0.09\text{ m/s})(0.5 \times 10^{-3}\text{ m})(0.85)}{(1.003 \times 10^{-6}\text{ m}^2\text{/s})} = 38.1$$

$$C_d = \frac{24}{38.1} + \frac{3}{\sqrt{38.1}} + 0.34 = 1.456$$

$$v_p = \sqrt{\frac{4(9.81\text{ m/s}^2)(2.65 - 1)(0.5 \times 10^{-3}\text{ m})}{3 \times 1.456}} = 0.086\text{ m/s}$$

Although the assumed settling velocity (0.09 m/s) and the calculated settling velocity (0.086 m/s) still do not agree, they are approaching closure. Successive iterations to calculate the actual settling velocity will be done as a homework problem.

Because the settling velocity used to compute the Reynolds number agrees with the settling velocity value from Newton's equation, the solution approach has been confirmed.

Discrete Particle Settling

In the design of sedimentation basins, the usual procedure is to select a particle with a terminal velocity v_c and to design the basin so that all particles that have a terminal velocity equal to or greater than v_c will be removed. The rate at which clarified water is produced is equal to

$$Q = Av_c \tag{5–25}$$

where Q = flowrate, L^3T^{-1} (m^3/s)
A = surface of the sedimentation basin, L^2 (m^2)
v_c = particle settling velocity, LT^{-1} (m/s)

Rearranging Eq. (5–25) yields

$$v_c = \frac{Q}{A} = \text{overflow rate}, LT^{-1}(\text{m}^3/\text{m}^2 \cdot \text{d})$$

Thus, the critical velocity is equivalent to the overflow rate or surface loading rate. A common basis of design for discrete particle settling recognizes that the flow capacity is independent of the depth.

For continuous-flow sedimentation, the length of the basin and the time a unit volume of water is in the basin (detention time) should be such that all particles with the design velocity v_c will settle to the bottom of the tank. The design velocity, detention time, and basin depth are related as follows:

$$v_c = \frac{\text{depth}}{\text{detention time}} \tag{5–26}$$

In actual practice, design factors must be adjusted to allow for the effects of inlet and outlet turbulence, short circuiting, sludge storage, and velocity gradients due to the operation of sludge-removal equipment. These design factors are discussed in Sec. 5–6. In the above discussion ideal settling conditions have been assumed.

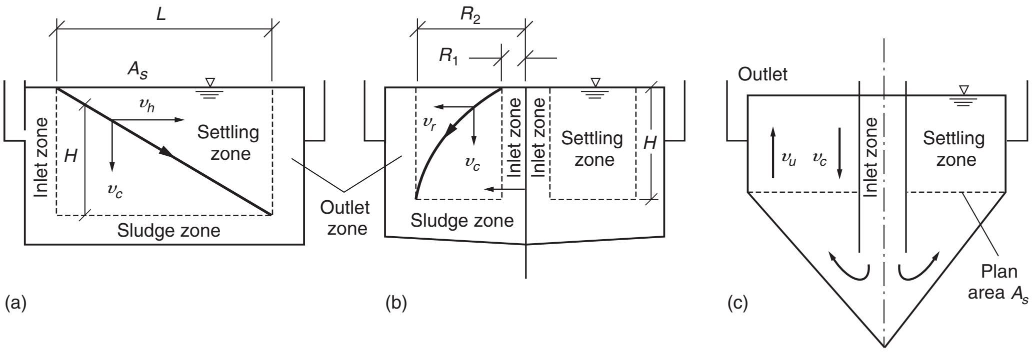


Figure 5–21

Definition sketch for the idealized settling of discrete particles in three different types of settling basins: (a) rectangular, (b) circular, and (c) upflow (Crites and Tchobanoglous, 1998).

Idealized discrete particle settling in three different types of settling basins is illustrated on Fig. 5–21. Particles that have a velocity of fall less than v_c will not all be removed during the time provided for settling. Assuming that the particles of various sizes are uniformly distributed over the entire depth of the basin at the inlet, it can be seen from an analysis of the particle trajectory on Fig. 5–22 that particles with a settling velocity less than v_c will be removed in the ratio

$$X_r = \frac{v_p}{v_c} \tag{5–27}$$

where X_r is the fraction of the particles with settling velocity v_p that are removed.

In most suspensions encountered in wastewater treatment, a large gradation of particle sizes will be found. To determine the efficiency of removal for a given settling time, it is necessary to consider the entire range of settling velocities present in the system. The settling velocities of the particles can be obtained by use of a settling column test. The particle settling data are used to construct a velocity settling curve as shown on Fig. 5–23.

For a given clarification rate Q where

$$Q = v_c A \tag{5–28}$$

Figure 5–22

Definition sketch for the analysis of ideal discrete particle settling.

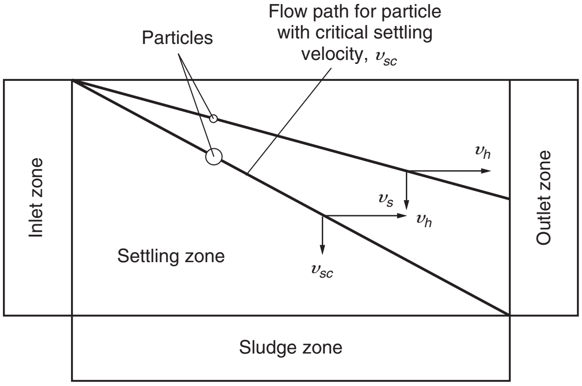
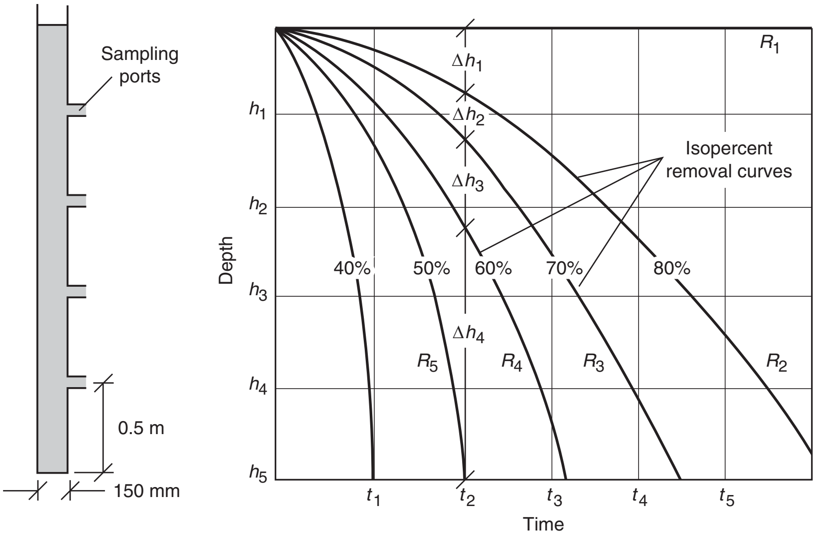


Figure 5–23

Definition sketch for the analysis of flocculent settling.



only those particles with a velocity greater than v_c will be removed completely. The remaining particles will be removed in the ratio v_p/v_c. The total fraction of particles removed for a continuous distribution is given by Eq. (5–29).

$$\text{Fraction removed} = (1 - X_c) + \int_0^{X_c} \frac{v_p}{v_c} dx \tag{5–29}$$

where $1 - X$ = fraction of particles with velocity v_p greater than v_c

$\int_0^{X_c} \frac{v_p}{v_c} dx$ = fraction of particles removed with v_p less than v_c

For discrete particles within a given settling velocity range, the following expression may be used

$$\text{Total fraction removed} = \frac{\sum_{i=1}^{n} \frac{v_{n_i}}{v_c}(n_i)}{\sum_{i=1}^{n} n_i} \tag{5–30}$$

where v_{n_i} = average velocity of particles in the ith velocity range

n_i = number of particles in the ith velocity range

The use of Eq. (5–30) is illustrated in Example 5–5.

EXAMPLE 5–5 Calculation of Removal Efficiency for a Primary Sedimentation Basin

Determine the removal efficiency for a sedimentation basin with a critical overflow velocity of 2 $m^3/m^2 \cdot h$ in treating a wastewater containing particles whose settling velocities are distributed as given in the table below. Plot the particle histogram for the influent and effluent wastewater.

Settling velocity, m/h	Number of particles per liter × 10^{-5}
0.0–0.5	30
0.5–1.0	50
1.0–1.5	90
1.5–2.0	110
2.0–2.5	100
2.5–3.0	70
3.0–3.5	30
3.5–4.0	20
Total	500

Solution

1. Create a table for calculating the percentage removal for each particle size. Enter the particle settling velocity ranges in column (1).

Settling velocity range, m/h (1)	Average settling velocity, m/h (2)	Number of particles in influent, × 10^{-5} (3)	Fraction of particles removed (4)	Number of particles removed, × 10^{-5} (5)	Particles remaining in effluent, × 10^{-5} (6)
0.0–0.5	0.25	30	0.125	3.75	26.25
0.5–1.0	0.75	50	0.375	18.75	31.25
1.0–1.5	1.25	90	0.625	56.25	33.75
1.5–2.0	1.75	110	0.875	96.25	13.75
2.0–2.5	2.25	100	1.000	100.0	0.0
2.5–3.0	2.75	70	1.000	70.0	0.0
3.0–3.5	3.25	30	1.000	30.0	0.0
3.5–4.0	3.75	20	1.000	20.0	0.0
Total		500		395.00	105.00

2. Calculate the average particle settling velocity for each velocity range by taking the average of the range limits, and enter the values in column (2). For the first velocity range, the average settling velocity is (0.0 + 0.5)/2 = 0.25 m/h.
3. Enter the number of influent particles for each velocity range in column (3).
4. Calculate the removal fraction for each velocity range by dividing the average settling velocity by the critical overflow velocity (2.0 m/h), and enter the result in column (4). For the first velocity range

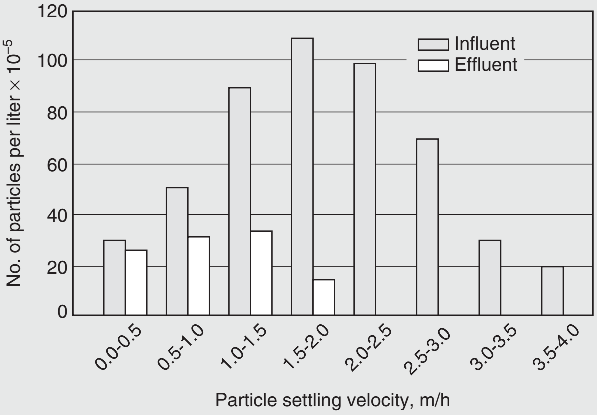
$$\text{Fraction removed} = \frac{v_{n_i}}{v_c} = \frac{0.25}{2.0} = 0.125$$

Where the result is greater than 1.0, enter a value of 1.0, because all of the particles are removed.

5. Determine the number of particles removed by multiplying the number of influent particles by the percent removal [column (3) × column (4)]. Enter the values in column (5).
6. Calculate the particles remaining by subtracting the particles removed from the number of influent particles [column (3) − column (5)]. Enter the result in column (6).
7. Compute the removal efficiency by calculating the sum of particles removed and dividing the sum by the total number of particles in the influent.

$$\text{Total fraction removed} = \frac{\sum_{i=1}^{n} \frac{v_{n_i}}{v_c}(n_i)}{\sum_{i=1}^{n} n_i} = \frac{395 \times 10^{-5}}{500 \times 10^{-5}} \times 100\% = 79\%$$

8. Plot the particle histogram for the influent and effluent wastewater.



Flocculent Particle Settling

Particles in relatively dilute solutions will not act as discrete particles but will coalesce during sedimentation. As coalescence or flocculation occurs, the mass of the particle increases, and it settles faster. The extent to which flocculation occurs is dependent on the opportunity for contact, which varies with overflow rate, depth of the basin, velocity gradients in the system, concentration of particles, and range of particle sizes. The effects of these variables can be determined only by sedimentation tests.

The settling characteristics of a suspension of flocculent particles can be obtained by using a settling column test. Such a column can be of any diameter but should be equal in height to the depth of the proposed tank. The solution containing the suspended matter should be introduced into the column in such a way that a uniform distribution of particle sizes occurs from top to bottom. Care should be taken to ensure that a uniform temperature is maintained throughout the test to eliminate convection currents. Settling should take place under quiescent conditions. The duration of the test should be equivalent to the settling time in the proposed tank.

At the conclusion of the settling time, the settled matter that has accumulated at the bottom of the column is drawn off, the remaining liquid is mixed, and the TSS of the liquid is measured. The TSS of the liquid is then compared to the sample TSS before settling to obtain the percent removal.

The more traditional method of determining settling characteristics of a suspension is to use a column similar to the one described above but with sampling ports inserted at approximately 0.5 m (1.5 ft) intervals. At various time intervals, samples are withdrawn from the ports and analyzed for suspended solids. The percent removal is computed for each sample analyzed and is plotted as a number against time and depth, as elevations are plotted on a survey grid. Curves of equal percent removal are drawn as shown on Fig. 5–23. From the curves shown on Fig. 5–23, the overflow rate for various settling is determined by noting the value where the curve intersects the x axis. The settling velocity v_c is

$$v_c = \frac{H}{t_c} \tag{5–31}$$

where H = height of the settling column, L (m)
t_c = time required for a given degree of removal to be achieved, T (min)

The fraction of particles removed is given by

$$R, \% = \sum_{h=1}^{n} \left(\frac{\Delta h_n}{H}\right)\left(\frac{R_n + R_{n+1}}{2}\right) \tag{5–32}$$

where R = TSS removal, %
n = number of equal percent removal curve
h_n = distance between curves of equal percent removal, L (m)
H = total height of settling column, L (m)
R_n = equal percent removal curve number n
R_{n+1} = equal percent removal curve number $n + 1$

The advantage of the more traditional method is that it is possible to obtain removal data at various depths of settling. The removal percentage obtained using the curve given on Fig. 5–23 is illustrated in Example 5–6.

EXAMPLE 5–6 **Removal of Flocculent Suspended Solids** Using the results of the settling test shown on Fig. 5–23, determine the overall removal of solids if the detention time is t_2 and the depth is h_5. Also demonstrate that the same result is obtained when the solids are measured after settling has occurred.

Solution

1. Determine the percent removal.

   Percent removal

$$= \frac{\Delta h_1}{h_5} \times \frac{R_1 + R_2}{2} + \frac{\Delta h_2}{h_5} \times \frac{R_2 + R_3}{2} + \frac{\Delta h_3}{h_5} \times \frac{R_3 + R_4}{2} + \frac{\Delta h_4}{h_5} \times \frac{R_4 + R_5}{2}$$

2. For the curves shown on Fig. 5–23, a total removal for quiescent settling is 65.7 percent. The computations follow.

$\dfrac{\Delta h_n}{h_5} \times \dfrac{R_n + R_{n+1}}{2}$	Percent removal
$0.20 \times \dfrac{100 + 80}{2} =$	18.00
$0.11 \times \dfrac{80 + 70}{2} =$	8.25
$0.15 \times \dfrac{70 + 60}{2} =$	9.75
$0.54 \times \dfrac{60 + 50}{2} =$	29.70
1.00	65.70

3. Determine the percent removal if the liquid had been mixed and the solids were measured.
   a. Assume the initial solids concentration is equal to 100 and that at the end of the settling period the concentration of the solids at the top of the column is equal to zero.
   b. Set up a computation table and determine the remaining solids after settling.

$\Delta h \times \dfrac{\text{TSS}_n + \text{TSS}_{n+1}}{2}$	Average TSS
$0.20 \times \dfrac{0 + 20}{2} =$	2.00
$0.11 \times \dfrac{20 + 20}{2} =$	2.75
$0.15 \times \dfrac{30 + 40}{2} =$	5.25
$0.54 \times \dfrac{40 + 50}{2} =$	24.30
	34.30

The percent removal is $R_t = 100 - 34.30 = 65.70$.

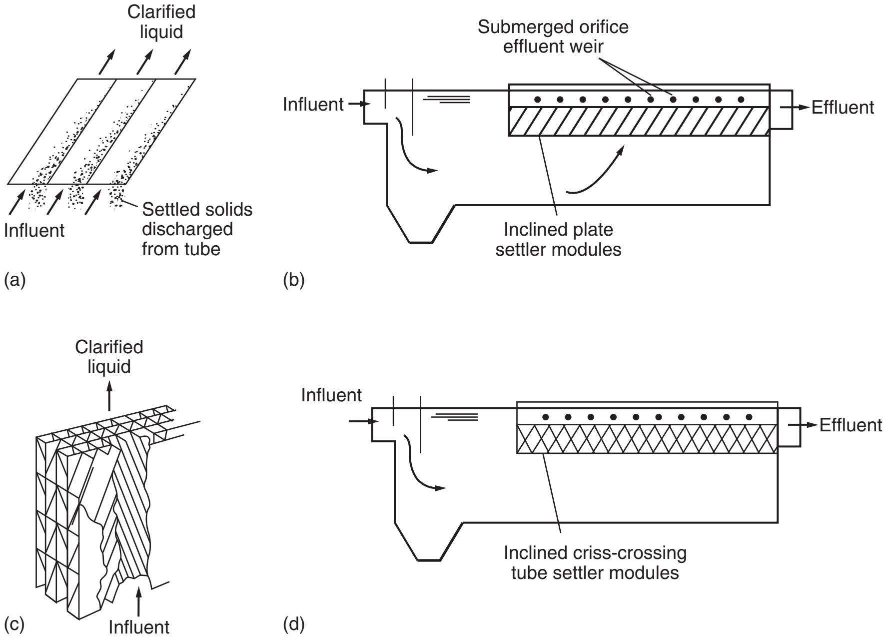
Comment To account for the less than optimum conditions encountered in the field, the design settling velocity or overflow rate obtained from column studies often is multiplied by a factor of 0.65 to 0.85, and the detention times are multiplied by a factor of 1.25 to 1.5.

Inclined Plate and Tube Settling

Inclined plate and tube settlers are shallow settling devices consisting of stacked offset trays or bundles of small plastic tubes of various geometries [see Figs. 5–24(a) and (c)] that are used to enhance the settling characteristics of sedimentation basins [see Figs. 5–24(b) and (d)]. Application of these devices are based on the theory that settling depends on the settling area rather than detention time. Although they are used predominantly in water-treatment applications, plate and tube settlers are used in wastewater treatment for primary, secondary, and tertiary sedimentation. In primary sedimentation applications, however, fine screening should be provided ahead of the settling operation to prevent plugging of the plates or tubes.

Figure 5–24

Plate and tube settlers: (a) module of inclined plates, (b) plates installed in a rectangular sedimentation tank, (c) module of criss-crossing tubes, and (d) criss-crossing tubes installed in a rectangular sedimentation tank.



To be self-cleaning, plate or tube settlers are usually set at an angle between 45 and 60° above the horizontal. When the angle is increased above 60°, the efficiency decreases. If the plates and tubes are inclined at angles less than 45°, solids will tend to accumulate within the plates or tubes. Nominal spacing between plates is 50 mm (2 in.), with an inclined length of 1 to 2 m (3 to 6 ft). To control biological growths and the production of odors (the principal problems encountered with their use), the accumulated solids must be flushed out periodically (usually with high-pressure water). The need for flushing poses a problem with the use of plate and tube settlers when the characteristics of the solids to be removed vary from day to day.

The main objective in inclined settler development has been to obtain settling efficiencies close to theoretical limits. Attention must be given to providing equal flow distribution to each settler, producing good flow distribution within each settler, and collecting settled solids while preventing resuspension. Inclined settling systems are generally constructed for use in one of three ways with respect to the direction of liquid flow relative to the direction of particle settlement: (1) countercurrent, (2) cocurrent, and (3) cross-flow. These three flow patterns are shown schematically on Fig. 5–25 along with definition sketches for their analysis.

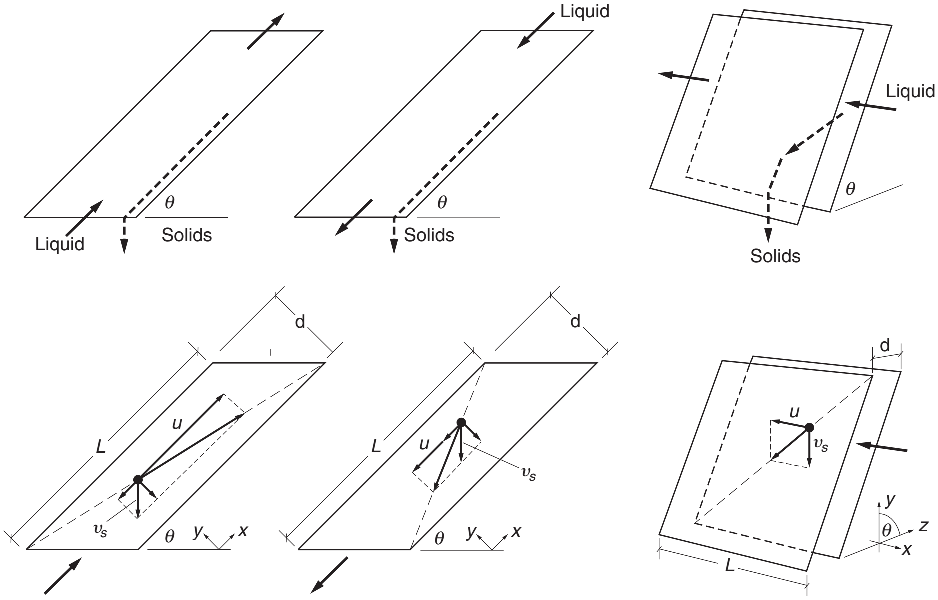
Countercurrent Settling. With countercurrent flow, wastewater suspension in the basin passes upward through the plate or tube modules and exits from the basin above the modules [see Fig. 5–25(a)]. The solids that settle out within the plates or tubes move by gravity countercurrently downward and out of the modules to the basin bottom. Tube settlers are mostly used in the countercurrent mode.

In countercurrent settling, the time, t, for a particle to settle the vertical distance between two parallel inclined surfaces is:

$$t = \frac{d}{v_s \cos\theta} \tag{5–33}$$

Figure 5–25

Alternative flow patterns and corresponding definition sketches for the analysis of plate and tube settlers: (a) countercurrent flow with respect to the movement of solids, (b) cocurrent flow with respect to the movement of solids, and (c) cross-flow.



(a) Countercurrent (b) Cocurrent (c) Cross-flow

where d = perpendicular distance between surfaces, L (m)
v_s = settling velocity, LT^{-1} (m/s)
θ = angle of the surface inclination from the horizontal

The length of surface, L, needed to provide this time, if the fluid velocity between the surfaces is u, is calculated as:

$$L = \frac{d(u - v_s \sin\theta)}{v_s \cos\theta} \tag{5–34}$$

By rearranging this equation, all particles with a settling velocity v_s and greater are removed if

$$v_s \geq \frac{u \cdot d}{L\cos\theta + d\sin\theta} \tag{5–35}$$

When many plates or tubes are used

$$u = \frac{Q}{Ndw} \tag{5–36}$$

where u = fluid velocity, LT^{-1} (m/s)
Q = flowrate, L^3T^{-1} (m^3/s)
N = number of channels made by N+1 plates
d = perpendicular distance between surfaces, L (m)
w = width of the channel, L (m)

A proprietary settler, the Lamella® Gravity Settler, manufactured by the Parkson Corporation, is based on countercurrent settling with modifications (see Fig. 5–26). The feed stream is introduced into the settler by means of a feed duct to the feed box, which is a bottomless channel between plate sections. The flow is directed downward toward individual side-entry plate slots. The feed is distributed across the width of the plates and flows upward under laminar flow conditions. The plates are inclined 55° from the horizontal. The solids settle on the plates and the clean supernatant exits the plates through orifice holes. The orifice holes are placed immediately above each plate and are sized to induce a

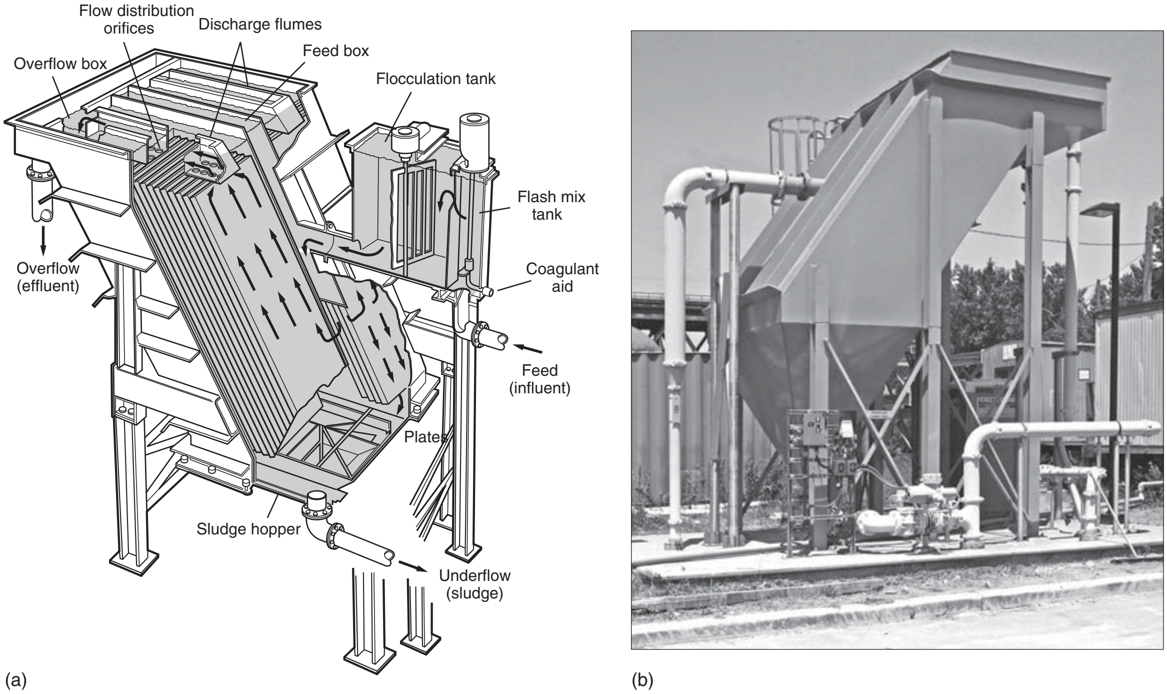


Figure 5–26

Example of a Lamella plate settler: (a) schematic (courtesy of Parkson Corporation) and (b) typical unit for pilot testing.

calculated pressure drop to ensure the feed is hydraulically distributed equally among the plates. The solids slide down the plates into a collection hopper. Further thickening of the solids occurs in the hopper due to compression in the quiescent zone made possible by feeding the plates from the side rather than from the bottom. Plate packs can also be retrofitted into existing clarifiers to improve performance.

Cocurrent Settling. In cocurrent settling, the solids suspension is introduced above the inclined surfaces and the flow is down through the tubes or plates [see Fig. 5–25(b)]. The time for a particle to settle the vertical distance between two surfaces is the same as for countercurrent settling. The length of surface needed, L_p, however, has to be based on downward and not upward liquid flow, as follows:

$$L = d\frac{(u - v_s \sin\theta)}{v_s \cos\theta} \tag{5–37}$$

Consequently, the condition for removal of particles is given by

$$v_s \geq \frac{u \cdot d}{L\cos\theta - d\sin\theta} \tag{5–38}$$

Cross-Flow Settling. In cross-flow settling, the liquid flow is horizontal and does not interact with the vertical settling velocity [see Fig. 5–25(c)]. The length of the surface L_p is determined by

$$L = \frac{u \cdot d}{v_s \cos\theta} \tag{5–39}$$

and

$$v_s \geq \frac{u \cdot d}{L \cos \theta} \tag{5–40}$$

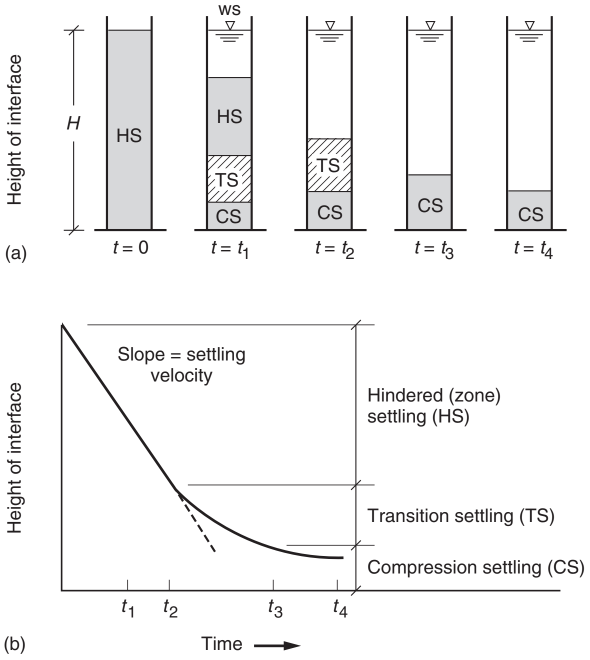
Hindered (Zone) Settling

In addition to discrete (free) and flocculent settling discussed previously, other forms of settling known as hindered or zone and compression settling usually occur in systems that contain a high concentration of suspended solids. The settling phenomenon that occurs when a concentrated suspension, initially of uniform concentration throughout, is placed in a graduated cylinder, is illustrated on Fig. 5–27. Because of the high concentration of particles, the liquid tends to move up through the interstices of the contacting particles. As a result, the contacting particles tend to settle as a zone or "blanket," maintaining the same relative position with respect to each other. The phenomenon is known as hindered settling. As the particles settle, a relatively clear layer of water is produced above the particles in the settling region. The scattered, relatively light particles remaining usually settle as discrete or flocculent particles, as discussed previously. In most cases, an identifiable interface develops between the upper region and the hindered settling region as shown on Fig. 5–27(a). The rate of settling in the hindered settling region is a function of the concentration of solids and their characteristics.

As settling continues, a compressed layer of particles begins to form on the bottom of the cylinder in the compression settling region. The particles apparently form a structure in which there is close physical contact between the particles. As the compression layer forms, regions containing successively lower concentrations of solids than those in the compression region extend upward in the cylinder. Thus, in actuality the hindered settling

Figure 5–27

Definition sketch for hindered (zone) settling: (a) settling column and (b) settling curve.



region contains a gradation in solids concentration from that found at the interface of the settling region to that found in the compression settling region.

Because of the variability encountered, settling tests are usually required to determine the settling characteristics of suspensions where hindered and compression settling are important considerations. On the basis of data derived from column settling tests, two different design approaches can be used to obtain the required area for the settling/thickening facilities. In the first approach, the data derived from one or more batch settling tests are used. In the second approach, known as the solids flux method, data from a series of settling tests conducted at different solids concentrations are used. Both methods are described in the following discussion. The solids flux method is considered further in Sec. 8–10 in Chap. 8. It should be noted that both methods have been used where existing plants are to be expanded or modified. These methods are, however, seldom used in the design of small treatment plants.

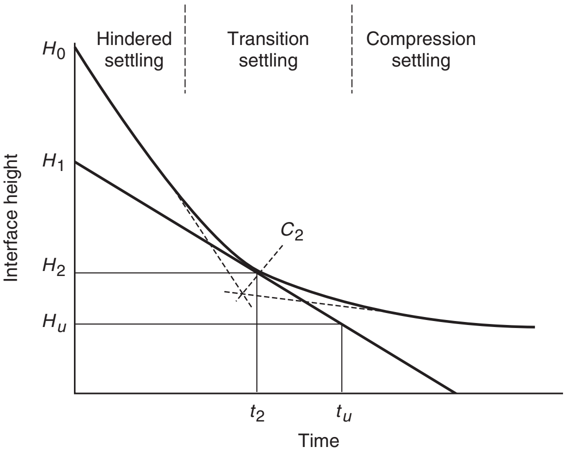
Area Requirement Based on Single-Batch Test Results. For purposes of design, the final overflow rate selected should be based on a consideration of the following factors: (1) the area needed for clarification, (2) the area needed for thickening, and (3) the rate of sludge withdrawal. Column settling tests, as previously described, can be used to determine the area needed for the free settling region directly. However, because the area required for thickening is usually greater than the area required for the settling, the rate of free settling rarely is the controlling factor. In the case of the activated-sludge process where stray, light, fluffy floc particles may be present, it is conceivable that the free flocculent settling velocity of these particles could control the design.

The area requirement for thickening is determined according to a method developed by Talmadge and Fitch (1955). A column of height H_o is filled with a suspension of solids of uniform concentration C_o. The position of the interface as time elapses and the suspension settles is given on Fig. 5–28. The rate at which the interface subsides is then equal to the slope of the curve at that point in time. According to the procedure, the area required for thickening is given by Eq. (5–41):

$$A = \frac{Qt_u}{H_o} \tag{5–41}$$

Figure 5–28

Graphical analysis of hindered (zone) interface settling curves.



where A = area required for sludge thickening, L^2 (m^2)
Q = flowrate into tank, L^3T^{-1} (m^3/s)
H_o = initial height of interface in column, L (m)
t_u = time to reach desired underflow concentration, T (s)

The critical concentration controlling the solids handling capability of the tank occurs at a height H_2 where the concentration is C_2. This point is determined by extending the tangents to the hindered settling and compression regions of the subsidence curve to the point of intersection and bisecting the angle thus formed, as shown on Fig. 5–28. The time t_u can be determined as follows:

1. Construct a horizontal line at the depth H_u that corresponds to the depth at which the solids are at the desired underflow concentration C_u. The value of H_u is determined using the following expression:

$$H_u = \frac{C_o H_o}{C_u} \tag{5–42}$$

2. Construct a tangent to the settling curve at the point indicated by C_2.
3. Construct a vertical line from the point of intersection of the two lines drawn in steps 1 and 2 to the time axis to determine the value of t_u.

With this value of t_u, the area required for the thickening is computed using Eq. (5–41). The area required for clarification is then determined. The larger of the two areas is the controlling value. Application of this procedure is illustrated in Example 5–7.

EXAMPLE 5–7 **Sizing an Activated-Sludge Settling Tank** The settling curve shown in the following diagram was obtained for an activated sludge with an initial solids concentration C_o of 3000 mg/L. The initial height of the interface in the settling column was at 0.75 m (2.5 ft). Determine the area required to yield a thickened solids concentration, C_u, of 12,000 mg/L with a total flow of 3800 m^3/d (1 Mgal/d). Determine also the solids loading rate (kg/m^2·d) and the overflow rate (m^3/m^2·d).

Solution

1. Determine the area required for thickening using Eq. (5–42).
   a. Determine the value of H_u

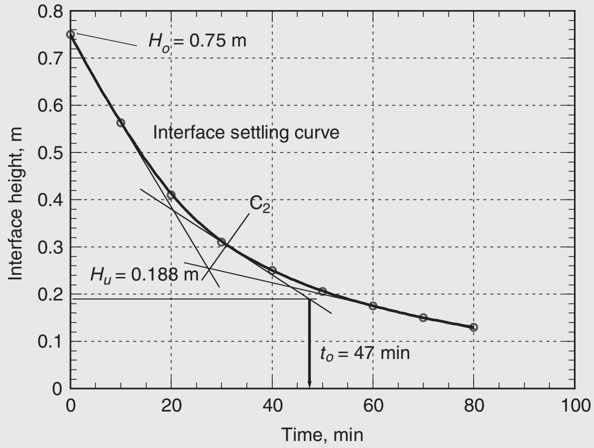
$$H = \frac{C_o H_o}{C_u}$$

$$= \frac{(3000 \text{ mg/L})(0.75 \text{ m})}{(12{,}000 \text{ mg/L})} = 0.188 \text{ m}$$

On the following settling curve, a horizontal line is constructed at H_u = 0.188 m. A tangent is constructed to the settling curve at C_2, the midpoint of the region between hindered and compression settling. Bisecting the angle formed where the two tangents meet determines point C_2. The intersection of the tangent at C_2 and the line H_u = 0.188 m determines t_u. Thus t_u = 47 min, and the required area is

$$A = \frac{Qt_u}{H_o} = \left[\frac{(3800 \text{ m}^3\text{/d})}{(24 \text{ h/d})(60 \text{ min/h})}\right]\left(\frac{47 \text{ min}}{0.75 \text{ m}}\right) = 165 \text{ m}^2$$

2. Determine the area required for clarification.



   a. Determine the interface subsidence velocity *v*. The subsidence velocity is determined by computing the slope of the tangent drawn from the initial portion of the interface settling curve. The computed velocity represents the unhindered settling rate of the sludge.

$$v = \left(\frac{0.75\text{ m} - 0.3\text{ m}}{29.5\text{ min}}\right)\left(\frac{60\text{ min}}{\text{h}}\right) = 0.92\text{ m/h}$$

   b. Determine the clarification rate. Because the clarification rate is proportional to the liquid volume above the critical sludge zone, it may be computed as follows:

$$Q = 3800\text{ m}^3\text{/d}\left(\frac{0.75\text{ m} - 0.188\text{ m}}{0.75\text{ m}}\right) = 2847\text{ m}^3\text{/d}$$

   c. Determine the area required for clarification. The required area is obtained by dividing the clarification rate by the settling velocity.

$$A = \frac{Q_c}{v} = \frac{(2847\text{ m}^3\text{/d})}{(24\text{ h/d})(0.92\text{ m/h})} = 129\text{ m}^2$$

3. The controlling area is the thickening area (165 m^2) because it exceeds the area required for clarification (129 m^2).
4. Determine the solids loading rate. The solids loading rate is computed as follows:

$$\text{Daily solids mass, kg/d} = \frac{(3800\text{ m}^3\text{/d})(3000\text{ g/m}^3)}{(10^3\text{ g/1 kg})} = 11{,}400\text{ kg/d}$$

$$\text{Solids loading rate} = \frac{(11{,}400\text{ kg/d})}{165\text{ m}^2} = 69.1\text{ kg/m}^2\text{·d}$$

5. Determine the hydraulic loading rate.

$$\text{Hydraulic loading rate} = \frac{(3800\text{ m}^3\text{/d})}{165\text{ m}^2} = 23.0\text{ m}^3\text{/m}^2\cdot\text{d}$$

Comment An alternative approach for sizing the secondary clarifiers using the initial settling velocity of the sludge is given in Sec. 8–10 in Chap. 8.

Area Requirements Based on Solids Flux Analysis. An alternative method of determining the area required for hindered settling is based on an analysis of the solids (mass) flux (Coe and Clevenger, 1916). In the solids flux method of analysis it is assumed that a settling basin is operating at steady state. Within the tank, the downward flux of solids is brought about by gravity (hindered) settling and by bulk transport due to the underflow that is being pumped out and recycled. The solids flux method of analysis is used to assess the performance of existing facilities and to obtain information for the design of new facilities to treat the same wastewater. Application of the solids flux method of analysis is illustrated in Sec. 8–7 in Chap. 8, and additional information may be found in the following references: Dick and Ewing (1967), Dick and Young (1972), Keinath (1985), Wahlberg and Keinath (1988), and Yoshika et al. (1957).

Compression Settling

The volume required for sludge in the compression region can also be determined by settling tests. The rate of consolidation has been found to be proportional to the difference in the depth at time *t* and the depth to which the sludge will settle after a long period of time. The long-term consolidation can be modeled as a first-order decay function, as given by Eq. (5–43).

$$H_t - H_\infty = (H_2 - H_\infty)e^{-i(t-t_2)} \tag{5–43}$$

where H_t = sludge height at time t, L
H_∞ = sludge depth after long settling period, on the order of 24 h, L
H_2 = sludge height at time t_2, L
i = constant for a given suspension

Stirring serves to compact solids in the compression region by breaking up the floc and permitting water to escape. Rakes are often used on sedimentation equipment to manipulate the solids and thus produce better compaction.

Gravity Separation in an Accelerated Flow Field

Sedimentation, as described previously, occurs under the force of gravity in a constant acceleration field. The removal of settleable particles can also be accomplished by taking advantage of a changing acceleration field. A number of devices that take advantage of both gravitational and centrifugal forces and induced velocities have been developed for the removal of grit from wastewater. The principles involved are illustrated on Fig. 5–29. In appearance, the separator looks like a large diameter cylinder with a conical bottom. Wastewater, from which grit is to be separated, is introduced tangentially near the top and exits through the opening in the top of the unit. The liquid is removed at the top. Grit is removed through an opening in the bottom of the unit.

Because the top of the separator is enclosed, the rotating flow creates a free vortex within the separator. The most important characteristic of a free vortex is that the product of the tangential velocity times the radius is a constant:

$$Vr = \text{constant} \tag{5–44}$$

where V = tangential velocity, LT^{-1} (m/s)
r = radius, L (m)

The significance of Eq. (5–44) can be illustrated by the following example. Assume the tangential velocity in a separator with a 1.5 m (5 ft) radius is 0.9 m/s (3 ft/s). The product of the velocity times the radius at the outer edge of the separator is equal to 1.35 m^2/s (15 ft^2/s). If the discharge port has a radius of 0.9 m (3 ft), then the tangential velocity at

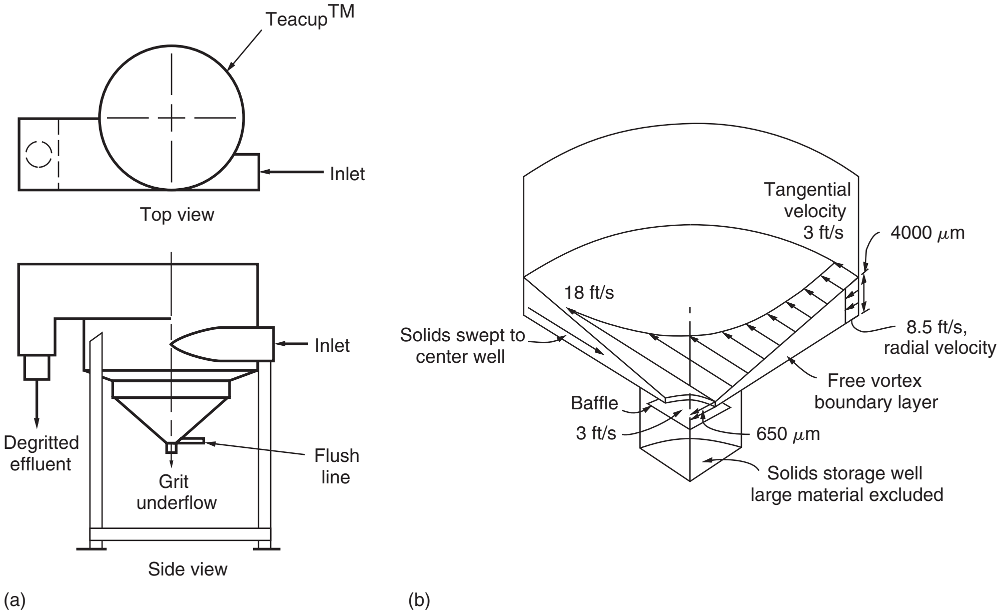


Figure 5–29

Accelerated gravity separator: (a) outline sketch and (b) definition sketch. (From Hydro International.)

the entrance to the discharge port is 4.5 m/s (15 ft/s). The centrifugal force experienced by a particle within this flow field is equal to the square of the velocity divided by the radius. Because the centrifugal force is also proportional to the inverse of the radius, a fivefold decrease in the radius results in a 125-fold increase in the centrifugal force.

Because of the high centrifugal forces near the discharge port, some of the particles, depending on their size, density, and drag, are retained within the body of the free vortex near the center of the separator, while other particles are swept out of the unit. Grit and sand particles will be retained while organic particles are discharged from the unit. Organic particles having the same settling velocity as sand will typically be from 4 to 8 times as large. The corresponding drag forces for these organic particles will be from 16 to 64 times as great. As a result, the organic particles tend to move with the fluid and are transported out of the separator. The particles held in the free vortex ultimately settle to the bottom of the unit under the force of gravity. Organic particles that sometimes settle usually consist of oil and grease attached to grit or sand particles.

5–5 GRIT REMOVAL

Grit in wastewater consists of sand, gravel, cinders, or other heavy solid materials that have subsiding velocities or specific gravities substantially greater than those of the organic putrescible solids in wastewater. Grit removal is necessary to (1) reduce formation of heavy deposits in aeration tanks, aerobic digesters, pipelines, channels, and conduits; (2) reduce the frequency of digester cleaning caused by excessive accumulations of grit; and (3) protect moving mechanical equipment from abrasion and accompanying abnormal wear. The removal of grit is essential for equipment with closely machined metal surfaces, such as comminutors, fine screens, centrifuges, heat exchangers, and high-pressure diaphragm pumps.

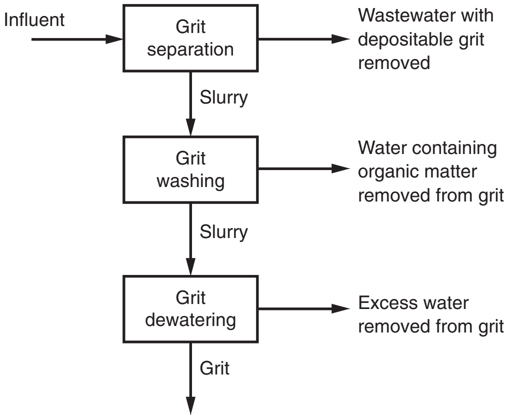


Figure 5–30

A complete grit removal system, including grit separation, washing, and dewatering. (Adapted from Wilson et al., 2007.)

The overall objectives of the grit removal system are to remove all depositable grit during normal and peak wet-weather flow and to produce an end product suitable for landfill disposal. As illustrated on Fig. 5–30, a complete grit removal system consists of three distinct unit processes each with a specific objective: (1) grit separation, (2) grit washing, and (3) grit dewatering. Before discussing these three distinct unit processes, it is appropriate to consider the characteristics of wastewater grit, as they will affect significantly the selection and design of the grit removal system. Grit separators for combined collection systems and stormwater are also reviewed along with degritting of primary sludge where grit removal is not used before primary sedimentation.

Wastewater Grit Characteristics

Traditionally, grit removal systems have been designed based on the assumption that grit consists of inorganic settleable solids ranging in size from 0.050 to 1.0 mm with settling characteristics similar to clean, spherical silica sand with a specific gravity of 2.65 and a particle size predominantly larger than 0.210 mm. As a consequence of these assumptions, many existing grit removal systems fall short of performance expectations resulting in excessive maintenance and operating costs.

Grit Composition. Grit consists of sand, gravel, cinders, and other heavy materials. It also includes organic matter such as eggshells, bone chips, seeds, and coffee grounds. As grit travels through the collection system, grit particles come in contact with organic matter and surface active agents (SAAs) that can adhere to the grit particles. The presence of these constituents can alter the size and settling characteristics of the grit particles.

Generally, what is removed from the wastewater flow as grit is predominantly inert and relatively dry material. However, the composition can be highly variable, with moisture content ranging from 13 to 65 percent, and volatile content from 1 to 56 percent with a wide range of specific gravity. Often, enough organics are present in the grit so that it quickly putrefies if not properly handled after removal from the wastewater.

Grit Particle Size. Based on the limited available information, the actual size distribution of collected grit exhibits wide variation due to differences in collection system characteristics, as well as variations in grit-removal efficiency (see Fig. 5–31). Generally, most collected grit particles are retained on a 0.15 mm (100-mesh) sieve, reaching nearly 100 percent retention in some instances; however, particle size can vary significantly. In urban, high-density environments, wastewater grit tends to be coarser. In coastal, lower density environments, grit tends to be finer. In the southeastern United States, where fine

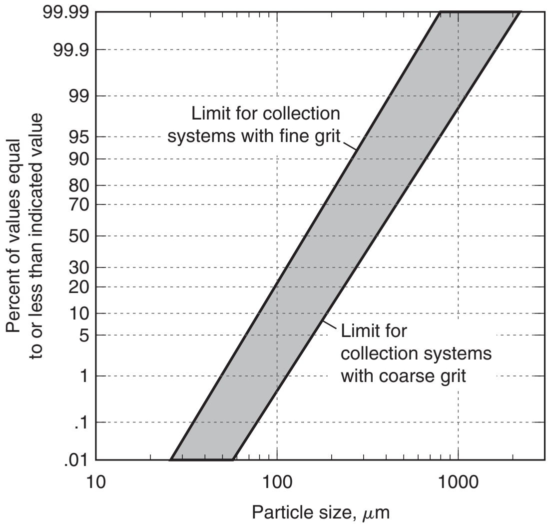


Figure 5–31

Typical grit particle size distribution range. Shaded region corresponds to the range of grit particle sizes measured at various wastewater treatment plants. (Adapted from Wilson et al., 2007.)

sand known as "sugar sand" constitutes a portion of the grit, less than 60 percent of the grit was retained on a 0.15 mm (100-mesh) screen in some cases.

Grit Quantities. The quantities of grit will vary greatly from one location to another, depending on the type of sewer system, the characteristics of the drainage area, the condition of the collection system, the frequency of street sanding to counteract icing conditions, the types of industrial wastes, the number of households with food waste grinders, and amount of infiltration in areas with sandy soils. Sand drift from beach areas in coastal cities is also a major contributor. Another factor in combined sewer systems is the occurrence of a first flush associated with rainfall events. The higher flowrates associated with wet weather events resuspend grit that settled in the collection system during dry weather. Thus, heavier grit loads are experienced in wastewater treatment plants soon after the start of a heavy rainfall.

Grit removal data are difficult to interpret because grit is poorly characterized and almost no data exists on relative removal efficiencies. The information on grit characteristics derives from what has been collected rather than the grit in the influent wastewater. Sieve analyses are not normally performed on grit chamber influents and effluents. For these reasons, the efficiencies of grit removal systems cannot be compared. A comparison of the reported quantities of grit removed from separate and combined sewer systems is presented in Table 5–15.

Grit Settling Characteristics. The settling characteristics of grit vary considerably from the point of entry in the collection system and its location in the treatment plant. Grit moves through gravity sewers within the collection system in three more-or-less distinct layers (see Fig. 5–32). The deposit limit for clean, inorganic grit particles for most collection systems is 0.225 mm. Consequently, particles above this size cannot be transported hydraulically under normal flow conditions and will deposit within the collection system. As SAAs in the wastewater pass over the deposited particles, some adhere to the deposited grit particles. When a sufficient amount of SAAs have accumulated, the coated grit particle buoyancy increases and deposited grit rises into the bed load. Here, the coated grit moves slowly to the treatment plant along the base of the collection system at less than the average wastewater velocity. Above the bed load is the suspended load consisting of the

Table 5–15

Comparison of quantities of grit removed from wastewater from separate and combined collection systems in aerated grit chambers

Type of collection system	Ratio of maximum day to average day	Average grit quantity	
		ft^3/Mgal	m^3/1000 m^3
Separate	1.5 to 3:1	0.5–5	0.004–0.037
Combined	3 to 15:1	0.5–27	0.004–0.20

lighter grit particles. This suspended grit load has a settling velocity of less than that of clean inorganic grit particles that are less than 0.225 mm in diameter and is the primary grit load that reaches the treatment plant under normal (day-to-day) conditions. Given that, traditionally, grit removal systems have been designed to remove clean inorganic grit particles that are greater than 0.210 mm, most grit passes through the grit removal flows under normal conditions. During periods of high flow, particularly in combined sewers, the heavier, deposited grit is re-suspended and the quantity of grit reaching the treatment plant increases substantially. It is, therefore, important that the grit removal system not only operate efficiently during normal flow conditions but also under sustained peak flows when the greatest volume of grit reaches the plant.

Because of SAAs, grit particles reaching the treatment plant under normal conditions are often lighter than anticipated during design. The impact of SAAs on particles size is demonstrated graphically on Fig. 5–33. Assuming that the specific gravity of clean sand is constant, the settling velocity of clean sand increases as particle size increases [see Fig. 5–33(a)]. However, the settling velocity of wastewater grit is independent of particle size because of the buoyant effect of SAAs [see Fig. 5–33(b)]. Consequently, wastewater grit passes through traditionally designed grit removal processes to the primary settling tanks where they settle out with the primary sludge or are passed to the aeration tanks. When exposed to biological activity, the SAAs decompose and the remaining high density grit particles settle rapidly. As a result, there is often excessive grit deposition in biological reactors, such as aeration tanks, aerobic digesters, and anaerobic digesters, that leads to frequent and costly cleaning to maintain process efficiency. Typically, two-thirds of grit that enters a WWTP is larger than the deposit limit size of clean sand.

Sand Equivalent Size. Many grit removal systems have been designed based on grit particles having a specific gravity of 2.65 (similar to silica sand) and a particle size predominantly larger than 0.210 mm and settling characteristics similar to clean sand. Yet, grit particles that are 0.210 mm and larger are often cited as the cause of most downstream problems. The sand equivalent size (SES) relates the settling velocity of wastewater grit to that of clean sand as shown on Fig. 5–33(c). As shown, a wastewater grit particle (coated

Figure 5–32

Distribution of suspended, bed load, and deposited grit particles in a collection system. (Adapted from Wilson et al., 2007.)

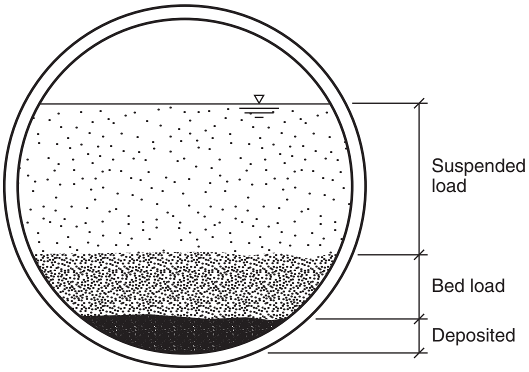
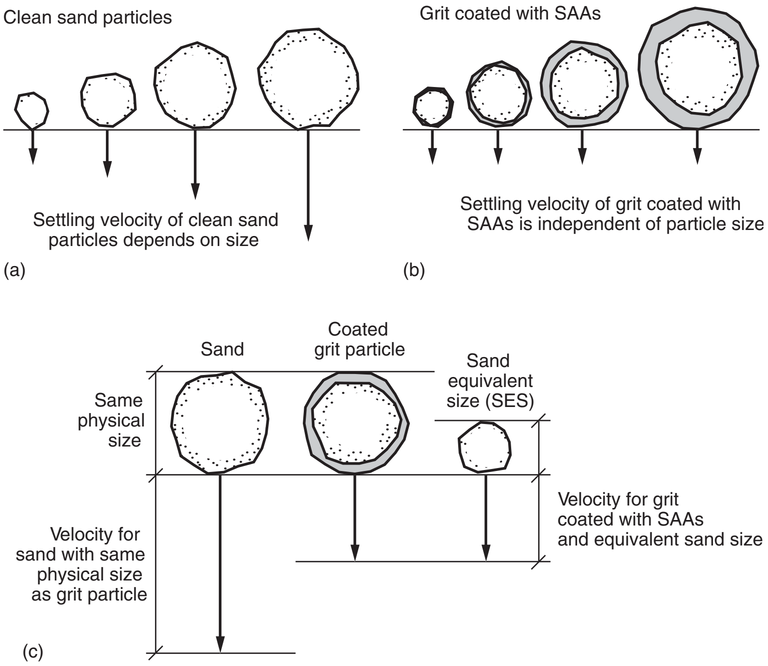


Figure 5–33

Settling velocities of grit particles reaching the treatment plant: (a) clean sand, (b) frothy grit coated with surface active agents, (c) equivalent size of coated grit particles as compared to clean sand particles. (Adapted from Wilson et al., 2007.)



with SAAs) with the same physical size as a clean particle of sand will have a lower settling velocity than the clean sand particle. The SES of a wastewater grit particle is the equivalent size to that of a clean sand particle size that has the same settling velocity.

Settling Velocity of Wastewater Grit. The level of grit that must be removed should be decided on a case-by-case basis. It is dependent on the characteristics of the wastewater grit entering the treatment plant and the level of removal that is required to protect the downstream treatment processes and equipment. While previous practice targeted clean particles of 0.225 mm, modern removal systems will typically target wastewater grit with a much lower SES. High, efficiency grit systems typically target grit with an SES ranging from 0.075 to 0.150 mm.

When possible, a grit study should be conducted to determine the SES of the incoming wastewater grit and the design SES to achieve the desired level of grit removal. If a grit study is not viable, regional data can be used as a substitute. When data are not available, design can be based on a qualitative evaluation of the collection system. The physical size of wastewater grit particles with their associated SES due to the impact of SAAs as measured at various locations are compared on Fig. 5–34(a). The physical size and the SES of wastewater grit begins to diverge at 0.106 mm with the disparity increasing as the physical size increases. For example, a physical particle size of 0.210 mm, the SES could vary from 0.106 to 0.210 mm. Consequently, to remove particles with a physical size of 0.210 mm, an SES as low as 0.106 mm may have to be used as a basis of design.

A design based on an SES of 0.106 mm will result in 90 percent grit removal for most collection systems. However, for those collection systems where fine silt is expected, the grit removal efficiency might be significantly lower (50 to 65 percent). If the goal is to achieve 90 percent grit removal, a design SES as low as 0.075 mm may be required.

Once the design SES has been identified, the surface loading rate (settling velocity) of the targeted grit particle can be established using the settling characteristics of clean sand [see Fig. 5–34(b)]. For an SES of 0.106 mm, the surface loading rate is 0.49 m/min

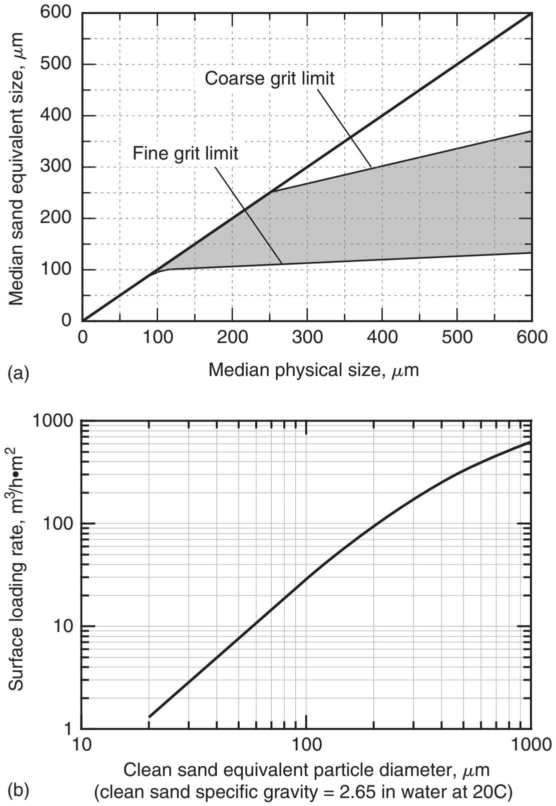


Figure 5–34

Process design information for vortex type grit removal units: (a) comparison of median physical size to SES at various US wastewater treatment plants (adapted from Hydro International) and (b) surface loading rate for sizing vortex grit units based on Eq. (5–18).

(12 gal/ft^2·min). For an SES of 0.075 mm, the required surface loading rate is 0.24 m/min (6 gal/ft^2·min), which doubles the size of the required grit removal system.

Grit Separators for Wastewater

The separation of grit from wastewater is usually accomplished in separate grit chambers designed to physically separate heavy grit particles from lighter organic solids. Grit chambers are most often located after the bar screens and before the primary sedimentation tanks to prevent screening debris from impacting the operation and maintenance of the grit removal equipment. For those treatment plants using comminution equipment, the grit chambers should be located upstream to reduce the wear on cutting blades. There are three general types of grit separation devices: horizontal-flow grit chambers, of either a rectangular or a square configuration, aerated grit chambers, or vortex grit chambers. Each type is considered separately below along with the associated washing and drying equipment that is typically provided. Grit washing drying and disposal are considered after the discussion of the grit separation processes.

Horizontal-Flow Grit Chambers. Rectangular and square horizontal-flow grit chambers have been used for many years. Their use, however, in new installations has been limited in favor of aerated and vortex-type grit chambers.

Rectangular Horizontal-Flow Grit Chambers. The oldest type of grit chamber used is the rectangular horizontal-flow, velocity-controlled type. Representative design data for

Table 5–16

Typical design information for horizontal-flow grit chambers

	U.S. customary units			SI units		
	Unit	Range	Typical	Unit	Range	Typical
Detention time	s	45–90	60	s	45–90	60
Horizontal velocity	ft/s	0.8–1.3	1.0	m/s	0.25–0.4	0.3
Settling velocity for removal of:						
0.21 mm (70-mesh) material	ft/min[a]	3.2–4.2	3.8	m/min[a]	1.0–1.3	1.15
0.149 mm (100-mesh) material	ft/min[a]	2.0–3.0	2.5	m/min[a]	0.6–0.9	0.75
Headloss in a control section as percent of depth in channel	%	30–40	36[b]	%	30–40	36[b]
Added length allowance for inlet and outlet turbulence	%	25–50	30	%	25–50	30

[a] If the specific gravity of the grit is significantly less than 2.65, lower velocities should be used.

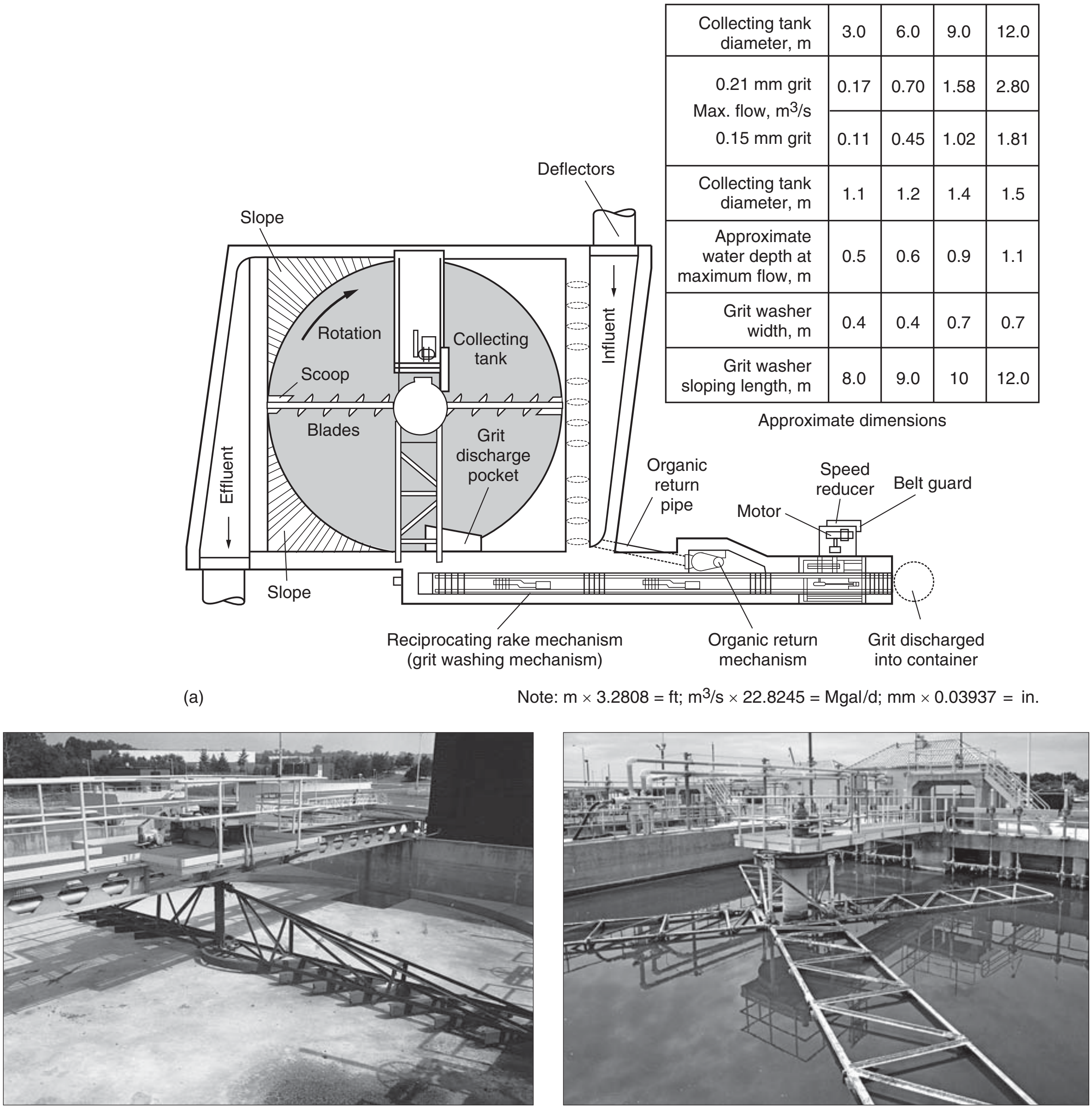
[b] For Parshall flume control.

rectangular horizontal-flow grit chambers are presented in Table 5–16. These units are designed to maintain a velocity as close to 0.3 m/s (1.0 ft/s) as practical and to provide sufficient time for grit particles to settle to the bottom of the chamber. The design velocity will carry most organic particles through the chamber and will tend to re-suspend any organic particles that settle but will permit the heavier grit to settle out.

The basis of design of rectangular horizontal-flow grit chambers is that, under the most adverse conditions, the lightest particle of depositable grit will reach the bed of the channel prior to its outlet end. Typically, grit chambers were designed to remove all grit particles that will be retained on a 0.21-mm-diameter (70-mesh) screen, although many chambers have been designed to remove grit particles retained by a 100-mesh screen (i.e., 0.15-mm-diameter). The settling velocity to be used should be based on the SES of the wastewater grit using the settling rates from Fig. 5–34(b). The length of channel is governed by the depth required by the settling velocity and the control section, and the cross-sectional area will be governed by the rate of flow and by the number of channels. Additional channel length should be added to allow for inlet and outlet turbulence.

Grit removal from horizontal-flow grit chambers is accomplished usually by a conveyor with scrapers, buckets, or plows. Screw conveyors or bucket elevators are used to elevate the removed grit for washing or disposal. In small plants, grit chambers are sometimes cleaned manually.

Square Horizontal-Flow Grit Chambers. Square horizontal-flow grit chambers, such as those shown on Fig. 5–35, have also been in use for over 60 y. Influent to the units is distributed over the cross section of the tank by a series of vanes or gates, and the



Collecting tank diameter, m		3.0	6.0	9.0	12.0
Max. flow, m^3/s	0.21 mm grit	0.17	0.70	1.58	2.80
	0.15 mm grit	0.11	0.45	1.02	1.81
Collecting tank diameter, m		1.1	1.2	1.4	1.5
Approximate water depth at maximum flow, m		0.5	0.6	0.9	1.1
Grit washer width, m		0.4	0.4	0.7	0.7
Grit washer sloping length, m		8.0	9.0	10	12.0

Approximate dimensions

(a)

Note: m × 3.2808 = ft; $m^3/s \times 22.8245$ = Mgal/d; mm × 0.03937 = in.

(b)

(c)

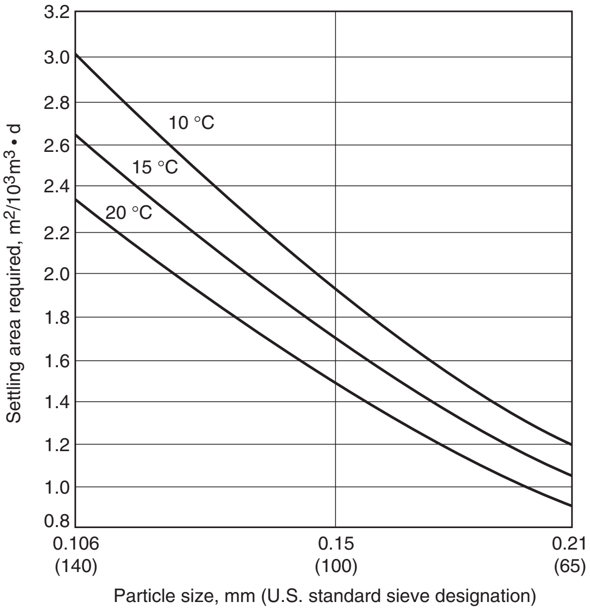
Figure 5–35

Typical square horizontal-flow grit chambers: (a) schematic with design information based on clean grit with a specific gravity of 2.65, (b) view of empty basin. The two rakes are used to move settled grit to the periphery for removal and (c) view of square grit chamber.

distributed wastewater flows in straight lines across the tank and overflows a weir in a free discharge. Where square grit chambers are used, it is generally advisable to use at least two units. These types of grit chambers are designed on the basis of overflow rates that are dependent on particle size and the temperature of the wastewater. Typically, these units have been designed to remove 95 percent of the 0.15-mm-diameter particles at peak flow.

Figure 5–36

Effect of wastewater temperature on the area required for settling grit particles with a specific gravity of 2.65 in a horizontal-flow grit chamber.



A typical set of design curves is shown on Fig. 5–36. The SES should be used for determining the settling area required.

In square grit chambers, the solids are removed by a rotating raking mechanism to a sump at the side of the tank (see Fig. 5–35). Settled grit may be moved up an incline by a reciprocating rake mechanism or an inclined screw conveyor. Grit may also be pumped from the tank through a cyclone degritter to separate the remaining organic material and concentrate the grit. The concentrated grit may then be washed again in a classifier using a submerged reciprocating rake or an inclined-screw conveyor. By either method, organic solids are separated from the grit and flow back into the basin, resulting in a cleaner, dryer grit.

Aerated Grit Chambers. In aerated grit chambers, air is introduced along one side of a rectangular tank to create a spiral flow pattern perpendicular to the flow through the tank (see Fig. 5–37). Wastewater, as shown on Fig. 5–37(b), will move through the tank in a spiral path and will make two to three passes across the bottom of the tank at maximum flow and more passes at average flow. The heavier grit particles settle to the bottom of the tank. Lighter, principally organic particles remain in suspension and pass through the tank. The velocity of roll or agitation governs the size of particles of a given specific gravity that will be removed. If the velocity is too high, grit will be carried out of the chamber; if it is too low, organic material will be removed with the grit. The quantity of air is easily adjusted. With proper adjustment, almost 100 percent of depositable grit will be removed and the grit will be well washed.

Design Considerations. Aerated grit chambers have been designed to remove 0.21-mm-diameter particles or larger, with 2- to 5-min detention periods at the peak hourly rate of flow. The cross section of the tank is similar to that provided for spiral circulation in activated-sludge aeration tanks, except that a grit hopper about 0.9 m (3 ft) deep with steeply sloping sides is located along one side of the tank under the air diffusers (see Fig. 5–37). The air diffusers are located about 0.45 to 0.6 m (1.5 to 2 ft) above the normal plane of the bottom. Influent and effluent baffles are used frequently for hydraulic control and improved grit-removal effectiveness. To determine the required headloss through the

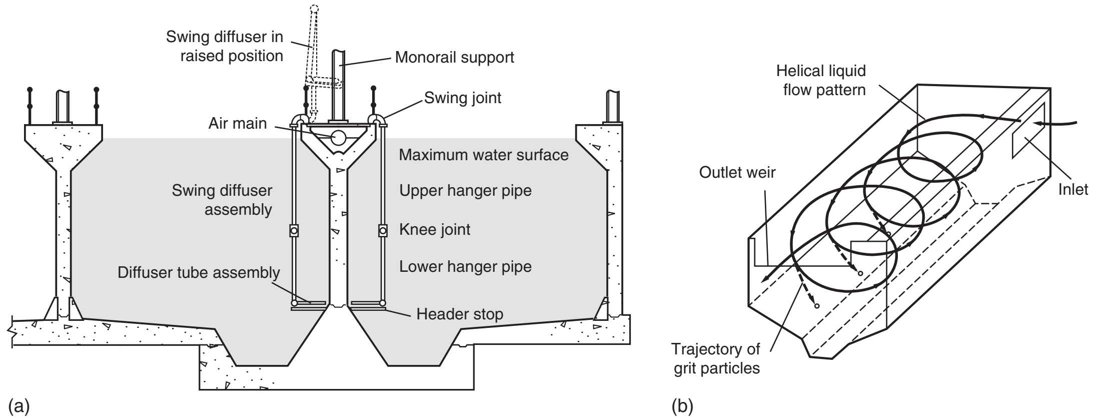


Figure 5–37

Typical aerated grit chamber: (a) cross-section through grit chamber and (b) schematic of helical flow pattern through an aerated grit chamber.

chamber, the expansion in volume caused by the air must be considered. Basic design data for aerated grit chambers are presented in Table 5–17. The design of aerated grit chambers is illustrated in Example 5–8.

Grit Removal Facilities. The removal of grit from aerated grit chambers can be accomplished using grab buckets traveling on monorails and centered over the grit collection and storage trough. Other installations have been equipped with chain-and-bucket conveyors, running the full length of the storage troughs, which move the grit to one end of the trough and elevate it above the wastewater level in a continuous operation. Screw conveyors, tubular conveyors, jet pumps, and airlifts have also been used to remove the collected grit. Grit-removal equipment for aerated grit chambers is subject to the same wear as experienced in the horizontal-flow units.

Table 5–17
Typical design information for aerated grit chambers

	U.S. customary units			SI units		
Item	**Unit**	**Range**	**Typical**	**Unit**	**Range**	**Typical**
Detention time at peak flowrate	min	2–5	3	min	2–5	3
Dimensions:						
Depth	ft	7–16		m	2–5	
Length	ft	25–65		m	7.5–20	
Width	ft	8–23		m	2.5–7	
Width-depth ratio	Ratio	1:1 to 5:1	1.5:1	Ratio	1:1 to 5:1	1.5:1
Length-width ratio	Ratio	3:1 to 5:1	4:1	Ratio	3:1 to 5:1	4:1
Air supply per unit of length	$ft^3/ft \cdot min$	3–8		$m^3/m \cdot min$	0.2–0.5	
Grit quantities[a]	$ft^3/Mgal$	0.5–27	2	$m^3/10^3\ m^3$	0.004–0.20	0.015

[a] From combined collection system.

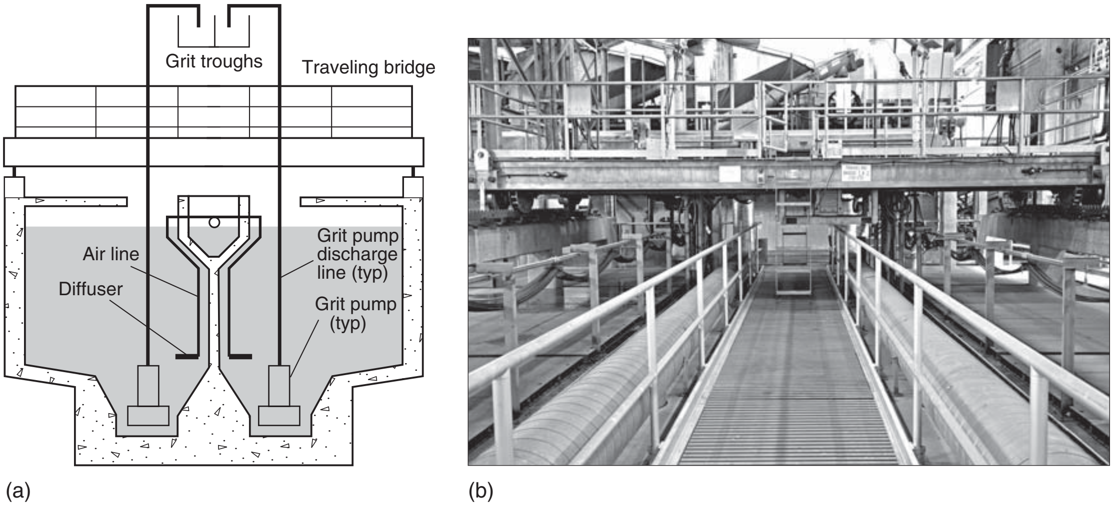


Figure 5–38

Aerated grit chamber with traveling bridge type grit removal system: (a) pumps mounted on the traveling bridge for removal of grit from the grit hopper [diffusers create the helical flow pattern as shown in Fig. 5–37(b)] and (b) view of covered aerated grit chamber equipped with traveling bridge.

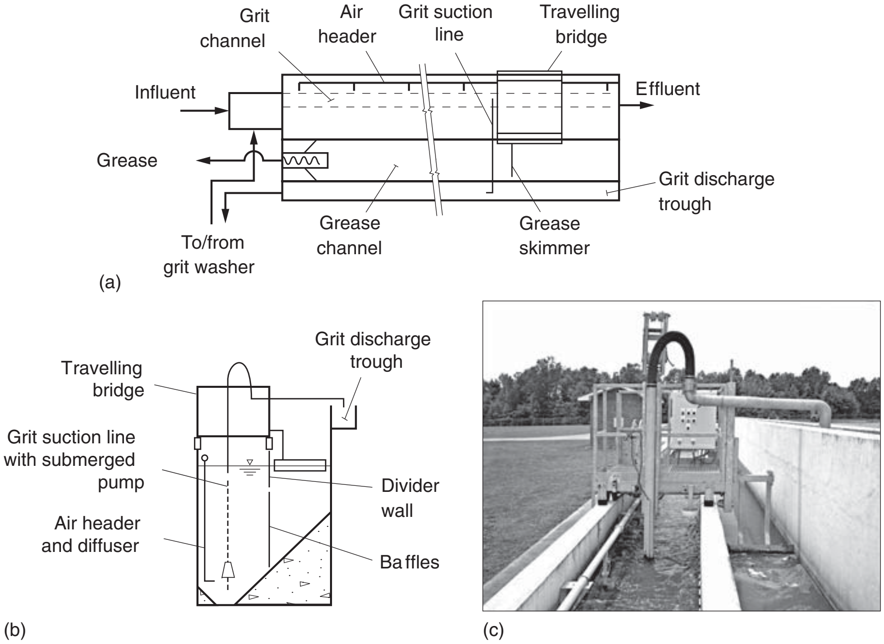
For large installations, traveling-bridge grit collectors, as shown on Fig. 5–38, are being used. Grit pumps are immersed in the grit chambers and travel the entire length, pumping grit into a stationary grit collection trough. The pumps can operate continuously or they can be programmed to run on cycles based on time or flow. A variation of this system uses a grit air lift pump installed on the traveling bridge for each grit tank rather than grit pumps. In another variation, a scraper blade attached to a traveling bridge is used in combination with grit air lifts to push heavier accumulations of grit to a pit at the inlet end of the grit chamber where it is removed using a grit pump. The scraper blade is installed on a pivoting arm and is raised during reverse travel of the bridge. Troughs receiving the pumped grit slurry must be designed for abrasion resistance.

Scum Removal. Aerated grit chambers may also be designed or existing grit chambers modified to provide FOG/scum removal (see Fig. 5–39). In this application, a baffle wall is provided that passes longitudinally through the aerated grit chamber. The baffle wall extends below the water surface and separates the tank into two parallel channels; a grit channel and a grease channel. The rising air buoys up entrained grease within the wastewater and brings it to the surface where the spiral roll moves the grease across the grit channel towards the longitudinal baffle. The lighter grease particles move through openings in the baffle to the grease channel. The heavier grit particles settle and move downward, where they collect in the channel hopper at the bottom, similar to a conventional aerated grit chamber. The grease collected in the grease channel is continuously moved by air/water skimming jets to the discharge end of the channel, where the collected grease is removed by a grease removal screw.

Emissions from Aerated Grit Chambers. Influent wastewater may contain constituents that when subjected to air agitation may be stripped from the wastewater. Where wastewater has been retained in the collection system for a long time, especially in warm

Figure 5–39

Aerated grit chamber with scum and FOG removal for small and intermediate size treatment plant: (a) plan view, (b) typical cross section, and (c) view of grit chamber (adapted from Schreiber).



climates, hydrogen sulfide and other odorous gases may be released from aerated grit chambers. In areas where industrial wastewater is discharged to the collection system, the release of volatile organic compounds (VOCs) may need to be considered as the release of significant amounts of VOCs can be a health risk to the treatment plant operators. Where emissions are an important consideration, covers may be required or non-aerated-type grit chambers used.

EXAMPLE 5–8 **Design of Aerated Grit Chamber** Design an aerated grit chamber for the treatment of municipal wastewater. The average flowrate is 0.5 m^3/s (11.4 Mgal/d), and the peaking factor curve given on Fig. 3–13 is applicable.

Solution

1. Establish the peak flowrate for design. Assume that the aerated grit chamber will be designed for the 1-d sustained peak flowrate. From Fig. 3–11, the peaking factor is found to be 2.75, and the peak design flowrate is

   $$\text{Peak flowrate} = 0.5\ \text{m}^3/\text{s} \times 2.75 = 1.38\ \text{m}^3/\text{s}$$

2. Determine the grit chamber volume. Because it will be necessary to drain the chamber periodically for routine maintenance, use two chambers. Assume the average detention time at the peak flowrate is 3 min.

   $$\text{Grit chamber volume, m}^3\ \text{(each)} = (1/2)(1.38\ \text{m}^3/\text{s})(3\ \text{min})(60\ \text{s/min}) = 124.2\ \text{m}^3$$

3. Determine the dimensions of each grit chamber. Use a width-to-depth ratio of 1.2:1 and assume that the depth is 3 m.
   a. Width = 1.2 (3 m) = 3.6 m

   $$\text{b. Length} = \frac{\text{volume}}{\text{width} \times \text{depth}} = \frac{124.2\ \text{m}^3}{3\ \text{m} \times 3.6\ \text{m}} = 11.5\ \text{m}$$

4. Determine the average detention time in each grit chamber at average flow.

$$\text{Detention time} = \frac{124.2\ \text{m}^3}{(0.25\ \text{m}^3/\text{s})} = 496.8\ \text{s}\left(\frac{1\ \text{min}}{60\ \text{s}}\right) = 8.28\ \text{min}$$

5 Determine the air supply requirement. Assume that 0.3 $\text{m}^3/\text{min}\cdot\text{m}$ of length will be adequate.

$$\begin{aligned}\text{Air required (length basis)} &= (11.5\ \text{m})(0.3\ \text{m}^3/\text{min}\cdot\text{m})\\ &= 3.45\ \text{m}^3/\text{min for each grit chamber}\end{aligned}$$

$$\text{Total air supply required} = 3.45 \times 2 = 6.9\ \text{m}^3/\text{min}\ (244\ \text{ft}^3/\text{min})$$

6. Estimate the daily quantity of grit. Assume a value of 0.05 $\text{m}^3/10^3\ \text{m}^3$.

$$\begin{aligned}\text{Volume grit} &= [(0.5\ \text{m}^3/\text{s})(86{,}400\ \text{s/d})(0.05\ \text{m}^3/10^3\ \text{m}^3)]\\ &= 2.16\ \text{m}^3/\text{d}\ (76\ \text{ft}^3/\text{d})\end{aligned}$$

Comment In designing aerated grit chambers, methods of regulating the air flowrate should be provided to control grit removal and improve the cleanliness of the grit.

Vortex-Type Grit Chambers. Grit is also removed in devices that use a vortex flow pattern. Three types of devices are shown on Fig. 5–40 and discussed below.

Mechanically Induced Vortex. In the mechanically induced vortex unit illustrated on Fig. 5–40(a), wastewater is directed to the vortex unit by a long, straight inlet channel that is designed to guide the wastewater flow into the vortex unit while the grit is directed downward. The toroidal motion caused by the tangential entry is enhanced by a rotating turbine impeller within the unit that causes the grit to move along the flat floor towards the center where it passes under a separating plate into a grit slurry hopper while the lighter organic materials are maintained in suspension and are guided to the surface where they leave the unit through the tangential exit. The grit hopper contents are kept fluidized by the axial flow propeller or by supplemental water jets. Grit may be removed by a grit slurry or air-lift pump. Typical design data are presented in Table 5–18. If more than two units are installed, special arrangements for flow splitting are required to prevent settling of grit upstream of grit chambers.

Hydraulically Induced Vortex. In the hydraulically induced vortex unit, the vortex, illustrated on Fig. 5–40(c), the vortex is generated by the flow entering the unit without any mechanical rotating devices. Wastewater is directed to the unit by a long, straight inlet channel and enters tangentially into the cylindrical unit, causing the contents to rotate slowly about the vertical axis. Flow spirals down the perimeter, allowing the grit and sand particles to settle. The internal components direct the main flow away from the perimeter and back to the middle where the degritted effluent is discharged near the center of the unit into an effluent channel. The heavy grit moves down a spiral path to the center where it passes under a center cone to the grit slurry hopper.

Headloss in the unit is a function of the size particle to be removed and increases significantly for very fine particles. Vortex grit-removal units are sized to handle peak flowrates up to 0.3 m^3/s (7 Mgal/d) per unit. Grit is removed from the unit by a cleated belt conveyor. Because of its overall height, this type of grit system requires a deep basement, or a lift station if it is installed above grade.

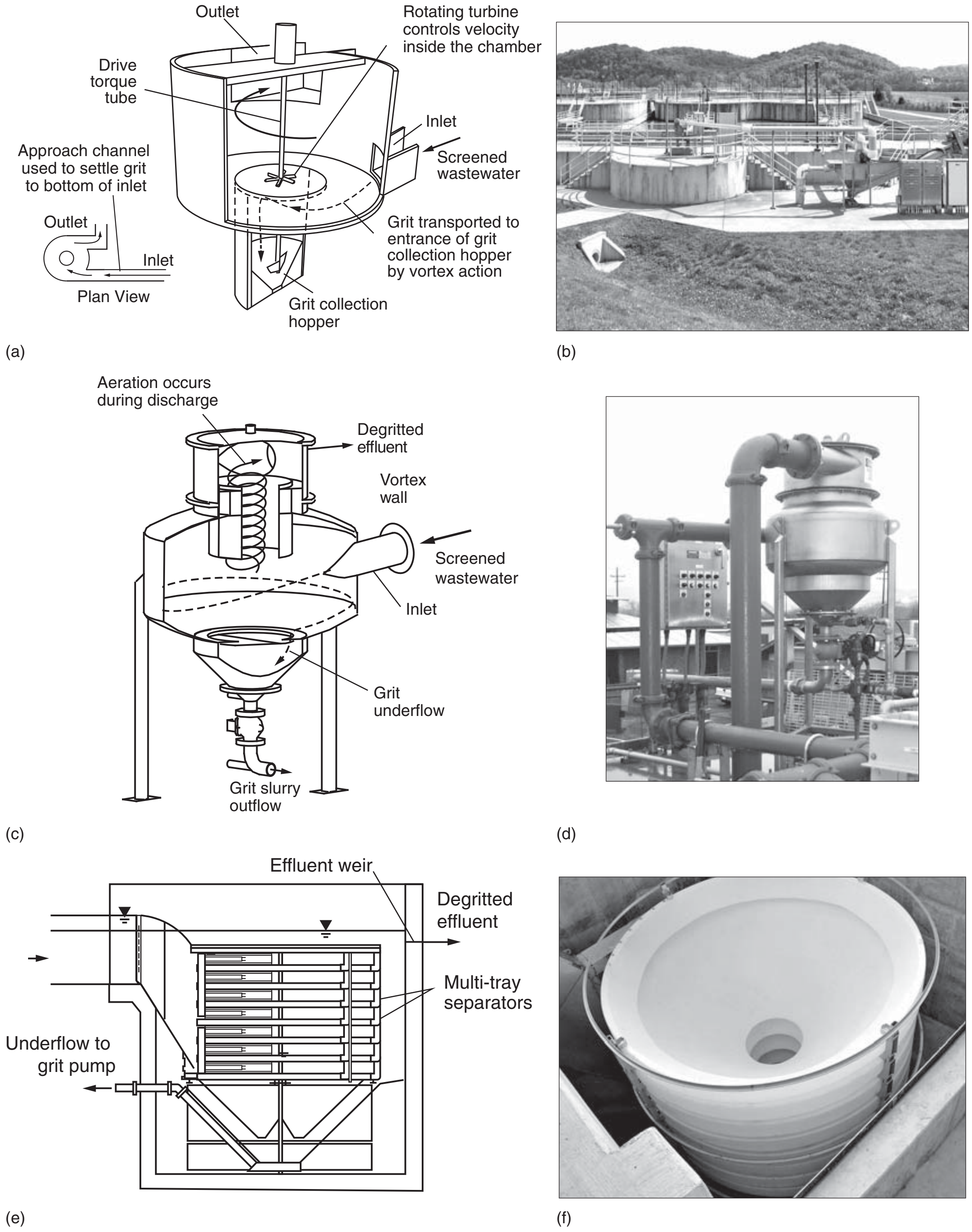


Figure 5–40

Vortex-type grit chambers: (a) schematic Pista® Grit Separator (adapted from Smith & Loveless), (b) view of typical installation (courtesy of Smith & Loveless) (c) schematic of Eutek TeaCup® separator (adapted from Hydro International), (d) view of Eutek TeaCup® separator (courtesy of Hydro International), (e) section through seven-tray Eutek HeadCell® grit separator, and (f) view of Eutek HeadCell of tray grit separator (courtesy of Hydro International).

Table 5–18

Typical design information for vortex-type grit chambers

	U.S. customary units			SI units		
Item	**Unit**	**Range**	**Typical**	**Unit**	**Range**	**Typical**
Detention time at average flowrate	s	20–30	30	s	20–30	30
Diameter						
Upper chamber	ft	4.0–24.0		m	1.2–7.2	
Lower chamber	ft	3.0–6.0		m	0.9–1.8	
Height	ft	9.0–16.0		m	2.7–4.8	
Removal rates[a]						
0.30 mm (50 mesh)	%	92–98	95+	%	92–98	95+
0.21 mm (70 mesh)	%	80–90	85+	%	80–90	85+
0.149 mm (100 mesh)	%	60–70	65+	%	60–70	65+

[a] Based on grit with a specific gravity of 2.5 to 2.65.

Multi-tray Vortex Grit Separator. The proprietary multi-tray vortex grit separator consists of multiple stacked trays which maximizes surface area and minimizing settling distances on Fig. 5–40(e). This allows for a very compact installation with low headloss. Flow is directed into the multiple tray unit by an influent distributor header which feeds a manifold which evenly distributes influent tangentially into the multiple-tray system. The tangential feed establishes a hydraulically forced vortex flow pattern where grit settles by gravity along the sloped surface of each tray and through a center opening which allows collected grit from each tray to settle into a single sump located below the center of the unit. Degritted effluent flows out of the space between the trays and into the surrounding basin where it then exits over a weir.

Grit Separators for Combined Wastewater and Stormwater

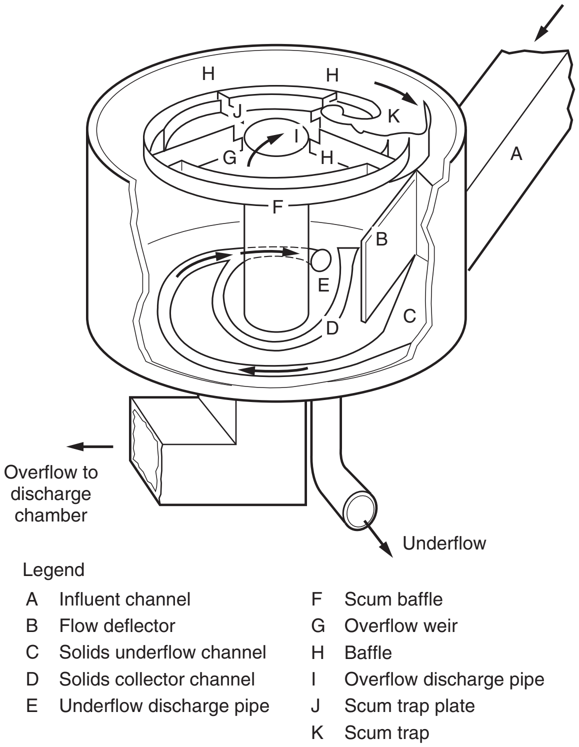
Solids-separation devices such as swirl concentrators and vortex separators have been used in Europe and, to a lesser extent, in the United States for the treatment of combined sewer overflows (CSOs) and stormwater. These devices are compact solids separation units with no moving parts.

Vortex-Type Separators. A typical vortex-type CSO solids-separation unit is illustrated on Fig. 5–41. Operation of vortex separators is based on the movement of particles within the unit. Water velocity moves the particles in a swirling action around the separator, additional flow currents move the particles toward the vortex, gravity pulls particles down, and a sweeping action moves heavier particles across the sloping floor toward the central drain.

During wet weather, the outflow from the unit is throttled, causing the unit to fill and to self-induce a swirling vortex-like flow regime. In the device shown on Fig. 5–41, secondary flow currents rapidly separate settleable grit and floatable matter. Concentrated foul matter is intercepted for treatment while the cleaner, treated flow is discharged to receiving waters. The device is intended to operate under extremely high flow regimes.

Figure 5–41

Typical vortex separator used for solids removal from combined sewer overflows.



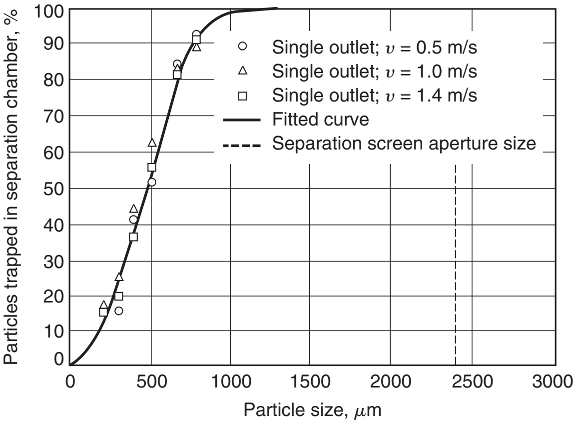
Continuous Deflection Separator. A device more recently developed and termed the continuous deflection separator (CDS) differs from the more traditional vortex separator in that it utilizes a filtration mechanism for solids separation and does not rely on secondary flow currents induced by the vortex action. The CDS system (pictured previously on Fig. 5–5(b) involves a single flow path and has one outlet point while other types of vortex separators discharge flows at the top and bottom of the units. The flow conditions within the CDS separation chamber have a different velocity profile. The surface velocity increases with increasing distance from the center of the separation chamber of the CDS unit, the reverse of that normally observed in conventional vortex separators. Solids separation is enhanced by a large expanded stainless-steel plate that acts as a filter screen with an outer volute outlet passage. The perforations in the separation screen are typically elongated in shape and are aligned with the longer axis in the vertical direction. The separation screen is installed so that the leading edge of each perforation extends into the flow stream of the containment chamber. Perforations in the screen can range from 1200 to 4700 mm (0.0475 to 0.185 in.). The CDS device is most appropriately used to capture the "first flush" and set up to divert all flows up to a threshold discharge. In tests conducted by Wong (1997), greater than 90 percent capture was reported for solids as small as 900 microns (see Fig. 5–42). Headloss through the separation unit varies depending on flowrate and screen openings.

Grit Washing

As some of the heavier organic matter remains normally with the grit, grit washers are used to provide a second stage of volatile solids separation. Grit separated from the main wastewater flow, is transported in a slurry to a washing process to remove organic material. Unwashed grit may contain 50 percent or more of putrescible, organic material, has

Figure 5–42

Solids capture from a continuous deflection separator (Wong, et al., 1997).



a distinctly disagreeable odor, and may attract insects and rodents. The objective of the grit washing process is to achieve a clean grit with a low volatile solids content. Typically, a grit washing system can be expected to result in a clean grit with a volatile solids content of less than 20 percent while retaining at least 95 percent of the depositable grit.

Two principal types of grit washers are available. One type relies on an inclined submerged rake that provides the necessary agitation for separation of the grit from the organic materials and, at the same time, raises the washed grit to a point of discharge above water level (similar to the inclined ramp shown on Fig. 5–35). Another type of grit washer (see Fig. 5–43) uses an inclined screw and moves the grit up the ramp. Both types can be equipped with water sprays to assist in the cleansing action. Hydrocyclone separators are often installed at the inlet to the grit washer to improve grit separation and organics removal.

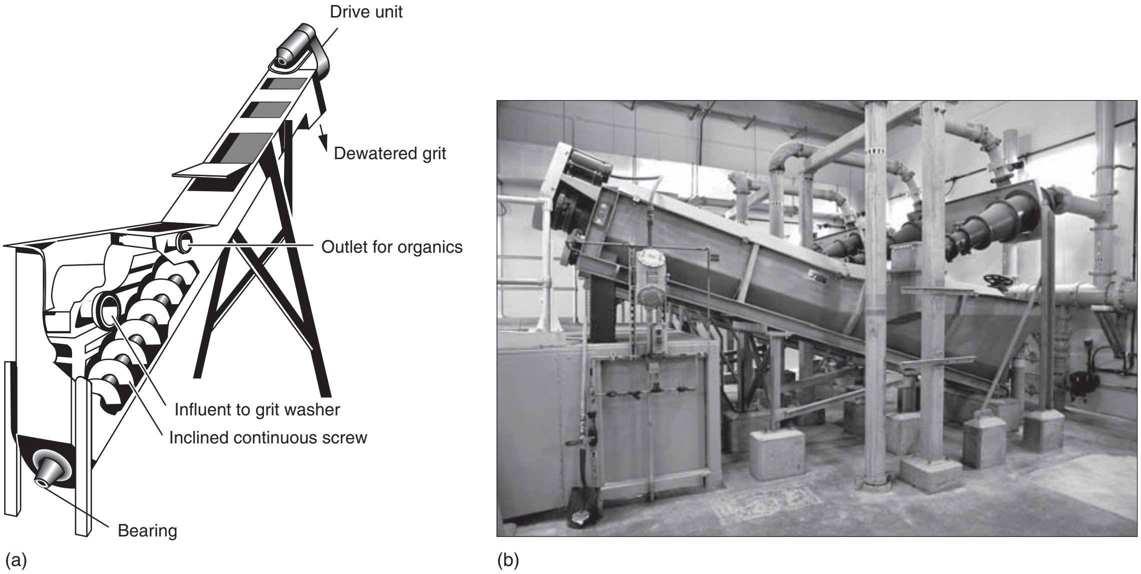


Figure 5–43

Example of grit separation and washing unit: (a) schematic and (b) view of typical installation.

Grit Drying

The clean grit must then be dewatered to remove all free water prior to disposal. Typically, grit is disposed in sanitary landfills and depending on local regulations may have to pass a paint filter liquid test (U.S. EPA, 2004). The objective of the drying process is to achieve a clean, dry grit with a total solids concentration greater than 60 percent while retaining at least 95 percent of the depositable grit.

Disposal of Grit

The most common method of grit disposal is transport to a landfill. In some large plants, grit is incinerated with solids. As with screenings, some states require grit to be lime stabilized before disposal in a landfill. Disposal in all cases should be done in conformance with the appropriate environmental regulations. In larger plants where trucks are used to transfer grit, elevated grit storage facilities may be provided with bottom-loading gates. Difficulties experienced in getting the grit to flow freely from the storage hoppers have been minimized by using steep slopes on the storage hoppers, by applying air beneath the grit, and by the use of hopper vibrators. Drainage facilities for collection and disposal of drippings from the bottom-loading gates are desirable. Grab buckets operating on a monorail system may also be used to load trucks directly from the grit chambers. Pneumatic conveyors are sometimes used to convey grit short distances. Advantages of pneumatic conveying include (1) no elevated storage hoppers are required and (2) attendant odor problems associated with storage are eliminated. The principal disadvantage is the considerable wear on piping, especially at bends.

Solids (Sludge) Degritting

Where grit chambers are not used and the grit is allowed to settle in the primary settling tanks, grit removal is accomplished by pumping dilute quantities of primary sludge to a cyclone degritter. The cyclone degritter acts as a centrifugal separator in which the heavy particles of grit and solids are separated by the action of a vortex and discharged separately from the lighter particles and the bulk of the liquid. The principal advantage of cyclone degritting is the elimination of the cost of constructing, operating, and maintaining grit chambers. The disadvantages are (1) pumping of dilute quantities of solids usually requires solids thickeners and (2) pumping of grit with the liquid primary solids increases the cost of operating and maintaining solids collectors and the primary sludge pumps.

5–6 PRIMARY SEDIMENTATION

The objective of treatment by sedimentation is to remove readily settleable solids and floating material found in wastewater and thus reduce the suspended solids content. Primary sedimentation is typically the first step in the further processing of the wastewater following the removal of coarse solids and grit. Efficiently designed and operated, from 50 to 70 percent of the suspended solids and from 25 to 40 percent of the BOD can be removed by primary sedimentation tanks.

Sedimentation tanks have also been used as stormwater retention tanks to provide a moderate detention period (10 to 30 min) for overflows from either combined sewers or storm sewers. The purpose of sedimentation is to remove a substantial portion of the organic solids that otherwise would be discharged directly to the receiving waters. Sedimentation tanks have also been sized to provide detention periods sufficient for effective disinfection of such overflows. The purpose of this section is (1) to describe the various

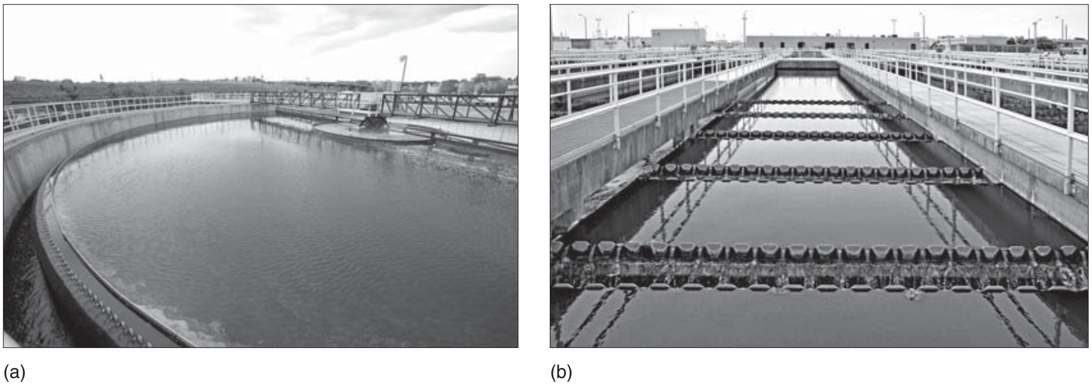


Figure 5–44

Typical sedimentation basins used at wastewater treatment plant: (a) circular with peripheral V-notch effluent weirs and (b) rectangular with inboard V-notch effluent weirs.

types of sedimentation facilities, (2) to consider their performance, and (3) to review important design considerations. Sedimentation tanks used for secondary treatment are considered in Chap. 8.

Description

Almost all treatment plants that have primary sedimentation use mechanically cleaned sedimentation tanks of standardized circular or rectangular design (see Fig. 5–44). The selection of the type of sedimentation unit for a given application is governed by the size of the installation, by rules and regulations of local control authorities, by local site conditions, and by the experience and judgment of the engineer. Two or more tanks should be provided so that the process may remain in operation while one tank is out of service for maintenance and repair work. At large plants, the number of tanks is determined largely by size limitations.

Rectangular Tanks. Rectangular sedimentation tanks may use either chain-and-flight solids collectors or traveling-bridge-type collectors. A rectangular tank that uses a chain-and-flight-type collector is shown on Fig. 5–45. Multiple rectangular tanks require less land area than multiple circular tanks and find application where site space is at a premium. Rectangular tanks also lend themselves to nesting with preaeration tanks and aeration tanks in activated-sludge plants, thus permitting common wall construction and reducing construction costs. They are also used generally where tank roofs or covers are required. While construction costs may be less, the chain-and-flight collector systems used in rectangular tanks generally require more maintenance than the rotating sludge collector mechanism used with circular settling tanks. Important issues with rectangular tanks include (1) sludge removal, (2) flow distribution, and (3) scum removal.

Sludge Removal. Equipment for settled solids removal generally consists of a pair of endless conveyor chains, manufactured of alloy steel, cast iron, or thermoplastic. Attached to the chains at approximately 3 m (10 ft) intervals are scraper flights made of wood or fiberglass, extending the full width of the tank or bay [see Fig. 5–45(d)]. The solids settling in the tank are scraped to solids hoppers in small tanks and to transverse troughs in large tanks. The transverse troughs are equipped with collecting mechanisms (cross collectors),

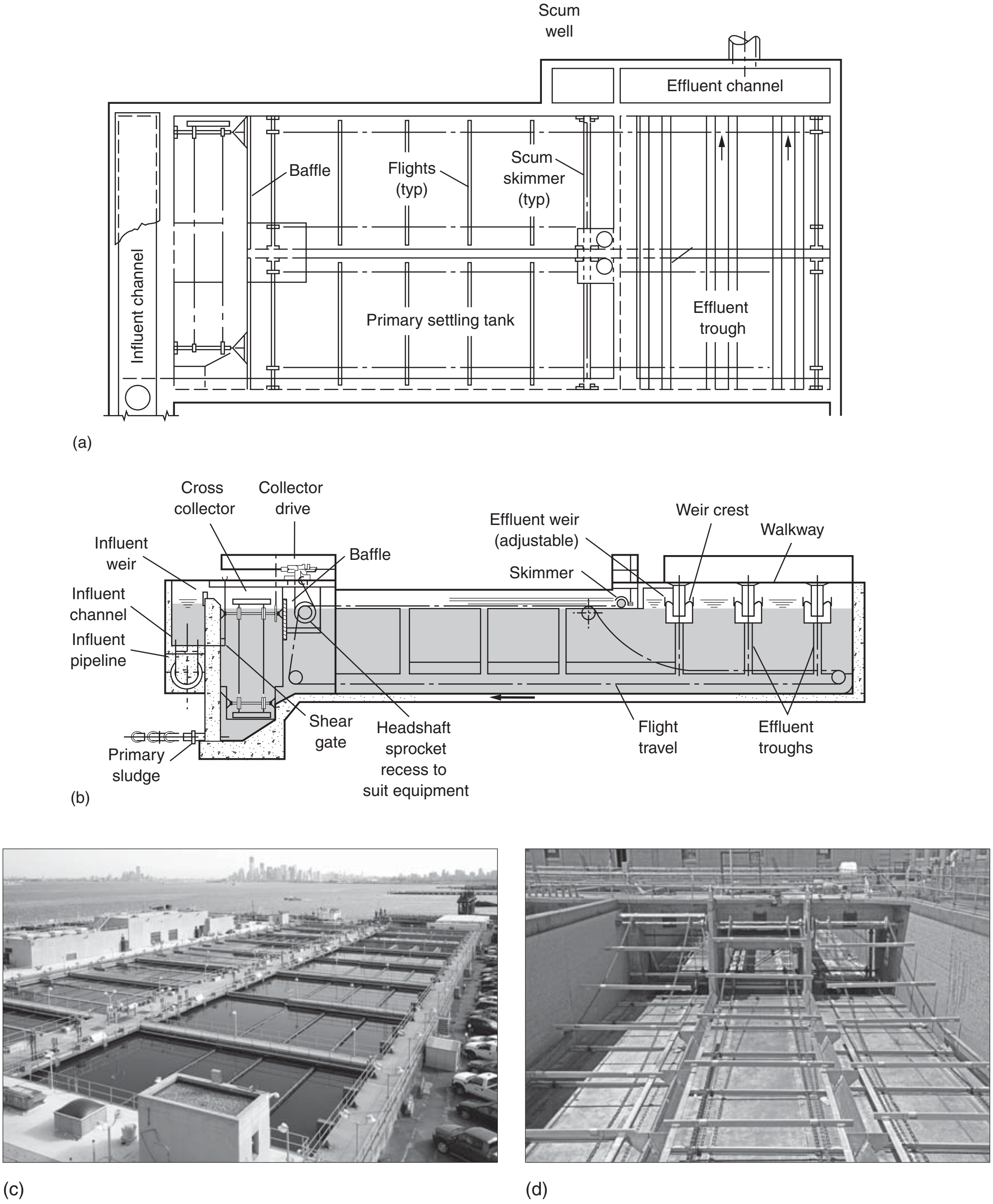


Figure 5–45

Typical rectangular primary sedimentation tank: (a) plan, (b) section, (c) view of large rectangular sedimentation tank with weirs similar to those shown on (b), and (d) view of empty tank with sludge removal mechanism.

usually either chain-and-flight or screw-type collectors, which convey solids to one or more collection hoppers. In very long units (over 50 m), two collection mechanisms can be used to scrape solids to collection points near the middle of the tank length.

Where possible, it is desirable to locate solids pumping facilities close to the collection hoppers. Where cross collectors are not provided, multiple solids hoppers must be installed. Solids hoppers have operating difficulties, notably solids accumulation on the slopes and in the corners and arching over the solids drawoff piping. Wastewater may also be drawn through the solids hopper, bypassing some of the accumulated solids and resulting in a “rathole” effect. A cross collector is more advisable, except possibly in small plants, because a more uniform and concentrated solids can be withdrawn and many of the problems associated with solids hoppers can be eliminated. Rectangular tanks may also be cleaned by a bridge-type mechanism that travels up and down the tank on rubber wheels or on rails supported on the sidewalls. One or more scraper blades are suspended from the bridge. Some of the bridge mechanisms are designed so that the scraper blades can be lifted clear of the solids blanket on the return travel.

Flow Distribution. Because flow distribution in rectangular tanks is critical, one of the following inlet designs is used: (1) full-width inlet channels with inlet weirs, (2) inlet channels with submerged ports or orifices, (3) or inlet channels with wide gates and slotted baffles. Inlet weirs, while effective in spreading flow across the tank width, introduce a vertical velocity component into the solids hopper that may re-suspend the solids particles. The weirs also introduce additional headloss in the plant hydraulic profile. Inlet ports can provide good distribution across the tank width if the velocities are maintained in the 3 to 9 m/min (10 to 30 ft/min) range. Inlet baffles are effective in reducing the high initial velocities and distribute flow over the widest possible cross-sectional area. Where full-width baffles are used, they should extend from 150 mm (6 in.) below the surface to 300 mm (12 in.) below the entrance opening.

For installations of multiple rectangular tanks, below-grade pipe and equipment galleries can be constructed integrally with the tank structure and along the influent end. The galleries are used to house the sludge pumps and sludge drawoff piping. The galleries also provide access to the equipment for operation and maintenance. Galleries can also be connected to service tunnels for access to other plant units.

Scum Removal. Scum is usually collected at the effluent end of rectangular tanks with the flights returning at the liquid surface. The scum is moved by the flights to a point where it is trapped by baffles before removal. Water sprays can also move the scum. The scum can be scraped manually up an inclined apron, or it can be removed hydraulically or mechanically, and for scum removal a number of means have been developed. For small installations, the most common scum drawoff facility consists of a horizontal, slotted pipe that can be rotated by a lever or a screw. Except when drawing scum, the open slot is above the normal tank water level. When drawing scum, the pipe is rotated so that the open slot is submerged just below the water level, permitting the scum accumulation to flow into the pipe. Use of this equipment results in a relatively large volume of scum liquor.

Another method for removing scum is by a transverse rotating helical wiper attached to a shaft. Scum is removed from the water surface and moved over a short inclined apron for discharge to a cross-collecting scum trough. The scum may then be flushed to a scum ejector or hopper ahead of a scum pump. Another method of scum removal consists of a chain-and-flight type of collector that collects the scum at one side of the tank and scrapes it up a short incline for deposit in scum hoppers, whence it can be pumped to disposal units. Scum is also collected by special scum rakes in rectangular tanks that are equipped with the carriage or bridge type of sedimentation tank equipment. In installations where appreciable amounts of scum are collected, the scum hoppers are usually equipped with mixers to provide a homogeneous mixture prior to pumping. Scum is usually disposed of with the solids and biosolids produced at the plant; however, separate scum disposal is used at many plants.

Circular Tanks. Circular sedimentation tanks are customarily arranged in groups of two or four. The flow is divided among the tanks by a flow-split structure, commonly located between the tanks. Solids are usually withdrawn by sludge pumps for discharge to the solids processing and disposal units. Important considerations include (1) method of introducing flow, (2) energy dissipation, and (3) sludge removal.

Flow Pattern. In circular tanks the flow pattern is radial (as opposed to horizontal in rectangular tanks). To achieve a radial flow pattern, the wastewater to be settled can be introduced in the center or around the periphery of the tank, as shown on Fig. 5–46. Both flow configurations have proved to be satisfactory generally, although the center-feed type is more commonly used, especially for primary treatment. In the center-feed design [see Fig. 5–46(a)], the wastewater is transported to the center of the tank in a pipe suspended from the bridge, or encased in concrete beneath the tank floor. At the center of the tank, the wastewater enters a circular well designed to distribute the flow equally in all directions [see Fig. 5–46(a)]. The center well has a diameter typically between 15 and

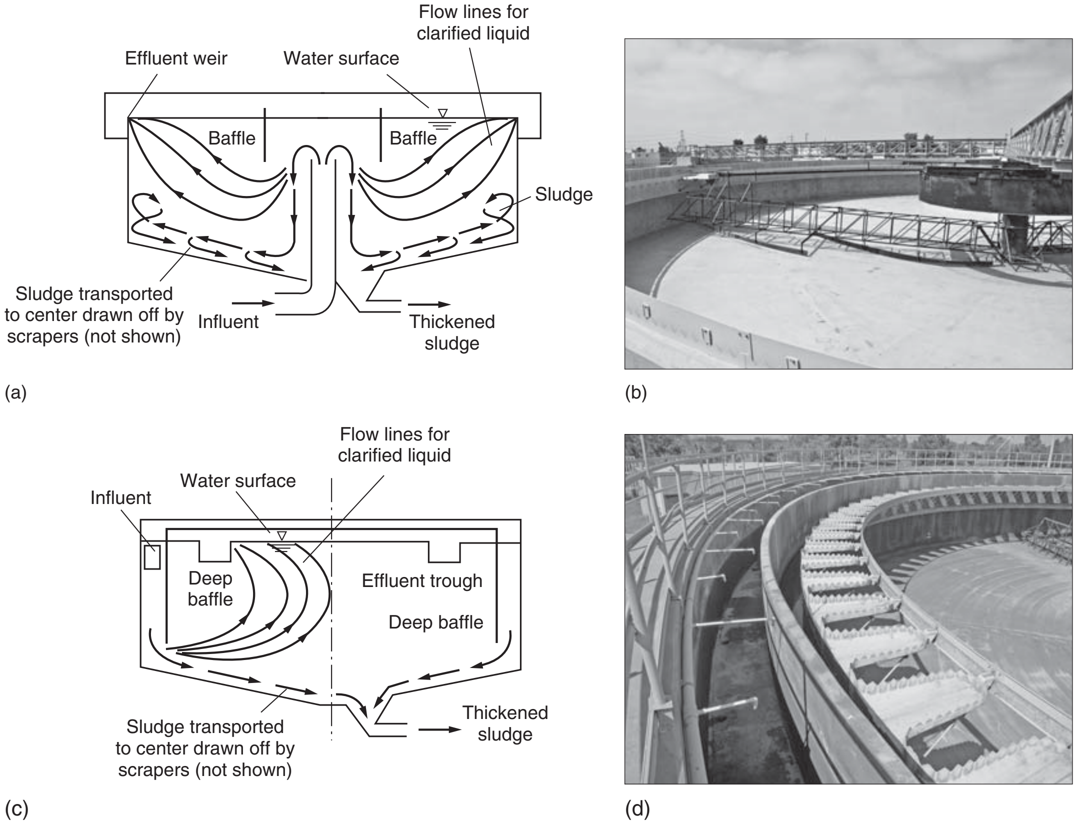
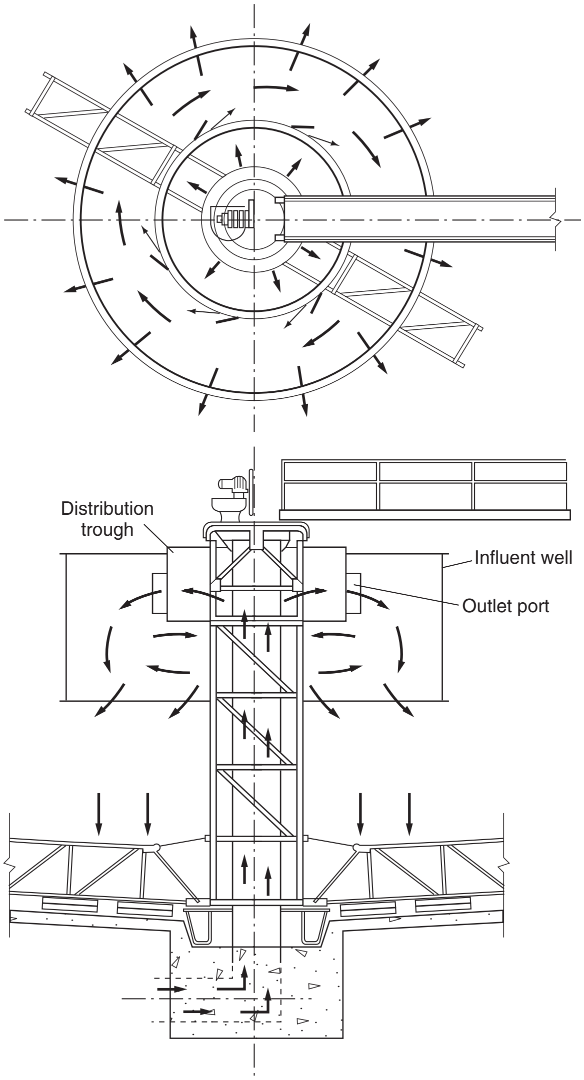


Figure 5–46

Typical circular sedimentation tanks: (a) schematic of center feed, (b) view of center feed unit, (c) schematic of peripheral feed, and (d) view of a peripheral feed unit.

Figure 5–47

Typical energy dissipating and flow distribution inlet for a center-feed sedimentation tank. The inner ring is used to create a tangential flow pattern (Randle, et al., 1992).



25 percent of the total tank diameter and ranges from 1 to 2.5 m (3 to 8 ft) in depth and should have a tangential energy-dissipating inlet within the feedwell.

Energy Dissipation. The energy-dissipating device (see Fig. 5–47) functions to collect influent from the center column and discharge it tangentially into the upper 0.5 to 0.7 m of the feedwell. The discharge ports are sized to produce a velocity of 0.75 m/s at maximum flow and 0.30 to 0.45 m/s at average flow. The feedwell should be sized so that the maximum downward velocity does not exceed 0.75 m/s. The depth of the feedwell should extend about 1 meter below the energy-dissipating inlet ports (Randall et al., 1992). An alternative approach to energy dissipation is to taper the inlet riser pipe so that the pipe increases in size from bottom to top. A center column formed in concrete can provide this configuration. The outlet of the riser pipe should terminate below the water surface. Vertical supports extending upward from the center column are used to support the center drive for the sludge collector mechanism. The tapered inlet with submerged outlet has the effect of reducing the inlet velocity of flow to the feedwell, thereby dissipating inlet energy.

In the peripheral-feed design [see Fig. 5–46(b)], a suspended circular baffle forms an annular space into which the inlet wastewater is discharged in a tangential direction. The wastewater flows spirally around the tank and underneath the baffle, and the clarified liquid is skimmed off over weirs on both sides of a centrally located weir trough. Grease and scum are confined to the surface of the annular space. Peripheral feed tanks are used generally for secondary clarification.

Sludge Removal. Circular tanks 3.6 to 9 m (12 to 30 ft) in diameter have the solids-removal equipment supported on beams spanning the tank. Tanks 10.5 m (35 ft) in diameter and larger have a central pier that supports the mechanism and is reached by a walkway or bridge. The bottom of the tank is sloped at about 1 in 12 (vertical:horizontal) to form an inverted cone, and the solids are scraped to a relatively small hopper located near the center of the tank. Airlift pumps can also be used to withdraw sludge minimizing the costs associated with operating and maintaining sludge pumps.

Combination Flocculator-Clarifier. Combination flocculator-clarifiers are often used in water treatment and sometimes used for wastewater treatment, especially in cases where enhanced settling, such as for industrial wastewater treatment or for biosolids concentration, is required. Inorganic chemicals or polymers can be added to improve flocculation. Circular clarifiers are ideally suited for incorporation of an inner, cylindrical flocculation compartment (see Fig. 5–48). Wastewater enters through a center shaft or well and flows into the flocculation compartment, which is generally equipped with a paddle-type or low-speed mixer. The gentle stirring causes flocculent particles to form. From the flocculation compartment, flow then enters the clarification zone by passing down and radially outward. Settled solids and scum are collected in the same way as in a conventional clarifier.

Stacked (Multilevel) Clarifiers. Stacked clarifiers originated in Japan in the 1960s where limited land area is available for the construction of wastewater treatment facilities. Since that time, stacked clarifiers have been used in the United States, the most notable installation of which is at the Deer Island Wastewater Treatment Plant constructed in Boston Harbor. Design of these types of clarifiers recognizes the importance of settling

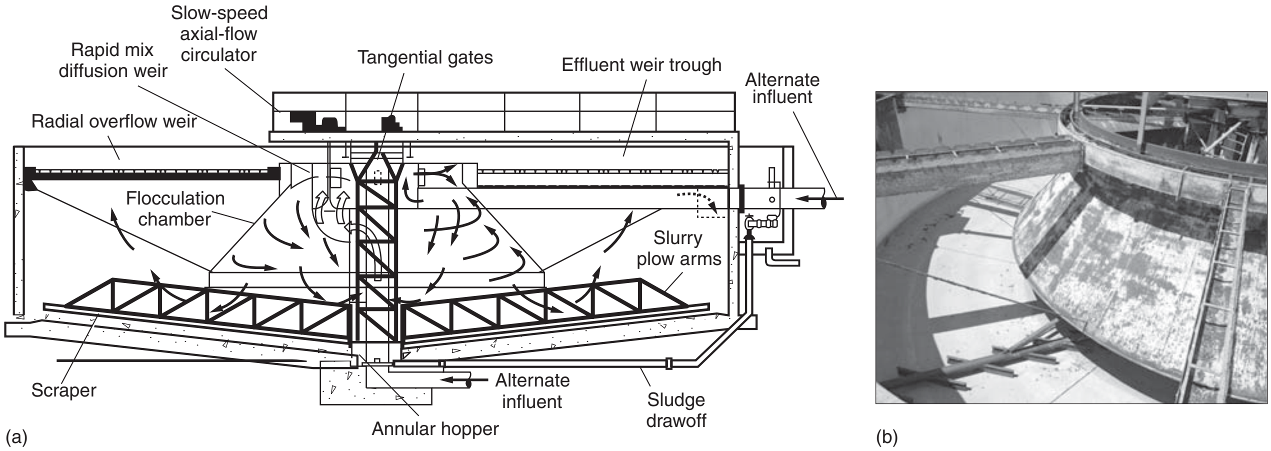
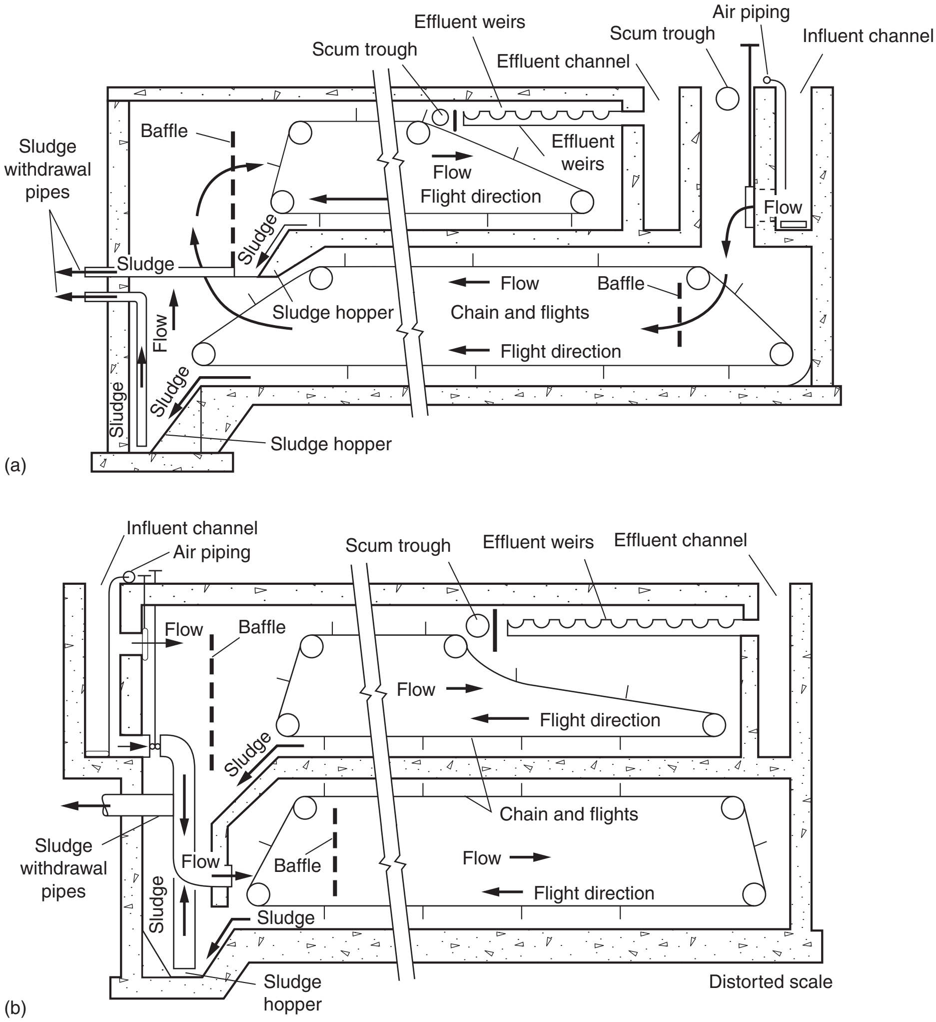


Figure 5–48

Typical flocculator-clarifier: (a) schematic and (b) view of empty tank. In some designs, turbine or propeller mixers are included in the flocculation chamber.

Figure 5–49

Typical section through a stacked clarifier (a) series flow and (b) parallel flow type used at the Deer Island Wastewater Treatment Plant, Boston, MA. Note: In the parallel flow type, the upper effluent weirs serve both the upper and lower clarifiers. Channels for the discharge of effluent from the lower to the upper clarifier are located on either side of the sludge collection mechanism in the upper clarifier.



area to settling efficiency. Operation of stacked rectangular clarifiers is similar to conventional rectangular clarifiers in terms of influent and effluent flow patterns and solids collection and removal. The stacked clarifiers are actually two (or more) tanks, one located above the other, operating on a common water surface (see Fig. 5–49). Each clarifier is fed independently, resulting in parallel flow through the lower and upper tanks. Settled solids are collected from each tank with chain and flight solids collectors, discharging to a common hopper. In addition to saving space, advantages claimed for stacked clarifiers include less piping and pumping requirements. Because the facilities are more compact and have less exposed surface area, better control of odors and volatile organic compound emissions is possible. Disadvantages include higher construction cost than conventional clarifiers and more complex structural design. Design criteria for stacked clarifiers, as regards overflow and weir rates, are similar to conventional primary and secondary clarifiers.

Sedimentation Tank Weirs. Settled effluent from sedimentation tanks is discharged over weirs into an effluent launder which, in turn, discharges into a conduit or open channel leading to the biological treatment process. Small rectangular settling tanks are often fitted with a single transverse weir on the discharge end wall. However, larger rectangular settling tanks require additional weir length to minimize headloss and avoid short circuiting. Headloss and short circuiting can be minimized by adding a series of inboard launders with double weirs that are aligned either longitudinally (parallel to the length of the settling tank) or transversely (perpendicular to the length of the settling tank) [see Fig. 5–45(c)] or both longitudinally and transversely [see Fig. 5–50(a)].

Figure 5–50

Typical sedimentation tank appurtenances used for sedimentation tanks: (a) internal longitudinal and transverse weirs in rectangular sedimentation tanks, (b) peripheral V-notch effluent weirs launders in peripheral feed circular basin, (c) and (d) covered peripheral weirs to limit algal growth, and (e) and (f) covers for primary clarifiers to eliminate the release of odors.

Because most circular settling tanks are designed with center feed, treated effluent is discharged over a series of V-notch weirs located peripherally at the tank wall [see Figs. 5–46(a) and (b)]. A similar arrangement is used for peripheral feed circular clarifier [see Fig. 5–50(b)]. For large circular settling tanks, inboard launders with double weirs may be provided to increase the number of V-notches and reduce the headloss at high flow [see Fig. 5–46(d)]. In these cases the weir plates must be designed to balance the flow from both weirs by proportioning the number of V-notches on each launder by the proportion of tank surface area that it serves.

The V-notches are spaced equally along the tank circumference and provide a simple means to detect short circuiting when the weir is not level. The V-notches are normally 90 degree Vs and have a total depth of 75 mm (3 in.). The number of V-notches used is based on limiting the depth from about 13 to 63 mm (0.5 to 2.5 in.) from low to high flow. Surface baffles are placed in front of the peripheral weir to retain surface scum [see Figs. 5–46(b) and Fig. 5–50(b)]. Density currents tend to form along the floor and wall of the settling tanks that can re-suspend solids. Consequently, the weir launder box is constructed on the inside of the perimeter wall or horizontal baffles are provided on the tank wall to redirect the density currents. To minimize maintenance cost and to limit algal growth, peripheral weirs are often covered [see Figs. 5–50(c) and (d)].

Covers for Sedimentation Tanks. In many locations where residential development has surrounded a pre-existing wastewater treatment plant, primary sedimentation facilities are often covered to limit odorous emissions. A variety of different types of covers have been used (see Fig. 5–50(e) and (f)). The use of covers and the processing of the gases released from the sedimentation tanks is considered in greater detail in Chap. 16.

Sedimentation Tank Performance

The efficiency of sedimentation basins with respect to the removal of BOD and TSS is reduced by (1) eddy currents formed by the inertia of the incoming fluid, (2) wind-induced circulation cells formed in uncovered tanks, (3) thermal convection currents, (4) cold or warm water causing the formation of density currents that move along the bottom of the basin and warm water rising and flowing across the top of the tank, and (5) thermal stratification in hot, arid climates (Fair and Geyer, 1954). Factors that affect performance are considered in the following discussion.

BOD and TSS Removal. Typical performance data for the removal of BOD and TSS in primary sedimentation tanks, as a function of the detention time and constituent concentration, are presented on Fig. 5–51. The curves shown on Fig. 5–51 are derived from observations of the performance of actual sedimentation tanks. The curvilinear relationships in the figure can be modeled as rectangular hyperbolas using the following relationship (Crites and Tchobanoglous, 1998):

$$R = \frac{t}{a + bt} \tag{5–45}$$

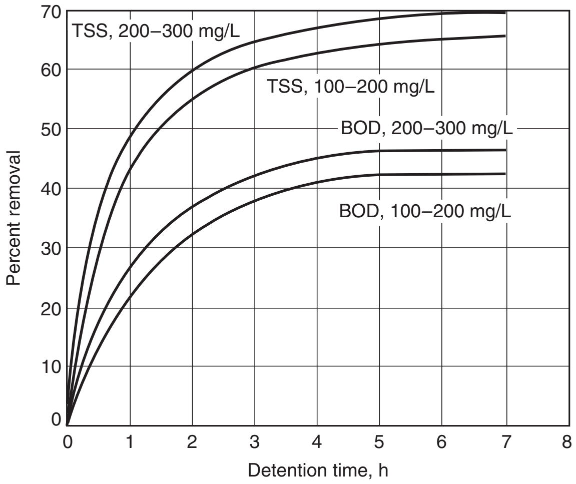
where R = expected removal efficiency
t = nominal detention time, T
a, b = empirical constants

Typical values for the empirical constants in Eq. (5–45) at 20°C are as follows:

Item	b	a
BOD	0.020	0.018
TSS	0.014	0.0075

Figure 5–51

Typical BOD and TSS removal in primary sedimentation tanks (Greeley, 1938).

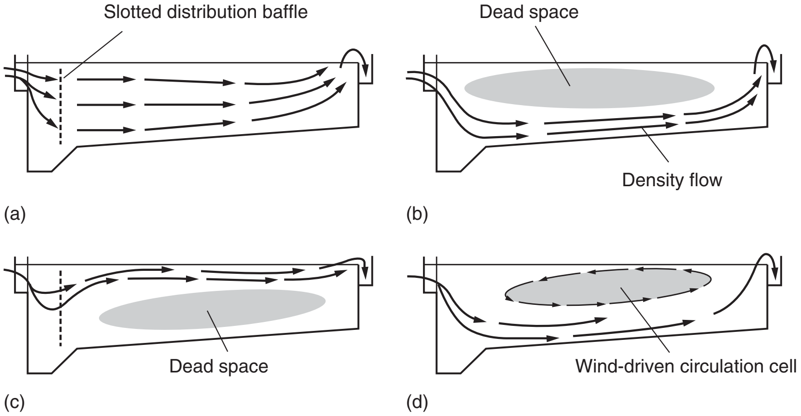


A fact that is often overlooked in sedimentation tank performance is the change in the wastewater characteristics that occurs through the sedimentation process. Larger, more slowly biodegradable suspended solids settle first, leaving a more volatile fraction in suspension that remains in the primary tank effluent. The strict use of removal curves, such as those given on Fig. 5–51, does not account for the transformation in wastewater characteristics that actually occurs. Where possible for domestic wastewater, primary tank influent and effluent should be characterized to determine concentration and composition of the constituents. Such characterization is important when determining the organic loading rate required to be treated by the succeeding biological treatment units. Further discussion on the effects of wastewater characterization on biological processes is contained in Chaps. 7 and 8.

Short Circuiting and Hydraulic Stability. In an ideal sedimentation basin, a given block of entering water should remain in the basin for the full detention time (see Fig. 5–52(a)). Short circuiting, which can lead to reduced treatment performance, can result from temperature differences [see Figs. 5–52(b) and (c)], wind driven circulation patterns [see Fig. 5–52(d)], and the presence of dead zones resulting from poor design, inadequate mixing and dispersion (see Fig. H-1 in Appendix H).

Figure 5–52

Typical flow patterns observed in rectangular sedimentation tanks: (a) ideal flow, (b) effect of density flow or thermal stratification (water in tank is warmer than influent), (c) effect of density stratification (influent wastewater is warmer than water in tank) and (d) formation of wind-driven circulation cell (Crites and Tchobanoglous, 1998).



Temperature Effects. Temperature effects can be significant in sedimentation basins. It has been shown that a 1°C temperature differential between the incoming wastewater and the wastewater in the sedimentation tank will cause a density current to form. The impact of the temperature effects on performance will depend on the material being removed and its characteristics. Temperature effects can be more pronounced in secondary settling tanks where a less dense sludge is processed.

Wind Effects. Wind blowing across the top of open sedimentation basins can cause circulation cells to form. When circulation cells form, the effective volumetric capacity of the basin is reduced. As with temperature effects, the impact of the reduced volume on performance will depend on the material being removed and its characteristics.

Modeling Basin Performance. To determine if short circuiting exists and to what extent, tracer studies should be performed, as discussed in Appendix H. Time-concentration curves should be developed for analysis. If in the repeated tests the time-concentration curves are similar, then the basin is stable. If the time-concentration curves (also known as residence time distribution (RTD) curves (see Appendix H) are not repeatable, the basin is unstable and the performance of the basin will be erratic (Fair and Geyer, 1954). The method of influent flow distribution, as discussed above, will also affect short circuiting. The effects of short circuiting on the performance of sedimentation tanks and other reactors was examined extensively by Morrill (1932), who based on his work developed the Morrill Dispersion Index (MDI) for the evaluation of reactor performance (see Appendix H). The application of the MDI method of analysis is illustrated in Example 12–8 in Chap. 12.

Design Considerations

If all solids in wastewater were discrete particles of uniform size, uniform density, uniform specific gravity, and uniform shape, the removal efficiency of these solids would be dependent on the surface area of the tank and time of detention. The depth of the tank would have little influence, provided that horizontal velocities would be maintained below the scouring velocity. However, the solids in most wastewaters are not of such regular character but are heterogeneous in nature, and the conditions under which they are present range from total dispersion to complete flocculation. Design parameters for sedimentation are considered below. Typical design information and dimensions for rectangular and circular sedimentation tanks used for primary treatment are presented in Tables 5–19 and 5–20, respectively. Additional details on the analysis and design of sedimentation tanks may be found in WPCF, 1985. A design procedure is illustrated in Example 5–9.

Detention Time. The bulk of the finely divided solids reaching primary sedimentation tanks is incompletely flocculated but is susceptible to flocculation. Flocculation is aided by eddying motion of the fluid within the tanks and proceeds through the coalescence of fine particles, at a rate that is a function of their concentration and of the natural ability of the particles to coalesce upon collision. As a general rule, coalescence of a suspension of solids becomes more complete as time elapses; thus, detention time is a consideration in the design of sedimentation tanks. The mechanics of flocculation are such, however, that as the time of sedimentation increases, less and less coalescence of remaining particles occurs.

Normally, primary sedimentation tanks are designed to provide 1.5 to 2.5 h of detention based on the average rate of wastewater flow. Tanks that provide shorter detention periods (0.5 to 1 h), with less removal of suspended solids, are sometimes used for preliminary

Table 5–19

Typical design information for primary sedimentation tanks[a]

	U.S. customary units			SI units		
Item	**Unit**	**Range**	**Typical**	**Unit**	**Range**	**Typical**
Primary sedimentation tanks followed by secondary treatment						
Detention time	h	1.5–2.5	2.0	h	1.5–2.5	2.0
Overflow rate						
Average flowrate	gal/ft^2·d	800–1200	1000	m^3/m^2·d	30–50	40
Peak hourly flowrate	gal/ft^2·d	2000–3000	2500	m^3/m^2·d	80–120	100
Weir loading rate	gal/ft·d	10,000–40,000	20,000	m^3/m·d	125–500	250
Primary settling with waste activated sludge return						
Detention time	h	1.5–2.5	2.0	h	1.5–2.5	2.0
Overflow rate						
Average flowrate	gal/ft^2·d	600–800	700	m^3/m^2·d	24–32	28
Peak hourly flowrate	gal/ft^2·d	1200–1700	1500	m^3/m^2·d	48–70	60
Weir loading rate	gal/ft·d	10,000–40,000	20,000	m^3/m·d	125–500	250

[a] Comparable data for secondary clarifiers are presented in Chap. 8.

Table 5–20

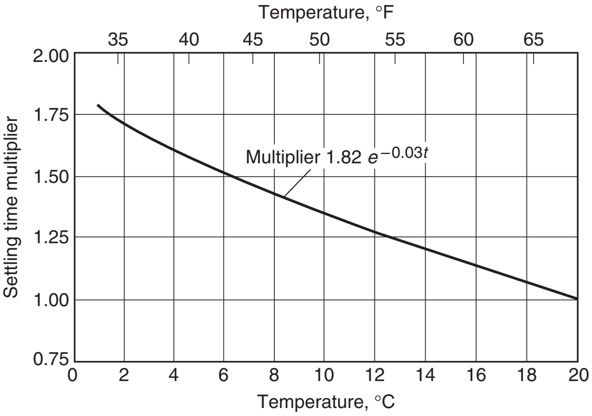
Typical dimensional data for rectangular and circular sedimentation tanks used for primary treatment of wastewater

	U.S. customary units			SI units		
Item	**Unit**	**Range**	**Typical**	**Unit**	**Range**	**Typical**
Rectangular:						
Depth	ft	10–16	14	m	3–4.9	4.3
Length	ft	50–300	80–130	m	15–90	24–40
Width[a]	ft	10–80	16–32	m	3–24	4.9–9.8
Flight speed	ft/min	2–4	3	m/min	0.6–1.2	0.9
Circular:						
Depth	ft	10–16	14	m	3–4.9	4.3
Diameter	ft	10–200	40–150	m	3–60	12–45
Bottom slope	in./ft	3/4–2	1.0	mm/mm	1/16–1/6	1/12
Flight speed	rev/min	0.02–0.05	0.03	rev/min	0.02–0.05	0.03

[a] If widths of rectangular mechanically cleaned tanks are greater than 6 m (20 ft), multiple bays with individual cleaning equipment may be used, thus permitting tank widths up to 24 m (80 ft) or more.

Figure 5–53

Curve of the increase in detention time required at cooler temperatures to achieve the same sedimentation performance as achieved at 20°C.



treatment ahead of biological treatment units. In cold climates, increases in water viscosity at lower temperatures retard particle settling in clarifiers and reduce performance at wastewater temperatures below 20°C (68°F). A curve showing the increase in detention time necessary to equal the detention time at 20°C is presented on Fig. 5–53 (WPCF, 1985). For wastewater having a temperature of 10°C, for example, the detention period is 1.38 times that required at 20°C to achieve the same efficiency. Thus, in cold climates, safety factors should be considered in clarifier design to ensure adequate performance.

Surface Loading Rates. Sedimentation tanks are normally designed on the basis of a surface loading rate (commonly termed *overflow rate*) expressed as cubic meters per square meter of surface area per day, $m^3/m^2 \cdot d$ (gallons per square foot of surface area per day, $gal/ft^2 \cdot d$). The selection of a suitable loading rate depends on the type of suspension to be separated. Typical values for various suspensions, without and with return activated sludge, are reported in Table 5–19. Designs for municipal plants must also meet the approval of state regulatory agencies, many of which have adopted standards for surface loading rates that must be followed. When the area of the tank has been established, the detention period in the tank is governed by water depth. Overflow rates in current use result in nominal detention periods of 2.0 to 2.5 h, based on average design flow.

The effect of the surface loading rate and detention time on suspended solids removal varies widely depending on the character of the wastewater, proportion of settleable solids, concentration of solids, and other factors. It should be emphasized that overflow rates must be set low enough to ensure satisfactory performance at peak rates of flow, which may vary from over 3 times the average flow in small plants to 2 times the average flow in large plants (see discussion of peak flowrates in Chap. 3).

Weir Loading Rates. In general, weir loading rates have little effect on the efficiency of primary sedimentation tanks and should not be considered when reviewing the appropriateness of clarifier design. For general information purposes only, typical weir loading rates are given in Table 5–19. The placement of weirs and baffles in secondary sedimentation applications is discussed in Sec. 8–8 of Chap. 8. Baffles should be placed ahead of effluent weirs in primary settling tanks to reduce hydraulic short circuiting. The baffles also prevent the discharge of scum into the effluent.

Scour Velocity. To avoid the resuspension (scouring) of settled particles, horizontal velocities through the tank should be kept sufficiently low. Using the results from studies by Shields (1936) and Camp (1946) developed the following equation for the critical velocity.

$$v_H = \left[\frac{8k(s - 1)gd}{f}\right]^{1/2} \tag{5–46}$$

where v_H = horizontal velocity that will just produce scour, LT^{-1} (m/s)
k = constant that depends on type of material being scoured (unitless)
s = specific gravity of particles
g = acceleration due to gravity, LT^{-2} (9.81 $\mathrm{m/s^2}$)
d = diameter of particles, L
f = Darcy-Weisbach friction factor (unitless)

Typical values of k are 0.04 for unigranular sand and 0.06 for more sticky, interlocking matter. The term f (the Darcy-Weisbach friction factor) depends on the characteristics of the surface over which flow is taking place and the Reynolds number. Typical values of f are 0.02 to 0.03. Either SI or U.S. customary units may be used in Eq. (5–46), so long as they are consistent, because k and f are dimensionless.

Computational fluid dynamic (CFD) modeling is sometimes used to optimize the design of settling tanks, particularly large diameter circular settling tanks. CFD modeling can be used to optimize feedwell diameter, feedwell depth, center column height and tank depth.

EXAMPLE 5–9 **Design of a Primary Sedimentation Basin** The average flowrate at a small municipal wastewater treatment plant is 20,000 $\mathrm{m^3/d}$. The highest observed peak daily flowrate is 50,000 $\mathrm{m^3/d}$. Design rectangular primary clarifiers with a channel width of 6 m (20 ft). Use a minimum of two clarifiers. Calculate the scour velocity, to determine if settled material will become resuspended. Estimate the BOD and TSS removal at average and peak flow. Use an overflow rate of 40 $\mathrm{m^3/m^2{\cdot}d}$ at average flow (see Table 5–19) and a side water depth of 4 m (13.1 ft).

Solution

1. Calculate the required surface area. For average flow conditions, the required area is

$$A = \frac{Q}{OR} = \frac{(20{,}000\ \mathrm{m^3/d})}{(40\ \mathrm{m^3/m^2\cdot d})} = 500\ \mathrm{m^2}$$

2. Determine the tank length.

$$L = \frac{A}{W} = \frac{500\ \mathrm{m}}{2 \times 6\ \mathrm{m}} = 41.7\ \mathrm{m}$$

However, for the sake of convenience, the surface dimensions will be rounded to 6 m by 42 m.

3. Compute the detention time and overflow rate at average flow.
Using the assumed sidewater depth of 4 m,

$$\text{Tank volume} = 4\ \mathrm{m} \times 2(42\ \mathrm{m} \times 6\ \mathrm{m}) = 2016\ \mathrm{m^2}$$

$$\text{Overflow rate} = \frac{Q}{A} = \frac{(20{,}000\ \mathrm{m^3/d})}{2(6\ \mathrm{m} \times 42\ \mathrm{m})} = 39.7\ \mathrm{m^3/m^2\cdot d}$$

$$\text{Detention time} = \frac{V}{Q} = \frac{(2016\ \mathrm{m^3})(24\ \mathrm{h/d})}{(20{,}000\ \mathrm{m^3/d})} = 2.42\ \mathrm{h}$$

4. Determine the detention time and overflow rate at peak flow.

$$\text{Overflow rate} = \frac{Q}{A} = \frac{(50{,}000\ \text{m}^3/\text{d})}{2(6\ \text{m} \times 42\ \text{m})} = 99.2\ \text{m}^3/\text{m}^2\cdot\text{d}$$

$$\text{Detention time} = \frac{V}{Q} = \frac{(2016\ \text{m}^3)(24\ \text{h/d})}{(50{,}000\ \text{m}^3/\text{d})} = 0.97\ \text{h}$$

5. Calculate the scour velocity (Eq. 5–46), using the following values:

Cohesion constant:	$k = 0.05$
Specific gravity:	$s = 1.25$
Acceleration due to gravity:	$g = 9.81\ \text{m/s}^2$
Diameter of particles:	$d = 100\ \mu\text{m} = 100 \times 10^{-6}\ \text{m}$
Darcy-Weisbach friction factor:	$f = 0.025$

$$v_{H(p)} = \left[\frac{8k(s-1)gd}{f}\right]^{1/2} = \left[\frac{(8)(0.05)(0.25)(9.81)(100 \times 10^{-6})}{0.025}\right]^{1/2} = 0.063\ \text{m/s}$$

6. Compare the scour velocity calculated in the previous step to the peak flow horizontal velocity (the peak flow divided by the cross-sectional area through which the flow passes).

The peak flow horizontal velocity through the settling tank is

$$v_{H(p)} = \frac{Q}{A_x} = \left[\frac{(50{,}000\ \text{m}^3/\text{d})}{2(6\ \text{m} \times 4\ \text{m})}\right]\left[\frac{1}{(24\ \text{h/d})(3600\ \text{s/h})}\right] = 0.012\ \text{m/s}$$

The horizontal velocity value, even at peak flow, is substantially less than the scour velocity. Therefore, settled matter should not be resuspended.

7. Use Eq. (5–45) and the accompanying coefficients to estimate the removal rates for BOD and TSS at average and peak flow.

a. At average flow:

$$\text{BOD removal} = \frac{t}{a+bt} = \frac{2.42}{0.018 + (0.020)(2.42)} = 36\%$$

$$\text{TSS removal} = \frac{t}{a+bt} = \frac{2.42}{0.0075 + (0.014)(2.42)} = 58\%$$

b. At peak flow:

$$\text{BOD removal} = \frac{t}{a+bt} = \frac{0.97}{0.018 + (0.020)(0.97)} = 26\%$$

$$\text{TSS removal} = \frac{t}{a+bt} = \frac{0.97}{0.0075 + (0.014)(0.97)} = 46\%$$

Characteristics and Quantities of Solids (Sludge) and Scum

Typical values of specific gravity and solids concentration of solids (sludge) and scum removed from primary sedimentation tanks are presented in Table 5–21. Scum consists of a variety of floatable materials, and solids concentrations vary widely. In primary sedimentation tanks used in activated-sludge plants, provision may be required for handling the excess activated sludge that may be discharged into the influent of the primary tanks for

Table 5–21

Typical values of specific gravity and solids concentration of solids and scum removed from primary sedimentation tanks

Type of solids (sludge)	Specific gravity	Solids concentration, %[a] Range	Solids concentration, %[a] Typical
Primary only:			
Medium strength wastewater	1.03	4–12	6
From combined sewer system	1.05	4–12	6.5
Primary and waste activated sludge	1.03	2–6	3
Primary and trickling filter humus sludge	1.03	4–10	5
Scum	0.95	*b*	—

[a] Percent dry solids.
[b] Range is highly variable.

settlement and consolidation with the primary sludge. For treatment plants where waste-activated sludge is returned to the primary sedimentation tanks, the primary sedimentation tanks should include provisions for light flocculent solids of 98 to 99.5 percent moisture and for concentrations ranging from 1500 to 10,000 mg/L in the influent mixed liquor.

The volume of solids produced in primary settling tanks must be known or estimated so that these tanks and subsequent solids pumping, processing, and disposal facilities can be properly designed. The solids volume will depend on (1) the characteristics of the untreated wastewater, including strength and freshness; (2) the period of sedimentation and the degree of purification to be effected in the tanks; (3) the condition of the deposited solids, including specific gravity, water content, and changes in volume under the influence of tank depth or mechanical solids-removal devices; and (4) the period between solids-removal operations. Additional information on the characteristics and quantities of solids produced during primary sedimentation and other treatment operations and processes is provided in Chap. 13.

5–7 HIGH-RATE CLARIFICATION

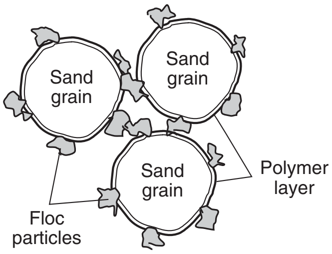
High-rate clarification employs physical/chemical treatment and utilizes special flocculation and sedimentation systems to achieve rapid settling. The essential elements of high-rate clarification are enhanced particle settling and the use of inclined plate or tube settlers. Advantages of high-rate clarification are (1) units are compact and thus reduce space requirements, (2) start-up times are rapid (usually less than 30 min) to achieve peak efficiency, and (3) a highly clarified effluent is produced. Enhanced particle flocculation and high-rate clarification applications are discussed in this section. Inclined plate and tube settlers were discussed previously in Sec. 5–4.

Enhanced Particle Flocculation

Enhanced particle flocculation has been used in Europe for more than 15 y but has only been introduced relatively recently in the United States. In its most basic form, enhanced particle flocculation involves the addition of an inert ballasting agent (usually silica sand or recycled chemically conditioned sludge) and a polymer to a coagulated and partially flocculated suspension. The polymer appears to coat the ballasting particles and forms the "glue" that binds the chemical floc to the ballasted particles (see Fig. 5–54). After contact with the ballasting agent, the mixture is stirred gently in a maturation tank that allows the floc particles to grow. The particles grow as the larger, faster-settling particles overtake and collide with slower-settling particles [see Fig. 5–11(b)]. The velocity gradient *G* for flocculation is important as a high gradient will cause a breakdown in the

Figure 5–54

Schematic of microsand ballasted floc particles. Polymer layer is used to absorb chemical flocs onto sand grains. (Adapted from Kruger.)



floc particles, and insufficient agitation will inhibit floc formation. Velocity gradients for enhanced particle settling of wastewater generally range from 200 to 400 s^{-1}.

Analysis of Ballasted Particle Flocculation and Settling

The settling velocity of the ballasted particle is increased, when compared to an unballasted floc particle, by (1) increasing the density of the particle, (2) decreasing the coefficient of drag and increasing the Reynolds number, and (3) decreasing the shape factor through the formation of more dense spherical-shaped particles [see Eq. (5–18)]. The ballasted floc particles appear to be more spherical than the floc particles alone. In effect, ballasted flocculent particles settle with a velocity closer to that of a discrete particle than that of flocculent particles that have very high shape factors. The comparative settling velocities of ballasted particles compared to other particles are illustrated in Example 5–10.

EXAMPLE 5–10 Calculation of Settling Velocities of Ballasted Floc and Other Particles

Determine the settling velocities at 20°C, for a ballasted floc particle, a spherical particle, and an irregularly shaped untreated wastewater particle, having the characteristics tabulated below. Use the alternative computation method involving the multiplication of the coefficient of drag times the shape factor.

	Particle type		
Particle characteristics	Ballasted floc	Spherical	Wastewater particle
Equivalent diameter, μm	200	150	500
Specific gravity	2.6	2.65	1.4
Shape factor, ϕ	2.5	1	18

Solution

Using Eq. (5–22) and the same computation procedure as Example 5–5, calculate the particle settling velocity for a ballasted floc particle and compare to spherical and wastewater floc particles.

1. Determine the terminal settling velocity for the ballasted floc particle.

$$\upsilon_p = \frac{g(sg_p - 1)d_p^2}{18\nu}$$

$$d_p = 200\ \mu\text{m} = 200 \times 10^{-6}\ \text{m}$$

$$\nu = 1.003 \times 10^{-6}\ \text{m}^2/\text{s at } 20°\text{C (from Appendix C)}$$

$$\upsilon_p = \frac{(9.81\ \text{m/s}^2)(2.6 - 1)(200 \times 10^{-6}\ \text{m})^2}{18(1.003 \times 10^{-6}\ \text{m}^2/\text{s})}$$

$$= 0.0348\ \text{m/s}$$

2. Check the Reynolds number (Eq. 5–20).

$$N_R = \frac{v_p d_p}{\nu} = \frac{(0.0348\ \text{m/s})(200 \times 10^{-6}\ \text{m})}{(1.003 \times 10^{-6}\ \text{m}^2/\text{s})} = 6.9$$

Because the Reynolds number > 1.0, Newton's Law [Eq. (5–18)] must be used to determine the settling velocity in the transition region (see Fig. 5–20). Follow the iterative procedure described in Example 5–5.

3. For the first assumed settling velocity, use the Stokes' Law settling velocity calculated above. Using the resulting Reynolds number, also determined previously, compute the drag coefficient.

$$C_d = \frac{24}{N_R} + \frac{3}{\sqrt{N_R}} + 0.34 = \frac{24}{6.9} + \frac{3}{\sqrt{6.9}} + 0.34 = 4.96$$

4. Use the drag coefficient in Newton's equation to determine the particle settling velocity in Eq. (5–21), which incorporates the shape factor.

$$v_p = \sqrt{\frac{4g(sg - 1)d}{3C_d\phi}} = \sqrt{\frac{4(9.81\ \text{m/s}^2)(2.6 - 1)(200 \times 10^{-6}\ \text{m})}{(3)(4.96)(2.5)}}$$
$$= 0.018\ \text{m/s}$$

Because the initial assumed settling velocity (0.035 m/s) does not equal the Newton's equation settling velocity (0.018 m/s), additional iterations are necessary.

5. For a following iteration, assume a settling velocity value of 0.012 m/s, and calculate the Reynolds number (the assumed value is based on several trial iterations for various velocities). Use the Reynolds number to determine the drag coefficient, and use the drag coefficient in Newton's equation to find the settling velocity.

$$N_R = \frac{(0.012\ \text{m/s})(200 \times 10^{-6}\ \text{m})}{(1.003 \times 10^{-6}\ \text{m}^2/\text{s})} = 2.39$$

$$C_d = \frac{24}{2.39} + \frac{3}{\sqrt{2.39}} + 0.34 = 12.3$$

$$v_p = \sqrt{\frac{4(9.81\ \text{m/s}^2)(2.6 - 1)(200 \times 10^{-6}\ \text{m})}{(3)(12.3)(2.5)}} = 0.012\ \text{m/s}$$

The assumed and calculated settling velocities (0.012 m/s) are in agreement.

6. Using the same computational procedure, calculate the settling velocities for the spherical and wastewater particles. The results of the calculations for the three particle settling velocities are summarized below:

Ballasted floc particle = 0.012 m/s = 43 m/h
Spherical particle = 0.0164 m/s = 59 m/h
Wastewater particle = 0.002 m/s = 7.2 m/h

Comment Although the particle settling velocities vary widely because of the particle characteristics (specific gravity, shape, and size), as illustrated in this example, by forming a ballasted floc particle the settling velocity can be enhanced significantly. Increasing the size and density of the wastewater particles by various means of ballasting is one of the premises of high-rate clarification. Because an iterative process is needed to calculate settling velocities in the transition zone, a spreadsheet program can expedite the computation process by allowing several assumed velocities to be tried to effect closure with the computed velocity.

Process Application

Three basic types of process used for high-rate clarification: (1) ballasted flocculation with lamella plate clarification, (2) three-stage flocculation with lamella plate clarification, and (3) dense-solids flocculation/clarification with lamella plate clarification are illustrated on Fig. 5–55. Each of these processes can operate at high overflow rates that allow significant reduction in the physical size of the sedimentation units. A summary of the principal

Figure 5–55

High-rate clarification processes: (a) ballasted flocculation, (b) lamella plate clarification, and (c) dense-sludge.

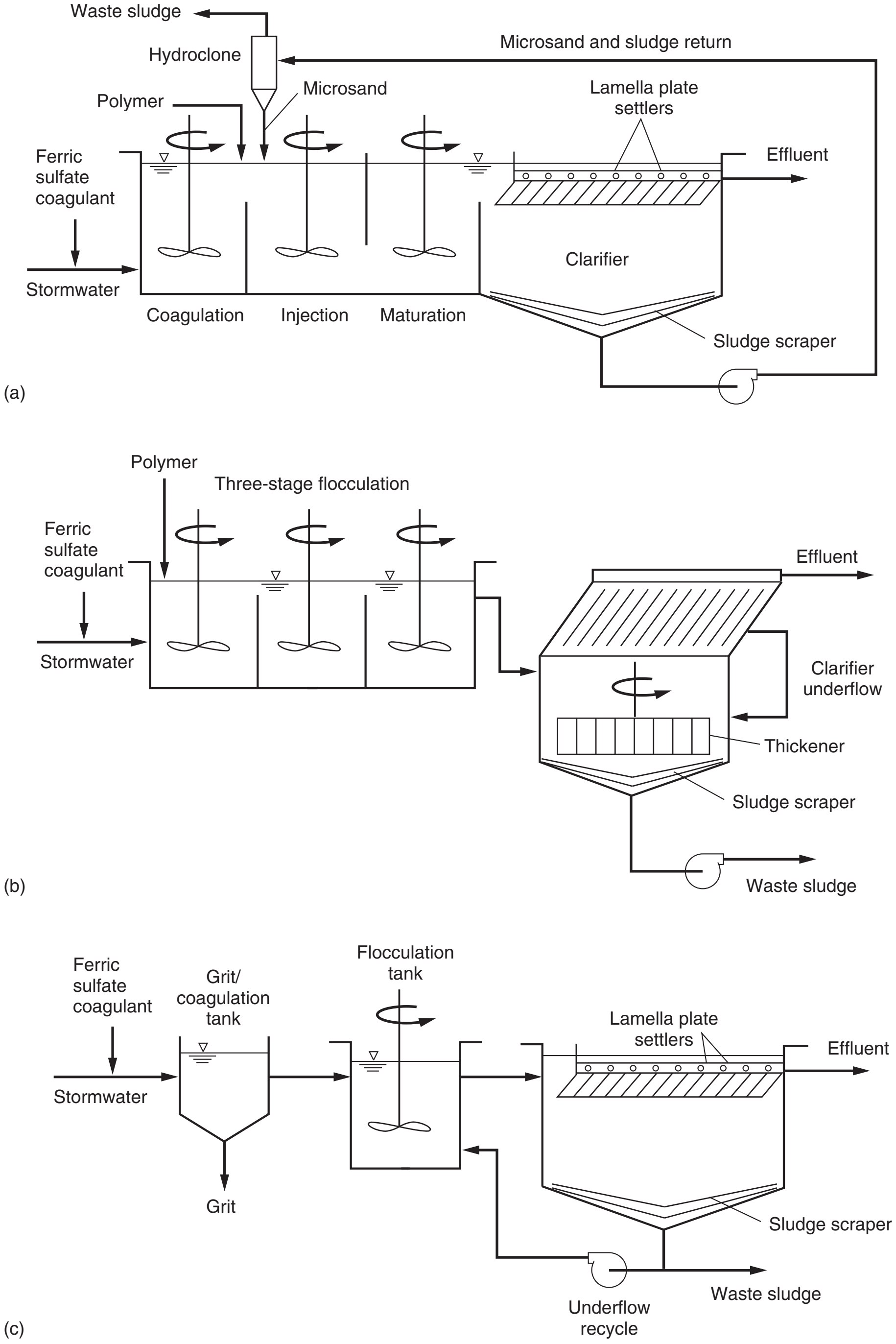


Table 5–22

Summary of features of high-rate clarification processes

Process	Features
Microsand ballasted flocculation and clarification [see Fig. 5–55(a)]	• Microsand provides nuclei for floc formation • Floc is dense and settles rapidly • Lamella clarification, when used, provides high rate settling in a small tank volume
Chemical addition, multi-stage flocculation, and lamella clarification [see Fig. 5–55(b)]	• Three-stage flocculation enhances floc formation • Lamella clarification provides high rate settling in a small tank volume
Air-mixing zone with grit separation and chemical addition followed by two-stage flocculation with chemically-conditioned recycled sludge, and lamella clarification [see Fig. 5–55(c)]	• Settled sludge solids are recycled to accelerate floc formation • Dense floc is formed that settles rapidly • Lamella clarification provides high rate settling in a small tank volume

features of each high-rate clarification process is presented in Table 5–22. Applications for high-rate clarification include (1) providing advanced primary treatment, (2) treating wet-weather flows and combined sewer overflows, (3) treating waste filter backwash water, and (4) treating return flows from solids-processing facilities. Ranges of overflow rates and BOD and TSS removals for treating wet-weather flows (domestic wastewater plus infiltration/inflow) are reported in Table 5–23 (Sawey, 1998).

Ballasted Flocculation. Ballasted flocculation employs a proprietary process, shown on Fig. 5–55(a), in which a flocculation aid and a ballasting agent (typically silica microsand) are used to form dense microfloc particles. The resulting floc particles are thus

Table 5–23

Ranges of overflow rates and BOD and TSS removals from high-rate clarification processes treating wet weather flows

Parameter/Process	Ballasted flocculation	Lamella plate clarification	Dense sludge
Overflow rates			
Low, $m^3/m^2 \cdot d$	1200–2900	880	2300
($gal/min \cdot ft^2$)	(20–50)	(15)	(40)
Medium, $m^3/m^2 \cdot d$	1800–3500	1200	2900
($gal/min \cdot ft^2$)	(30–60)	(20)	(50)
High, $m^3/m^2 \cdot d$	2300–4100	1800	3500
($gal/min \cdot ft^2$)	(40–70)	(30)	(60)
BOD removals, %			
At low overflow rates	35–50	45–55	25–35
At medium overflow rates	40–60	35–40	40–50
At high overflow rates	30–60	35–40	50–60
TSS removals, %			
At low overflow rates	70–90	60–70	80–90
At medium overflow rates	40–80	65–75	70–80
At high overflow rates	30–80	40–50	70–80

"ballasted" and settle rapidly. The treatment system consists of three compartments or zones: a mixing zone, maturation zone, and settling zone. Depending on the manufacturer of the process equipment, separate, serial compartments can be used to perform the process functions, or the functions can be combined in a single vessel. Either lamella plate settling or conventional gravity clarification can be used.

Typically, screened wastewater is introduced to the ballasted flocculation reactor where a chemical coagulant (typically an iron salt) is injected to destabilize the solids. The wastewater then enters a mixing zone where microsand and polymer are injected to maximize the efficiency of flocculation and enhance settling of suspended solids. In the mixing zone, the polymer acts as a bonding agent for adhering the destabilized solids to the microsand. The maturation zone follows and is used to keep the solids in suspension while floc particles continue to develop and grow. Once developed, the ballasted floc particles settle rapidly to the bottom of the clarifier. Sand and floc particles removed from the clarified water are pumped to a cyclone separator (hydroclone) for separation of the sand. The separated sand is returned to the injection tank, and solids from the hydroclone are sent to the biosolids processing facilities. The microsand usually ranges in size from 100 to 150 mm for treating wastewater and combined wastewater flows and has a specific gravity greater than 2.6 to enhance settling.

Lamella Plate Clarification. Lamella plate clarification uses chemical addition followed by three-stage flocculation and a lamella plate clarifier [see Fig. 5–55(b)]. Coagulant and polymer are injected into the influent wastewater prior to entrance into the flocculation zone. When chemically conditioned wastewater passes through each of the three flocculation zones, the mixing energy gradient is decreased as the wastewater proceeds from one stage to the next. The chemically conditioned/flocculated wastewater then passes to the lamella clarifier for solids separation. A portion of the clarifier underflow can be recycled to the influent of the process to enhance settling, or the entire underflow can be sent to a thickening tank and the solids processing facilities.

Dense-Sludge Process. The dense-sludge system is a proprietary process and differs from ballasted flocculation in that recycled chemically conditioned solids are used to form microfloc particles with the incoming wastewater instead of microsand. As shown on Fig. 5–55(c), the influent wastewater enters an air-mixing zone where grit separation occurs and coagulant (usually ferric sulfate) is injected. After mixing, the wastewater flows into the first stage of a two-stage flocculation tank where polymer is added together with chemically conditioned, recirculated solids. Recirculated solids accelerate the flocculation process and ensure the formation of dense, homogeneous floc particles. In the second stage of flocculation, grease and scum begin separating and are removed. Flow from the flocculation tank enters a presettling zone and then passes into a lamella plate settler. Most of the suspended flocculated solids are separated directly in the presettling zone; the residual flocculated particles are removed in the lamella settler. A portion of the settled solids is recirculated, and the remainder is sent to the solids processing and disposal system.

5–8 FLOTATION

Flotation is a unit operation used to separate solid or liquid particles from a liquid phase. Separation is brought about by introducing fine gas (usually air) bubbles into the liquid phase. The bubbles attach to the particulate matter, and the buoyant force of the combined particle and gas bubbles is great enough to cause the particle to rise to the surface. Particles that have a higher density than the liquid can thus be made to rise. The rising of particles with lower density than the liquid can also be facilitated (e.g., oil suspension in water).

In wastewater treatment, flotation is used principally to remove suspended matter and to concentrate biosolids (see Chap. 14). The principal advantages of flotation over sedimentation are that very small or light particles that settle slowly can be removed more completely and in a shorter time. Once the particles have been floated to the surface, they can be collected by a skimming operation.

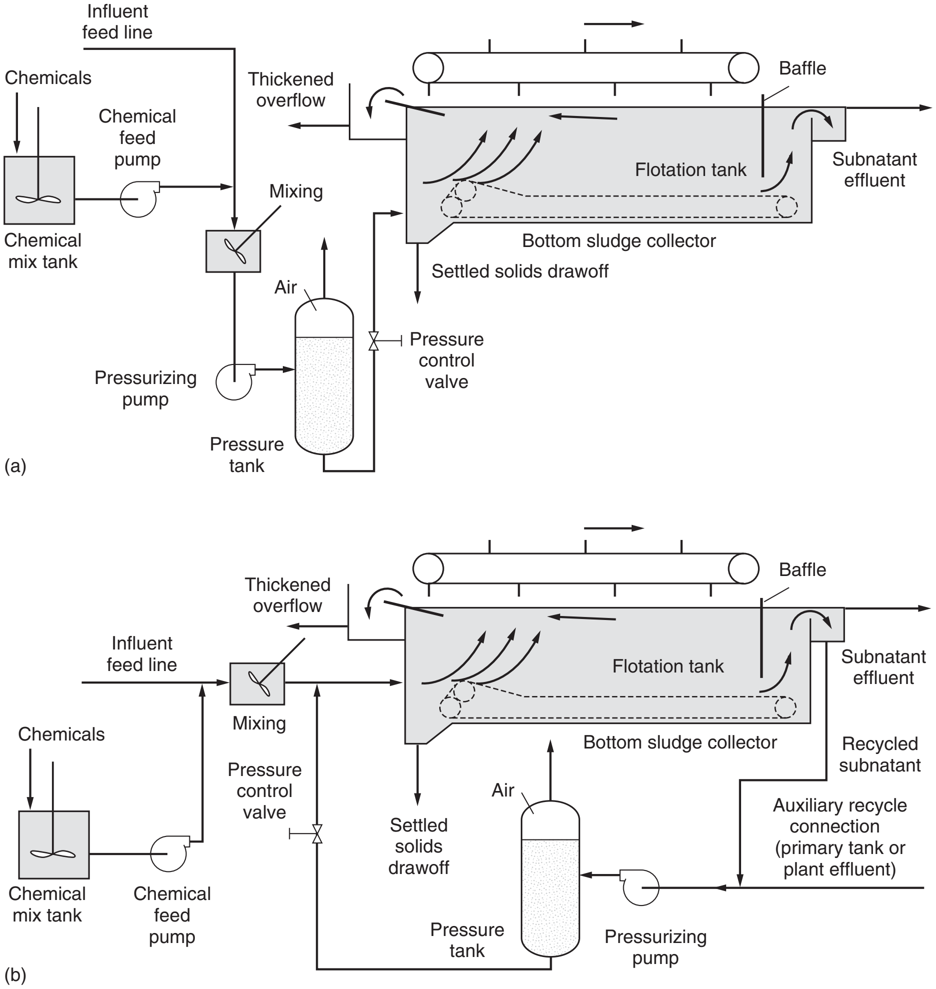
Description

The present practice of flotation as applied to wastewater treatment is confined to the use of air as the flotation agent. Air bubbles are added or caused to form by (1) injection of air while the liquid is under pressure, followed by release of the pressure (dissolved-air flotation), and (2) aeration at atmospheric pressure (dispersed-air flotation). In these systems, the degree of removal can be enhanced through the use of various chemical additives. In municipal wastewater treatment, dissolved-air flotation is frequently used, especially for thickening of waste biosolids.

Dissolved-Air Flotation. In dissolved-air flotation (DAF) systems, air is dissolved in the wastewater under a pressure of several atmospheres, followed by release of the pressure to the atmospheric level (see Fig. 5–56). In small pressure systems, the entire flow

Figure 5–56

Schematic of dissolved-air flotation systems: (a) without recycle in which the entire flow is passed through the pressurizing tank and (b) with recycle in which only the recycle flow is pressurized. The pressurized flow is mixed with the influent before being released into the flotation tank.



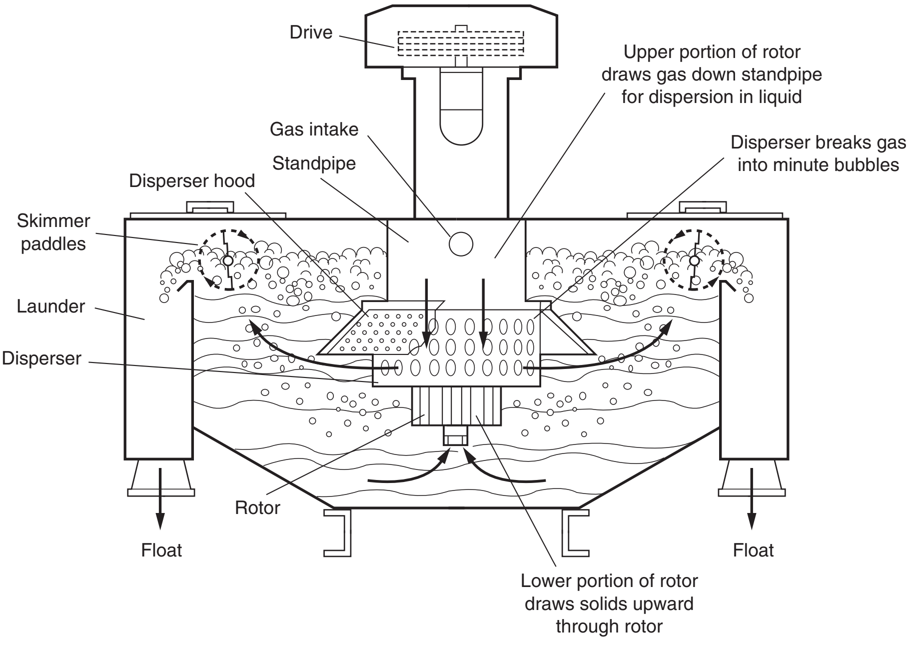
may be pressurized by means of a pump to 275 to 350 kPa (40 to 50 $lb_f/in.^2$ gage) with compressed air added at the pump suction [see Fig. 5–56(a)]. The entire flow is held in a retention tank under pressure for several minutes to allow time for the air to dissolve. It is then admitted through a pressure-reducing valve to the flotation tank, where the air comes out of solution in very fine bubbles.

In the larger units, a portion of the DAF effluent (15 to 120 percent) is recycled, pressurized, and semisaturated with air [see Fig. 5–56(b)]. The recycled flow is mixed with the unpressurized main stream just before admission to the flotation tank, with the result that the air comes out of solution in contact with particulate matter at the entrance to the tank. Pressure types of units have been used mainly for the treatment of industrial wastes and for the concentration of solids.

Dispersed-Air Flotation. Dispersed-air (sometimes referred to as induced-air) flotation is seldom used in municipal wastewater treatment, but it is used in industrial applications for the removal of emulsified oil and suspended solids from high-volume waste or process waters. In dispersed-air flotation systems, air bubbles are formed by introducing the gas phase directly into the liquid phase through a revolving impeller. The spinning impeller acts as a pump, forcing fluid through disperser openings and creating a vacuum in the standpipe (see Fig. 5–57). The vacuum pulls air (or gas) into the standpipe and thoroughly mixes it with the liquid. As the gas/liquid mixture travels through the disperser, a mixing force is created that causes the gas to form very fine bubbles. The liquid moves through a series of cells before leaving the unit. Oil particles and suspended solids attach to the bubbles as they rise to the surface. The oil and suspended solids gather in dense froth at the surface and are removed by skimming paddles. The advantages of a dispersed-air flotation system are (1) compact size, (2) lower capital cost, and (3) capacity to remove relatively free oil and suspended solids. The disadvantages of induced-air flotation include higher connected power requirements than the pressurized system, performance is dependent on strict hydraulic control, and less flocculation flexibility. The quantities of float

Figure 5–57

Dispersed-air flotation unit. Air is induced and dispersed into the liquid by pumping action of the inductors. (Courtesy of Eimco.)



skimmings are significantly higher than the pressurized unit: 3 to 7 percent of the incoming flow as compared to less than 1 percent for dissolved-air systems (Eckenfelder, 2000).

Chemical Additives. Chemicals are commonly used to aid the flotation process. These chemicals, for the most part, function to create a surface or a structure that can easily absorb or entrap air bubbles. Inorganic chemicals, such as the aluminum and ferric salts and activated silica, can be used to bind the particulate matter together and, in so doing, create a structure that can easily entrap air bubbles. Various organic polymers can be used to change the nature of either the air-liquid interface or the solid-liquid interface, or both. These compounds usually collect on the interface to bring about the desired changes.

Design Considerations for Dissolved-Air Flotation Systems

Because flotation is very dependent on the type of surface of the particulate matter, laboratory and pilot-plant tests should be performed to yield the necessary design criteria. Factors that must be considered in the design of flotation units include the concentration of particulate matter, quantity of air used, the particle-rise velocity, and the solids loading rate. In the following analysis, dissolved-air flotation is discussed because it is the method most commonly used.

The performance of a dissolved-air flotation system depends primarily on the ratio of the volume of air to the mass of solids (A/S) required to achieve a given degree of clarification. The ratio will vary with each type of suspension and must be determined experimentally using a laboratory flotation cell. A typical laboratory flotation cell is shown on Fig. 5–58. Procedures for conducting the necessary tests may be found in Higbie (1935), WEF (1988c), and Edzwald and Haarhoff (2012). Typical A/S ratios encountered in the thickening of solids and biosolids in wastewater treatment plants vary from about 0.005 to 0.060.

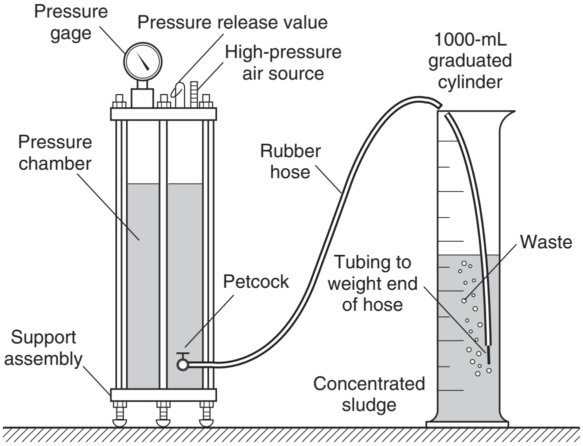
The relationship between the A/S ratio and the solubility of air, the operating pressure, and the concentration of solids for a system in which all the flow is pressurized is given in Eq. (5–47).

$$\frac{A}{S} = \frac{1.3 s_a (fP - 1)}{S_a} \tag{5–47}$$

where A/S = air to solids ratio, mL air/mg solids
s_a = air solubility, mL/L
f = fraction of air dissolved at pressure P, usually 0.5

Figure 5–58

Schematic of dissolved-air flotation test apparatus.



P = pressure, atm

$$= \frac{p + 101.35}{101.35} \text{ (SI units)}$$

$$= \frac{p + 14.7}{14.7} \text{ (U.S. customary units)}$$

p = gage pressure, kPa ($\text{lb}_\text{f}/\text{in.}^2$ gage)
S_a = influent suspended solids, g/m^3 (mg/L)

Temp., °C	0	10	20	30
s_a, mL/L	29.2	22.8	18.7	15.7

The corresponding equation for a system with only pressurized recycle is

$$\frac{A}{S} = \frac{1.3 s_a (fP - 1)R}{S_a Q} \tag{5–48}$$

where R = pressurized recycle, m^3/d (Mgal/d)
Q = mixed-liquor flow, m^3/d (Mgal/d)

In both of the foregoing equations, the numerator represents the weight of air and the denominator the weight of the solids. The factor 1.3 is the weight in milligrams of 1 mL of air and the term (1) within the brackets accounts for the fact that the system is to be operated at atmospheric conditions. The use of these equations is illustrated in Example 5–11. Additional information of the theory and application of flotation may be found in Eckenfelder (2000) and Edzwald and Haarhoff (2012).

The required area of the thickener is determined from a consideration of the rise velocity of the solids, 8 to 160 L/m²·min (0.2 to 4.0 gal/min·ft²), depending on the solids concentration, degree of thickening to be achieved, and the solids loading rate (see Table 14–20).

EXAMPLE 5–11 **Flotation Thickening of Activated-Sludge Mixed Liquor** Design a flotation thickener without and with pressurized recycle to thicken the solids in activated-sludge mixed liquor from 0.3 to about 4 percent. Assume that the following conditions apply:

1. Optimum A/S ratio = 0.008 mL/mg
2. Temperature = 20°C
3. Air solubility = 18.7 mL/L
4. Recycle-system pressure = 275 kPa
5. Fraction of saturation = 0.5
6. Surface-loading rate = 8 L/m²·min
7. Sludge flowrate = 400 m³/d

Solution (without recycle)

1. Compute the required pressure using Eq. (5–47).

$$\frac{A}{S} = \frac{1.3 s_a (fP - 1)}{S_a}$$

$$0.008 \text{ mL/mg} = \frac{1.3(18.7 \text{ mL/L})(0.5P - 1)}{(3000 \text{ mg/L})}$$

$$0.5\,P = 0.99 + 1$$

$$P = 3.98\ \text{atm} = \frac{p + 101.35}{101.35}$$

$$p = 302\ \text{kPa}\ (43.8\ \text{lb}_f/\text{in.}^2\text{gage})$$

2. Determine the required surface area.

$$A = \frac{(400\ \text{m}^3/\text{d})(10^3\ \text{L}/1\ \text{m}^3)}{(8\ \text{L/m}^2 \cdot \text{min})(1440\ \text{min/d})} = 34.7\ \text{m}^2$$

3. Check the solids loading rate.

$$\text{kg/m}^2 \cdot \text{d} = \frac{(400\ \text{m}^3/\text{d})(3000\ \text{g/m}^3)}{(34.7\ \text{m}^2)(10^3\ \text{g}/1\ \text{kg})} = 34.6\ \text{kg/m}^2$$

Solution (with recycle)

1. Determine pressure in atmospheres.

$$P = \frac{275 + 101.35}{101.35} = 3.73\ \text{atm}$$

2. Determine the required recycle rate using Eq. (5–48).

$$\frac{A}{S} = \frac{1.3 s_a (fP - 1)R}{S_a Q}$$

$$0.008\ \text{mL/mg} = \frac{1.3(18.7\ \text{mL/L})[0.5(3.73) - 1]R}{(3000\ \text{mg/L})(400\ \text{m}^3/\text{d})}$$

$$R = 461.9\ \text{m}^3/\text{d}$$

3. Determine the required surface area.

$$A = \frac{(461.9\ \text{m}^3/\text{d})(10^3\ \text{L/m}^3)}{(8\ \text{L/m}^2 \cdot \text{min})(1440\ \text{min/d})} = 40.1\ \text{m}^2$$

Comment Alternatively, the recycle flowrate could have been set and the pressure determined. In an actual design, the costs associated with the recycle pumping, pressurizing systems, and tank construction can be evaluated to find the most economical combination.

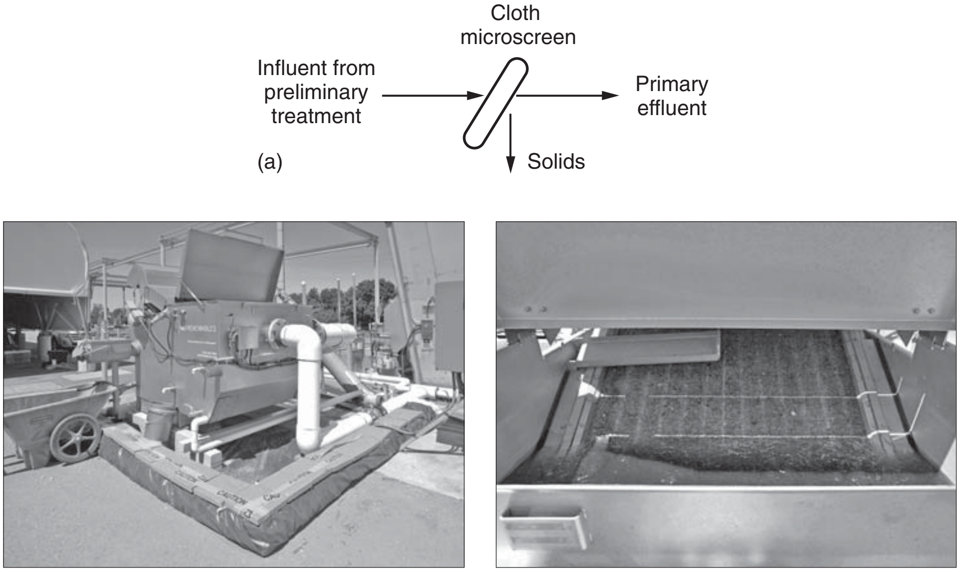
5–9 NEW APPROACHES FOR PRIMARY TREATMENT

In addition to the well-established physical treatment processes, as described previously in this chapter, a number of other physical processes have been or are being tested for various primary treatment applications. Three processes which have been demonstrated successfully are introduced in the following discussion. Those processes are (1) microscreening of raw wastewater, (2) charged bubble flotation, and (3) primary effluent filtration.

Given the current emphasis on energy conservation and recovery, three applications for these technologies are as follows. The first application is for the reduction of the organic loading rate to the biological treatment process to reduce the energy requirement for the oxidation of the carbonaceous organic material. The solids removed would be sent to the solids processing facility, most commonly to an anaerobic digester for the recovery of energy. In the second application, these technologies would be used to achieve load equalization to the biological treatment process by diverting organic material during peak energy demand periods and shifting the load to the early morning hours. In the third

Figure 5–59

Cloth screen used for the filtration of untreated wastewater: (a) schematic, (b) view of cloth screen filter unit with front cover open, and (c) excess water being removed by gravity from solids accumulated on the belt.



(b)

(c)

application, the diverted organic matter could be sent to a fermenter for the production of volatile fatty acids, which in turn could be used for phosphorus removal.

Microscreening of Raw Wastewater

Two cloth screens have been developed that can be used to filter raw wastewater following the removal of coarse solids. Views of a typical screen are shown on Fig. 5–59. Operationally, as shown on Fig. 5–59(a) untreated wastewater is introduced above the rotating screen which moves up the incline. Initially, when the clean screen comes in contact with the wastewater, the filtration rate is essentially the same as the clean water filtration rate. As the screen moves upward along the incline, the solids accumulate and auto-filtration (i.e., the removed solids form a filter) occurs in addition to screening. As the screen moves out of the water, excess water is removed gravimetrically [see Fig. 5–59(c)].

The partially dewatered accumulated solids are removed as the screen passes over the upper roller. Depending on the nature of the raw wastewater, water jets, located above the upper roller, are used to enhance the removal of the accumulated solids. The solids removed from the roller can be processed further (e.g., thickened) or fed to a digester directly. The performance data for the removal of BOD (25 to 35 percent) and TSS (60 to 70 percent) are similar to or somewhat better than that achieved with primary sedimentation. The fact that the screen alters the particle size distribution of the solids to be treated and the small footprint are significant relative to conventional primary clarification.

Charged Bubble Flotation

The principal components of the charged bubble flotation (CBF) process are illustrated on Fig. 5–60. The CBF process uses an externally generated suspension of micron-sized (about 7 to 50 μm) bubbles in water, achieving 40 to 50 percent volumetric air content. Each of the bubbles is coated (encapsulated) with a thin soap film made from an electrically charged surfactant, either anionic or cationic depending on the application. The charged bubbles provide a large interfacial area for the absorption of charged and hydrophobic molecules

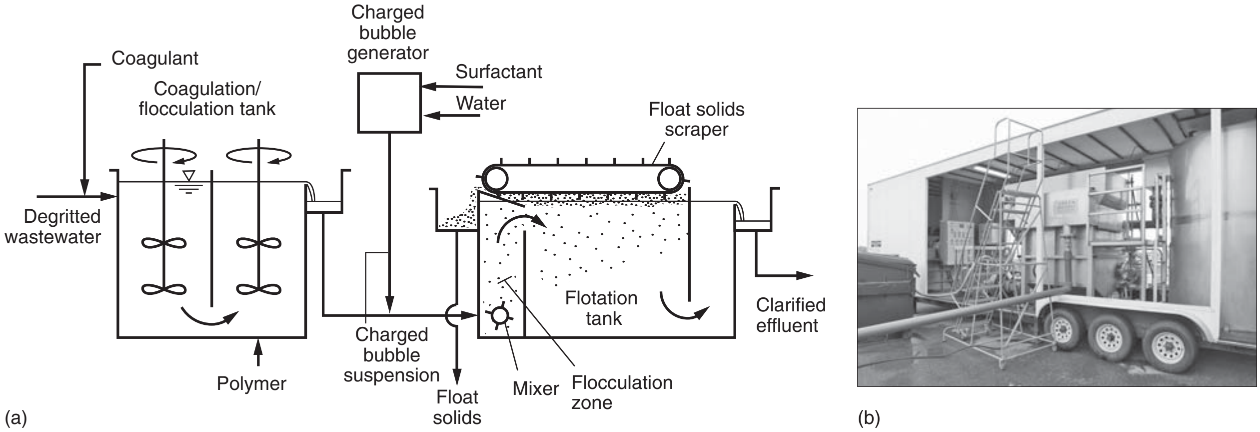


Figure 5–60

Charged bubble flotation process: (a) schematic flow diagram and (b) view of trailer mounted pilot-scale test unit.

(Jauregi and Varley, 1999). In the literature, these micron-sized bubbles in suspension have been called *colloidal gas aphrons,* a term coined by the inventor (Sebba, 1987). Because the air bubbles are charged, they do not coalesce and are attracted readily to oppositely charged flocculated wastewater solids. For the flotation of screened raw wastewater, a coagulant and polymer are used, as the natural charge density of the solids is too low for effective flotation with CBF.

Operationally, coagulant is dosed to the wastewater and conditioned for a few minutes. Polymer and charged bubble suspension are then introduced in the flocculation chamber in the flotation tank where they are contacted with the solids in the wastewater. The flocculated wastewater is introduced into the flotation tank where the float rises to the surface and is removed by skimming, and the clarified underflow passes underneath a baffle and over a weir into a launder. Typical performance data for the removal of BOD and TSS from raw wastewater are 50 to 70 and 70 to 99 percent, respectively, depending on the type of conditioning chemicals used and the corresponding dosages.

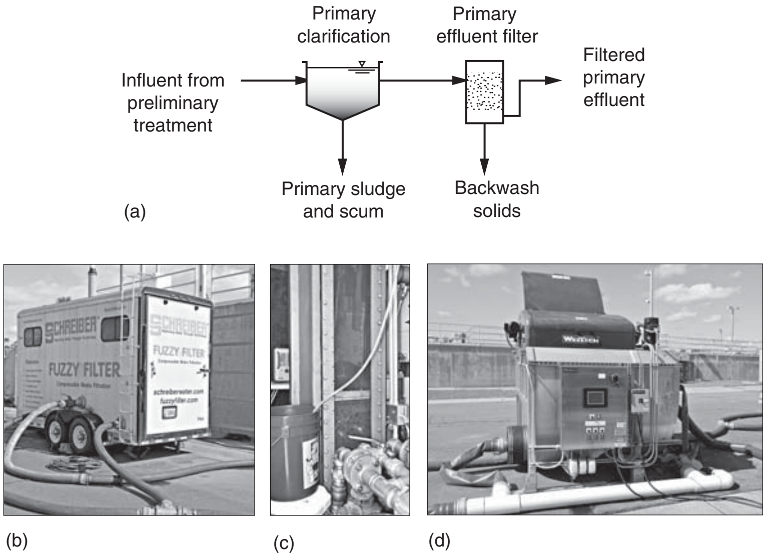
The CBF process, used for screened raw wastewater, can replace three unit processes: grit removal (except for the largest and densest particles), primary clarification, and primary scum handling. The CBF process can also be employed as an alternative to chemically enhanced primary clarification or primary effluent filtration. Other applications for the CBF process are in secondary clarification, waste activated sludge thickening, clarification of algae-laden waters (especially as pretreatment for tertiary filtration), and digested sludge thickening. The principal advantages of the CBF process include smaller footprint (less than one-fifth the size of a conventional primary clarifier), low power requirement, the ability to handle high concentrations of suspended solids (up to 15,000 mg/L), and high solids separation efficiency.

Primary Effluent Filtration

Primary effluent filtration (PEF) involves the filtering of the effluent from primary sedimentation tanks. First studied and introduced in 1980 (Matsumoto et al., 1980 and 1982; England et al., 1994) and one or more full-scale plants were built in the early 1980s. The process was effective but did not catch on because market energy costs were so low at the time that the return on investment was not favorable. More recently, the process has been

Figure 5–61

Primary effluent filtration:
(a) schematic flow diagram,
(b) and (c) view of test filters.
Information on the Schreiber and WesTech test filters can be found in Sec. 11–5 in Chap. 11.



investigated using new filter technologies including the Fuzzy Filter (compressible filter medium) and the WesTech disk cloth filter. Both filters have been tested successfully (see Fig. 5–61). Typical performance data for the removal of BOD and TSS are 25 to 35 and 45 to 75 percent, respectively. It is interesting to note that the backwash water percentage is extremely low as compared to the filtration of settled secondary effluent. The reason secondary effluent is more difficult to filter is due to the presence of extracellular polymers produced during biological treatment that are associated with the residual solids that do not settle and must be removed by filtration.

5–10 GAS LIQUID MASS TRANSFER

In the unit separation processes described in the previous sections, change is brought about by the force of gravity. There are, however, a number of important physical unit processes used for the treatment of wastewater involving the transfer of material (e.g., mass) from one phase to another, as summarized in Table 5–24. Physical unit processes such as aeration (Secs. 5–11, 8–9, and 16–4), carbon adsorption (Sec. 11–8), gas stripping (Secs. 11–8, 15–5 and 16–4), reverse osmosis (Sec. 11–6) and ion exchange (Sec. 11–10) are examples of processes that involve mass transfer. The transfer of gas to or from a liquid is introduced in this section. Following a brief history of the development of gas transfer theories, the two-film theory of gas transfer is introduced, and the basic concepts are applied to absorption and desorption of a gas from a liquid. Aeration systems are introduced in the following section.

Historical Development of Gas Transfer Theories

Over the past 50 y a number of mass transfer theories have been proposed to explain the mechanism of gas transfer across gas-liquid interfaces. The simplest and most commonly used is the two-film theory proposed by Lewis and Whitman (1924). The penetration model proposed by Higbie (1935) and the surface-renewal model proposed by Danckwerts (1951) are more theoretical and take into account more of the physical phenomena involved. The two-film theory remains popular because in more than 95 percent of the situations encountered, the results obtained are essentially the same as those obtained with the more complex theories. Even in the 5 percent where there is some disagreement between the

Table 5–24

Principal applications of mass transfer operations and processes in wastewater treatment

Type of reaction	Phase equilibria	Application
Absorption	Gas → liquid	Addition of gases to water (e.g., O_2, O_3, CO_2, Cl_2, SO_2), NH_3 scrubbing in acid
Adsorption	Gas → solid	Removal of organics with activated carbon
	Liquid → solid	Removal of organics with activated carbon, dechlorination
Desorption	Solid → liquid	Sediment scrubbing
	Solid → gas	Reactivation of spent activated carbon
Drying (evaporation)	Liquid → gas	Drying of sludges
Gas stripping (also known as desorption)	Liquid → gas	Removal of gases (e.g., CO_2, O_2, H_2S, NH_3, volatile organic compounds, NH_3 from digester supernatant)
Ion exchange	Liquid → solid	Selective removal of chemical constituents, demineralization

[a] Adapted from Crittenden, et al. (2012).

two-film theory and other theories, it is not clear which approach is more correct. Thus, the focus of the following discussion is on the two-film theory.

The Two-Film Theory of Gas Transfer

The two-film theory is based on a physical model in which two films exist at the gas-liquid interface, as shown on Fig. 5–62. Two conditions are shown on Fig. 5–62: (a) "absorption,"

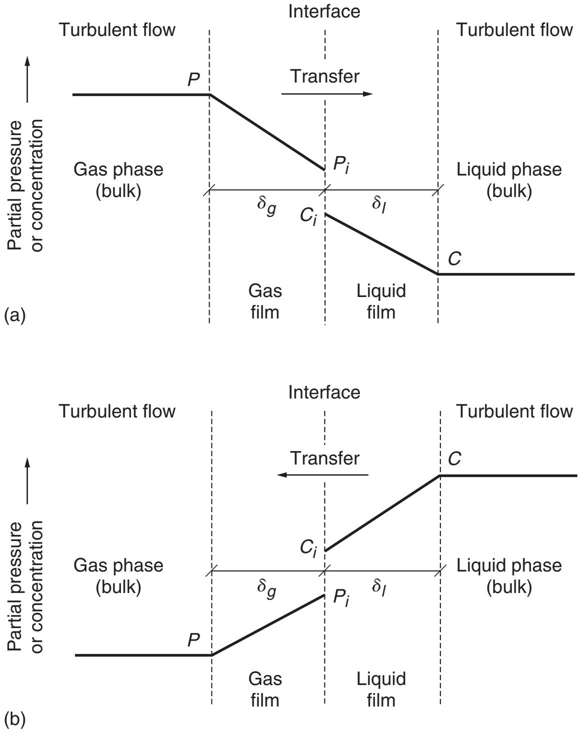


Figure 5–62

Definition sketch for the two-film theory of gas transfer: (a) absorption and (b) desorption.

in which a gas is transferred from the gas phase to the liquid phase, and (b) "desorption," in which a gas is transferred out of the liquid phase into the gas phase. The two films, one liquid and one gas, provide the resistance to the passage of gas molecules between the bulk-liquid and the bulk-gaseous phases. It is very important to note that in the application of the two-film theory, it is assumed that the concentration and partial pressure in both the bulk-liquid and bulk-gas phase are uniform (i.e., mixed completely).

Steady-State Mass Transfer. Under steady-state conditions, the rate of mass transfer of a gas through the gas film must be equal to the rate transfer through the liquid film. Using Fick's first law [Eq. (1–54)], the mass flux for each phase for absorption (gas addition) is written as follows (Lewis and Whitman, 1924):

$$r = k_G(P_G - P_i) = k_L(C_i - C_L) \tag{5–49}$$

where r = rate of mass transferred per unit area per unit time
k_G = gas film mass transfer coefficient
P_G = partial pressure of constituent A in the bulk of the gas phase
P_i = partial pressure of constituent A at the interface in equilibrium with concentration C_i of constituent A in liquid
k_L = liquid film mass transfer coefficient
C_i = concentration of constituent A at the interface in equilibrium with partial pressure P_i of constituent A in the gas
C_L = concentration of constituent A in the bulk liquid phase

It should be noted that the gas and liquid film mass transfer coefficients depend on the conditions at the interface. The terms $(P_G - P_i)$ and $(C_i - C_L)$ represent the driving force causing transfer in the gas and liquid phase, respectively. If the terms $(P_G - P_i)$ and $(C_i - C_L)$ are divided by their respective film thickness values (δ_G and δ_L), the driving force can be expressed in terms of unit thickness. Thus, the degree of mass transport can be enhanced by reducing the thickness of the film, depending on which is the controlling film.

Overall Mass Transfer Coefficients. Because it is difficult to measure the values of k_G and k_L at the interface, it is common to use overall coefficients K_G and K_L, depending on whether the resistance to mass transfer is on the gas or liquid side. If it is assumed that essentially all of the resistance to mass transfer is caused by the liquid film, then the rate of mass transfer can be defined as follows in terms of the overall liquid mass transfer coefficient:

$$r = K_L(C_s - C_L) \tag{5–50}$$

where r = rate of mass transferred per unit area per unit time
K_L = overall liquid mass transfer coefficient
C_s = concentration of constituent A at the interface in equilibrium with the partial pressure of constituent A in the bulk gas phase
C_L = concentration of constituent A in the bulk liquid phase

If the two expressions given by Eqs. (5–50) and (5–49) above are equated, the following relationship can be derived between the overall liquid mass transfer coefficient and the gas and liquid film coefficients:

$$r = K_L(C_s - C_L) = k_G(P_G - P_i) = k_L(C_i - C_L) \tag{5–51}$$

Because it was assumed that essentially all of the resistance to mass transfer is caused by the liquid film, the following relationships, based on Henry's law (see Chap. 2), must apply at the interface:

$$P_G = HC_s \quad \text{and} \quad P_i = HC_i$$

It will now be noted that the overall driving force $(C_s - C_L)$ in Eq. (5–51) can be written as

$$(C_s - C_L) = (C_s - C_i) + (C_i - C_L) \tag{5–52}$$

Substituting for P_G and P_i in Eq. (5–51) and combining Eq. (5–51) and Eq. (5–52), the following relationship is obtained where the liquid film controls the mass transfer:

$$\frac{r}{K_L} = \frac{r}{k_L} + \frac{r}{Hk_G} \quad \text{or} \quad \frac{1}{K_L} = \frac{1}{k_L} + \frac{1}{Hk_G} \tag{5–53}$$

In a similar manner it can be shown that the following relationship holds if the transfer of mass is controlled by the gas film:

$$\frac{1}{K_G} = \frac{1}{k_G} + \frac{H}{k_L} \tag{5–54}$$

The relationship between the overall liquid and gas phase transfer coefficients is

$$\frac{1}{K_L} = \frac{1}{K_G H} \tag{5–55}$$

It should be noted that in Eqs. (5–54) and (5–55) the overall transfer coefficients include the resistance to mass transfer offered by both the gas and liquid phases. The fact that the overall resistance to mass transfer is the sum of the gas and liquid phase resistances was first demonstrated by Lewis and Whitman (1924). Referring to Eq. (5–53), it is interesting to note that if the Henry's constant is large, then the liquid phase resistance will typically control the mass transfer process. For the transfer of gas molecules from the gas phase to the liquid phase, slightly soluble gases (e.g., O_2, N_2, and CO_2 in water) encounter the primary resistance to transfer from the liquid film, and very soluble gases (e.g., NH_3 in water) encounter the primary resistance to transfer from the gaseous film. Gases of intermediate solubility (e.g., H_2S in water) encounter significant resistance from both films.

Flux of a Slightly Soluble Gas. To estimate the flux of a slightly soluble gas from the gas to the liquid phase (liquid film controls transfer rate), Eq. (5–50) can be approximated by substituting C_t for C_L as follows:

$$r = K_L(C_s - C_t) \tag{5–56}$$

where r = rate of mass transferred per unit area per unit time, $ML^{-2}T^{-1}$
K_L = overall liquid mass transfer coefficient, LT^{-1}
C_t = concentration in liquid bulk phase at time t, ML^{-3}
C_s = concentration in equilibrium with gas as given by Henry's law, ML^{-3}

The corresponding rate of mass transfer per unit volume per unit time is obtained by multiplying Eq. (5–56) by the area A and dividing by the volume V.

$$r_V = K_L\frac{A}{V}(C_s - C_t) = K_La(C_s - C_t) \tag{5–57}$$

where r_V = rate of mass transfer per unit volume per unit time, $ML^{-3}T^{-1}$
K_La = volumetric mass transfer coefficient, T^{-1}
A = area through which mass is transferred, L^2
V = volume in which constituent concentration is increasing, L^3
a = interfacial area for mass transfer per unit volume, A/V, L^{-1}

The term K_La, known as the volumetric mass transfer coefficient, depends on water quality and the type of aeration equipment and is unique for each situation. Numerical values for K_La are usually determined experimentally (see Sec. 5–11). Equation (5–57) is the basic relationship used in solving problems involving the addition of oxygen to water as in aeration, the removal of volatile organics from wastewater by bubbling air through the wastewater, and for the stripping of dissolved constituents such as ammonia from digested supernatant.

Absorption of Gases Under Turbulent Conditions

The application of the gas-liquid mass transfer relationship developed above will be illustrated by considering the absorption of a gas in a turbulent liquid (see Fig. 5–63a). Consider, for example, a storage basin open to the atmosphere with surface area *A* and depth *h*. If the concentration of dissolved oxygen in the basin is initially undersaturation, how long would it take for the oxygen concentration to increase by a given amount? The approach to this mass transfer problem can be outlined as follows.

First, a mass balance is written for the open basin as follows:

1. General word statement:

$$\begin{array}{c} \text{Rate of accumulation} \\ \text{of a gas within the} \\ \text{system boundary} \end{array} = \begin{array}{c} \text{rate of flow of} \\ \text{a gas into the} \\ \text{system boundary} \end{array} - \begin{array}{c} \text{rate of flow of} \\ \text{a gas out of the} \\ \text{system boundary} \end{array} + \begin{array}{c} \text{amount of gas} \\ \text{absorbed through} \\ \text{system boundary} \end{array} \quad (5\text{–}58)$$

2. Simplified word statement:

$$\text{Accumulation} = \text{inflow} - \text{outflow} + \text{increase due to absorption} \quad (5\text{–}59)$$

3. Symbolic representation at equilibrium (refer to Fig. 5–62a):

$$\frac{dC}{dt}(V) = 0 - 0 + r_V V \quad (5\text{–}60)$$

where dC/dt = change in concentration with time, $\text{ML}^{-3}\text{T}^{-1}$, ($\text{g/m}^3\cdot\text{s}$)
V = volume in which constituent concentration is increasing, L^3, (m^3)
r_V = mass of constituent transferred per unit volume per unit time, $\text{ML}^{-3}\text{T}^{-1}$, ($\text{g/m}^3\cdot\text{s}$)

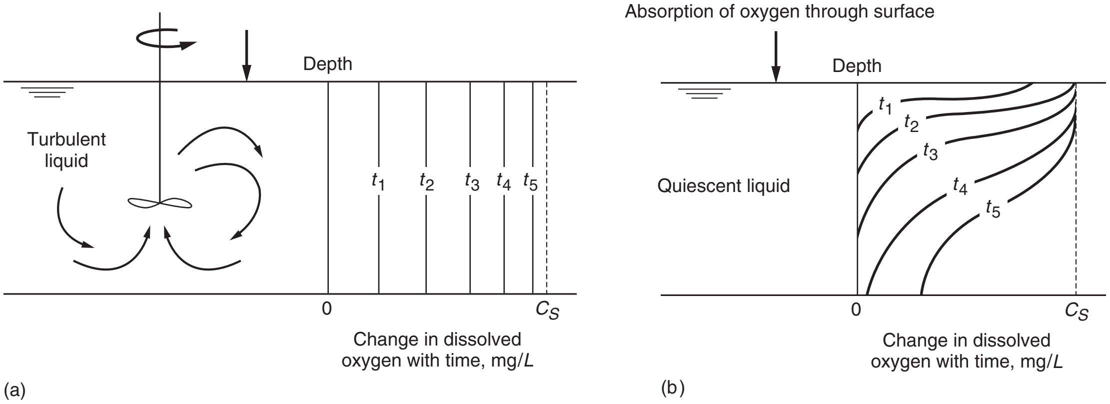


Figure 5–63

Definition sketch for the absorption of a gas: (a) under turbulent conditions where the concentration of gas in the gaseous and liquid phases is uniform and (b) under quiescent conditions. (Adapted from Tchobanoglous and Schroeder, 1985.)

Using Eq. (5–57) to describe the mass transfer through the surface of the basin, Eq. (5–60) can be written as follows, which is the same as Eq. (5–57):

$$\frac{dC}{dt} = K_L a(C_s - C_t) \tag{5–61}$$

Integrating Eq. (5–61) between the limits of $C = C_o$ and $C = C_t$ and $t = 0$ and $t = t$, where C_o is the initial concentration and C_t is the concentration at some time t

$$\int_{C_o}^{C_t} \frac{dC}{C_s - C_t} = K_L a \int_0^t dt \tag{5–62}$$

yields

$$\frac{C_s - C_t}{C_s - C_o} = e^{-(K_L a)t} \tag{5–63}$$

In Eq. (5–63), the term $(C_s - C_t)$ represents, as noted above, the degree of undersaturation at any time t and the term $(C_s - C_o)$ represents the initial degree of undersaturation. The application of Eq. (5–63) is illustrated in Example 5–12.

EXAMPLE 5–12 **Time Required to Absorb a Gas** Dechlorinated secondary effluent is placed in a storage basin until needed for reuse. If the initial dissolved oxygen concentration is 1.5 mg/L, estimate the time required for the dissolved oxygen concentration to increase to 8.5 mg/L due to surface reaeration assuming the water in the storage basin is circulated and not stagnant. Assume the K_L value for oxygen is equal to 0.03 m/h. The surface area of the storage basin is 400 m^2 and the depth is 3 m. Assume the temperature is 20°C and that the saturation value for DO is 9.09 (see Appendix E).

Solution

1. The time required for the concentration of oxygen to be increased from 1.5 to 8.5 can be estimated using Eq. (5–63).

   $$\frac{C_s - C_t}{C_s - C_o} = e^{-(K_L a)t}$$

   a. The oxygen saturation value from Example 2–6 is 9.09 mg/L.
   b. Solve for the term $(K_L a)t$.

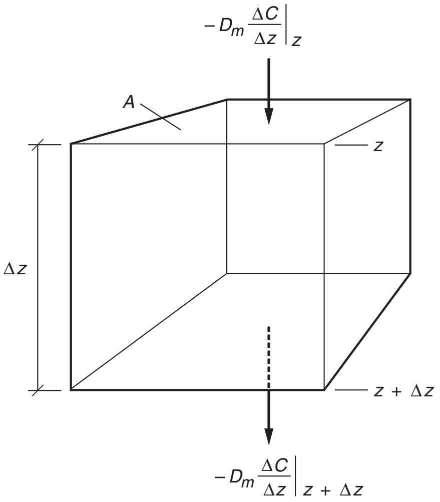
   $$\ln\left(\frac{9.09 - 8.5}{9.09 - 1.5}\right) = -2.55 = -(K_L a)t$$

2. The time required is
   a. The value for a, the interfacial area for mass transfer per unit volume:
   $a = A/V = 400\ \text{m}^2/(400\ \text{m}^2 \times 3\ \text{m}) = 0.33/\ \text{m}$
   b. Solve for t.
   $t = 2.55/[(0.03\ \text{m/h})(0.33\ /\text{m})] = 257\ \text{h} = 10.7\ \text{d}$

Comment The importance of the surface area exposed to the atmosphere is illustrated in this example. The larger the surface area relative to the depth, the greater the rate of oxygen transfer. In biological wastewater treatment either many small gas bubbles, released at the bottom of a reactor, are used to transfer oxygen to the active biomass or small droplets of water containing the active biomass are sprayed into the atmosphere to maximize the rate of oxygen transfer.

Figure 5–64

Definition sketch for the absorption of a gas under quiescent conditions due to molecular diffusion.



Absorption of Gases Under Quiescent Conditions

The transfer of a slightly soluble gas into a liquid under quiescent conditions (see Fig. 5–63b) occurs as a result of molecular diffusion. Applying the materials-balance approach to the control volume shown on Fig. 5–64, the quiescent transfer of a gas across the open surface can be modeled as follows:

First, a mass balance is written for the open basin as follows:

1. General word statement:

$$\begin{array}{c} \text{Rate of accumulation} \\ \text{of a gas within} \\ \text{the system boundary} \end{array} = \begin{array}{c} \text{rate of diffusion} \\ \text{of a gas into the} \\ \text{system boundary} \end{array} - \begin{array}{c} \text{rate of diffusion} \\ \text{of gas out of the} \\ \text{system boundary} \end{array} \tag{5–64}$$

2. Simplified word statement:

$$\text{Accumulation} = \text{inflow} - \text{outflow} \tag{5–65}$$

3. Symbolic representation at equilibrium (refer to Fig. 5–63):

$$\frac{\partial C}{\partial t}(A\Delta z) = -D_m A \frac{\Delta C}{\Delta z}\bigg|_z + D_m A \frac{\Delta C}{\Delta z}\bigg|_{z+\Delta z} \tag{5–66}$$

where $\partial C/\partial t$ = change in concentration per unit time, $ML^{-3}T^{-1}$, (g/m³·s)
A = surface area through which mass is transferred, L^2,(m²)
Δz = distance in z direction, L, (m)
D_m = coefficient of molecular diffusion, L^2T^{-1},(m²/s)
$\Delta C/\Delta z$ = change in concentration with distance, $ML^{-3}L^{-1}$, (g/m³·s)

Taking the limit as Δz approaches zero yields

$$\frac{\partial C}{\partial t} = D_m A \frac{\partial^2 C}{\partial z^2} \tag{5–67}$$

Equation (5–67) is also known as Fick's second law of diffusion (Crank, 1957). Typical values for the coefficient of molecular diffusion for gases of low solubility are reported in Table 5–25. It should be noted that the mechanism of adsorption of moderately soluble gases in water was studied extensively by Lewis and Whitman (1924), mentioned

Table 5–25

Approximate coefficients of molecular diffusion and coefficients of diffusion for gases of low solubility in water at 20°C[a]

Gas	Coefficient of molecular diffusion, cm²/h	Gas transfer coefficient, cm/h	Estimated film thickness, cm
Oxygen, O_2	6.7×10^{-2}	$32.3 \times 1.018^{T-20}$	$\sim 2 \times 10^{-3}$
Nitrogen, N_2	6.4×10^{-2}	$34.0 \times 1.019^{T-20}$	$\sim 2 \times 10^{-3}$
Carbon dioxide, CO_2	$\sim 6.5 \times 10^{-2}$		$\sim 2 \times 10^{-3}$
Air		$32.1 \times 1.019^{T-20}$	$\sim 2 \times 10^{-3}$

[a] Adapted from Adeney and Becker (1920), Becker (1924).

previously, and by Adeney and Becker (1919), and Becker (1924) among others and that their work is still valid today. Solutions to Eq. (5–67) for a variety of boundary conditions may be found in Carlslaw and Jaeger (1947), Crank (1957), Danckwertz (1970), and Thibodeaux (1996).

Desorption (Removal) of Gases

The application of the gas-liquid mass transfer relationship for the removal of a gas from a liquid will be illustrated by considering the volatilization of a constituent from a liquid. The same approach used for the addition of a gas will be followed, but noting that for the removal of a gas from a liquid Eq. (5–61) is written as follows:

$$\frac{dC}{dt} = -K_L a(C_t - C_s) \qquad (5\text{–}68)$$

where the term $(C_t - C_s)$ represents the degree of supersaturation at any time t. If Eq. (5–68) is integrated between the limits of $C = C_s$ and $C = C_t$ and $t = 0$ and $t = t$, the integrated form of Eq. (5–68), corresponding to Eq. (5–63), for the volatilization of a gas from a supersatured liquid is given by

$$\frac{C_t - C_s}{C_o - C_s} = e^{-(K_L a)t} \qquad (5\text{–}69)$$

In Eq. (5–69), the term $(C_o - C_s)$ represents the initial degree of supersaturation. The application of Eq. (5–69) is illustrated in Example 5–13.

EXAMPLE 5–13 **Time Required for a Gas to Volatilize from a Liquid** A quantity of benzene was spilled accidentally into a treated wastewater storage basin. Estimate the time required for the concentration of benzene to drop by 50 percent from the initial concentration due to volatilization. Assume the $K_L a$ value for benzene is 0.144/h.

Solution

1. If it is assumed that the concentration of the specific volatile chemical is not common to the atmosphere, then $C_s \sim 0$, and Eq. (5–69) can be written as follows:

$$\frac{C_t}{C_o} = e^{-(K_L a)t}$$

2. Knowing the value of K_La, the time for the concentration to dissipate to one half of the initial concentration can be determined by rewriting the equation developed in Step 1 as:

$$\frac{0.5C_o}{1.0C_o} = e^{-(K_La)t_{1/2}}$$

Solving for $t_{1/2}$ yields

$$t_{1/2} = \frac{0.69h}{K_La}$$

3. Using a K_La value of 0.144 m/h for benzene, the time for 50 percent of the initial concentration to dissipate is

$$t_{1/2} = \frac{(0.69)(2\,\text{m})}{(0.144\,\text{m/h})} = 9.6\,\text{h}$$

The application of gas-liquid mass transfer is considered further in the following section on aeration systems.

5–11 AERATION SYSTEMS

The functioning of aerobic processes, such as activated sludge, biological filtration, and aerobic digestion, depends on the availability of sufficient quantities of oxygen. Because of the low solubility of oxygen in water and the consequent low rate of oxygen transfer, sufficient oxygen to meet the requirements of aerobic waste treatment does not enter water through normal surface air-water interfaces. To transfer the large quantities of oxygen that are needed, additional interfaces must be formed. Either air or oxygen can be introduced into the liquid, or the liquid in the form of droplets can be exposed to the atmosphere.

Oxygen can be supplied by mechanical devices designed to increase surface air-water interfaces or by means of submerged diffusion of air or pure-oxygen bubbles to the water to create additional gas-water interfaces. Submerged-bubble aeration is most frequently accomplished by dispersing air bubbles in the liquid at depths up to 10 m (30 ft). Mechanical devices may also be used to create smaller bubbles and more air-water interfaces by impinging a flow of liquid at an orifice to break up the air bubbles into smaller sizes. Turbine mixers may be used to disperse air bubbles introduced below the center of the turbine; they are designed both to mix the liquid in the basin and to expose it to the atmosphere in the form of small liquid droplets.

Oxygen Transfer

Commercial equipment used for oxygen transfer are rated in accordance with established test procedures using clean water and reported in terms of oxygen transfer under standard conditions. The definition of standard conditions varies among industries and geographical areas. The definition used in this textbook is defined as follows:

Standard temperature, $T_s = 20°\text{C}$ (68°F)
Standard pressure, $P_s = 1.0$ atm [101.325 kPa (14.7 $\text{lb}_f/\text{in.}^2$)]

It is important to determine the standard conditions being used for equipment rating and when dealing with foreign equipment manufacturers.

Application of Correction Factors. The actual amount of oxygen required must be obtained by applying factors to a standard oxygen requirement in clean water that reflect the effects of the wastewater (or process water) characteristic, such as the concentration of solids, salinity, surface tension, temperature, elevation, diffused depth (for diffused aeration systems), the desired oxygen operating level, and the effects of mixing intensity and basin configuration. The interrelationship of these factors is given by the following expression:

$$\text{OTR}_f = (\text{SOTR})\left[\frac{(\tau\beta\Omega C^*_{\infty 20} - C)}{C^*_{\infty 20}}\right][(\theta)^{t-20}](\alpha)(F) \tag{5–70}$$

Where: OTR_f = field oxygen transfer rate estimated for the system operating under process conditions at an average DO concentration, C, and temperature, T, kg O_2/h

SOTR = oxygen transfer rate under standard conditions (20°C, 1 atm, $C = 0$ mg/L), kg O_2/h

τ = temperature correction factor $= C^*_{\text{st}}/C^*_{\text{s20}}$

C^*_{st} = dissolved oxygen surface saturation concentration at operating temperature, mg/L (see Appendix E)

C^*_{s20} = dissolved oxygen surface saturation concentration at standard temperature (20°C), mg/L (see Appendix E)

β = relative DO saturation to clean water, typically 0.95 to 0.98.
$= C^*_{\infty\,(\text{wastewater})}/C^*_{\infty\,(\text{tap water})}$ (see following discussion)

C^*_{∞} = steady-state DO saturation concentration obtained from nonlinear regression analysis of clean water test results (also available from aeration equipment manufacturers)

Ω = pressure correction factor
$= P_b/P_s$

P_b = barometric pressure at test site, m, kPa (see Appendix B)

P_s = standard barometric pressure (1.00 atm, 10.33 m, 101.325 kPa)

$C^*_{\infty,20}$ = saturated DO value at sea level and standard temperature (20°C) for diffused aeration, mg/L. It is higher than C_{st} as it is affected by oxygen transfer from bubbles under pressure in water column. The value of $C^*_{\infty,20}$ can be estimated using the following equation (U.S. EPA, 1989):

$$C^*_{\infty 20} = C^*_{\text{s20}}\left[1 + d_e\left(\frac{D_f}{P_s}\right)\right]$$

d_e = mid-depth correction factor; may vary from 0.25–0.45 (0.40)

D_f = depth of diffusers in basin, m

C = average dissolved oxygen concentration within the entire process water volume, mg/L

θ = empirical temperature correction factor, typically 1.024 (see following discussion)

T = field temperature, °C

T_s = standard temperature, °C

α = relative oxygen transfer rate in process water versus clean water (see following discussion)
$= K_La_{f\,20(\text{wastewater})}/K_La_{20(\text{tap water})}$

F = fouling factor, typically 0.65 to 0.9 (For surface aerators, $F = 1$)

Note that the OTR_f and SOTR values given above can also be expressed as transfer efficiencies. The fouling factor F is used to account for both internal and external fouling of air diffusers. Internal fouling is caused by impurities in the compressed air, whereas external fouling is caused by the formation of biological slimes and inorganic precipitants. The oxygen necessary for the biological process can be supplied by using air or pure oxygen. Three commonly used methods to introduce oxygen into the liquid in the aeration tank are

(1) mechanical aeration, (2) injection of diffused air, and (3) injection of high-purity oxygen. The application of Eq. (5–70) for aeration in biological treatment systems is illustrated in Example 8–3 in Chap. 8.

Evaluation of Alpha (α) Correction Factor

For a given volume of water being aerated, aeration devices are evaluated on the basis of the quantity of oxygen transferred per unit of air introduced to the water for equivalent conditions (temperature and chemical composition of the water, depth at which the air is introduced, etc.). The evaluation of the oxygen transfer coefficient in clean water and wastewater is considered in the following discussion.

Oxygen Transfer in Clean Water. The accepted procedure for determining the overall oxygen transfer coefficient in clean water, as detailed in ASCE (1992), may be outlined as follows. The accepted test method involves the removal of dissolved oxygen (DO) from a known volume of water by the addition of sodium sulfite followed by reoxygenation to near the saturation level. The DO of the water volume is monitored during the reaeration period by measuring DO concentrations at several different points selected to best represent the contents of the tank. The minimum number of points, their distribution, and range of DO measurements made at each determination point are specified in the procedure (ASCE, 1992).

The data obtained at each determination point are then analyzed by a simplified mass transfer model (Eq. (5–63) given previously:

$$\frac{C_s - C_t}{C_s - C_o} = e^{-(K_L a)t}$$

where $K_L a$ = overall liquid film coefficient
C_t = concentration in liquid bulk phase at time t, mg/L
C_s = concentration in equilibrium with gas as given by Henry's law
C_o = initial concentration

Equation (5–63) is used to estimate the apparent volumetric mass transfer coefficient $K_L a$ and the equilibrium concentration C_x^*, obtained as the aeration period approaches infinity. The term C_x^* is substituted for the term C_s in Eq. (5–63). A nonlinear regression analysis is employed to fit Eq. (5–63) to the DO profile measured at each determination point during reoxygenation test period. In this way, estimates of $K_L a$ and C_x^* are obtained at each determination point. These estimates are adjusted to standard conditions and the standard oxygen transfer rate (mass of oxygen dissolved per unit time at a hypothetical concentration of zero DO) is obtained as the average of the products of the adjusted point $K_L a$ values, the corresponding adjusted point C_x^* values, and the tank volume (ASCE, 1992).

Oxygen Transfer in Wastewater. In an activated-sludge system, the $K_L a$ value can be determined by considering the uptake of oxygen by microorganisms. Typically, oxygen is maintained at a level of 1 to 3 mg/L and the microorganisms use the oxygen as rapidly as it is supplied. In equation form,

$$\frac{dC}{dt} = K_L a(C_s - C) - r_M \tag{5–71}$$

where C = concentration of oxygen in solution
r_M = rate of oxygen used by the microorganisms

Typical values of r_M vary from 2 to 7 g/d·g MLVSS (gram of mixed-liquor volatile suspended solids). If the oxygen level is maintained at a constant level, dC/dt is zero and

$$r_M = K_L a\,(C_s - C) \tag{5–72}$$

C in this case is constant also.

Values of r_M can be determined in a laboratory by using a respirometer. In this case, $K_L a$ can easily be determined as follows:

$$K_L a = \frac{r_M}{(C_s - C)} \tag{5–73}$$

Prediction of in-process oxygen transfer rates is nearly always based on an oxygen rate model. ASCE has developed standard guidelines for performing in-process tests (ASCE, 1997). The overall oxygen mass transfer coefficient $K_L a$ is usually determined in test or full-scale facilities. If pilot-scale facilities are used to determine $K_L a$ values, scale-up must be considered. The mass transfer coefficient $K_L a$ is also a function of temperature, intensity of mixing (and hence of the type of aeration device used and the geometry of the mixing chamber), and constituents in the water (Tchobanoglous and Schroeder, 1985). The effects of temperature, mixing intensity, tank geometry, and wastewater characteristics and the application of correction factors are discussed below. Determination of $K_L a$ is illustrated in Example 5–14.

Effect of Temperature on Oxygen Transfer. Temperature effects are treated in the same manner as they were treated in establishing the BOD rate coefficient (i.e., by using an exponential function to approximate the van't Hoff–Arrhenius relationship):

$$K_L a_{(T)} = K_L a_{(20°\text{C})} \theta^{T-20} \tag{5–74}$$

where $K_L a_{(T)}$ = oxygen mass transfer coefficient at temperature T, s^{-1}
$K_L a_{(20°\text{C})}$ = oxygen mass transfer coefficient at 20°C, s^{-1}

Reported values for θ vary with the test conditions. Typical θ values are in the range of 1.015 to 1.040. The θ value of 1.024 is typical for both diffused and mechanical aeration devices.

EXAMPLE 5–14 **Determination of $K_L a$ Value** The following data have been obtained from a surface aeration test. Using the data, determine the $K_L a$ value at 20°C using a linear regression analysis. The temperature of the water was 15°C.

Time, min	DO conc., mg/L
4	0.8
7	1.8
10	3.3
13	4.5
16	5.5
19	6.2
22	7.3

Solution

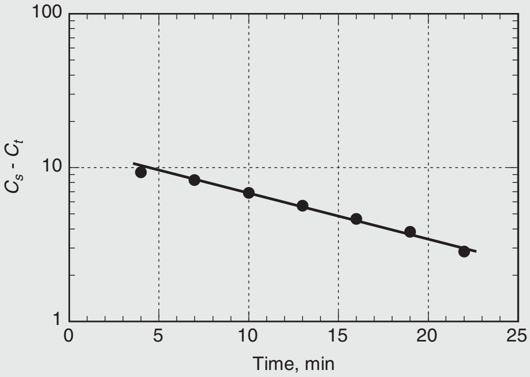
1. To analyze the field data, rewrite Eq. (5–63) in a linear form.

$$\log(C_s - C_t) = \log(C_s - C_o) - \frac{K_L a}{2.303} t$$

2. Determine $C_s - C_t$, and plot $C_s - C_t$ versus t on semilog paper.
   a. $C_{s(15°C)} = 10.08$ (see Appendix E)

Time, min	$C_s - C_t$, mg/L
4	9.28
7	8.28
10	6.78
13	5.58
16	4.58
19	3.88
22	2.78

   b. Plot $C_s - C_t$ versus t. See the following plot:



3. Determine the value of K_La at 20°C.
   a. From the plot, the value of K_La at 15°C is

$$K_La = 2.303 \frac{\log C_{t_1} - \log C_{t_2}}{t_2 - t_1}(60)$$

$$K_La = 2.303 \frac{\log 8.28 - \log 2.78}{22 - 7}(60)$$

$$K_La = 4.37/\text{h}$$

   b. The approximate value of K_La at 20°C is

$$K_La_{20} = (4.37)\ 1.024^{20-15}$$
$$= 4.92/\text{h}$$

Comment The value of K_La determined in this example is approximate because a linear regression analysis was used. To obtain a more accurate value of K_La, the nonlinear method outlined in ASCE (1992) should be used.

Effects of Mixing Intensity and Tank Geometry. Effects of mixing intensity and tank geometry are difficult to deal with on a theoretical basis but must be considered in the design process because aeration devices are often chosen on the basis of efficiency. Efficiency is strongly related to the K_La value associated with a given aeration unit. In most cases an aeration device is rated for a range of operating conditions using tap water having a low TDS concentration. A correction factor is used to estimate the K_La value in the actual system:

$$\alpha = \frac{K_La\,(\text{wastewater})}{K_La\,(\text{tap water})} \tag{5–75}$$

where alpha (α) is the correction factor. Values of α vary with the type of aeration device, the concentration of MLVSS, the basin geometry, the degree of mixing, and other wastewater characteristics, as discussed above. Values of α vary from about 0.3 to 1.2. Typical values for diffused and mechanical aeration equipment, discussed in the following section, are in the range of 0.4 to 0.8 and 0.6 to 1.2, respectively. EPA's Design Manual – Fine Pore Aeration Systems (U.S. EPA, 1989) includes α values collected during full-scale tests on fine pore aeration systems. If the basin geometry in which the aeration device is to be used is significantly different from that used to test the device, great care must be exercised in selecting an appropriate α value.

Evaluation of Beta (β) Correction Factor. The correction factor β is used to correct the test system oxygen transfer rate for differences in oxygen solubility due to constituents in the water such as salts, particulates, and surface-active substances:

$$\beta = \frac{C_s\,(\text{wastewater})}{C_s\,(\text{tap water})} \tag{5–76}$$

The ratio can vary from approximately 0.8 to 1.0 and is generally close to 1.0 for municipal wastewater. Because the ratio cannot be measured by a membrane probe and because many wastewaters contain substances that interfere with wet methods (Winkler method), it is difficult to measure accurately. For this reason the value of β is calculated as the ratio of the surface saturation DO concentration in the process water to the surface saturation DO concentration in clean water. The values given in Appendix E can be used for C_s in clean water with varying amounts of salinity and at various elevations.

Types of Aeration Systems

The various types of aeration systems used and their applications are described in Table 5–26. The principal types, diffused-air systems, mechanical aeration, and high-purity oxygen systems, are discussed in the following paragraphs. Postaeration, which is a special application for aeration, is also discussed in the latter part of this section.

Diffused-Air Aeration

The two basic methods of aerating wastewater are (1) to introduce air or pure oxygen into the wastewater with submerged diffusers or other aeration devices, or (2) to agitate the wastewater mechanically so as to promote solution of air from the atmosphere. A diffused-air system consists of diffusers that are submerged in the wastewater, header pipes, air mains, and the blowers and appurtenances through which the air passes. The following discussion covers the selection of diffusers, the design of blowers, and air piping design.

Diffusers. In the past, the various diffusion devices have been classified as either fine bubble or coarse bubble, with the connotation that fine bubbles are more efficient in transferring oxygen. The definition of terms and the demarcation between fine and coarse bubbles, however, have not been clear, but they continue to be used. The current preference is to categorize the diffused aeration systems by the physical characteristics of the

Table 5–26
Description of commonly used devices for wastewater aeration

Classification	Description	Use or application
Submerged:		
Diffused air		
Fine bubble (fine-pore system	Bubbles generated with ceramic, plastic, or flexible membranes (domes, tubes , disks, plates, or panel configurations)	All types of activated sludge processes
Coarse bubble (non-porous) system	Bubbles generated with orifices, injectors and nozzles, or shear plates	All types of activated sludge processes, channel and grit chamber aeration, and aerobic digestion
Sparger turbine	Low-speed turbine and compressed air injection	All types of activated sludge processes and aerobic digestion
Mixer/Aerator	Hyperboloid mixer rotation mixes the air in the dispersing tunnels turning it into fine bubbles by special shear fins	All types of activated sludge processes and channel aeration
Static tube mixer	Short tubes with internal baffles designed to retain air injected at bottom of tube in contact with liquid	Aerated lagoons and activated sludge processes
Jet	Compressed air injected into mixed liquor as it is pumped under pressure through jet device	All types of activated sludge processes, equalization tank mixing and aeration, and deep tank aeration
Surface:		
Low-speed turbine aerator	Large diameter turbine used to expose liquid droplets to the atmosphere	Conventional activated sludge processes, aerated lagoons, and aerobic digestion
High-speed floating aerator	Small diameter propeller used to expose liquid droplets to the atmosphere	Aerated lagoons and aerobic digestion
Aspirating	Inclined propeller assembly	Aerated lagoons
Rotor-brush or rotating disk assembly	Blades or disks mounted on a horizontal central shaft are rotated through the liquid. Oxygen is induced into the liquid by the splashing action of the rotor and by exposure of liquid droplets to the atmosphere	Oxidation ditch, channel aeration, and aerated lagoons
Cascade	Wastewater flows over a series of steps in sheet flow	Post aeration

equipment. Three categories are defined: (1) porous or fine-pore diffusers, (2) nonporous diffusers, and (3) other diffusion devices such as jet aerators, aspirating aerators, and U-tube aerators. The various types of diffused-air devices are described in Table 5–27.

Porous Diffusers. Porous diffusers are made in many shapes, the most common being discs, domes, plates, tubes, and membranes. Tubes are also used. Ceramic plates were once the most popular but are costly to install and difficult to maintain. Porous ceramic domes and discs, and membranes, discs, and tubes have largely supplanted ceramic plates in

Table 5–27

Description of commonly used air diffusion devices

Type of diffuser or device	Transfer efficiency	Description	See Fig. No.
Porous			
Disk	High	Rigid ceramic discs mounted on air distribution pipes near the tank floor	5–65(a), (c)
Dome	High	Dome-shaped ceramic diffusers mounted on air distribution pipes near the tank floor	5–65(b)
Membrane	High	Flexible porous membrane supported on disk mounted on an air distribution grid	5–65(d)
Panel	Very high	Rectangular panel with a flexible plastic perforated membrane	5–67
Nonporous			
Fixed orifice			
Orifice	Low	Devices usually constructed of molded plastic and mounted on air distribution pipes	5–68(a)
Slotted tube	Low	Stainless steel tubing containing perforations and slots to provide a wide band of diffused air	5–68(b)
Static tube	Low	Stationary vertical tube mounted on basin bottom and functions like an air lift pump	5–69(a)

newer installations (see Fig. 5–65). Several manufacturers of plate diffusers using membrane media are currently on the market.

Domes, discs, or tube diffusers are mounted on or screwed into air manifolds, which may run the length of the tank close to the bottom and along one or two sides, or short manifold headers may be mounted on movable drop pipes on one side of the tank. Dome and disc diffusers may also be installed in a grid pattern on the bottom of the aeration tank to provide uniform aeration throughout the tank (see Fig. 5–66).

Numerous materials have been used in the manufacture of porous diffusers. These materials generally fall into the categories of rigid ceramic and plastic materials and flexible plastic, rubber, or cloth sheaths. The ceramic materials consist of rounded or irregular-shaped mineral particles bonded together to produce a network of interconnecting passageways through which compressed air flows. As the air emerges from the surface pores, pore size, surface tension, and air flowrate interact to produce the bubble size. Rigid porous plastic materials are similar to the ceramic materials in that the plastics contain a number of interconnecting channels or pores through which the compressed air can pass. Thin, flexible sheaths made from soft plastic or synthetic rubber have also been developed and adapted to discs, tubes, and plates. Air passages are created by punching minute holes in the sheath material. When the air is turned on, the sheath expands and each slot acts as a variable aperture opening; the higher the air flowrate, the greater the opening.

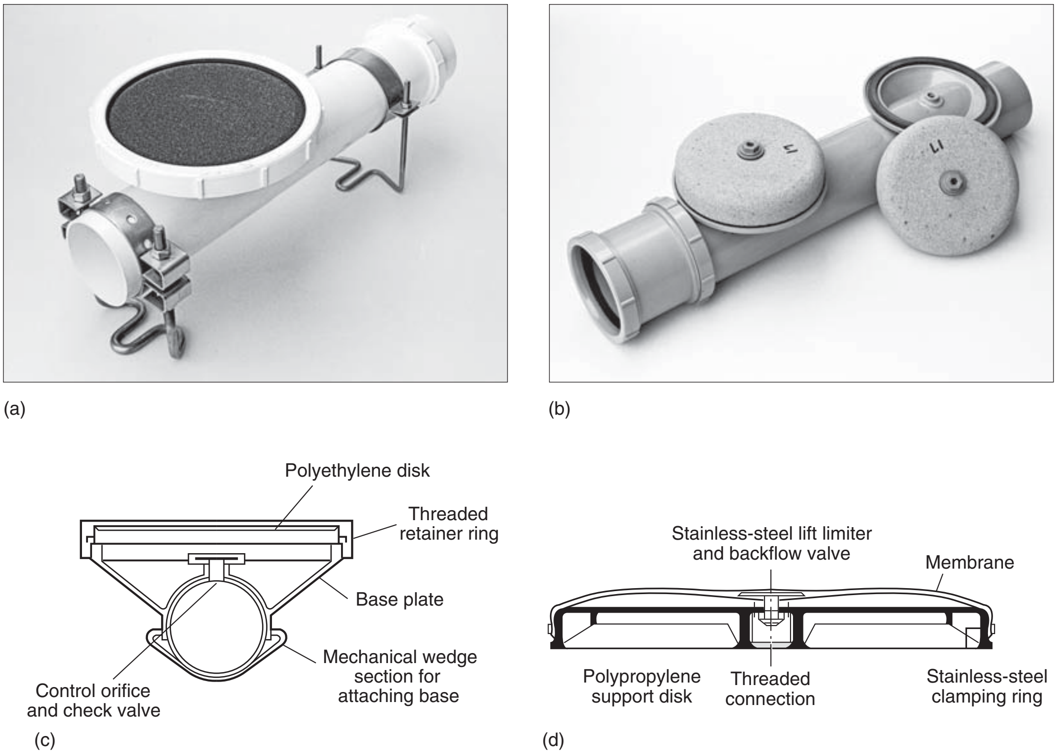


Figure 5–65

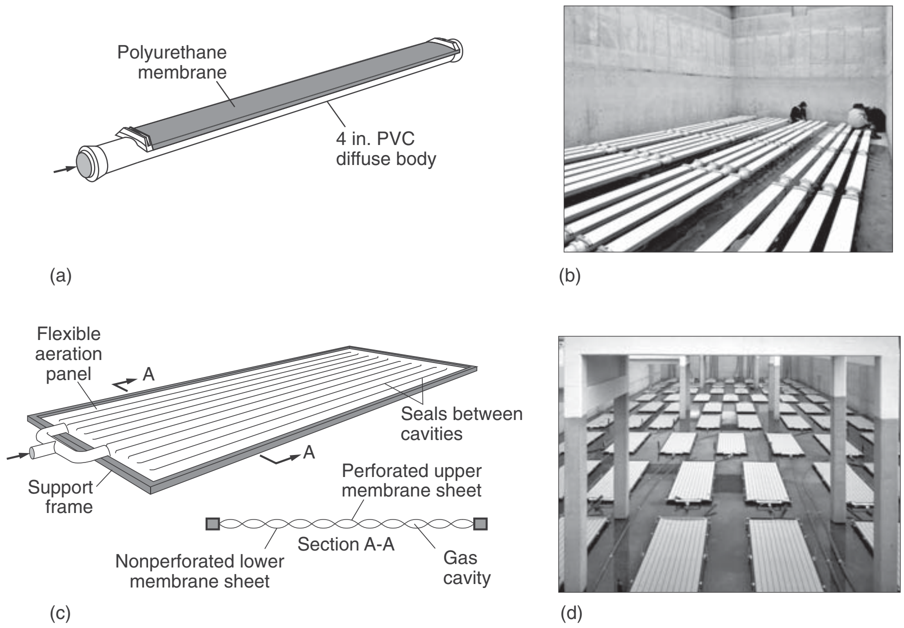
Typical porous air diffusers: (a) view of ceramic disk diffuser, (b) view of ceramic dome, (c) cross-section through polyethylene disk diffuser, and (d) perforated membrane.

Figure 5–66

Plug-flow aeration tank equipped with ceramic disk aeration devices.

Figure 5–67

Ultra-fine pore membrane aeration devices: (a) schematic of aeration panel (adapted from Xylem Water Solutions), (b) view of aeration panels in bottom of an activated sludge reactor aeration (courtesy Xylem Water Solutions), (c) schematic of aeration panel (adapted from Parkson), and (d) view of aeration panels placed in bottom of an activated sludge reactor (courtesy of Parkson).



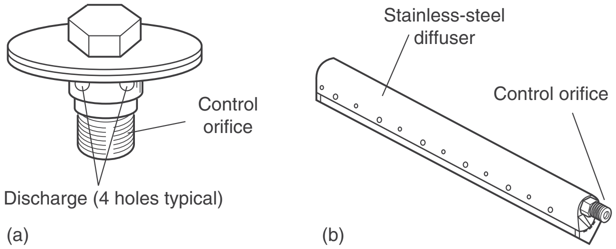
Rectangular panels or strips that use a flexible polyurethane sheet (see Fig. 5–67) are also used in activated-sludge aeration. The panels are constructed with a stainless-steel frame and are placed on or close to the bottom of the tank and anchored. Advantages cited for aeration panels are (1) ultra-fine bubbles are produced that significantly improve oxygen transfer and system energy efficiency, (2) large areas of the tank floor can be covered, which facilitates mixing and oxygen transfer, and (3) foulants can be dislodged by "bumping," i.e., increasing the airflow to flex the membrane. Disadvantages are (1) the membrane has a higher headloss, which may affect blower performance in retrofit applications, and (2) increased blower air filtration is required to prevent internal fouling.

With all porous diffusers, it is essential that the air supplied be clean and free of dust particles that might clog the diffusers. Air filters, often consisting of viscous-impingement and dry-barrier types, are commonly used. Precoated bag filters and electrostatic filters have also been used. The filters should be installed on the blower inlet.

Nonporous Diffusers. Several types of nonporous diffusers are available (see Figs. 5–68 and 5–69). Nonporous diffusers, also known as coarse bubble diffusers, produce larger bubbles than porous diffusers and consequently have lower aeration efficiency; but the advantages of lower cost, less maintenance, and the absence of stringent air-purity

Figure 5–68

Nonporous coarse bubble diffusers: (a) orifice and (b) tube.



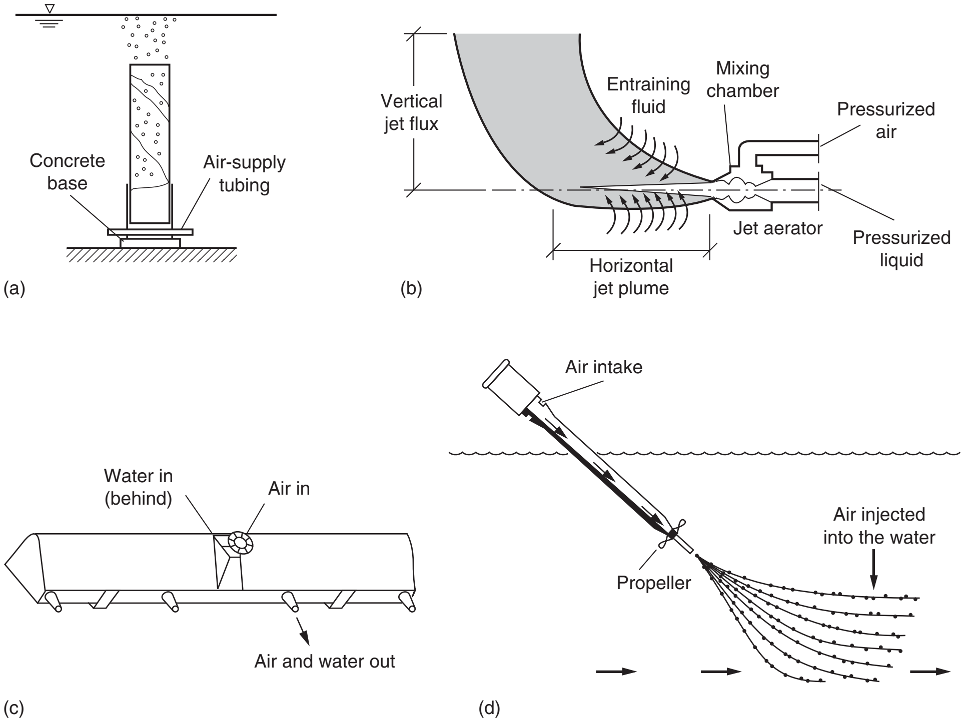


Figure 5–69

Other devices used for the transfer of oxygen: (a) static tube mixer where air is introduced at the base of the aerator that contains mixing elements, (b) jet reactor in which pressurized air and liquid are combined in a mixing chamber (as the jet is emitted, the surrounding liquid is entrained to enhance oxygen transfer), (c) jet aerator in a manifold arrangement, and (d) aspirating aerator.

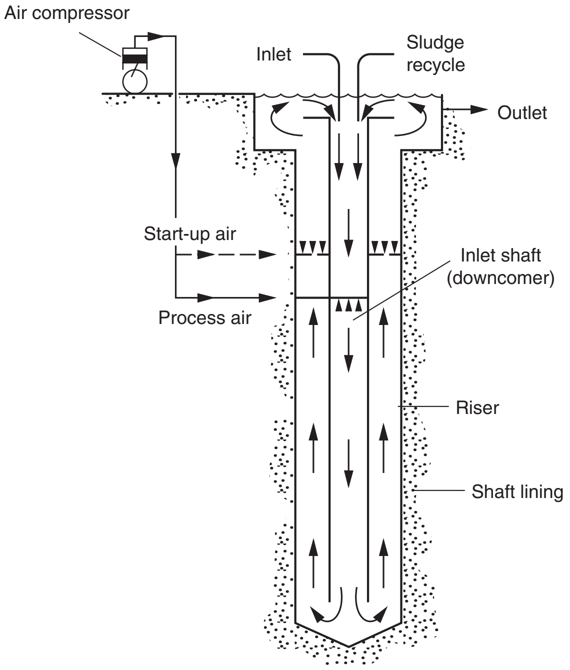
requirements may offset the lower oxygen transfer efficiency and energy cost. Typical system layouts for orifice diffusers closely parallel the layouts for porous dome and disk diffusers; however, single- and dual-roll spiral patterns using narrow- or wide-band diffuser placement are the most common. Applications for orifice and tube diffusers include aerated grit chambers, channel aeration, flocculation basin mixing, aerobic digestion, and industrial waste treatment (WEF, 1998b).

In the static tube aerator [see Fig. 5–69(a)], air is introduced at the bottom of a circular tube that can vary in height from 0.5 to 1.25 m (1.5 to 4.0 ft). Internally, the tubes are fitted with alternately placed deflection plates to increase the contact of the air with the wastewater. Mixing is accomplished because the tube aerator acts as an airlift pump. Static tubes are normally installed in a grid-type floor coverage pattern.

Other Air-Diffusion Devices. Jet aeration [see Figs. 5–69(b) and (c)] combines liquid pumping with air diffusion. The pumping system recirculates liquid in the aeration basin, ejecting it with compressed air through a nozzle assembly. This system is particularly suited for deep (8 m) tanks. Aspirating aeration [see Fig. 5–69(d)] consists of a motor-driven aspirator pump. The pump draws air in through a hollow tube and injects it underwater where both high velocity and propeller action create turbulence and diffuse the air bubbles. The aspirating device can be mounted on a fixed structure or on pontoons. U-tube aeration consists of a deep shaft that is divided into two zones (see Fig. 5–70). Air is added to the influent wastewater in the downcomer under high pressure; the mixture travels to the bottom of the tube and then back to the surface. The great depth to which the air-water mixture is subjected results in high oxygen transfer efficiencies because the high pressure forces all the oxygen into solution. U-tube aeration has particular application for high-strength wastes.

Figure 5–70

Typical schematic of a U-tube aerator.



Diffuser Performance. The efficiency of oxygen transfer depends on many factors, including the type, size, and shape of the diffuser; the air flowrate; the depth of submersion; tank geometry including the header and diffuser location; and wastewater characteristics. Aeration devices are conventionally evaluated in clean water and the results adjusted to process operating conditions through widely used conversion factors. Typical clean water transfer efficiencies and air flowrates for various diffused-air devices are reported in Table 5–28. Typically, the standard oxygen transfer efficiency (SOTE) increases with depth; the transfer efficiencies in Table 5–28 are shown for the 4.5-m (15-ft) depth, the most common depth of submergence. Data on the variation of SOTE with water depth for various diffuser types can be found in WPCF (1988). The variation of oxygen transfer efficiencies with the type of diffuser and diffuser arrangement are also illustrated in Table 5–28. Additional data on the effects of diffuser arrangement on transfer efficiency are reported in U.S. EPA (1989).

Oxygen transfer efficiency (OTE) of porous diffusers may also decrease with use due to internal clogging or exterior fouling. Internal clogging may be due to impurities in the compressed air that have not been removed by the air filters. External fouling may be due to the formation of biological slimes or inorganic precipitants. The effect of fouling on OTE is described by the term F in Eq. (5–70). The rate at which F decreases with time is designated f_F, which is expressed as the decimal fraction of OTE lost per unit time. The rate of fouling will depend on the operating conditions, changes in wastewater characteristics, and the time in service. The fouling rates are important in determining the loss of OTE and the expected frequency of diffuser cleaning. Fouling and the rate of fouling can be estimated by (1) conducting full-scale OTE tests over a period of time, (2) monitoring aeration system efficiency (see Fig. 5–71), and (3) conducting OTE tests of fouled and new diffusers.

Blowers. There are four types of blowers commonly used for wastewater aeration: single-stage centrifugal, multi-stage centrifugal, high speed turbo, and positive displacement. Centrifugal blowers [see Figs. 5–72(a) and (b)] are almost universally used where the

Table 5–28

Typical information on the clean water oxygen transfer efficiency of various air diffuser systems

Diffuser type and placement	Air flowrate/diffuser ft^3/min	m^3/min	SOTE (%) at 4.5 m (15 ft) submergence[a]
Ceramic discs - grid	0.4–3.4	0.01–0.1	25–35
Ceramic domes - grid	0.5–2.5	0.015–0.07	27–37
Ceramic plates - grid	2.0–5.0[b]	0.6–1.5[c]	26–33
Rigid porous plastic tubes			
Grid	2.4–4.0	0.07–0.11	28–32
Dual spiral roll	3.0–11.0	0.08–0.3	17–28
Single spiral roll	2.0–12.0		13–25
Nonrigid porous plastic tubes			
Grid	1.0–7.0	0.03–0.2	26–36
Single spiral roll	2.0–7.0	0.06–0.2	19–37
Perforated membrane tubes			
Grid	1.0–4.0	0.03–0.11	22–29
Quarter points	2.0–6.0	0.6–0.17	19–24
Single spiral roll	2.0–6.0	0.6–0.17	15–19
Perforated membrane panels	N/A	N/A	38–43[d]
Jet aeration			
Side header	54–300	1.5–8.5	15–24
Nonporous diffusers			
Dual spiral roll	3.3–10	0.1–0.28	12–13
Mid width	4.2–45	0.12–1.25	10–13
Single spiral roll	10–35	0.28–1.0	9–12

[a] SOTE = Standard oxygen transfer efficiency. Standard conditions: tap water 20°C (68°F); at 101.325 kN/m^2 (14.7 $lb_f/in.^2$); and initial dissolved oxygen = 0 mg/L.

[b] Units are ft^3/ft^2 of diffuser·min.

[c] Units are m^3/m^2 of diffuser·min.

[d] Personal communication, Parkson Corporation.

N/A = Not applicable.

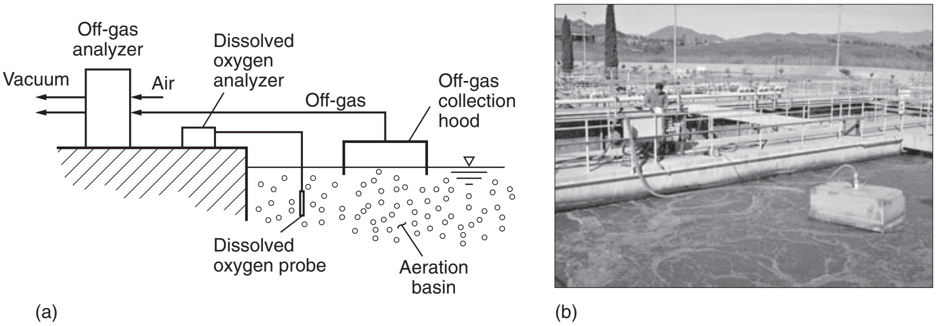
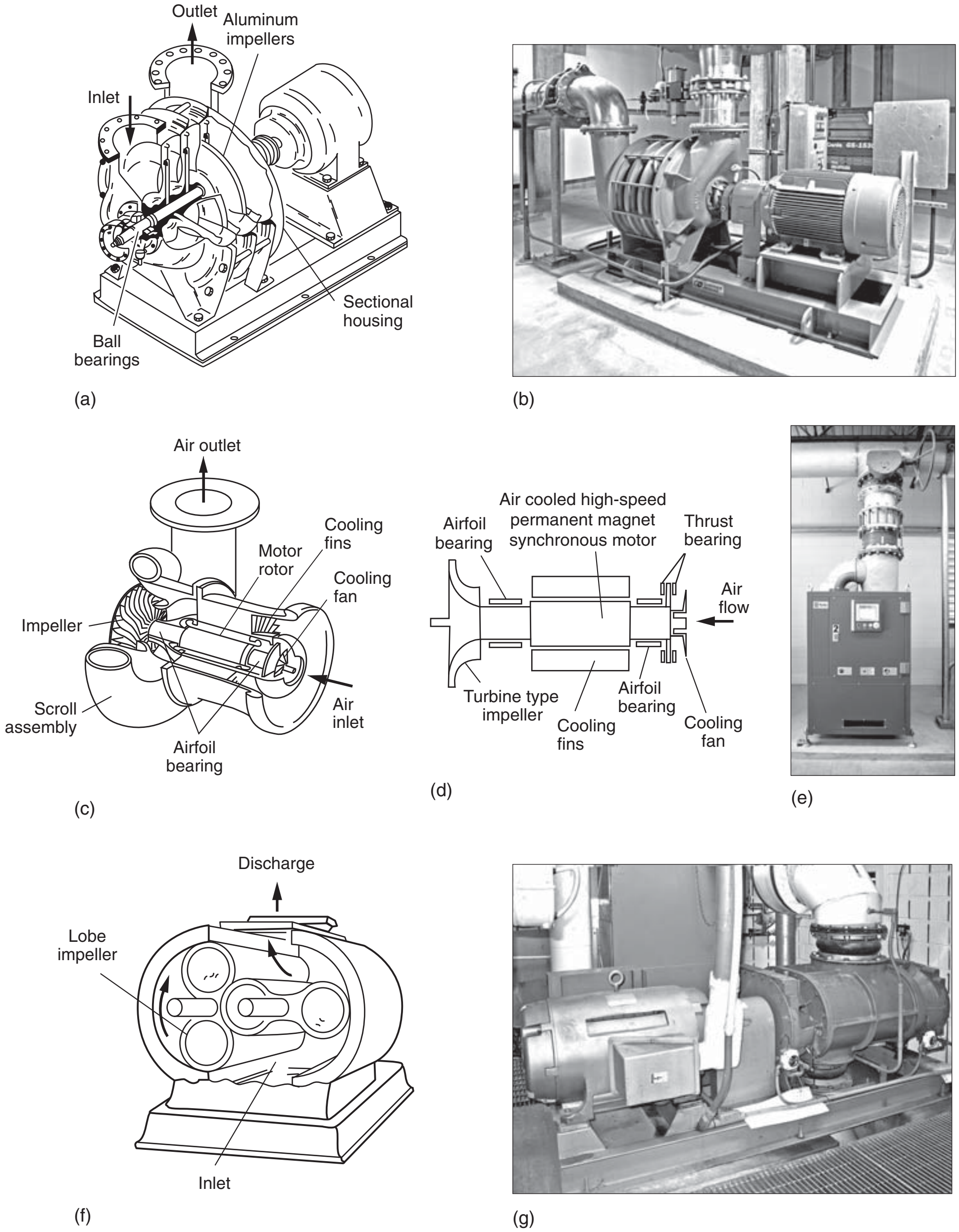


Figure 5–71

Aeration hood used to measure oxygen transfer rates in a biological wastewater treatment reactor: (a) schematic of measurement setup. The composition of the gases trapped in the hood is analyzed to assess aeration system performance (adapted from Rosso et al., 2011) and (b) hood is moved to multiple locations in each basin to obtain average performance data.

Figure 5–72

Typical blowers used for diffused-air aeration: (a) schematic of multistage centrifugal blower, (b) view of multi-stage centrifugal blower, (c) schematic of high speed turbo blower (adapted from APG Neuros), (d) schematic cross section through high speed turbo blower (adapted from APG Neuros), (e) view of high speed turbo blower installation in self contained sound-dampening enclosure, (f) schematic of rotary-lobe positive displacement blower, and (g) view of rotary-lobe positive displacement blower.



unit capacity is greater than 425 m^3/min (15,000 ft^3/min) of free air. Rated discharge pressures range normally from 48 to 62 kN/m^2 (7 to 9 lb$_f$/in.^{2}) for single stage centrifugal blowers and as high as 138 kN/m^2 (20 lb$_f$/in.^{2}) for multistage centrifugal blowers. Centrifugal blowers have operating characteristics similar to a low-specific-speed centrifugal pump. The discharge pressure rises from shutoff to a maximum at about 50 percent of capacity and then drops off. The operating point of the blower is determined, similar to a centrifugal pump, by the intersection of the head-capacity curve and the system curve.

In wastewater treatment plants, the blowers must supply a wide range of air flowrates with a relatively narrow pressure range under varied environmental conditions. A blower usually can only meet one particular set of operating conditions efficiently. Because it is necessary to meet a wide range of air flowrates and pressures at a wastewater treatment plant, provisions have to be included in the blower system design to regulate or turn down the blowers. Methods to achieve regulation or turndown are (1) flow blowoff or bypassing, (2) inlet throttling, (3) adjustable discharge diffuser, (4) variable-speed driver, and (5) parallel operation of multiple units. Inlet throttling and an adjustable discharge diffuser are applicable only to centrifugal blowers; variable-speed drivers are more commonly used on positive-displacement blowers. Flow blowoff and bypassing is also an effective method of controlling surging of a centrifugal blower, a phenomenon that occurs when the blower operates alternately at zero capacity and full capacity, resulting in vibration and overheating. Surging occurs when the blower operates in a low volumetric range.

With greater emphasis on energy efficiency, high speed turbo blowers [see Figs. 5–72(c), (d), and (e)] are now being used for wastewater aeration applications. Using advancements in blower technology (adapted from the aircraft industry), these blowers are provided as complete packages that feature (1) aerodynamically designed impellers using turbine engine technology, (2) high speed direct-drive permanent magnet synchronous motors (PMSM) capable of speeds up to 75,000 rev/min (both air and water cooled), (3) air foil bearings that eliminate gears and contact between the stator and the drive shaft, and (4) integrated variable frequency drives. As a result, these units can operate efficiently over a wider range of operating conditions, as compared to conventional blower technologies. High speed turbo blowers are provided for capacities of up to 567 m^3/min (20,000 ft^3/min) and pressures of 103 kN/m^2 (15 lb_f/in.^{2}). Energy savings as high as 40 percent have been reported.

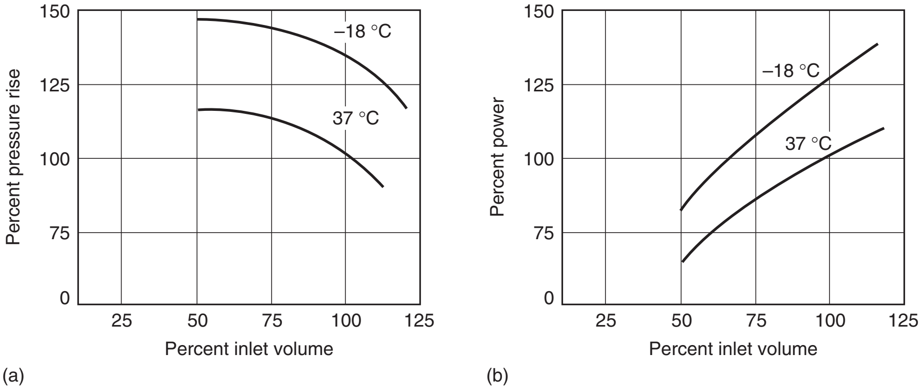
For discharge pressure applications above 55 kN/m^2 (8 lb_f/in.^{2}), for variable discharge pressure applications (such as sequencing batch reactors), and for capacities smaller than 425 m^3/min (15,000 ft^3/min) of free air per unit, rotary-lobe positive-displacement blowers are commonly used [see Figs. 5–72(f) and (g)]. The positive-displacement blower is a machine of constant capacity with variable pressure. The units cannot be throttled, but capacity control can be obtained by the use of multiple units or a variable-speed drive. Rugged inlet and discharge silencers are essential.

Blowers with inlet guide vane-variable diffusers mitigate some of the problems and considerations associated with standard centrifugal blowers. The design is based on a centrifugal blower operation that incorporates actuators to position the inlet guide vane and variable diffusers to vary blower flowrate and optimize efficiency. The blowers are especially well suited to applications with medium to high fluctuations in inlet temperature, discharge pressure, and flowrate. Blower capacities range from 85 to 1700 m^3/min (3000 to 60,000 ft^3/min) at pressures up to 170 kN/m^2 (25 lb_f/in.^{2}). Turndown rates of up to 40 percent of maximum capacity are possible without significant reduction in operating efficiency over the range of operation. Principal disadvantages are high initial cost and a sophisticated computer control system to ensure efficient operation.

The performance curve for a centrifugal blower in which pressure is plotted versus inlet air volume resembles the performance curve for a centrifugal pump. The performance curve typically is a falling-head curve where the pressure decreases as the inlet volume increases. Blowers are rated at standard air conditions, defined as a temperature of 20°C (68°F), a pressure of 760 mm Hg (14.7 lb_f/in.^{2}), and a relative humidity of 36 percent. Standard air has a specific weight of 1.20 kg/m^3 (0.0750 lb/ft^3). The air density affects the performance of a centrifugal blower; any change in the inlet air temperature and barometric pressure will change the density of the compressed air. The greater the gas density, the higher the pressure will rise. As a result, greater power is needed for the compression process (see Fig. 5–73). (Typical values for the specific weight of ambient air are presented

Figure 5–73

Characteristic curves for centrifugal blower at various inlet air temperatures: (a) pressure rise as a function inlet volume and (b) power rise as a function inlet volume.



in Appendix B.) Blowers must be selected to have adequate capacity for a hot summer day, and be provided with a driver with adequate power for the coldest winter weather. The power requirement for adiabatic compression is given in Eq. (5–77).

$$P_w = \frac{wRT_1}{28.97\,n\,e}\left[\left(\frac{p_2}{p_1}\right)^n - 1\right] \qquad \text{(SI units)} \qquad (5\text{–}77a)$$

$$P_w = \frac{wRT_1}{550\,n\,e}\left[\left(\frac{p_2}{p_1}\right)^n - 1\right] \qquad \text{(U.S. customary units)} \qquad (5\text{–}77b)$$

where P_w = power requirement of each blower, kW (hp)
w = weight of air flowrate, kg/s (lb/s)
R = universal gas constant for air, 8.314 J/mole·K (SI units)
53.3 ft·lb/(lb air)·°R (U.S. customary units)
T_1 = absolute inlet temperature, K (°R)
p_1 = absolute inlet pressure, atm ($lb_f/in.^2$)
p_2 = absolute outlet pressure, atm ($lb_f/in.^2$)
n = $(k - 1)/k$ where k is the specific heat ratio. For single-stage centrifugal blower power calculations a value of 1.395 is used for k for dry air and $n = 0.283$.
28.97 = molecular weight of dry air
550 = conversion factor from ft·lb/s to hp
e = efficiency (usual range for compressors is 0.70 to 0.90)

Air Piping. Air piping consists of mains, valves, meters, and other fittings that transport compressed air from the blowers to the air diffusers. Because the pressures are low [less than 70 kN/m^2 (10 $lb_f/in.^2$)], lightweight piping can be used.

The piping should be sized so that losses in air headers and diffuser manifolds are small in comparison to the losses in the diffusers. Typically, if headlosses in the air piping between the last flow-split device and the farthest diffuser are less than 10 percent of the headloss across the diffusers, good air distribution through the aeration basin can be maintained. Valves and control orifices are an important consideration in piping design (WEF, 1998b). Typical velocities in air piping are given in Table 5–29.

Friction losses in air piping can be calculated using the Darcy-Weisbach equation written in the following form:

$$h_L = f\frac{L}{D}h_i \qquad (5\text{–}78)$$

Table 5–29

Typical air velocities in aeration header pipes

Pipe diameter		Velocity[a]	
in.	mm	ft/min	m/min
1–3	75–225	1200–1800	360–540
4–10	100–250	1800–3000	540–900
12–24	300–600	2700–4000	800–1200
30–60	750–1500	3800–6500	1100–2000

[a] At standard conditions.

where h_L = friction loss, mm (in.) of water
f = dimensionless friction factor obtained from Moody diagram (see Appendix G) based on relative roughness. It is recommended that f be increased by at least 10 percent to allow for an increase in friction factor as the pipe ages.
L = equivalent length of pipe, m (ft)
D = pipe diameter, m (ft)
h_i = velocity head of air, mm (in.) of water

The friction factor for steel pipes carrying air can be approximated by Eq. (5–79) (McGhee, 1991).

$$f = \frac{0.029(D)^{0.027}}{Q^{0.148}} \tag{5–79}$$

where Q = airflow, m^3/min under prevailing pressure and temperature conditions
D = pipe diameter, m

Headloss in a straight pipe can be computed by substituting Eq. (5–79) in Eq. (5–78), which yields

$$h_L = 9.82 \times 10^{-8}\left(\frac{fLTQ^2}{PD}\right) \tag{5–80}$$

where P = air discharge pressure, atm
T = temperature in pipe, K [from Eq. (5–81)]

$$T = T_o(P/P_o)^n \tag{5–81}$$

where T_o = ambient air temperature, K (maximum summer air temperature)
P_o = ambient barometric pressure, atm

Losses in elbows, tees, valves, etc., can be computed as a fraction of velocity head using headloss coefficient K values given in the companion volume to this text (Metcalf & Eddy, 1981) or in standard hydraulic texts. Minor losses can also be computed as equivalent lengths of straight pipe, as follows:

$$L = 55.4\ CD^{1.2} \tag{5–82}$$

where L = equivalent length of pipe, L (m)
D = pipe diameter, L (m)
C = resistance factor (see Table 5–30)

Meter losses can be estimated as a fraction of the differential velocity head across the meter, depending on the type of meter. Losses in air filters, blower silencers, and check

Table 5–30

Resistance factors for fittings in aeration piping systems

Fitting	*C* factor
Long radius ell or run of standard tee	0.33
Medium radius ell or run of tee reduced by 25 percent	0.42
Standard ell or run of tee reduced 50 percent	0.67
Tee through side outlet	1.33
Gate valves	0.25
Globe valve	2.00
Angle valve	0.90

valves should be obtained from equipment manufacturers, but approximate values given in Table 5–31 can be used as a guide (Qasim, 1999).

The discharge pressure at the blowers will be the sum of the above losses, the depth of water over the air diffusers, and the loss through the diffusers.

Because of the high temperature of the air discharged by blowers [60 to 80°C (140 to 180°F)], condensation in the air piping is not a problem, except where piping is submerged in the wastewater. It is essential, however, that provisions be made for pipe expansion and contraction. Where porous diffusers are used, pipes must be made of nonscaling materials or must be lined with material that will not corrode. Pipe materials are often stainless steel, fiberglass, or plastics suitable for higher temperatures. Other materials used include mild steel or cast iron with external coatings (e.g., coal-tar epoxy or vinyl). Interior surfaces include cement lining or coal tar or vinyl coatings.

Mechanical Aerators

Mechanical aerators are commonly divided into two groups based on major design and operating features: aerators with vertical axis and aerators with horizontal axis. Both groups are further subdivided into surface and submerged aerators. In surface aerators, oxygen is entrained from the atmosphere; in submerged aerators, oxygen is entrained from the atmosphere and, for some types, from air or pure oxygen introduced in the tank bottom. In either case, the pumping or agitating action of the aerators helps to keep the contents of the aeration tank or basin mixed. In the following discussion, the various types of aerators will be described, along with aerator performance and the energy requirement for mixing.

Surface Mechanical Aerators with Vertical Axis. Surface mechanical aerators with a vertical axis are designed to induce either updraft or downdraft flows through

Table 5–31

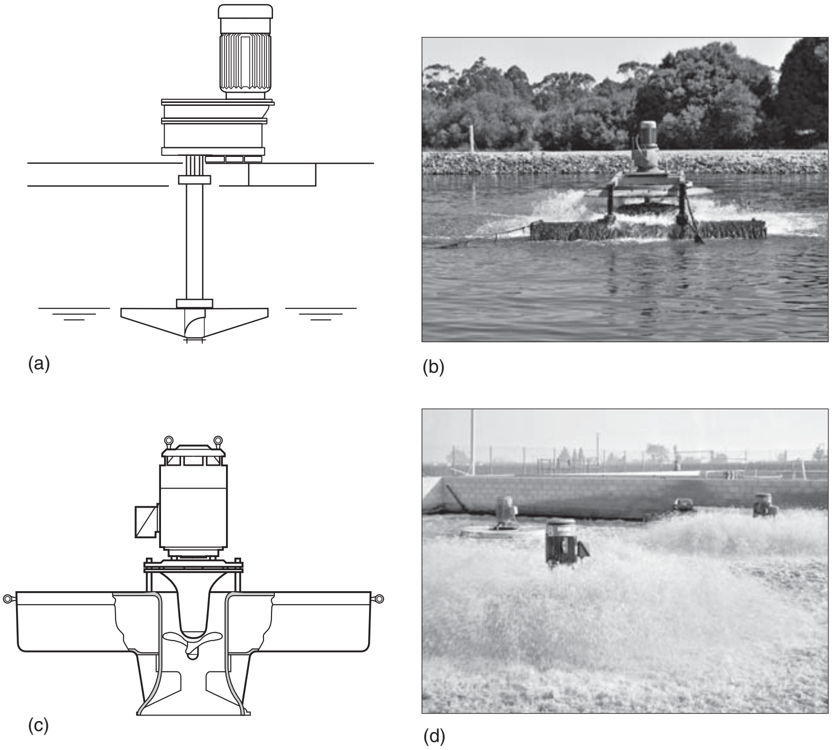
Typical headlosses through air filters, blower silences, and check valves[a]

Device	Headloss	
	in.	mm
Air filter	0.5–3	13–76
Silencer		
Centrifugal blower	0.5–1.5	13–38
Positive displacement blower	6–8.5	152–216
Check valve	0.8–8	20–203

[a] Adapted from Qasim (1999).

Figure 5–74

Typical mechanical aerators: (a) schematic low-speed surface aerator, (b) view of low-speed surface aerator mounted on floats used in biological treatment system, (c) schematic high-speed surface aerator, and (d) view of high-speed surface aerators in a lined earthen aerobic digester.



a pumping action (see Fig. 5–74). Surface aerators consist of submerged or partially submerged impellers that are attached to motors mounted on floats or on fixed structures. The impellers are fabricated from steel, cast iron, noncorrosive alloys, and fiberglass-reinforced plastic and are used to agitate the wastewater vigorously, entraining air in the wastewater and causing a rapid change in the air-water interface to facilitate dissolution of the air. Surface aerators may be classified according to the type of impeller used—centrifugal, radial-axial, or axial—or the speed of rotation of the impeller: low and high speed. Centrifugal impellers belong to the low-speed category; the axial-flow impeller type aerators operate at high speed. In low-speed aerators, the impeller is driven through a reduction gear by an electric motor [see Fig. 5–74(a)]. The motor and gearbox are usually mounted on a platform that is supported either by piers extending to the bottom of the tank or by beams that span the tank. Low-speed aerators may also be mounted on floats. In high-speed aerators, the impeller is coupled directly to the rotating element of the electric motor [see Fig. 5–74(c)]. High-speed aerators are almost always mounted on floats. These units were originally developed for use in ponds or lagoons where the water surface elevation fluctuates, or where a rigid support would be impractical. Surface aerators may be obtained in sizes from 0.75 to 100 kW (1 to150 hp).

Submerged Mechanical Aerators with Vertical Axis. Most surface mechanical aerators are upflow types that rely on violent agitation of the surface and air entrainment for their efficiency. With submerged mechanical aerators, however, air or pure oxygen may also be introduced by diffusion into the wastewater beneath the impeller or downflow of radial aerators [see Fig. 5–75(a)]. The impeller is used to disperse the air bubbles and mix the tank contents. A draft tube may be used with either upflow or downflow models to control the flow pattern of the circulating liquid within the aeration tank [see Fig. 5–75(b)]. The draft tube is a cylinder, usually with flared ends, mounted

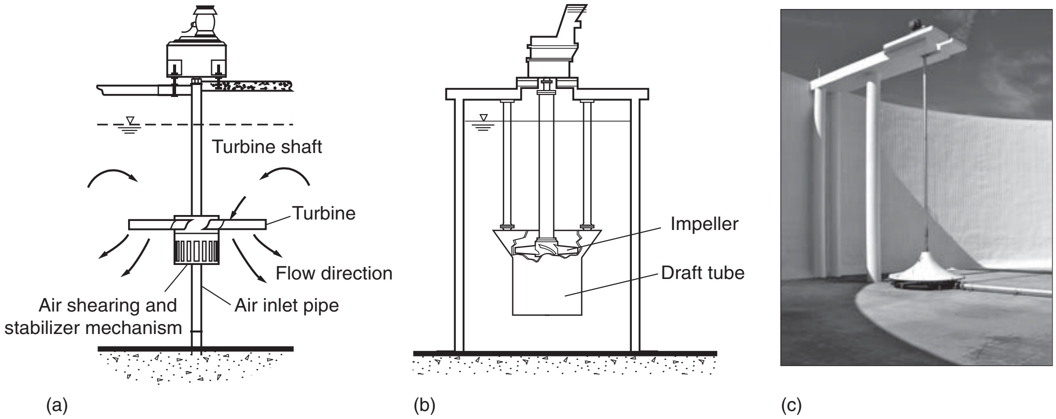


Figure 5–75

Typical submerged mechanical aerators: (a) turbine type with supplementary air or oxygen feed introduced below the turbine, (b) draft tube turbine aerator equipped with an air sparger (adapted from Philadelphia Mixer), and (c) hyperboloid mixer/aerator (courtesy of INVENT Env. Technologies, Inc.).

concentrically with the impeller. The length of the draft tube depends upon the aerator manufacturer. Submerged mechanical aerators of such type may be obtained in sizes from 0.75 to 100kW (1 to 150 hp).

The hyperboloid mixer/aerator [see Fig. 5–75(c)] is a downflow impeller with separate air or gas supply via blower or system pressure. The separation from mixing, gas dispersion and gas feed results in significantly higher efficiencies compared to self-inducing aerators. Hyperboloid mixer/aerators are available in sizes from 7.5 to 37 kW (10 to 50 hp).

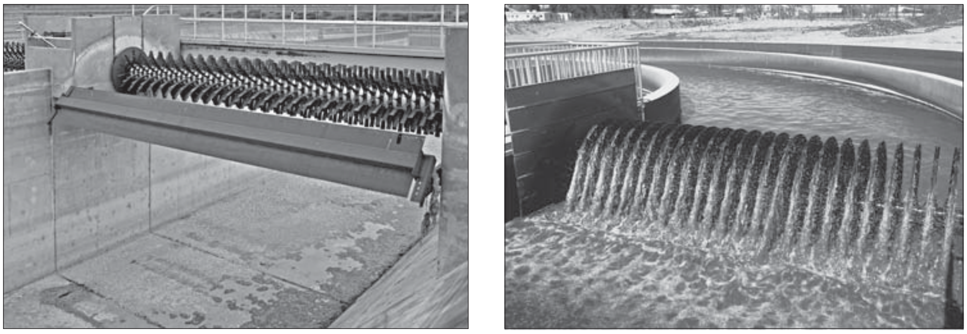
Mechanical Aerators with Horizontal Axis. Mechanical aerators with horizontal axis are divided into two groups: surface and submerged aerators. The surface aerator is patterned after the original Kessener brush aerator, a device used to provide both aeration and circulation in oxidation ditches. The brush-type aerator had a horizontal cylinder with bristles mounted just above the water surface. The bristles were submerged in the water and the cylinder was rotated rapidly by an electric motor drive, spraying wastewater across the tank, promoting circulation, and entraining air in the wastewater. Angle steel, steel of other shapes, or plastic bars or blades are now used instead of bristles. A typical horizontal-axis surface aerator is shown on Fig. 5–76.

Submerged horizontal-axis aerators are similar in principle to surface aerators except disks or paddles attached to rotating shafts are used to agitate the water. The disk aerator has been used in numerous applications for channel and oxidation ditch aeration. The disks are submerged in the wastewater for approximately one-eighth to three-eighths of the diameter and enter the water in a continuous, nonpulsating manner. Recesses in the disks introduce entrapped air beneath the surface as the disk turns. Spacing of the disks can vary depending on the oxygen and mixing requirements of the process. Typical power requirements are reported as 0.1 to 0.75 kW/disk (0.15 to 1.00 hp/disk) (WPCF, 1988).

Aerator Performance. Mechanical aerators are rated in terms of their oxygen transfer rate expressed as kilograms of oxygen per kilowatt-hour (pounds of oxygen per

Figure 5–76

Horizontal axis aerators: (a) rotary brush (also known as a Kessener brush) and (b) disk aerators.


(a) (b)

horsepower-hour) at standard conditions. Standard conditions exist when the temperature is 20°C, the dissolved oxygen is 0.0 mg/L, and the test liquid is tap water. Testing and rating are normally done under non-steady-state conditions using fresh water, deaerated with sodium sulfite. Commercial-size surface aerators range in efficiency from 1.20 to 2.4 kg O_2/kW·h (2 to 4 lb O_2/hp·h). Oxygen transfer data for various types of mechanical aerators are reported in Table 5–32. Efficiency claims for aerator performance should be accepted by the design engineer only when they are supported by actual test data for the actual model and size of aerator under consideration.

Energy Requirement for Mixing in Aeration Systems

As with diffused-air systems, the size and shape of the aeration tank are very important if good mixing is to be achieved. Aeration tanks may be square or rectangular and may contain one or more aerators. The depth and width of the aeration tanks for mechanical surface aerators are dependent on aerator size, and typical values are given in Table 5–33. Depths up to 11 m (35 ft) have been used with submerged-draft tube and hyperboloid mixers.

Table 5–32

Typical ranges of oxygen transfer capabilities for various types of mechanical aerators

	Transfer rate, lb O_2/hp·h		Transfer rate, kg O_2/kW·h	
Aeration system	**Standard**[a]	**Field**[b]	**Standard**[a]	**Field**[b]
Surface low speed	2.5–3.5	1.2–2.4	1.5–2.1	0.7–1.5
Surface low speed with draft tube	2.0–4.6	1.2–2.1	1.2–2.8	0.7–1.3
Surface high speed	1.8–2.3	1.2–2.0	1.1–1.4	0.7–1.2
Submerged turbine with draft tube	2.0–3.3	1.2–1.8[c]	1.2–2.0	0.6–1.1
Submerged turbine	1.8–3.5		1.1–2.1	
Submerged turbine with sparger	2.0–3.3	1.2–1.8[c]	1.2–2.0	0.7–1.0
Horizontal rotor	1.5–3.6	0.8–1.8	1.5–2.1	0.5–1.1

[a] Standard conditions: tap water 20°C (68°F); at 101.325 kN/m^2 (14.7 lb$_f$/in.^{2}); and initial dissolved oxygen = 0 mg/L.
[b] Field conditions: wastewater, 15°C (59°F); altitude 150 m (500 ft), $\alpha = 0.85$, $\beta = 0.9$; operating dissolved oxygen = 2 mg/L.
[c] The α values may be lower than 0.85; reported ranges vary from 0.3 to 1.1.

Table 5–33

Typical aeration tank dimensions for mechanical surface aerators

Aerator size		Tank dimensions			
		U.S. customary units		SI units	
hp	kW	Depth, ft	Width, ft	Depth, m	Width, m
10	7.5	10–12	30–40	3–3.5	9–12
20	15	12–14	35–50	3.5–4	10–15
30	22.5	13–15	40–60	4–4.5	12–18
40	30	12–17	45–65	3.5–5	14–20
50	37.5	15–18	45–75	4.5–5.5	14–23
75	55	15–20	50–85	4.5–6	15–26
100	75	15–20	60–90	4.5–6	18–27

In diffused-air systems, the air requirement to ensure good mixing varies from 20 to 30 $m^3/10^3\ m^3\cdot min$ (20 to 30 $ft^3/10^3\ ft^3\cdot min$) of tank volume, for a spiral-roll aeration pattern. For a grid system of aeration in which the diffusers are installed uniformly along the aeration basin bottom, mixing rates of 10 to 15 $m^3/10^3\ m^3\cdot min$ (10 to 15 $ft^3/10^3\ ft^3\cdot min$) have been suggested (WPCF, 1988). Fine pore diffusers installed in a grid Pattern typically require a minimum aeration rate of 0.12 scfm/square foot for mixing (U.S. EPA, 1989). Typical power requirements for maintaining a completely mixed flow regime with mechanical aerators vary from 20 to 40 $kW/10^3\ m^3$ (0.75 to 1.50 $hp/10^3\ ft^3$), depending on the type and design of the aerator, the nature and concentration of the suspended solids, the temperature, and the geometry of the aeration tank, lagoon, or basin. In the design of aerated lagoons for the treatment of domestic wastewater, it is extremely important that the mixing power requirement be checked because, in most instances, it will be the controlling factor.

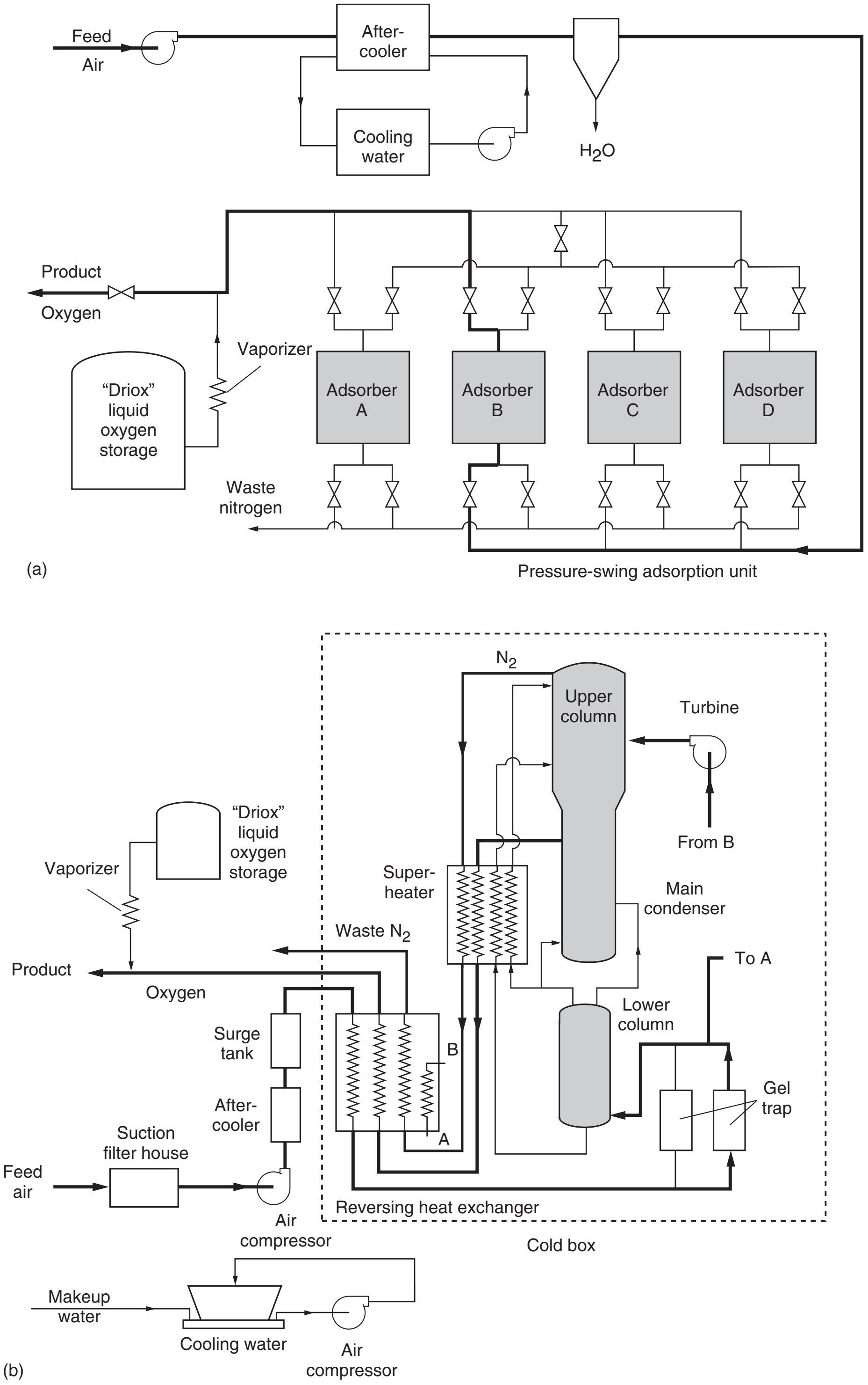
Generation and Dissolution of High-Purity Oxygen

After the quantity of oxygen required is determined, it is necessary, where high-purity oxygen is to be used, to specify the type of oxygen generator that will best serve the needs of the plant. There are two basic oxygen generator designs: (1) a pressure swing adsorption (PSA) system for smaller and more common plant sizes (less than 150,000 m^3/d (40 Mgal/d)), and (2) the traditional cryogenic air-separation process for large applications. Liquid oxygen can also be trucked in and stored onsite in a liquid form.

Pressure-Swing Adsorption. The pressure-swing adsorption system uses a multibed adsorption process to provide a continuous flow of oxygen gas (U.S. EPA, 1974). A schematic diagram of the four-bed system is shown on Fig. 5–77(a). The operating principle of the pressure-swing adsorption generator is that the oxygen is separated from the feed air by adsorption at high pressure, and the adsorbent is regenerated by "blowdown" to low pressure. The process operates on a repeated cycle having two basic steps, adsorption and regeneration. During the adsorption step, feed air flows through one of the adsorber vessels until the adsorbent is partially loaded with impurity. At that time the feed-air flow is switched to another adsorber, and the first adsorber is regenerated. During regeneration, the impurities are cleaned from the adsorbent so that the bed will be available again for the adsorption step. Regeneration is carried out by depressurizing to atmospheric pressure, purging with some of the oxygen, and repressurizing back to the pressure of the feed air.

Figure 5–77

Schematic diagrams for the generation of oxygen used in high-purity oxygen activated sludge process: (a) pressure-swing adsorption and (b) cryogenic air separation.



Cryogenic Air Separation. The cryogenic air separation process involves the liquefaction of air, followed by fractional distillation to separate it into its components (mainly nitrogen and oxygen). A schematic diagram of this process is shown on Fig. 5–77(b). First, the entering air is filtered and compressed, and then it is fed to the reversing heat exchangers, which perform the dual function of cooling and removing the water vapor and carbon dioxide by freezing these mixtures out into the exchanger surfaces. Periodically switching or reversing the feed air and the waste nitrogen streams through identical passes of the exchangers to regenerate their water vapor and carbon dioxide removal capacity accomplishes this process.

Next, the air is processed through "cold and gel traps," which are adsorbent beds that remove the final traces of carbon dioxide as well as most hydrocarbons from the feed air. The processed air is then divided into two streams. The first stream is fed directly to the lower column of the distillation unit. The second stream is returned to the reversing heat exchangers and partially warmed to provide the required temperature difference across the exchanger. This stream is then passed through an expansion turbine and fed into the upper column of the distillation unit. An oxygen-rich liquid exits from the bottom of the lower column and the liquid nitrogen exits from the top. Both streams are then subcooled and transferred to the upper column. In this column, the descending liquid phase becomes progressively richer in oxygen, and the liquid that subsequently collects in the condenser reboiler is the oxygen product stream. This oxygen is recirculated continually through an adsorption trap to remove all possible residual traces of hydrocarbons. The waste nitrogen exits from the top portion of the upper column and is heat exchanged along with the oxygen product to recover all available refrigeration and to regenerate the reversing heat exchangers.

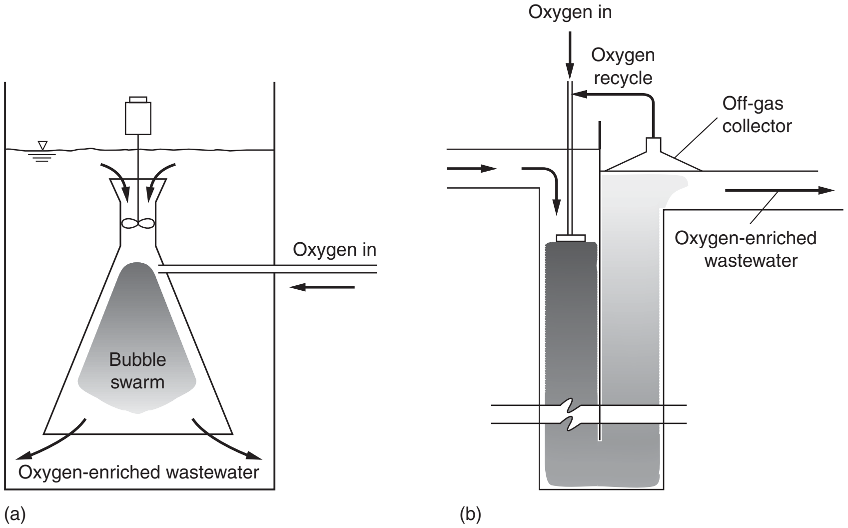
Dissolution of Commercial Oxygen. Oxygen is very insoluble in water—even pure oxygen—and requires special considerations to ensure high absorption efficiency. Oxygen dissolution equipment designed for air only optimizes energy consumption because the air is free and efficient oxygen absorption is not relevant. However, because of the cost of commercial oxygen, the facilities used for its dissolution must be designed to both efficiently absorb the commercial oxygen as well as minimize the unit energy consumption. These requirements rule out the more common aeration equipment alternatives.

Dissolution Time. A key feature that must be incorporated into a commercial oxygen dissolution system is oxygen retention time. To optimize the absorption of pure oxygen it has been found that a detention time of about 100 s is required. Further, two-phase flow must be maintained to avoid the coalescence of the oxygen bubbles to maintain absorption efficiency. Unfortunately, some pure oxygen dissolution systems consume as much energy to dissolve a ton of pure oxygen as standard surface aerators consume in dissolving a ton of oxygen from air.

Speece Cone (Downflow Bubble Contactor). A system that incorporates prolonged oxygen bubble contact time and high rates of oxygen transfer is a cone-shaped chamber, downflow bubble contact aerator termed the Speece cone [see Fig. 5–78(a)]. Water enters the chamber at the apex with a velocity of approximately 3 m/s (10 ft/s). This inlet velocity provides the energy to maintain a two-phase bubble swarm in the cone, ensuring a very high bubble/water interface, resulting in a proportionately high gas transfer rate. The expanding horizontal cross section of the cone reduces the downward flow velocity of the water to less than 0.3 m/s (1 ft/s). Because the bubbles have a nominal buoyant velocity of about 0.3 m/s, if the downflow velocity of the water is reduced below the buoyant velocity of the bubbles, they will remain indefinitely in the cone, thus satisfying the required bubble residence time. The water, however, has a residence time of about 10 s,

Figure 5–78

Pure oxygen dissolution systems: (a) Speece cone (downflow bubble contactor) and (b) U-tube contactor (Speece and Tchobanoglous, 1990).



reflecting a relatively small volume of reactor cone. This system incorporates the desired features of relatively small size, high rate of oxygen transfer, and more than adequate bubble residence time (Speece and Tchobanoglous, 1990).

U-Tube Contactor. Another oxygen transfer system that incorporates desirable features for efficient dissolution of commercial oxygen with low unit energy consumption is the U-tube [see Fig. 5–78(b)]. At a depth of 30 m (100 ft) and a throughput velocity of 2.5 m/s (8 ft/s), the residence time is 25 s. Because a contact time of 25 s is low, off-gas recycle back through the system can be used to increase the contact time to about 100 s where efficient absorption occurs. The energy requirements are low because the bubble/water mixture is pumped through a filled U-tube pipe that is hydrostatically pressurized by its vertical configuration. Use of the U-tube enhances gas transfer significantly. Energy consumption is about 54 kWh/Mg O_2 (60 kWh/ton) while producing an effluent dissolved oxygen of 60 mg/L (Speece and Tchobanoglous, 1990).

Conventional Diffused Aeration. Conventional diffused aeration or surface aerators must operate in a closed headspace to absorb commercial oxygen efficiently. A concrete cover usually is placed over the aeration tank to enclose the headspace.

Postaeration

The requirement for postaeration systems has developed in recent years with the introduction of effluent standards and permits that include high dissolved oxygen levels (5 to 8 mg/L). Dissolved oxygen levels have become standard for discharge to water-quality-limited stream sections and to effluent dominated waters. The regulatory intent is to ensure that low dissolved oxygen levels in the treated effluent do not cause immediate depression after mixture with the waters of the receiving stream. To meet postaeration requirements, three methods are most commonly used: (1) diffused air, (2) mechanical aeration, (3) cascade aeration, and (4) The Speece cone, described above, can also be used for postaeration and for reaerating reclaimed water-storage reservoirs.

Diffused-Air Aeration. In larger treatment plants, diffused aeration systems may be more appropriate [see Fig. 5–79(a)]. Coarse- or fine-bubble diffusers may be used. Depending

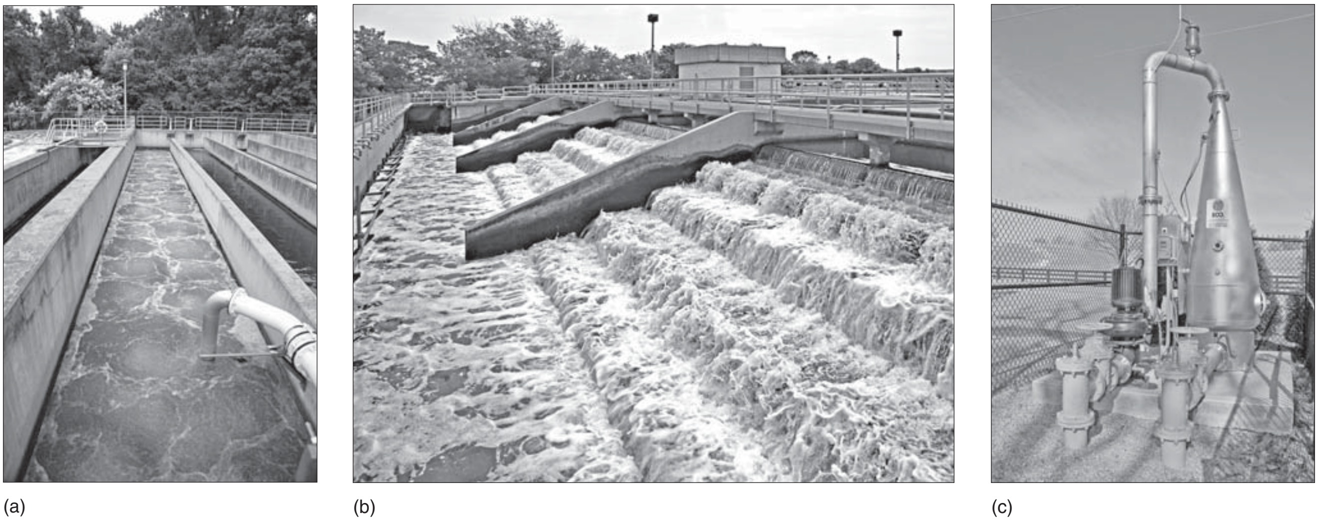

(a) (b) (c)

Figure 5–79

Postaeration aeration systems: (a) diffused air, (b) cascade step aeration, and (c) Speece cone.

on the depth of submergence, transfer efficiencies of 5 to 8 percent may be attained with non-porous diffusers, and 15 to 25 percent with fine-pore diffusers. After secondary treatment, the alpha factors should range from 0.85 to 0.95 for nonporous systems, and from 0.70 to 0.85 for fine-pore systems.

EXAMPLE 5–15 **Derivation of Equation for Estimating Diffused-Air Requirements for Post-aeration** Develop an expression that can be used to estimate the diffused-air requirement for the postaeration of effluent following chlorination. Assume that aeration will be accomplished in a plug-flow reactor. Using the expression developed, estimate the amount of air in m^3/s to increase the DO from 1.5 to 5 g/m^3 at 20°C. The wastewater flowrate is 3800 m^3/d (1.0 Mgal/d). Assume the oxygen transfer efficiency is 6 percent and that C_s at 20°C is equal to 9.09 g/m^3 (from Appendix E).

Solution—Part 1 Expression Development

1. Write an equation for the oxygen-solution rate. The appropriate expression, similar to Eq. (5–57), is

$$r_m = \frac{dm}{dt} = K'_T(C_s - C)$$

where K'_T = overall mass transfer coefficient for the given conditions
$K'_T = K'_{20} \times (1.024)^{T-20}$

2. Write an expression for the oxygen-transfer efficiency. The efficiency may be defined as

$$E = \frac{(dm/dt)_{20°\text{C},C=0}}{M}$$

where E = oxygen transfer efficiency
$(dm/dt)_{20°C,C=0}$ = oxygen-solution rate at 20°C and zero dissolved oxygen
M = mass rate at which oxygen is introduced

3. Develop a differential expression for the mass rate at which oxygen is introduced. The mass rate at which oxygen is introduced is given by

$$M = \frac{1}{E}\left(\frac{dm}{dt}\right)_{20°\mathrm{C},C=0}$$
$$= \frac{1}{E}\left(\frac{dm}{dt}\right)_T \frac{(dm/dt)_{20°\mathrm{C},C=0}}{(dm/dt)_T}$$

Substituting for $(dm/dt)_{20,\,C=0}$ and $(dm/dt)_T$ yields

$$M = \frac{1}{E}\left(\frac{dm}{dt}\right)_T \frac{(C_s)_{20°\mathrm{C}}}{(C_s - C)_T\,(1.024)^{T-20}}$$

If the expression is applied to an infinitesimal transverse segment of the tank and QdC is substituted for dm/dt (note that $V(dC/dt) = dm/dt$ and $Q = V/dt$), then the differential form of the above expression can be rewritten as

$$dm = \frac{Q(C_s)_{20°\mathrm{C}}}{E(1.024)^{T-20}}\left(\frac{dC}{C_s - C}\right)_T$$

4. Integrate the differential expression derived in step 3. Integrating the expression from the inlet of the tank where $C = C_i$ to the outlet of the tank where $C = C_o$ and $M = 0$ to $M = M$:

$$\int_0^M dM = \frac{Q(C_s)_{20°\mathrm{C}}}{E(1.024)^{T-20}}\int_{C_i}^{C_o}\frac{dC}{C_s - C}$$

$$M = \frac{Q(C_s)_{20°\mathrm{C}}}{E(1.024)^{T-20}}\left[\ln\left(\frac{C_s - C_i}{C_s - C_o}\right)_T\right]$$

5. Rewrite the equation derived in step 4 in a more practical format. Note that the density of air is 1.23 kg/m³ and air contains about 23 percent oxygen by weight. The mass rate at which oxygen is introduced, expressed as g/s, is

$$M,\ \mathrm{g/s} = \left(Q, \frac{\mathrm{m^3\ air}}{\mathrm{s}}\right)\left(\frac{1.23\ \mathrm{kg}}{\mathrm{m^3\ O_2}}\right)\left(\frac{0.23\ \mathrm{kg\ O_2}}{\mathrm{kg\ air}}\right)\left(\frac{10^3\ \mathrm{g}}{1\ \mathrm{kg}}\right)$$

Thus, the rate of oxygen input, in terms of the equivalent air flowrate, expressed in m³/s, is equal to

$$Q_a = (3.53 \times 10^{-3})\frac{Q(C_s)_{20°\mathrm{C}}}{E(1.024)^{T-20}}\left[\ln\left(\frac{C_s - C_i}{C_s - C_o}\right)\right]$$

where Q_a = required air flowrate, m³/s
Q = wastewater flowrate, m³/s
C_s = saturation concentration of oxygen at 20°C, g/m³

Solution—Part 2 Estimate the mass air flowrate

6. Estimate the amount of air required using the expression developed in Step 5 above.

$$Q_a = (3.53 \times 10^{-3})\frac{Q(C_s)_{20°\mathrm{C}}}{E(1.024)^{T-20}}\left[\ln\left(\frac{C_s - C_i}{C_s - C_o}\right)\right]$$

a. Summarize known values
Q = 3800 m³/d = 0.044 m³/s
C_s = 9.09 g/m³
C_i = 1.5 g/m³

$C_o = 5.0\ \text{g/m}^3$
$E = 0.06$

b. Substitute known values and solve for Q_a.

$$Q_a = (3.53 \times 10^{-3})\frac{(0.044\ \text{m}^3/\text{s})(9.09)}{0.06(1.024)^{20-20}}\left[\ln\left(\frac{9.09 - 1.5}{9.09 - 5.0}\right)\right] = 1.30\ \text{m}^3/\text{s}$$

$$= 78\ \text{m}^3/\text{min}$$

Comment The value of Q_a is usually multiplied by a factor of 1.1 to account for the fact that the saturation value of oxygen in wastewater is about 95 percent of that in distilled water, and to account for the difference in the transfer rates.

Cascade Aeration. If site constraints and hydraulic conditions permit gravity flow, the least costly method to raise dissolved oxygen levels is with the use of cascade aeration [see Fig. 5–79(b)]. Cascade aeration consists of using the available discharge head to create turbulence as the wastewater falls in a thin film over a series of concrete steps. Performance depends on the initial dissolved oxygen level, required discharge dissolved oxygen, and wastewater temperature. Typical design information is given in Table 5–34. Where the cascade aeration facility joins the chlorine contact basin, the postaeration structure may be made equal to the chlorine contact basin width for ease of construction.

The most commonly used method for determining required cascade height is based on the following equations developed by Barrett (1960).

$$H = \frac{R - 1}{0.361\,ab(1 + 0.046 \times T)} \quad \text{SI units} \tag{5–83a}$$

$$H = \frac{R - 1}{0.11\,ab(1 + 0.046 \times T)} \quad \text{U.S. customary units} \tag{5–83b}$$

where R = deficit ratio = $\dfrac{C_s - C_o}{C_s - C}$

C_s = dissolved oxygen saturation concentration of the wastewater at temperature T, mg/L
C_o = dissolved oxygen concentration of the postaeration influent, mg/L
C = required final dissolved oxygen level after postaeration, mg/L
a = water-quality parameter equal to 0.8 for a wastewater-treatment plant effluent
b = weir geometry parameter for a weir, $b = 1.0$; for steps, $b = 1.1$; for step weir, $b = 1.3$

Table 5–34
Typical design information for a cascade type post aeration system

	U.S. customary units			SI units		
Parameter	**Units**	**Range**	**Typical**	**Units**	**Range**	**Typical**
Hydraulic loading rate at average design flow	gal/ft of width·d	100,000–500,000	240,000	m^3/m of width·d	1240–6200	3000
Step dimensions						
Height	in.	6–12	8	mm	150–300	200
Length	in.	12–24	18	mm	300–600	450
Cascade height	ft	6–16		m	2–5	

T = water temperature, °C
H = height through which water falls, m (ft)

A key element is the proper selection of the critical wastewater temperature, which affects the dissolved oxygen saturation concentration C_s. The effect of temperature is illustrated in Example 5–16.

EXAMPLE 5–16 **Calculation of Cascade Aeration Height** Calculate the height of a step cascade aeration system for a wastewater treatment plant in a warm climate where the wastewater temperature averages 20°C in the winter and 25°C in the summer. The dissolved oxygen in the influent to the postaeration system C_o is 1.0 mg/L and the required final dissolved oxygen concentration C is 6.0 mg/L.

Solution

1. Determine the dissolved oxygen saturation concentration C_s at the wastewater temperatures.
    a. From Appendix E, the dissolved oxygen at 20°C = 9.09 mg/L; and 8.26 mg/L at 25°C.
2. Calculate the cascade height for T = 20°C using Eq. (5–83a).
    a. Calculate the deficit ratio.

$$R = \text{deficit ratio} = \frac{C_s - C_o}{C_s - C} = \frac{9.09 - 1.0}{9.09 - 6.0} = 2.62$$

    b. Calculate the cascade height, assuming steps ($b = 1.1$).

$$H = \frac{R - 1}{0.361\,ab\,(1 + 0.046T)}$$

$$H = \frac{2.62 - 1}{0.361\,(0.8)(1.1)\,(1 + 0.046 \times 20)}$$

$$H = 2.66\ \text{m}\ (8.73\ \text{ft})$$

3. Calculate the cascade height for T = 25°C.
    a. Calculate the deficit ratio

$$R = \frac{8.26 - 1.0}{8.26 - 6.0} = 3.21$$

    b. Calculate the cascade height, assuming steps and using the same computation procedure as in 2b above.

$$H = 3.62\ \text{m}\ (11.9\ \text{ft})$$

Comment The increased wastewater temperature increases the dissolved oxygen deficit ratio and therefore affects the height of the cascade. The maximum wastewater temperatures should be checked to ensure the cascade height is not underdesigned. The physical dimensions of each step will depend on the site conditions and should be configured relative to the typical values given in Table 5–34. If a step height of 300 mm (12 in.) is selected, the required number of steps would be 12.

Mechanical Aeration. Two major types of mechanical aeration equipment are commonly used for postaeration systems: low-speed surface aerators and submerged turbine aerators. Low-speed surface aerators are preferred because they are usually the most economical, except where high oxygen transfer rates are required. For high oxygen transfer rates, submerged turbine units are preferred. Most installations consist of two or more aerators in rectangular basins. Detention times for postaeration using either mechanical or diffused-air aeration are usually 10 to 20 min at peak flowrates.

Aeration Using the Speece Cone. In postaeration with a Speece cone, a sidestream of the main flow is oxygenated and then reblended with the main flow before discharge. In open reservoirs used for storing reclaimed water, problems occur due to temperature stratification, low dissolved oxygen, and release of odors, principally hydrogen sulfide (see Sec. 13–4). The Speece cone, which has a very high oxygen transfer rate and is shown on Fig. 5–79(c), is ideally suited for reservoir aeration and temperature destratification. For reservoir aeration applications, compressed air can be used instead of high-purity oxygen.

PROBLEMS AND DISCUSSION TOPICS

5–1 A vertical bar screen is constructed of 20-mm diameter bars that have a clear spacing of 25 mm. Determine the headloss and velocity through a clean screen when the channel approach velocity is 1.0 m/s. If the screen is inclined at a 50, 55, and 60° angle from the horizontal (angle to be selected by instructor), how will the headloss and velocity through the clean screen be impacted?

5–2 Size the channel for a mechanically cleaned coarse bar rack for a flowrate of 40,000 m^3/d. The bar rack will be set at an angle of 75 degrees from horizontal and will have clear bar spacing of 12 mm. The bar rack uses 12-mm-wide by 25-mm deep bars. The upstream channel velocity should be greater than 0.4 m/s. Determine the headloss through the bar rack for the clean screen and for a screen that is 50 percent clogged.

5–3 The contents of a tank are to be mixed with a turbine impeller that has six flat blades. The diameter of the impeller is 3 m, and the impeller is installed 1.25 m above the bottom of a 6-m deep tank. If the temperature is 30°C and the impeller is rotated at 30 rev/min, what will be the power consumption? Find the Reynolds number using Eq. (5–11).

5–4 It is desired to flash-mix some chemicals with incoming wastewater that is to be treated. Mixing is to be accomplished using a flat-paddle mixer 500 mm in diameter having six blades. If the temperature of the incoming water is 10°C and the mixing chamber power number is 1.70, determine, (a) the speed of rotation when the Reynolds number is approximately 100,000, (b) why it is desirable to have a high Reynolds number in most mixing operations, (c) the required mixer motor size, assuming an efficiency factor of 20 percent.

5–5 Assuming that a given flocculation process can be defined by a first-order reaction ($r_N = -kN$), complete the following table (instructor to select data set) assuming the process is occurring in a plug-flow reactor with a detention time of 10 min. What would the value be after 5 min if a batch reactor were used instead, assuming the rate constant is the same?

	Particles, no/unit volume		
Time, min	Data set 1	Data set 2	Data set 3
0	10	40	20
5	?	?	?
10	3	5	2

5–6 If the steady-state effluent from a complete-mix reactor used as a flocculator contained 3 particles/unit volume, determine the concentration of particles in the effluent 5 min after the process started before steady-state conditions are reached. Assume that the influent contains 10 particles/unit volume, the detention time in the reactor is equal to 10 min, and that first-order kinetics apply ($r_N = -kN$).

5–7 An air flocculation system is to be designed. If a G value of 60 s^{-1} is to be used, estimate the air flowrate that will be necessary for a 200 m^3 flocculation chamber. Assume the depth of the flocculation basin is to be 4 m and that the air is released 0.25 m above the tank bottom.

5–8 Determine the required air flowrate to accomplish the flocculation operation in Example 5–3 pneumatically. Assume the air will be released at a depth of 3 m.

5–9 Derive Stokes' law by equating Eq. (5–23) to the particle mass.

5–10 Determine the settling velocity in m/s of a sand particle with a specific gravity of 2.65 and a diameter of 1 mm. Assume that the Reynolds number is 275.

5–11 Prepare a spreadsheet for computing the settling velocity of a spherical particle for the data given in Example 5–4. What are the values of Reynolds number and coefficient of drag necessary to have the assumed velocities agree with the calculated velocities?

5–12 Design a vortex type grit removal facility to handle a peak flowrate of 40,000 m^3/d. No grit sampling program has been conducted, and the design is to be based on removal of grit particles with an assumed SES of 106 μm. If this facility is located in an area that is known to have fine grit, what would be the expected percent grit removed? To achieve 90 percent grit removal, what adjustments would be necessary in the design of the facility.

5–13 Design an aerated grit chamber for a plant with an average flowrate of 15,000 m^3/d and a peak hourly flowrate of 40,000 m^3/d. Determine the amount of air required and the pressure at the discharge of the blowers. Allow a 250-mm loss in the diffusers, and add the submergence plus 400-mm for loss in piping and valves. Determine the power required. Use a blower efficiency of 70 percent. Determine the monthly power bill, assuming a motor efficiency of 90 percent and a power cost of \$0.12/kWh.

5–14 Design an aerated grit chamber installation for an average wastewater flowrate of 0.3 m^3/s and a peak flowrate of 1.0 m^3/s. Assume that the detention time at peak = 3.0 min, 4.0 min, and 5.0 min (detention time to be selected by instructor). Determine the dimensions of the grit chambers and the total air required.

5–15 Discuss the advantages and disadvantages of aerated grit chambers versus vortex-type grit chambers. Cite three references.

5–16 Determine the removal efficiency for a sedimentation basin with a critical velocity V_o of 2 m/h in treating a wastewater containing particles whose settling velocities are distributed as given in the table below (sample number to be selected by instructor). Plot the particle histogram for the influent and effluent wastewater.

	No. of particles		
	Sample		
V_{avg}, m/h	1	2	3
0.0–0.5	10	20	20
0.5–1.0	29	100	40
1.0–1.5	47	130	80
1.5–2.0	65	100	120
2.0–2.5	74	70	100
2.5–3.0	60	45	70
3.0–3.5	28	28	20
3.5–4.0	13	16	10
4.0–4.5	5	7	3

5–17 The rate of flow through an ideal clarifier is 8000 m^3/d, the detention time is 1 h, the depth is 3 m. If a full-length movable horizontal tray is set 1 m below the surface of the water, determine the percent removal of particles having a settling velocity of 1 m/h. Could the removal efficiency of the clarifier be improved by moving the tray? If so, where should the tray be located and what would be the maximum removal efficiency? What effect would moving the tray have if the particle settling velocity were equal to 0.3 m/h?

5–18 In Example 5–4, the assumed and calculated particle settling velocities did not reach closure. Continue the calculation iterations to find the actual settling velocity.

5–19 For a flocculent suspension, determine the removal efficiency for a basin 3 m deep with an overflow rate V_o equal to 3 m/h using the laboratory settling data presented in the following table.

	Percent suspended solids removed at indicated depth, m				
Time, min	0.5	1.0	1.5	2	2.5
20	61				
30	71	63	55		
40	81	72	63	61	57
50	90	81	73	67	63
60		90	80	74	68
70			86	80	75
80				86	81

5–20 A sedimentation basin has been retrofitted with 2.0-m (6.6-ft) square inclined plates spaced 75 mm (3.0 in.) apart. The angle of inclination of the plates can be set from 40, 50, or 60°. Assuming that the operation of the sedimentation basin can be either countercurrent or cocurrent, determine which arrangement is most efficient.

5–21 The following data were obtained from a settling test in a 3 m cylinder. The initial solids concentration was 3600 mg/L. Determine the thickener area required for sample (to be selected by instructor) for a concentration C_u of 11,000 or 12,000 mg/L (to be selected by instructor) with a flowrate of 1500 m^3/d.

	Height of interface, m		
	Sample		
Time, min	1	2	3
0	3.0	3.0	3.0
5	2.0	2.2	2.3
10	1.35	1.6	1.75
20	0.75	1.0	1.25
30	0.5	0.7	0.9
40	0.35	0.5	0.7
50	0.25	0.4	0.6
60	0.2	0.32	0.5

5–22 Design a circular radial-flow sedimentation tank for a town with a projected population of 45,000. Assume that the wastewater flowrate is 400 L/capita·d. Design for 2 h detention at the average flow. Determine the tank depth and diameter to produce an overflow rate of 36 $m^3/m^2 \cdot d$ for average flow. Assume standard tank dimensions to fit mechanisms made in diameters with increments of 1.0 m and in depth increments of 0.5 m.

5–23 A rectangular settling tank has an overflow rate of 30 $m^3/m^2 \cdot d$ and dimensions of 2.75 m deep by 6 m wide by 15 m long. Determine whether or not particles with a diameter of

0.1 mm and a specific gravity of 2.5 will be scoured from the bottom. Use $f = 0.03$ and $k = 0.04$.

5–24 Determine the percentage increase in the hydraulic and solids loading rates of the primary settling facilities of a treatment plant when 200 m^3/d of settled waste activated sludge containing 2000 mg/L of suspended solids is discharged to the existing primary facilities for thickening. The average plant flowrate is 20,000 m^3/d and the influent suspended solids concentration is about 350 mg/L. The design overflow rate for the primary settling tanks without the waste solids is 32 $m^3/m^2 \cdot d$ and the detention time is 2.8 h. Will the added incremental loading rate affect the performance of the primary settling facilities? Document the basis for your answer.

5–25 Prepare a table and compare the data from a minimum of three references with regard to the following primary sedimentation tank design parameters: (1) detention time (with and without preaeration); (2) expected BOD removal; (3) expected suspended-solids removal; (4) mean horizontal velocity; (5) surface loading rate in $m^3/m^2 \cdot d$; (6) effluent weir overflow rate per unit length; (7) size of organic particle removed; (8) length-to-width ratio (rectangular tanks); (9) average depth. List all references.

5–26 A medium-size treatment plant is being designed, and circular and rectangular primary sedimentation tanks are being considered. What factors should be considered in the evaluation and selection of the type of tank? List the advantages and disadvantages for each type. Cite at least three recent references (since 2000).

5–27 Contrast dissolved-air flotation with sedimentation discussing the following parameters: (a) detention time, (b) surface-loading rate, (c) power input, (d) efficiency, and (e) most favorable application for each type.

5–28 Using one of the following three data sets (to be selected by instructor), design a dissolved air flotation thickening system to thicken the solids to a 4 percent solids concentration.

		Data set		
		1	2	3
Item	Unit	Mixed liquor	Settled activated sludge	Primary + activated sludge
Solids conc.	mg/L	2500	7500	10,000
Optimum A/S ratio	ratio	0.02	0.03	0.03
Temperature	°C	20	20	20
Surface loading rate	$L/m^2 \cdot min$	10	15	8
Flowrate	m^3/d	1200	400	800

5–29 The following data were obtained from a test program designed to evaluate a new diffused-air aeration system. Using these data, determine the value of K_La at 20°C and the equilibrium dissolved-oxygen concentration in the test tank. The test program was conducted using tap water at a temperature of 24°C.

C, mg/L	dC/dt, mg/L·h
1.5	8.4
2.7	7.5
3.9	5.3
4.8	4.9
6.0	4.2
7.0	2.8
8.2	2.0

5–30 If the volume of the test tank used to evaluate the aeration system in Prob. 5–29 were equal to 100 m^3 and the air flowrate were equal to 2 m^3/min, determine the maximum oxygen-transfer efficiency at 20°C and 1.0 atmosphere.

5–31 Using the equation developed in Example 5–15, estimate the air flowrate in m^3/min required to increase the oxygen content of chlorinated effluent from zero to 4 mg/L. The effluent flowrate is equal to 20,000 m^3/d. Assume that the transfer efficiency is 6 percent and the temperature is 15°C. What is the air requirement when the temperature is 25°C?

5–32 Alternative diffused-air aeration devices are being considered for installation with a submergence of 4.5 m in an aeration tank at an activated-sludge treatment plant. The oxygen required for biological treatment is 7000 kg/d. Determine the standard oxygen transfer rate and theoretical air requirements for ceramic dome diffusers installed in a grid pattern as compared to nonporous diffusers installed for a dual-spiral roll. The wastewater temperature is 20°C and the α factors are 0.64 for ceramic domes and 0.75 for the nonporous diffusers, respectively.

5–33 Using the data in Prob. 5–29 and a wintertime wastewater temperature of 10°C, determine the theoretical air requirements for cold-weather operation. How will equipment selection be affected in meeting both summer and winter operation?

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6

Chemical Unit Processes

WORKING TERMINOLOGY

Term	Definition
Absorption	The process by which atoms, ions, molecules, and other constituents are transferred from one phase and are distributed uniformly in another phase (see also adsorption).
Activated carbon	A substance used commonly in adsorption processes for the removal of various constituents from water and air. Activated carbon is derived from an organic base material, prepared (activated) using a high temperature and pressure pyrolysis process, resulting in properties conducive to mass transfer.
Adsorption	The process by which atoms, ions, molecules, and other constituents are transferred from one phase and accumulate on the surface of another phase (see also absorption). Substances accumulate on the solid phase due to a number of physical attraction and chemical bonding forces.
Adsorbate	The compound in a liquid or gas phase suspension that is deposited onto the surface of an adsorbent material.
Adsorbent	The solid material onto which adsorption takes place.
Advanced oxidation	A chemical oxidation process that relies upon the hydroxyl radical (HO•) for the destruction of trace organic constituents found in water. A number of processes have been identified that are able to produce HO•, including combinations of ozone, hydrogen peroxide and ultraviolet light.

Term	Definition
Desorption	The release of a volatile gas phase from a liquid phase as in gas stripping or the release of a previously adsorbed compound from an adsorbent material.
Electrical potential	The driving force for the exchange of electrons between constituents during a redox reaction, reported in volts with respect to the standard hydrogen electrode.
Extinction coefficient	The fractional amount of UV radiation attenuated as UV light passes through water that contains dissolved substances that absorb the energy. The extinction coefficient is also known as the molar absorptivity.
Gas stripping	A process to remove a volatile constituent from a liquid phase, such as the removal of ammonia from water in a packed column using air as the gas phase.
Ion exchange	A process used for the removal of dissolved ionic constituents where ions of a given species are displaced from a solid phase material by ions of a different species from solution.
Isotherm	A mathematical function used to relate the amount of a given constituent adsorbed from water per concentration of adsorbent at a given temperature.
Mineralization	The complete oxidation and conversion of organic substances into inorganic forms, such as carbon dioxide, water, and mineral acids, through the action of chemically and biologically mediated redox reactions.
Natural organic matter (NOM)	A bulk parameter used to describe the complex matrix of organic constituents of biological origin found in water, quantified typically as total organic carbon (TOC). See also Chap. 2.
Oxidation reactions	The component of a redox reaction involving the loss of electrons. The oxidation reaction can be obtained from tabulated half reactions (by convention, half reactions are reported as reduction reactions) by reversing the direction of the half reaction and multiplying the electrical potential by -1.
Photolysis	A process used for the treatment of trace organic constituents, where a UV light source is used to supply photons that are absorbed by the constituent and subsequently become unstable and react or split apart.
Quantum yield	A quantity used to describe the frequency at which photon absorption results in a photolysis reaction and is specific to the compound and the wavelength.
Quenching	The use of physical or chemical means to stop a chemical reaction.
Redox reaction	The overall reaction resulting from the combination of a reduction and oxidation reaction.
Reduction reactions	The component of a redox reaction involving the gain of electrons. The reduction reaction can be obtained directly from tabulated half reactions.
Reactivation	The desorption of constituents from an adsorbent material followed by the combustion of the remaining sorbed constituents, resulting in restoration of the adsorptive capacity.
Regeneration	The desorption of constituents from an adsorbent material for the partial restoration of the adsorptive capacity.
Reverse osmosis (RO)	The rejection of dissolved constituents by preferential diffusion using a pressure-driven, semipermeable membrane (see Chap. 11).
Scavengers	In advanced oxidation systems, substances that preferentially react with the oxidant and radical species, typically reducing the degradation rate for the compound of interest and overall efficiency of the process.
Separation processes	Physical and chemical mass transfer processes used in water reclamation that bring about treatment by the isolation of particular constituents. The isolated constituents are concentrated into a waste stream that must be managed (see Chap. 11).

Term	Definition
Synthetic organic compounds (SOCs)	Compounds of synthetic origin used extensively in industrial processes and contained in numerous manufactured consumer products. The presence of SOCs in drinking water as well as reclaimed water is of concern due to toxicity and unknown effects.
Trace constituent	A diverse classification of constituents found at low concentrations in untreated wastewater and not readily removed using conventional secondary treatment, and thus also found in treated effluent and environmental water systems. Trace constituents are of concern due to known or suspected toxicity associated with many of these compounds, which may need to be removed during water reclamation depending on the reuse requirements.
Water matrix	The term used to refer to a given water and all of its constituents that result in the physical, chemical, and biological properties of the solution.

Processes used for the treatment of wastewater in which change is brought about by means of or through chemical reactions are known as *chemical unit processes*. In the field of wastewater treatment, chemical unit processes are used commonly in conjunction with the physical unit processes discussed in Chap. 5, and the biological unit processes discussed in Chaps. 7 through 10, to meet specific treatment objectives. The purpose in this chapter is to present and discuss (1) the role of chemical unit processes in wastewater treatment; (2) fundamentals of chemical coagulation; (3) the precipitation reactions that occur when various chemicals are added to improve the performance of wastewater treatment facilities; (4) the chemical reactions involved in the precipitation of phosphorus from wastewater; (5) precipitation of ammonium and phosphorus; (6) precipitation of heavy metals and dissolved inorganic substances; (7) chemical oxidation; (8) advanced oxidation; (9) photolysis; (10) chemical neutralization, scale control, and stabilization; and (11) chemical storage, feeding, piping, and control systems. Ion exchange and adsorption, sometimes classified as chemical unit processes, but which, more correctly, should be classified as separation processes as they are based on mass transfer, are considered in Chap. 11. Chemical disinfection, an important and much-used chemical unit process, is considered separately in Chap. 12.

6–1 ROLE OF CHEMICAL UNIT PROCESSES IN WASTEWATER TREATMENT

The principal chemical unit processes used for wastewater treatment include (1) chemical coagulation, (2) chemical precipitation, (3) chemical oxidation, (4) advanced oxidation, (5) photolysis, and (6) chemical neutralization, scale control, and stabilization. The applications and the limitations involved in the use of these processes are considered in the following discussion.

Applications of Chemical Unit Processes

Applications of chemical unit processes for the management and treatment of wastewater are reported in Table 6–1. Chemical processes, in conjunction with various physical operations, have been used in nearly all aspects of wastewater treatment, including physical-chemical treatment systems for the complete treatment of untreated (raw) wastewater. In some cases, chemical processes have been developed for the advanced treatment of secondary effluent (see Fig. 6–1). The most important applications of chemical unit processes in

Table 6–1

Applications of chemical unit processes in wastewater treatment

Process	Application	See section or chapter
Advanced oxidation processes	Removal of refractory organic compounds	6–8
Chemical coagulation	The chemical destabilization of particles in wastewater to bring about their aggregation during perikinetic and orthokinetic flocculation	6–2
Chemical disinfection	Disinfection with chlorine, chlorine compounds, bromine, and ozone	Chap. 12
	Control of biofilm growths in sewers	Chap. 12
	Control of odors	Chap. 16
Chemical neutralization	Control of pH	6–10
Chemical oxidation	Removal of BOD, grease, etc.	6–7
	Removal of ammonium	6–7
	Destruction of microorganisms	Chap. 12
	Control of odors in sewers, pump stations, and treatment plants	Chap. 16
	Removal of resistant organic compounds	6–7, 6–8, 6–9
Chemical precipitation	Enhanced removal of total suspended solids and BOD in primary sedimentation facilities (chemically enhanced primary treatment, CEPT).	6–3
	Removal of phosphorus	6–4, 6–5
	Removal of ammonium	6–5
	Removal of heavy metals	6–6
	Physical-chemical treatment	6–3
	Corrosion control in sewers due to H_2S	6–7
Chemical scale control	Control of scaling due to calcium carbonate and related compounds	6–10
Chemical stabilization	Stabilization of treated effluents	6–10
Ion exchange	Removal of ammonium (NH_4^+), heavy metals, total dissolved solids	Chap. 11
	Removal of organic compounds	Chap. 11

modern wastewater treatment have included: (1) the disinfection of wastewater; (2) the removal of phosphorus; (3) the coagulation of particulate matter; (4) pH control, neutralization, and alkalinity supplementation; and (5) water stabilization and stability control, which is increasingly important in wastewater reuse applications, e.g., cooling tower makeup. However, the need to remove trace constituents prior to potable reuse has stimulated the rapid development and implementation of advanced oxidation and photolysis processes. Chemical unit processes are also being implemented to separate wastewater constituents from untreated wastewater for the recovery of energy and agricultural nutrients.

Considerations in the Use of Chemical Unit Processes

In considering the application of chemical unit processes, it is important to remember that one of the inherent disadvantages associated with some chemical unit processes, as compared with the physical unit operations, is that they are additive processes (i.e., something

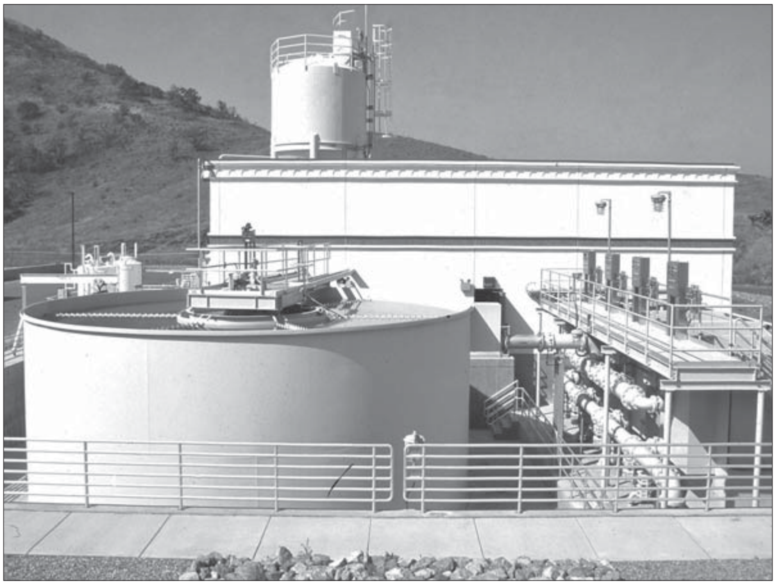


Figure 6–1

Typical lime clarification facilities following secondary treatment used as pretreatment step for advanced treatment of wastewater using reverse osmosis. Lime storage is in the silo shown behind the building that is used to house the lime slaking facilities and the reverse osmosis units used for advanced treatment. Granular medium depth filters are shown to the right of the lime clarifier in the foreground.

is added to the wastewater to achieve the removal of something else). As a result, there is often a net increase in certain dissolved wastewater constituents from a given process. For example, where coagulants are added to enhance the removal efficiency of particulate matter in sedimentation, the total dissolved solids (TDS) concentration of the wastewater is increased generally. Depending on the specific chemicals used, the concentration of aluminum or iron in the process effluent or solids may also be increased. Similarly, when chlorine is added to wastewater, the TDS of the effluent is increased and disinfection byproducts may appear. If the treated wastewater is to be reused or strict discharge limits exist for specific constituents, the increase in dissolved constituents can be a significant factor. This additive aspect is in contrast to the physical unit operations and the biological unit processes, which may be described as being subtractive, in that wastewater constituents are removed from the wastewater without a corresponding increase in the dissolved constituents. A potential disadvantage of chemical precipitation processes is the handling, treatment, and disposal of the sludge that is produced. Another disadvantage of chemical unit processes is that the cost of most chemicals is related to the cost of energy. As a result, the use of chemicals affects the overall wastewater treatment process carbon footprint because of energy inputs required to produce and transport chemicals. The impacts of chemical usage on the carbon footprint and overall process sustainability are considered in Chap. 16. In some cases, such as meeting low discharge limits for nutrients or specific constituents, the use of chemicals will be unavoidable.

6–2 FUNDAMENTALS OF CHEMICAL COAGULATION

Colloidal particles found in wastewater typically have a net negative surface charge. The size of colloids (about 0.001 to 1 μm) is such that the attractive body forces between particles are considerably less than the repelling forces of the electrical charge. Under these stable conditions, Brownian motion keeps the particles in suspension. Brownian motion (i.e., random movement) results from the constant bombardment of the colloidal particles by the relatively small water molecules, driven by random thermal motion that surrounds them. Coagulation is the process of destabilizing colloidal particles so that particle growth can occur as a result of particle collisions. The complete theory of chemical coagulation reactions is beyond the scope of this textbook. The simplified reactions used in this and other textbooks to describe coagulation and chemical precipitation processes can only be considered approximations, as the reactions may not necessarily proceed as indicated.

Coagulation reactions are often incomplete, and numerous side reactions with other substances in wastewater may take place depending on the characteristics of the water matrix, which will vary diurnally as well as seasonally. To introduce the subject of chemical coagulation, the following topics are discussed in this section: (1) basic definitions for coagulation and flocculation, (2) the nature of particles in wastewater, (3) the development and measurement of surface charge, (4) consideration of particle-particle interactions, (5) particle destabilization with potential determining ions and electrolytes, (6) particle destabilization and aggregation with polyelectrolytes, and (7) particle destabilization and removal with hydrolyzed metal ions. The following discussion is meant to serve as an introduction to the nature of the phenomena and processes involved in the coagulation process.

Basic Definitions

The term *chemical coagulation* as used in this text includes all of the reactions and mechanisms involved in the chemical destabilization of particles and in the formation of larger particles through perikinetic flocculation (aggregation of particles in the size range from 0.001 to 1 μm). Coagulant and flocculant are terms that will also be encountered in the literature on coagulation. In general, a coagulant is the chemical that is added to destabilize the colloidal particles in wastewater so that floc formation can result. A flocculant is a chemical, typically organic, added to enhance the flocculation process. Typical coagulants include natural and synthetic organic polymers, metal salts such as alum or ferric chloride, and prehydrolized metal salts such as polyaluminum chloride (PACl), polyiron chloride (PICl), and polydiallyldimethylammonium chloride (polyDADMAC). Flocculants, especially organic polymers, are also used to enhance the performance of granular medium filters and in the dewatering of digested biosolids. In these applications, the flocculant chemicals are often identified as filter aids. Note that inorganic substances, such as bentonite, are used to improve the performance of filtration system in some industrial applications.

The term *flocculation* is used to describe the process whereby the size of particles increases as a result of particle collisions. As noted in Chap. 5, there are two types of flocculation: (1) *microflocculation* (also known as *perikinetic flocculation*) in which particle aggregation is brought about by the random thermal motion of fluid molecules known as Brownian motion or movement, and (2) *macroflocculation* (also known as *orthokinetic flocculation*) in which particle aggregation is brought about by inducing velocity gradients and mixing in the fluid containing the particles to be flocculated. Another form of macroflocculation is brought about by differential settling in which large particles overtake small particles to form larger particles. The purpose of flocculation is to produce particles, by means of aggregation, that can be removed by inexpensive particle-separation procedures such as gravity sedimentation and filtration. Again as noted in Chap. 5, macroflocculation is ineffectual until the colloidal particles reach a size of 1 to 10 μm through contacts produced by Brownian motion and gentle mixing.

Nature of Particles in Wastewater

The particles in wastewater may, for practical purposes, be classified as suspended and colloidal. In practice, the distinction between colloidal and suspended particles is blurred because the particles removed by gravity settling will depend on the design of the sedimentation facilities. Because colloidal particles cannot be removed by sedimentation in a reasonable period of time, chemical methods (i.e., the use of chemical coagulants and flocculant aids) must be used to help bring about the removal of these particles.

To understand the role that chemical coagulants and flocculant aids play in bringing about the removal of colloidal particles, it is important to understand the characteristics of the colloidal particles found in wastewater. Important factors that contribute to the characteristics of colloidal particles in wastewater include: (1) particle size and number, (2) particle shape and flexibility, (3) surface properties including electrical characteristics, (4) particle-particle interactions, and (5) particle-solvent interactions (Shaw, 1966). Particle size, particle shape and flexibility, and particle-solvent interactions are considered below. Because of their importance, the development and measurement of surface charge and particle-particle interactions are considered separately.

Particle Size and Number. The size of colloidal particles in wastewater considered in this text is typically in the range from 0.01 to 1 μm. As noted in Chap. 2, some researchers have classified the size range for colloidal particles as varying from 0.001 to 1 μm. The number of colloidal particles in untreated wastewater and after primary sedimentation is typically in the range from 10^6 to 10^{12}/mL. It is important to note that the number of colloidal particles will vary depending on the location where the sample is taken within a treatment plant. The number of particles, as will be discussed later, is of importance with respect to the method to be used for their removal.

Particle Shape and Flexibility. Particle shapes found in wastewater can be described as: spherical, semi-spherical, ellipsoids of various shapes (e.g., prolate and oblate), rods of various length and diameter (e.g., *E. coli*), disk and disk-like, strings of various lengths, and random coils. Large organic molecules are often found in the form of coils which may be compressed, uncoiled, or almost linear. The shape of some larger floc particles is often described as fractal. The particle shape will vary depending on the location within the treatment process that is being evaluated. The shape of the particles will affect the electrical properties, the particle-particle interactions, and particle-solvent interactions. Because of the many shapes of particles encountered in wastewater, the theoretical treatment of particle-particle interactions is an approximation, at best.

Particle-Solvent Interactions. There are three general types of colloidal particles in liquids: hydrophobic or "water-hating," hydrophilic or "water-loving," and association colloids. The first two types are based on the attraction of the particle surface for water. Hydrophobic particles have relatively little attraction for water while hydrophilic particles have a great attraction for water. It should be noted, however, that water can interact to some extent even with hydrophobic particles. Some water molecules will generally adsorb on the typical hydrophobic surface, but the reaction between water and hydrophilic colloids occurs to a much greater extent. The third type of colloid is known as an association colloid, typically made up of surface-active agents such as soaps, synthetic detergents, and dyestuffs that form organized aggregates known as micelles.

Development and Measurement of Surface Charge

An important factor in the stability of colloids is the presence of a surface charge. Surface charge develops in a number of different ways, depending on the chemical composition of the water matrix (wastewater in this case) and the nature of the colloid. Surface charge develops most commonly through (1) isomorphous replacement, (2) structural imperfections,

(3) preferential adsorption, and (4) ionization, as defined below. Regardless of how it develops, the surface charge, which promotes stability, must be overcome if these particles are to be aggregated (flocculated) into larger particles with enough mass to settle easily.

Isomorphous Replacement. Charge development through isomorphous replacement occurs in clay and other soil particles in which ions in the lattice structure are replaced with ions from solution (e.g., the replacement of Si^{4+} with Al^{3+}).

Structural Imperfections. In clay and similar particles, charge development can occur because of broken bonds on the crystal edge and imperfections in the formation of the crystal.

Preferential Adsorption. When oil droplets, gas bubbles, or other chemically inert substances are dispersed in water, they will acquire a negative charge through the preferential adsorption of anions (particularly hydroxyl ions).

Ionization. In the case of substances such as proteins or microorganisms, surface charge is acquired through the ionization of carboxyl and amino groups (Shaw, 1966). This ionization can be represented as follows, where *R* represents the bulk of the solid (Fair et al., 1968):

$$\underset{\text{at high pH}}{R^{\mathrm{COO^-}}_{\mathrm{NH_2}}} \qquad \underset{\text{at isoelectric point}}{R^{\mathrm{COO^-}}_{\mathrm{NH_3^+}}} \qquad \underset{\text{at low pH}}{R^{\mathrm{COOH}}_{\mathrm{NH_3^+}}} \tag{6–1}$$

The Electrical Double Layer. When the colloid or particle surface becomes charged, some ions of the opposite charge (known as counter-ions) become attached to the surface (see Fig. 6–2). Ions are held there through electrostatic and van der Waals forces of attraction strongly enough to overcome thermal agitation. Surrounding this fixed layer of ions is a diffuse layer of ions, which is prevented from forming a compact double layer by thermal agitation, as illustrated schematically on Fig. 6–2. The electrical double layer consists of a compact layer (Stern) in which the potential drops from ψ_o to ψ_s, and a diffuse layer in which the potential drops from ψ_s to 0 in the bulk solution.

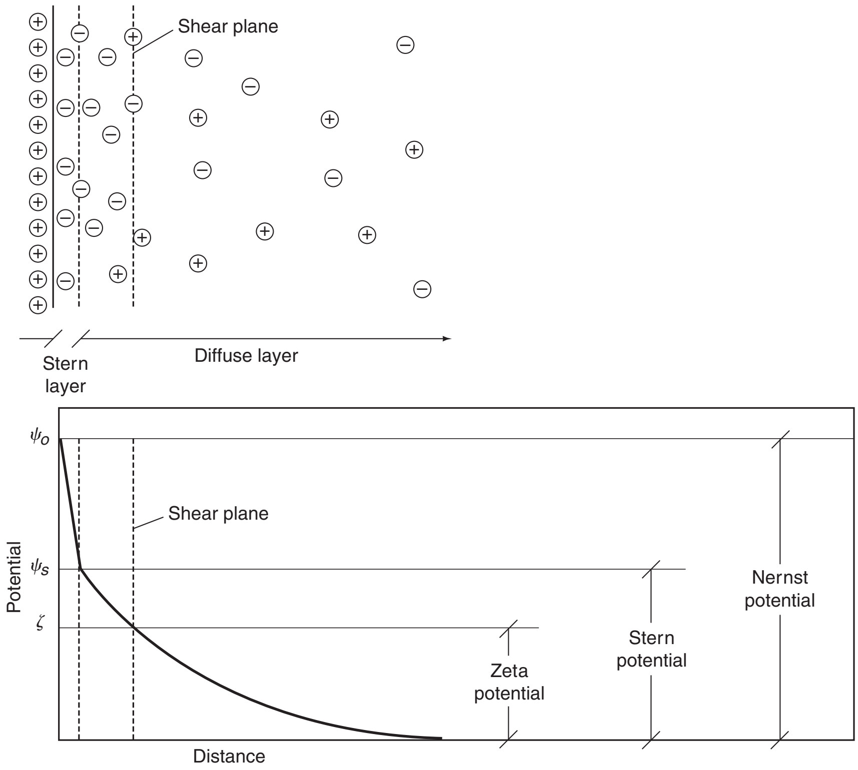
Measurement of Surface Potential. If a particle is placed in an electrolyte solution, and an electric current is passed through the solution, the particle, depending on its surface charge, will be attracted to one or the other of the electrodes, dragging with it a cloud of ions. The potential at the surface of the cloud (called the surface of shear) is sometimes measured in wastewater treatment operations. The measured value is often called the zeta potential. Theoretically, however, the zeta potential should correspond to the potential measured at the surface enclosing the fixed layer of ions attached to the particle, as shown on Fig. 6–2. The use of the measured zeta potential value is limited because it will vary with the nature of the water matrix.

Particle-Particle Interactions

Particle-particle interactions are important in bringing about aggregation by means of Brownian motion. The theory that has been developed to describe particle-particle interactions is based on the consideration of interaction between two charged flat plates and between two charged spheres (Deryagin and Landau, 1941; Verwey and Overbeek, 1948). As neither of these developments is directly applicable to the particles found in wastewater,

Figure 6–2

Stern model of electrical double layer (Shaw, 1966).



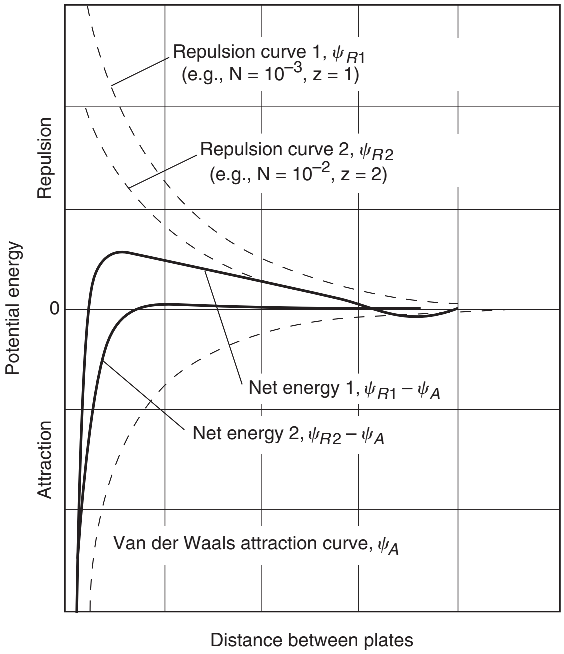
as described previously, the analysis for two charged flat plates will be used for illustrative purposes. The interaction between two plates is illustrated on Fig. 6–3. As shown on Fig. 6–3, the two principal forces involved are the forces of repulsion, due to the electrical properties of the charged plates, and the van der Waals forces of attraction. It should be noted that the van der Waals forces of attraction do not come into play until the two plates are brought together in close proximity to each other.

The net total energy shown on Fig. 6–3 by the solid lines is the difference between the forces of repulsion and attraction. The two conditions, with respect to the forces of repulsion, are illustrated on Fig. 6–3. As shown for condition 1, the forces of attraction will predominate at short and long distances. The net energy curve for condition 1 contains a repulsive maximum that must be overcome if the particles, represented as the two plates, are to be held together by the van der Waals force of attraction. In condition 2, there is no energy barrier to overcome. Clearly, if colloidal particles are to be removed by microflocculation, the repulsive force must be reduced. Although floc particles can form at long distances as shown by the net energy curve for condition 2, the net force holding these particles together is weak and the floc particles that are formed can be ruptured easily.

Particle Destabilization with Potential-Determining Ions and Electrolytes. To bring about particle aggregation through microflocculation, steps must be taken to reduce particle charge or to overcome the effect of this charge. The effect of the charge can be overcome by (1) the addition of potential-determining ions, which will be taken up by or will react with the colloid surface to lessen the surface charge and (2) the addition of electrolytes, which have the effect of reducing the thickness of the diffuse electric layer and, thereby, reduce the zeta potential.

Figure 6–3

Definition sketch for particle-particle interactions based on the repulsion due to particle surface charge and van der Waals forces of attraction. N = concentration, z = charge.



Use of Potential-Determining Ions. The addition of potential-determining ions to promote coagulation can be illustrated by the addition of strong acids or bases to reduce the charge of metal oxides or hydroxides to near zero so that coagulation can occur. The effect of adding potential determining ions in a solution containing charged particles is illustrated on Fig. 6–4. The magnitude of the effect will depend on the concentration of

Figure 6–4

Definition sketch for the effects of the addition of counter-ions and electrolytes to solutions containing charged colloidal particles.

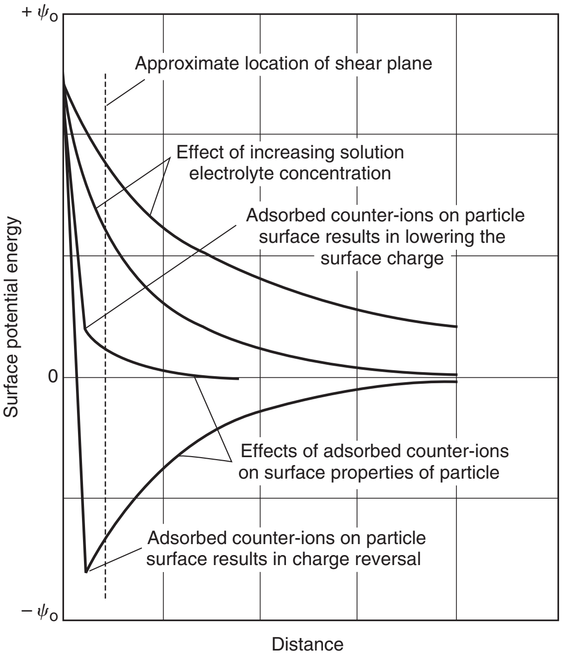
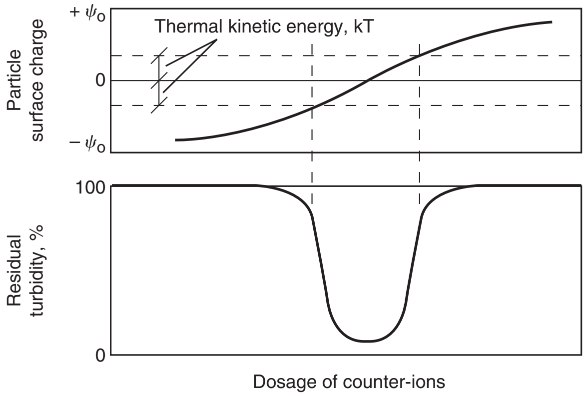


Figure 6–5

Definition sketch for the reversal of particle surface charge due to the addition of counter-ions.



potential determining ions added. The following ratios, known as the Shultz-Hardy rule, can be used to assess the effectiveness of potential determining or counter-ions:

$$1:\frac{1}{2^6}:\frac{1}{3^6} \quad \text{or} \quad 100:1.6:0.13 \tag{6–2}$$

It is interesting to note that, depending on the concentration and nature of the counter-ions added, it is possible to reverse the charge of the double layer and develop a new stable particle.

The effect of adding counter-ions to a solution containing charged particles is illustrated on Fig. 6–5. The upper curve on Fig. 6–5 represents the surface charge of the particle as a function of the concentration of counter-ions added. The lines designated kT represent the thermal kinetic energy of the particle. The lower diagram is a plot of the turbidity that would result if the particles that have been destabilized and have undergone microflocculation were removed by settling. As shown, when the surface charge (either positive or negative) is greater than the thermal kinetic energy of the particles, the particles will not flocculate and the original turbidity is observed.

Additional details on the use of counter-ions may be found in Shaw (1966). The use of potential determining ions is not feasible in either water or wastewater treatment because of the massive concentration of ions that must be added to bring about sufficient compression of the electrical double layer to effect perikinetic flocculation.

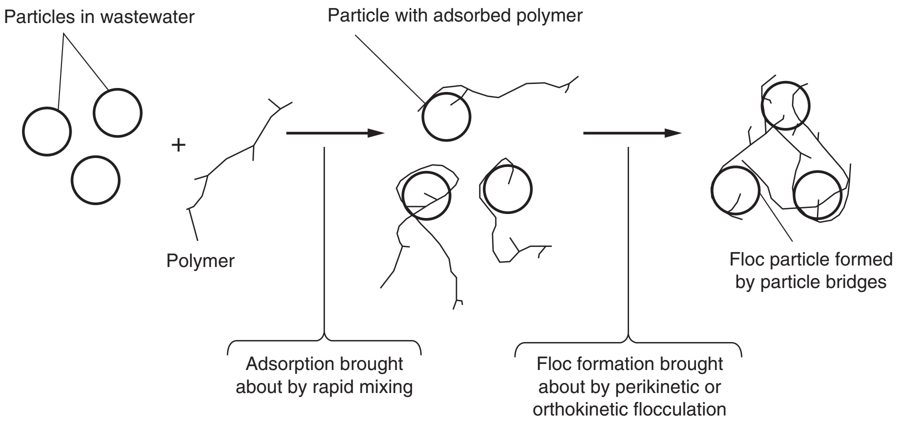
Use of Electrolytes. Electrolytes can also be added to coagulate colloidal suspensions. Increased concentration of a given electrolyte will cause a decrease in zeta potential and a corresponding decrease in repulsive forces as illustrated in condition 2 on Fig. 6–3 and on Fig. 6–4. The concentration of an electrolyte that is needed to destabilize a colloidal suspension is known as the *critical coagulation concentration* (CCC). Increasing the concentration of an indifferent electrolyte will not result in the restabilization of the colloidal particles. As with the addition of potential determining ions, the use of electrolytes is also not feasible in wastewater treatment. As discussed subsequently, a change in the particle charge will occur when chemicals are added to adjust the pH of the wastewater to optimize the performance of hydrolyzed metal ions used as coagulants.

Particle Destabilization and Aggregation with Polyelectrolytes

Polyelectrolytes may be divided into two categories: natural and synthetic. Important natural polyelectrolytes include polymers of biological origin and those derived from starch products such as cellulose derivatives, and alginates. Synthetic polyelectrolytes

Figure 6–6

Definition sketch for inter-particle bridging with organic polymers.



consist of simple monomers that are polymerized into high-molecular-weight substances. Depending on whether their charge, when placed in water, is negative, positive, or neutral, these polyelectrolytes are classified as anionic, cationic, and nonionic, respectively. The action of polyelectrolytes may be divided into the following three general categories.

Charge Neutralization. In the first category, polyelectrolytes act as coagulants that neutralize or lower the charge of the wastewater particles. Because wastewater particles normally are charged negatively, cationic polyeletrolytes are used for this purpose. In this application, the cationic polyelectrolytes are considered to be primary coagulants. To affect charge neutralization, the polyelectrolyte must be adsorbed to the particle. Because of the large number of particles found in wastewater, the mixing intensity must be sufficient to bring about the adsorption of the polymer onto the colloidal particles. With inadequate mixing, the polymer will eventually fold back on itself and its effectiveness in reducing the surface charge will be diminished. Further, if the number of colloidal particles is limited, it will be difficult to remove them with low polyelectrolyte dosages.

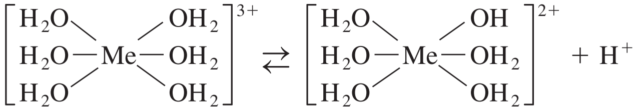
Polymer Bridge Formation. The second mode of action of polyelectrolytes is interparticle bridging (see Fig. 6–6). In this case, polymers that are anionic and nonionic (usually anionic to a slight extent when placed in water) become attached at a number of adsorption sites to the surface of the particles found in the wastewater. A bridge is formed when two or more particles become adsorbed along the length of the polymer. Bridged particles become intertwined with other bridged particles during the flocculation process. The size of the resulting three-dimensional particles grows until they can be removed easily by sedimentation. Where particle removal is to be achieved by the formation of particle-polymer bridges, the initial mixing of the polymer and the wastewater containing the particles to be removed must be accomplished in a matter of seconds. Instantaneous initial mixing is usually not required, as the polymers are already formed, which is not the case with the polymers formed by metal salts (see discussion of hydrolyzed metal ions given below). As noted above, the mixing intensity must be sufficient to bring about the adsorption of the polymer onto the colloidal particles. If inadequate mixing is provided, the polymer will eventually fold back on itself, in which case it is not possible to form polymer bridges.

Charge Neutralization and Polymer Bridge Formation. The third type of polyelectrolyte action may be classified as a charge neutralization and bridging phenomenon, which results from using cationic polyelectrolytes of high molecular weight. Besides lowering the surface charge on the particle, these polyelectrolytes also form particle bridges as described above.

Particle Destabilization and Removal with Hydrolyzed Metal Ions

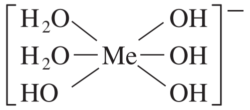
In contrast with the aggregation brought about by the addition of chemicals that act as counter-ions, electrolytes, and polymers, aggregation brought about by the addition of alum or ferric compounds is a more complex process. To understand particle destabilization and the removals achieved with hydrolyzed metal ions, it will be instructive to consider first the formation of metal ion hydrolysis products. Operating ranges for action of metal salts and the importance of initial mixing are also considered in light of the formation of these particles.

Formation of Hydrolysis Products. In the past, it was thought that free Al^{3+} and Fe^{3+} were responsible for the effects observed during particle aggregation; it is now known, however, that their hydrolysis products are responsible (Stumm and Morgan, 1962; Stumm and O'Melia, 1968). Although the effect of these hydrolysis products is only now appreciated, it is interesting to note that their chemistry was first elucidated in the early 1900s by Pfeiffer (1902–1907), Bjerrum (1906–1920), and Werner (1907) (Thomas, 1934). In the early 1900s, Pfeiffer proposed that the hydrolysis of trivalent metal salts, such as chromium, aluminum, and iron, could be represented as



(6–3)

with the extent of the dissociation depending on the anion associated with the metal and on the physical and chemical characteristics of the solution. Further, it was proposed that, upon the addition of sufficient base, the dissociation can proceed to produce a negative ion (Thomas, 1934), such as



(6–4)

It should be noted that the complex compounds given in Eqs. (6–3) and (6–4) are known as a *coordination compounds*, which are defined as a central metal ion (or atom) attached to a group of surrounding molecules or ions by coordinate covalent bonds. The surrounding molecules or ions are known as *ligands*, and the atoms attached directly to the metal ion are called ligand donor atoms (McMurry and Fay, 2011). Ligand compounds of interest in wastewater treatment include: carbonate (CO_3^{2-}), chloride (Cl^-), hydroxide (OH^-), ammonium (NH_4^+), and water (H_2O).

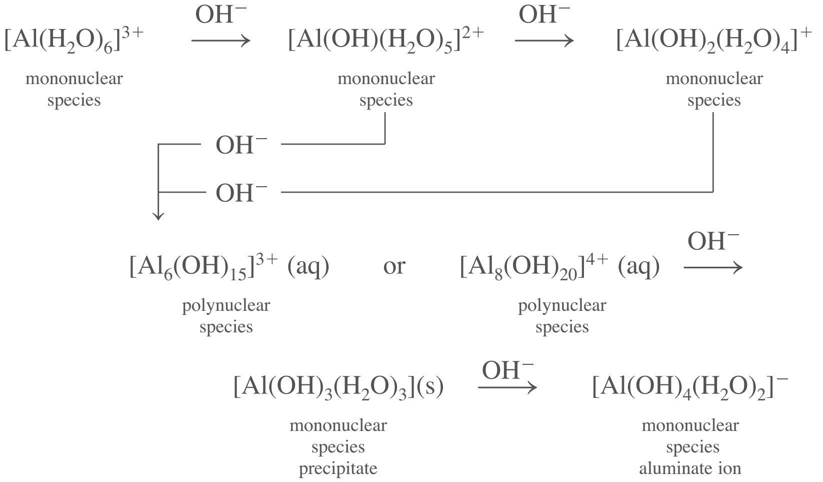
In addition, a number of the coordination compounds are also *amphoteric* in that they can exist both in strong acids and in strong bases (McMurry and Fay, 2011). For example, aluminum hydroxide behaves as follows in acidic and basic solutions:

$$\text{In acid:}\ \ Al(OH)_3(s) + 6H_3O^+(aq) \rightleftarrows Al^{3+}(aq) + 6H_2O \qquad (6\text{–}5)$$

$$\text{In base:}\ \ Al(OH)_3(s) + OH^-(aq) \rightleftarrows Al(OH)_4^-(aq) \qquad (6\text{–}6)$$

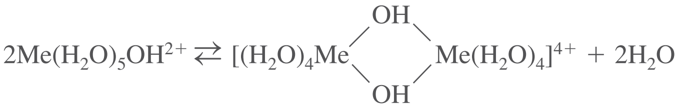
As shown in Eq. (6–5), $Al(OH)_3$ will dissolve in the presence of excess acid to form aqueous Al^{3+}. In the presence of excess hydroxide [see Eq. (6–6)], $Al(OH)_3$ will dissolve to form the aluminate ion, $Al(OH)_4^-$. The acid and base properties of the hydroxides and the nature of the covalent bonds will depend on the position of the element on the periodic table. Further, it should be noted some basic hydroxides will dissolve in strong acid, but not in a strong base (McMurry and Fay, 2011).

Over the past 50 years, it has been observed that the intermediate hydrolysis reactions of Al(III) are much more complex than would be predicted on the basis of a model in which a base is added to the solution. At the present time the complete chemistry for the formation of hydrolysis reactions and products is not well understood (Letterman et al., 1999). A hypothetical model [see Eq. (6–7)], proposed by Stumm (Fair et al., 1968) for Al(III), is useful for the purpose of illustrating the complex reactions involved. A number of alternative formation sequences have also been proposed (Letterman, 1991).



(6–7)

Before the reaction proceeds to the point where a negative aluminate ion is produced, polymerization as depicted in the following formula will usually take place (Thomas, 1934).



$$2Me(H_2O)_5OH^{2+} \rightleftarrows [(H_2O)_4Me(OH)_2Me(H_2O)_4]^{4+} + 2H_2O \tag{6–8}$$

As illustrated by Eqs. (6–7) and (6–8), the possible combinations of the various hydrolysis products are endless, and their enumeration is not the purpose here. What is important, however, is the realization that one or more of the hydrolysis products and or polymers may be responsible for the observed action of aluminum or iron.

Further, because the hydrolysis reactions follow a stepwise process, the effectiveness of aluminum and iron will vary with time. For example, an alum slurry that has been prepared and stored will behave differently from a freshly prepared solution when it is added to a wastewater. For a more detailed review of the chemistry involved, the excellent articles on this subject by Stumm and Morgan (1962) and Stumm and O'Melia (1968) are recommended. Additional details on the chemistry of aluminum and iron may be found in Benefield et al. (1982), Morel and Hering (1993), Pankow (2012), Snoeyink and Jenkins (1980), Sawyer et al. (2002), and Stumm and Morgan (1981).

Action of Hydrolyzed Metals Ions. The action of hydrolyzed metal ions in bringing about the destabilization and removal of colloidal particles may be divided into the following three categories:

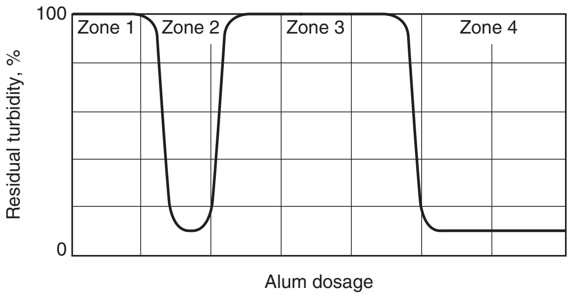
1. Adsorption and charge neutralization
2. Adsorption and interparticle bridging
3. Enmeshment in sweep floc

Adsorption and charge neutralization involves the adsorption of mononuclear and polynuclear metal hydrolysis species [see Eq. (6–7)] on the colloidal particles found in wastewater. It should be noted that it is also possible to get charge reversal with metal salts, as describe previously with the addition of counter-ions (see Fig. 6–5). Adsorption and interparticle bridging involves the adsorption of polynuclear metal hydrolysis species and polymer species [see Eqs. (6–7) and (6–8)] which, in turn, will ultimately form particle-polymer bridges, as described previously. As the coagulant requirement for adsorption and charge neutralization is satisfied, metal hydroxide precipitates and soluble metal hydrolysis products will form as defined by Eq. (6–5). If a sufficient concentration of metal salt is added large amounts of metal hydroxide floc will form. Following macroflocculation, large floc particles will be formed that will settle readily. In turn, as these floc particles settle, they sweep through the water containing colloidal particles. The colloidal particles that become enmeshed in the floc will thus be removed from the wastewater. In most wastewater applications, the sweep floc mode of operation is used most commonly where particles are to be removed by sedimentation.

The sequence of reactions and events that occur in the coagulation and removal of particles can be illustrated pictorially as shown on Fig. 6–7. In zone 1, sufficient coagulant has not been added to destabilize the colloidal particles, even though some reduction in surface charge may occur due to the presence of Fe^{3+} and some mononuclear hydrolysis species. In zone 2, the colloidal particles have been destabilized by the adsorption of mono- and polynuclear hydrolysis species, and, if allowed to flocculate and settle, the residual turbidity would be lowered as shown. In zone 3, as more coagulant is added, the surface charge of the particles has reversed due to the continued adsorption of mono- and polynuclear hydrolysis species (see Fig. 6–7). As the colloidal particles are now positively charged, they cannot be removed by perikinetic flocculation. As more coagulant is added, zone 4 is reached where large amounts of hydroxide floc will form. As the floc particles settle, the colloidal particles will be removed by the sweep action of the settling floc particles, and the residual turbidity will be lowered as shown. The coagulant dosage required to reach any of the zones will depend on the nature of the colloidal particles and the pH and temperature of the wastewater. Specific constituents (e.g., organic matter) will also have an effect on the coagulant dose.

Figure 6–7

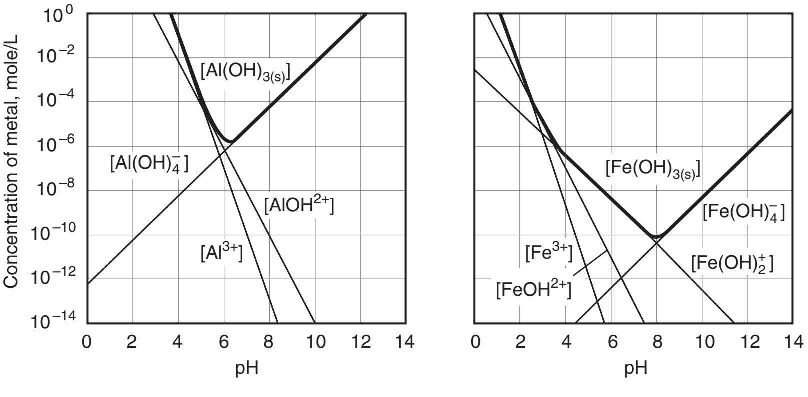
Definition sketch for the effects of the continued addition of a coagulant (e.g., alum) on the destabilization and flocculation of colloidal particles.



It is also important to note that the example reaction sequence given by Eq. (6–7), and the coagulation process illustrated on Fig. 6–7, is time dependent. For example, if it is desired to destabilize the colloidal particles in wastewater with mono- and polynuclear species, then *rapid* and *intense* initial mixing of the metal salt and the wastewater containing the particles to be destabilized is of critical importance. If the reaction is allowed to proceed to the formation of metal hydroxide floc, it will be difficult to contact the chemical and the particles. As discussed below, it has been estimated that the formation of the mono- and polynuclear and polymer hydroxide species occurs in a fraction of second.

Solubility of Metal Salts. To further appreciate the action of the hydrolyzed metal ions it will be useful to consider the solubility of the metal salts. The solubility of the various aluminum [Al(III)] and iron [Fe(III)] species are illustrated on Figs. 6–8(a) and 6–8(b), respectively in which the log molar concentrations have been plotted as a function of pH. In preparing these diagrams, only the mononuclear species for alum and iron have been plotted. The various mononuclear species for alum and iron are given in Table 6–2, along with the corresponding range of acid solubility products reported in the literature. The formation of some of the mononuclear species is also illustrated in Eq. (6–7). It should be noted that Hayden and Rubin (1974) compared experimental and predicted values and concluded that $Al(OH)_2^+$ is not an important mononuclear species. Accordingly, $Al(OH)_2^+$ has not been included in the development of Fig. 6–8(a). The solid lines trace the approximate total concentration of residual soluble alum [see Fig. 6–8(a)] and iron [see Fig. 6–8(b)] after precipitation.

Aluminum hydroxide and ferric hydroxide are precipitated within the areas above the bold line, and polynuclear and polymeric species are formed outside at higher and lower pH values. The region within the square boxes is where most precipitation operations are conducted when these coagulants are used in a sweep floc mode of operation. As shown the operating region for alum precipitation is from a pH range of 5 to about 7, with minimum solubility occurring at a pH of 6.0, and from about 7 to 9 for iron precipitation, with minimum solubility occurring at a pH of 8.0.



$\log[Al^{3+}] = 10.8 - 3pH$

$\log[AlOH^{2+}] = 5.8 - 2pH$

$\log[Al(OH)_4^-] = -12.2 + pH$

$C_T = [Al^{3+}] + [AlOH^{2+}] + [Al(OH)_4^-]$

(a)

$\log[Fe^{3+}] = 3.2 - 3pH$

$\log[FeOH^{2+}] = 1.0 - 2pH$

$\log[Fe(OH)_2^+] = -2.5 - pH$

$\log[Fe(OH)_4^-] = -18.4 + pH$

$C_T = [Fe^{3+}] + [FeOH^{2+}] + [Fe(OH)_2^+] + [Fe(OH)_4^-]$

(b)

Figure 6–8

Solubility diagram for alum [Al(III)] and iron [Fe(III)]. It should be noted that only the mononuclear species have been plotted. The polynuclear species are extremely dependent on the chemistry of the wastewater. The mononuclear species $Al(OH)_2^+$ has not been included in the development of Fig. 6–8(a). Further, because of the wide variation in the solubility and formation constants for the various metal hydroxides, the curves presented in this figure should only be used as a reference guide.

Table 6–2

Reactions and associated equilibrium constants for aluminum and iron species in equilibrium with amorphorous aluminum hydroxide and ferric hydroxide[a]

	Acid equilibrium constants		
Reaction	**Equilibrium constant**	**Range[a]**	**Used for Fig. 6–8**
Aluminum, Al(III)			
$Al(OH)_{3(s)} + 3H^+ \rightleftarrows Al^{3+} + 3H_2O$	$^*K_{s0}$	9.0–10.8	10.8
$Al(OH)_{3(s)} + 2H^+ \rightleftarrows AlOH^{2+} + 2H_2O$	$^*K_{s1}$	4.0–5.8	5.8
$Al(OH)_{3(s)} + H^+ \rightleftarrows Al(OH)_2^+ + H_2O$[b]	$^*K_{s2}$	1.5	1.5
$Al(OH)_{3(s)} \rightleftarrows Al(OH)_3$	$^*K_{s3}$	−4.2	−4.2
$Al(OH)_{3(s)} + H_2O \rightleftarrows AlOH_4^- + H^+$	$^*K_{s4}$	−7.7–(−12.5)	−12.2
Species not considered: $Al^2(OH)_2{}^{4+}$; $Al_8(OH)_{20}^{4+}$; $Al_{13}O_4(OH)_{24}^{7+}$; $Al_{14}(OH)_{32}^{10+}$			
Iron, Fe(III)			
$Fe(OH)_{3(s)} + 3H^+ \rightleftarrows Fe^{3+} + 3H_2O$	$^*K_{s0}$	3.2–4.891	3.2
$Fe(OH)_{3(s)} + 2H^+ \rightleftarrows FeOH^{2+} + 2H_2O$	$^*K_{s1}$	0.91–2.701	1.0
$Fe(OH)_{3(s)} + H^+ \rightleftarrows Fe(OH)_2^+ + H_2O$	$^*K_{s2}$	−0.779–(−2.5)	−2.5
$Fe(OH)_{3(s)} \rightleftarrows Fe(OH)_3$	$^*K_{s3}$	−8.709–(−12.0)	−12.0
$Fe(OH)_{3(s)} + H_2O \rightleftarrows FeOH_4^- + H^+$	$^*K_{s4}$	−16.709–(−19)	−18.4
Species not considered: $Fe_2(OH)_2^{4+}$; $Fe_3(OH)_4^{5+}$			

[a] Abstracted from Benefield et al.(1982), McMurray and Fay (2011), Morel and Hering (1993), Pankow (2012), Snoeyink and Jenkins, (1980), Sawyer et al. (2002), and Stumm and Morgan (1981).

[b] Hayden and Rubin (1974) compared experimental and predicted values and concluded that $Al(OH)_2^+$ is not an important mononuclear species.

Operating Regions for Action of Metal Salts. Because the chemistry of the various reactions is so complex, there is no complete theory to explain the action of hydrolyzed metal ions. To quantify qualitatively the application of alum as a function of pH, taking into account the action of alum as described above, Amirtharajah and Mills (1982) developed the diagram shown on Fig. 6–9. Although Fig. 6–9 was developed for water treatment applications, it has been found to apply reasonably well to most wastewater applications, with minor variations. The approximate regions in which the different phenomenon associated with particle removal in conventional sedimentation and filtration processes are operative are plotted on Fig. 6–9 as a function of the alum dose and the pH of the treated effluent after alum has been added. For example, optimum particle removal by sweep floc occurs in the pH range of 7 to 8 with an alum dose of 20 to 60 mg/L. Generally, for many wastewater effluents that have high pH values (e.g., 7.3 to 8.5), low alum dosages in the range of 5 to 10 mg/L will not be effective. With proper pH control it is possible to operate with low alum dosages. Because the characteristics of wastewater will vary from treatment plant to treatment plant, bench scale and/or pilot plant tests must be conducted to establish the appropriate chemical dosages.

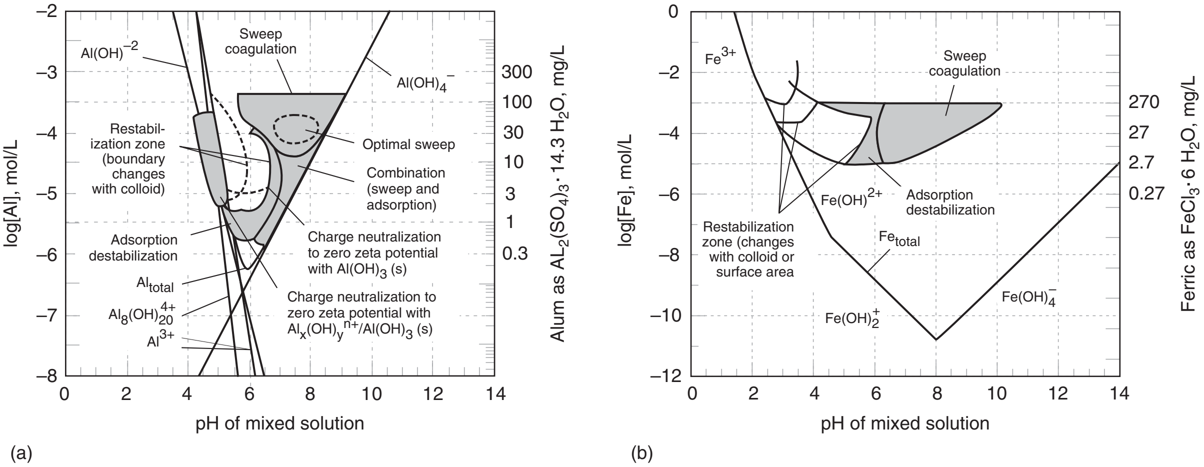


Figure 6–9

Typical operating ranges for coagulation with (a) alum and (b) ferric chloride. (Adapted from Crittenden et al., 2012; Amirtharajah and Mills, 1978).

Importance of Initial Chemical Mixing with Metal Salts. Perhaps the least appreciated fact about chemical addition of metal salts is the importance of the rapid initial mixing of the chemicals with the wastewater to be treated. In a 1967 article, Hudson and Wolfner (1967) noted that "coagulants hydrolyze and begin to polymerize in a fraction of a second after being added to water." Hahn and Stumm (1968) studied the coagulation of silica dispersions with Al(III). They reported that the time required for the formation of mono- and polynuclear hydroxide species appears to be on the order of 10^{-3} s. The time of formation for the polymer species was on the order of 10^{-2} s. Further, they found that the rate limiting step in the coagulation process was the time required for the colloidal transport step brought about by Brownian motion (i.e., perikinetic flocculation) which was estimated to be on the order of 1.5 to 3.3×10^{-3} s. The importance of initial and rapid mixing is also discussed by Amirtharajah and Mills (1982) and Vrale and Jorden (1971). Clearly, based on the literature and actual field evaluations, the instantaneous rapid and intense mixing of metal salts is of critical importance, especially where the metal salts are to be used as coagulants to lower the surface charge of the colloidal particles. It should be noted that although achieving suitable mixing times in large treatment plants is often difficult, adequate mixing times can be achieved by using multiple mixers. Typical mixing times for various chemicals are reported in Table 6–24 in Sec. 6–11.

6–3 CHEMICAL PRECIPITATION FOR IMPROVED PLANT PERFORMANCE

Chemical precipitation, as noted previously, involves the addition of chemicals to alter the physical state of dissolved and suspended solids and facilitate their removal by sedimentation. In the past, chemical precipitation was often used to enhance the degree of TSS and BOD removal (1) where there were seasonal variations in the concentration of the wastewater (such as in cannery wastewater), (2) where an intermediate degree of treatment was required, and (3) as an aid to the sedimentation process. Since about 1970, the need to provide more complete removal of the organic compounds and nutrients (nitrogen and phosphorus) contained in wastewater has brought about renewed interest in chemical precipitation. In current practice, chemical precipitation is used (1) as a means of improving the performance of primary settling facilities, (2) as a basic step in the independent physical-chemical treatment of wastewater, (3) for the removal of phosphorus, (4) for the removal of heavy metals, and (5) for improving the quality of water to be reused by softening.

Aside from the determination of the required chemical dosages and specification of the rapid mixing and flocculation facilities, the principal design considerations related to the use of chemical precipitation involve the analysis and design of the necessary sludge processing facilities, and the selection and design of the chemical storage, feeding, piping, and control systems (see Sec. 6–11).

Chemical Reactions in Wastewater Precipitation Applications

Over the years a number of different substances have been used as precipitants (Metcalf and Eddy, 1935). The degree of clarification obtained depends on the quantity of chemicals used and the care with which the process is controlled. It is possible by chemical precipitation to obtain a clear effluent, substantially free from matter in suspension or in the colloidal state. The chemicals added to wastewater interact with substances that are either normally present in the wastewater or added for this purpose. The most common chemicals are listed in Table 6–3. The reactions involved with (1) alum, (2) lime, (3) ferrous sulfate (copperas) and lime, (4) ferric chloride, (5) ferric chloride and lime, and (6) ferric sulfate and lime are considered in the following discussion.

Table 6–3

Inorganic chemicals used most commonly for coagulation and precipitation processes in wastewater treatment

Chemical	Formula	Molecular weight	Equiv. weight	Availability	
				Form	Percent
Alum	$Al_2(SO_4)_3 \cdot 18H_2O^a$	666.5		Liquid Lump	8.5 (Al_2O_3) 17 (Al_2O_3)
	$Al_2(SO_4)_3 \cdot 14H_2O^a$	594.4	114	Liquid Lump	8.5 (Al_2O_3) 17 (Al_2O_3)
Aluminum chloride	$AlCl_3$	133.3	44	Liquid	
Calcium hydroxide (lime)	$Ca(OH)_2$	56.1 as CaO	40	Lump Powder Slurry	63–73 (CaO) 85–99 (CaO) 15–20 [$Ca(OH)_2$]
Ferric chloride	$FeCl_3$	162.2	91	Liquid Lump	20 (Fe) 20 (Fe)
Ferric sulfate	$Fe_2(SO_4)_3$	400	51.5	Granular	18.5 (Fe)
Ferrous sulfate (copperas)	$FeSO_4 \cdot 7H_2O$	278.1	139	Granular	20 (Fe)
Sodium aluminate	$Na_2Al_2O_4$	163.9	100	Flake	46 (Al_2O_3)

[a] Number of bound water molecules will typically vary from 14 to 18.

Alum. When alum is added to wastewater containing calcium and magnesium bicarbonate alkalinity, a precipitate of aluminum hydroxide will form. The overall reaction that occurs when aluminum sulfate is added to water may be illustrated as follows:

$$\underset{\substack{\text{Calcium}\\ \text{bicarbonate}\\ \text{(Soluble)}}}{\overset{3\times 100\ (\text{as }\mathrm{CaCO_3})}{3\mathrm{Ca(HCO_3)_2}}} + \underset{\substack{\text{Aluminum}\\ \text{sulfate}\\ \text{(Soluble)}}}{\overset{666.5}{\mathrm{Al_2(SO_4)_3{\cdot}18H_2O}}} \rightleftarrows \underset{\substack{\text{Aluminum}\\ \text{hydroxide}\\ \text{(Insoluble)}}}{\overset{2\times 78}{2\mathrm{Al(OH)_3}}} + \underset{\substack{\text{Calcium}\\ \text{sulfate}\\ \text{(Soluble)}}}{\overset{3\times 136}{3\mathrm{CaSO_4}}} + \underset{\substack{\text{Carbon}\\ \text{dioxide}\\ \text{(Soluble)}}}{\overset{6\times 44}{6\mathrm{CO_2}}} + \overset{18\times 18}{18\mathrm{H_2O}} \tag{6–9}$$

The numbers above the chemical formulas are the combined molecular weights of the different substances and, therefore, denote the quantity of each one involved. The precipitation reaction given above also occurs with the addition of aluminum chloride ($AlCl_3$). The insoluble aluminum hydroxide is a gelatinous floc that settles slowly through the wastewater, sweeping out suspended material and producing other changes. The reaction is exactly analogous when magnesium bicarbonate is substituted for the calcium salt.

Because alkalinity in Eq. (6–9) is reported in terms of calcium carbonate ($CaCO_3$), the molecular weight of which is 100, the quantity of alkalinity required to react with 10 mg/L of alum is

$$(10.0\ \text{mg/L})\left[\frac{3(100\ \text{g/mole})}{(666.5\ \text{g/mole})}\right] = 4.5\ \text{mg/L}$$

If less than this amount of alkalinity is available, it must be added. Lime is commonly used for this purpose when necessary, but it is seldom required in the treatment of wastewater.

Lime. When lime alone is added as a precipitant, the principles of clarification are explained by the following reactions for carbonic acid [Eq. (6–10)] and alkalinity [Eq. (6–11)]:

$$\underset{\substack{\text{Carbonic}\\ \text{acid}\\ \text{(Soluble)}}}{\overset{44\text{ as }\mathrm{CO_2}}{\mathrm{H_2CO_3}}} + \underset{\substack{\text{Calcium}\\ \text{hydroxide}\\ \text{(Slightly}\\ \text{soluble)}}}{\overset{56\ (\text{as CaO})}{\mathrm{Ca(OH)_2}}} \rightleftarrows \underset{\substack{\text{Calcium}\\ \text{carbonate}\\ \text{(Somewhat}\\ \text{soluble)}}}{\overset{100}{\mathrm{CaCO_3}}} + \overset{2\times 18}{2\mathrm{H_2O}} \tag{6–10}$$

$$\underset{\substack{\text{Calcium}\\ \text{bicarbonate}\\ \text{(Soluble)}}}{\overset{100\ (\text{as }\mathrm{CaCO_3})}{\mathrm{Ca(HCO_3)_2}}} + \underset{\substack{\text{Calcium}\\ \text{hydroxide}\\ \text{(Slightly}\\ \text{soluble)}}}{\overset{56\text{ as CaO}}{\mathrm{Ca(OH)_2}}} \rightleftarrows \underset{\substack{\text{Calcium}\\ \text{carbonate}\\ \text{(Somewhat}\\ \text{soluble)}}}{\overset{2\times 100}{2\mathrm{CaCO_3}}} + \overset{2\times 18}{2\mathrm{H_2O}} \tag{6–11}$$

A sufficient quantity of lime must, therefore, be added to combine with all the free carbonic acid and with the carbonic acid of the bicarbonates (half-bound carbonic acid) to produce calcium carbonate. Much more lime is generally required when it is used alone than when sulfate of iron is also used (see the following discussion). Where industrial wastes introduce mineral acids or acid salts into the wastewater, these must be neutralized before precipitation can take place. Typical lime clarification facilities were shown on Fig. 6–1.

Ferrous Sulfate and Lime. In most cases, ferrous sulfate cannot be used alone as a precipitant because lime must be added at the same time to form a precipitate. When ferrous sulfate alone is added to wastewater the following reactions occur:

$$\overset{278}{\underset{\substack{\text{Ferrous}\\ \text{sulfate}\\ \text{(Soluble)}}}{\mathrm{FeSO_4{\cdot}7H_2O}}} + \overset{100(\text{as CaCO}_3)}{\underset{\substack{\text{Calcium}\\ \text{bicarbonate}\\ \text{(Soluble)}}}{\mathrm{Ca(HCO_3)_2}}} \rightleftarrows \overset{178}{\underset{\substack{\text{Ferrous}\\ \text{bicarbonate}\\ \text{(Soluble)}}}{\mathrm{Fe(HCO_3)_2}}} + \overset{136}{\underset{\substack{\text{Calcium}\\ \text{sulfate}\\ \text{(Soluble)}}}{\mathrm{CaSO_4}}} + \overset{7\times18}{\mathrm{7H_2O}} \tag{6–12}$$

$$\overset{178}{\underset{\substack{\text{Ferrous}\\ \text{bicarbonate}\\ \text{(Soluble)}}}{\mathrm{Fe(HCO_3)_2}}} \rightarrow \overset{4\times89.9}{\underset{\substack{\text{Ferrous}\\ \text{hydroxide}\\ \text{(Very slightly}\\ \text{soluble)}}}{\mathrm{Fe(OH)_2}}} + \underset{\substack{\text{Carbon}\\ \text{dioxide}\\ \text{(Soluble)}}}{\mathrm{CO_2}} \tag{6–13}$$

If sufficient alkalinity is not available, lime is often added in excess in conjunction with ferrous sulfate. The reaction of ferrous sulfate with lime is as follows

$$\overset{178}{\underset{\substack{\text{Ferrous}\\ \text{bicarbonate}\\ \text{(soluble)}}}{\mathrm{Fe(HCO_3)_2}}} + \overset{2\times56\text{ as CaO}}{\underset{\substack{\text{Calcium}\\ \text{hydroxide}\\ \text{(Slightly}\\ \text{soluble)}}}{\mathrm{2Ca(OH)_2}}} \rightleftarrows \overset{89.9}{\underset{\substack{\text{Ferrous}\\ \text{hydroxide}\\ \text{(Very slightly}\\ \text{soluble)}}}{\mathrm{Fe(OH)_2}}} + \overset{2\times100}{\underset{\substack{\text{Calcium}\\ \text{carbonate}\\ \text{(Somewhat}\\ \text{soluble)}}}{\mathrm{2CaCO_3}}} + \overset{2\times18}{\mathrm{2H_2O}} \tag{6–14}$$

The ferrous hydroxide can be oxidized to ferric hydroxide, the final form desired, by oxygen dissolved in the wastewater. The reaction is

$$\overset{89.9}{\underset{\substack{\text{Ferrous}\\ \text{hydroxide}\\ \text{(Very slightly}\\ \text{soluble)}}}{\mathrm{Fe(OH)_2}}} + \overset{1/4\times32}{\underset{\substack{\text{Oxygen}\\ \text{(Soluble)}}}{\mathrm{1/4O_2}}} + \overset{1/2\times18}{\mathrm{1/2\ H_2O}} \rightleftarrows \overset{106.9}{\underset{\substack{\text{Ferric}\\ \text{hydroxide}\\ \text{(Insoluble)}}}{\mathrm{Fe(OH)_3}}} \tag{6–15}$$

The insoluble ferric hydroxide is formed as a bulky, gelatinous floc similar to the alum floc. The alkalinity required for a 10 mg/L dosage of ferrous sulfate [see Eq. (6–12)] is

$$(10.0\text{ mg/L})\left[\frac{(100\text{ g/mole})}{(278\text{ g/mole})}\right] = 3.6\text{ mg/L}$$

The lime required (see Eq. 6–14) is

$$(10.0\text{ mg/L})\left[\frac{2(56\text{ g/mole})}{(278\text{ g/mole})}\right] = 4.0\text{ mg/L}$$

The oxygen required (see Eq. 6–15) is

$$(10.0\text{ mg/L})\left[\frac{(32\text{ g/mole})}{4\,(278\text{ g/mole})}\right] = 0.29\text{ mg/L}$$

Because the formation of ferric hydroxide is dependent on the presence of dissolved oxygen, the reaction given in Eq. (6–15) cannot be completed in most wastewaters, and, as a result, ferrous sulfate is not used commonly for wastewater.

Ferric Chloride. Because of the many problems associated with the use of ferrous sulfate, ferric chloride is the iron salt used most commonly in precipitation applications. When ferric chloride is added to wastewater the following reaction takes place:

$$\begin{array}{ccccccccc}
2\times 162.2 & & 3\times 100\text{ (as }CaCO_3) & & 2\times 106.9 & & & & \\
2FeCl_3 & + & 3Ca(HCO_3)_2 & \rightleftarrows & 2Fe(OH)_3 & + & 3CaCl_2 & + & 6CO_2 \\
\text{Ferric} & & \text{Calcium} & & \text{Ferric} & & \text{Calcium} & & \text{Carbon} \\
\text{chloride} & & \text{bicarbonate} & & \text{hydroxide} & & \text{chloride} & & \text{dioxide} \\
\text{(Soluble)} & & \text{(Soluble)} & & \text{(Insoluble)} & & \text{(Soluble)} & & \text{(Soluble)}
\end{array} \qquad (6\text{–}16)$$

Ferric Chloride and Lime. If lime is added to supplement the natural alkalinity of the wastewater, the following reaction can be assumed to occur:

$$\begin{array}{ccccccc}
2\times 162.2 & & 3\times 56\text{ (as CaO)} & & 2\times 106.9 & & 3\times 111 \\
2FeCl_3 & + & 3Ca(OH)_2 & \rightleftarrows & 2Fe(OH)_3 & + & 3CaCl_2 \\
\text{Ferric} & & \text{Calcium} & & \text{Ferric} & & \text{Calcium} \\
\text{chloride} & & \text{hydroxide} & & \text{hydroxide} & & \text{chloride} \\
\text{(Soluble)} & & \text{(Slightly} & & \text{(Insoluble)} & & \text{(Soluble)} \\
 & & \text{soluble)} & & & &
\end{array} \qquad (6\text{–}17)$$

Ferric Sulfate and Lime. The overall reaction that occurs when ferric sulfate and lime are added to wastewater may be represented as follows:

$$\begin{array}{ccccccc}
399.9 & & 3\times 56\text{ (as CaO)} & & 2\times 106.9 & & 3\times 136 \\
Fe_2(SO_4)_3 & + & 3Ca(OH)_2 & \rightleftarrows & 2Fe(OH)_3 & + & 3CaSO_4 \\
\text{Ferric} & & \text{Calcium} & & \text{Ferric} & & \text{Calcium} \\
\text{sulfate} & & \text{hydroxide} & & \text{hydroxide} & & \text{sulfate} \\
\text{(Soluble)} & & \text{(Slightly} & & \text{(Insoluble)} & & \text{(Soluble)} \\
 & & \text{soluble)} & & & &
\end{array} \qquad (6\text{–}18)$$

Chemically Enhanced Primary Treatment (CEPT)

The degree of clarification obtained when chemicals are added to untreated wastewater depends on the type and quantity of chemicals used, mixing times, and the care with which the process is monitored and controlled. With chemical precipitation, it is possible to remove 80 to 90 percent of the total suspended solids (TSS) including some colloidal particles and 50 to 80 percent of the BOD. Comparable removal values for well-designed and well-operated primary sedimentation tanks without the addition of chemicals are 50 to 70 percent of the TSS and 25 to 40 percent of the BOD. Enhanced removal of solids and BOD during primary treatment is a critical issue in energy management at wastewater treatment facilities. Solids from primary treatment have a high energy value, some of which can be recovered by anaerobic digestion or other thermal conversion processes (see Chap. 14). In addition, the removal of constituents in primary treatment that exert an oxygen demand reduces the amount of aeration energy required for secondary treatment. Thus, primary treatment has been

Table 6–4

Recommended surface loading rates for sedimentation tanks for various chemical suspensions

	Overflow rate			
	$m^3/m^2 \cdot d$		$gal/ft^2 \cdot d$	
Suspension	**Typical range**	**Peak hour**	**Typical range**	**Peak hour**
Alum floc[a]	30–70	80	700–1700	2000
Iron floc[a]	30–70	80	700–1700	2000
Lime floc[a]	35–80	90	900–2000	2200
Untreated wastewater	30–70	80	700–1700	2000

[a] Mixed with the settleable suspended solids in the untreated wastewater and colloidal or other suspended solids swept out by the floc.

Note: $m^3/m^2 \cdot d \times 24.5424 = gal/ft^2 \cdot d$.

identified as a key area of research in improving the sustainability of wastewater treatment processes.

Because of the variable characteristics of wastewater, the effectiveness of alternative coagulants and the required chemical dosages should be evaluated based on the results of bench- or pilot-scale tests. For example, dosages of $FeCl_3$ for coagulation of raw or screened wastewater range typically from about 15 to 40 mg/L, after a short reaction time anionic polymer may be added at a dosage of 0.1 to 1 mg/L to enhance floc development. Recommended surface loading rates for various chemical suspensions to be used in the design of the sedimentation facilities are given in Table 6–4.

Independent Physical-Chemical Treatment

In some localities, industrial wastewater has been found to be difficult to treat by biological means. In such situations, physical-chemical treatment in the absence of a biological treatment step may be an alternative approach. This method of treatment has met with limited success because of its lack of consistency in meeting discharge requirements, high costs for chemicals, handling and disposal of the great volumes of sludge resulting from the addition of chemicals, and numerous operating problems. Based on typical performance results of full scale plants using activated carbon, the activated carbon columns removed only 50 to 60 percent of the applied total BOD, and the plants did not meet consistently the effluent standards for secondary treatment. In some instances, substantial process modifications have been required to reduce the operating problems and meet performance requirements, or the process has been replaced by biological treatment. Because of these reasons, new applications of physical-chemical treatment for municipal wastewater are rare. Physical-chemical treatment is used more extensively for the treatment of industrial wastewater. Depending on the treatment objectives, the required chemical dosages and application rates should be determined from bench- or pilot-scale tests.

A flow diagram for the physical-chemical treatment of untreated wastewater is presented on Fig. 6–10. As shown, after first stage precipitation and pH adjustment by recarbonation (if required), the wastewater is passed through a granular medium filter to remove any residual floc and then through carbon columns to remove dissolved organic compounds. The filter is shown as optional, but its use is recommended to reduce the blinding and headloss buildup in the carbon columns. The treated effluent from the carbon column is usually chlorinated before discharge to the receiving waters.

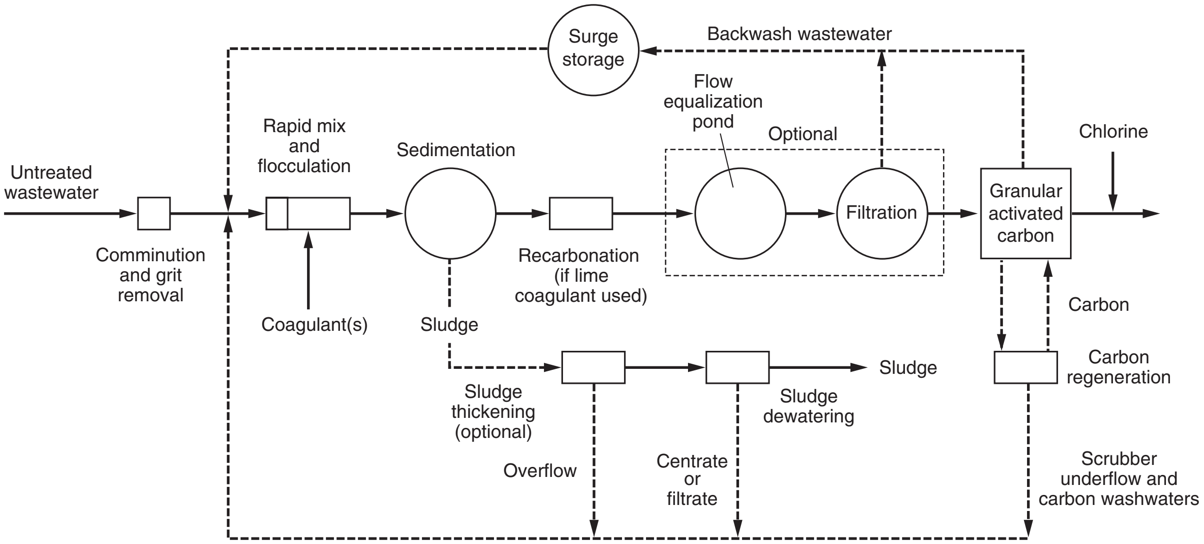


Figure 6–10

Typical flow diagram of an independent physical-chemical treatment plant.

Estimation of Sludge Quantities from Chemical Precipitation

The handling and disposal of the sludge resulting from chemical precipitation is one of the greatest difficulties associated with chemical treatment. Sludge is produced in great volume from most chemical precipitation operations, often reaching 0.5 percent of the volume of wastewater treated when lime is used. The computational procedures involved in estimating the quantity of sludge resulting from chemical precipitation with ferric chloride and lime are illustrated in Example 6–1.

EXAMPLE 6–1 **Estimation of Sludge Volume from Chemical Precipitation of Untreated Wastewater** Estimate the mass and volume of sludge produced from untreated wastewater without and with the use of ferric chloride for the enhanced removal of TSS. Also estimate the amount of lime required for the specified ferric chloride dose. Assume that 60 percent of the TSS is removed in the primary settling tank without the addition of chemicals, and that the addition of ferric chloride results in an increased removal of TSS to 85 percent. Assume that the following data apply to this situation:

1.	Wastewater flowrate, m^3/d	1000
2.	Wastewater TSS, mg/L	220
3.	Wastewater alkalinity as $CaCO_3$, mg/L	136
4.	Ferric chloride ($FeCl_3$) added, kg/1000 m^3	40
5.	Raw sludge properties	
	Specific gravity	1.03
	Moisture content, %	94

6. Chemical sludge properties (from Chap. 13)

Specific gravity	1.05
Moisture content, %	92.5

Solution

1. Compute the mass of TSS removed without and with chemicals.
   a. Determine the mass of TSS removed without chemicals.

$$M_{\text{TSS}} = \frac{0.6\,(220\ \text{g/m}^3)(1000\ \text{m}^3/\text{d})}{(10^3\ \text{g/1 kg})} = 132.0\ \text{kg/d}$$

   b. Determine the mass of TSS removed with chemicals.

$$M_{\text{TSS}} = \frac{0.85\,(220\ \text{g/m}^3)(1000\ \text{m}^3/\text{d})}{(10^3\ \text{g/1 kg})} = 187.0\ \text{kg/d}$$

2. Using Eq. (6–16), determine the mass of ferric hydroxide [$Fe(OH)_3$] produced from the addition of 40 kg/1000 m^3 of $FeCl_3$.

$$Fe(OH)_3\ \text{formed} = 40 \times \left(\frac{2 \times 106.9}{2 \times 162.2}\right) = 26.4\ \text{kg/1000 m}^3$$

3. Using Eq. (6–17), determine the mass of lime required to convert the ferric chloride to ferric hydroxide $Fe(OH)_3$.

$$\text{Lime required} = 40 \times \left(\frac{3 \times 56}{2 \times 162.2}\right) = 20.7\ \text{kg/1000 m}^3$$

$$\text{Lime required, expressed as alkalinity} = 20.7 \times \left(\frac{100}{56}\right) = 37\ \text{kg/1000 m}^3$$

$$\begin{aligned}\text{Alkalinity available per 1000 m}^3 &= (136\ \text{g/m}^3)(1000\ \text{m}^3)/(10^3\ \text{g/1 kg})\\ &= 136\ \text{kg/1000 m}^3\end{aligned}$$

   Because there is sufficient natural alkalinity no lime addition will be required.

4. Determine the total amount of sludge on a dry basis resulting from chemical precipitation.

   Total dry solids = 187 + 26.4 = 213.4 kg/1000 m^3

5. Determine the total volume of sludge resulting from chemical precipitation, assuming that the sludge has a specific gravity of 1.05 and a moisture content of 92.5 percent.

$$V_s = \frac{(213.4\ \text{kg/d})}{(1.05)(10^3\ \text{kg/m}^3)(1 - 0.925)} = 2.71\ \text{m}^3/\text{d}$$

6. Determine the total volume of sludge without chemical precipitation, assuming that the sludge has a specific gravity of 1.03 and a moisture content of 94 percent.

$$V_s = \frac{(132\ \text{kg/d})}{(1.03)(10^3\ \text{kg/m}^3)(1 - 0.94)} = 2.13\ \text{m}^3/\text{d}$$

7. Prepare a summary table of the sludge mass and volume produced without and with chemical precipitation.

	Sludge	
Treatment	Mass, kg/d	Volume, m^3/d
Without chemical precipitation	132.0	2.13
With chemical precipitation	213.4	2.71

Comment The magnitude of the sludge disposal problem when chemicals are used is evident from a review of the data presented in the summary table given in Step 7. Even larger volumes of sludge are produced when lime is used as the primary precipitant (see Example 6–2).

6–4 CHEMICAL PHOSPHORUS REMOVAL

Phosphorus in wastewater can be incorporated into either biological solids (e.g., microorganisms) or removed by the addition of chemicals. The fundamentals of biological phosphorus removal are considered in Chap. 8. The chemical removal of phosphorus is introduced in this section. The topics to be considered include (1) the chemicals used for phosphorus removal, (2) phosphorus removal from liquid stream with metal salts, (3) phosphorus removal from the liquid stream with calcium, and (4) strategies for chemical phosphorus removal. The removal of phosphorus from return flows and sidestreams is considered in Chap. 15. General consideration that may affect the implementation of chemical phosphorus removal systems are summarized in Table 6–5.

Chemicals Used for Phosphorus Removal

The principal chemicals used for the removal of phosphorus from the liquid stream include: aluminum [Al(lll)], ferric iron [Fe(lll)], ferrous iron [Fe(ll)], and calcium [Ca(ll)]. Polymers have been used effectively in conjunction with metal salts and lime as flocculant

Table 6–5

Factors affecting the implementation of chemical phosphorus removal[a]

- Influent phosphorus concentration and speciation
- Wastewater suspended solids
- Alkalinity
- Chemical cost (including transportation)
- Reliability of chemical supply
- Sludge handling facilities
- Ultimate disposal methods
- Compatibility with other treatment processes
- Required effluent phosphorus concentration
- Seasonal permit requirements
- Efficacy of other options (e.g., biological P removal)
- Management of sidestreams
- Phosphorus recovery objectives

[a] Adapted in part from Kugelman (1976).

aids. Because the chemistry of phosphate removal with aluminum and iron is quite different than with calcium, the two different types of chemical removal are considered separately in the following discussion.

Phosphate Removal with Aluminum and Iron. The removal of phosphate by the addition of metal salts is thought to occur in a number of different ways, including (WEF, 1998, 2011):

1. Formation of hydrous ferric or aluminum oxides which serve as a substrate for phosphate adsorption
2. Incorporation of phosphate into the hydrous oxide structure
3. Formation of mixed cation phosphates (e.g., Fe or Al phosphates)
4. Formation of ferric or aluminum phosphate

It should be noted that phosphate removal is only accomplished if, after the formation of the above compounds, they are removed by sedimentation (precipitation) or by filtration.

Historically, the following two reactions were used to describe the formation of ferric or aluminum phosphate.

Phosphate precipitation with aluminum:

$$\mathrm{Al}^{3+} + \mathrm{H}_n\mathrm{PO}_4^{3-n} \rightleftarrows \mathrm{AlPO}_4\downarrow + n\mathrm{H}^+ \tag{6–19}$$

Phosphate precipitation with iron:

$$\mathrm{Fe}^{3+} + \mathrm{H}_n\mathrm{PO}_4^{3-n} \rightleftarrows \mathrm{FePO}_4\downarrow + n\mathrm{H}^+ \tag{6–20}$$

Unfortunately, these reactions are deceptively simple and generally do not apply. It has been found that ferric phosphate occurs near a pH value of 3.5 and does not occur above pH 5 (Smith et al., 2008). Further, any precipitation reaction must be considered in light of the many competing reactions and their associated equilibrium constants, and the effects of alkalinity, pH, trace elements, and ligands found in wastewater. Based on more recent studies (Sedlak, 1991; WEF, 2011) it appears that the following overall reaction may provide a better description of what actually happens when metal salts are added for phosphorus removal.

$$r\mathrm{Me}^{3+} + \mathrm{HPO}_4^{-} + (3r - 1)\mathrm{OH}^- \rightarrow \mathrm{Me}_r\cdot\mathrm{H}_2\mathrm{PO}_4(\mathrm{OH})_{3r-1}(s) \tag{6–21}$$

where $r = 1.6$ for Fe(III) and 0.8 for Al(III).

However, because of the many competing reactions, Eq. (6–21) cannot be used to estimate the required chemical dosages directly. Therefore, dosages are generally established on the basis of bench-scale tests, occasionally by full-scale tests, especially if polymers are used, and from data from other operating treatment plants. For example, for equimolar initial concentrations of Al(III), Fe(III), and phosphate, the total concentration of soluble phosphate in equilibrium with both insoluble FePO_4 and AlPO_4 is shown on Fig. 6–11. The solid lines trace the concentration of residual soluble phosphate after precipitation. Metal phosphates are precipitated within the shaded area; mixed complex polynuclear species are formed outside of the shadowed area toward higher and lower pH values.

Phosphate Removal with Calcium. Calcium is usually added in the form of lime $\mathrm{Ca(OH)_2}$. From the equations presented previously, it will be noted that when lime is added to water it reacts with the natural bicarbonate alkalinity to precipitate $\mathrm{CaCO_3}$. As the pH value of the wastewater increases beyond about 10, excess calcium ions will then react

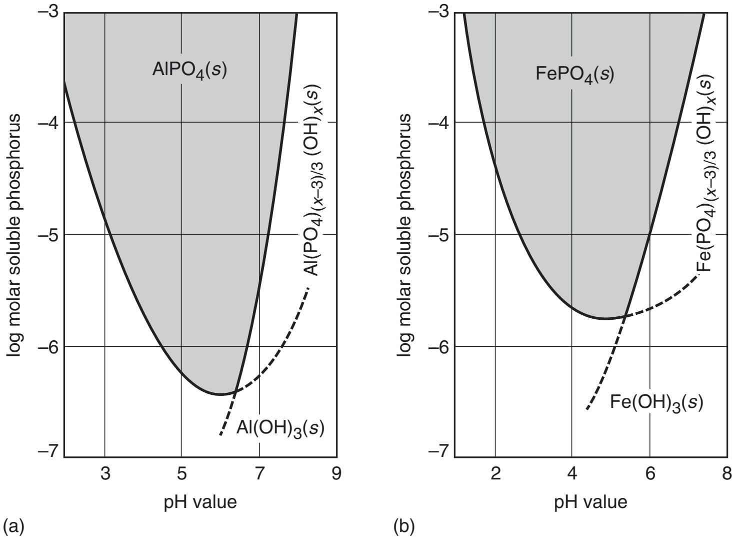


Figure 6–11

Concentration of aluminum and ferric phosphate in equilibrium with soluble phosphorus:
(a) Al(III)-phosphate;
(b) Fe(III)-phosphate.

with the phosphate, as shown in Eq. (6–22), to precipitate hydroxylapatite $Ca_{10}(PO_4)_6(OH)_2$. In practice, partial removal of phosphate has been accomplished by adding lime upstream of primary clarifiers to a target pH around 9.

$$10Ca^{2+} + 6PO_4^{3-} + 2OH^- \rightleftarrows \underset{\text{Hydroxylapatite}}{Ca_{10}(PO_4)_6(OH)_2} \tag{6–22}$$

Because of the reaction of lime with the alkalinity of the wastewater, the quantity of lime required will, in general, depend primarily on the alkalinity of the wastewater (see Fig. 6–12).

Estimate Chemical Requirements for Phosphate Removal. The chemical requirement for the removal of phosphate with metal salts and calcium are as follows.

For Aluminum and Iron.

$$Al_{dose} = (Al/P)(C_{P,in} - C_{P,res})[(26.98\ g/mole\ Al)/(30.97\ g/mole\ P)] \tag{6–23}$$

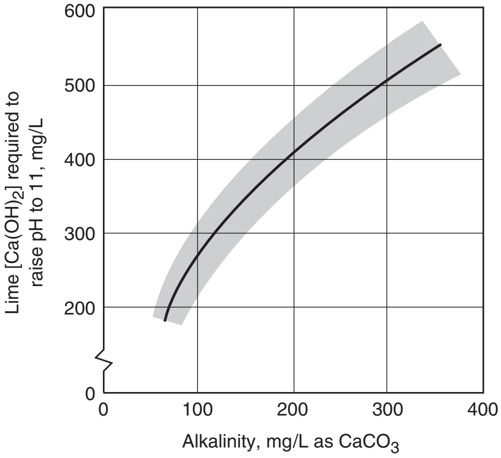
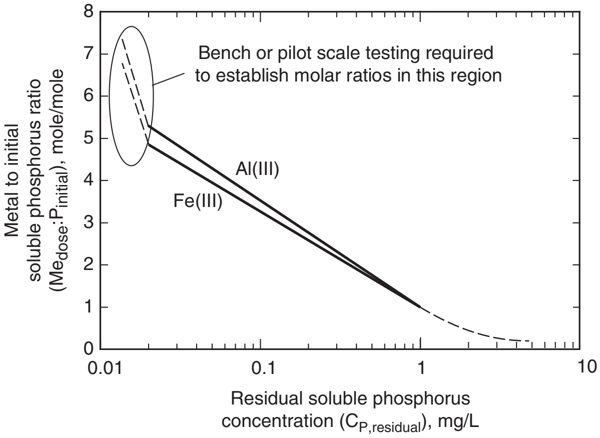


Figure 6–12

Lime dosage required to raise the pH to 11 as a function of untreated wastewater alkalinity.

Figure 6–13

Soluble phosphorus removal by aluminum and iron addition. (Adapted from Szabo et al., 2008.)



$$\text{Fe(III)}_{\text{dose}} = (\text{Fe/P})(C_{\text{P,in}} - C_{\text{P,res}})[(55.85\ \text{g/mole Fe})/(30.97\ \text{g/mole P})] \tag{6–24}$$

where $C_{p,in}$ = influent phosphate concentration, mg/L
$C_{p,res}$ = residual phosphate concentration, mg/L

The alum or ferric dose corresponds to the total amount that would be added for the removal of orthophosphate across the treatment plant. Molar ratios for aluminum and iron are given on Fig. 6–13 for various levels of residual orthophosphate. Typically, dosages of aluminum and iron salts on a molar ratio basis usually fall in the range of 1 to 3 if the residual phosphorus in the secondary effluent is on the order of 0.5 mg/L. The exact application rate is determined by on-site testing and varies with the characteristics of the wastewater and the desired variability in phosphorus removal rates.

For Calcium. The quantity of lime required to precipitate the phosphorus in wastewater is typically about 1.4 to 1.5 times the total alkalinity expressed as $CaCO_3$. When lime is added to raw wastewater or to secondary effluent, pH adjustment is usually required before subsequent treatment or disposal. Recarbonation with carbon dioxide (CO_2) is used to lower the pH value. Because of the additional expense associated with the use of lime, metal salts are now used most commonly for the chemical removal of phosphorus.

EXAMPLE 6–2 Determination of Ferric Chloride Dosage for Phosphorus Removal

Determine the amount of ferric chloride required to precipitate phosphorus from untreated wastewater with the characteristics given below. Also determine the required ferric chloride storage capacity if a 15 d supply is to be stored at the treatment facility and the added quantity of sludge generated from the ferric chloride addition.

1. Wastewater flowrate, m^3/d	3800
2. Wastewater TSS, mg/L	220
3. TSS removal without iron addition, %	60
4. TSS removal with iron addition, %	75
5. Influent total P, g/m^3	7

6.	Influent PO_4^{3-} as P, mg/L	5
7.	Effluent PO_4^{3-} as P, mg/L	0.1
8.	Wastewater alkalinity as $CaCO_3$, mg/L	240
9.	Ferric chloride solution, %	37
10.	Ferric chloride, unit weight	1.35 kg/L
11.	Raw sludge properties	
	Specific gravity	1.03
	Moisture content, %	94
12.	Chemical sludge properties (from Chap. 13)	
	Specific gravity	1.05
	Moisture content, %	92.5

Solution

1. Determine the weight of iron required to remove orthophosphate.
   a. From Fig. 6–13 for an effluent PO_4^{3-} concentration of 0.1 mg/L the required (Fe/P) mole ratio is approximately 3.3.
   b. Using Eq. 6–24, the required ferric chloride dose is

$$\text{Fe(III)}_{\text{dose}} = (\text{Fe/P})(C_{\text{P,in}} - C_{\text{P,res}})[(55.85\,\text{g/mole Fe})/(30.97\,\text{g/mole P})]$$

Substitute known values and solve for the dose

$$\begin{aligned}\text{Fe(III)}_{\text{dose}} &= (3.3)(5 - 0.1)[(55.85\text{ g/mole Fe})/(30.97\text{g/mole P})]\\ &= 29.2\text{ mg/L}\end{aligned}$$

2. Determine primary effluent P concentration.

$$\text{P, mg/L} = 7 - (5 - 0.1) = 2.1\text{ mg/L}$$

3. Determine the amount of ferric iron required per day.

$$\text{Fe Dose} = (3800\text{ m}^3\text{/d})(29.2\text{ mg/L})(1\text{ kg/}10^3\text{ g}) = 111.0\text{ kg/d}$$

4. Determine the amount of ferric chloride solution required per day and the 15 d storage requirement.
   a. Determine the percent ferric iron in $FeCl_3$.

$$\text{Percent Fe in FeCl}_3 = (55.85/162.3) \times 100 = 34.4\%$$

   b. Determine amount of a 34.4% solution of ferric chloride required per day.

$$\text{FeCl}_3\text{ solution} = [(111.0\text{ kg/d})/34.4](100) = 322.7\text{ kg/d}$$

   c. Determine volume of required $FeCl_3$ solution per day.

$$\text{FeCl}_3\text{ volume} = [(322.7\text{ kg/d})/0.0.37 \times 1.35\ ](1\text{ L/kg}) = 646.0\text{ L/d}$$

   d. Determine 15-d storage requirement based on average flowrate.

$$\text{15-d storage requirement} = (646.0\text{ L/d})(1\text{ m}^3/10^3\text{ L})(15) = 10.3\text{ m}^3$$

5. Determine the total mass of sludge on a dry basis resulting from chemical precipitation.
   a. Estimate the additional TSS removal resulting from the addition of $FeCl_3$ for P removal.

$$\begin{aligned}\text{Additional sludge} &= (0.15)(220\text{ g/m}^3)\ (3800\text{ m}^3\text{/d})(1\text{ kg/}10^3\text{ g})\\ &= 125.4\text{ kg/d}\end{aligned}$$

   b. Estimate the additional sludge resulting from the precipitate formed with P using Eq. (6–21).

$$1.6\text{Fe}^{3+} + \text{HPO}_4^{-} + 3.8\text{OH} \rightarrow \text{Fe}_{1.6}\cdot\text{H}_2\text{PO}_4(\text{OH})_{3.8}(\text{s})$$

Fe dose = (29.2 mg Fe/L)(1 g/10^3 mg)/(55.85 g/mole) = 0.52×10^{-3} mole/L

P removed = [(5 − 0.1) mg P/L)](1 g/10^3 mg)/(30.97 g/mole)
= 0.16×10^{-3} mole/L

$Fe_{1.6} \cdot H_2PO_4(OH)_{3.8}$ sludge = (0.16×10^{-3} mole/L)(251 g/mole)(10^3 mg/1 g)
= 40.2 mg/L

c. Estimate the additional sludge resulting from $Fe(OH)_3$.

Excess Fe added = 0.52×10^{-3} mole/L – 1.6 (0.16×10^{-3} mole/L)
= 0.264×10^{-3} mole/L

$Fe(OH)_3$ sludge = (0.264×10^{-3} mole/L)(106.8 g/mole)(10^3 mg/1 g)
= 28.2 mg/L

d. Estimate total chemical sludge resulting $FeCl_3$ addition.

Excess sludge = 40.2 mg/L + 28.2 mg/L = 68.4 mg/L

Excess sludge = (3800 m^3/d)(68.4 mg/L)(1 kg/10^3 g) = 259.9 kg/d

d. Estimate total excess sludge resulting $FeCl_3$ addition.

Total excess sludge = 125.4 kg/d + 259.9 kg/d = 385.3 kg/d

6. Compare total sludge production without and with chemical addition.
   a. Without chemical addition

   Sludge = (3800 m^3/d)(220.0 mg/L)(0.6)(1 kg/10^3 g) = 501.6 kg/d

   b. Total with chemical addition

   Total = 501.6 kg/d + 385.3 kg/d = 886.9 kg/d

7. Determine the total volume of sludge without chemical precipitation, assuming that the sludge has a specific gravity of 1.03 and a moisture content of 94 percent.

$$V_s = \frac{(501.6 \text{ kg/d})}{(1.03)(10^3 \text{ kg/m}^3)(0.06)} = 8.1 \text{ m}^3\text{/d}$$

8. Determine the total volume of sludge resulting from chemical precipitation, assuming that the sludge has a specific gravity of 1.05 and a moisture content of 92.5 percent.

$$V_s = \frac{(886.9 \text{ kg/d})}{(1.05)(10^3 \text{ kg/m}^3)(0.075)} = 11.3 \text{ m}^3\text{/d}$$

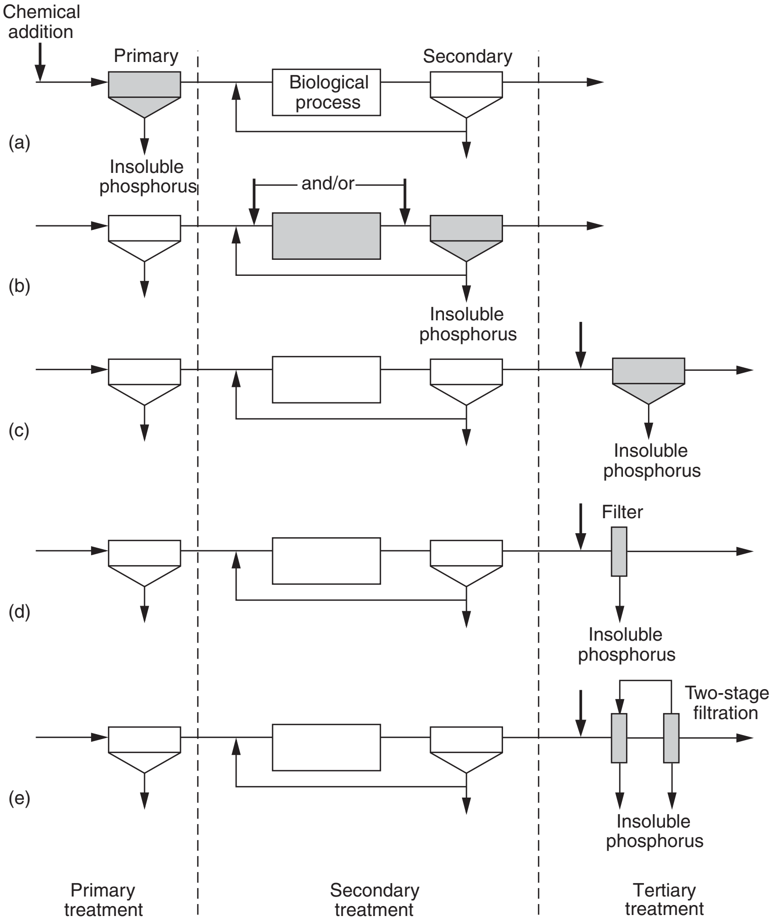
9. Prepare a summary table of sludge masses and volumes without and with chemical precipitation.

	Sludge	
Treatment	Mass, kg/d	Volume, m^3/d
Without chemical precipitation	501.6	8.1
With chemical precipitation	886.9	11.3

Comment The additional BOD and TSS removals afforded by chemical addition to primary treatment may also solve overloading problems on downstream biological systems, or may allow seasonal or year-round nitrification, depending on biological system designs. Alternatively, the total amount of ferric chloride could have been added at multiple locations.

Figure 6–14

Addition of chemicals at a single dosing point at various locations for phosphorus removal: (a) before primary sedimentation, (b) before and/or following biological treatment, (c) following secondary treatment, (d) chemical addition prior to single-stage filtration, and (e) chemical addition prior to dual-stage filtration.



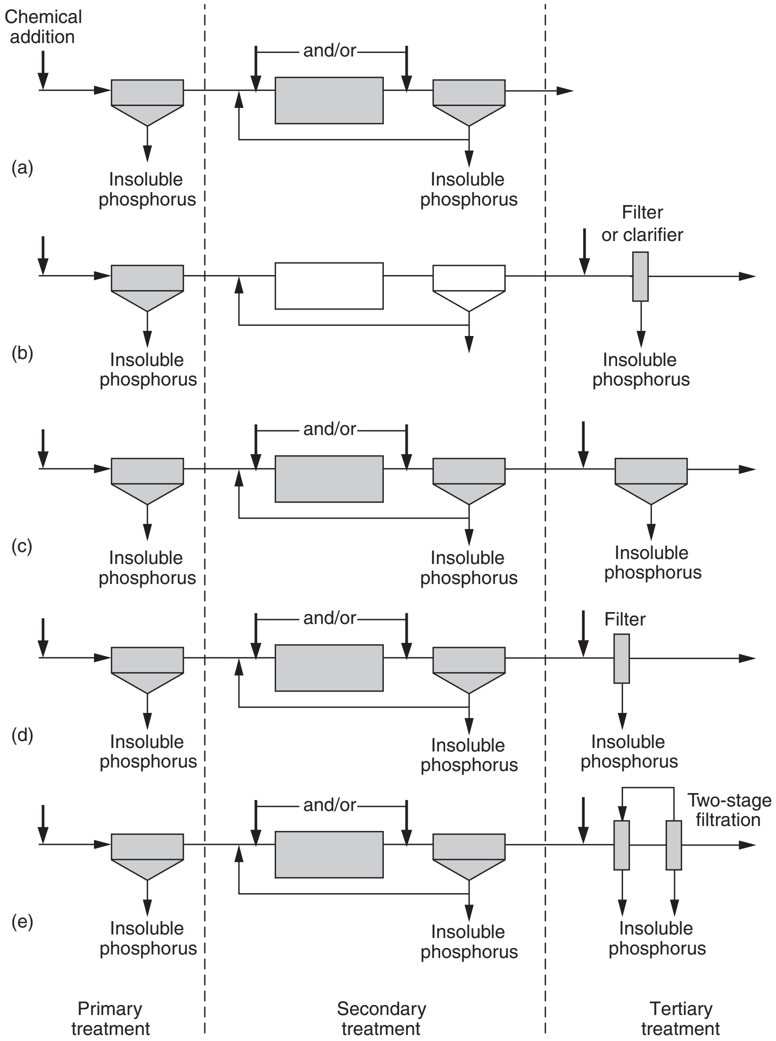
Phosphorus Removal from the Liquid Stream with Metal Salts

Phosphorus is removed from the liquid stream, as noted above, through a combination of precipitation, adsorption, exchange, and agglomeration followed by collection and removal with the process sludge. Metal salt, as illustrated on Figs. 6–14 and 6–15, can be added at a variety of different points in the treatment process, but because polyphosphates and organic phosphorus are less easily removed than orthophosphorus, adding aluminum or iron salts after secondary treatment (where organic phosphorus and polyphosphorus are transformed into orthophosphorus) may be necessary for the most complete removal of residual phosphorus. A number of the important features of adding metal salts and polymers at different points in the treatment process are discussed in this section.

Metal Salt Addition to Primary Sedimentation Tanks. When aluminum or iron salts are added to untreated wastewater, they react with the soluble orthophosphate to produce a precipitate. Organic phosphorus and polyphosphate are also removed partially by more complex reactions and by adsorption onto floc particles. The insolubilized

Figure 6–15

Addition of chemicals at multiple dosing points in a treatment process for phosphorus removal: (a) before primary and before and/or after biological treatment; (b) before primary and before tertiary treatment; (c), (d) and (e) at several locations in a treatment process (known as "split treatment").



phosphorus, as well as considerable quantities of BOD and TSS, are removed from the system as primary sludge. Adequate initial mixing and flocculation are necessary upstream of primary facilities, whether separate basins are provided or existing facilities are modified to provide these functions (Lijklema, 1980). Polymer addition may be required to aid in settling. In low alkalinity waters, the addition of a base is sometimes necessary to keep pH in the 5 to 7 range.

Metal Salt Addition to Secondary Treatment. Metal salts can be added in the influent to the activated sludge process, the activated sludge aeration tank, or the final clarifier influent channel. In trickling filter systems, the salts are added to the primary effluent or the trickling filter effluent which is to be recirculated. Multipoint additions have also been used. Theoretically, the minimum solubility of $AlPO_4$ occurs at about pH 6.3, and that of $FePO_4$ occurs at about pH 5.3; however, practical applications have yielded good phosphorus removal anywhere in the range of pH 6.5 to 7.0, which is compatible with most biological treatment processes. The use of ferrous salts is limited because they produce low phosphorus levels only at high pH values. In low alkalinity waters, either sodium aluminate and alum or ferric plus lime, or both, can be used to maintain the pH higher than 5.5. Improved settling and lower effluent BOD result from chemical addition, particularly if polymer is also added to the final clarifier.

Metal Salt Addition to Secondary Clarifiers. In certain cases, such as trickling filtration and extended aeration activated sludge processes, solids may not flocculate and settle well in the secondary clarifier. This settling problem may become acute in plants that are overloaded. The addition of aluminum or iron salts will cause the precipitation of metallic hydroxides or phosphates, or both. Aluminum and iron salts, along with certain organic polymers, can also be used to coagulate colloidal particles and to improve removals on filters. The resultant coagulated colloids and precipitates will settle readily in the secondary clarifier, reducing the TSS in the effluent and effecting phosphorus removal.

Metal Salt Addition to Effluent Filtration. Depending on the quality of the settled secondary effluent, chemical addition has been used to improve the performance of effluent filters. Chemical addition has also been used to achieve specific treatment objectives including the removal of specific contaminants such as phosphorus, metal ions, and humic substances. The removal of phosphorus by chemical addition to the contact filtration process is used in many parts of the country to remove phosphorus from wastewater treatment plant effluents which are discharged to sensitive water bodies. A two-stage filtration process (see Table 11–10 and discussion in Chap. 11) has also proven to be effective for the removal of phosphorus.

Phosphorus Removal from the Liquid Stream with Calcium

The use of calcium (lime) for phosphorus removal is declining because of (1) the substantial increase in the mass of sludge to be handled compared to metal salts and (2) the operation and maintenance problems associated with the handling, storage, and feeding of lime. When lime is used, the principal variables controlling the dosage are the degree of removal required and the alkalinity of the wastewater. The operating dosage is usually determined by on-site testing. Lime has been used customarily either as a precipitant in the primary sedimentation tanks or following secondary clarification.

Lime Addition to Primary Sedimentation Tanks. Both low and high lime treatment can be used to precipitate a portion of the phosphorus (usually about 65 to 80 percent). When lime is used, both the calcium and the hydroxide react with the orthophosphorus to form an insoluble hydroxyapatite [$Ca_5(OH)(PO_4)_3$]. A residual phosphorus level of 1.0 mg/L can be achieved with the addition of effluent filtration facilities to which chemicals can be added. In the high lime system, sufficient lime is added to raise the pH to about 11. After precipitation, the effluent must be recarbonated before biological treatment. In activated sludge systems, the pH of the primary effluent should not exceed 9.5 or 10; higher pH values can result in biological process upsets. In the trickling filter process, the carbon dioxide generated during treatment is usually sufficient to lower the

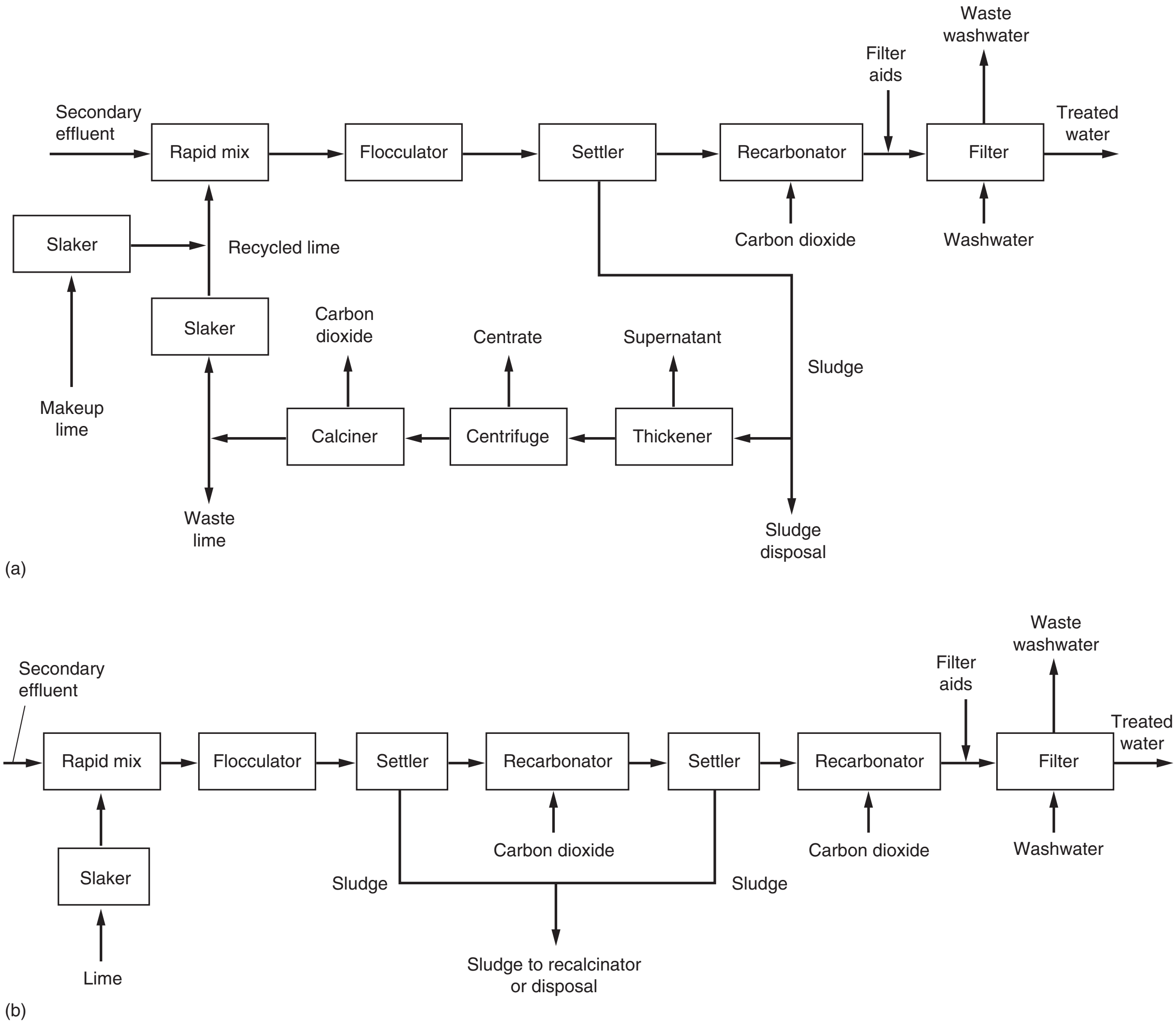


Figure 6–16

Typical lime treatment process flow diagrams for phosphorus removal: (a) single-stage system and (b) two-stage system.

pH without recarbonation. The dosage for low lime treatment is usually in the range of 75 to 250 mg/L as $Ca(OH)_2$ at pH values of 8.5 to 9.5. In low lime systems, however, the conditions required for precipitation are more specialized; the Ca^{2+}/Mg^{2+} mole ratio is $\leq 5/1$ (Sedlak, 1991).

Lime Addition Following Secondary Treatment. Lime can be added to the waste stream after biological treatment to reduce the level of phosphorus and TSS. Single-stage process and two-stage process flow diagrams for lime addition are shown on Fig. 6–16. On Fig. 6–16(a), a single-stage lime precipitation process is used for the treatment of secondary effluent. In the first stage clarifier of the

Figure 6–17

Large reactor clarifier used for the lime precipitation of secondary effluent. The settled effluent is treated further by microfiltration before reuse in industrial applications.

two-stage process shown on Fig. 6–16(b), sufficient lime is added to raise the pH above 11 to precipitate the soluble phosphorus as basic calcium phosphate (apatite). The calcium carbonate precipitate formed in the process acts as a coagulant for TSS removal. An example of a large lime precipitation unit is shown on Fig. 6–17. The excess soluble calcium is removed in the second stage clarifier as a calcium carbonate precipitate by adding carbon dioxide gas to reduce the pH to about 10. Generally, there is a second injection of carbon dioxide to the second stage effluent to reduce the formation of scale. To remove the residual levels of TSS and phosphorus, the secondary clarifier effluent is passed through a multimedia filter or a membrane filter. Care should be taken to limit excess calcium in the filter feed to ensure cementing of the filter media will not occur.

Lime Recalcination. Although lime recalcination lowers chemical costs, it is a feasible alternative only for large plants. Where a lime recovery system is required for a cost-effective operation, it includes a thermal regeneration facility, which converts the calcium carbonate in the sludge to lime by heating to 980°C (1800°F). The carbon dioxide from this process or other on-site stack gas (containing 10 to 15 percent carbon dioxide) is generally used as the source of recarbonation for pH adjustment of the wastewater.

Strategies for Chemical Phosphorus Removal

Depending on the characteristics of the wastewater and treatment facilities a number different operating strategies can be employed for the chemical removal of phosphorus. In general, the strategies can be divided into categories: single point addition or multiple points of addition. The advantages and disadvantages of each of these approaches for the removal of phosphorus are summarized in Table 6–6. It is recommended that each alternative point of application be evaluated carefully.

Table 6–6

Single and multiple chemical dosing strategies for the removal of phosphorus from the liquid stream with metal salts[a]

Point of application	Advantages	Disadvantages
Single point of application (see Fig. 6–14)		
Primary	Applicable to most plants; increased BOD and suspended solids removal; can achieve good removal of influent phosphate at Me:P ratios of 0.5 to 1.	Polymer may be required for flocculation; sludge more difficult to dewater than primary sludge; must leave residual P for biological process.
Secondary bioreactor	Improved stability of activated sludge; polymer not required.	Overdose of metal may cause low pH toxicity; with low alkalinity wastewaters, a pH control system may be necessary; cannot use lime because of excessive pH; inert solids added to activated sludge mixed liquor, reducing the percentage of volatile solids.
Secondary clarifier	Improved settling in secondary clarifier.	Metal carry over may impact disinfection with UV.
Tertiary clarifier	Low effluent phosphorus achievable at high Me:P ratios.	High capital cost.
Tertiary filter	Low cost; can be combined with the removal of residual suspended solids.	Length of filter run may be reduced with single-stage filtration. Additional expense with two-stage filtration process.
Multiple points of addition (see Fig. 6–15)		
Two locations (e.g., raw wastewater and secondary sedimentation or settled secondary effluent before filtration	Enhanced control over phosphorus concentration in effluent. Multiple dosing method is most suitable for locations with restrictive effluent discharge limits.	High capital cost, as two separate chemical dosing stations are required. Metal carry over may impact disinfection with UV.
Three or more locations.	As above.	As above.

[a] Refer also to Table 6–5. Adapted in part from U.S. EPA (1976).

6–5 CHEMICAL FORMATION OF STRUVITE FOR AMMONIUM AND PHOSPHORUS REMOVAL

The removal of phosphorus from wastewater by precipitation was considered in the previous section. The combined removal of ammonium and phosphorus is considered in this section. In wastewater treatment, one of the more serious problems in the processing of primary sludge and waste activated sludge in anaerobic digesters is the formation of magnesium ammonium phosphate hexahydrate, $MgNH_4PO_4 \cdot 6H_2O$, commonly known as struvite. Other precipitates that occur in anaerobic digesters include calcium phosphate [$Ca(PO_4)_2 \cdot nH_2O$], vivianite [$Fe_3(PO_4)_2 \cdot 8H_2O$], and variscite ($AlPO_4 \cdot 2H_2O$).

Figure 6–18

Image of struvite precipitation in wastewater piping.

The formation and accumulation of struvite crystals and vivianite can cause problems in process pipelines, pumps, and dewatering facilities (see Fig. 6–18). The formation of struvite, the means to control the formation of struvite, and the recovery of nitrogen and phosphorus using the basic principles of struvite formation are considered in this section. The practical application of struvite formation for nutrient recovery is considered in Chap. 15.

Chemistry of Struvite Formation

During anaerobic digestion, magnesium, ammonium, and phosphate are released from the digestion of primary and waste activated sludge. If the concentrations of soluble magnesium, ammonium, and orthophosphate exceed the solubility limit for the formation of struvite at a given pH, crystals of struvite will form. The formation of struvite can be described by the following general reaction.

$$Mg^{2+} + NH_4^+ + PO_4^{3-} + 6H_2O \rightleftarrows MgNH_4PO_4 \cdot 6H_2O \tag{6–25}$$

Although the struvite reaction is deceptively simple, the process is complex depending on the ionic strength, pH, alkalinity, and temperature of the waste stream. Once initiated, the growth of struvite crystals will continue to form as long as favorable conditions exist including the presence of the three constituents in a molar ratio of 1 : 1 : 1 of Mg^{2+} : NH_4^+ : PO_4^{3-}. The corresponding solubility product constant for struvite is given by the following equation,

$$\{Mg^{2+}\}\{NH_4^+\}\{PO_4^{3-}\} = K_{so} \quad \text{(struvite)} \tag{6–26}$$

where the term within the ellipses corresponds to the ion activity concentration of the constituent.

The principal side reactions involved in the precipitation of struvite are summarized in Table 6–7. The conditional solubility product, P_s, used to account for side reactions involving the constituents, ion activity, and ionic strength is given by the following expression.

Table 6–7

Chemical reactions involved in struvite chemistry and expressions for the total concentration of magnesium, ammonium, and phosphate in solution.

Reaction	p*K* Range	p*K* Typical
$NH_4^+ \rightleftarrows NH_{3(aq)} + H^+$	9.25–9.3	9.25
$H_3PO_4 \rightleftarrows H_2PO_4^- + H^+$	2.1	2.1
$H_2PO_4^- \rightleftarrows H_2PO_4^{2-} + H^+$	7.2	7.2
$HPO_4^{2-} \rightleftarrows PO_4^{3-} + H^+$	12.3	12.3
$MgOH^+ \rightleftarrows Mg^{2+} + OH^-$	2.56	2.56
$MgH_2PO_4^+ \rightleftarrows H_2PO_4^- + Mg^{2+} + OH^-$	0.45	0.45
$MgHPO_4 \rightleftarrows H_2PO_4^{2-} + Mg^{2+}$	2.91	2.91
$MgPO_4^- \rightleftarrows PO_4^{3-} + Mg^{2+}$	4.8	4.8
$MgNH_4PO_4{\cdot}6H_2O \rightleftarrows Mg^{2+} + NH_4^+ + PO_4^{2+} + Mg^{2+} + H_2O$	12.6–13.26	13.0
$AlPO_{4(s)} \rightleftarrows Al^{3+} + PO_4^{3-}$	21	21
$FePO_{4(s)} \rightleftarrows Fe^{3+} + PO_4^{3-}$	21.9–23	22.0

$C_{T,Mg} = [Mg^{2+}] + [MgOH^+] + [MgH_2PO_4^+] + [MgHPO_4] + [MgPO_4^-]$

$C_{T,NH_3} = [NH_4^+] + [NH_3]$

$C_{T,P} = [PO_4^{3-}] + [H_3PO_4] + [H_2PO_4^-] + [HPO_4^{2-}] + [MgH_2PO_4^+] + [MgHPO_4] + [MgPO_4^-]$

$$P_s = C_{T,\,Mg}C_{T,\,NH_3}C_{T,\,PO_4} = \frac{K_{so}}{\alpha_{Mg^{2+}}\alpha_{NH_4^+}\alpha_{PO_4^{3-}}\gamma_{Mg^{2+}}\gamma_{NH_4^+}\gamma_{PO_4^{3-}}} \tag{6–27}$$

where C_{Mg}, C_{NH_3}, C_{PO_4} = total analytical concentration of individual constituents

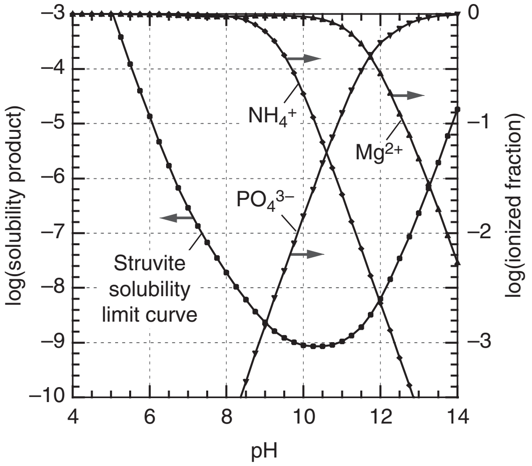
$\alpha_{Mg^{2+}}$, $\alpha_{NH_4^+}$, $\alpha_{PO_4^{3-}}$ = ionization fraction of the individual constituents

$\gamma_{Mg^{2+}}$ $\gamma_{NH_4^+}$ $\gamma_{PO_4^{3-}}$ = ionic strength of the individual constituents

The ionization fraction is defined as the ratio of the free constituent concentration of magnesium, ammonium, and phosphate over the total concentration in solution (e.g., $\alpha = [Mg^{2+}]/C_{T,\,Mg}$). Expressions for the total concentration for magnesium, ammonium, and phosphate in solution are also given in Table 6–7. Using Eq. (6–27), the equations and corresponding pK values given in Table 6–7 for the various equilibrium relationships, the minimum solubility for struvite, as illustrated in Fig. 6–19, occurs at a pH of about 10.3 (Ohlinger et al., 1998).

Figure 6–19

Plot of struvite solubility limit curve and ionization fraction developed using MINTEQA2 with an ionic strength value set constant at μ = 0.1 (Ohlinger et al., 1998).



The struvite solubility limit curve shown on Fig. 6–19 can be used to determine if struvite formation is likely to occur. The P_s value is calculated from the analytical concentrations of magnesium, ammonium, and phosphate using the left side of Eq. 6–27, and then plotted along with the solubility limit curve corresponding to the relevant ionic strength. If the P_s values fall outside of the area bounded within the curve, then struvite precipitation is not expected to occur.

The chemical system composed of the species and equilibrium constants presented in Table 6–7 can be modeled using chemical speciation software such as MINEQL+ or MINTEQA2. The model output, consisting of constituent concentration data, can then be used to compute P_s with the right side of Eq. 6–27. The MINTEQA2 software package, used for the generation of Fig. 6–19, was developed by U.S. EPA and is available for download at http://www.epa.gov/ceampubl/mmedia/minteq/. Details on the development and use of the solubility product equation can be found in Snoeyink and Jenkins (1980) and Ohlinger et al. (1998). Application of the curve given in Fig. 6–19 is illustrated in Example 6–3.

EXAMPLE 6–3 **Assessment of Potential for Struvite Precipitation** The following analytical test results were reported by Snoeyink and Jenkins (1980) for sludge obtained from the Hyperion wastewater treatment plant in Los Angeles, CA. Using the plot of struvite solubility given on Fig. 6–18, determine if struvite precipitation is likely to occur.

		Constituent concentration, mole/L[a]		
Sample	pH	$C_{T,Mg}$	$C_{T,NH3}$	$C_{T,PO4}$
Raw sludge	5.5	0.005	0.005	0.04
Digested sludge	7.5	0.005	0.1	0.07
Diluted digested sludge	7.5	0.001	0.025	0.02

[a] Data adapted from Snoeyink and Jenkins (1980).

Solution

1. Compute the conditional solubility product as the product of the analytical concentrations for each of the sludge samples using the left side of Eq. 6–27:

$$P_s = (C_{T,\,Mg})(C_{T,\,NH_3})(C_{T,\,PO_4})$$

a. Raw sludge

$$P_s = (0.005)(0.005)(0.04) = 0.000001$$

$$\log(P_s) = \log(0.000001) = -6$$

b. Digested sludge

$$P_s = (0.005)(0.1)(0.07) = 0.000035$$

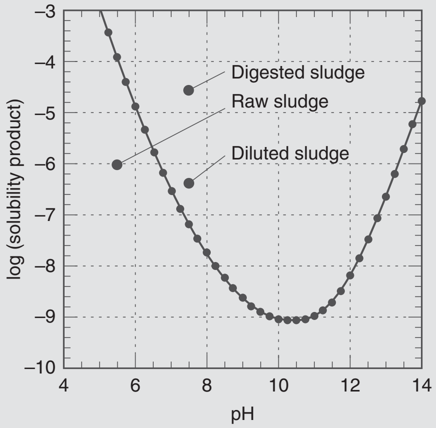
$$\log(P_s) = \log(0.000035) = -4.5$$

c. Diluted digested sludge

$$P_s = (0.001)(0.025)(0.02) = 0.0000005$$

$$\log(P_s) = \log(0.000001) = -6.3$$

2. Plot the conditional solubility product, log(P_s), values as a function of pH on the solubility limit curve to determine if struvite precipitation is expected to occur. The struvite solubility limit curve from Step 1 is shown below with the data computed in Step 2.



Comment As shown in the plot presented in Step 2, the raw sludge sample is not within the region where struvite precipitation would be expected. However, both the as-received and diluted digested sludge are both subject to potential struvite precipitation. Further dilution of the digested sludge or reduction in the sludge pH may be recommended to inhibit struvite formation.

Control and/or Mitigation Measures for the Formation of Struvite

The control of struvite is not only important from the standpoint of its accumulation in process equipment, it is also important because of the beneficial effects on wastewater treatment that accrue as a result of the elimination of phosphorus and ammonium in the return flows to the treatment process. The formation of struvite can be controlled by chemical addition and/or harvesting and removal. The chemical methods that have been used over the years are considered below. Struvite harvesting is considered following the discussion of chemical methods. The principal methods of struvite control are summarized in Table 6–8. As reported in Table 6–8, the principal chemical methods involve the use of alum and iron salts, lime, and proprietary chemicals. In the precipitation methods, the fundamental concept is to reduce one or more of the constituents involved in the formation of struvite below the conditional solubility product and, thus, limit the formation of struvite. Other chemical means include the use of proprietary chemicals to suppress the formation of struvite and the addition of chemicals to lower the pH and, thus, limit the formation of struvite. Additional details of the formation and control of struvite are discussed in Chap. 15.

Enhanced Struvite Formation for Nutrient Removal

From a review of Eq. (6–27) and the equilibrium expressions given in Table 6–7, if the conditions controlling the formation of struvite were optimized, it should be possible to

Table 6–8

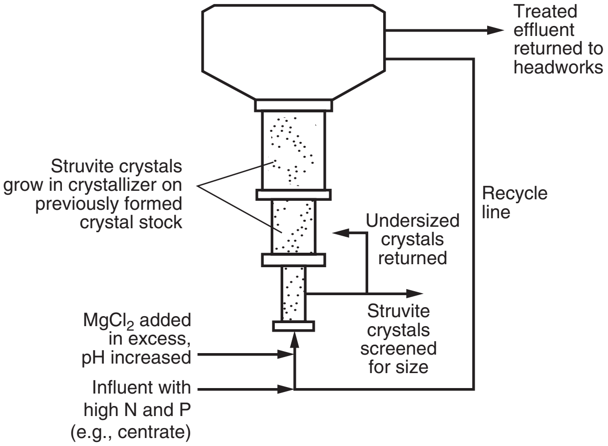
Methods used to control and/or limit the formation of struvite.

Method	Description
Alum [$Al_2(SO_4)_3 \cdot 14H_2O$] addition	Used to precipitate phosphorus. Because alum addition depletes alkalinity, lime is used to add the necessary alkalinity. Can be added upstream of anaerobic digestion, to anaerobic digester, and downstream of anaerobic digestion.
Ferric chloride ($FeCl_3$) addition	Used to precipitate phosphorus. Because ferric addition depletes alkalinity, lime is used to add the necessary alkalinity. Can be added upstream of anaerobic digestion, to anaerobic digester, and downstream of anaerobic digestion.
Ferrous chloride ($FeCl_2$) addition	Used to precipitate phosphorus. Can be added upstream of anaerobic digestion, to anaerobic digester, and downstream of anaerobic digestion.
Ferric sulfate[$Fe_2(SO_4)_3$] addition	Used to precipitate phosphorus. Generally not recommended because of the potential to form hydrogen sulfide during the digestion process.
Lime [$Ca(OH)_2$] addition	Used to precipitate phosphorus and to adjust the alkalinity.
Addition of proprietary chemicals	Propriety chemicals are also available that suppress the formation of struvite. The fate of the phosphorus is unknown.
Addition of anti-scalent chemicals	Propriety anti-scalent chemicals are available that suppress the formation of struvite.
pH reduction	Acid or carbon dioxide (CO_2) can be added to reduce the pH of the waste stream. Carbon dioxide is favored over acid addition for safety reasons and because it is difficult to overdose.
Minimization of carbon dioxide release	Carbon dioxide can be released at points of turbulence in the process piping. When carbon dioxide is released, the pH goes up and struvite precipitation can occur.
Routine maintenance	Depending on where struvite forms, routine flushing of process lines with high-pressure water jets has proven effective in controlling the formation of struvite.

enhance the removal of ammonium and phosphate. Worldwide there are more than 10 technologies in various states of development for the removal of nutrients in the form of struvite. The method used most commonly for the removal of ammonium and phosphate involves the formation of recoverable struvite crystals in an upflow fluidized bed reactor (see Fig. 6–20). As shown on Fig. 6–20, the formation of struvite crystals is accelerated by the addition of an excess amount of Mg^{2+} which is added to the waste stream at an elevated pH and temperature. Other technologies employ complete-mix reactors in a variety of configurations. Based on numerous literature citations, optimum conditions for the removal of phosphorus (and ammonium) occur when the molar ratio Mg^{2+}: PO_4^{3-} is between 1.1 and 1.6, the pH is between 9 and 10.5, and the temperature is about 25°C. The specific values will depend on the chemical composition and ionic strength of the wastewater streams and local conditions. The removal of nutrients is considered further in Chap. 15.

Figure 6–20

Reactor used for struvite precipitation.



6–6 CHEMICAL PRECIPITATION FOR REMOVAL OF HEAVY METALS AND DISSOLVED SUBSTANCES

The technologies available for the removal of heavy metals from wastewater include chemical precipitation, carbon adsorption, ion exchange, and reverse osmosis. Of these technologies, chemical precipitation is most commonly employed for most of the metals. Common precipitants include hydroxide (OH) and sulfide (S^{2-}). Carbonate (CO_3^{2-}) has also been used in some special cases. Metals may be removed separately or co-precipitated with phosphorus. Adsorption, ion exchange, and RO are discussed in Chap. 11

Precipitation Reactions

Metals of interest include arsenic (As), barium (Ba), cadium (Cd), copper (Cu), mercury (Hg), nickel (Ni), selenium (Se), and zinc (Zn). Most of these metals can be precipitated as hydroxides or sulfides. Solubility products for free metal concentrations in equilibrium with hydroxide and sulfide precipitates are reported in Table 6–9.

In wastewater treatment facilities, metals are precipitated most commonly as metal hydroxides through the addition of lime or caustic to a pH of minimum solubility. However, several of these compounds are, as discussed previously, amphoteric (i.e., capable of either accepting or donating a proton) and exhibit a point of minimum solubility. The pH value at minimum solubility varies with the metal in question as illustrated on Fig. 6–21 for hydroxide precipitation. The solid line shown on Fig. 6–21 represents the total metal in solution in equilibrium with the precipitate. The curves were developed based on the mononuclear hydroxide species using the same procedures as illustrated on Fig. 6–8 for Al^{3+} and Fe^{3+}. It is important to remember that the location of the minimum solubility will vary depending on the constituents in the wastewater. The curves given on Fig. 6–21 are useful in establishing the pH ranges for testing.

Metals can also be precipitated as sulfides as illustrated on Fig. 6–22. The minimum effluent concentration levels that can be achieved in the chemical precipitation of heavy metals are reported in Table 6–10. In practice, the minimum achievable residual metal concentrations will also depend on the nature and concentration of the organic matter in the wastewater as well as the temperature. It should also be noted that some metals, such

Table 6–9

Solubility products for free metal ion concentrations in equilibrium with hydroxides and sulfides [a,b]

Metal species	Half reaction	pK_{sp}
Cadmium hydroxide	$Cd(OH)_2 \rightleftarrows Cd^{2+} + 2OH^-$	13.93
Cadmium sulfide	$CdS \rightleftarrows Cd^{2+} + S^{2-}$	28
Chromium hydroxide	$Cr(OH)_3 \rightleftarrows Cr^{3+} + 3OH^-$	30.2
Copper hydroxide	$Cu(OH)_2 \rightleftarrows Cu^{2+} + 2OH^-$	19.66
Copper sulfide	$CuS \rightleftarrows Cu^{2+} + S^{2-}$	35.2
Iron (II) hydroxide	$Fe(OH)_2 \rightleftarrows Fe^{2+} + 2OH^-$	14.66
Iron (II) sulfide	$FeS \rightleftarrows Fe^{2+} + S^{2-}$	17.2
Lead hydroxide	$Pb(OH)_2 \rightleftarrows Pb^{2+} + 2OH^-$	14.93
Lead sulfide	$PbS \rightleftarrows Pb^{2+} + S^{2-}$	28.15
Mercury hydroxide	$Hg(OH)_2 \rightleftarrows Hg^{2+} + 2OH^-$	23
Mercury sulfide	$HgS \rightleftarrows Hg^{2+} + S^{2-}$	52
Nickel hydroxide	$Ni\ (OH)_2 \rightleftarrows Ni^{2+} + 2OH^-$	15
Nickel sulfide	$NiS \rightleftarrows Ni^{2+} + S^{2-}$	24
Silver hydroxide	$AgOH \rightleftarrows Ag^+ + OH^-$	14.93
Silver sulfide	$Ag_2S \rightleftarrows 2Ag^+ + S^{2-}$	28.15
Zinc hydroxide	$Zn(OH)_2 \rightleftarrows Zn^{2+} + 2OH^-$	16.7
Zinc sulfide	$ZnS \rightleftarrows Zn^{2+} + S^{2-}$	22.8

[a] Adapted from Bard (1966).

[b] To obtain the complete solubility of a metal, all of the complex species must be considered such as reported in Table 6–2 for aluminum and iron.

Figure 6–21

Residual soluble metal concentration as a function of pH for the precipitation of metals as hydroxides. Because of the wide variation in the solubility and formation constants for the various metal hydroxides, the curves presented in this figure should only be used as a reference guide (see also Table 6–9).

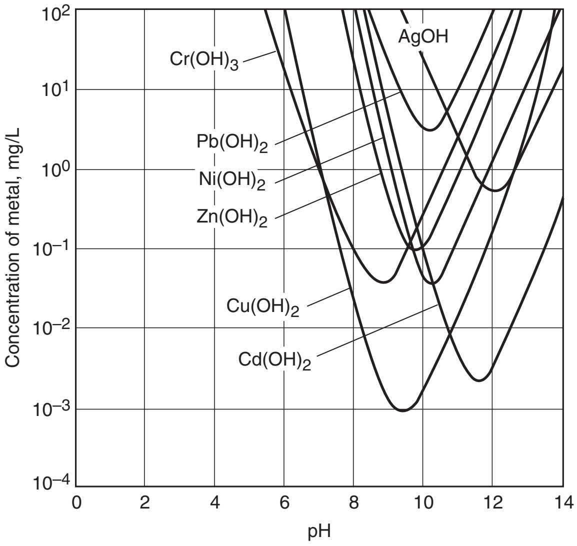
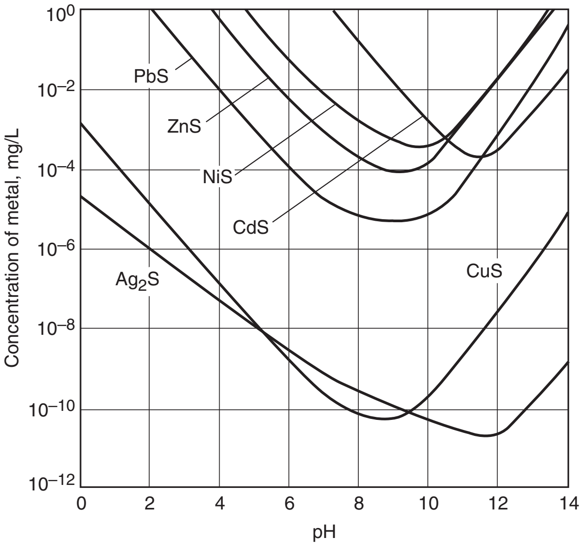


Figure 6–22

Residual soluble metal concentration as a function of pH for the precipitation of metals as sulfides. Because of the wide variation in the solubility and formation constants for the various metal sulfides, the curves presented in this figure should only be used as a reference guide (see also Table 6–9).



as Hg and As, have multiple oxidation states that can impact the efficiency of precipitation. Because of the many uncertainties associated with the precipitation of metals, laboratory bench scale or pilot plant testing should be conducted.

Co-precipitation with Phosphorus

As discussed previously, precipitation of phosphorus in wastewater is usually accomplished by the addition of coagulants, such as alum, iron, or calcium. Coincidentally with

Table 6–10

Practical effluent concentration levels achievable in heavy metals removal by precipitation[a]

Metal	Achievable effluent concentration, mg/L	Type of precipitation and technology
Arsenic	0.05	Sulfide precipitation with filtration
	0.005	Ferric hydroxide co-precipitation
Barium	0.5	Sulfide precipitation
Cadmium	0.05	Hydroxide precipitation at pH 10–11
	0.05	Co-precipitation with ferric hydroxide
	0.008	Sulfide precipitation
Copper	0.02–0.07	Hydroxide precipitation
	0.01–0.02	Sulfide precipitation
Mercury	0.01–0.02	Sulfide precipitation
	0.001–0.01	Alum co-precipitation
	0.0005–0.005	Ferric hydroxide co-precipitation
	0.001–0.005	Ion exchange
Nickel	0.12	Hydroxide precipitation at pH 10
Selenium	0.05	Sulfide precipitation
Zinc	0.1	Hydroxide precipitation at pH 11

[a] From Eckenfelder et al. (2009).

the addition of these chemicals for the removal of phosphorus, removal of various inorganic ions, principally some of the heavy metals, occurs as co-precipitation. Dissolved metals may be adsorbed onto the hydroxide complex, while particulate and colloidal forms may be incorporated into flocculated material. Where both industrial and domestic wastes are treated together, it may be necessary to add chemicals to the primary settling facilities, especially if on-site pretreatment measures prove to be ineffective. When chemical precipitation is used, anaerobic digestion for sludge stabilization may not be possible because of the toxicity of the precipitated heavy metals. As noted previously, one of the disadvantages of chemical precipitation is that it usually results in a net increase in the total dissolved solids of the wastewater that is being treated.

6–7 CONVENTIONAL CHEMICAL OXIDATION

Conventional chemical oxidation in wastewater treatment typically involves the use of oxidizing agents such as ozone (O_3), hydrogen peroxide (H_2O_2), permanganate (MnO_4), chlorine dioxide (ClO_2), chlorine (Cl_2 or HOCl), and oxygen (O_2), to bring about change in the chemical composition of a compound or a group of compounds. Advanced oxidation processes (AOPs) in which the free hydroxyl radical (HO•) is used as a strong oxidant to destroy specific organic constituents and compounds that cannot be oxidized easily by conventional oxidants are discussed in Sec. 6–8.

Applications for Conventional Chemical Oxidation

The principal applications of conventional chemical oxidation in wastewater treatment are for (1) odor control, (2) hydrogen sulfide control, (3) color removal, (4) iron and manganese removal, (5) disinfection, (6) control of biofilm growth and biofouling in treatment processes and distribution system components, and (7) oxidation of selected trace organic constituents. Some of the more important applications of chemical oxidation in wastewater management are summarized in Table 6–11. Chemical oxidation is especially effective for the elimination of odorous compounds (e.g., oxidation of sulfides and mercaptans) and is covered further in Chap. 16. Because of its importance, chemical disinfection is considered separately in Chap. 12.

In addition to the applications reported in Table 6–11, chemical oxidation is now commonly used to (1) improve the treatability of nonbiodegradable (refractory) organic compounds, (2) eliminate the inhibitory effects of certain organic and inorganic compounds to microbial growth, and (3) reduce or eliminate the toxicity of certain organic and inorganic compounds to microbial growth and aquatic flora. Additional details on the applications of conventional oxidation may be found in Rakness (2005), Crittenden et al. (2012), U.S. EPA (1999), and Black and Veatch Corporation (2010).

Oxidants Used in Chemical Oxidation Processes

Oxidants that are used frequently in wastewater treatment include (1) chlorine, (2) ozone, (3) chlorine dioxide, (4) permanganate, and (5) hydrogen peroxide. The oxidation kinetics of oxygen are usually too slow to be of practical use beyond secondary biological treatment. Chemical oxidants are usually added at specific points during treatment (e.g., to control odors or membrane fouling) or at the final stage of treatment before discharge or reuse (e.g., disinfection). The rate of oxidation typically follows the trend given below; however, there will be exceptions depending on the characteristics of the solution (e.g., pH) and type of compound that is to be oxidized.

$$HO\cdot > O_3 > H_2O_2 > HOCl > ClO_2 > MnO_4^- > O_2 > OCl^- \tag{6–28}$$

Table 6–11

Typical applications of chemical oxidation in wastewater collection, treatment, and dispersal

Application	Chemicals used[a]	Remarks
Collection		
Slime-growth control	Cl_2, H_2O_2	Control of fungi and slime-producing bacteria
Corrosion control (H_2S)	Cl_2, H_2O_2, O_3	Control brought about by oxidation of H_2S
Odor control	Cl_2, H_2O_2, O_3	Especially in pumping stations and long, flat sewers
Treatment		
Grease removal	Cl_2	Added before preaeration
BOD reduction	Cl_2, O_3	Oxidation of organic substances
Ferrous sulfate oxidation	Cl_2	Production of ferric sulfate and ferric chloride
Filter-ponding control	Cl_2	Maintaining residual at filter nozzles
Filter-fly control	Cl_2	Maintaining residual at filter nozzles during fly season
Sludge-bulking control	Cl_2, H_2O_2, O_3	Temporary control measure
Control of filamentous microorganisms	Cl_2	Dilute chlorine solution sprayed on foam caused by filamentous organisms
Digester supernatant oxidation	Cl_2	
Digester foaming control	Cl_2	
Ammonium oxidation	Cl_2	Conversion of ammonia to nitrogen gas
Odor control	Cl_2, H_2O_2, MnO_4, O_3	
Oxidation of refractory organic compounds	O_3	
Reuse/dispersal		
Bacterial reduction	Cl_2, ClO_2, H_2O_2, O_3	Plant effluent, overflows, and stormwater
Odor control	Cl_2, H_2O_2, MnO_4, O_3	
Color removal	Cl_2, H_2O_2, MnO_4, O_3	

[a] Cl_2 = chlorine, ClO_2 = chlorine dioxide, H_2O_2 = hydrogen peroxide, MnO_4 = permanganate, O_3 = ozone.

The behavior of the hydroxyl radical, HO•, discussed in detail in Sec. 6–8, is introduced briefly here as related to its formation from ozonation. Of the conventional chemical oxidants, ozonation is effective for the destruction of organic compounds either by direct reactions with O_3, or indirect reactions with HO•, as shown in Eq. (6–29).

$$O_3 \rightarrow \left[\begin{array}{l} \xrightarrow[O_3]{\text{Direct Pathway}} O_3 + R \rightarrow \text{Product 1} \\ \xrightarrow[\text{NOM}]{\text{Indirect Pathway}} HO\cdot + R \rightarrow \text{Product 2} \end{array} \right. \qquad (6\text{–}29)$$

The reaction of ozone with natural organic matter (NOM) to produce HO•, is among the most important mechanisms used to destroy target compounds (Elovitz and von Gunten,

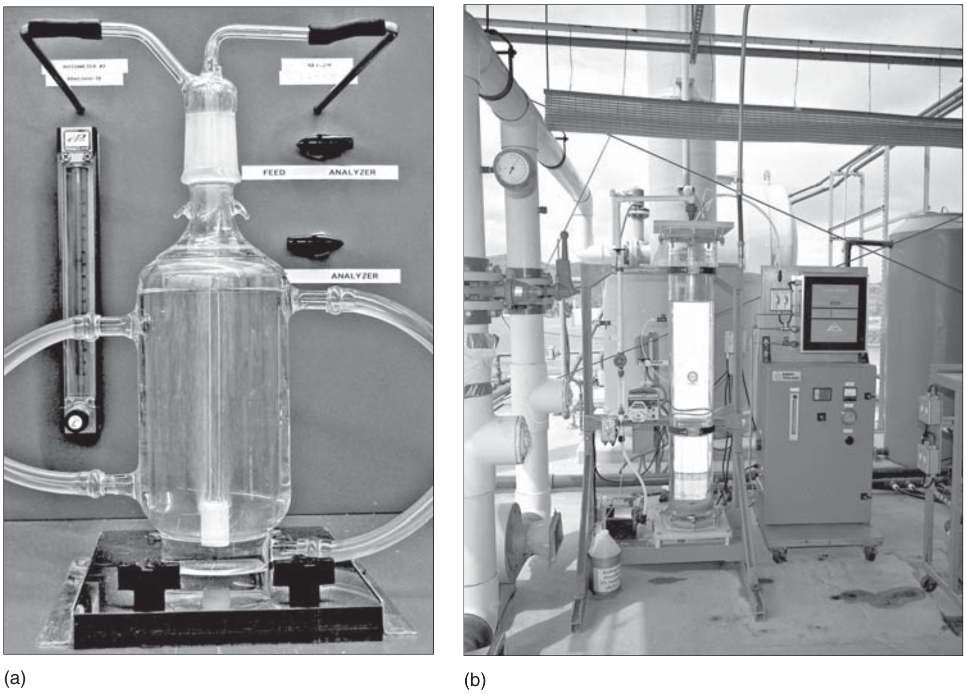


Figure 6–23

Contactors used for evaluation of chemical oxidation using ozone (a) bench scale and (b) pilot scale.

1999; Westerhoff et al., 1999). However, substantial removals of residual pharmaceuticals in low-DOC reclaimed water are possible by direct ozonation, even at low ozone dosages (Huber et al., 2005). Views of ozone contactors used for bench and pilot scale evaluation are shown on Fig. 6–23. Transformation of some pharmaceuticals is also possible during chlorine disinfection, however, the effectiveness will depend on the structure of the chemical compound, the form of chlorine, the contact time, and application of dechlorination (Pinkston and Sedlak, 2004).

Fundamentals of Chemical Oxidation

The purpose of the following discussion is to introduce the basic concepts involved in chemical oxidation reactions. The topics to be discussed include (1) oxidation-reduction reactions, (2) half reaction potentials, (3) reaction potentials, (4) equilibrium constants for redox equations, and (5) rate of oxidation-reduction reactions.

Oxidation-Reduction Reactions. Oxidation-reduction reactions (known as redox reactions) take place between an *oxidizing agent* and a *reducing agent*. In redox reactions electrons are exchanged and there is a change in the oxidation states of the constituents involved in the reaction. While an oxidizing agent causes the oxidation to occur, it is reduced in the process. Similarly, a reducing agent which causes a reduction to occur is oxidized in the process. For example, consider the following reaction.

$$Cu^{2+} + Zn \rightleftarrows Cu + Zn^{2+} \tag{6–30}$$

In the above reaction copper (Cu) changes from a +2 to zero oxidation state and the zinc (Zn) changes from a zero to a +2 state. Because of the electron gain or loss, oxidation-reduction reactions can be separated into two half reactions. The oxidation half reaction

Table 6–12

Chemicals that have been used for the disinfection of wastewater, along with their their oxidation half reaction and standard electrode potential[a]

Disinfectant	Half reaction	Oxidation potential[b], V
Ozone	$O_3 + 2H^+ + 2e^- \rightleftarrows O_2 + H_2O$	+2.07
Hydrogen peroxide	$H_2O_2 + 2H^+ + 2e^- \rightleftarrows 2H_2O$	+1.78
Permanganate	$MnO_4^- + 4H^+ + 3e^- \rightleftarrows MnO_2 + 2H_2O$	+1.67
Chlorine dioxide	$ClO_2 + e^- \rightleftarrows ClO_2^-$	+1.50
Hypochlorous acid	$HOCl + H^+ + 2e^- \rightleftarrows Cl^- + H_2O$	+1.49
Hypoiodous acid	$HIO + H^+ + e^- \rightleftarrows 1/2I_2 + H_2O$	+1.45
Chlorine gas	$Cl_2 + 2e^- \rightleftarrows 2Cl^-$	+1.36
Oxygen	$O_2 + 4H^+ + 4e^- \rightleftarrows 2H_2O$	+1.23
Bromine	$Br_2 + 2e^- \rightleftarrows 2Br^-$	+1.09
Hypochlorite	$ClO^- + H_2O + 2e^- \rightleftarrows Cl^- + 2OH^-$	+0.90
Chlorite	$ClO_2^- + 2H_2O + 4e^- \rightleftarrows Cl^- + 4OH^-$	+0.76
Iodine	$I_2 + 2e^- \rightleftarrows 2I^-$	+0.54

[a] Derived in part from Bard (1966) and Black and Veatch Corporation (2010).

[b] Reported values will vary depending on source. The electrical potential or voltage (V) is used commonly to characterize various oxidation reduction reactions.

involves the loss of electrons while the reduction half reaction involves the gain of electrons. The two half reactions that comprise Eq. (6–30) are as follows:

$$Zn - 2e^- \rightleftarrows Zn^{2+} \qquad \text{(oxidation)} \tag{6–31}$$

$$Cu^{2+} + 2e^- \rightleftarrows Cu \qquad \text{(reduction)} \tag{6–32}$$

Referring to the above equations there is a two electron change.

Half Reaction Potentials. Because of the almost infinite number of possible reactions, there are no summary tables of equilibrium constants for oxidation-reduction reactions. What is done instead is the chemical and thermodynamic characteristics of the half reactions, such as those given by Eqs. (6–31) and (6–32), are determined and tabulated so that any combination of reactions can be studied. Half reactions for disinfection processes are given in Table 6–12 and other representative half reactions are given in Table 6–13. Of the many properties that can be used to characterize oxidation-reduction reactions, the electrical potential (i.e., voltage) of the half reaction is used most commonly. Thus, every half reaction involving an oxidation or reduction has a *standard potential* $E°$ associated with it. The potentials for the half reactions given by Eqs. (6–33) and (6–34) are as follows:

$$Zn + 2e^- \rightleftarrows Zn^{2+} \qquad E° = -0.763 \text{ V} \tag{6–33}$$

$$Cu^{2+} + 2e^- \rightleftarrows Cu \qquad E° = 0.340 \text{ V} \tag{6–34}$$

The potentials for a number of half reactions are given in Table 6–13. The half reaction potential is a measure of the tendency of a reaction to proceed to the right. Half reactions with large positive potential, $E°$, tend to proceed to the right. Conversely, half reactions with large negative potential, $E°$, tend to proceed to the left.

Table 6–13

Selected standard electrode potentials for oxidation-reduction half reactions[a]

Half reaction	Oxidation potential[b], V
$Li^+ + e^- \rightarrow Li$	−3.03
$K^+ + e^- \rightarrow K$	−2.92
$Ba^{2+} + 2e^- \rightarrow Ba$	−2.90
$Ca^{2+} + 2e^- \rightarrow Ca$	−2.87
$Na^+ + e^- \rightarrow Na$	−2.71
$Mg(OH)_2 + 2e^- \rightarrow Mg + 2OH^-$	−2.69
$Mg^{2+} + 2e^- \rightarrow Mg$	−2.37
$Al^{3+} + 3e^- \rightarrow Al$	−1.66
$MnO_4^- + 8H^+ + 5e^- \rightarrow Mn^{2+} + 4H_2O$	−1.51
$Mn^{2+} + 2e^- \rightarrow Mn$	−1.18
$2H_2O + 2e^- \rightarrow H_2 + 2OH^-$	−0.828
$Zn^{2+} + 2e^- \rightarrow Zn$	−0.763
$Fe^{2+} + 2e^- \rightarrow Fe$	−0.440
$Cd^{2+} + 2e^- \rightarrow Cd$	−0.40
$Ni^{2+} + 2e^- \rightarrow Ni$	−0.250
$S + 2H^+ + 2e^- \rightarrow H_2S$	−0.14
$Pb^{2+} + 2e^- \rightarrow Pb$	−0.126
$2H^+ + 2e^- \rightarrow H_2$	0.000
$Cu^{2+} + e^- \rightarrow Cu^+$	+0.15
$N_2 + 4H^+ + 3e^- \rightarrow NH_4{}^+$	+0.27
$Cu^{2+} + 2e^- \rightarrow Cu$	+0.34
$I_2 + 2e^- \rightarrow 2I^-$	+0.54
$O_2 + 2H^+ + 2e^- \rightarrow H_2O_2$	+0.68
$Fe^{3+} + e^- \rightarrow Fe^{2+}$	+0.771
$Ag^+ + e^- \rightarrow Ag$	+0.799
$ClO^- + H_2O + 2e^- \rightarrow Cl^- + 2OH^-$	+0.90
$Br_2(aq) + 2e^- \rightarrow 2Br^-$	+1.09
$O_2 + 4H^+ + 4e^- \rightarrow 2H_2O$	+1.229
$Cl_2(g) + 2e^- \rightarrow 2Cl^-$	+1.360
$H_2O_2 + 2H^+ + 2e^- \rightarrow 2H_2O$	+1.776
$O_3 + 2H^+ + 2e^- \leftrightarrow O_2 + H_2O$	+2.07
$F_2 + 2H^+ + 2e^- \rightarrow 2HF$	+2.87

[a] Adapted in part from Bard (1966) and Benefield et al. (1982).

[b] Reported values will vary depending on source. The electrical potential or voltage (V) is used commonly to characterize various oxidation reduction reactions.

Reaction Potentials. The half-reaction potentials, discussed above, can be used to predict whether a reaction comprised of two half-reactions will proceed as written. The tendency of a reaction to proceed is obtained by determining the $E^\circ_{\text{reaction}}$ for the entire reaction as given by the following expression.

$$E^\circ_{\text{reaction}} = E^\circ_{\text{reduction}} - E^\circ_{\text{oxidation}} \tag{6–35}$$

Where $E^\circ_{\text{reaction}}$ = the potential of the overall reaction, V
$E^\circ_{\text{reduction}}$ = the potential of the reduction half reaction, V
$E^\circ_{\text{oxidation}}$ = the potential of the oxidation half reaction, V

For example, for the reaction between copper and zinc (see Eq. 6–30) the $E^\circ_{\text{reaction}}$ is determined as follows:

$$E^\circ_{\text{reaction}} = E^\circ_{\text{Cu}^{2+},\text{Cu}} - E^\circ_{\text{Zn}^{2+},\text{Zn}} \tag{6–36}$$

$$E^\circ_{\text{reaction}} = 0.340 - (-0.763) = +1.103 \text{ V} \tag{6–37}$$

The positive value for the $E^\circ_{\text{reaction}}$ is taken as an indication that the reaction will proceed as written. The magnitude of the value, as will be illustrated subsequently, can be taken as measure of the extent to which the reaction as written will proceed. For example, if Eq. (6–30) had been written as follows:

$$\text{Cu} + \text{Zn}^{2+} \rightleftarrows \text{Cu}^{2+} + \text{Zn} \tag{6–38}$$

The corresponding $E^\circ_{\text{reaction}}$ for this reaction is:

$$E^\circ_{\text{reaction}} = E^\circ_{\text{Zn}^{2+},\text{Zn}} - E^\circ_{\text{Cu}^{2+},\text{Cu}} \tag{6–39}$$

$$E^\circ_{\text{reaction}} = (-0.763) - 0.340 = -1.103 \text{ V} \tag{6–40}$$

Because the $E^\circ_{\text{reaction}}$ for the reaction is negative, the reaction will proceed in the opposite direction from what is written.

EXAMPLE 6–4 **Determination of Reaction Potential** Determine whether hydrogen sulfide (H_2S) can be oxidized with hydrogen peroxide (H_2O_2). The pertinent half reactions from Table 6–13 are as follows:

$$H_2S \rightleftarrows S + 2H^+ + 2e^- \qquad E^\circ = -0.14 \text{ V}$$

$$H_2O_2 + 2H^+ + 2e^- \rightleftarrows 2H_2O \qquad E^\circ = +1.78 \text{ V}$$

Solution

1. Determine the overall reaction by adding the two half reactions.

$$\begin{array}{l} H_2S \rightleftarrows S + 2H^+ + 2e^- \\ \underline{H_2O_2 + 2H^+ + 2e^- \rightleftarrows 2H_2O} \\ H_2S + H_2O_2 \rightleftarrows S + 2H_2O \end{array}$$

2. Determine the $E^\circ_{\text{reaction}}$ for overall reaction.

$$E^\circ_{\text{reaction}} = E^\circ_{H_2O_2,\, H_2O} - E^\circ_{H_2S,\, S}$$

$$E^\circ_{\text{reaction}} = (1.78) - (-0.14) = +1.92 \text{ V}$$

Because the $E^\circ_{\text{reaction}}$ for the reaction is positive, the reaction is expected to proceed as written.

Equilibrium Constants for Redox Equations. The equilibrium constant for oxidation reduction reactions is calculated using the Nernst equation as defined below.

$$\ln K = \frac{nFE^\circ_{\text{reaction}}}{\text{RT}} \tag{6–41}$$

$$\log K = \frac{nFE^\circ_{\text{reaction}}}{2.303\,RT} \tag{6–42}$$

Where K = equilibrium constant
n = number of electrons exchanged in the overall reaction
F = Faraday's constant
= 96,485 C/g-eq (Note: C = coulomb)
$E^\circ_{\text{reaction}}$ = reaction potential
R = universal gas constant
= 8.3144 J (abs)/g-mole
T = temperature, K (273.15 + °C)

For example, at 25°C

$$\log K = \frac{n\,(96{,}485\text{ C/g eq})\,E^\circ_{\text{reaction}}}{(2.303)(8.3144\text{ J/mole}\cdot\text{K})[(273.15 + 25)K]} = \frac{nE^\circ_{\text{reaction}}}{0.0592}$$

The application of the Nernst equation is illustrated in Example 6–5.

EXAMPLE 6–5 **Determination of Equilibrium Constant for Oxidation-Reduction Reaction** Determine the equilibrium constant for the following oxidation-reduction reactions:

$$Cu^{2+} + Zn \rightleftarrows Cu + Zn^{2+}$$

$$H_2S + H_2O_2 \rightleftarrows S + 2H_2O$$

Solution

1. Determine the equilibrium constant for the following equation using Eq. (6–42).

$$Cu^{2+} + Zn \rightleftarrows Cu + Zn^{2+}$$

As computed above, the $E^\circ_{\text{reaction}}$ for the reaction is +1.1 volts, and the number of electrons exchanged is 2. Using this information, the K value is determined as follows:

$$\log K = \frac{nE^\circ_{\text{reaction}}}{0.0592} = \frac{2(1.10)}{0.0592} = 37.2$$

$$K = 1.58 \times 10^{37} = \frac{[Zn^{2+}]}{[Cu^{2+}]}$$

2. Determine the equilibrium constant for the following equation using Eq. (6–42).

$$H_2S + H_2O_2 \rightleftarrows S + 2H_2O$$

From Example 6–4, the $E^\circ_{\text{reaction}}$ for the above reaction is $+1.92$ volts. The value of the equilibrium constant is

$$\log K = \frac{nE^\circ_{\text{reaction}}}{0.0592} = \frac{2(1.92)}{0.0592} = 64.9$$

$$K = 7.94 \times 10^{64} = \frac{[\text{S}]}{[\text{H}_2\text{S}][\text{H}_2\text{O}_2]}$$

Rate of Oxidation-Reduction Reactions. As noted previously, the half-reaction potentials can be used to predict whether a reaction will proceed as written. Unfortunately, the reaction potential provides no information about the rate at which the reaction will proceed. Chemical oxidation reactions often require the presence of one or more catalysts for the reaction to proceed or to increase the rate of reaction. Transition metal cations, enzymes, pH adjustment, and a variety of proprietary substances have been used as catalysts.

Chemical Oxidation of Organic Constituents

The overall reaction for the oxidation of organic molecules comprising residual COD, for example, with ozone, can be represented as follows:

$$\begin{matrix}\text{Organic} & & \text{Intermediate} & & \text{Simple} \\ \text{molecule} & \xrightarrow{O_3} & \text{oxygenated} & \xrightarrow{O_3} & \text{end products} \\ \text{(e.g., COD)} & & \text{molecules} & & \text{(e.g., CO}_2\text{, H}_2\text{O, etc.)}\end{matrix} \qquad (6\text{–}43)$$

Multiple arrows in the direction of the reaction are used to signify that a number of steps are involved in the overall reaction sequence. The use of oxidizing agents such as oxygen, chlorine, ozone, and hydrogen peroxide is termed "simple oxidation." In general the overall reaction rates are usually too slow to be applicable generally for the removal of recalcitrant COD during wastewater treatment (SES, 1994). Advanced oxidation processes (AOPs), which typically involve the use of the hydroxyl radical for the oxidation of complex organic molecules, are considered in Sec. 6–8.

Typical chemical dosages for both chlorine and ozone for the oxidation of the organics in wastewater are reported in Table 6–14. The dosages increase with the degree of treatment, which is reasonable when it is considered that the organic compounds that remain after biological treatment are typically composed of low molecular weight polar organic compounds and complex organic compounds built around the benzene ring structure.

It is recommended that bench and/or pilot plant studies be conducted when chlorine, chlorine dioxide, or ozone is to be used for the oxidation of organics. Because ozone can be generated conveniently at treatment plants that use the high purity oxygen activated sludge process, it is anticipated that its use may become more common at these locations in the future.

Chemical Oxidation of Ammonium

The chemical process in which chlorine is used to oxidize the ammonium nitrogen in solution to nitrogen gas and other stable compounds is known as breakpoint chlorination. Perhaps the most important advantage of this process is that, with proper control, all the

Table 6–14

Typical chemical dosages for the oxidation of organics in wastewater[a]

Chemical	Use	Dosage, kg/kg destroyed Range	Typical
Chlorine	BOD reduction		
	Settled wastewater	0.5–2.5	1.75
	Secondary effluent	1.0–3.0	2.0
Ozone	COD reduction		
	Settled wastewater	2.0–4.0	3.0
	Secondary effluent	3.0–8.0	6.0

[a] Derived in part from Black and Veatch Corporation (2010).

ammonium nitrogen in the wastewater can be oxidized. However, because the process has a number of disadvantages including cost, the buildup of acid (HCl) which will react with the alkalinity, the buildup of total dissolved solids, and the formation of unwanted chloro-organic compounds, ammonium oxidation is seldom used for effluents that have not undergone nitrification. Where nitrification is practiced, however, breakpoint chlorination can be an effective tool to achieve high levels of disinfection and eliminate residual ammonium.

Although the theory and practice of breakpoint chlorination is described in greater detail in Chap 12, it is, nevertheless, instructive to assess whether the reaction is feasible using the oxidation-reduction concepts presented above. The pertinent half reactions are as follows:

$$\mathrm{HOCl} + \mathrm{H}^+ + 2\mathrm{e}^- \rightleftarrows \mathrm{Cl}^- + \mathrm{H_2O} \qquad E^\circ = +1.49\ \mathrm{V} \tag{6–44}$$

$$\mathrm{N_2} + 8\mathrm{H}^+ + 6\mathrm{e}^- \rightleftarrows 2\mathrm{NH_4^+} \qquad E^\circ = +0.27\ \mathrm{V} \tag{6–45}$$

Rewrite Eq. (6–45) as a reduction

$$2\mathrm{NH_4^+} \rightleftarrows \mathrm{N_2} + 8\mathrm{H}^+ + 6\mathrm{e}^- \qquad E^\circ = -0.27\ \mathrm{V} \tag{6–46}$$

Combining Eqs. (6–44) and (6–46) yields

$$\begin{array}{l} 2\mathrm{NH_4^+} \rightleftarrows \mathrm{N_2} + 8\mathrm{H}^+ + 6\mathrm{e}^- \\ \underline{(3)\mathrm{HOCl} + (3)\mathrm{H}^+ + (3)2\mathrm{e}^- \rightleftarrows (3)\mathrm{Cl}^- + (3)\mathrm{H_2O}} \\ 3\mathrm{HOCl} + 2\mathrm{NH_4^+} \rightleftarrows \mathrm{N_2} + 3\mathrm{HCl} + 2\mathrm{H}^+ + 3\mathrm{H_2O} \end{array} \tag{6–47}$$

Determine the $E^\circ_{\mathrm{reaction}}$ for overall reaction

$$E^\circ_{\mathrm{reaction}} = E^\circ_{\mathrm{HOC1,Cl^-}} - E^\circ_{\mathrm{NH_4^+,N_2}}$$

$$E^\circ_{\mathrm{reaction}} = (1.49) - (-0.27) = +1.96\ \mathrm{V}$$

Because the $E^\circ_{\mathrm{reaction}}$ for the reaction is positive, the reaction will proceed as written. The stoichiometric mass ratio of chlorine as $\mathrm{Cl_2}$ to ammonium as N, as computed using Eq. (6–47) is 7.6:1. In practice, the ratio has been found to vary from 8:1 to 10:1.

Chemical Oxidation Process Limitations

Aside from the expense of chemical addition, the primary concern with any chemical oxidation process is the potential for the formation of toxic byproducts due to incomplete oxidation. While the hydroxyl radical oxidation processes discussed in Sec. 6–8 are able to achieve complete mineralization of many constituents under optimal conditions, conventional chemical oxidation is typically not powerful enough to reach this endpoint. Therefore, subsequent treatment processes, such as adsorption, will be needed to remove the oxidation byproducts. Further, chemical oxidation will increase the biodegradability of some constituents, potentially requiring the use of a biological process to remove residual biodegradable organic material. Byproduct formation may be controlled by removal of the byproduct precursors before application of the oxidant and careful control of the oxidant dose.

The properties of chemical oxidants that justify their use in wastewater treatment also contribute to their potential to be corrosive under certain conditions. Thus, careful control of oxidant dosage and the use of compatible materials are important factors to prevent corrosion of facilities and equipment. A number of methods are available to evaluate the potential, type, and rate of corrosion by chemical oxidants with a given material under certain conditions, including thermodynamics, electrokinetics (mixed-potential models), and experimental testing.

6–8 ADVANCED OXIDATION

Advanced oxidation processes (AOPs) can be used to destroy trace constituents that cannot be oxidized completely by conventional oxidants, including constituents that are known to affect the endocrine system (Rosenfeldt and Linden, 2004). Treated effluent water following tertiary treatment typically contains a variety of natural and synthetic organic chemicals at low concentrations that may need to be removed or destroyed to protect public health and the environment, especially in potable reuse applications. The conventional oxidants described in Sec. 6–7 are able to remove some of the constituents of interest; however, there is uncertainty regarding the formation of toxic byproducts following conventional chemical oxidation. Moreover, some trace constituents may be found in the permeate from reverse osmosis treatment.

The advantage of advanced oxidation is the ability to generate elevated concentrations of hydroxyl radical ($HO\cdot$), a strong oxidant capable of the complete oxidation of most organic compounds into carbon dioxide, water, and mineral acids (e.g., HCl). The dot is added after the radical species to denote that an unpaired electron is present in the outer orbital of the molecule. Because of the unpaired electron, hydroxyl radicals are reactive electrophiles (electron-loving) that react rapidly with nearly all electron-rich organic compounds. The reactions with hydroxyl radical are second order because the reactions depend on the concentration of the constituent that is oxidized and the concentration of the hydroxyl radical species. The second order hydroxyl radical rate constants for many dissolved organic compounds are on the order of 10^8 to 10^9 L/mole·s (Buxton and Greenstock, 1988), three to four orders of magnitude greater than second order rate constants for other oxidants.

Applications for Advanced Oxidation

The relative oxidizing power of the hydroxyl radical expressed as electrochemical oxidation potential, along with other common oxidants, is summarized in Table 6–15. As shown, with the exception of fluorine, the hydroxyl radical is one of the most active oxidants known.

Table 6–15
Oxidants, forms, and applications methods

Oxidant	Forms	Applications[a]	Application methods	Electrical potential, V
Fluorine	Not used	—	Not used	2.87
Hydroxyl radical	Generated in specially designed reactors at the moment of use due to short life	A, B, C, D	See Sec. 6–8	2.80
Ozone	Ozone is a gas that is generated on site by passing dry compressed air or pure oxygen across a high voltage electrode	A, B, C, D	Ozone is applied to water as a gas. Mass transfer is an important issue and ozone contactors usually use fine bubble diffusers and water depth to ensure high transfer efficiency	2.08
Peracetic acid	Stabilized liquid solution	A, D	Concentrated solution mixed with water to be treated	—
Hydrogen peroxide	Liquid solution	A	Concentrated solution mixed with water to be treated	1.78
Permanganate	Available in bulk as granules	A, B	Added as a dry chemical using a feeder or as a concentrated solution (no more than 5 percent by weight due to its limited solubility)	1.67
Chlorine, free	Chlorine gas, NaOCl solution	A, D	Gas eductors and spray jets	1.36
Chlorine, combined (chloramines)	Addition of ammonia: anhydrous ammonia gas, ammonium sulfate, aqua ammonia (20 to 30 percent ammonia solution)	A, D	Gas eductors, dry chemical feeders, spray jets	—
Chlorine dioxide	Chlorine dioxide gas is produced on-site using a 25 percent sodium chlorite solution. The sodium chlorite solution is reacted with the following constituents to form $ClO_{2\,(g)}$: (1) gaseous chlorine (Cl_2), (2) aqueous chlorine (HOCl), or acid (usually hydrochloric acid, HCl)	A	Gas eductors	1.27
Oxygen	Gas and Liquid	—	Pure oxygen or the oxygen in air is applied as a gas	1.23

[a] A = Oxidation of reduced inorganic species such as soluble metals, complexed metal species, and destruction of odor causing compounds.
B = Oxidation of organics such as toxic organic compounds, colors, reduce TOC and NOM.
C = Improve coagulation.
D = Used as a biocide to control algae in reservoirs and basins, for disinfection, to control growth in distributions system.

Advanced oxidation processes differ from the other advanced treatment processes discussed (such as adsorption, ion exchange, or stripping) because organic compounds in water are degraded rather than concentrated or transferred into a different phase. Further, compounds that are not adsorbable or only partially adsorbable may be destroyed by reaction with hydroxyl radical. Because a secondary waste stream is not generated, there is no additional cost to dispose of or regenerate materials. Hydroxyl radicals are capable of oxidizing almost all reduced materials present without restriction to specific classes or groups of compounds, as compared to other oxidants. In addition to being non-selective, many AOPs operate at normal temperature and pressures. Other processes that can generate hydroxyl radicals, but require elevated temperature and/or pressure, include catalytic oxidation, gas-phase combustion, supercritical oxidation, and wet oxidation processes. Additional details on AOPs may be found in Singer and Reckhow (1999) and Crittenden et al. (2012).

Degree of Degradation. Depending on the application, it may not be necessary to oxidize completely a given compound or group of compounds, as partial oxidation may be sufficient to render specific compounds more amenable to subsequent biological treatment or to reduce their toxicity. The oxidation of specific compounds may be characterized by the extent of degradation of the final oxidation products as follows (Rice, 1996).

1. *Primary degradation.* A structural change in the parent compound.
2. *Acceptable degradation (defusing).* A structural change in the parent compound to the extent that toxicity is reduced.
3. *Ultimate degradation (mineralization).* Conversion of organic carbon to inorganic CO_2.
4. *Unacceptable degradation (fusing).* A structural change in the parent compound resulting in increased toxicity.

Oxidation of Refractory Organic Compounds. Hydroxyl radicals are used most commonly for the oxidation of trace amounts of refractory organic compounds found in highly treated effluents (e.g., following reverse osmosis). The hydroxyl radicals, once generated can attack organic molecules by (1) radical addition, (2) hydrogen abstraction, (3) electron transfer, and (4) radical combination (SES, 1994) as described below.

1. By radical addition:
   The addition of the hydroxyl radical to an unsaturated aliphatic or aromatic organic compound (e.g., C_6H_6) results in the production of a radical organic compound that can be oxidized further by compounds such as oxygen or ferrous iron to produce stable oxidized end products. Radical addition is much more rapid than hydrogen abstraction. In the following reactions the abbreviation R is used to denote the reacting organic compound.

   $$R + HO\cdot \rightarrow ROH\cdot \tag{6–48}$$

2. By hydrogen abstraction:
   The hydroxyl radical can be used to remove a hydrogen atom from organic compounds. The removal of a hydrogen atom results in the formation of a radical organic compound, initiating a chain reaction where the radical organic compound reacts with oxygen, producing a peroxyl radical, which can react with another organic compound, and so on.

   $$R + HO\cdot \rightarrow R\cdot + H_2O \tag{6–49}$$

3. By electron transfer:
   Electron transfer results in the formation of ions of a higher valance. Oxidation of a monovalent negative ion will result in the formation of an atom or a free radical. In the following reaction, n is used to denote the charge on the reacting organic compound R.

   $$R^n + HO\cdot \rightarrow R^{n-1} + OH^- \tag{6–50}$$

4. By radical combination:
Two radicals can combine to form a stable product.

$$HO\cdot + HO\cdot \rightarrow H_2O_2 \qquad (6\text{–}51)$$

The reaction of HO· with organic compounds by radical addition reactions with double bonds and hydrogen abstraction are among the most common processes. In general, the reaction of hydroxyl radicals with organic compounds, at completion, will produce water, carbon dioxide, and mineral salts; this process is also known as *mineralization*.

Disinfection. Because it was recognized that free radicals generated from ozone were more powerful oxidants than ozone alone, it was reasoned that the hydroxyl free radicals could be used effectively to oxidize microorganisms in wastewater. Unfortunately, because the half-life of the hydroxyl free radicals is short, on the order of microseconds, it is not possible to develop hydroxyl radicals in high concentrations. With extremely low concentrations, the required detention times for microorganism disinfection, based on the C_Rt concept (see Chap. 12), are prohibitive. However, AOPs that incorporate high dosages of UV energy (1000 to 2000 mJ/cm^2) to initiate photolysis reactions may be of sufficient intensity to accomplish significant levels of disinfection. Challenge testing using pilot or full scale installations may be used to determine the actual level of disinfection accomplished.

Processes for Advanced Oxidation

Based on numerous studies, it has been found that AOPs are more effective than any of the individual oxidizing agents (e.g., ozone, UV, hydrogen peroxide). Several technologies are available to produce HO· in the aqueous phase (U.S. EPA, 1998). Selected technologies are summarized in Table 6–16. In water reclamation, AOPs are usually applied to low COD wastewaters (typically following treatment by reverse osmosis) because of the cost of ozone and/or H_2O_2 required to generate the hydroxyl radicals. Of the technologies reported in Table 6–16, the commercially available AOPs for water reclamation are ozone/UV, ozone/hydrogen peroxide, and hydrogen peroxide/UV.

The major advantages and disadvantages of various AOPs are also provided in Table 6–16. It should be noted that following oxidation, constituents that were previously resistant to degradation may be transformed into biodegradable compounds that will require further biological treatment.

Ozone/UV. Production of the free radical HO· with UV light can be illustrated by the following reactions for the photolysis of ozone (Glaze et al., 1987; Glaze and Kang, 1990). The first step of ozone/ultraviolet light (O_3/UV) process is the formation of H_2O_2 by photolysis of ozone.

$$O_3 + H_2O + UV(\lambda < 310\text{ nm}) \rightarrow O_2 + HO\cdot + HO\cdot \rightarrow O_2 + H_2O_2 \qquad (6\text{–}52)$$

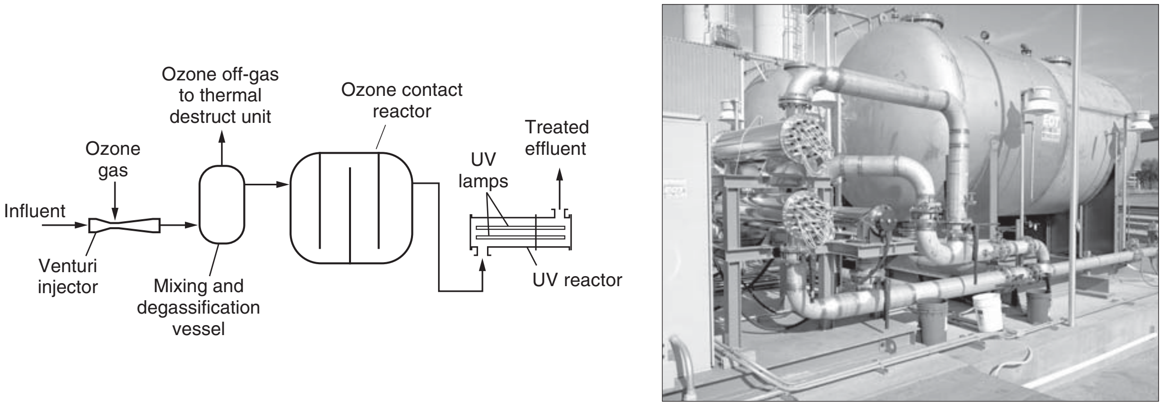
As shown in Eq. (6–52), the photolysis of ozone in wet air results in the formation of hydroxyl radicals. In water, the photolysis of ozone leads to the formation of hydrogen peroxide, which is subsequently photolyzed or reacted with O_3 to form hydroxyl radicals. The ozone/UV process can degrade compounds through direct ozonation, photolysis, or reaction with the hydroxyl radical, resulting in a process that is more effective when the compounds of interest can be degraded through the absorption of the UV irradiation as well as through the reaction with the hydroxyl radicals. Basic components of the ozone/UV process include ozone gas generation, ozone injection facilities, and UV photolysis reactors. A schematic flow diagram and view of a typical ozone/UV oxidation process is illustrated on Fig. 6–24.

Table 6–16

Advantages and disadvantages of various oxidation processes that produce hydroxyl radicals[a]

Advanced oxidation process	Advantages	Disadvantages
Commercially available AOPs for water reclamation		
Hydrogen peroxide/ultraviolet light	H_2O_2 is fairly stable and can be stored on site temporarily prior to use	H_2O_2 has poor UV absorption characteristics and if the water matrix absorbs UV light energy then most of the light input to the reactor will be wasted Special reactors which are designed for UV illumination are required Residual H_2O_2 must be removed Potential for UV lamp fouling
Hydrogen peroxide/ozone	Waters with poor UV-light transmission may be treated Special reactors designed for UV illumination are not required Volatile organics will be stripped from the ozone contactor (VOCs may require treatment)	Production of O_3 can be an expensive and inefficient process Gaseous ozone which is present in the off gas of the ozone contactor must be removed Maintaining and determining the proper O_3/H_2O_2 dosages may be difficult Low pH is detrimental to the process
Ozone/UV	No need to maintain precise dosages of O_3/H_2O_2 Residual oxidant will degrade rapidly (typical half life of O_3 is 7 minutes) Ozone absorbs more UV light than an equivalent dosage of hydrogen peroxide. (~200 times more at 254 nm) Volatile compounds will be stripped from the process (VOCs may require treatment)	Must use O_3 and UV light to produce H_2O_2, which is the primary means of producing HO· and using O_3 to produce H_2O_2 is very inefficient as compared to just adding H_2O_2 Special reactors which are designed for UV illumination are required Ozone in the off-gas must be removed Potential for UV lamp fouling
Other selected AOPs		
Ozone/UV/H_2O_2	Commercial processes that utilize the technology are available H_2O_2 promotes ozone mass transfer Volatile compounds will be stripped from the process. (VOCs may require treatment)	Special reactors that are designed for UV illumination are required Ozone in the off-gas must be removed Potential for UV lamp fouling
Fenton's reactions (Fe/hydrogen peroxide, photo-Fenton's or Fe/ozone)	Some effluents may contain sufficient Fe to drive the Fenton's reaction Commercial processes are available that utilize the technology	Process requires low pH
Titanium dioxide/UV	Activated with near UV light; consequently greater light transmission is achievable	Fouling of the catalyst may occur When used as a slurry, the TiO_2 must be recovered Potential for UV lamp fouling
Ozone at elevated pH (8 to >10)	Does not require the addition of UV light or hydrogen peroxide	Ozone is the off-gas must be removed pH adjustment is not practical in most wastewater applications There are no commercial applications Process does not yield an appreciable destruction for contaminants for reason provided in Sec. 6–7

[a] Adapted from Crittenden et al. (2012).



(a)

(b)

Figure 6–24

Advanced oxidation process involving the use of ozone and UV radiation (a) schematic representation (ozone contactor shown without chimneys, see also Chap. 12) and (b) view of full scale installation.

While it is possible for UV light to split H_2O_2 into HO•, the extinction coefficient for O_3 is greater than that for H_2O_2 at λ_{254} nm. Thus, using ozone to produce H_2O_2, which in turn reacts with O_3 to produce HO•, may not be the most efficient way to produce HO• radicals because of the large amount of energy required to form ozone onsite. Processes involving ozone and UV dosages in the range of 16 to 24 mg/L and 810 to 1610 mJ/cm^2, respectively, have been found to mineralize a portion of the TOC and to reduce disinfection byproduct formation potential as compared to the use of either UV or ozone alone (Chin and Berube, 2005). As with all UV processes, fouling of the UV lamp sleeve, lamp replacement costs, and energy consumption are important considerations.

Ozone/Hydrogen Peroxide. For compounds that do not adsorb UV or where the transmittance of the water to be treated inhibits photolysis, AOPs involving ozone/H_2O_2 may be more effective than ozone/UV. Processes using O_3/H_2O_2 have been used to reduce the concentration of assorted VOCs, petroleum compounds, industrial solvents, and pesticides in water (Karimi et al., 1997; Mahar et al., 2004; Chen et al., 2006). The overall reaction for the production of hydroxyl radicals using hydrogen peroxide and ozone is as follows.

$$H_2O_2 + 2O_3 \rightarrow HO\cdot + HO\cdot + 3O_2 \qquad (6\text{–}53)$$

According to Eq. (6–53), 0.5 moles of H_2O_2 are needed for every mole of O_3 or a mass ratio of 0.354 kg of H_2O_2 is needed for every kg of O_3. However, there are several issues that impact the proper dosages of H_2O_2 and O_3. First, O_3 tends to be more reactive with background organic matter and inorganic species than with H_2O_2. As a result, the required O_3 dosage will be higher than estimated from stoichiometry. Typical ozone and hydrogen peroxide concentrations range from 5 to 30 mg/L and 5 to 15 mg/L, respectively. Pilot studies are usually conducted to determine the chemical dosage required for a given level of trace constituent removal. However, an excess O_3 dosage has the potential of wasting O_3, forming oxidation byproducts (e.g., bromate), and quenching HO• radicals via the following reaction:

$$O_3 + HO\cdot \rightarrow HO_2\cdot + O_2 \qquad (6\text{–}54)$$

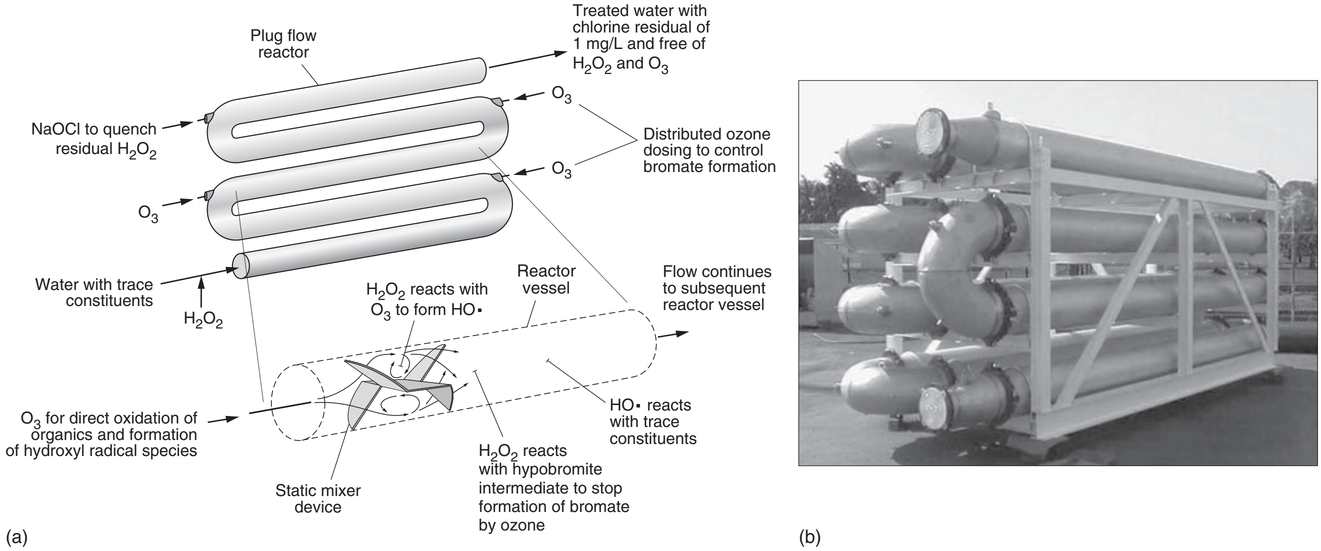


Figure 6–25

Advanced oxidation process involving the use of ozone and hydrogen peroxide (a) schematic of HiPOx® reactor and (b) view of reactor described in (a) (courtesy of Applied Process Technology, Inc., Pleasant Hill, CA.).

The $HO_2\cdot$ radical, formed by Eq. (6–54), may react to produce additional $HO\cdot$. To overcome the problem of byproduct formation and quenching of $HO\cdot$, new reactor designs have incorporated the addition of H_2O_2 or O_3 at multiple points in a single reactor and by using multiple reactors in series. A schematic flow diagram and view of a reactor used for reacting hydrogen peroxide and ozone is shown on Fig. 6–25. Excess H_2O_2 is also detrimental to the H_2O_2/O_3 AOP because it may scavenge $HO\cdot$. Further, the H_2O_2 residual can be more problematic than ozone because hydrogen peroxide is more stable than ozone and, in some applications, it may be necessary to remove residual hydrogen peroxide before reuse. Hydrogen peroxide reacts quickly with hypochlorite to form water, oxygen, and chloride ion.

Hydrogen Peroxide/UV. Hydroxyl radicals are also formed when water containing H_2O_2 is exposed to UV light (200 to 280 nm). The following reaction can be used to describe the photolysis of H_2O_2.

$$H_2O_2 + \text{UV (or } h\nu,\ \lambda \approx 200 \text{ to } 280 \text{ nm)} \rightarrow HO\cdot + HO\cdot \tag{6–55}$$

In some cases the use of the hydrogen peroxide/UV process has not been feasible because H_2O_2 has a small molar extinction coefficient, requiring high concentrations of H_2O_2 and high UV dosages. A schematic flow diagram and a typical installation of the hydrogen peroxide/UV process is shown on Fig. 6–26.

The basic elements of the hydrogen peroxide/ultraviolet light (H_2O_2/UV) process includes hydrogen peroxide injection and mixing followed by a reactor that is equipped with UV lights (see Fig. 6–26) Typical H_2O_2/UV reactors configurations include inline stainless steel reactors with low pressure (low and high intensity) or medium pressure UV lamps arranged parallel to the flow, perpendicular to the flow, or in upflow columns with crisscrossing lamps oriented perpendicular to the direction of flow.

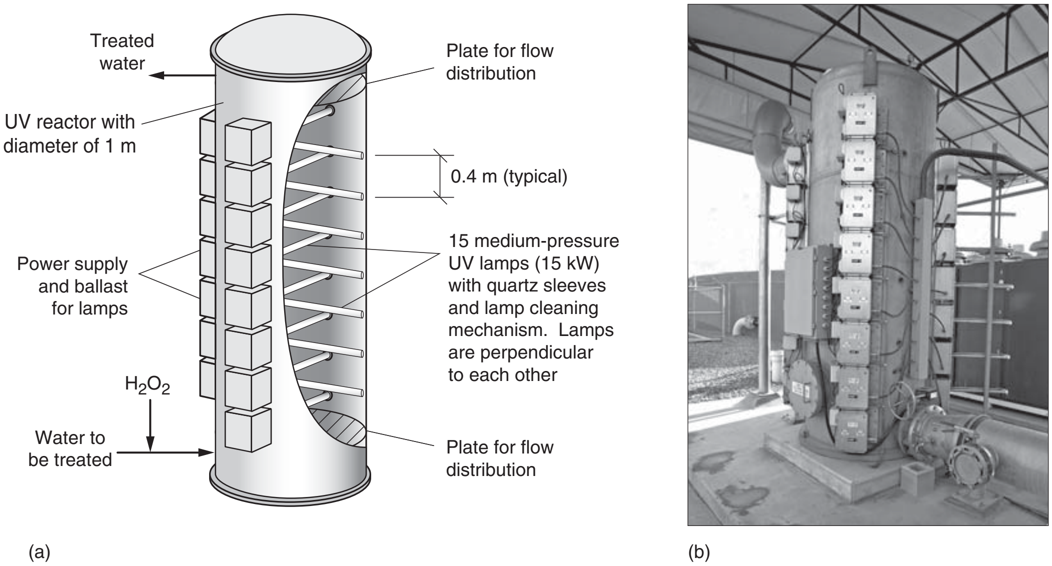


Figure 6–26

Hydrogen peroxide and UV radiation advanced oxidation process (a) schematic diagram (from Crittenden et al., 2012) and (b) photograph of typical vertical flow UV reactor.

The H_2O_2/UV process has not been used commonly for potable water treatment because it normally results in high effluent H_2O_2 concentrations. However, the residual hydrogen peroxide is not a concern in water reclamation. High effluent H_2O_2 concentrations occur because high initial dosages of H_2O_2 are required to efficiently utilize the UV light and produce hydroxyl radical. The residual H_2O_2 will consume chlorine and interfere with disinfection. In some instances, where high UV doses are required, as in the photolyisis of NDMA (see Chap. 2), H_2O_2 may be added to achieve advanced oxidation of other constituents that are resistant to photolysis alone (Linden et al., 2004). This method of operation is now being used in a number of water reclamation applications. The details required for modeling the H_2O_2/UV process can be found in Crittenden et al. (1999). As discussed in Chap. 12, UV processes are subject to fouling of the UV lamp sleeve, lamp replacement costs, and high energy consumption.

Other Processes. Other reactions which yield HO· include the reactions of H_2O_2 and UV with Fenton's reagent and the adsorption of UV by semiconductor metal oxides such as TiO_2 suspended in water which acts as a catalyst. Still others are currently under development.

Basic Considerations for Advanced Oxidation Processes

The engineering of an advanced oxidation process consists of the selection of a process to generate hydroxyl radicals, estimation of the reaction kinetics with the compounds of interest, and design of a reactor in which the reaction will take place. In addition, the presence of background organic and inorganic matter that reacts with the HO· will reduce the process efficiency for the target compound. Therefore, bench and/or pilot studies are always necessary to determine process efficiency for a given water matrix.

Commercially available advanced oxidation processes are rated for a given hydroxyl radical output. Reported field HO· concentrations range from 10^{-11} to 10^{-9} moles/L (Glaze et al., 1987; Glaze and Kang, 1990). The second order hydroxyl radical rate constants for

Table 6–17

Hydroxyl rate constants for selected constituents[a]

Compound name	**HO· rate constant, L/mole·s**	**Compound name**	**HO· rate constant, L/mole·s**
Ammonia	9.00×10^7	Hypobromous acid	2.0×10^9
Arsenic trioxide	1.0×10^9	Hypoiodous acid	5.6×10^4
Bromide ion	1.10×10^{10}	Iodide ion	1.10×10^{10}
Carbon tetrachloride	2.0×10^6	Iodine	1.10×10^{10}
Chlorate ion	1.00×10^6	Iron	3.2×10^8
Chloride ion	4.30×10^9	Methyl tertiary butyl ether (MTBE)	1.6×10^9
Chloroform	5×10^6	Nitrite ion	1.10×10^{10}
CN^-	7.6×10^9	N-Dimethylnitrosamine (NDMA)	4×10^8
CO_3^{2-}	3.9×10^8	Ozone	1.1×10^8
Dibromochloropropane	1.5×10^8	p-Dioxane	2.8×10^9
1,1-Dichloroethane	1.8×10^8	Tetrachloroethylene	2.6×10^9
1,2-Dichloroethane	2.0×10^8	Tetrachloroethylene	1.0×10^7
H_2O_2	2.7×10^7	Tribromomethane	1.8×10^8
HCN	6.0×10^7	Trichloroethylene	4.2×10^9
HCO_3^-	8.5×10^6	Trichloromethane	5.0×10^6
Hydrogen sulfide	1.5×10^{10}	Vinyl chloride	1.2×10^{10}

[a] Adapted from Crittenden et al. (2012).

several compounds of interest are presented in Table 6–17. As mentioned previously, the reactions are second order because they depend on the concentration of the hydroxyl radical and the compound undergoing oxidation. The reaction between HO· and an organic compound of interest, R, is represented as follows:

$$HO\cdot + R \rightarrow \text{byproducts} \tag{6–56}$$

The second order rate law, r_R, corresponding to the reaction shown in Eq. 6–56, is given by the following expression.

$$r_R = -K_R C_{HO}\cdot C_R \tag{6–57}$$

where r_R = second order rate law, mole/L·s
K_R = second order rate constant for the destruction of R with HO· radicals, L/mole·s
$C_{HO\cdot}$ = concentration of hydroxyl radical, mole/L
C_R = concentration of the target organic R, mole/L

The half-life of the target organic compounds may be calculated assuming that the HO· is constant and equal to a typical field value or manufacturer specification. The expression for the half-life of an organic compound is obtained by substituting the rate law into a mass balance on a batch reactor whose contents are mixed completely and solving and rearranging the result, as follows:

$$\frac{dC_R}{dt} = -k_R C_{HO\cdot} C_R \qquad (6–58)$$

$$t_{1/2} = \frac{\ln(2)}{k_R C_{HO\cdot}} \qquad (6–59)$$

where $t_{1/2}$ = the half life of the organic compound R, s.

The use of Eqs. (6–58) and (6–59) is shown in the following example.

EXAMPLE 6–6 **Advanced Oxidation Process for Removal of NDMA** NDMA ($C_2H_6N_2O$) is a compound of concern present in many secondary and tertiary treated wastewater effluents. From Table 6–17, the second order rate constant of HO· for NDMA is 4×10^8 L/mole·s. Compute the time required to lower the concentration of NDMA from 200 μg/L to 20 μg/L for a HO· concentration of 10^{-9} mole/L using an ideal plug flow reactor.

Solution

1. Develop an expression of the concentration of NDMA as a function of time in a completely mixed batch reactor (CMBR). Note that the residence time for an ideal plug flow reactor is equivalent to the residence time in a completely mixed batch reactor.
   a. Using Eq. (6–58), the rate expression for a CMBR, where C_R represents the concentration of NDMA is given by

$$r_R = \frac{dC_R}{dt} = -k_R C_{HO\cdot} C_R = -k' C_R$$

   where $k' = k_R C_{HO\cdot}$

   b. The integrated form of the rate expression for a CMBR is:

$$\int_{C_{RO}}^{C_R} \frac{dC_R}{C_R} = -\int_o^t k't$$

$$C_R = C_{RO} e^{-k't}$$

2. Calculate the time it would take to achieve a concentration of 20 μg/L using the equation developed in step 1.
   a. Rearrange the above equation to solve for t.

$$t = \frac{1}{k'} \ln\left(\frac{C_{RO}}{C_R}\right)$$

   b. Solve for t.

   The value of k' from step 1 is

$$k' = k_R C_{HO\cdot} = (4 \times 10^8 \text{ L/mole·s})(10^{-9} \text{ mole/L}) = 0.4 \text{ 1/s}$$

$$t = \frac{1}{0.4} \ln\left(\frac{200}{20}\right) = 5.8 \text{ s}$$

Comment Advanced oxidation of NDMA appears to be feasible given the short contact time required for the reaction. Because some organic compounds of interest react more slowly with HO•, these compounds will require longer reaction times and/or high concentrations of HO•. The presence of background organic matter, carbonate, bicarbonate, and pH will also reduce the efficiency of the AOP and must be considered during process design. As discussed in Chap. 4, bench and/or pilot testing will be required to determine site specific process design and operational parameters.

Advanced Oxidation Process Limitations

The feasibility and efficiency of AOPs are determined by a number of factors, including interferences and the production of byproducts, as described below. Means of overcoming most process limitations in water reuse applications are also considered.

Byproducts of Advanced Oxidation Processes. Advanced oxidation processes (and processes that use ozone), have been found to produce brominated byproducts and bromate (BrO_3^-) in waters containing bromide ion. The concentration of bromide ion, total organic carbon (TOC) concentration, and pH determine the quantity of brominated byproducts formed. Some AOPs have been designed to minimize bromate formation by pH control or ammonia addition (Crittenden et al., 2012).

Both hydrogen abstraction and radical addition produce reactive organic radicals. The organic radicals undergo subsequent oxidation and may combine with dissolved oxygen to form peroxy organic radicals (ROO•), which subsequently undergo radical chain reactions that produce a variety of oxygenated byproducts. A general pattern of oxidation is presented in Eq. (6–60) (Bolton and Cater, 1994).

$$\begin{matrix}\text{Organic}\\\text{compound}\end{matrix} \rightarrow \text{Aldehydes} \rightarrow \begin{matrix}\text{Carboxylic}\\\text{acids}\end{matrix} \rightarrow \begin{matrix}\text{Carbon dioxide}\\\text{and mineral acids}\end{matrix} \qquad (6\text{–}60)$$

Carboxylic acids are of particular concern as the second order rate constants for these compounds are much lower than for most other organics, which may inhibit degradation of these constituents. Other byproducts that may be of concern are the halogenated acetic acids, formed from the oxidation of halogenated alkenes such as trichloroethene (Crittenden et al., 2012).

Impact of Bicarbonate and Carbonate. High concentrations of carbonate and bicarbonate in some wastewater can react with HO• and reduce the efficiency of advanced oxidation treatment processes. Bicarbonate and carbonate ions are known scavengers of HO• radicals and reduce the rate of organics destruction significantly. Unfortunately, the concentrations of HCO_3^- and CO_3^{2-} are often three orders of magnitude higher than the organic pollutants targeted for destruction. Even low alkalinities (50 mg/L) reduce the rate of TCE destruction by a factor of 10 at a pH of 7 (Crittenden et al., 2012). However, at high pH a given alkalinity is more detrimental because the second order rate constant with CO_3^{2-} is much larger than HCO_3^-. Wastewater with high pH and alkalinity are more difficult to treat using AOPs. To overcome these difficulties and improve the effectiveness of AOPs, pretreatment processes such as softening or reverse osmosis are used to remove the alkalinity.

Impact of pH. The pH affects AOP performance because it determines the distribution of the carbonate species, HCO_3^- and CO_3^{2-}, as discussed above. The pH will also control the concentration of HO_2^- (H_2O_2 has a pK_a of 11.6), which is important in the H_2O_2 type advanced oxidation processes. For example, in the H_2O_2/UV process, HO_2^- has about 10 times the UV absorbance at 254 nm (228 L/mole·cm) than does H_2O_2; consequently, H_2O_2/UV may be more effective at higher pH. Raising the pH to improve the process performance would only be practical if the pH was raised for other purposes such as softening. Finally, pH affects the charge on the organic compounds if they are weak acids or bases. The reactivity and light absorption properties of the compound can be affected by its charge, an observation to be considered in the design of AOPs (Crittenden et al., 2012).

Impact of Metal Ions. Metal ions in reduced oxidation states, such as Fe(II) and Mn(II), can consume a significant quantity of chemical oxidants as well as scavenge HO· radicals. Consequently, the concentration of reduced metal ions should be measured as part of any treatability study, and the dosage of oxidants needed should include the COD of the reduced metal species.

Impact of Other Factors. Other factors that also affect the treatment process include suspended material (which affects light transmission) and the type and nature of the residual TOC and COD. For example, NOM, which reacts with hydroxyl radicals, can have a great influence on the reaction rate. Because the chemistry of the water matrix is different for each reclaimed water, bench and/or pilot testing is almost always required to test the technical feasibility, to obtain usable design data and information, and to obtain operating experience with a specific AOP.

Means of Overcoming Process Limitations. To overcome the problems noted above, advanced oxidation processes are typically applied following treatment by reverse osmosis. Further, if adequate reaction time is provided, $>$99 percent of the organic constituents (as measured by a TOC mass balance) are mineralized (Stefan and Bolton, 1998; Stefan et al., 2000).

6–9 PHOTOLYSIS

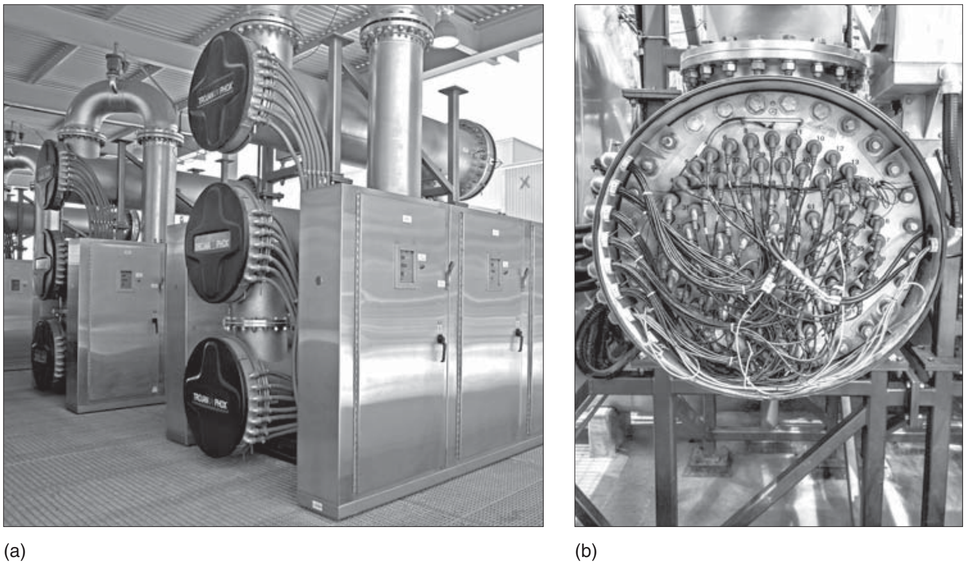
Photolysis is a process by which constituents are broken down by exposure and absorption of photons from a light source. As with AOPs, the primary use of photolysis is for the removal of trace organic compounds in water reuse applications. In natural systems, sunlight is the light source for photolysis reactions, however, in engineered systems, ultraviolet (UV) lamps are used to produce the photonic energy. The photons that are absorbed cause the electrons in the outer orbital of some compounds to become unstable and split or become reactive. The effectiveness of the photolysis process depends, in part, on the characteristics of the reclaimed water, structure of the compounds, design of the photolysis reactor, and dose and wavelength of the applied light. The photolysis rate can be estimated from the rate at which the compound absorbs light and the photonic efficiency of the reaction (quantum yield).

Applications for Photolysis

Photolysis may be used for the removal of various compounds, such as NDMA (see Chap. 3) and other trace organic constituents. It should be noted that many compounds are not removed using photolysis alone, and that the addition of hydrogen peroxide can

Figure 6–27

View of photolysis reactors for advance oxidation for the removal of constituents such as NDMA from reclaimed water: (a) three reactors in series each containing 72 UV lamps and (b) end cover removed from a reactor showing wiring to the 72 UV lamps.


(a)
(b)

enhance the degradation of these constituents. However, the addition of hydrogen peroxide may actually reduce the photolysis of some compounds, such as NDMA (Linden et al., 2004). As described in Sec. 6–8, the photolysis of hydrogen peroxide, resulting in the formation of hydroxyl radicals, is an advanced oxidation process known for its effective destruction of most organic compounds.

Photolysis Processes

Engineered photolysis reactions are conducted in specially designed reactors optimized with respect to UV dose. Photolysis reactors are typically comprised of a stainless steel column or pipe containing UV lamps arranged parallel to the flow, perpendicular to the flow, or in a crisscrossing pattern perpendicular to the direction of flow. An example of a reactor used for photolysis is shown on Fig. 6–27. Fouling that may occur on the outside of the UV lamp protective quartz sleeve may be managed using an automatic cleaning system consisting of a collar that periodically moves along the lamp to remove precipitates and intercepted matter. When used in conjunction with reverse osmosis pretreatment, the chemicals added to reduce the pH for scale control in the membranes will also reduce the potential for precipitation on the UV lamps.

Photolysis reactions are caused by light emission in the ultraviolet range (200 to 400 nm, see also Fig. 12–33 in Chap. 12). There are three types of UV lamps used for photolysis processes: (1) low-pressure low-intensity, (2) low-pressure high-intensity, and (3) medium-pressure high-intensity lamps. Low-pressure lamps emit much of their energy at a wavelength of 254 nm while medium pressure lamps emit energy at multiple wavelengths (see Fig. 12–33 in Chap. 12). The type of lamp used and reactor configuration will depend on the constituent to be removed as well as the water matrix and site-specific conditions.

Basic Considerations for Photolysis Processes

Photolysis occurs when an electron in the outer orbital of a constituent absorbs a photon and forms an unstable compound that splits apart or becomes reactive. Many non-target constituents may be present in reclaimed water, depending on the prior treatment applied, that absorb light during the photolysis process. As an introduction to the photolysis process, the concepts involved in photolysis are presented in the following discussion for the case of a single absorbing solute. The fundamentals of photolysis consist of (1) absorption of UV light by a compound in water, (2) rate of photolysis, (3) electrical efficiency, and (4) photolysis process limitations.

Absorption of UV Light. The absorption of light by a compound in water or other aqueous solution can be described using the Beer-Lambert Law. The absorbance of a solution is a measure of the amount of light absorbed by constituents in the solution using a spectrophotometer at a specified wavelength and over a fixed path length.

$$A(\lambda) = -\log\left(\frac{l}{I_o}\right) = \varepsilon(\lambda)Cx = k(\lambda)x \tag{6–61}$$

where $A(\lambda)$ = absorbance, dimensionless
l = light intensity after passing through solution containing constituents of interest at wavelength, λ, einsteins/cm^2·s (note: an einstein is equal to one mole of photons)
I_o = light intensity after passing through a blank solution (i.e., distilled water) of known depth (typically 1.0 cm) at wavelength, λ, einstein/cm^2·s
$\varepsilon(\lambda)$ = base 10 extinction coefficient or molar absorptivity of light absorbing solute at wavelength λ, L/mole·cm
λ = wavelength, nm
C = concentration of light absorbing solute, mole/L
x = length of light path, cm
$k(\lambda)$ = absorptivity (base 10), 1/cm

The extinction coefficient is a function of wavelength because as the wavelength decreases more energetic photons are absorbed and the absorptivity of a light absorbing compound increases. Values of the extinction coefficients for several compounds at various wavelengths are given in Table 6–18. The use of Eq. (6–61) is presented in Example 6–7.

Table 6–18

Selected quantum yields and extinction coefficients for compounds commonly found in water.

Compound	Primary quantum yield in aqueous phase, mole/einstein	Extinction coefficient at 253.7 nm, L/mole·cm
NO_3^-	—	3.8
HOCl (at 330 nm)	0.23	15
OCl^-	0.23	190
HOCl	—	53.4
OCl^-	0.52	155
O_3	0.5	3300
ClO_2	0.44	108
Sodium chlorite	0.72	—
TCE	0.54	9
PCE	0.29	205
NDMA	0.3	1974
Water	—	0.0000061

[a] Adapted from Crittenden et al. (2012).

EXAMPLE 6–7 **UV Absorbance by NDMA** The chemical compound NDMA (see Chap. 2) is commonly found at low concentrations in reclaimed water, even after treatment by reverse osmosis. Estimate the absorptivity of NDMA ($C_2H_6N_2O$) at a wavelength of 254 nm, assuming NDMA is present at a concentration of 30 ng/L.

Solution

1. Convert the mass concentration of NDMA in solution to mole/L.
   Using the periodic table from the back inside cover of this textbook, the molecular weight of NDMA is 74.09 g/mole. The concentration is determined as follows:

$$C = \frac{(30 \text{ ng/L})}{(74.09 \text{ g/mole})}(1 \text{ g}/10^9 \text{ ng}) = 4.05 \times 10^{-10} \text{ mole/L}$$

2. Compute the absorptivity of NDMA using Eq. (6–61).
   a. Determine the extinction coefficient, $\varepsilon(\lambda)$, of NDMA from Table 6–18. The extinction coefficient of NDMA at a wavelength of 254 nm is 1974 L/mole•cm.
   b. The absorptivity, $k(\lambda)$, of NDMA is:

$$\begin{aligned} k(\lambda) &= \varepsilon(\lambda_{254})C = (1974 \text{ L/mole}\cdot\text{cm})(4.05 \times 10^{-10} \text{ mole/L}) \\ &= 8.0 \times 10^{-7} \text{ cm}^{-1} \end{aligned}$$

Comment Because of the low concentration of NDMA in the water, the absorptivity is also low. If photolysis was to be used for removal of NDMA, other constituents in the water matrix that will absorb photons would need to be considered as well as the background absorbance of the water matrix.

Light absorption by a single compound in water was described in the previous analysis. In practice, however, a number of absorbing compounds will be present in solution. The absorption of light as it passes through a solution containing several different compounds may be determined by summing the absorption that would result from each individual compound as shown in the following expression:

$$\ln\left(\frac{I}{I_o}\right) = -\left[\sum \varepsilon'(\lambda)_i C_i\right] x \tag{6–62}$$

where $\varepsilon'(\lambda)_i$ = extinction coefficient of compound i at wavelength λ (base e), L/(mole•cm); Note: $\varepsilon'(\lambda)_i = 2.303\varepsilon(\lambda)_i$.
C_i = concentration of compound i, mole/L

Other terms as defined previously.

The relationship shown in Eq. (6–62) is based on a single incident wavelength, such as with low-pressure UV lamps. Multiple wavelength situations, such as with the use of medium-pressure UV lamps, can be determined using a similar approach, that is, summing adsorption of each compound for each wavelength.

Energy Input for Photolysis. The lamp output and reactor size can be used to estimate the energy input for the photolysis reaction. The theoretical maximum photonic energy input per unit volume of the reactor can be determined using the following expression:

$$P_R = \frac{P\eta}{N_P\, V\, hv} \tag{6–63}$$

where P_R = photonic energy input per unit volume of the reactor, einstein/L•s
P = lamp power, J/s (W)
h = Planck's constant, 6.62×10^{-34} J•s
η = output efficiency at the wavelength of interest (as a fraction)
N_P = number photons per mole (as einstein), 6.023×10^{23} 1/einstien
V = reactor volume, L
$v = \dfrac{c}{\lambda}$ = frequency of light, 1/s
c = the speed of light, 3.00×10^{8} m/s
λ = wavelength of the light, m

While the above analysis is satisfactory for a theoretical assessment, the actual performance of a photoreactor is expected to be lower than computed using Eq. (6–63) due to light being absorbed by the reactor walls or blocked by the precipitate that forms on the lamp sleeve. While a safety factor specific to a particular system could be applied to compensate for these inefficiencies, pilot studies are used to obtain more reliable design criteria.

Rate of Photolysis. The rate at which a compound is photolyzed depends on the rate and frequency of photon absorption. The volumetric photon absorption rate, derived from Eq. (6–62), is:

$$I_v = -\frac{dl}{dx} = \varepsilon'(\lambda)\cdot C\cdot I_o\cdot e^{-\varepsilon'(\lambda)Cx} \tag{6–64}$$

where I_v = rate that photons are absorbed per volume of solution at a particular point, einstein/cm^3•s
$\varepsilon'(\lambda)$ = base *e* extinction coefficient or molar absorptivity of light absorbing solute at wavelength $\lambda = 2.303\varepsilon(\lambda)$, L/mole•cm

Other terms as defined previously.

The quantum yield is a quantity used to describe the frequency at which photon absorption results in a photolysis reaction and is specific to the type of compound and the wavelength. The quantum yield, $\phi(\lambda)$, is defined as being equal to the number of photolysis reactions divided by the number of photons absorbed by the molecule as follows:

$$\phi(\lambda) = \frac{-r_R}{I_V} = \frac{\text{Reaction rate}}{\text{Rate of photon absorption}} \tag{6–65}$$

where $\phi(\lambda)$ = quantum yield at wavelength λ, mole/einstein
r_R = photolysis rate, mole/(cm^3•s)

As a general rule, the quantum yield increases as wavelength decreases (increasing photonic energy). Selected quantum yields at a wavelength 254 nm are summarized in Table 6–18.

Typically, the light absorption by the component that is targeted for removal is minor as compared to the light absorption by the background water matrix (Crittenden et al., 2012). The pseudo first order rate law for the photolysis reaction is

$$r_{\text{avg}} = \left[\phi(\lambda) P_R \frac{\varepsilon'(\lambda)}{k'(\lambda)}\right] C = kC \tag{6–66}$$

where r_{avg} = overall average photolysis rate of the constituent in the reactor, mole/L·s
$\phi(\lambda)$ = quantum yield of the constituent at wavelength λ, mole/einstein
$\varepsilon'(\lambda)$ = the extinction coefficient of the constituent (base e), L/mole·cm
C_i = the concentration of the constituent, mole/L
$k'(\lambda)$ = measured absorptivity of the water matrix at wavelength (base e) λ, 1/cm
k = pseudo-first order rate coefficient, 1/s

After obtaining the rate law, r_{avg}, an appropriate reactor model may be used for determination of the expected performance.

Electrical Efficiency. The electrical energy requirement for photolytic reactions is significant due to the process inefficiencies. Consequently, it is important to compare process efficiency on the basis of electrical usage per amount of compound destruction. One such measure is the electrical efficiency per log order (EE/O) of compound destruction (Bolton and Cater, 1994). The definition of EE/O is the electrical energy (in kWh) required to reduce the concentration of a constituent by one order of magnitude for 3785 L (1000 U.S. gallons) of water.

$$\text{EE/O} = \frac{Pt}{V \log\left(\dfrac{C_i}{C_f}\right)} \text{ (for batch systems)} \tag{6–67}$$

$$\text{EE/O} = \frac{P}{Q \log\left(\dfrac{C_i}{C_f}\right)} \text{ (for continuous flow systems)} \tag{6–68}$$

where EE/O = electrical efficiency per log order reduction, kWh/m^3
P = lamp power output, kW
t = irradiation time, h
V = reactor volume, m^3
C_i = initial concentration, mg/L
C_f = final concentration, mg/L
Q = water flowrate, m^3/h

For a flow through system, the power input can be divided by the EE/O to obtain an estimate of the flowrate that can be treated in a given reaction and achieve one order of magnitude reduction in concentration. Consequently, EE/O is a convenient measure because it can be used to estimate the energy that is required to reduce the contaminant concentration by one order of magnitude.

Because of the variability in wastewater characteristics, the required EE/O value for a one log order of reduction (i.e., 100 to 10) of NDMA can range from 21 to 265 kWh/10^3 m^3·log order (0.08 to 1.0 kWh/10^3 gal·log order) with a 5 to 6 mg/L dose of H_2O_2, although testing should be conducted to determine if peroxide is necessary (Soroushian et al., 2001).

EXAMPLE 6–8 **Design of Direct Photolysis Process for NDMA** A water reclamation plant produces 1.9×10^4 m^3/d (5 Mgal/d) of reverse osmosis (RO) effluent containing 50 ng/L of NDMA. Determine the number of photolysis reactors needed to reduce the NDMA concentration of the RO effluent to 1 ng/L prior to groundwater injection. The photolysis reactors under evaluation are 0.5 m in diameter and 1.5 m long with an effective water volume of 242 L. Each reactor has 72 lamps rated at 200 W per lamp and an output efficiency of 30 percent at 254 nm. Assume that the hydraulic detention time, τ, of the reactor can be described using the tanks in series model, $\tau = n[(C_e/C_o)^{1/n} - 1]/k$, where k is the reaction rate constant and n is the number of tanks in series. Use three tanks in series and neglect all other losses. The RO water has an absorptivity measured at a wavelength of 254 nm of $k'(\lambda) = 0.02\ \text{cm}^{-1}$. Calculate the EE/O and daily energy usage for the photolysis process.

Solution

1. Calculate the photonic energy input per unit volume of the reactor.
   a. Calculate the total lamp power.

$$P = (72 \text{ lamps} \times 200 \text{ W/lamp}) = 14{,}400 \text{ W} = 14{,}400 \text{ J/s}$$

   b. Calculate the photonic energy input for the reactor using Eq. (6–63).

$$P_R = \frac{(14{,}400 \text{ J/s})(0.3)(254 \times 10^{-9} \text{ m})}{(6.023 \times 10^{23} \text{ 1/einstein})(6.62 \times 10^{-34} \text{J·s})(3.0 \times 10^{8} \text{ m/s})(242 \text{ L})}$$

$$= 3.80 \times 10^{-5} \text{ einstein/L·s}$$

2. Calculate the rate constant for NDMA.
   a. The extinction coefficient of NDMA at 254 nm can be obtained from Table 6–18.

$$\varepsilon(\lambda_{254}) = 1974 \text{ L/mole·cm}$$

$$\varepsilon'(\lambda_{254}) = 2.303[\varepsilon(\lambda_{254})] = 2.303 \times 1974 = 4546 \text{ L/mole·cm}$$

   b. The quantum yield for NDMA can be obtained from Table 6–18.

$$\phi(\lambda_{\text{NDMA}}) = 0.3 \text{ mole/einstein}$$

   c. Compute k_{NDMA} using Eq. (6–66).

$$k_{\text{NDMA}} = \phi(\lambda_{\text{NDMA}})P_R \frac{\varepsilon'(\lambda_{\text{NDMA}})}{k'(\lambda)}$$

$$= (0.3 \text{ mole/einstein})(3.80 \times 10^{-5} \text{ einstein/L·s})\left[\frac{(4546 \text{ L/mole·cm})}{(0.01/\text{cm})}\right]$$

$$= 2.59 \text{ 1/s}$$

3. Calculate the flowrate, that can be treated per reactor.
   a. Calculate hydraulic detention time for the reactor.

$$\tau = \frac{n[(C_{\text{NDMA,o}}/C_{\text{NDMA,e}})^{1/n} - 1]}{k_{\text{NDMA}}} = \frac{3[(50/1)^{1/3} - 1]}{(2.59 \text{ 1/s})} = 3.11 \text{ s}$$

   b. Calculate the flowrate, that can be processed by one reactor.

$$Q = \frac{V}{\tau} = \frac{242 \text{ L}}{3.11 \text{ s}} = 77.7 \text{ L/s}$$

4. Determine the number of reactors needed to treat the full flow.
   a. The total flow to be treated is $1.9 \times 10^4\ \mathrm{m^3/d} = 219\ \mathrm{L/s}$.
   b. The number of reactors needed is (219 L/s) / (77.7 L/s) = 2.8 (use 3).
   c. The actual number of reactors needed will be greater than the computed value to compensate for lamp failure, fouling, and so that one or more reactors can be taken off line for lamp maintenance without interrupting the flow. It should be noted that the extra reactors will not be in continuous operation, but will only be used when needed or in a service rotation to reduce costs.
5. Calculate the EE/O for the photolysis process.

$$\mathrm{EE/O} = \frac{P}{Q \log\left(\dfrac{C_i}{C_f}\right)}$$

$$= \frac{(14.4\ \mathrm{kW})(10^3\ \mathrm{L/1\ m^3})}{(77.7\ \mathrm{L/s})\left\{\log\left[\dfrac{(50\ \mathrm{ng/L})}{(1\ \mathrm{ng/L})}\right]\right\}(3600\ \mathrm{s/h})} = 0.0303\ \mathrm{kWh/m^3}$$

The computed EE/O value is low compared to the typical range for ground and surface waters because of the high quality effluent from the RO process. Reverse osmosis removes or reduces many of the constituents that would interfere with photolysis of specific constituents and can produce effluent with low absorbance, improving the efficiency of the photolysis process.

6. Estimate the overall daily energy usage for the process.
   For the two operational reactors, the estimated energy usage is

$$3\ \text{reactors} \times 14.4\ \mathrm{kW} \times 24\ \mathrm{h/d} = 1037\ \mathrm{kWh/d}$$

Comment The photolysis reactors sized in the design example represent a minimum size and do not include correction factors for nonideal flow, variability in lamp output, and other inefficiencies. Bench and/or pilot studies are always required to determine actual design parameters.

Photolysis Process Limitations

The efficiency of the photolysis process depends, in part, on the characteristics of the water matrix and compounds targeted for degradation. For example, the extinction coefficient for residual organic matter varies over a wide range and may interfere with the photolysis of other compounds. In addition, the light energy input may be absorbed by the other constituents, there may be photon losses upon reflection off the reactor wall, and the precipitate that builds up on the exterior surface of sleeves that cover the lamps due to the elevated temperature will block light transmission. For some constituents, the performance of direct photolysis processes have been improved by the addition of hydrogen peroxide (Linden et al., 2004), as described in Sec. 6–8. To overcome the limitations associated with photolysis processes related to absorbance of UV energy by non-target constituents in water reuse, pretreatment using reverse osmosis should be used to remove most of the interfering compounds and improve the overall process performance. Pilot studies should be conducted to characterize the expected efficiency of photolysis and the rate and characteristics of fouling of the lamp sleeve.

6–10 CHEMICAL NEUTRALIZATION, SCALE CONTROL, AND STABILIZATION

The removal of excess acidity or alkalinity by treatment with a chemical of the opposite composition is termed *neutralization*. In general, all treated wastewaters with excessively low or high pH will require neutralization before they can be dispersed to the environment. Scaling control is required for nanofiltration and reverse osmosis treatment to control the formation of scale, which can severely impact performance. Chemical stabilization is often required for highly treated wastewaters to control their aggressiveness with respect to corrosion. These subjects are considered briefly below.

pH Adjustment

In a variety of wastewater treatment operations and processes, there is often a need for pH adjustment. Because a number of chemicals are available that can be used, the choice will depend on the suitability of a given chemical for a particular application and prevailing economics. General information on the chemicals used most commonly for pH adjustment is given in Table 6–19. Wastewater that is acidic can be neutralized with any number of

Table 6–19

Chemicals used most commonly for the control of pH (neutralization)

				Availability	
Chemical	**Formula**	**Molecular weight**	**Equiv weight**	**Form**	**Percent**
Chemicals used to raise pH					
Calcium carbonate	$CaCO_3$	100.0	50.0	Powder crushed	96 to 99
Calcium hydroxide (lime)	$Ca(OH)_2$	74.1	37.1	Powder granules	82 to 95
Calcium oxide	CaO	56.1	28.0	Lump, pebble, ground	90 to 98
Dolomitic hydrated lime	$[Ca(OH)_2]_{0.6}$ $[Mg(OH)_2]_{0.4}$	67.8	33.8	Powder	58 to 65
Dolomitic quicklime	$(CaO)_{0.6}(MgO)_{0.4}$	49.8	24.8	Lump, pebble, ground	55–58 CaO
Magnesium hydroxide	$Mg(OH)_2$	58.3	29.2	Powder	
Magnesium oxide	MgO	40.3	20.2	Powder, granules	99
Sodium bicarbonate	$NaHCO_3$	84.0	84.0	Powder, granules	99
Sodium carbonate (soda ash)	Na_2CO_3	106.0	53.0	Powder	99.2
Sodium hydroxide (caustic soda)	$NaOH$	40.0	40.0	Solid flake, ground flake, liquid	98
Chemicals used to lower pH					
Carbonic acid	H_2CO_3	62.0	31.0	Gas (CO_2)	
Hydrochloric acid	HCl	36.5	36.5	Liquid	27.9, 31.45, 35.2
Sulfuric acid	H_2SO_4	98.1	49.0	Liquid	77.7 (60° Be) 93.2 (66° Be)

[a] Adapted in part from Eckenfelder et al. (2009).

basic chemicals, as reported in Table 6–19. Sodium hydroxide (NaOH, also known as caustic soda) and sodium carbonate, although somewhat expensive, are convenient and are used widely by small plants or for treatment where small quantities are adequate. Lime, which is cheaper, but somewhat less convenient, is the most widely used chemical. Lime can be purchased as quicklime or slaked hydrated lime, high-calcium or dolomitic lime, and in several physical forms. Limestone and dolomitic limestone are cheaper, but less convenient to use and slower in reaction rate. Because they can become coated in certain waste treatment applications, their use is limited. Calcium and magnesium chemicals often form sludges that require disposal.

Alkaline wastes are less of a problem than acid wastes, but nevertheless often require treatment. If acidic waste streams are not available or are not adequate to neutralize alkaline wastes, sulfuric acid is commonly employed. In some treatment plants, carbon dioxide (CO_2) in the form of flue gas has been used to neutralize alkaline wastewaters as illustrated by the following reactions.

$$2OH^- + CO_2 \rightarrow CO_3^{2-} + H_2O \tag{6–69}$$

$$CO_3^{2-} + CO_2 + H_2O \rightarrow 2HCO_3^- \tag{6–70}$$

Based on the chlorine dose used for disinfection, the pH of the disinfected effluent will be lower than that allowed for reuse applications and for dispersal to the environment. In such cases, neutralization is controlled by automatic instruments using a feedback loop, and the final effluent pH is recorded. Depending on the sensitivity of the environment, two-stage neutralization may be required. The reagent chemicals can be fed automatically, in the form of solutions, slurries, or dry materials. If the reaction rate is slow, instrumentation and control design must take this factor into account.

Analysis of Scaling Potential

With the increasing use that is being made of nanofiltration, reverse osmosis, and electrodialysis in wastewater reuse applications, adjustment of the scaling characteristics of the effluent to be treated is important to avoid calcium carbonate and sulfate scale formation. Depending on the recovery rate, the concentration of salts can increase by a factor of up to ten within the treatment module. When such a salt concentration increase occurs, it is often possible to exceed the solubility product of calcium carbonate and other scale-forming compounds. The formation of scale within the treatment module will cause deterioration in the performance, ultimately leading to the failure of the membrane module.

Over the past 75 years, a number of theories of scale formation (principally calcium carbonate scale) has been advanced and applied. Generally, these theories have resulted in the formation of relative indexes that are used to assess whether a water is undersaturated, balanced, or supersaturated with respect to calcium carbonate. An undersaturated water will tend to dissolve an existing calcium carbonate film; a balanced water will neither dissolve or form a film; a supersatured water will tend to form a film. Representative indexes that are used to assess the potential for scale formation are presented in Table 6–20. It is interesting to note that most of the available indexes were developed for the water supply and petroleum industry. The Langelier saturation index and the Ryzner stability index, used commonly in the wastewater field, are discussed in greater detail below. It should be noted that a variety of computer programs are available that can be used to solve the index equations, many of which are quite complex, such as the

Table 6–20

Representative indexes used to assess the stability of water

Index	General form of relationship	Undersaturated systems	Saturated systems	Oversaturated systems
Aggressiveness Index (AI)[a,b]	AI $= \mathrm{pH} + \log_{10}[(\mathrm{Ca})(\mathrm{Alk})]$ Ca and Alk expressed as mg $CaCO_3$/L	AI < 10 highly aggressive	$10 <$ AI < 12 moderately aggressive	AI > 12 nonaggressive (i.e. . protective)
Calcium saturation index (CSI)[b]	CSI $= \mathrm{pH} + \mathrm{Tf} + \mathrm{Af} + \mathrm{Cf} - 12.1$ Where Tf = temperature, Af = alkalinity, and Cf = calcium factors taken from tables	CSI < -0.3 scale may form	CSI ~ 0 water is balanced	CSI > 0 water will not form scale
Calcium carbonate precipitation potential (CCPP)[c,b]	Complex computer based model the solution			
Driving force index (DFI)[d,b]	DFI $= (Ca^{2+})(CO_3^{2-})/K's \cdot 10^{10}$ Ca^{2+} and CO_3^{2-} expressed as mg $CaCO_3$/L	DFI < 1	DF $= 1$	DFI < 1
Langelier saturation index (LSI)[e,b]	LSI $= \mathrm{pH} - \mathrm{pH}_s$	LSI < 0	LSI $= 0$	LSI > 0 LI
Larson-Skold index[f]	$L - SI = (Cl^- + SO_4^{2-})/(HCO_3^- + CO_3^{2-})$ all expressed as equivalents per million (epm)	$>>0.6$ high corrosion rates expected	$>> 0.2$ but $<<0.6$ chlorides and sulfate may interfere with film formation	$<<0.2$ chlorides and sulfate will not interfere with film formation
Saturation level (SL)	$\mathrm{SL} = \dfrac{\gamma_{Ca}[Ca]\gamma_{CO_3}[CO_3]}{K_{sp}}$	<1	1	>1
Puckorius scaling index (PSI)[g,b]	PSI $= 2(\mathrm{pH}_s) - \mathrm{pH}_{eq}$	$<<6$ scale forming tendence	6 water is stable	$>>6 - 7$ scale should not form
Ryznar stability index (RSI)[h,b]	RSI $= 2\mathrm{pH}_s - \mathrm{pH}$	RSI > 6.8	$6.2 <$ RSI < 6.8	RI < 6.2
Momentary excess (ME)[i,b]	$[Ca - X][CO_3 - X]\ K_{spc}$ where X is the precipitation required to restore equilibrium	ME < 0	ME ~ 0	ME > 0

[a] Millette et al. (1980), [b]Temkar et al. (1990), [c]Standard Methods (2012), [d]McCauley (1960),
[e] Langelier (1936), [f]Larson and Skold (1958), [g]Puckorius and Brooke (1991), [h]Ryzner (1944), [i]Dye (1952).

calcium carbonate precipitation potential (Truesdell and Jones, 1973; Merrill and Sanks, 1977a, b and 1978; Ball and Nordstrom, 1991; WaterCycle®, 2012). Computer programs are also available that can be used to compare the results obtained with the various models (WaterCycle®, 2012).

Another approach to assess water chemistry involves the use of chemical equilibrium models, such as MINTEQA2 and MINEQL+. With these models, the chemical components and concentrations are selected and used as inputs to the model. Thermodynamic relationships are used to solve a system of equations that determine which species may form, what the resulting species concentration is expected to be, and what precipitates may form. Additional information of chemical equilibrium models is presented in Standard Methods (2012).

Langelier Saturation Index. The tendency to develop calcium carbonate ($CaCO_3$) scale during the advanced treatment of treated effluent can be approximated by calculating the Langelier Saturation Index (LSI) of the concentrate stream (Langelier, 1946).

$$\text{LSI} = \text{pH} - \text{pH}_s \tag{6–71}$$

where pH = measured pH in concentrate stream water sample
pH_s = saturation pH for calcium carbonate

The scaling criteria for the Langelier Saturation Index are

$\text{LSI} > 0$	Water is supersaturated with respect to calcium carbonate ($CaCO_3$) and scaling may occur.
$\text{LSI} < 0$	Water is undersaturated with respect to calcium carbonate. Undersaturated water has a tendency to remove existing calcium carbonate protective coatings in pipelines and equipment.
$\text{LSI} = 0$	Water is considered to be neutral (i.e., neither scale forming or scale removing).

It should be noted that undersatured water is also sometimes referred to as *corrosive*, but use of the term corrosive is incorrect, as the LSI index only applies to the presence or absence of a calcium carbonate scale. The diagram presented in Fig. 6–28 can also be used to estimate LSI.

Ryzner Stability Index. An alternative index known as the *stability index* was proposed by Ryzner (1944) and is used in a number of industrial applications. The Ryzner Stability Index (RSI) is given by the following expression:

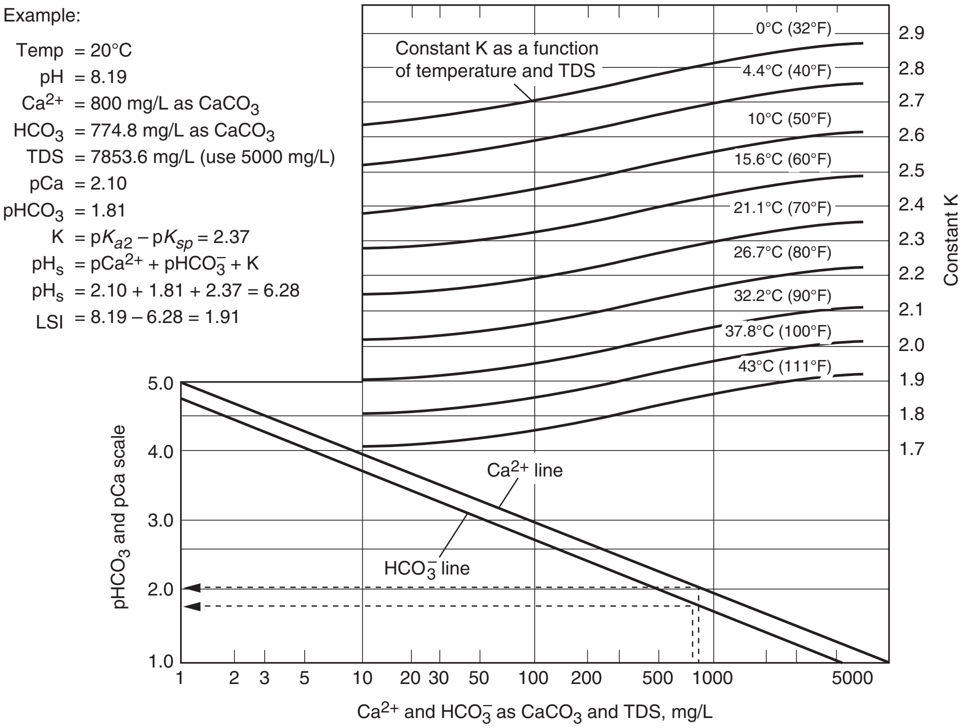
$$\text{RSI} = 2\text{pH}_s - \text{pH} \tag{6–72}$$

The scaling criteria for the Ryzner Index are as follows:

$\text{RSI} < 5.5$	Heavy scale will form
$5.5 < \text{RSI} < 6.2$	Scale will form
$6.2 < \text{RSI} < 6.8$	No difficulties
$6.8 < \text{RSI} < 8.5$	Water is aggressive
$\text{RSI} > 8.5$	Water is very aggressive

Figure 6–28

Chart for determining the value of the Langelier saturation index. (Adapted from DuPont Company, 1992.)



Application of Indexes. The saturation pH_s used in the above equations can be computed using the following expression:

$$pH_s = -\log\left(\frac{K_{a2}\gamma_{Ca^{2+}}[Ca^{2+}]\gamma_{HCO_3^-}[HCO_3^-]}{K_{sp}}\right) \tag{6–73}$$

where K_{a2} = equilibrium constant for the dissociation of bicarbonate
$\gamma_{Ca^{2+}}$ = activity coefficient for calcium
$[Ca^{2+}]$ = concentration of calcium, mole
$\gamma_{HCO_3^-}$ = activity coefficient for bicarbonate
$[HCO_3^-]$ = concentration of bicarbonate, mole
K_{sp} = solubility product constant for the dissociation of calcium carbonate

The activity coefficient can be estimated using Eq (2–10) given in Chap. 2.

$$\log \gamma = -0.5\,(Z_i)^2\left(\frac{\sqrt{I}}{1+\sqrt{I}} - 0.3I\right) \tag{2–10}$$

where Z_i = charge on ionic species
I = ionic strength

The ionic strength can be estimated using Eq (2–12) given in Chap. 2.

$$I = 2.5 \times 10^{5-} \times \text{TDS} \tag{2–12}$$

Table 6–21
Carbonate equilibrium constants as function of temperature[a]

Temperature, °C	Equilibrium constant[b]		
	$K_{a1} \times 10^7$	$K_{a2} \times 10^{11}$	$K_{sp} \times 10^9$
5	3.020	2.754	8.128
10	3.467	3.236	7.080
15	3.802	3.715	6.02
20	4.169	4.169	5.248
25	4.467	4.477	4.571
40	5.012	6.026	3.090

[a] Adapted from Snoeyink and Jenkins (1980) and Pankow (2012).
[b] The reported values have been multiplied by the indicated exponents. Thus, the value K_2 at 20°C is equal to 4.169×10^{-11}.

The saturation pH_s for calcium carbonate ($CaCo_3$) solubility in the pH range from 6.5 to 9.0 is given by

$$pH_s = pK_{a2} - pK_{sp} + p[Ca^{2+}] + p[HCO_3^-] - \log \gamma_{Ca^{2+}} - \log \gamma_{HCO_3} \tag{6–74}$$

where pK_{a2} = negative logarithm of the equilibrium constant for the dissociation of bicarbonate
pK_{sp} = negative logarithm of equilibrium constant for the dissociation of calcium carbonate
$p[Ca^{2+}]$ = negative logarithm of calcium concentration
$p[HCO_3^-]$ = negative logarithm of bicarbonate concentration

Values of K_1, K_2, and K_{sp} for the carbonate system are given in Table 6–21 as a function of temperature. The application of these equations is illustrated in Example 6–9.

EXAMPLE 6–9 **Analysis of Scaling Potential** Estimate the scaling potential using both the Langelier and Ryzner indexes for a treated wastewater with the following chemical characteristics:

	Concentration	
Constituent	g/m³	mole/L
Ca^{2+}	5	0.125×10^{-3}
HCO_3^-	10	0.164×10^{-3}
TDS	20	
pH	7.7	

Solution

1. Estimate the ionic strength of the treated water using Eq. (2–11).

$$I = 2.5 \times 10^{-5} \times \text{TDS}$$
$$I = 2.5 \times 10^{-5} \times 20 = 50 \times 10^{-5}$$

2. Determine the activity coefficients for calcium and bicarbonate using Eq. (2–9).
   a. For calcium

$$\log \gamma_{\mathrm{Ca}^{2+}} = -0.5(Z_i)^2\left(\frac{\sqrt{I}}{1+\sqrt{I}} - 0.3I\right)$$

$$= -0.5(2)^2\left[\frac{\sqrt{50 \times 10^{-5}}}{1+\sqrt{50 \times 10^{-5}}} - 0.3(50 \times 10^{-5})\right] = -0.0434$$

$$\gamma \mathrm{Ca}^{2+} = 0.905$$

   b. For bicarbonate

$$\log \gamma_{\mathrm{HCO}_3^-} = -0.5(Z_i)^2\left(\frac{\sqrt{I}}{1+\sqrt{I}} - 0.3I\right)$$

$$= -0.5(1)^2\left[\frac{\sqrt{50 \times 10^{-5}}}{1+\sqrt{50 \times 10^{-5}}} - 0.3(50 \times 10^{-5})\right] = -0.0109$$

$$\gamma_{\mathrm{HCO}_3^-} = 0.975$$

3. Determine the saturation pH_s using Eq. (6–46).

$$\mathrm{pH}_s = -\log\left(\frac{K_{a2}\gamma_{\mathrm{Ca}^{2+}}[\mathrm{Ca}^{2+}]\gamma_{\mathrm{HCO}_3^-}[\mathrm{HCO}_3^-]}{K_{sp}}\right)$$

$$\mathrm{pH}_s = -\log\left[\frac{(4.17 \times 10^{-11})(0.905)(0.125 \times 10^{-3})(0.975)(0.164 \times 10^{-3})}{5.25 \times 10^{-9}}\right]$$

$$\mathrm{pH}_s = -\log(1.43 \times 10^{-10}) = 9.84$$

4. Determine the Langelier and Ryzner indexes.
   a. Langelier Saturation Index

   $\mathrm{LSI} = \mathrm{pH} - \mathrm{pH}_s = 7.7 - 9.84 = -2.14$

   $\mathrm{LSI} < 0$ (Water is undersaturated with respect to calcium carbonate)

   b. Ryzner Index

   $\mathrm{RSI} = 2\mathrm{pH}_s - \mathrm{pH} = 2(9.84) - 7.7 = 11.98$

   $(\mathrm{RSI} = 11.98) < 8.5$ (Water is very aggressive)

Comment Although both indexes are used, the Langelier index is used most commonly in the water and wastewater field, whereas the Ryzner index is used most commonly in industrial applications.

Scale Control

Usually, $\mathrm{CaCO_3}$ scale control can be achieved using one or more of the following methods:

- Acidifying to reduce pH and alkalinity
- Reducing calcium concentration by ion exchange or lime softening
- Adding a scale inhibitor chemical (antiscalant) to increase the apparent solubility of $\mathrm{CaCO_3}$ in the concentrate stream (see Chap. 11)
- Lowering the RO product recovery rate (see Chap. 11)

Because it is not possible to predict a priori the value of pH in water treated with reverse osmosis, it is usually necessary to conduct bench and/or pilot scale studies using the same modules that will be used in the full scale installation.

Stabilization

Wastewater effluent that is demineralized with reverse osmosis will generally require pH and calcium carbonate adjustment (stabilization) to prevent metallic corrosion, due to the contact of the demineralized water with metallic pipes and equipment. Corrosion occurs because material from the solid is removed (solubilized) to satisfy the various solubility products. Demineralized water typically is stabilized by adding lime to adjust the LSI, using the procedure outlined above.

6–11 CHEMICAL STORAGE, FEEDING, PIPING, AND CONTROL SYSTEMS

The design of chemical treatment operations involves not only the sizing of the various unit operations and processes but also the necessary appurtenances. Because of the corrosive nature of many of the chemicals used and the different forms in which they are available, special attention must be given to the design of chemical storage, feeding, piping, and control systems. The following discussion is intended to serve as an introduction to this subject.

In domestic wastewater treatment systems, the chemicals employed can be in a solid, liquid, or gaseous form. Coagulants in the dry solid form generally are converted to solution or slurry form prior to introduction into the wastewater. Coagulants in the liquid form are usually delivered to the plant in a concentrated form and have to be diluted prior to introduction into the wastewater. Chemicals in the gas form (generally stored as a liquid), typically used for disinfection purposes, are either dissolved in water before injection or are injected directly into the wastewater. The types of chemical feed systems used for these chemicals are termed *dry, liquid* (also known as wet), or *gas* feed. The various types of feeders are classified on Fig. 6–29. Chemical feeders are generally designed to be: (1) proportioning, feeding chemical in proportion to the influent wastewater flowrate and (2) constant feed, designed to deliver chemical at a fixed rate regardless of the influent flowrate.

Chemical Storage and Handling

General information on the handling, storage, and feeding requirements for various chemicals is presented in Table 6–22. The specific storage facilities required will depend on the

Figure 6–29

Classification of chemical feed systems.

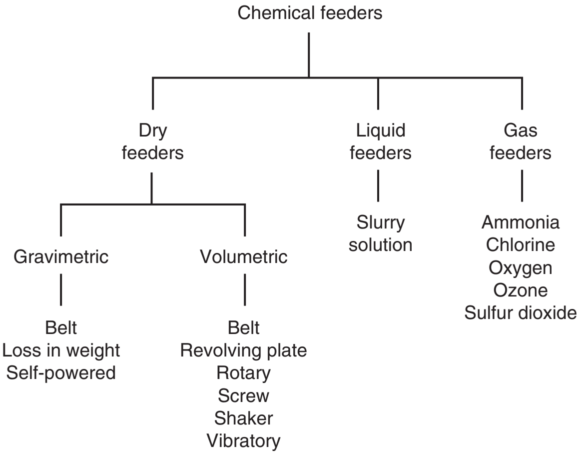


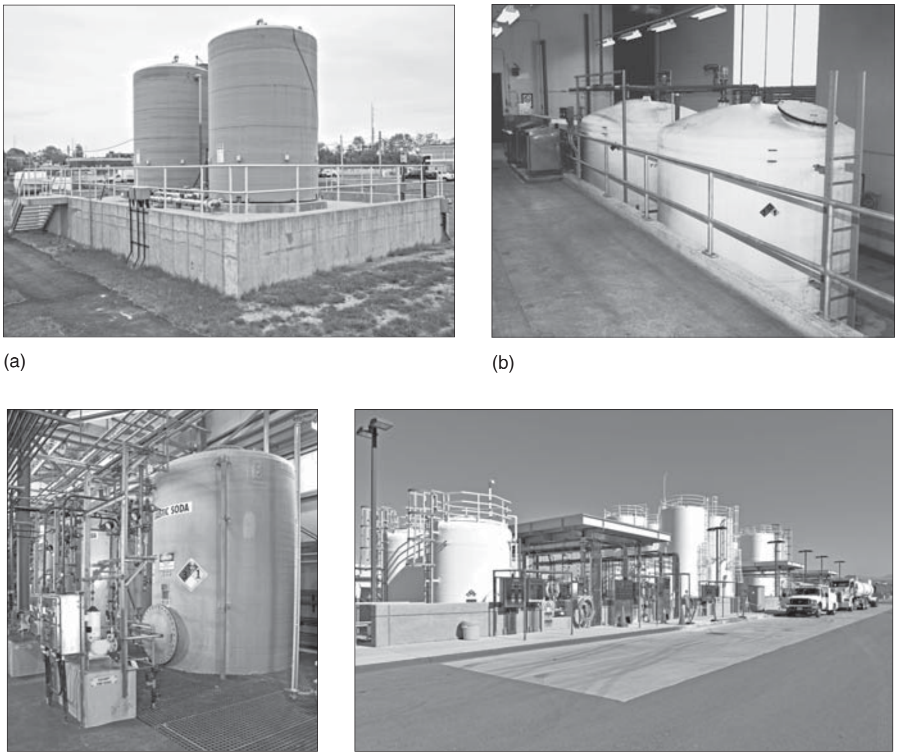
Table 6–22

Handling, storage and feeding requirements for various chemicals used in wastewater treatment[a]

Chemical	Formula	Shipping form	Feeding form	Feeding type	Accessory equipment	Handling materials
Chemicals used for precipitation						
Aluminum sulfate	$Al_2(SO_4)_3 \cdot 18H_2O$	Lump, ground or powdered solutions	Liquid	Metering pump	Slurry tank, slaker	Iron, steel
Aluminum chloride	$AlCl_3$	Liquid	Liquid	Metering pump	Storage tank	Hastelloy B, plastic
Calcium hydroxide (lime)	$Ca(OH)_2$	Bags, barrels, and bulk	Liquid	Metering pump	Slurry tank	Iron, steel
Ferric chloride	$FeCl_3$	Bags, carboys, and bulk	Liquid	Metering pump	Slurry tank	Iron, steel
Ferric sulfate	$Fe_2(SO_4)_3$	Bags, barrels, and bulk	Liquid	Metering pump	Slurry tank	Iron, steel
Copperas	$FeSO_4 \cdot 7H_2O$		Liquid	Metering pump	Slurry tank	Iron, steel
Chemicals used for neutralization						
Calcium carbonate	$CaCO_3$	Bags, drums or bulk	Slurry, dry slurry in fixed beds	Metering pump, Volumetric pump	Slurry tank	Iron, steel
Calcium oxide	CaO	Bags (22.5 kg), barrels, or bulk	Dry or slurry, slaked to $Ca(OH)_2$	Metering pump	Slurry tank, slaker	Iron, steel, plastic, rubber hose
Sodium bicarbonate	$NaHCO_3$	Bags or drums	Dry or slurry	Metering pump	Dissolving tank	Iron, steel, plastic, rubber hose
Sodium carbonate	Na_2CO_3	Bags (45.5 kg), bulk	Dry or slurry	Metering pump	Dissolving tank	Iron, steel, plastic, rubber hose
Sodium hydroxide	$NaOH$	Drum (45.5, 204.5, 367.5 kg)	Dry or slurry	Metering pump	Solution tank	Iron, steel, plastic, rubber hose
Carbonic acid	H_2C0_3		Gas (CO_2)	Metering pump		
Hydrochloric acid	HCl	Barrels, drums, bulk	Liquid	Metering pump	Dilution tank	Hastelloy A, plastic, rubber[b]
Sulfuric acid	H_2S0_4	Carboys, drums, and bulk	Liquid	Metering pump		Iron, steel, plastic, rubber hose

[a] Adapted in part from Eckenfelder et al. (2009).

[b] Selected types.



(a) (b) (c) (d)

Figure 6–30

Typical chemical storage facilities: (a) outdoor facility at a small wastewater treatment plant. The chemical storage tanks are located within a containment structure in case of a chemical spill (b) chemical storage tanks located indoors in a containment area, (c) chemical storage tanks located indoors on grated floor over chemical containment area below, and (d) large chemical storage facilities located within chemical containment structures. The chemical feed station is located between storage tanks.

form in which the chemical is available locally. For small treatment plants the available forms are usually limited. Typical indoor and outdoor storage facilities for chemicals at wastewater treatment plants are shown on Fig. 6–30. In all cases some form of containment is provided to limit the spread of any chemical should any type of leak develop.

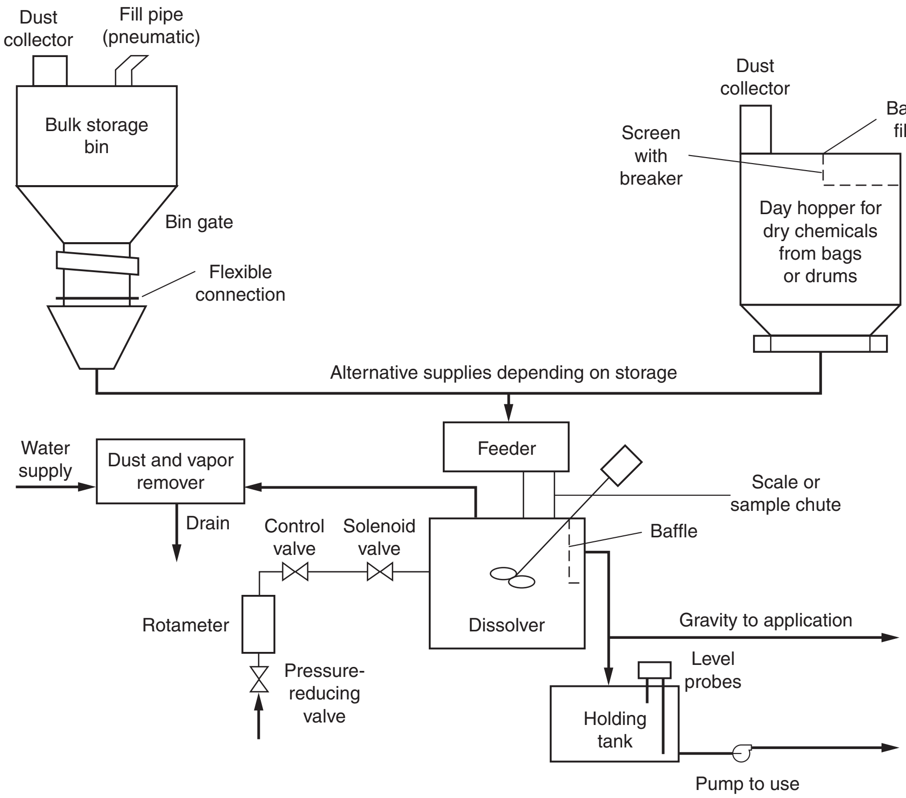
Dry Chemical Feed Systems

A dry chemical feed system typically consists of a storage hopper, dry chemical feeder, a dissolving tank, and a pumped or gravity distribution system (see Figs. 6–31 and 6–32). The units are sized according to the volume of wastewater, treatment rate, and optimum length of time for chemical feeding and dissolving. Hoppers used with powdered chemicals that are compressible and can form an arch such as lime are equipped with positive agitators and a dust collection system. Dry chemical feeders are either of the volumetric or gravimetric type. In the volumetric type the volume of the dry chemical fed is measured, whereas in the gravimetric type the weight of chemical fed is measured. A brief description and illustration of the chemical feeders used most commonly is presented in Table 6–23.

With a dry feed system, the dissolving operation is critical. The capacity of the dissolving tank is based on the detention time, which is directly related to the wettability or rate of solution of the chemical. When the water supply is controlled for the purpose of forming a constant strength solution, mechanical mixers should be used. Depending on the flow pattern within the mixing tank, it may be necessary to add baffles for effective mixing. In smaller mixing tanks, the mixer can be set at an angle to avoid the use of baffles. Solutions or slurries are often stored after dissolving and discharged to the application point at metered rates by chemical feed pumps.

Figure 6–31

Schematic of typical dry chemical-feed system.



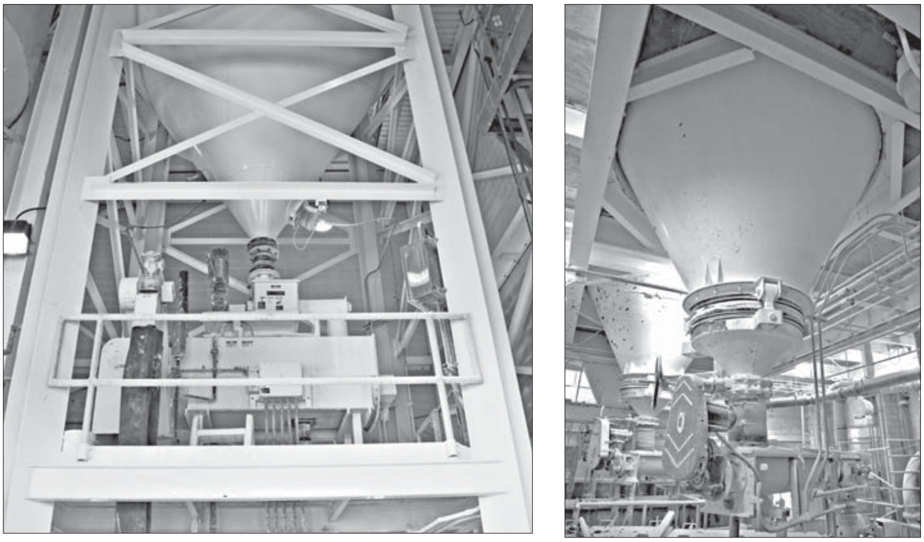
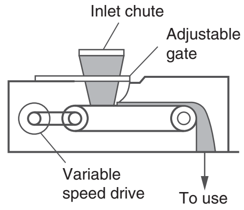
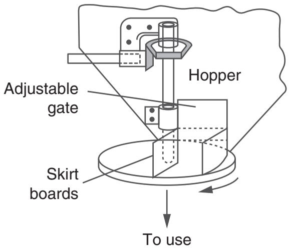
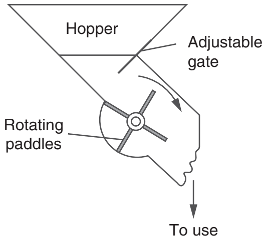
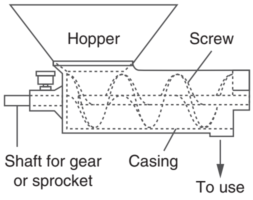
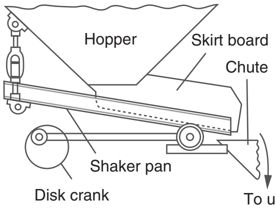


Figure 6–32

Typical dry chemical-feed system (a) single storage hopper and (b) two chemical storage hoppers coupled through flexible connectors to individual chemical feeders and dissolvers as illustrated schematically on Fig. 6–31.

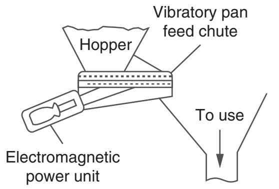
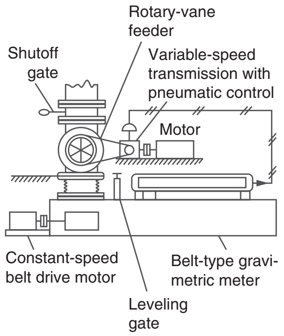
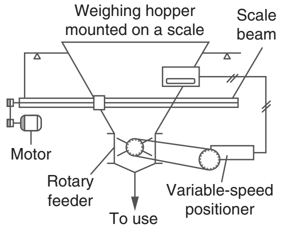
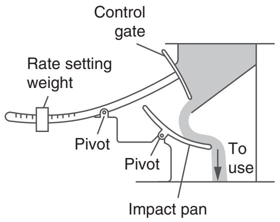
Table 6–23

Typical characteristics of chemical feeders[a]

Type of feeder	Description	Illustration
Volumetric		
Conveyor belt	Consists of a belt located below a hopper. The feed rate is adjusted by varying the speed of the belt.	
Revolving plate	Consists of a rotating plate below the storage hopper. As the plate is rotated, material to be feed is drawn from the hopper. The amount of material feed is controlled by the rate of rotation.	
Rotary	Consists of rotating shaft with vanes that form pockets. The amount of material feed is controlled by the rate of rotation.	
Screw	Consists of a variable pitch screw mounted below a feed hopper. The amount of material feed is controlled by the rate of rotation of the screw.	
Shaker	Consists of a shaker pan mounted below a storage hopper. As the pan oscillates, the material to be feed is moved forward and dropped into the feed chute.	

(*continued*)

Table 6–23 (*Continued*)

Type of feeder	Description	Illustration
Vibratory	Consists of vibrating pan or chute positioned below a chemical storage hopper. The pan or chute which vibrates back and forth by the oscillating electromagnetic driver delivers the material to be feed forward. The amount of material feed can be controlled by adjusting the rate of oscillation.	
Gravimetric		
Belt	Consists of volumetric feeder that transfers the material to be feed from the hopper to the weigh belt. The signal generated from the weigh belt is used to control the volumetric feeder.	
Loss in weight	Consists of a feed hopper mounted on scale and a chemical feeder. The chemical feed rate can be controlled with a screw or vibratory feeder. The feed rate is controlled by the loss in weight measured by the scale.	
Self-powered	Consists of a counter balanced control gate mounted below a storage hopper. The weight of the material in the hopper is counterbalanced by the setting on the beam balance. The rate at which material is feed is controlled by the impact pan. Although not accurate, this device does not require any power source.	

[a] Adapted from Liptak (1974).

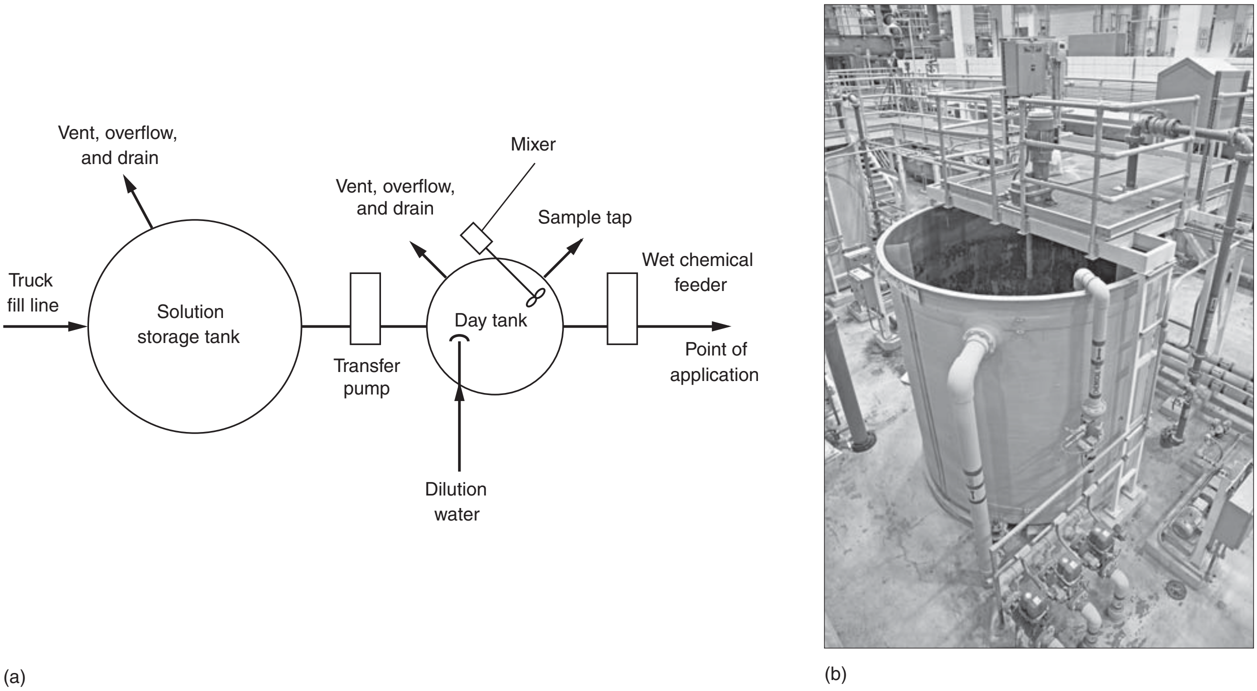


Figure 6–33

Typical liquid chemical-feed system (a) schematic and (b) view of large day tank with mixer. Also, note the internal deflection baffle to enhance mixing (see Fig. 5–14).

Liquid Chemical Feed Systems

Liquid chemical feed systems typically include a solution storage tank, transfer pump, day tank for diluting the concentrated solution, and chemical feed pump for distribution to the application point (see Figs. 6–33 and 6–34). In general, liquid feed systems provide for better initial contact and dispersion of the chemical and the wastewater. In systems where the liquid chemical does not require dilution, the chemical feed pumps draw liquid directly from the solution storage tank. The storage tank is sized based upon the stability of the chemical, feed rate requirements, regulatory requirements, delivery constraints (cost, size of tank truck, etc.), and availability of the supply. Solution feed pumps are usually of the positive displacement type for accurate metering of the chemical feed. Spill or secondary containment is usually required around the storage tanks; chemicals that are highly reactive should not be stored next to each other.

Gas Chemical Feed Systems

Chemicals that are used as a gas include ammonia, chlorine, oxygen, ozone, and sulfur dioxide. Gas feed systems are used mostly for feeding chemicals used for disinfection and dechlorination (see Chap. 12). Chlorine, a commonly used chemical for disinfection, is often supplied in a liquid form within the storage container and evaporates continuously as the gas is drawn from the headspace above the liquid in the storage container. Feed systems for disinfection chemicals are illustrated in Chap. 12 which deals with disinfection.

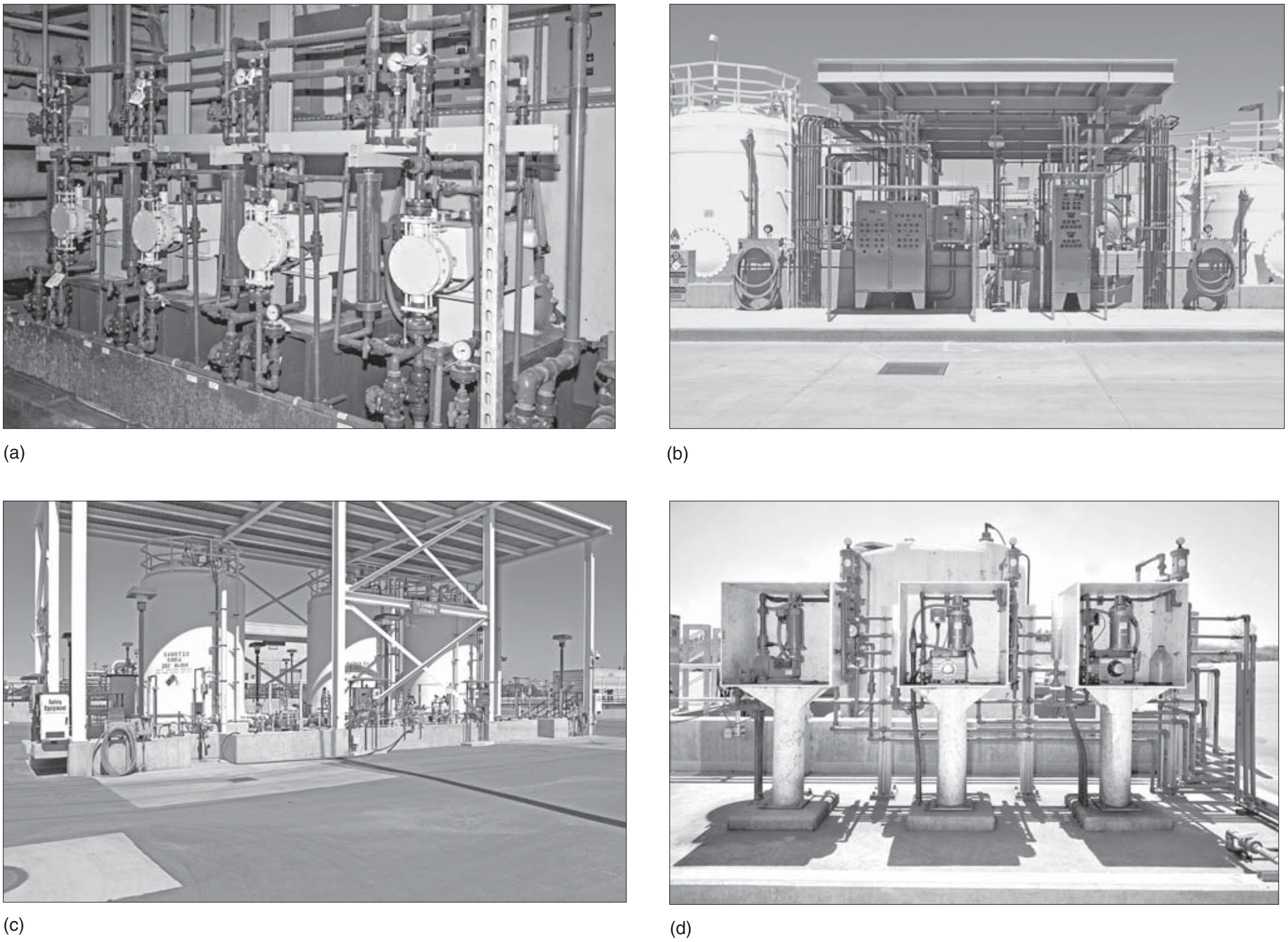

(a) (b) (c) (d)

Figure 6–34

Typical liquid chemical-feed systems (a) indoor chlorine metering pumps and (b), (c) and (d) typical outdoor shaded chemical feed stations.

Initial Chemical Mixing

Perhaps the least appreciated fact about chemical addition is the importance of both the initial and uniform mixing of the chemical with the wastewater to be treated. The optimal time for mixing can, as discussed in this section, vary from a fraction of a second to several seconds or more. Because of the difficulties in achieving rapid mixing in large treatment plants with a single mixing device, the use of multiple mixing devices is recommended. The particular mixing device selected for a given application must be based on a consideration of the reaction times and operative mechanisms for the chemicals that are being used. Typical mixing times for the chemicals used in wastewater treatment facilities are reported in Table 6–24. Typical devices used for mixing chemicals in wastewater treatment plants are discussed in Chap. 5. Approximate mixing times achievable with various mixing devices may be found in Table 5–9 in Chap. 5.

Table 6–24

Typical mixing times for various chemicals used in wastewater treatment facilities

Chemical	Applications	Recommended mixing times, s
Alum, Al^{3+}; Ferric chloride, Fe^{3+}	Coagulation of colloidal particles	<1
Alum, Al^{3+}; Ferric chloride, Fe^{3+}	Sweep floc precipitation	1–10
Lime, $Ca(OH)_2$	Chemical precipitation	10–30
Chlorine, Cl_2	Chemical disinfection	<1
Chloramine, NH_2Cl	Chemical disinfection	5–10
Cationic polymers	Destabilization of colloidal particles	<1
Anionic polymers	Particle bridging	1–10
Polymers, nonionic	Filter aids	1–10

PROBLEMS AND DISCUSSION TOPICS

6–1 To aid sedimentation in the primary settling tank, 15, 25, 40, or 60 g/m^3 value to be selected by instructor) of ferrous sulfate ($FeSO_4 \cdot 7H_2O$) is added to the wastewater. Determine the minimum alkalinity required to react initially with the ferrous sulfate. How many grams of lime should be added as CaO to react with $Fe(HCO_3)_2$ and the dissolved oxygen in the wastewater to form insoluble $Fe(OH)_3$?

6–2 Iron sulfate in the form ($FeSO_4 \cdot 7H_2O$) is to be added at a rate of 15, 25, 30, or 40 kg/1000 m^3 (value to be selected by instructor) to a wastewater to improve the efficiency of an existing primary sedimentation tank. Assuming sufficient alkalinity is present as $Ca(HCO_3)_2$ determine:

(a) How many kg of lime should be added as CaO to complete the reaction,

(b) The concentration of oxygen needed in the wastewater to oxidize the ferrous hydroxide that is formed,

(c) The mass of sludge that will result per 1000 m^3 of wastewater, and

(d) The amount (kg) of alum needed to obtain the same quantity of sludge as in Part (c), assuming $Al(OH)_3$ is the precipitate that is formed.

6–3 Assume that 40, 45, 50, and 55 kg (value to be selected by instructor) of (a) alum (mole wt 666.5) and (b) ferrous sulfate and lime as $Ca(OH)_2$ is added per 4000 m^3 of wastewater. Also assume that all insoluble and very slightly soluble products of the reactions, with the exception of 15 g/m^3 $CaCO_3$, are precipitated as sludge. How many kg of sludge/1000 m^3 will result in each case?

6–4 Raw wastewater is to be treated chemically for the removal of total suspended solids and phosphorus through coagulation, flocculation, and sedimentation. The wastewater characteristics are as follows: $Q = 0.75$ m^3/s; orthophosphate $= 10$ g/m^3 as P; alkalinity $= 200$ g/m^3 expressed as $CaCO_3$ [essentially all due to the presence of $Ca(HCO_3)_2$]; total TSS $= 220$ g/m^3.

a. Determine the sludge production in kg dry wt/d and m^3/d under the following conditions: (1) Alum [$Al_2(SO_4)_3 \cdot 14.3\ H_2O$] dosage of 120, 130, 140, or 150 g/m^3 (to be selected by instructor); (2) 100 percent removal of orthophosphate as insoluble $AlPO_4$; (3) 95 percent removal of original TSS; (4) all alum not required for reaction with phosphate reacts with alkalinity to form $Al(OH)_3$, which is 100 percent removed; (5) wet sludge has a water content of 93 percent and a specific gravity of 1.04.

b. Determine the sludge production in a kg dry wt/d and m^3/d under the following conditions: (1) Lime [$Ca(OH)_2$] dosage of 450 g/m^3 to give pH of approximately 11.2; (2) 100 percent removal of orthophosphate as insoluble hydroxylapatite [$Ca_{10}(PO_4)_6(OH)_2$]; (3) 95 percent removal of original TSS; (4) added lime (i) reacts with phosphate, (ii) reacts with all alkalinity to form $CaCO_3$, 20 g/m^3 of $CaCO_3$ is soluble and remains in solution and the rest is 100 percent removed, and (iii) remainder stays in solution; (5) wet sludge has a water content of 92 percent and a specific gravity of 1.05.

c. Determine the net increase in calcium hardness in g/m^3 as $CaCO^3$ for the treatment specified in part b.

6–5 Verify the correctness of one of the plots given on Fig. 6–8, using the equations given below the figures.

6–6 Obtain equilibrium values from the literature and verify one of the solubility curves given on Fig. 6–21.

6–7 Obtain equilibrium values from the literature and verify one of the solubility curves given on Fig. 6–22.

6–8 Using the potentials for the following two half reactions given in the Table 6–12, estimate the equilibrium constant for the ionization of water.

$$2H_2O + 2e^- \rightarrow H_2 + 2OH^-$$

$$2H^+ + 2e^- \rightarrow H_2$$

6–9 Using the half reactions given in Table 6–12, evaluate the feasibility of the following reaction:

$$2Fe^{2+} + 2H_2O_2 \rightleftarrows 2Fe^{3+} + 2H_2O$$

6–10 Using the half reactions given in Table 6–12, evaluate the feasibility of the following reaction:

$$2Fe^{2+} + Cl_2 \rightleftarrows 2Fe^{3+} + 2Cl^-$$

6–11 Using the half reactions given in Table 6–12, evaluate the feasibility of the following reaction:

$$H_2S + Cl_2 \rightleftarrows S + 2HCl$$

6–12 Using the half reactions given in Table 6–12, evaluate the feasibility of the following reaction:

$$H_2S + O_3 \rightleftarrows S + O_2 + H_2O$$

6–13 Estimate the scaling potential of one of the treated wastewater samples (to be selected by instructor) using both the Langelier and Ryzner indexes. Use a temperature of 20°C.

		Wastewater sample			
Constituent	Unit	1[a]	2	3	4
Ca^{2+}	mg/L as $CaCo_3$	5	12	245	15
HCO_3^-	mg/L as $CaCO_3$	7	9	200	16
TDS	mg/L	30	275	600	500
pH	unitless	6.5	8.0	6.9	6.5

[a] Typical values are for runoff from snow melt (adapted from Benefield et al. (1982).

6–14 Estimate the scaling potential of one of the treated wastewater samples (to be selected by instructor) using both the Langelier and Ryzner indexes. The measured pH values for the four samples are 1 = 7.2, 2 = 6.9, 3 = 7.3, and 4 = 6.8.

	Concentration, mg/L					Concentration, mg/L			
Cation	1	2	3	4	Anion	1	2	3	4
Ca^{2+}	121.3	64.0	42.1	44.0	HCO_3^-	280	96.0	158.7	91.5
Mg^{2+}	36.2	15.1	14.6	25.2	SO_4^{2-}	116	80	48.0	57.6
Na^+	8.1	20.5	46.0	4.6	Cl^-	61	17.3	63.8	17.7
K^+	12	10.0	11.7		NO_3^-	15.6	5		
					H_2CO_3				8.8
					CO_3^{2-}			12.0	

6–15 Review the literature concerning the Langelier (1946) and Ryzner (1944) indexes and discuss the difference in approach to scaling used in the development of these two indexes.

6–16 Determine the hydroxyl radical concentration required to remove each of the following compounds using an advanced oxidation process with a contact time of 10 s. Comment on the feasibility of removing each of the compounds under the given conditions.

	Concentration, μg/L			
	Water 1		Water 2	
Compound	Influent	Effluent	Influent	Effluent
Clorobenzene	100	5	120	7
Chloroethene	100	5	150	5
TCE	100	5	180	10
Toluene	100	5	200	15

6–17 Design an advanced oxidation process to achieve a 95 percent reduction of one of the following compounds for a flowrate of 3800 m^3/d, specify reactor dimensions and hydroxyl radical concentration required for the process.

Compound	Initial concentration, μg/L
1	25
2	10
3	100
4	75

6–18 A water reclamation plant produces 1×10^5 m^3/d of effluent containing 100 ng/L of NDMA. Determine the number of photolysis reactors needed to reduce the NDMA concentration of the RO effluent to 10 ng/L prior to indirect potable reuse using absorptivity values of $k'(\lambda) = 0.01$, 0.05, and 0.1 cm^{-1} (measured at a wavelength of 254 nm). The photolysis reactors under evaluation are 0.5 m in diameter and 1.5 m long with an effective water volume of 250 L. Each reactor has 25 lamps rated at 500 W per lamp and an output efficiency of 30 percent at 254 nm. Assume that the reactors operate as four mixed tanks in series and neglect all other losses, lamp fouling, and process inefficiencies. Calculate EE/O and daily energy usage for the photolysis process. Comment on the importance of absorptivity and recommend an appropriate pretreatment process.

6–19 Estimate the electricity cost (based on the current price of electricity) to treat a flowrate of 3800 m^3/d with a NDMA concentration of 100 mg/m^3 using a photolysis unit.

6–20 For each compound in the following list, which of the treatment methods discussed in this chapter, if any, are suitable for use to reduce the concentration from 100 to 10 mg/L?

Benzene
Chloroform
Dieldrin
Heptachlor
N-Nitrosodimethylamine
Trichloroethylene (TCE)
Vinyl chloride

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7

Fundamentals of Biological Treatment

WORKING TERMINOLOGY

Term	**Definition**
Acetogenesis	Biological conversion of volatile acids to acetic acid and hydrogen.
Activated sludge process	Biological treatment process that involves the conversion of organic matter and/or other constituents in the wastewater to gases and cell tissue by a large mass of aerobic microorganisms maintained in suspension by mixing and aeration. The microorganisms form flocculent particles that are separated from the process effluent in a sedimentation tank (clarifier) or by membranes and are returned subsequently to the aeration process or wasted.
Aerobic (oxic) process	Biological treatment process that occur in the presence of free dissolved oxygen; oxygen is consumed by aerobic microorganisms to drive metabolic reactions.
Anaerobic fermentation	See fermentation.
Anaerobic process	Biological treatment process that occur in the absence of free dissolved oxygen and oxidized compounds.
Anammox process	An anaerobic biological treatment process in which nitrite is used by specialized planctomycete bacteria as the electron acceptor to convert ammonia and nitrite to nitrogen gas primarily.
Anoxic process	Biological treatment process that occurs in the absence of free dissolved oxygen where oxidized compounds such as nitrate and nitrite are used to drive metabolic reactions; denitrification is an example of an anoxic process.
Attached-growth process (also known as fixed-film processes)	Biological treatment process in which the microorganisms responsible for the conversion of organic matter or other constituents in the wastewater to gases and cell tissue are attached to some inert medium such as rocks, slag or specially designed ceramic or plastic materials.

Term	Definition
Bioenergetics	The study of energy transformations that take place in living organisms.
Biological nutrient removal (BNR)	The term applied to the removal of nitrogen and phosphorus in biological treatment processes.
Biofilm	An accumulation of biological growth on the surface of an object such as attached-growth media.
Biomass	The mass of microorganisms in a biological treatment process.
Carbonaceous BOD removal	Biological conversion of the carbonaceous organic matter in wastewater to cell tissue and various gaseous end products. In the conversion, it is assumed that the nitrogen present in the various compounds is converted to ammonia.
Combined aerobic/ anoxic/anaerobic processes	Various combination of aerobic, anoxic and anaerobic biological treatment processes grouped together to achieve biological nitrogen and phosphorus removal.
Denitritation	Biological reduction of nitrite to nitrogen gas and other gaseous end products containing nitrogen.
Denitrification	Biological reduction of nitrate to nitrogen gas and other gaseous end products containing nitrogen.
Enhanced biological phosphorus removal	The term applied to phosphorus removal by certain bacteria with phosphorus storage capabilities greater than normally found in biological wastewater treatment.
Energetics	The scientific study of energy flows and transformations.
Facultative processes	Biological conversion processes which can occur in the presence or absence of molecular oxygen.
Fermentation (also referred to as acidogenesis)	The conversion of organic matter to carbon dioxide and other low molecular weight compounds.
Greenhouse gas	Gaseous compounds that contribute to global warming. Methane and nitrous oxide are potent greenhouse gases that can be produced in biological treatment processes.
Hybrid process	Term used to describe combined suspended and attached growth biological process.
Methanogenesis	Biological conversion of acetic acid or hydrogen and carbon dioxide to methane.
Nitratation	Biological oxidation of nitrite to nitrate.
Nitritation	Biological oxidation of ammonia to nitrite.
Nitrification	The two-step biological oxidation of nitrogen (mostly in the form of ammonia) to nitrite and then to nitrate.
Substrate	Wastewater or solids constituents used to promote biological growth.
Suspended growth process	Biological treatment process in which the microorganisms responsible for the conversion of the organic matter or other constituents in the wastewater to gases and cell tissues are maintained in suspension within the liquid.
Yield	The amount of biological solids produced relative to the amount of substrate removed.

With proper analysis and environmental control, almost all wastewaters containing biodegradable constituents can be treated biologically. Therefore, it is essential that the environmental engineer understand the characteristics of each biological process to ensure that the proper environment is produced and controlled effectively. The principal purposes of this chapter are (1) to provide fundamental background information on the microorganisms used to treat wastewater and (2) to consider the application of biological process fundamentals for the biological treatment of wastewater. The information presented in this chapter provides the necessary background material needed for the design of biological treatment processes discussed in Chaps. 8 through 10. For ease of computation, constituent concentrations in this chapter and in Chaps. 8, 9, and 10 are expressed in g/m^3 instead of mg/L because flowrate is given in units of m^3/s or m^3/d.

The fundamentals of biological treatment introduced in the first seven sections of this chapter include (1) an overview of biological wastewater treatment, (2) the composition and classification of the microorganisms used for wastewater treatment, (3) an introduction to important aspects of microbial metabolism, (4) bacterial growth and energetics, (5) microbial growth kinetics, (6) modeling suspended growth treatment processes, and (7) modeling attached-growth treatment processes. Following the presentation of fundamentals, the remaining eight sections deal with an introduction to the general classes of biological processes used for the treatment of wastewater. The material in these sections will also serve as an introduction to the material presented in Chaps. 8, 9 and 10. Because the focus of this book is on quantifiable treatment processes, the role of algae in wastewater treatment, principally in lagoon systems, is not considered. Information on algae is available in publications from the U.S. EPA, WEF, and several textbooks. Similarly, constructed wetlands are not considered because of space constraints and the fact that there are numerous publications including several books devoted solely to the analysis, design, and implementation of these systems.

7–1 OVERVIEW OF BIOLOGICAL WASTEWATER TREATMENT

The objectives of biological treatment, some useful definitions, the role of microorganisms in the biological treatment of wastewater, and biological processes used for wastewater treatment are introduced in this section to provide a perspective for the material to be presented in this chapter.

Objectives of Biological Treatment

The overall objectives of the biological treatment of domestic wastewater are to (1) transform (i.e., oxidize) dissolved and particulate biodegradable constituents into acceptable end products, (2) capture and incorporate suspended and nonsettleable colloidal solids into a biological floc or biofilm, (3) transform or remove nutrients, such as nitrogen and phosphorus, and (4) in some cases, remove specific trace organic constituents and compounds. For industrial wastewater, the objective is to remove or reduce the concentration of organic and inorganic compounds. Because some of the constituents and compounds found in industrial wastewater are toxic to microorganisms, pretreatment may be required before the industrial wastewater can be discharged to a municipal collection system. For agricultural irrigation return (drainage) water containing nutrients, the objective is to remove the nutrients, specifically nitrogen and phosphorus, that are capable of stimulating the growth of aquatic plants.

Role of Microorganisms in Wastewater Treatment

The removal of dissolved and particulate carbonaceous BOD and the stabilization of organic matter found in wastewater is accomplished biologically using a variety of microorganisms, principally bacteria. Microorganisms are used to oxidize (i.e., convert) the dissolved and particulate carbonaceous organic matter into simple end products and additional biomass, as represented by the following equation for the aerobic biological oxidation of organic matter.

$$v_1\ (\text{organic material}) + v_2\text{O}_2 + v_3\text{NH}_3 + v_4\text{PO}_4^{3-} \xrightarrow{\text{microorganisms}} v_5\ (\text{new cells}) + v_6\text{CO}_2 + v_7\text{H}_2\text{O} \tag{7–1}$$

where v_i = the stoichiometric coefficient, as defined previously in Sec. 1–9 in Chap. 1.

In Eq. (7–1), oxygen (O_2), ammonia (NH_3), and phosphate (PO_4^{3-}) are used to represent oxygen and the nutrients needed for the conversion of the organic matter to simple end products [i.e., carbon dioxide (CO_2) and water]. The term shown over the directional arrow is used to denote the fact that microorganisms are needed to carry out the oxidation process. The term *new cells* is used to represent the biomass produced as a result of the oxidation of the organic matter. Microorganisms are also used to remove nitrogen and phosphorus in wastewater treatment processes. Specific bacteria are capable of oxidizing ammonia (nitrification) to nitrite and nitrate, while other bacteria can reduce the oxidized nitrogen to gaseous nitrogen. For phosphorus removal, biological processes are configured to encourage the growth of bacteria with the ability to take up and store large amounts of inorganic phosphorus.

Because the biomass has a specific gravity slightly greater than that of water, the biomass can be removed from the treated liquid by gravity settling. It is important to note that unless the biomass produced from the organic matter is removed on a periodic basis, complete treatment has not been accomplished because the biomass, which itself is organic, will be measured as BOD in the effluent. Without the removal of biomass from the treated liquid, the only treatment achieved is that associated with the bacterial oxidation of a portion of the organic matter originally present.

Types of Biological Processes for Wastewater Treatment

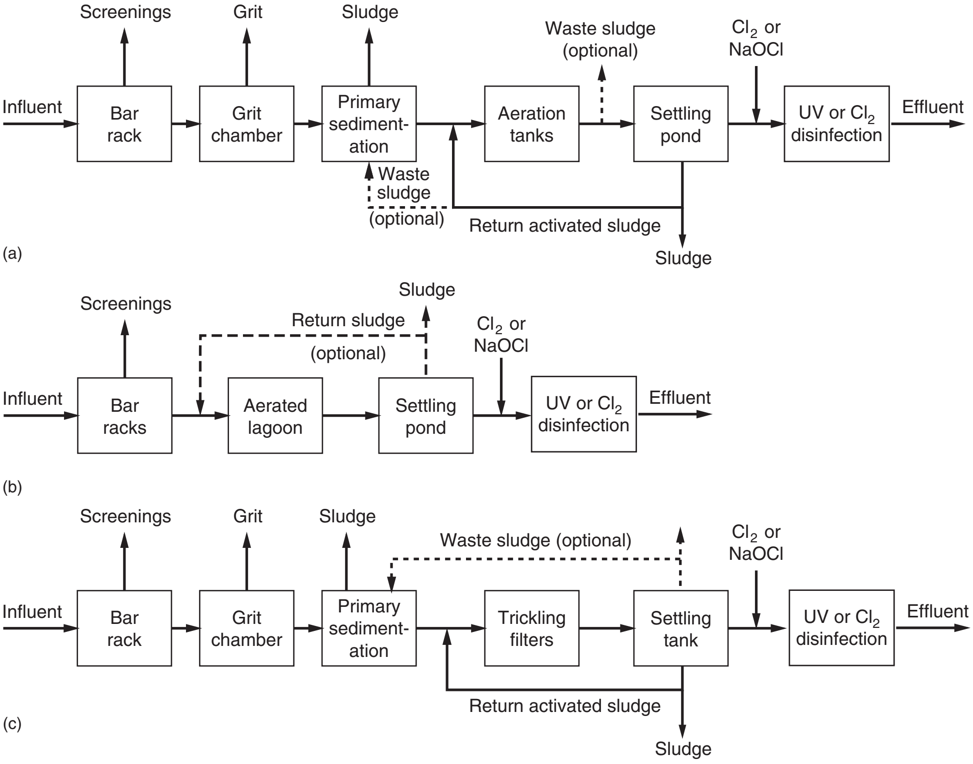
The principal biological processes used for wastewater treatment can be divided into two main categories: *suspended growth* and *attached growth* (or *biofilm*) processes. Examples of commonly used suspended and attached growth biological treatment processes are shown in Fig. 7–1. Typical process applications for suspended and attached growth biological treatment processes are given in Table 7–1, along with other treatment processes. The successful design and operation of the processes listed in Table 7–1 requires an understanding of the types of microorganisms involved, the specific reactions that they perform, the environmental factors that affect their performance, their nutritional needs, and their reaction kinetics. These subjects are considered in the sections that follow.

Suspended Growth Processes. In suspended growth processes, the microorganisms responsible for treatment are maintained in liquid suspension by appropriate mixing methods. Many suspended growth processes used in municipal and industrial wastewater treatment for biodegradation of organic substances are operated with dissolved oxygen (aerobic) or nitrate/nitrite (anoxic) utilization, but applications exist where suspended growth anaerobic (no oxygen present) reactors are used, such as for high organic concentration industrial wastewaters and organic sludges. The most common suspended growth process used for municipal wastewater treatment is the activated sludge process shown on Fig. 7–2 and discussed below.

Early Developments. The activated sludge process was developed around 1913 at the Lawrence Experiment Station in Massachusetts by Clark and Gage (Metcalf and Eddy, 1930)

Figure 7–1

Typical (simplified) flow diagrams for biological processes used for wastewater treatment:
(a) activated sludge process,
(b) aerated lagoons, and
(c) trickling filters.



and by Ardern and Lockett (1914) at the Davyhulme Sewage Works, in Manchester, England. The activated sludge process was so named because it involved the production of an activated mass of microorganisms capable of stabilizing a waste under aerobic conditions. In the aeration tank, contact time is provided for mixing and aerating influent wastewater with the microbial suspension, generally referred to as the mixed liquor suspended solids (MLSS) or mixed liquor volatile suspended solids (MLVSS). Mechanical equipment is used to provide the mixing and transfer of oxygen into the process (see Sec. 5–11, Chap. 5). The mixed liquor then flows to a clarifier where the microbial suspension is settled and thickened. The settled biomass, described as *activated sludge* because of the presence of active microorganisms, is returned to the aeration tank to continue biodegradation of the influent organic material. A portion of the thickened solids is removed daily or periodically as the process produces excess biomass that would accumulate with nonbiodegradable solids contained in the influent wastewater. If the accumulated solids are not removed, the solids level in the clarifier would increase until solids eventually exit via the system effluent flow.

An important feature of the activated sludge process is the formation of floc particles, ranging in size from 50 to 200 mm, which can be removed by gravity settling, leaving a relatively clear liquid as the treated effluent. Typically, greater than 99 percent of the suspended solids can be removed in the clarification step. As will be discussed in Chap. 8, the characteristics and thickening properties of the flocculent particles will affect the clarifier design and performance.

Treatment Objective and Process Developments. The treatment objectives and process configurations for the activated sludge process have changed considerably since its

Table 7–1

Major biological treatment processes used for wastewater treatment

Type	Common name	Use[a]
Aerobic processes:		
Suspended growth	Activated sludge process(es)	Carbonaceous BOD removal, nitrification
	Aerated lagoons	Carbonaceous BOD removal, nitrification
	Aerobic digestion	Stabilization, carbonaceous BOD removal
	Membrane bioreactor	Carbonaceous BOD removal, nitrification
	Nitritation process	Nitritation
Attached growth	Biological aerated filters	Carbonaceous BOD removal, nitrification
	Moving bed bioreactor	Carbonaceous BOD removal, nitrification
	Packed-bed reactors	Carbonaceous BOD removal, nitrification
	Rotating biological contactors	Carbonaceous BOD removal, nitrification
	Trickling filters	Carbonaceous BOD removal, nitrification
Hybrid processes	Trickling filter/activated sludge	Carbonaceous BOD removal, nitrification
	Integrated fixed film activated sludge (IFAS)	Carbonaceous BOD removal, nitrification
Anoxic processes:		
Suspended growth	Suspended-growth denitrification	Denitrification
Attached growth	Attached growth denitrification filter	Denitrification
Anaerobic processes:		
Suspended growth	Anaerobic contact processes	Carbonaceous BOD removal
	Anaerobic digestion	Stabilization, solids destruction, pathogen kill
	Anammox process	Denitritation, ammonia removal
Attached growth	Anaerobic packed and fluidized bed	Carbonaceous BOD removal, waste stabilization, denitrification
Sludge blanket	Upflow anaerobic sludge blanket	Carbonaceous BOD removal, especially high strength wastes
Hybrid	Upflow sludge blanket/attached growth	Carbonaceous BOD removal
Combined aerobic, anoxic, and anaerobic processes:		
Suspended growth	Single- or multi-stage processes, Various proprietary processes	Carbonaceous BOD removal, nitrification, denitrification, and phosphorus removal
Hybrid	Single- or multi-stage suspended growth processes with fixed film media	Carbonaceous BOD removal, nitrification, denitrification, and phosphorus removal
Lagoon processes:		
Aerobic lagoons	Aerobic lagoons	Carbonaceous BOD removal, nitrification
Maturation (tertiary) lagoons	Maturation (tertiary) lagoons	Carbonaceous BOD removal, nitrification
Facultative lagoons	Facultative lagoons	Carbonaceous BOD removal
Anaerobic-lagoons	Anaerobic-lagoons	Carbonaceous BOD removal (waste stabilization)

[a] Adapted from Tchobanoglous and Schroeder (1985).

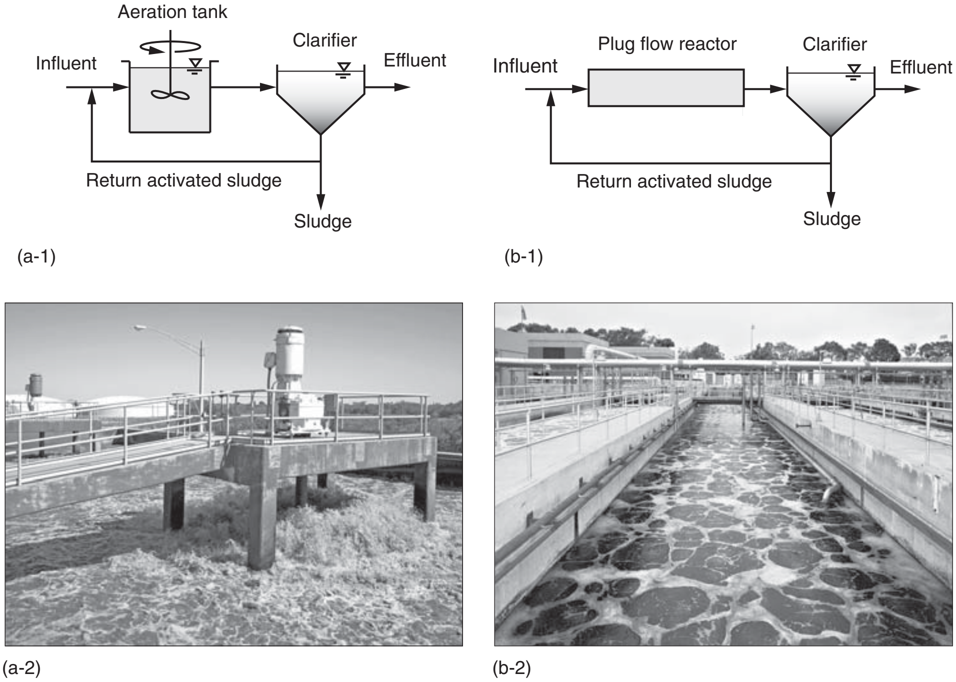


Figure 7–2

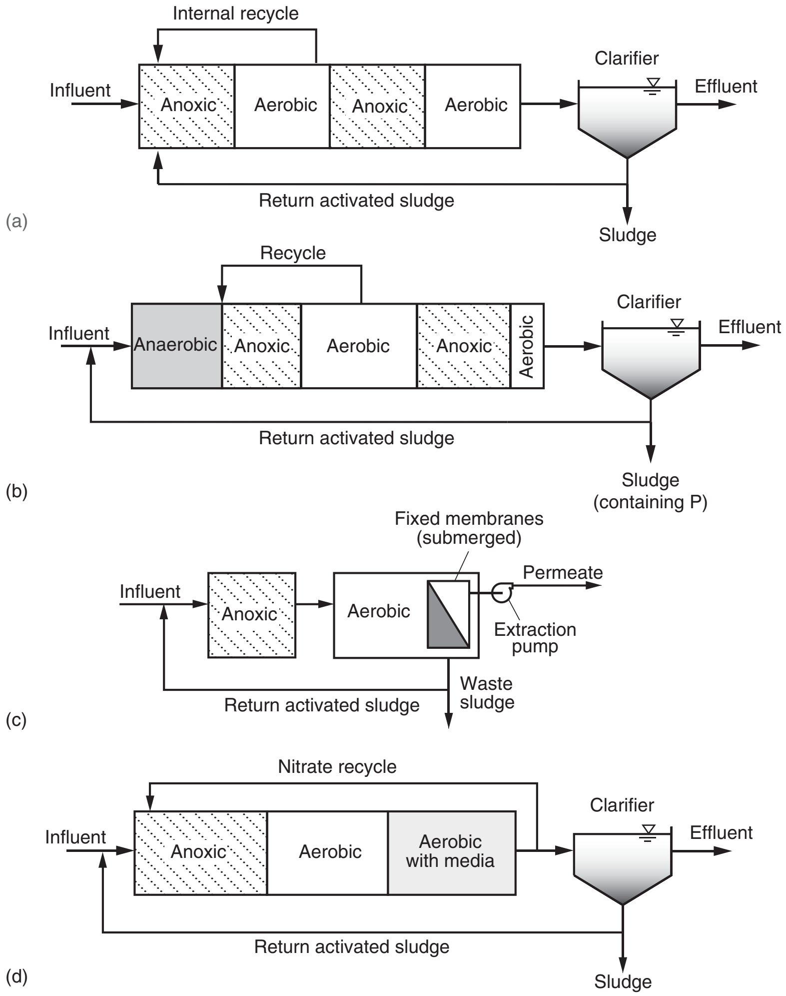
Suspended growth biological treatment process: (a-1) schematic and (a-2) view of complete-mix activated sludge process and (b-1) schematic and (b-2) view of plug-flow activated sludge process.

development and application in the early 1900s. Significant changes have been brought about by enhanced understanding of the process microbiology, improved characterization of the individual microorganisms involved, improvements in aeration technology, and new equipment innovations. As a result of these discoveries and innovations, the process capabilities have been expanded and treatment objectives have become more rigorous. In the late 1950s, Pasveer developed a race track configuration for activated sludge treatment (Hao et al., 1997) that led to the popular oxidation ditch system with biological nitrogen removal that is employed at hundreds of wastewater treatment plants around the world today. A common hindrance to operation and performance over the first 60 y of using activated sludge was settling problems associated with the growth of filamentous bacteria. Work by Chudoba et al. (1973) led to the concept of biological selectors and staged reactor configurations to prevent filamentous growth. Staged activated sludge reactor configurations proposed by Barnard (1974) led to activated sludge process applications for biological removal of nitrogen and phosphorus in addition to carbonaceous BOD removal.

Processes with a sequence of anoxic-aerobic-anoxic-aerobic zones for biological nitrogen removal and the addition of an anaerobic contact zone to promote enhanced biological phosphorus removal are shown on Figs. 7–3(a) and (b). Another major change in activated sludge process design occurred in the late 1990s when membrane materials and manufacturing technology had advanced sufficiently so that membranes could be used in activated sludge for effluent liquid-solids separation instead of clarifiers. A membrane bioreactor (MBR) system is depicted in Fig. 7–3(c). The membrane separation unit is immersed in an aerated activated sludge compartment and a sufficient head is provided by gravity or pumping so that the effluent or permeate can be discharged essentially free of suspended solids. As will be shown in Chap. 8, the MBR process can also be incorporated with biological nutrient removal designs.

Figure 7–3

Progression of activated sludge processes: (a) anoxic-aerobic activated sludge for nitrogen removal, (b) anaerobic-anoxic-aerobic-anoxic-aerobic process for nitrogen and phosphorus removal, (c) anoxic-aerobic treatment in membrane bioreactor process with nitrogen removal, and (d) integrated fixed film activated sludge process with nitrogen removal.



Attached Growth Processes. In attached growth processes, the microorganisms responsible for the conversion of organic material or nutrients are attached to an inert packing material. The organic material and nutrients are removed from the wastewater flowing past the attached growth, also known as a biofilm. Packing materials used in attached growth processes include rock, gravel, slag, sand, redwood, and a wide range of plastic and other synthetic materials. Attached growth processes can also be operated as aerobic or anaerobic processes. The packing can be submerged completely in liquid or partially submerged, with air or gas space above the biofilm liquid layer.

The most common aerobic attached growth process used is the trickling filter in which wastewater is distributed over the top area of a vessel containing nonsubmerged packing material (see Fig. 7–4). Historically, rock was used most commonly as the packing material for trickling filters [see Fig. 7–4(a)], with typical depths ranging from 1.25 to 2 m (4 to 6 ft). Most modern trickling filters vary in height from 5 to 10 m (16 to 33 ft) and are filled with a plastic packing material for biofilm attachment [see Fig. 7–4(b)]. The plastic packing material is designed such that about 90 to 95 percent of the volume in the tower consists of void space. Air circulation in the void space, by either natural draft or blowers,

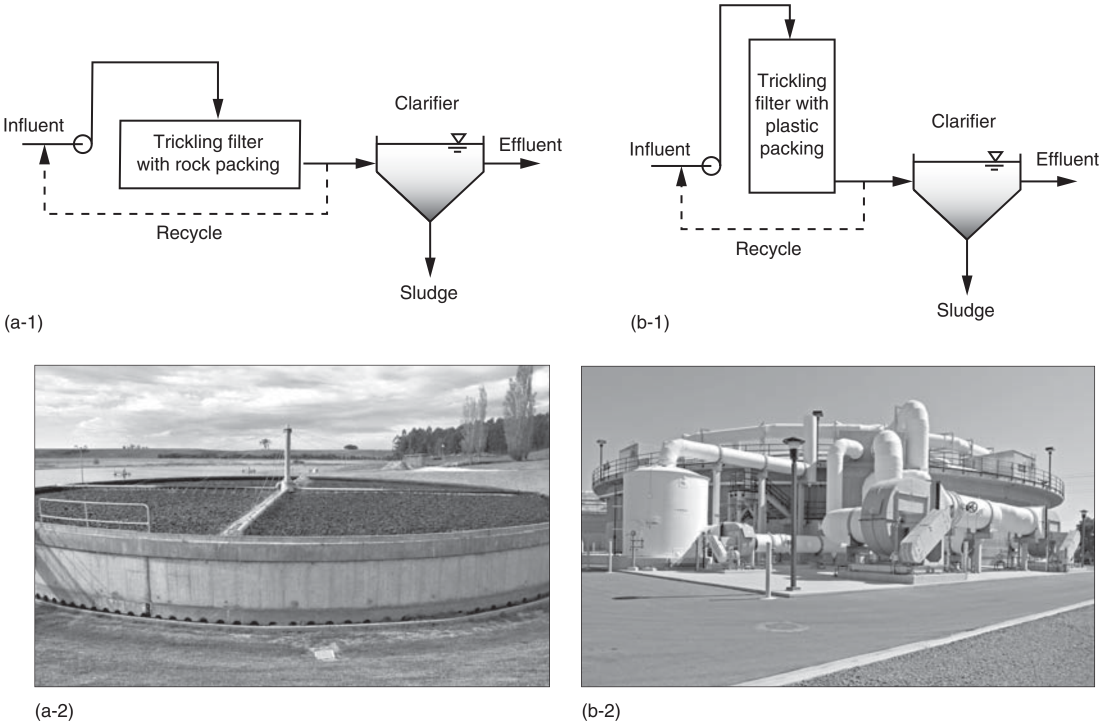


Figure 7–4

Attached growth biological treatment process: (a-1) schematic and (a-2) view of trickling filter with rock packing; and (b-1) schematic and (b-2) view of covered tower trickling filter with plastic packing. The air injection and odor control facilities are shown on the foreground. The tower filter is 10 m high and 50 m in diameter.

provides oxygen for the microorganisms growing as an attached biofilm. Influent wastewater is distributed over the packing and flows as a nonuniform liquid film over the attached biofilm. Excess biomass sloughs from the attached growth periodically and some form of liquid/solids separation is needed to provide an effluent with an acceptable suspended solids concentration. The solids are collected at the bottom of the clarifier and removed for waste-sludge processing.

The popularity of trickling filters had declined with the increasing applications for biological nutrient removal. Advances in the 1990s involved the combination of fixed film and activated sludge treatment, termed an integrated fixed film activated sludge process (IFAS) [see Fig. 7–3(d)]. Other fixed film processes that save space and/or can be used with biological nutrient removal have received more attention and are presented in Chap. 9.

7–2 COMPOSITION AND CLASSIFICATION OF MICROORGANISMS

Biological processes for wastewater treatment consist of mixed microbial communities which may include bacteria, archaea, protozoa, fungi, rotifers, and possibly algae. The basic characteristics and important roles of these organisms have been described in Chap. 2. In some cases, biological treatment goals can only be accomplished by the presence of a specific microbial species. To provide a basic understanding of the nature of microorganisms, the topics introduced in this section are (1) cell components, (2) cell composition, (3) environmental factors that affect microbial activity, and (4) methods used to identify and classify microorganisms. The focus here is mainly on prokaryotes, bacteria and archaea (see Sec. 2–8 in Chap. 2), because of their major role in biological wastewater treatment.

Cell Components

The important components of the prokaryotic cell and their functions were illustrated on Fig. 2–28(a) and are described in Table 7–2; the eukaryotic cell is illustrated on Fig. 2–28(b). Key components that relate to the cell's genetic information and specific enzymes produced, which determine the capability of the microorganism in wastewater treatment, are deoxyribose nucleic acid (DNA) and the ribosomes. Ribosomes are the sites of protein synthesis, which are necessary for enzyme production, and the DNA provides the genetic information used to determine the protein structure synthesized. To understand how DNA codes for the cell proteins, the DNA structure and nucleotide sequence and the structure and role of ribose nucleic acid (RNA) are reviewed.

Nucleic Acids. The nucleic acids, DNA and RNA, are composed of a series of nucleotides. Each nucleotide consists of a five-carbon sugar compound, a nitrogen base,

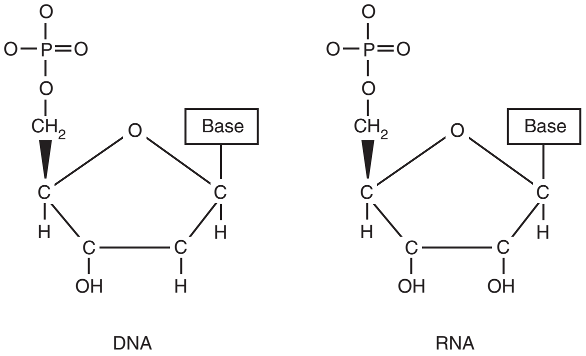
Table 7–2

Description of prokaryote cell components

Cell Component	Function
Cell wall	Provides strength to maintain the cell shape and protects the cell membrane. Some bacteria can produce a sticky polysaccharide layer outside the cell wall, called a capsule or slime layer.
Cell membrane	Controls the passage of dissolved organics and nutrients into the cell and the waste materials and metabolic by products out of the cell.
Cytoplasm	Contains the material within the cell to carry out cell functions and includes water, nutrients, enzymes, ribosomes, and small organic molecules.
Cytoplasmic inclusions	Contains storage material that can provide carbon, nutrients, or energy. These may be carbohydrate deposits, such as polyhydroxybutyrate (PHB) or glycogen, polyphosphates, lipids, and sulfur granules.
Deoxyribonucleic acid (DNA)	A double-stranded helix-shaped molecule that contains genetic information which determines the nature of the cell protein and enzymes that are produced.
Plasmid DNA	Small circular DNA molecules that can also provide genetic characteristics for the bacteria.
Ribosomes	Particles in the cytoplasm that are composed of ribonucleic acid (RNA) and protein and are the sites where proteins are produced.
Flagella	Protein hair-like structures that extend from the cytoplasm membrane several bacteria lengths out from the cell and provide mobility by rotating at high speeds.
Fimbriae and pili	Short protein hair-like structures (Pili is longer) that enable bacteria to stick to surfaces. Pili also enable bacteria to attach to each other.

Figure 7–5

Nucleotide structure of deoxyribose nucleic acid (DNA) and ribose nucleic acid (RNA).



and a phosphate molecule (see Fig. 7–5). To form the nucleotide chain of DNA or RNA, the phosphate group bonds to the third carbon (clockwise from the oxygen bond) of the sugar molecule. The nitrogen bases for DNA can be one of four pyrimidine or purine compounds: cytosine (C), thymine (T), adenine (A), and guanine (G). For RNA, four nitrogen bases are also involved, including A, C, and G, and thymine is replaced with uracil (U). The DNA is a double-strand helix structure with bonding between the nitrogen bases of each strand. Base bonding is very specific with bonds only between G and C, and A and T. The RNA is a single strand of a nucleotide sequence of various combinations of A, C, G, and U. The sequence of nucleotides in DNA contains the necessary genetic codes for the cell, which determine the specific proteins and enzymes that the microorganism can produce. The number of nucleotides in DNA is very high, and the size of the DNA molecule is described in terms of the number of thousands of nucleotide bases (kilobase) per molecule. The bacterium *Escherichia coli* has 4.7 million nucleotides in each DNA strand or 4700 kilobase pairs.

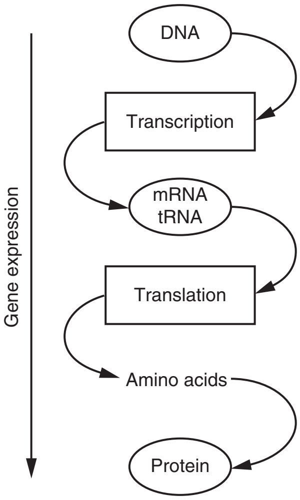


Figure 7–6

Gene expression leads to the formation of proteins by transcription of a segment of the DNA genetic code and translation in the ribosome via messenger RNA (mRNA) and transfer RNA (tRNA) to form a series of amino acids, which form polypeptides and finally protein.

Gene Expression. Gene expression as illustrated on Fig. 7–6 involves the transcription and translation of a segment (gene) of the DNA to form a specific protein. The first step in the process is transcription, in which a small segment of the DNA is unraveled to form a single strand that is used to form a single strand of RNA by complementing base pairing of the nitrogen bases in the DNA nucleotides. For example, adenine (A) in the DNA strand pairs with uracil (U), and guanine (G) with cytosine (C) in the mRNA. The order of the nucleotides in the mRNA will determine the order of amino acids that form polypeptides and the protein structure produced. Translation of the mRNA occurs in the ribosome by tRNA. Each of the nucleotides on the mRNA in series is matched by complementary base pairing with the tRNA in the ribosome, and for each match, another segment of the tRNA, containing three nucleotides, selects a specific amino acid. The three nucleotide sequences in the mRNA are called *codons*. In essence, each codon selects for a specific amino acid, and there is more than one codon for each of the possible 21 amino acids that can be found in living cells. Thus, the length and nucleotide sequence expressed in the DNA represents a gene that determines what specific protein is formed. Because protein is an essential component of cellular enzymes, the DNA gene composition then determines the microbial cell functions and degradative capabilities. Additional details on gene expression may be found in Madigan et al. (2012).

Cell Enzymes. Cell enzymes, consisting of protein and a cofactor such as a metal ion (e.g., zinc, iron, copper, manganese, or nickel), determine the metabolic capability of microorganisms in wastewater treatment. Enzymes are large organic molecules with molecular weights in the range of 10,000 to 1,000,000. Enzymes catalyze biological reactions necessary for cell functions, such as hydrolysis, oxidation-reduction reactions, and cell synthesis reactions. Cells may also produce enzymes for activity outside the cell wall (extracellular enzymes). An example of the function of extracellular enzymes is the hydrolysis of particulates and large molecules so that the material can be transported across the cell membrane for use by the cell. Enzymes can be *constitutive* or *inducible*. Constitutive enzymes are produced continuously by the cell, while inducible enzymes are produced in response to the presence of a particular compound. The rate of enzyme activity is affected by temperature and pH.

Cell Composition

To support microbial growth in biological systems, appropriate nutrients must be available. Reviewing the composition of a typical microbial cell will provide a basis for understanding the nutrients needed for growth. Prokaryotes are composed of about 80 percent water and 20 percent dry material, of which 90 percent is organic and 10 percent is inorganic. Typical values for the composition of prokaryote cells are reported in Table 7–3. The most widely used empirical formula for the organic fraction of cells is $C_5H_7O_2N$, first proposed by Hoover and Porges (1952). About 53 percent by weight of the organic fraction is carbon. The formulation $C_{60}H_{87}O_{23}N_{12}P$ can be used when phosphorus is also considered. It should be noted that both formulations are approximations and may vary with time and species, but they are used for practical purposes. Nitrogen and phosphorus are considered macronutrients because they are required in comparatively large amounts. Prokaryotes also require trace amounts of metallic ions, or micronutrients, such as zinc, manganese, copper, molybdenum, iron, and cobalt. Because all of these elements and compounds must be derived from the environment, a shortage of any of these substances would limit and, in some cases, alter growth.

Environmental Factors

Environmental conditions such as temperature and pH have an important effect on the selection, survival, and growth of microorganisms. In general, optimal growth for a particular microorganism occurs within a fairly narrow range of temperature and pH, although most microorganisms can survive within much broader limits. Temperatures below the optimum typically have a more significant effect on growth rate than temperatures above the optimum; it has been observed that growth rates double with approximately every 10°C increase in temperature until the optimum temperature is reached. According to the temperature range in which they function best, prokaryotes may be classified as *psychrophilic*, *mesophilic*, or *thermophilic*. Typical temperature ranges for microorganisms in each of these categories are presented in Table 7–4. For a more detailed discussion of the organisms in the various temperature ranges, see Madigan et al. (2012).

The pH of the environment is also a key factor in the growth of organisms. Most bacteria cannot tolerate pH levels above 9.5 or below 4.0. Generally, the optimum pH for bacterial growth lies between 6.5 and 7.5. Different archaea are able to grow at thermophilic and ultrathermophilic (60 to 80°C) temperatures, extremely low pH, and high salinity.

Table 7–3

Typical composition of bacteria cells[a]

Constituent or element	Percent of dry weight
Major cellular material	
Protein	55.0
Polysaccharide	5.0
Lipid	9.1
DNA	3.1
RNA	20.5
Other (sugars, amino acids)	6.3
Inorganic ions	1.0
As cell elements	
Carbon	50.0
Oxygen	22.0
Nitrogen	12.0
Hydrogen	9.0
Phosphorus	2.0
Sulfur	1.0
Potassium	1.0
Sodium	1.0
Calcium	0.5
Magnesium	0.5
Chlorine	0.5
Iron	0.2
Other trace elements	0.3

[a] Adapted from Madigan et al. (2012).

Table 7–4

Temperature classification of biological processes

Type	Temperature range, °C	Optimum range, °C
Psychrophilic	10–30	12–18
Mesophilic	20–50	25–40
Themophilic	35–75	55–65

Microorganism Identification and Classification

With the use of more sophisticated biological treatment processes and developments in the use of molecular tools for microbial applications, environmental engineers have expanded their interests from the general functionality of microorganisms to understanding the presence and role of specific microbial species in biological treatment systems. For example, engineered systems can be controlled to create selective pressures that will favor the growth of preferred microorganisms. Thus, the identification of microorganisms and their metabolic characteristics is very important and is described here along with molecular methods used in biological processes.

In the past, the methods used to identify microorganisms relied on physical taxonomic (morphologic) and metabolic characteristics (phenotypic analysis). With modern tools from molecular biology, microorganims identification is now based on cellular genetic information. The basic taxonomic unit in the identification of bacteria, for example, is the species, which represents a collection of similar strains of bacteria that exhibit characteristics significantly different from other groups of bacteria. Species that share one or more major properties are placed in a group termed a genus (plural genera). All bacteria are given a genus and species name. The genus name is capitalized and placed before the species name and can be abbreviated ahead of the species name after first identified. The genus and species names are italicized. For example, the genus *Bacillus* contains several species including *B. subtilis*, *B. cerus*, and *B. stearothermophilis*, based on different morphological, physiological, and ecological traits (Madigan et al., 2012).

Taxonomic Classification. Conventional taxonomic methods used to identify a bacterium, for example, rely on physical properties of the bacteria and metabolic characteristics. To apply this approach, a pure culture must first be isolated. The culture may be isolated by serial dilution and growth in selective growth media. The cells are harvested and grown as pure culture using sterilization techniques to prevent contamination. In some cases, isolation of a species is not possible, which may be due to the need for synergistic growth with other species or the lack of a specific growth factor. Historically, the types of tests that are used to characterize a pure culture include (1) microscopic observations, to determine morphology (size and shape); (2) gram staining, to determine if the bacteria cell wall will absorb crystal violet dye; (3) the type of electron acceptor (i.e., oxygen, CO_2, etc.) used in oxidation-reduction reactions; (4) the type of carbon source used for cell growth; (5) the ability to use various nitrogen and sulfur sources; (6) nutritional needs; (7) cell wall chemistry; (8) cell characteristics including pigments, segments, cellular inclusions, and storage products; (9) resistance to antibiotics; and (10) environmental effects of temperature and pH. An alternative to taxonomic classification is a newer method, termed *phylogeny*.

Phylogenetic Classification. In the late 1970s, microbiologists began to use tools that allowed them to study microorganisms at the molecular level and to observe genetic relationships related to the evolutionary history of living cells. The characterization of microorganisms based on genetic information and evolutionary location in time is termed *phylogeny*, which is the more current method of identification and classification. To identify accurately the microbes and determine the true evolutionary relationships among species, the choice of cellular genetic material is critical. The genetic code for ribosomal RNA was chosen as the evolutionary chronometer for cell identification because the code (1) is of evolutionary significance, (2) is present in all known forms of life, (3) is well conserved across broad phylogenetic distances, and (4) contains sufficient nucleotide sequences so that similarity in sequences between two organisms indicates a phylogenetic relationship (Pace et al., 1986).

Ribosomal RNA. Ribosomal RNA can be separated into two components, 30S (Svedberg units) and 50S, based on different centrifugal forces in ultracentrifugation. The 30S units consist of 16S rRNA (18S rRNA in eukaryotes), containing about 1500 nucleotides and 21 proteins. While the 16S rRNA can be extracted from cells for nucleotide sequencing using molecular techniques, more often the section of DNA that encodes the rRNA genes is used. This method includes genomic DNA extraction from the cell material, followed by a polymerase chain reaction (PCR) procedure that uses DNA primers and a thermally-stable DNA polymerase enzyme to reproduce and amplify artificially a section of the DNA material by a factor of 10^6 or more from the small amount of DNA extracted from the cell. The amplified 16S rRNA gene is then sequenced to determine the order of nucleotide bases in 16S rRNA gene. The sequencing result is compared to the ribosome sequences available in a database to determine its phylogenetic relationship to known organisms.

Molecular Phylogeny. Molecular phylogeny involves a systematic organization and classification of microorganisms based on their genetic traits. The phylogenetic tree of life as determined from 16S and 18S rRNA sequencing is shown on Fig. 7–7. Life is divided into three basic domains, two composed of prokaryotic cells (Archaea and Bacteria), as discussed previously, and the third composed of Eukarya cells. The gene used to build the tree of life is highly conserved, meaning that changes to the genomic sequence that encode this gene occur slowly. Thus, the similarity of 16S rRNA gene sequences can be used as an evolutionary timeline to estimate the distance to a common ancestor (defined as the nodes or branches in the tree of life). The evolutionary history of the organisms does not necessarily indicate the evolution of the organisms' full set of genes, which can be passed between unrelated species by several mechanisms. Therefore, metabolism of the organisms cannot always be inferred by position on the tree of life. This situation has been shown to be particularly true for genes encoding sulfate reduction and ammonia oxidation, which are found in discrete branches within both the bacterial and archaeal domains. In addition to having different rRNA sequencing properties, the archaea have a number of different phenotypic properties compared to bacteria. These differences are found in cell wall composition, cell membrane lipid chemistry, RNA polymerase composition, and protein synthesis mechanisms in the ribosome.

Domains of Life. Representatives from all three domains of life are involved in wastewater treatment. Nitification (Sec. 7–9) is an important process in biological nutrient

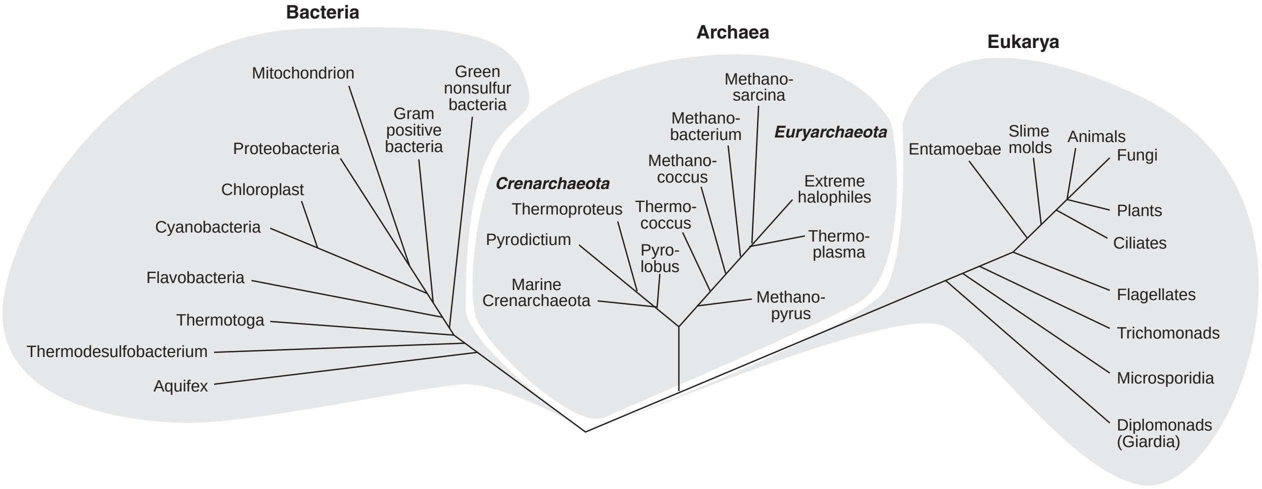


Figure 7–7

Phylogenetic tree of life.

removal and involves a first step of biological oxidation of ammonium. Aerobic ammonia oxidizing prokaryotes (AOP) are found within the Bacteria and Archaea domains with the ammonia oxidizing bacteria (AOB) in the genus *Nitrosomonas* and *Nitrosospira* within the β-proteobacteria class of the phylum, Proteobacteria. The ammonia oxidizing archaea (AOA) were originally recognized as Crenarchaeota and have now been reclassified to a new phylum named Thaumarchaeota.

The more recently found "anammox" bacteria (Sec. 7–11), which can oxidize ammonia under anaerobic conditions, have been classified within the phylum Planctomycetes. Bacteria that are important for the removal of phosphorus without chemical addition in enhanced biological phosphorus removal (EBPR) processes (Sec. 7–12) appear to be similar to the phosphate accumulating bacteria Candidatus *Accumulibacter phosphatis*, also a member of the β-proteobacteria. Nitrite oxidizing bacteria (NOB) within the genus *Nitrobacter* are within the α-proteobacteria and *Nitrospira* are within the phylum Nitrospirae. Methylotrophic bacteria important for using methanol in denitrification applications include the proteobacteria *Hyphomicrobium spp*. and *Methyloversatilis spp*. All the methanogens are Archaea within the phylum Euryarchaeaota. Eukaryotes within wastewater treatment include amoebae, ciliates, and rotifers.

Use of Molecular Tools

In addition to identifying and classifying microorganisms, molecular tools are used to provide information about active microbial communities and to monitor water or wastewater treatment plant effluents for specific pathogenic organisms. Molecular techniques use DNA, RNA, and proteins to identify, track, and quantify the presence and activities of microorganisms. Some common molecular tools that are applied to understand relationships between microbial community composition and biological process performance are considered below. The use of molecular tools is a rapidly growing area of microbiology. The continual development of new tools in the next few years for application to research and operations of biological wastewater treatment processes may be expected.

Polymerase Chain Reaction (PCR). Many molecular tools are based on the polymerase chain reaction (PCR) process in which small sections of DNA are amplified using naturally occurring enzymes involved in cellular DNA replication. The DNA is recovered from the complex microbial communities that comprise biological processes by first disrupting the cells either physically (e.g., shaking with glass micro-beads) or chemically (e.g., by adding phenol, lysozyme, or detergent). The recovered DNA is cleaned and concentrated by sorption to a column in the presence of ethanol. The clean DNA is eluted into purified water (see Fig. 7–8). The PCR process proceeds by mixing the sample (i.e., template) DNA, PCR primers (short oligonucleotides that complement a section of the target DNA sequence), DNA polymerase (a naturally occurring enzyme that creates copies of DNA during cell replication), a mixture of nucleotides (the molecular building-blocks for constructing new DNA), and a pH buffer containing Mg^{2+}.

A typical PCR reaction uses three temperature steps (see Fig. 7–9): (1) the temperature is increased to about 95°C causing the double stranded DNA to separate into two single strands, (2) the temperature is lowered to allow PCR primers to anneal to the DNA template, and (3) the temperature is increased to about 72°C and DNA polymerase extends the copy of the template DNA. The amount of amplified DNA doubles with each repeated temperature cycle. The process is done in instruments, called thermocyclers, specifically designed to modulate rapidly between the required temperatures, allowing much greater process precision. The most commonly used DNA polymerase was originally obtained

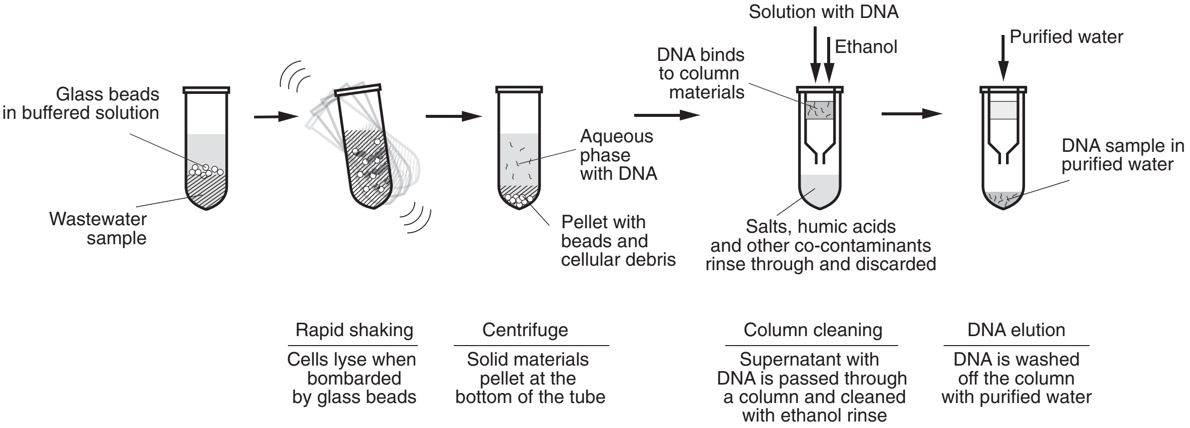


Figure 7–8

Extraction of DNA from wastewater samples. A vortex mixer or an instrument specifically designed for cell disruption by rapid shaking is used for the rapid shaking process.

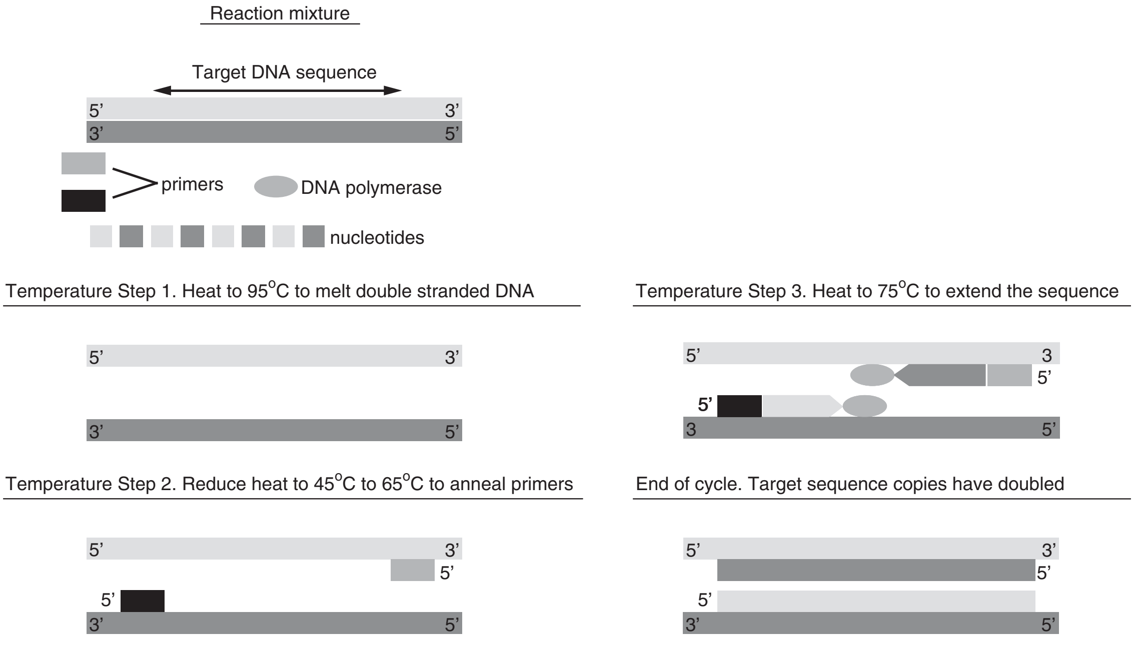


Figure 7–9

Sequence of a polymerase chain reaction (PCR) process.

from the thermophilic Bacteria *Thermus aquaticus* (Taq) and remains active when exposed to the high temperatures used to melt the template DNA. A great advantage of the PCR process is that the primers can be designed to target DNA sequences that are either conserved across domains or are unique to individual species. For example, primers have been designed that target all Bacteria, all Archaea, *dsr* genes (associated with sulfate reduction), *amoA* genes (associated with ammonia oxidation), or even 16S rRNA gene sequences of a single species.

Modifications of PCR Technologies. PCR allows genetic identification without requiring prior knowledge of a microbial community composition and has led to the discovery that less than 1 percent of microbial diversity was previously documented. Even now, studies of biological processes lead to new species discoveries. Some applications of PCR to wastewater treatment processes are described in the following paragraphs.

Quantification of Microbial Populations in Mixed Communities. By modifying the PCR approach, different microbial populations within mixed communities can be quantified using quantitative PCR (qPCR). The qPCR procedure uses a dye that fluoresces when bound to double-stranded DNA, which allows measurement of the amount of PCR product at the end of each PCR cycle. Through comparison to standards with known DNA concentrations, qPCR is used to quantify relative bacterial or functional gene concentrations.

Profiles of Mixed Communities. Another modification of the PCR technique can be used to generate profiles of mixed microbial communities, sometimes called community "finger prints." For example, terminal restriction fragment length polymorphism (TRFLP) modifies PCR by adding a fluorescent tag to one of the primers. The PCR products are cut using a restriction enzyme. The distance from the PCR priming site to the restriction enzyme cut site will differ among microbial populations. The resulting fragments are size separated using conventional sequencing instruments and a chromatogram of florescence intensities is recorded corresponding to the differing fragment lengths.

Sequencing PCR Products. Sequencing PCR products can provide information about microbial phylogenetic identity. When the PCR product is from a mixed microbial community, cloning reactions of the PCR products can be used to separate amplified DNA from individual organisms. The PCR product is incorporated into plasmid vectors (small circular DNA). The vectors are transformed into "competent" *E. coli* cells. When the *E. coli* are grown on solid media, each colony starting from a single *E. coli* cell will have identical copies of one original PCR product strand. These strands can then be sequenced and the sequence results compared to public databases, such as the Basic Local Alignment Search Tool (BLAST) at the National Center for Biotechnology Information (NCBI) website (http://blast.ncbi.nlm.nih.gov), to reveal the microorganisms that comprised the original mixed community.

Other Molecular Methods. Several molecular methods do not require a PCR step. One such method is fluorescent in-situ hybridization (FISH). An oligonucleotide probe (similar in design to PCR primers) is labeled so that it can be detected by fluorescence. The probes are allowed to penetrate intact cells, where they hybridize to the rRNA of active microorganisms. Multiple probes can be applied with each, using different fluorescens. When viewed through a microscope with fluorescence capabilities, FISH allows visual observation of how different microbial populations interact and allows quantification of the population through cell counting (Maier et al., 2000). The use of the FISH technique to identify dense clusters (up to 3000 cells) of ammonia-oxidizing (*Nitrosomonas*) bacteria in several activated sludge plant sample using the NEU probe was demonstrated by Wagner et al. (1995). A probe targeting *E. coli* that was developed to study UV disinfection was illustrated previously on Fig. 2–36 in Chap. 2.

One of the distinct advantages of FISH is that the distribution of a species in its environment can be observed. When FISH techniques are employed with more than one nucleic acid probe added to the sample, a number of bacterial species or strains can be identified and their relative distribution within the matrix can be observed. Microradiography-FISH (MAR-FISH) modifies the approach to provide a link between phylogeny and physiology by first incubating an active culture in the presence of a C^{14}-labeled substrate. Following fluorescence to identify bacteria, microscope slides are overlaid onto film and bacteria that took up the labeled compounds are identified based on film exposure.

Metagenomics. As sequencing technologies have become more efficient and affordable, it has become possible to sequence community DNA without first requiring a PCR step. This technique is known as metagenomics and provides information not only on 16S rRNA sequence data but also on all genomics data (much of which still has no assigned function). This approach was used to obtain a full genome sequence of "Candidatus *Accumulibacter phosphatis"* (Martín et al., 2006), a bacterium involved in enhanced biological phosphorus removal (EBPR) that has thus far eluded culture attempts. By studying the genome sequence, new hypotheses have been generated about the physiological and adaptive capabilities of this important bacterium that will help direct future EBPR designs.

Proteomics. Another method that has been applied to many environments is proteomics, used to examine protein profiles from mixed samples. The amino acids that comprise each protein are compared to genome sequence information to identify which genes were active at the time the sample was collected. Comparing protein profiles obtained under a different set of conditions (for example, with and without a contaminant or at different reactor temperatures) may provide insights into how operational changes impact microbial activities.

7–3 INTRODUCTION TO MICROBIAL METABOLISM

Basic to the design of a biological treatment process, or to the selection of the type of biological process to be used, is an understanding of the biochemical activities of microorganisms. The classification of microorganisms by sources of cell carbon, electron donor, electron acceptor, and end products is summarized in Table 7–5. Different microorganisms can use a wide range of electron acceptors, including oxygen, nitrite, nitrate, iron (III), sulfate, organic compounds, and carbon dioxide. The two major topics considered in this section are (1) the general nutritional requirements of the microorganisms commonly encountered in wastewater treatment, and (2) the nature of microbial metabolism based on the need for molecular oxygen.

Carbon and Energy Sources for Microbial Growth

To continue to reproduce and function properly, an organism must have sources of energy, carbon for the synthesis of new cellular material, and inorganic elements (nutrients) such as nitrogen, phosphorus, sulfur, potassium, calcium, and magnesium. Organic nutrients (growth factors) may also be required for cell synthesis. Carbon and energy sources, usually referred to as *substrates*, and nutrient and growth factor requirements for various types of organisms are considered in the following discussion.

Table 7–5

Classification of microorganisms by electron donor, electron acceptor, sources of cell carbon, and end products. All reactions shown support new biomass growth

Type of bacteria	Common reaction name	Carbon source	Electron Donor (substrate oxidized)	Electron acceptor	Products
Aerobic heterotrophic	Aerobic oxidation	Organic compound	Organic compound	O_2	CO_2, H_2O
Aerobic autotrophic	Nitrification	CO_2	NH_4^+, NO_2^-	O_2	NO_2^-, NO_3^-
	Iron oxidation	CO_2	Fe (II)	O_2	Ferric Iron Fe (III)
	Sulfur oxidation	CO_2	H_2S, S°, $S_2O_3^{2-}$	O_2	SO_4^{2-}
Facultative heterotrophic	Denitrification anoxic reaction	Organic compound	Organic compounds	NO_2^-, NO_3^-	N_2, CO_2, H_2O
Anaerobic heterotrophic	Acid fermentation	Organic compound	Organic compounds	Organic compounds	Volatile fatty acids (VFAs) (acetate, propionate, butyrate)
	Iron reduction	Organic compound	Organic compounds	Fe (III)	Fe (II), CO_2, H_2O
	Sulfate reduction	Organic compound	Organic compounds	SO_4	H_2S, CO_2, H_2O
	Methanogenesis	Organic compound	Volatile fatty acids (VFAs)	CO_2	Methane
Anaerobic autotrohpic	Anammox	CO_2	NH_4^+	NO_2^-	N_2, NO_3^-

Carbon Sources. Microorganisms obtain their carbon for cell growth from either organic matter or carbon dioxide. Organisms that use organic carbon for the formation of new biomass are called *heterotrophs*, while organisms that derive cell carbon from carbon dioxide are called *autotrophs*. The conversion of carbon dioxide to cellular carbon compounds is an anabolic process, which requires a net input of energy. Autotrophic organisms must therefore spend more of their energy for synthesis than do heterotrophs, resulting in generally lower yields of cell mass and growth rates.

Energy Sources. The energy needed for cell synthesis may be supplied by light or by a chemical oxidation reaction. Organisms that are able to use light as an energy source are called *phototrophs*. Phototrophic organisms may be either heterotrophic (certain sulfur-reducing bacteria) or autotrophic (algae and photosynthetic bacteria). Organisms that derive their energy from chemical reactions are known as *chemotrophs*. As with the phototrophs, chemotrophs may be either heterotrophic (protozoa, fungi, and most bacteria) or autotrophic (i.e., nitrifying bacteria). *Chemoautotrophs* obtain energy from the oxidation of reduced *inorganic* compounds, such as ammonia, nitrite, ferrous iron, and sulfide. *Chemoheterotrophs* usually derive their energy from the oxidation of *organic* compounds.

Oxidation-Reduction Reaction. The energy-producing chemical reactions by chemotrophs are oxidation-reduction reactions that involve the transfer of electrons from an electron donor to an electron acceptor. The electron donor is oxidized and the electron acceptor is reduced. The electron donors and acceptors can be either organic or inorganic compounds, depending on the microorganism. The electron acceptor may be available within the cell during metabolism (endogenous), or it may be obtained from outside the cell (i.e., dissolved oxygen) (exogenous). Organisms that generate energy by enzyme-mediated electron transport to an external electron acceptor are said to have a *respiratory metabolism*. The use of an internal electron acceptor is termed *fermentative metabolism* and is a less efficient energy-yielding process than respiration. Heterotrophic organisms that are strictly fermentative are characterized by lower growth rates and cell yields than respiratory heterotrophs.

Utilization of Oxygen. When oxygen is used for the electron acceptor the reaction is termed *aerobic*, and reactions involving other electron acceptors are considered anaerobic. The term *anoxic* is used to distinguish the use of nitrite or nitrate for electron acceptors from the others under anaerobic conditions. Under anoxic conditions nitrite or nitrate reduction to gaseous nitrogen occurs, and this reaction is also referred to as biological denitrification. Organisms that can only meet their energy needs with oxygen are called *obligate aerobic* microorganisms. Some bacteria can use oxygen or nitrate/nitrite as electron acceptors when oxygen is not available. These bacteria are called *facultative aerobic* bacteria.

Energy for Fermentation. Organisms that generate energy by fermentation and that can exist only in an environment that is devoid of oxygen are *obligate anaerobes*. *Facultative anaerobes* have the ability to grow in either the presence or absence of molecular oxygen and fall into two subgroups, based on their metabolic abilities. True facultative anaerobes can shift from fermentative to aerobic respiratory metabolism, depending upon the presence or absence of molecular oxygen. *Aerotolerant anaerobes* have a strictly fermentative metabolism but are relatively insensitive to the presence of molecular oxygen.

Nutrient and Growth Factor Requirements

Nutrients, rather than carbon or energy sources, may at times be the limiting material for microbial cell synthesis and growth. The principal inorganic nutrients needed by microorganisms are N, S, P, K, Mg, Ca, Fe, Na, and Cl. Minor nutrients of importance include Zn, Mn, Mo, Se, Co, Cu, and Ni (Madigan et al., 2012). Required organic nutrients, known as growth factors, are compounds needed by an organism as precursors or constituents of organic cell material, which cannot be synthesized from other carbon sources. Although growth factor requirements differ from one organism to another, the major growth factors fall into the following three classes: (1) amino acids, (2) nitrogen bases (i.e., purines and pyrimidines), and (3) vitamins.

For municipal wastewater treatment sufficient nutrients are generally present, but for industrial wastewaters nutrients may need to be added to the biological treatment processes. The lack of sufficient nitrogen and phosphorus is common especially in the treatment of food-processing wastewaters or wastewaters high in organic content. Using the formula $C_{12}H_{87}O_{23}N_{12}P$ (given previously) for the composition of cell biomass, about 12.2 g of nitrogen and 2.3 g of phosphorus are needed per 100 g of cell biomass.

7–4 BACTERIAL GROWTH, ENERGETICS, AND DECAY

In the description of microbial metabolism it was noted that as microorganisms consume substrate and carry out oxidation-reduction reactions, growth occurs by the production of additional cells. Thus, in wastewater treatment applications biomass is produced continuously as the substrate in the wastewater is consumed and biodegraded. Topics considered in this section include (1) bacterial reproduction, (2) bacterial growth patterns in a batch reactor, (3) bacterial growth and biomass yield, (4) methods used to measure biomass growth, (5) estimating cell yield and oxygen requirements from stoichiometry, (6) estimating cell yield from bioenergetics, and (7) observed versus synthesis yield. The material presented in this section will serve as a basis for the sections that follow and the material presented in Chaps. 8, 9, and 10 on the various treatment processes.

Bacterial Reproduction

Bacteria can reproduce, as noted in Chap. 2, by binary fission, by asexual mode, or by budding. Generally, bacteria reproduce by binary fission, in which the original cell becomes two new organisms. The time required for each division, which is termed the generation time, can vary from days to less than 20 min. For example, if the generation time is 30 min, one bacterium would yield 16,777,216 (i.e., 2^{24}) bacteria after a period of 12 h. Assuming spherical-shaped bacteria with a 1 μm diameter and specific gravity of 1.0, the weight of 1 cell is approximately 5.0×10^{-13} g. In 12 h the bacteria mass would be about 8.4×10^{-6} g or 8.4 μg; thus the number of cells is quite large compared to the mass. This rapid change in biomass with time is a hypothetical example. In biological treatment systems bacteria would not continue to divide indefinitely because of environmental limitations, such as substrate and nutrient availability.

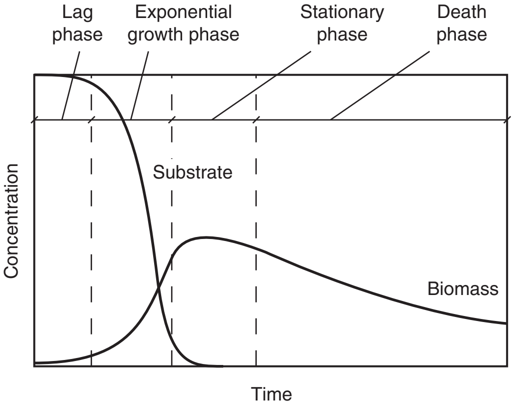
Bacterial Growth Patterns in a Batch Reactor

Bacterial growth in a batch reactor (see Sec. 1–7) is characterized by identifiable phases as illustrated on Fig. 7–10. The curves shown on Fig. 7–10 represent what occurs in a batch reactor in which, at time zero, substrate and nutrients are present in excess and only a very small population of biomass exists. As substrate is consumed, four distinct growth phases develop sequentially.

1. The lag phase. Upon addition of the biomass, the lag phase represents the time required for the organisms to acclimate to their new environment before significant cell division and biomass production occur. During the lag phase enzyme induction may be occurring and/or the cells may be acclimating to changes in salinity, pH, or temperature. The apparent extent of the lag phase may also be affected by the ability to measure the low biomass concentration during the initial batch phase.
2. The exponential-growth phase. During the exponential-growth phase, bacterial cells are multiplying at their maximum rate, as there is no limitation due to substrate

Figure 7–10

Batch process biomass growth phases with changes in substrate and biomass versus time.



or nutrients. The biomass growth curve increases exponentially during this period. With unlimited substrate and nutrients the only factor that affects the rate of exponential growth is temperature.

3. The stationary phase. During this phase, the biomass concentration remains relatively constant with time. In this phase, bacterial growth is no longer exponential, and the amount of growth is offset by the death of cells.
4. The death phase. In the death phase, the substrate has been depleted so that no growth is occurring, and the change in biomass concentration is due to cell death. An exponential decline in the biomass concentration is often observed as an approximate constant fraction of the biomass remaining that is lost each day.

Bacterial Growth and Biomass Yield

In biological treatment processes, cell growth occurs concurrent with the oxidation of organic or inorganic compounds, as described above. The ratio of the amount of biomass produced to the amount of substrate consumed (g biomass/g substrate) is defined as the *biomass synthesis yield* and typically is defined relative to the electron donor used.

$$\text{Biomass synthesis yield, } Y = \frac{\text{g biomass produced}}{\text{g substrate consumed}} \tag{7–2}$$

For example, for aerobic heterotrophic reactions with organic substrates, the yield is expressed as g biomass/g organic substrate; for nitrification the yield is expressed as g biomass/g NH_4-N oxidized; and for the anaerobic degradation of volatile fatty acids (VFAs) to produce methane, the yield is expressed as g biomass/g VFAs used. Where specific compounds are measured and known, such as ammonia, the yield is quantified relative to the amount of compound used. For aerobic or anaerobic treatment of municipal and industrial wastewater containing a large number of organic compounds, the yield is based on a measurable parameter reflecting the overall organic compound consumption, such as COD or BOD. Thus, the yield would be g biomass/g COD removed or g biomass/g BOD removed.

Measuring Biomass Growth

Because biomass is mostly organic material, an increase in biomass can be measured by volatile suspended solids (VSS) or particulate COD (total COD minus soluble COD). Other more direct parameters that are used to indicate biomass growth are protein content, DNA, and adenosine triphosphate (ATP), a cellular nucleotide compound involved in energy transfer. Of these growth measurement parameters, VSS is the parameter used most commonly to follow biomass growth in full-scale biological wastewater treatment systems because its measurement is simple, and minimal time is required for analysis. It should be noted that the VSS measured includes other particulate organic matter in addition to biomass. Most wastewaters contain some amount of nonbiodegradable VSS and possibly influent VSS that may be degraded slowly in the biological reactor. These solids are included with biomass in the VSS measurement. Nevertheless the VSS measurement is used as an apparent indicator of biomass production and also provides a useful measurement of reactor solids in general.

For laboratory research on biological treatment processes, growth parameters that can be related to true microbial mass are often used. Of these, protein is the most popular growth parameter due to the relative ease of measurement and the fact that about 50 percent of biomass dry weight is protein. Both ATP and DNA have also been used, especially where the reactor solids contain proteins and other solids that are not associated with biomass. Where very low biomass concentrations are involved, turbidity measurements may be used to provide a rapid and simple means of observing cell growth. Bacterial cell counts have also been used to enumerate the biomass population. A portion of a diluted sample is applied to an agar growth plate, and after incubation, the number of colonies formed are counted and used to determine the number of bacterial cells in the culture. It should be noted, however, that not all bacteria are culturable.

Estimating Biomass Yield and Oxygen Requirements from Stoichiometry

As given by Eq. (7–1), a definite stoichiometric relationship exists between the substrate removed, the amount of oxygen consumed during aerobic heterotrophic biodegradation, and the observed biomass yield. The most common approach used to define the fate of the substrate is to prepare a COD mass balance. The COD is used because the substrate concentration in the wastewater can be defined in terms of its oxygen equivalence, which can be accounted for by being conserved in the biomass or oxidized.

Biomass Yield. In general, the exact stoichiometry involved in the biological oxidation of a mixture of wastewater compounds is never known. However, for the purpose of illustration, it is assumed that organic matter can be represented as $C_6H_{12}O_6$ (glucose) and new cells can be represented as $C_5H_7NO_2$ (Hoover and Porges, 1952). Thus, neglecting nutrients other than nitrogen, Eq. (7–1) can be written as

$$\underset{3(180)}{3C_6H_{12}O_6} + \underset{8(32)}{8O_2} + \underset{2(17)}{2NH_3} \rightarrow \underset{2(113)}{2C_5H_7NO_2} + 8CO_2 + 14H_2O \tag{7–3}$$

As given by the above equation, the substrate used (glucose in this case) is divided between that found in new cells and that oxidized to produce CO_2 and H_2O. The yield based on the glucose consumed can be obtained as follows:

$$Y = \frac{\Delta(C_5H_7NO_2)}{\Delta(C_6H_{12}O_6)} = \frac{2(113\text{ g/mole})}{3(180\text{ g/mole})}$$
$$= 0.42\text{ g cells/g glucose used}$$

In practice, COD and VSS are used to represent the organic matter and the new cells, respectively. To express the yield on a COD basis, the COD of glucose must be determined. The COD of glucose can be determined by writing a balanced stoichiometric reaction for the oxidation of glucose to carbon dioxide as follows:

$$\underset{(180)}{C_6H_{12}O_6} + \underset{(32)}{6O_2} \rightarrow 6CO_2 + 6H_2O \tag{7–4}$$

The COD of glucose is

$$\text{COD} = \frac{\Delta(O_2)}{\Delta(C_6H_{12}O_6)} = \frac{6(32\text{ g/mole})}{(180\text{ g/mole})} = 1.07\text{ g }O_2\text{/g glucose}$$

The theoretical yield expressed in terms of COD, accounting for the portion of the substrate converted to new cells, is

$$Y = \frac{\Delta(C_5H_7NO_2)}{\Delta(C_6H_{12}O_6\text{ as COD})} = \frac{2(113\text{ g/mole})}{3(180\text{ g/mole})(1.07\text{ g COD/g glucose})}$$
$$= 0.39\text{ g cells/g COD used or }0.39\text{ g VSS/g COD used}$$

It should be noted that the actual observed yield in a biological treatment process will be less than the value given above, because a portion of the substrate incorporated into the cell mass will be later oxidized with time by the bacteria to obtain energy for cell maintenance.

Oxygen Requirements. The quantity of oxygen utilized can be accounted for by considering (1) the oxygen used for substrate oxidation to CO_2 and H_2O, (2) the COD of the biomass, and (3) the COD of any substrate not degraded. For the formula $C_5H_7NO_2$, the oxygen equivalent of the biomass (typically measured as VSS) is approximately 1.42 g COD/g biomass VSS, as given below.

$$\underset{(113)}{C_5H_7NO_2} + \underset{5(32)}{5O_2} \rightarrow 5CO_2 + NH_3 + 2H_2O \tag{7–5}$$

The COD of cell tissue is

$$\frac{\Delta(O_2)}{\Delta(C_5H_7NO_2)} = \frac{5(32\text{ g/mole})}{(113\text{ g/mole})} = 1.42\text{ g }O_2/\text{g cells} = 1.42\text{ g COD/g VSS}$$

Based on the above relationships, the oxygen consumed per unit of COD utilized for the reaction given by Eq. (7–3) can be determined from a mass balance on COD.
The COD removed is oxidized or accounted for in cell growth.

$$\text{COD}_r = \text{COD}_{\text{cells}} + \text{COD}_{ox} \tag{7–6}$$

where COD_r = COD utilized, g COD/d
$\text{COD}_{\text{cells}}$ = COD incorporated into cell synthesis, g COD/d
COD_{ox} = COD oxidized, g COD/d

The COD of oxidized substrate is equal to the oxygen consumed; thus

$$\text{Oxygen consumed} = \text{COD utilized} - \text{COD cells} \tag{7–7}$$

$$\begin{aligned} &= \left(\frac{1.07\text{ g }O_2}{\text{g glucose}}\right)\left(3\text{ mole} \times \frac{180\text{ g glucose}}{\text{mole}}\right) \\ &\quad - \left(\frac{1.42\text{ g }O_2}{\text{g cells}}\right)\left(2\text{ mole} \times \frac{113\text{ g cells}}{\text{mole}}\right) \\ &= 577.8\text{ g }O_2 - 320.9\text{ g }O_2 = 256.9\text{ g }O_2 \end{aligned}$$

Thus, the oxygen consumed per unit of COD used is

$$\begin{aligned} \frac{\text{Oxygen consumed}}{\text{Glucose as COD}} &= \frac{256.9\text{ g }O_2}{3\text{ mole}\,(1.07\text{ g COD/g glucose})(180\text{ g glucose/mole})} \\ &= 0.44\text{ g }O_2/\text{g COD used} \end{aligned}$$

The amount of oxygen required based on the COD balance, as given above, is in agreement with the oxygen use based on the stoichiometry as defined by Eq. (7–3) in which 8 moles of oxygen are required for 3 moles of glucose.

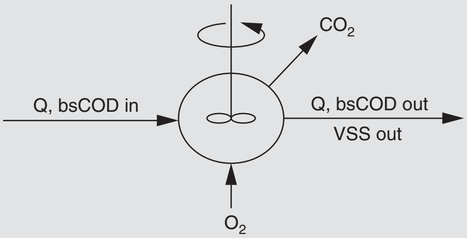
$$\begin{aligned} \frac{\text{Oxygen used}}{\text{Glucose as COD}} &= \frac{8(32\text{ g }O_2/\text{mole})}{3(180\text{ g/mole})(1.07\text{ g COD/g glucose})} \\ &= 0.44\text{ g }O_2/\text{g COD used} \end{aligned}$$

Note that the cell yield as 0.39 g VSS/gCOD used is equal to 0.56 g cells as COD/g COD used, and the sum of the g oxygen used for energy yielding oxidation/g COD used (0.44) plus the g cells as COD/g COD used = 1.0 g COD/g COD used. The COD is accounted for by the oxygen used and by the COD of biomass produced.

The relationship of the observed biomass yield to the oxygen consumed for substrate oxidation by aerobic heterotrophic biomass based on typical measurements made at wastewater-treatment plants is illustrated in Example 7–1.

EXAMPLE 7–1 **Observed Biomass Yield and Oxygen Consumption** An aerobic complete-mix biological treatment process without recycle, as shown below, receives wastewater with a biodegradable soluble COD (bsCOD) concentration of 500 g/m^3. The flowrate is 1000 m^3/d and the reactor effluent bsCOD and VSS concentrations are 10 and 200 g/m^3, respectively. Based on these data, determine the:

1. Observed yield in g VSS/g COD removed?
2. Amount of oxygen used in g O_2/g COD removed and in g/d?



Solution

1. Determine the observed yield. Assume the following general reaction is applicable.

$$\underset{500\ \text{g COD/m}^3}{\text{Organic matter}} + O_2 + \text{nutrients} \rightarrow \underset{200\ \text{g VSS/m}^3}{C_5H_7NO_2} + CO_2 + H_2O$$

   a. The g VSS/d produced is:

$$\text{g VSS/d} = 200\ \text{g/m}^3\ (1000\ \text{m}^3/\text{d}) = 200{,}000\ \text{g VSS/d}$$

   b. The g bsCOD/d removed is:

$$\begin{aligned}\text{g COD/d} &= (500 - 10)\ \text{g COD/m}^3\ (1000\ \text{m}^3/\text{d})\\ &= 490{,}000\ \text{g COD/d}\end{aligned}$$

   c. The observed yield is:

$$Y_{obs} = \frac{(200{,}000\ \text{g VSS/d})}{(490{,}000\ \text{g COD/d})} = 0.41\ \text{g VSS/g COD removed}$$

2. Determine the amount of oxygen used per g bsCOD removed.
   a. Prepare a steady-state COD mass balance around the reactor

$$\text{accumulation} = \text{inflow} - \text{outflow} + \text{conversion}$$

$$0 = \text{COD}_{in} - \text{COD}_{out} - \text{oxygen used (expressed as COD)}$$

$$\text{Oxygen used} = \text{COD}_{in} - \text{COD}_{out}$$

$$\text{COD}_{in} = 500\ \text{g COD/m}^3\ (1000\ \text{m}^3/\text{d}) = 500{,}000\ \text{g COD/d}$$

$$\text{COD}_{out} = \text{bsCOD}_{out} + \text{biomass COD}_{out}$$

$$\text{bsCOD}_{out} = 10\ \text{g/m}^3\ (1000\ \text{m}^3/\text{d}) = 10{,}000\ \text{g COD/d}$$

$$\text{Biomass COD}_{\text{out}} = 200{,}000\ \text{g VSS/d}\,(1.42\ \text{g COD/g VSS})$$
$$= 284{,}000\ \text{g COD/d}$$

$$\text{Total COD}_{\text{out}} = 10{,}000\ \text{g/d} + 284{,}000\ \text{g/d} = 294{,}000\ \text{g COD/d}$$

b. The oxygen used is

$$\text{Oxygen used} = 500{,}000\ \text{g COD/d} - 294{,}000\ \text{g COD/d}$$
$$= 206{,}000\ \text{g COD/d} = 206{,}000\ \text{g O}_2\text{/d}$$

c. Amount of oxygen used per unit COD removed is

$$\text{Oxygen/COD} = (206{,}000\ \text{g/d})/(490{,}000\ \text{g/d}) = 0.42\ \text{g O}_2\text{/g COD}$$

Comment The general COD balance that accounts for the cell yield fraction of COD oxidized is

$$\text{g COD cells} + \text{g COD oxidized} = \text{g COD removed}$$

$$(0.41\ \text{g VSS/g COD})(1.42\ \text{g O}_2\text{/g VSS}) + 0.42\ \text{g O}_2\text{/g COD} = 1.0\ \text{g O}_2\text{/g COD}$$

Estimating Biomass Yield from Bioenergetics

Most cell yield values are based on measurements from laboratory reactors, pilot plants, or full-scale systems. However, an approach that has been developed to estimate cell yield, based on bioenergetics, involves the application of thermodynamic principles to biological reactions. An introduction to bioenergetics and the application of bioenergetics to estimate the biomass yield for different types of biological reactions is provided in this section (McCarty 1971, 1975).

Gibbs Free Energy. Chemical reactions, which involve changes in energy, can be described thermodynamically by a change in the free energy G°, known as the *Gibbs free energy*. The change in energy due to the reaction is termed ΔG°. The superscript is used to indicate that the free energy values were obtained at standard conditions of $\text{pH} = 7.0$ and 25°C. The net Gibbs free energy, positive or negative, can be evaluated for reactants and products based on standard free energy values available for the half reactions. Half reactions describe the transfer of 1 mole of electron in oxidation-reduction and synthesis reactions. Free energy changes for various half reactions are listed in Table 7–6. Reactions that result in a negative change in the free energy are those that release energy and are called *exergonic* reactions. These reactions will proceed spontaneously in the direction shown. However, if the free energy change results in a positive value, the reaction is termed *endergonic*, and such a reaction will not occur spontaneously. Reactions with positive free energy values require energy to proceed in the direction indicated.

The basis of the analysis of free energy changes for reactions is that in oxidation-reduction reactions one compound loses electrons (electron donor) and the other compound gains electrons (electron acceptor) (see discussion in Sec. 6–6 in Chap. 6). The moles of compound used as an electron acceptor and electron donor per mole of electron (e^-) transferred from the half reactions along with the change in free energy are used to develop energy balances in bioenergetic analyses. Determination of the free energy change resulting from the oxidation of hydrogen by oxygen is illustrated in Example 7–2.

Table 7–6

Half-reactions for biological systems[a]

Reaction number		Half-reaction	ΔG^o (W),[b] kJ per electron equivalent
Reactions for bacterial cell synthesis (R_s)			
	Ammonia as nitrogen source:		
1.	$\frac{1}{5}CO_2 + \frac{1}{20}HCO_3^- + \frac{1}{20}NH_4^+ + H^+ + e^-$	$= \frac{1}{20}C_5H_7O_2N + \frac{9}{20}H_2O$	
	Nitrate as nitrogen source:		
2.	$\frac{1}{28}NO_3^- + \frac{5}{28}CO_2 + \frac{29}{28}H^+ + e^-$	$= \frac{1}{28}C_5H_7O_2N + \frac{11}{28}H_2O$	
Reactions for electron acceptors (R_a)			
	Nitrite:		
3.	$\frac{1}{3}NO_2^{2-} + \frac{4}{3}H^+ + e^-$	$= \frac{1}{6}N_2 + \frac{2}{3}H_2O$	−93.23
	Oxygen:		
4.	$\frac{1}{4}O_2 + H^+ + e^-$	$= \frac{1}{2}H_2O$	−78.14
	Nitrate:		
5.	$\frac{1}{5}NO_3^- + \frac{6}{5}H^+ + e^-$	$= \frac{1}{10}N_2 + \frac{3}{5}H_2O$	−71.67
	Sulfite:		
6.	$\frac{1}{6}SO_3^{2-} + \frac{5}{4}H^+ + e^-$	$= \frac{1}{12}H_2S + \frac{1}{12}HS^- + \frac{1}{2}H_2O$	13.60
	Sulfate:		
7.	$\frac{1}{8}SO_4^{2-} + \frac{19}{16}H^+ + e^-$	$= \frac{1}{16}H_2S + \frac{1}{16}HS^- + \frac{1}{2}H_2O$	21.27
8.	$\frac{1}{8}CO_2 + H^+ + e^-$	$= \frac{1}{8}CH_4 + \frac{1}{4}H_2O$	24.11
Reactions for electron donors (R_d)			
	Organic donors (heterotrophic reactions)		
	Domestic wastewater:		
9.	$\frac{9}{50}CO_2 + \frac{1}{50}NH_4^+ + \frac{1}{50}HCO_3^- + H^+ + e^-$	$= \frac{1}{50}C_{10}H_{19}O_3N + \frac{9}{25}H_2O$	31.80
	Protein (amino acids, proteins, nitrogenous organics)		
10.	$\frac{8}{33}CO_2 + \frac{2}{33}NH_4^+ + \frac{31}{33}H^+ + e^-$	$= \frac{1}{66}C_{16}H_{24}O_5N_4 + \frac{27}{66}H_2O$	32.22
	Formate		
11.	$\frac{1}{2}HCO_3^- + H^+ + e^-$	$= \frac{1}{2}HCOO^- + \frac{1}{2}H_2O$	48.07
	Glucose		
12.	$\frac{1}{4}CO_2 + H^+ + e^-$	$= \frac{1}{24}C_6H_{12}O_6 + \frac{1}{4}H_2O$	41.96
	Carbohydrate (cellulose, starch, sugars):		
13.	$\frac{1}{4}CO_2 + H^+ + e^-$	$= \frac{1}{4}CH_2O + \frac{1}{4}H_2O$	41.84

(*Continued*)

Table 7–6 (*Continued*)

Reaction number	Half-reaction	ΔG° (W),[b] kJ per electron equivalent
	Methanol:	
14.	$\frac{1}{6}CO_2 + H^+ + e^- = \frac{1}{6}CH_3OH + \frac{1}{6}H_2O$	37.51
	Pyruvate	
15.	$\frac{1}{5}CO_2 + \frac{1}{10}HCO_3 + H^+ + e^- = \frac{1}{10}CH_3COCOO^- + \frac{2}{5}H_2O$	35.78
	Ethanol:	
16.	$\frac{1}{6}CO_2 + H^+ + e^- = \frac{1}{12}CH_3CH_2OH + \frac{1}{4}H_2O$	31.79
	Propionate:	
17.	$\frac{1}{7}CO_2 + \frac{1}{14}HCO_3^- + H^+ + e^- = \frac{1}{14}CH_3CH_2COO^- + \frac{5}{14}H_2O$	27.91
	Acetate:	
18.	$\frac{1}{8}CO_2 + \frac{1}{8}HCO_3 + H^+ + e^- = \frac{1}{8}CH_3COO^- + \frac{3}{8}H_2O$	27.68
	Grease (fats and oils):	
19.	$\frac{4}{23}CO_2 + H^+ + e^- = \frac{1}{46}C_8H_{16}O + \frac{15}{46}H_2O$	27.61
	Inorganic donors (autotrophic reactions):	
20.	$Fe^{3+} + e^- = Fe^{2+}$	−74.40
21.	$\frac{1}{2}NO_3^- + H^+ + e^- = \frac{1}{2}NO_2^- + \frac{1}{2}H_2O$	−40.15
22.	$\frac{1}{8}NO_3^- + \frac{5}{4}H^+ + e^- = \frac{1}{8}NH_4^+ + \frac{3}{8}H_2O$	−34.50
23.	$\frac{1}{6}NO_2^- + \frac{4}{3}H^+ + e^- = \frac{1}{6}NH_4^+ + \frac{1}{3}H_2O$	−32.62
24.	$\frac{1}{6}SO_4^{2-} + \frac{4}{3}H^+ + e^- = \frac{1}{6}S + \frac{2}{3}H_2O$	19.48
25.	$\frac{1}{8}SO_4^{2-} + \frac{19}{16}H^+ + e^- = \frac{1}{16}H_2S + \frac{1}{16}HS^- + \frac{1}{2}H_2O$	21.28
26.	$\frac{1}{4}SO_4^{2-} + \frac{5}{4}H^+ + e^- = \frac{1}{8}S_2O_3^{2-} + \frac{5}{8}H_2O$	21.30
27.	$\frac{1}{6}N_2 + \frac{4}{3}H^+ + e^- = \frac{1}{3}NH_4^+$	27.47
28.	$H^+ + e^- = \frac{1}{2}H_2$	40.46
29.	$\frac{1}{2}SO_4^{2-} + H^+ + e^- = SO_3^{2-} + H_2O$	44.33

[a] Adapted from McCarty (1975) and Sawyer et al. (2003).

[b] Reactants and products at unit activity except $[H^+] = 10^{-7}$.

EXAMPLE 7–2 Free Energy Change from Hydrogen Oxidation by Molecular Oxygen

Determine the free energy change resulting from the oxidation of hydrogen by molecular oxygen.

Solution

1. Identify the electron donor and acceptor.
   Electron donor: hydrogen
   Electron acceptor: oxygen

2. Determine the change in free energy. From Table 7–6, the free energy change values for the half reactions are

		G^o, kJ/mole e^-
Reaction No. 28	$\frac{1}{2}H_2 \rightarrow H^+ + e^-$	-40.46
Reaction No. 4	$\frac{1}{4}O_2 + H^+ + e^- \rightarrow \frac{1}{2}H_2O$	-78.14
Overall	$\frac{1}{2}H_2 + \frac{1}{4}O_2 \rightarrow \frac{1}{2}H_2O$	$\Delta G = -118.60$

Because the ΔG value is negative for this oxidation-reduction reaction, energy is released, and the overall reaction will proceed as written.

Exergonic Reactions. Exergonic reactions are catalyzed by enzymes within the microbial cell, making energy available to support cell growth. Only a portion (40 to 80 percent) of the energy produced is captured by the bacteria, while the rest escapes as heat. McCarty (1971) assumed 60 percent energy capture efficiency, but the exact amount varies. The energy that is not captured or released can result in an elevated temperature in the surrounding liquid, where high biomass concentrations exist and high reaction rates are occurring. An example is autothermal aerobic digestion in which liquid temperatures are increased from 20°C to as high as 60°C due to biological oxidation and energy release. The volatile solids concentration in autothermal aerobic digesters may be in the range of 20 to 40 g/L.

Bioenergetic Methods of Analysis. The key steps in bioenergetics analysis are to (1) identify the electron donor (substrate oxidized) and electron acceptor, (2) determine the energy produced from the bacteria oxidation-reduction reaction, (3) determine the amount of energy needed for converting the growth carbon source into cell matter, and (4) calculate the cell yield based on a balance between energy produced and energy needed for cell yield. The energy production step was illustrated above for oxidation of hydrogen with oxygen as the electron acceptor.

The amount of energy required for cell synthesis depends on the specific carbon and nitrogen compounds used for growth. The bioenergetics analysis discussed here is for heterotrophic bacteria. A different procedure is used for the case of autotrophic bacteria, and additional details can be found in McCarty (1971, 1975) and Rittman and McCarty (2001). For heterotrophic bacteria, many carbon sources may be used for growth with different energetic effects. In the analysis it is assumed that pyruvate is the intermediate organic compound used for cell synthesis, and energy will be either produced or consumed depending on the free energy of the organic compound relative to pyruvate. Pyruvate was selected by McCarty (1971) because it is at the end of the glycolysis pathway and just ahead of the Krebs cycle. When CO_2 is used for the carbon source, as for autotrophic bacteria, a considerable amount of energy is needed to incorporate CO_2 into cell mass. If nitrogen is not available in the form of ammonia, additional energy is needed to convert the nitrogen source to ammonia.

The energy required for cell synthesis is estimated as follows, with pyruvate as the organic intermediate for cellular carbon constituents:

$$\Delta G_s = \frac{\Delta G_P}{K^m} + \Delta G_c + \frac{\Delta G_N}{K} \tag{7–8}$$

where ΔG_s = free energy to convert 1 electron equivalent (e^- eq) of the carbon source to cell material

ΔG_P = free energy to convert 1 e^- eq of the carbon source to the pyruvate intermediate

K = fraction of energy transfer captured

m = $+1$ if ΔG_P is positive and -1 if energy is produced

ΔG_c = free energy to convert 1 e^- eq of pyruvate intermediate to 1 e^- eq of cells

ΔG_N = free energy per e^- eq of cells to reduce nitrogen to ammonia

The value for ΔG_c is $+31.41$ kJ/e^- eq of cells (McCarty, 1971) and the ΔG_N for the following nitrogen sources are $+17.46$, $+13.61$, $+15.85$, and 0.00 kJ/e^- eq of cells for NO_3^-, NO_2^-, N_2, and NH_4^+, respectively. The value for ΔG_P is estimated by using the free energy half reactions to convert the carbon source to the pyruvate intermediate.

The electron donor used for heterotrophic reactions is divided between portions oxidized to produce energy or used in cell synthesis. The energy balance relative to the substrate used is illustrated in the following equation in which the energy made available (on the left side of the equation) equals the energy used for cell growth (right side of the equation).

$$K\Delta G_R\left(\frac{f_e}{f_s}\right) = -\Delta G_S \tag{7–9}$$

$$f_e + f_s = 1 \tag{7–10}$$

where K = fraction of energy captured

ΔG_R = energy released from oxidation-reduction reactions, kJ/mole e^- transferred

f_e = e^- mole of substrate oxidized per e^- mole of substrate used

f_s = e^- mole of substrate used for cell synthesis per e^- mole of substrate used

ΔG_S = energy used for cell growth, kJ/mole e^- transfer for cell growth

Equations (7–9) and (7–10) are used with half reactions and their free energy values to estimate cell yield by solving for f_e and f_s. The terms f_e and f_s represent the fraction of substrate oxidized or used in cell synthesis, respectively. The substrate is expressed as COD because a mole of COD contains a set quantity of electron moles of oxygen transfer. Thus, the values for f_e and f_s can also be expressed as COD fractions. The oxidation of acetate by heterotrophic bacteria with different acceptors is used in Example 7–3 to illustrate the bioenergetics analysis.

EXAMPLE 7–3 Estimate Biomass Yield Using Energetics

Compare the cell yield in g cells as COD/g COD used and g cells as VSS/g COD used for acetate utilization by heterotrophic bacteria with oxygen and then carbon dioxide as the electron acceptor, and ammonia as the nitrogen source. Assume 60 percent energy capture efficiency.

Solution

Part A—Oxygen as the electron acceptor

1. Solve for the energy produced and captured ($K\Delta G_R$) using reaction No. 18 for acetate oxidation and reaction No. 4 for oxygen reduction from Table 7–6.

		kJ/mole e^-
No. 18	$\frac{1}{8}CH_3COO^- + \frac{3}{8}H_2O \rightarrow \frac{1}{8}CO_2 + \frac{1}{8}HCO_3^- + H^+ + e^-$	−27.68
No. 4	$\frac{1}{4}O_2 + H^+ + e^- \rightarrow H_2O$	−78.14
	$\frac{1}{8}CH_3COO^- + \frac{3}{8}H_2O \rightarrow \frac{1}{8}CO_2 + \frac{1}{8}HCO_3^- + \frac{1}{8}H_2O$	$\Delta G = -105.82$

Energy captured by cell:

$K(\Delta G_R) = 0.60(-105.82) = -63.42$ kJ/mole e^-

2. Solve for the energy needed per electron mole of cell growth (ΔG_S).

$\Delta G_C = 31.41$ kJ/mole e^- cells

$\Delta G_N = 0$

ΔG_P [acetate (reaction No. 18) to pyruvate (reaction No. 15)]

		ΔG kJ/mole e^-
No. 18	$\frac{1}{8}CH_3COO^- + \frac{3}{8}H_2O \rightarrow \frac{1}{8}CO_2 + \frac{1}{8}HCO_3^- + H^+ + e^-$	−27.68
No. 15	$\frac{1}{5}CO_2 + \frac{1}{10}HCO_3^- + H^+ + e^- \rightarrow \frac{1}{10}CH_3COCOO^- + \frac{2}{5}H_2O$	+35.78
	$\frac{1}{8}CH_3COO^- + \frac{3}{40}CO_2 \rightarrow \frac{1}{10}CH_3COCOO^- + \frac{1}{40}HCO_3^- + \frac{1}{40}H_2O$	$\Delta G_p = +8.10$

Because ΔG_P is positive, energy is required and m = +1

$$\Delta G_S = \left[\frac{+8.10}{(0.6)^{1.0}} + 31.41 + 0\right] = 44.91 \text{ kJ/mole e}^-$$

3. Determine f_e and f_s using Eq. (7–9).

$$\frac{f_e}{f_s} = \frac{-\Delta G_S}{K\Delta G_R} = \frac{(-44.91 \text{ kJ/mole e}^-)}{(-63.42 \text{ kJ/mole e}^-)}$$

$$\frac{f_e}{f_s} = 0.708$$

$$f_e + f_s = 1.0$$

Solve for f_e and f_s:

$$f_e = 0.41$$

$$f_s = 0.59 \frac{\text{g cell COD}}{\text{g COD used}}$$

4. Determine the yield based on COD.

For biomass ($C_5H_7NO_2$), 1 g cells = 1.42 g COD

Thus, the yield is

$$Y = \frac{(0.59 \text{ g COD/g COD})}{(1.42 \text{ g COD/g VSS})} = 0.42 \text{ g VSS/g COD}$$

5. Determine the yield based on BOD assuming a conversion factor of 1.6 g COD/g BOD (see discussion in Sec. 8–2 in Chap. 8).

Thus, the yield is

$$Y = \frac{(0.42 \text{ g VSS/g COD})}{(\text{g BOD}/1.6 \text{ g COD})} = 0.67 \text{ g VSS/g BOD}$$

Solution Part B—Carbon dioxide as the electron acceptor

1. Solve for energy produced and captured ($K\Delta G_R$) using reaction No. 18 for acetate oxidation and reaction No. 8 for CO_2 reduction to methane from Table 7–6.

		kJ/mole e^-
No. 18	$\frac{1}{8}CH_3COO^- + \frac{3}{8}H_2O \rightarrow \frac{1}{8}CO_2 + \frac{1}{8}HCO_3^- + H^+ + e^-$	-27.68
No. 8	$\frac{1}{8}CO_2 + H^+ + e^- \rightarrow \frac{1}{8}CH_4 + \frac{1}{4}HCO_3^-$	$+24.11$
	$\frac{1}{8}CH3COO^- + \frac{1}{8}H_2O \rightarrow \frac{1}{8}CH_4 + \frac{1}{8}HCO_3^-$	$\Delta G = -3.57$

Energy captured by cell:

$(K\Delta G_R) = 0.60(-3.57) = -2.142$ kJ/mole e^-

2. Solve for the energy needed per electron mole at cell growth ΔG_S.

$\Delta G_C = 31.41$ kJ/mole e^- cells

$\Delta G_N = 0$

ΔG_P (same as for acetate/O_2)

$\Delta G_S = 44.94$ kJ/mole e^-

3. Determine the values of fe and fs using Eq. (7–9).

$$\frac{f_e}{f_s} = \frac{-\Delta G_S}{K\Delta G_R} = \frac{-44.94}{-2.142} = 21.0$$

$f_e + f_s = 1.0$

Solve for f_e and f_s.

$f_e = 0.954 \qquad f_s = 0.046$ g cell COD/g COD used

4. Determine the yield based on COD.

$$Y = \frac{(0.046 \text{ g cell COD/g COD used})}{(1.42 \text{ g COD/g VSS})} = 0.032 \text{ g VSS/g COD}$$

5. Compare yields for acetate oxidation.

Electron acceptor	Yield, g VSS/g COD	Product
O_2	0.42	CO_2, H_2O
CO_2	0.032	CH_4

Comment Based on bioenergetics calculations, estimated yield values for anaerobic reactions using carbon dioxide as the electron acceptor are much lower as compared to oxygen. The lower yield values are due to the much lower energy production with carbon dioxide as the electron acceptor in lieu of oxygen. The cell synthesis yield values calculated for these electron acceptors are very similar to yield values reported in the literature.

Stoichiometry of Biological Reactions

With the values determined for f_e and f_s, the stoichiometry of the biological reactions can be described according to the following relationship (McCarty, 1971, 1975):

$$R = f_e R_a + f_s R_{CS} - R_d \tag{7–11}$$

where R = overall balanced reaction
f_e = fraction of electron donor used for energy
R_a = half reaction for electron acceptor
f_s = fraction of electron donor used for cell synthesis
R_{cs} = half reaction for synthesis of cell tissue
R_d = half reaction for electron donor
$f_s + f_e = 1$

The minus sign in Eq. (7–11) means that the electron donor equation given in Table 7–6 must be reversed and then added to the other two equations. In the first equation given in Table 7–6, the term $C_5H_7O_2N$ (Hoover and Porges, 1952), is used to represent bacterial cell tissue. Application of Eq. (7–11) is illustrated in Example 7–4.

EXAMPLE 7–4 **Write a Balanced Reaction for the Biological Oxidation of Acetate Using Oxygen** Using Eq. (7–11) and the half reactions given in Table 7–6, write a balanced reaction for the biological oxidation of acetate with oxygen. Use the values for f_e and f_s determined in Example 7–3, Part A ($f_e = 0.41$ and $f_s = 0.59$). The COD of acetate is 1.07 g COD/g acetate, computed as illustrated in Eq. (7–4).

Solution

1. Develop the balanced stoichiometric reaction for the oxidation of acetate.

$$R = f_e R_a + f_s R_{CS} - R_d$$

$$R = 0.41(\text{No. } 4) + 0.59(\text{No. } 1) - \text{No. } 18$$

$$(0.41)(\text{No. } 4) = 0.103O_2 + 0.41H^+ + 0.4e^- + 0.205H_2O$$

$$(0.59)(\text{No. } 1) = 0.118CO_2 + 0.0295HCO_3^- + 0.0295NH_4^+ + 0.59H^+ + 0.59e^-$$
$$\rightarrow 0.0295C_5H_7O_2N + 0.2655H_2O$$

$$-\text{No. } 18 = \quad 0.125CH_3COO^- + 0.375H_2O \rightarrow 0.125CO_2 + 0.125HCO_3^- + H^+ + e^-$$

$$R = \quad 0.125CH_3COO^- + 0.0295NH_4^+ + 0.103O_2$$
$$\rightarrow 0.0295C_5H_7O_2N + 0.0955H_2O + 0.095HCO_3^- + 0.007CO_2$$

2. Determine the cell yield from the stoichiometry.
   a. Cells produced from oxidation of acetate

      Cells produced = 0.0295 mole (113 g VSS/mole) = 3.334 g VSS

   b. Acetate utilized for cell production

      Acetate used = 0.125 mole acetate (60 g/mole)(1.07 g COD/g acetate)
      = 8.03 g COD

c. Determine the cell yield

$$Y = \frac{3.334 \text{ g VSS}}{8.03 \text{ g COD}} = 0.42 \text{ g VSS/g COD}$$

The results are the same as in Example 7–3.

If nitrate (NO_3^-) is used for the electron acceptor instead of oxygen for acetate degradation, denitrification occurs with the NO_3^- reduced to nitrogen gas (N_2). The energetics calculation using the same procedure as above results in an f_s value of 0.57, only slightly less than that for oxygen as the electron acceptor (Rittman and McCarty, 2001). This f_s value leads to a biomass yield coefficient determination very close to that with oxygen (0.41 versus 0.42 g VSS/g COD). However, Muller et al. (2004) showed that the heterotrophic bacteria yield coefficient under anoxic conditions is much lower, approximately 80 percent of the aerobic yield. A similar ratio is recommended in the IWA ASM3 model (Gujer et al., 1999). In denitrification studies with pure cultures of *Paracoccus denitrificans* and *Pseudomonas stutzeri*, Strohm et al. (2007) found that much less energy is converted via ATP into cell mass than expected from energetic calculations. For acetate degradation with NO_3^- as the electron acceptor a cell yield of 0.28 to 0.32 g VSS/g COD was observed, which is 67 to 78 percent of that predicted by the above energetic calculation. Henze et al. (2008) applied a lower biomass yield by about 70 percent for NO_3^- reduction versus aerobic degradation in considering carbon addition for nitrate removal.

Biomass Synthesis Yields for Different Growth Conditions

In Example 7–3, it was demonstrated that the biomass synthesis yield is related to the energy produced by the electron transfer from the electron donor (acetate) to the electron acceptor (oxygen). From a review of the half reaction ΔG° values in Table 7–6, it can be seen that the energy production that occurs from the oxidation-reduction reactions is reduced as the electron acceptor is changed from oxygen to nitrate to sulfate and to carbon dioxide. Accordingly, a lower cell yield would be predicted using the half reactions given in Table 7–6. A range of typical synthesis yield coefficients are given in Table 7–7 for common electron donors and acceptors in wastewater treatment.

Biomass Decay

As shown for the batch process in Fig. 7–10, after the substrate is depleted and with continual aeration, the biomass concentration decreases. Terms used to define this biomass

Table 7–7

Typical bacteria synthesis yield coefficients for common biological reactions in wastewater treatment

Growth Condition	Electron Donor	Electron acceptor	Synthesis yield
Aerobic	Organic compound	Oxygen	0.45 g VSS/g COD
Aerobic	Ammonia	Oxygen	0.12 g VSS/g NH_4-N
Anoxic	Organic compound	Nitrate	0.30 g VSS/g COD
Anaerobic	Organic compound	Organic compound	0.06 g VSS/g COD
Anaerobic	Acetate	Carbon dioxide	0.05 g VSS/g COD

decrease are *decay*, *endogenous decay*, or *endogenous respiration*. A number of factors contribute to the biomass loss, including cell maintenance energy needs, cell lysis due to death or stress from environmental factors, and predation (Hao et al., 2010). Examples of maintenance energy needs are energy for resynthesis of essential cell components and for controlling osmotic pressure. Energy may come from oxidation of cellular organic material or substrates from the bulk liquid. If substrate is not available, the cell mass may decrease and cellular activity can approach a resting state. Cell death and lysis can be due to virus, environmental stress (pH, temperature and toxins), or to program cell death related to cell age (Rice and Bayle 2008).

Cell lysis results in the release of cellular substrates, which are mainly biodegradable proteins and polysaccharides. In addition, cell debris, which remains after cell lysis, has generally been assumed as inert volatile solids that accumulate in the bioreactor, but more recently cell debris has been found to have a very slow biodegradation rate (Ramdani et al., 2012). Protozoan and rotifers consume bacteria that are not well captured in activated sludge floc. Predation losses by protozoan and rotifers are greater for systems with free bacteria or more dispersed growth conditions. The parameter, specific endogenous decay, is commonly used to account for all of these biomass loss mechanisms, which may be occurring to a varying degree, depending on cell age and operating conditions. The specific endogenous decay coefficient is used to describe a specific rate of biomass loss, which can be in the range from 0.08 to 0.20 g biomass loss/g biomass present.

Observed versus Synthesis Yield

In the evaluation and modeling of biological treatment systems a distinction is made between the observed yield and the synthesis yield (or true yield). The observed biomass yield is based on the actual measurements of the net biomass production and substrate consumption and is actually less than the synthesis yield, because of cell loss by biomass decay (Sec. 7–6) concurrent with cell growth. In full-scale wastewater treatment processes the term solids production (or solids yield) is also used to describe the amount of VSS generated in the treatment process. The term is different from the synthesis biomass yield values because it contains other organic solids from the wastewater that are measured as VSS and have not been biologically degraded.

The synthesis yield is the amount of biomass produced immediately upon consumption of the growth substrate or oxidation of the electron donor in the case of autotrophic bacteria. The synthesis yield is seldom measured directly and is often interpreted from evaluating biomass production data for reactors operating under different conditions. Synthesis yield values for bacterial growth are affected by the energy that can be derived from the oxidation-reduction reaction, by the growth characteristics of the carbon source, by the nitrogen source, and by environmental factors such as temperature, pH, and osmotic pressure. As illustrated in this section, the synthesis yield can be estimated if the stoichiometry or the amount of energy produced in the oxidation-reduction reaction is known.

7–5 MICROBIAL GROWTH KINETICS

The performance of biological processes used for wastewater treatment depends on the dynamics of substrate utilization and microbial growth. Effective design and operation of such systems requires an understanding of the biological reactions occurring and an understanding of the basic principles governing the growth of microorganisms. Further, the need to understand all of the environmental conditions that affect the substrate utilization and microbial growth rate cannot be overemphasized, and it may be necessary to control such conditions as pH, dissolved oxygen (DO), and nutrients to provide effective treatment.

The purpose of this section is to present an introduction to microbial growth kinetics. The topics considered in this section include (1) microbial growth kinetics terminology, (2) rate of utilization of soluble substrate, (3) other rate expressions for the utilization of soluble substrate, (4) rate of soluble substrate production from biodegradable particulate organic matter, (5) the rate of biomass growth with soluble substrates, (6) kinetic coefficients for substrate utilization and biomass growth, (7) oxygen utilization rates, (8) effects of temperature, (9) total volatile suspended solids and active biomass, (10) rate of biomass decay and soluble substrate production, and (11) net biomass growth and observed yield.

Microbial Growth Kinetics Terminology

The kinetics of microbial growth govern the oxidation (i.e., utilization) of substrate and the production of biomass, which contributes to the total suspended solids concentration in a biological reactor. Common terms used to describe the transformations that occur in substrate oxidation and biomass growth as well as in biological treatment processes are defined in the Working Terminology. Because municipal and industrial wastewaters contain numerous substrates, the concentration of organic compounds is defined, most commonly, by the *biodegradable COD* (bCOD) or UBOD, both of which are comprised of soluble (dissolved), colloidal, and particulate biodegradable components. Both bCOD and UBOD represent measurable quantities that apply to all of the compounds. In the formulation of kinetic expressions in this chapter *biodegradable soluble COD* (bsCOD) will be used to quantify the fate of biodegradable organic compounds because it is related to the stoichiometry of substrate oxidized or used in cell growth [see Eq. (7–7)]. Some of the biodegradable COD in wastewater is in particulate or colloidal form, which must be hydrolyzed to bsCOD first before biological consumption. Kinetic expressions are also used to describe the rate of bsCOD production from particulate and colloidal material.

The biomass solids in a laboratory bioreactor fed only bsCOD are commonly measured as *total suspended solids* (TSS) and *volatile suspended solids* (VSS). These measurements are not equal to the active biomass, as the solids will contain cell debris material from bacteria decay.

In activated sludge systems used to treat municipal and industrial wastewater, the mixture of solids resulting from combining recycled sludge with influent wastewater in the bioreactor is termed *mixed liquor suspended solids* (MLSS) and *mixed liquor volatile suspended solids* (MLVSS). The solids are comprised of biomass, *nonbiodegradable volatile suspended solids* (nbVSS), and *inert inorganic total suspended solids* (iTSS). The nbVSS is derived from the influent wastewater and is also produced as cell debris from bacteria decay. The iTSS originates in the influent wastewater. Additional wastewater characterization terminology is considered in Sec. 8–2 in Chap. 8.

Rate of Utilization of Soluble Substrates

In the introduction to this chapter, it was noted that one of the principal concerns in wastewater treatment is the removal of substrate. Stated another way, the goal in biological wastewater treatment is, in most cases, to reduce the electron donor (i.e., organic compounds in aerobic oxidation) to minimal levels. For heterotrophic bacteria the electron donors are the organic substances being degraded; for autotrophic nitrifying bacteria it is ammonia or nitrite or other reduced inorganic compounds. The substrate utilization rate for soluble substrates in biological systems can be modeled with Eq. (7–12), in which the rate of substrate utilization increases as the reactor substrate concentration increases for a given biomass concentration.

$$r_{su} = \frac{kXS}{K_s + S} \tag{7–12}$$

where r_{su} = substrate utilization rate per unit of reactor volume, g/m^3·d
k = maximum specific substrate utilization rate, g substrate/g microorganisms·d
X = biomass (microorganism) concentration, g/m^3
S = growth-limiting substrate concentration in solution, g/m^3
K_s = half-velocity constant, substrate concentration at one-half the maximum specific substrate utilization rate, g/m^3

Equation (7–12) will be recognized as a saturation-type equation, as described previously in Chap. 1. Though Eq. (7–12) is an empirical model, based on using coefficients derived from biological reactor data, it is often referred to as the Michaelis-Menten equation (Bailey and Ollis, 1986), which was formulated from an enzymatic-substate model. A plot of r_{su} versus the substrate concentration is shown on Fig. 7–11. As shown on Fig. 7–11, the maximum substrate utilization rate occurs at high substrate concentrations. Further, as the substrate concentration decreases below some critical value, the value of r_{su} also decreases almost linearly. In practice, biological treatment systems are designed to produce an effluent with extremely low substrate concentrations.

Equation (7–12) is also of the form proposed by Monod for the specific growth rate of bacteria in which the limiting substrate is available to the microorganisms in a dissolved form (Monod, 1942, 1949).

$$r_g = \frac{\mu_m XS}{(K_s + S)} \tag{7–13}$$

where r_g = bacteria growth rate from substrate utilization, g/m^3·d
μ_m = maximum specific bacteria growth rate, g biomass/g biomass·d

As bacteria consume the substrate (electron donor), the energy produced from the substrate oxidation is used to process carbon and nutrients to produce new biomass. The new growth is directly proportional to the substrate used and is termed the true yield or synthesis yield coefficient as g biomass produced/g substrate used. The bacteria growth rate can thus be equated to the substrate utilization rate as follows:

$$r_g = Yr_{su} \tag{7–14}$$

and $$r_{su} = \frac{\mu_m XS}{Y(K_s + S)} \tag{7–15}$$

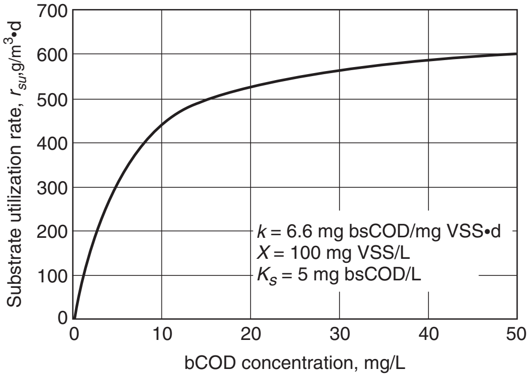
where Y = synthesis yield coefficient, g biomass/g substrate used

When the substrate is being used at its maximum rate, the bacteria are also growing at their maximum rate. The maximum specific growth rate of the bacteria is thus related to the maximum specific substrate utilization rate as follows.

$$\mu_m = kY \tag{7–16}$$

Figure 7–11

Rate of change of substrate utilization versus biodegradable soluble COD concentration based on the saturation-type model [see Eq. (7–12)].



Other Rate Expressions for Soluble Substrate Utilization

In reviewing kinetic expressions used to describe substrate utilization and biomass growth rates, it is important to remember that the expressions used to model biological processes are all empirical, based on coefficient values determined experimentally. Besides the substrate limited relationship presented above, other expressions that have been used to describe soluble substrate utilization rates include the following:

$$r_{su} = kS \tag{7–17}$$

$$r_{su} = kXS \tag{7–18}$$

$$r_{su} = kX\frac{S}{S_o} \tag{7–19}$$

The particular rate expression used to define kinetics of substrate utilization depends mainly on the experimental data available to fit the kinetic equations and the application of the kinetic model. In many cases, the pseudo first-order model as given by Eq. (7–18) is satisfactory for describing substrate utilization rates when the biological treatment process is operated at relatively low substrate concentrations. Fundamental in the use of any rate expression is its application in a mass-balance analysis as discussed in the following section. Also, with regard to modeling biological treatment processes, kinetic models should not be applied outside of the range of the conditions used to develop model coefficients.

Rate of Soluble Substrate Production from Biodegradable Particulate Organic Matter

The rate expressions for substrate utilization and biomass growth presented thus far are based on the utilization of soluble substrates. In municipal wastewater treatment only about 20 to 50 percent of the degradable organic material enters as soluble compounds, and for some industrial wastewaters the soluble organic material may be a low to moderate fraction of the total degradable organic substrates. Bacteria cannot consume the particulate substrates directly and must employ extracellular enzymes to hydrolyze the particulate organics to soluble substrates. The particulate substrate conversion rate is also a rate-limiting process that is dependent on the particulate substrate and biomass concentrations. A rate expression for particulate substrate conversion is shown as follows (Gujer et al., 1999):

$$r_{X_s} = -\frac{k_h(X_S/X_H)X_H}{(K_X + X_S/X_H)} \tag{7–20}$$

where r_{X_s} = rate of hydrolysis of particulate substrate to conversion to soluble substrate, g/m^3·d
k_h = maximum specific hydrolysis rate, g X_S/g X_H·d
X_S = particulate substrate concentration, g/m^3
X_H = heterotrophic biomass concentration, g/m^3
K_X = half-velocity hydrolysis coefficient, g/g

The particulate degradation concentration is expressed relative to the biomass concentration, because the particulate hydrolysis is related to the relative contact area between the nonsoluble organic material and the biomass. The effect of particulate organic constituents is considered further in Chap. 8.

Net Biomass Growth Rate

The net biomass growth rate is the biomass growth rate less the endogenous respiration rate. In Sec. 7–4, the biomass growth rate was shown to be proportional to the substrate utilization rate by the synthesis yield coefficient, and biomass decay was shown to be proportional to the biomass present. Thus, the following relationship between the net biomass growth rate and the substrate utilization rate is applicable in both batch and continuous culture systems.

$$r_X = Yr_{su} - bX \tag{7–21}$$

$$r_X = Y\frac{kXS}{K_s + S} - bX \tag{7–22}$$

where r_X = net biomass growth rate per unit reactor volume, g VSS/m^3·d
b = specific endogenous decay coefficient, g VSS/g VSS·d
Other terms are as defined above.

If both sides of Eq. (7–22) are divided by the biomass concentration *X*, the net specific growth rate is defined as follows:

$$\mu_{net} = \frac{r_X}{X} = Y\frac{kS}{K_s + S} - b \tag{7–23}$$

where μ_{net} = net specific biomass growth rate, g VSS/g VSS·d

As shown, the net specific growth rate corresponds to the change in biomass per day relative to the amount of biomass present and is a function of the reactor substrate concentration and the specific endogenous decay coefficient.

As noted in Sec. 7–4, the specific endogenous decay coefficient accounts for the loss in cell mass due to oxidation of internal storage products for energy for cell maintenance, cell death, and predation by organisms higher in the food chain. These factors are usually lumped together under endogenous decay, and it is assumed that the decrease in cell mass caused by them is proportional to the biomass concentration present. In Eq. (7–21), the coefficient *b* is the *specific endogenous decay* rate coefficient. An alternative approach used to describe the endogenous decay known as a *lysis-regrowth* model is described in Sec. 8–10 in Chap. 8. In biological treatment processes, both the substrate utilization and biomass growth rates are controlled by some limiting substrate, as given by Eqs. (7–12) and (7–22). The growth limiting substrate can be any of the essential requirements for cell growth (i.e., electron donor, electron acceptor, or nutrients), but often it is the electron donor that is limiting, as other requirements are usually available in excess. Thus, when the term substrate is used to describe growth kinetics, it generally refers to the electron donor.

For many substrate utilization modeling applications, it is convenient to use a general equation that includes factors other than the electron donor substrate and the substrate utilization rate. In some cases, low dissolved oxygen concentration or low nutrient concentrations can be of concern. For general model applications it is convenient to describe these effects as a multiple of rate expressions for each important factor affecting the soluble substrate utilization rate. For example, Eq. (7–24), used in the International Water Association ASM2d and ASM3 (Gujer et al., 1999) models, is used to account for possible effects of low dissolved oxygen and ammonia-nitrogen concentrations limiting the bacteria growth rate and thus the substrate utilization rate. At very low ammonia-nitrogen concentrations (less than 0.05 mg/L), the availability of nitrogen for cell synthesis may limit the biomass growth rate.

$$r_{su} = \left[\frac{\mu_{H,\max} S_S}{Y_H(K_s + S_S)}\right]\left(\frac{S_o}{K_o + S_o}\right)\left(\frac{S_{NH}}{K_{NH} + S_{NH}}\right)X_H \qquad (7\text{–}24)$$

where r_{su} = substrate utilization rate, g/m^3·d
$\mu_{H,\max}$ = maximum specific growth rate of heterotrophic bacteria, g VSS/g VSS·d
Y_H = heterotrophic bacteria synthesis yield coefficient, g VSS/g COD used
S_S = bsCOD concentration, g/m^3
S_o = dissolved oxygen concentration, g/m^3
S_{NH} = ammonia-nitrogen concentration, g/m^3
K_S = bsCOD half-velocity constant, g/m^3
K_o = dissolved oxygen half-velocity constant, g/m^3
K_{NH} = ammonia-nitrogen half-velocity constant, g/m^3
X_H = heterotrophic bacteria concentration, g VSS/m^3

Kinetic Coefficients for Substrate Utilization and Biomass Growth

The values used for the coefficients k, K_s, Y, and b to predict the rate of substrate utilization and biomass growth can vary as a function of the wastewater source, microbial population, and temperature. Kinetic coefficient values are determined from bench-scale testing or by fitting a model to full-scale plant test results. For modeling substrate removal in municipal and industrial wastewater, the coefficient values represent the net effect of microbial kinetics on the simultaneous degradation of a variety of different wastewater constituents by a variety of microorganisms. Typical values for kinetic coefficients are reported in Table 7–8 for the aerobic oxidation of BOD in domestic wastewater. Additional kinetic coefficient values are given in Chaps. 8, 9, and 10.

Rate of Oxygen Uptake

The rate of oxygen uptake is related stoichiometrically to the organic utilization rate and growth rate (see Sec. 7–4). Thus, the oxygen uptake rate can be defined as

$$r_o = r_{su} - 1.42 r_x \qquad (7\text{–}25)$$

where r_o = oxygen uptake rate, g O$_2$/m^3·d
r_{su} = substrate utilization rate per unit of reactor volume, g bsCOD/m^3·d
1.42 = the COD of biomass, g COD/g VSS
r_x = rate of biomass growth, g VSS/m^3·d

The factor 1.42 represents the COD of biomass as defined previously by Eq. (7–5).

Table 7–8

Typical kinetic coefficients for the activated sludge process for the removal of BOD from domestic wastewater

Coefficient	Unit	Value[a] Range	Value[a] Typical
k	g bsCOD/g VSS·d	4–12	6
K_s	mg/L BOD	20–60	30
	mg/L bsCOD	5–30	15
Y	mg VSS/mg BOD	0.4–0.8	0.6
	mg VSS/mg COD	0.4–0.6	0.45
b	g VSS/g VSS·d	0.06–0.15	0.10

[a] Values reported are for 20°C.

Effects of Temperature

The temperature dependence of the biological reaction-rate constants is very important in assessing the overall efficiency of a biological treatment process. Temperature not only influences the metabolic activities of the microbial population but also has a profound effect on such factors as gas-transfer rates and the settling characteristics of the biological solids. The effect of temperature on the reaction rate of a biological process is expressed using the same type of relationship developed previously in Chap. 1 [see Eq. (1–44)] and repeated here for ease of reference.

$$k_T = k_{20}\theta^{(T-20)} \tag{1–44}$$

where k_T = reaction-rate coefficient at temperature T, °C
k_{20} = reaction-rate coefficient at 20°C
θ = temperature-activity coefficient
T = temperature, °C

Values for θ in biological systems can vary from 1.02 to 1.25. Temperature correction factors for various kinetic coefficients are given in Chap. 8.

Total Volatile Suspended Solids and Active Biomass

The kinetic expressions used to describe biological kinetics and growth are related to the active biomass concentration X in the treatment reactor. In reality the VSS in a reactor consists of more than active biomass, and the fraction of active biomass can vary depending on the wastewater characteristics and operating conditions. The other components that contribute to the VSS concentration are cell debris, resulting from endogenous decay, and non-biodegradable VSS (nbVSS) in the influent wastewater fed to the biological reactor.

During cell death, cell lysis occurs with the release of cellular materials into the liquid for consumption by other bacteria. A portion of the cell mass (cell wall) is not dissolved and remains as nonbiodegradable particulate matter in the system. The remaining nonbiodegradable material is referred to as cell debris and represents about 10 to 15 percent of the original cell weight. Cell debris is also measured as VSS and contributes to the total VSS concentration measured in the reactor mixed liquor. The rate of production of cell debris is directly proportional to the endogenous decay rate.

$$r_{x,i} = f_d(b)X \tag{7–26}$$

where $r_{x,i}$ = rate of cell debris production, g VSS/m³·d
f_d = fraction of biomass that remains as cell debris, 0.10 – 0.15 g VSS/g biomass VSS depleted by decay

Other terms as defined previously.

The inert VSS concentration resulting from cell debris is typically a relatively small fraction of the VSS in a bioreactor used to treat municipal and some industrial wastewaters. As noted above, a variable amount of MLVSS that is not biomass originates from the nbVSS in the influent wastewater. For typical untreated municipal wastewaters the nbVSS concentration may be in the range from 60 to 100 mg/L, and following primary treatment may range from 10 to 40 mg/L.

Total Volatile Suspended Solids. The VSS production rate in the aeration tank can be defined as the sum of the biomass production rate given by Eq. (7–21), the nbVSS production given by Eq. (7–26), and the nbVSS in the influent wastewater:

$$r_{X_T,\text{VSS}} = \underbrace{Yr_{su} - bX}_{\substack{\text{net biomass}\\ \text{VSS from}\\ \text{soluble bCOD}}} + \underbrace{f_d(b)X}_{\substack{\text{nbVSS}\\ \text{from cells}}} + \underbrace{QX_{o,i}/V}_{\substack{\text{nbVSS}\\ \text{in influent}}} \tag{7–27}$$

where $r_{X_T,\text{VSS}}$ = total VSS production rate, g/m^3·d
Q = influent flowrate, m^3/d
$X_{o,i}$ = influent nbVSS concentration, g/m^3
V = volume of reactor, m^3
Other terms are as defined previously.

Active Biomass. From Eq. (7–27), the fraction of active biomass in the mixed liquor VSS (MLVSS) is the ratio of the net active biomass production rate, r_X in Eq. (7–21) divided by the total MLVSS production:

$$F_{X,\text{act}} = (Yr_{su} - bX)/r_{X_T,\text{VSS}} \tag{7–28}$$

where $F_{X,\text{act}}$ = active fraction of biomass in MLVSS, g VSS/g VSS

Net Biomass Yield and Observed Yield

The term true yield was defined in Sec. 7–4 as the amount of biomass produced during cell synthesis relative to the amount of substrate degraded. In the design and analysis of biological treatment processes, two other yield terms are important: (1) the net biomass yield and (2) the observed solids yield. The first is used as an estimate of the amount of active microorganisms in the system, and the second as the amount of sludge production.

Net Biomass Yield. The net biomass yield is the ratio of the net biomass growth rate in Eq. (7–21), and the substrate utilization rate:

$$Y_{\text{bio}} = r_x/r_{su} \tag{7–29}$$

where Y_{bio} = net biomass yield, g biomass/g substrate used

Observed Yield. The observed yield accounts for the actual solids production that would be measured for the system and is shown as follows:

$$Y_{\text{obs}} = r_{X_T,\text{VSS}}/r_{su} \tag{7–30}$$

where Y_{obs} = observed yield, g VSS produced/g substrate removed

A determination of the net biomass is considered in Example 7–5.

EXAMPLE 7–5 **Determine Biomass and Solids Yields** An aerobic complete-mix treatment process is used to treat an industrial wastewater. The amount of bsCOD in the influent wastewater is 300 g/m^3 and the influent nbVSS concentration is 50 g/m^3. The influent flowrate is 1000 m^3/d, the aerobic tank biomass concentration is 2000 g/m^3, the reactor bsCOD concentration is 2.4 g/m^3, and the reactor volume is 335 m^3. If the cell debris fraction f_d is 0.10, determine the net biomass yield, the observed solids yield, and the biomass fraction in the MLVSS. Use the kinetic coefficients given in Table 7–8.

Solution

1. Determine the net biomass yield using Eq. (7–29).

$$Y_{\text{bio}} = r_X / r_{su}$$

a. Solve for r_{su} using Eq. (7–12) and the information given in Table 7–8

$$\begin{aligned} r_{su} &= \frac{kXS}{K_s + S} \\ &= -\frac{(6/d)(2000\ \text{g/m}^3)(2.4\ \text{g bsCOD/m}^3)}{(15 + 2.4)\ \text{g/m}^3} \\ &= 1655.2\ \text{g bsCOD/m}^3\cdot\text{d} \end{aligned}$$

b. Determine the net biomass production rate r_X using Eq. (7–21)

$$\begin{aligned} r_X &= Yr_{su} - bX \\ &= (0.45\ \text{g VSS/g bsCOD})(1655.2\ \text{g bsCOD/m}^3\cdot\text{d}) \\ &\quad -(0.10\ \text{g VSS/g VSS}\cdot\text{d})(2000\ \text{g VSS/m}^3) \\ &= 544.8\ \text{g VSS/m}^3\cdot\text{d} \end{aligned}$$

c. Calculate the net biomass yield

$$\begin{aligned} Y_{\text{bio}} &= r_X/r_{su} = (544.8\ g\ \text{VSS/m}^3\cdot\text{d})/(1655.2\ g\ \text{bsCOD/m}^3\cdot\text{d}) \\ &= 0.33\ g\ \text{VSS/g bsCOD} \end{aligned}$$

2. Determine VSS production rate using Eq. (7–26).

$$\begin{aligned} r_{X_T,VSS} &= Yr_{su} - bX + f_d(b)X + QX_{o,i}/V \\ &= 544.8\ \text{g VSS/m}^3\cdot\text{d} \\ &\quad + (0.10\ \text{g VSS/g VSS})(0.10\ \text{g VSS/g VSS}\cdot\text{d})(2000\ \text{g VSS/m}^3) \\ &\quad + (1000\ \text{m}^3\text{/d})(50\ \text{g VSS/m}^3)/335\ \text{m}^3 \\ &= (544.8 + 20 + 149.3)\ \text{g VSS/m}^3\cdot\text{d} \\ &= 714\ \text{g VSS/m}^3\cdot\text{d} \end{aligned}$$

3. Calculate the observed solids yield using Eq. (7–30).

$$\begin{aligned} Y_{\text{obs}} &= r_{X_T,\text{VSS}}/r_{su} \\ &= (714\ g\ \text{VSS/m}^3\cdot\text{d})/(1655.2\ g\ \text{bsCOD/m}^3\cdot\text{d}) \\ &= 0.43\ g\ \text{VSS/g bsCOD} \end{aligned}$$

4. Calculate the active biomass fraction in the MLVSS. Using Eq. (7–28).

$$\begin{aligned} F_{X,\text{act}} &= (Yr_{su} - bX)/r_{X_T,\text{VSS}} \\ &= (544.8\ \text{g VSS/m}^3\cdot\text{d})/(714\ \text{g VSS/m}^3\cdot\text{d}) \\ &= 0.76 \end{aligned}$$

Comment Thus, accounting for the nbVSS in the wastewater influent and cell debris produced, the MLVSS contains 76 percent active biomass.

7–6 MODELING SUSPENDED GROWTH TREATMENT PROCESSES

Before discussing the individual biological processes used for the treatment of wastewater as given in Secs. 7–8 through 7–15, it will be helpful to review the general application of the kinetics of biological growth and substrate removal. The purpose here is to illustrate (1) the development of biomass and substrate balances, (2) the prediction of effluent biomass and soluble substrate concentrations, (3) the prediction of the reactor biomass and MLSS/MLVSS concentrations and amount of waste sludge produced daily, and (4) the prediction of the oxygen requirements. Attached growth processes are considered in Sec. 7–7.

Description of Suspended Growth Treatment Processes

The complete-mix reactor with recycle will be considered in the following discussion as a model for suspended growth processes. The schematic flow diagrams shown on Fig. 7–12 include the nomenclature used in the following mass-balance equations. A similar complete-mix reactor may be used in laboratory studies to assess wastewater treatability and to obtain model kinetic coefficients.

All biological treatment reactor designs are based on using mass balances across a defined volume for each specific constituent of interest (i.e., biomass, substrate, etc.). The mass balance includes the flowrates for the mass of the constituent entering and/or leaving the system and appropriate reaction rate terms for the depletion or production of the constituent within the system. The units for a mass balance are usually given in mass per volume per time. For all mass balances a check of the units is recommended to assure that the mass-balance equations are correct.

Solids Retention Time

Before proceeding with the mass balance for biomass and substrate in the completely mixed activated sludge (CMAS) process shown on Fig. 7–12, it is important to first select the system solids retention time (SRT) that will impact the solids production and is a key

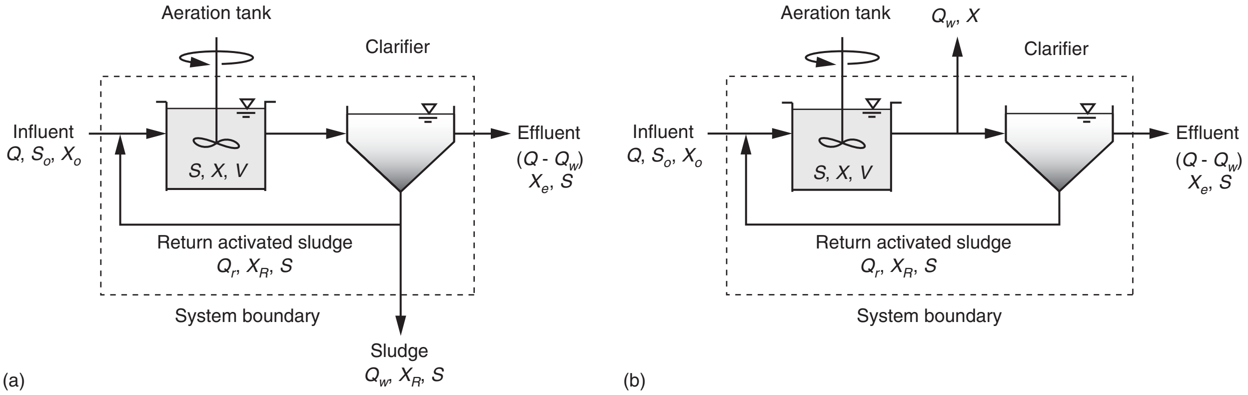


Figure 7–12

Schematic diagram of activated sludge process with model nomenclature: (a) with wasting from the sludge return line and (b) with wasting from the aeration tank.

operating and design parameter for activated sludge processes (Lawrence and McCarty, 1970). The SRT is the average time the activated sludge solids are in the system. Assuming that the solids inventory in the clarifier shown on Fig. 7–12(a) is negligible compared to that in the aeration tank, the SRT is determined by dividing the mass of solids in the aeration tank by the solids removed daily via the effluent and by wasting for process control. For many activated sludge processes, where good flocculation occurs and the clarifier is designed properly, the effluent VSS is typically less than 15 g/m^3. Where the effluent VSS is low, excess solids must be removed from the system by wasting. Wasting is accomplished most commonly by removing biomass (sludge) from the clarifier underflow recycle line as shown on Fig. 7–12(a). Alternatively, wasting can be accomplished from the aeration tank as shown on Fig. 7–12(b).

The average SRT for the process flow diagram shown on Fig. 7–12(a) is given as

$$\text{SRT} = \frac{VX}{(Q - Q_w)X_e + Q_w X_R} \tag{7–31}$$

where SRT = solids retention time, d
V = reactor volume (i.e., aeration tank), m^3
Q = influent flowrate, m^3/d
X = concentration of biomass in the aeration tank, g VSS/m^3
Q_w = waste sludge flowrate, m^3/d
X_e = concentration of biomass in the effluent, g VSS/m^3
X_R = concentration of biomass in the return activated sludge line from the clarifier, g VSS/m^3

Based on Eq. (7–31), the SRT can be controlled by the wasting rate. Increasing the value for Q_w in Eq. (7–31) results in a lower SRT. Similarly, it can be shown that by wasting from the aeration tank, the SRT can be controlled by wasting a given percentage of the aeration tank volume each day.

The inverse of the SRT is the solids wasted per day divided by the solids present.

$$\frac{1}{\text{SRT}} = \frac{(Q - Q_w)X_e + Q_w X_R}{VX} \tag{7–32}$$

At steady state operation, where the influent flowrate and substrate concentration is constant, the reactor biomass concentration is constant, and the net biomass growth rate per day is equal to the solids wasting rate, [the numerator in Eq. (7–32)]. If the product of r_x, the net biomass growth rate per unit volume (g/m^3·d) [see Eq (7–21)] and the volume, V, is substituted for the numerator in Eq. (7–32), it can be shown that the inverse of the SRT is the net specific biomass growth rate.

$$\frac{1}{\text{SRT}} = \frac{Vr_X}{VX} = \frac{r_X}{X} = \mu_{\text{net}} \tag{7–33}$$

Thus, based on Eq. (7–33), controlling the SRT by sludge wasting affects the net specific biomass growth rate, and the reactor substrate concentration. For a CMAS system the reactor effluent dissolved substrate concentration is equal to the reactor concentration.

Biomass Mass Balance

A mass balance for the mass of microorganisms in the complete-mix reactor shown on Fig. 7–12(a) can be written as follows:

1. General word statement:

$$\begin{array}{c} \text{Rate of accumulation} \\ \text{of microorganism} \\ \text{within the system} \\ \text{boundary} \end{array} = \begin{array}{c} \text{rate of flow of} \\ \text{microorganism} \\ \text{into the system} \\ \text{boundary} \end{array} - \begin{array}{c} \text{rate of flow of} \\ \text{microorganism} \\ \text{out of the system} \\ \text{boundary} \end{array} + \begin{array}{c} \text{net growth of} \\ \text{microorganism} \\ \text{within the} \\ \text{boundary} \end{array} \qquad (7\text{–}34)$$

2. Simplified word statement:

Accumulation = inflow – outflow + net growth (7–35)

3. Symbolic representation:

$$\frac{dX}{dt}V = QX_o - \ [(Q - Q_w)X_e] - (Q_w X_R) + r_x V \qquad (7\text{–}36)$$

where dX/dt = rate of change of biomass concentration in reactor measured as g VSS/m^3·d
V = reactor volume (i.e., aeration tank), m^3
Q = influent flowrate, m^3/d
X_o = concentration of biomass in influent, g VSS/m^3
Q_w = waste sludge flowrate, m^3/d
X_e = concentration of biomass in effluent, g VSS/m^3
X_R = concentration of biomass in return line from clarifier, g VSS/m^3
r_x = net rate of biomass production, g VSS/m^3·d

If it is assumed that the concentration of microorganisms in the influent can be neglected and that steady-state conditions exist (dX/dt =0), Eq. (7–36) can be simplified to

$$(Q - Q_W)X_e + Q_W X_R = r_X V \qquad (7\text{–}37)$$

If Eq. (7–37) is combined with Eq. (7–21), the result is

$$\frac{(Q - Q_w)X_e + Q_w X_R}{VX} = Y\frac{r_{su}}{X} - b \qquad (7\text{–}38)$$

where X = concentration of the biomass in the reactor, g/m^3

The inverse of the term on the left-hand side of Eq. (7–38) is defined as the average solids retention time (SRT) as given above in Eq. (7–31). Thus,

$$\frac{1}{\text{SRT}} = Y\frac{r_{su}}{X} - b \qquad (7\text{–}39)$$

The term, r_{su}, defined as the substrate utilization rate per unit of reactor volume [see Eq. (7–12)] is the amount of substrate removed in the reactor divided by the reactor volume:

$$r_{su} = \frac{Q(S_o - S)}{V} \qquad (7\text{–}40)$$

Combining Eq. (7–39) and (7–40):

$$\frac{1}{\text{SRT}} = \frac{YQ(S_o - S)}{XV} - b \qquad (7\text{–}41)$$

If Eq. (7–41) is rearranged as given by Eq. (7–42), the reactor biomass concentration is a function of the system SRT, the aerobic aeration tank hydraulic retention time, τ (V/Q), the synthesis yield coefficient, the amount of substrate removed ($S_o - S$), and the specific endogenous decay coefficient.

$$X = \left(\frac{\text{SRT}}{\tau}\right)\left[\frac{Y(S_o - S)}{1 + b(\text{SRT})}\right] \tag{7–42}$$

A substrate mass balance is needed to determine the effluent substrate concentration as a function of the reactor influent feed and operating conditions.

Substrate Mass Balance

The mass balance for substrate utilization in the aeration tank [see Fig. 7–12(a)] is

Accumulation = inflow − outflow + generation − utilization

$$\frac{dS}{dt}V = QS_o - QS + r_{su}V \tag{7–43}$$

where S_o = influent soluble substrate concentration, g/m^3

Substituting the value for r_{su} [Eq. (7–12)] and assuming steady-state conditions ($dS/dt = 0$), Eq. (7–43) can be rewritten as

$$S_o - S = \left(\frac{V}{Q}\right)\left(\frac{kXS}{K_s + S}\right) \tag{7–44}$$

If Eq. (7–41) is substituted for X in Eq. (7–44), the resulting expression is

$$S_o - S = \left(\frac{V}{Q}\right)\left(\frac{kS}{K_s + S}\right)\left(\frac{\text{SRT}}{V}\right)\left[\frac{QY}{1 + b(\text{SRT})}\right]S_o - S \tag{7–45}$$

Eliminating terms and solving for S yields

$$S = \frac{K_s[1 + b(\text{SRT})]}{\text{SRT}(Yk - b) - 1} \tag{7–46}$$

It should be noted that in Eq. (7–46), the effluent soluble substrate concentration for a complete-mix activated sludge process is only a function of the SRT and the kinetic coefficients for growth and decay. The effluent substrate concentration is not related to the influent soluble substrate concentration, but as shown previously in Eq. (7–42), the influent substrate concentration affects the reactor biomass concentration.

The same equations can be applied to describe an activated sludge process with no clarifier and thus no return sludge flow. For the case with no return sludge, all of the solids produced are present in the effluent from the aeration tank, and the SRT equals the τ.

$$\text{SRT} = VX/QX = \tau \tag{7–47}$$

The importance of the system SRT in determining the effluent soluble substrate concentration and aeration tank biomass concentration is clear from an examination of Eqs. (7–46) and (7–42).

Mixed Liquor Solids Concentration and Solids Production

The solids production from a biological reactor represents the mass of material that must be removed each day to maintain the process. It is of interest to quantify the solids production in terms of TSS, VSS, and biomass. By definition, the SRT also provides a convenient expression to calculate the total sludge produced daily from the activated sludge process:

$$P_{X_T,\text{VSS}} = \frac{X_T V}{\text{SRT}} \tag{7–48}$$

where $P_{X_T,\text{VSS}}$ = total solids wasted daily, g VSS/d
X_T = total MLVSS concentration in aeration tank, g VSS/m^3
V = volume of reactor, m^3
SRT = solids retention time, *d*

Because the 1/SRT in Eq. (7–32) represents the fraction of solids wasted per day and the mixed liquor can be assumed to be a homogeneous mixture of biomass and other solids, Eq. (7–48) can be used to calculate the amount of solids wasted for any of the mixed liquor components. For the amount of biomass wasted per day (P_X), the biomass concentration X can be used in place of X_T in Eq. (7–48).

Mixed Liquor Solids Concentration. The total MLVSS in the aeration tank equals the biomass concentration X plus the nbVSS concentration X_i:

$$X_T = X + X_i \tag{7–49}$$

A mass balance is needed to determine the nbVSS concentration in addition to the active biomass VSS concentration. The MLVSS nbVSS concentration is affected by the amount of nbVSS in the influent wastewater, the amount of nbVSS wasted per day, and the amount of cell debris produced from cell decay. A materials balance on the inert material is as follows:

Accumulation = inflow – outflow + generation

$$(dX_i/dt)V = QX_{o,i} - X_i V/\text{SRT} + r_{X,i}V \tag{7–50}$$

where $X_{o,i}$ = nbVSS concentration in influent, g/m^3
X_i = nbVSS concentration in aeration tank, g/m^3
$r_{X,i}$ = rate of nbVSS production from cell debris, g/m^3·d

At steady-state ($dX_i/dt = 0$) and substituting Eq. (7–26) for $r_{X,i}$ in Eq. (7–50) yields

$$0 = QX_{o,i} - X_i V/\text{SRT} + (f_d)(b)XV \tag{7–51}$$

$$X_i = X_{o,i}(\text{SRT})/\tau + (f_d)(b)X(\text{SRT}) \tag{7–52}$$

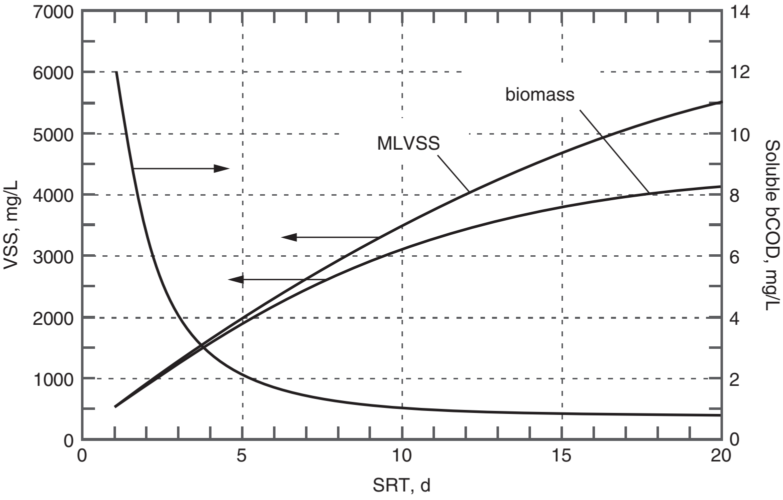
Substituting Eq. (7–42) for X and Eq. (7–52) for X_i in Eq. (7–49) produces the following equation that can be used to determine the total MLVSS concentration:

$$X_T = \underbrace{\left(\frac{\text{SRT}}{\tau}\right)\left[\frac{Y(S_o - S)}{1 + b(\text{SRT})}\right]}_{\substack{(A)\\ \text{Heterotrophic}\\ \text{biomass}}} + \underbrace{(f_d)(b)(X)(\text{SRT})}_{\substack{(B)\\ \text{Cell debris}}} + \underbrace{\frac{(X_{o,i})(\text{SRT})}{\tau}}_{\substack{(C)\\ \text{Nonbiodegradable}\\ \text{VSS in influent}}} \tag{7–53}$$

Solids Production. The amount of VSS produced and wasted daily can be estimated by substituting Eq. (7–53) into Eq. (7–48), and replacing τ with V/Q. The resulting

Figure 7–13

Biodegradable soluble COD, biomass, and MLVSS concentrations versus SRT for complete-mix activated sludge process.



equation is expressed as a function of substrate degraded, influent nbVSS, and kinetic coefficients.

$$P_{X,\text{VSS}} = \underbrace{\frac{QY(S_o - S)}{1 + b(\text{SRT})}}_{\substack{(A)\\ \text{Heterotrophic}\\ \text{Biomass}}} + \underbrace{\frac{(f_d)(b)YQ(S_o - S)\text{SRT}}{1 + b(\text{SRT})}}_{\substack{(B)\\ \text{Cell debris}}} + \underbrace{QX_{o,i}}_{\substack{(C)\\ \text{Nonbiodegradable}\\ \text{VSS in influent}}} \tag{7–54}$$

The effect of SRT on the performance of an activated sludge system for soluble substrate removal is illustrated on Fig. 7–13. In addition to the soluble substrate concentration, the total VSS concentration which includes nbVSS is also shown. As the SRT increases, more biomass decays and thus more cell debris accumulates, so that the difference between MLVSS and biomass VSS concentration increases with SRT. Also illustrated on Fig. 7–13 is the fact that the soluble substrate concentration is very low (bsCOD <5 mg/L) at SRTs above 2 d. The low substrate concentration is typical of the activated sludge process when used for the treatment of municipal wastewaters and illustrates how effectively the organic compounds are degraded in the activated sludge process. As will be shown in Chap. 8, organic substrate degradation is usually not the major factor in selecting a design SRT value.

The total mass of dry solids wasted per day is based on the TSS, which includes the VSS plus inorganic solids. Inorganic solids are present in the influent wastewater (TSS – VSS) and in the biomass, which contains 10 to 15 percent inorganic solids by dry weight. The influent inorganic solids are not soluble, and are assumed captured in the mixed liquor solids and removed in the wasted solids. To calculate the solids production in terms of TSS Eq. (7–54) is modified by adding the influent inorganic solids and by calculating the biomass in terms of TSS by assuming a typical biomass VSS/TSS ratio of 0.85. The ratio of VSS/TSS may vary from 0.80 to 0.90.

$$P_{X,\text{TSS}} = \frac{A}{0.85} + \frac{B}{0.85} + C + Q(\text{TSS}_o - \text{VSS}_o) \tag{7–55}$$

where $P_{X,\text{TSS}}$ = net waste activated sludge produced each day, measured in terms of total suspended solids, kg/d

TSS_o = influent wastewater TSS concentration, g/m^3

VSS_o = influent wastewater VSS concentration, g/m^3

A, *B*, and *C* are as defined in Eq. (7–54).

The mass of MLVSS and MLSS can be obtained by using Eqs. (7–54) and (7–55), respectively, with Eq. (7–48) as follows:

$$\text{Mass of MLVSS} = (X_{\text{VSS}})(V) = (P_{\text{X,VSS}})\,\text{SRT} \tag{7–56}$$

$$\text{Mass of MLSS} = (X_{\text{TSS}})(V) = (P_{\text{X,TSS}})\,\text{SRT} \tag{7–57}$$

By selecting an appropriate MLSS concentration, the aeration volume can be determined from Eq. (7–57). Typical MLSS concentrations in the range of 2000 to 4000 mg/L may be selected, and they must be compatible with the sludge settling characteristics and clarifier design as discussed in Secs. 8–10 and 8–11 in Chap. 8.

The Observed Yield

The observed yield Y_{obs} is based on the amount of solids production measured relative to the substrate removal, and may be calculated in terms of g TSS/g bsCOD or g BOD, or relative to VSS as g VSS/g bsCOD or g BOD. The measured solids production is the sum of the solids in the system effluent flow and the solids intentionally wasted, which equals the term P_X defined in Eqs. (7–54) and (7–55). The observed yield for VSS can be calculated by dividing Eq. (7–54) by the substrate removal rate, which is $Q(S_o - S)$:

$$Y_{\text{obs}} = \underbrace{\frac{Y}{1 + b(\text{SRT})}}_{\substack{(A)\\ \text{Heterotrophic}\\ \text{biomass}}} + \underbrace{\frac{(f_d)(b)(Y)(\text{SRT})}{1 + b(\text{SRT})}}_{\substack{(B)\\ \text{Cell debris}}} + \underbrace{\frac{X_{o,i}}{S_o - S}}_{\substack{(C)\\ \text{Nonbiodegradable}\\ \text{VSS in influent}}} \tag{7–58}$$

where Y_{obs} = g VSS/g substrate removed

For wastewaters with no nbVSS in the influent the solids production consists of only active biomass and cell debris, and the observed yield for VSS is as follows:

$$Y_{\text{obs}} = \frac{Y}{1 + b(\text{SRT})} + \frac{(f_d)(b)(Y)(\text{SRT})}{1 + b(\text{SRT})} \tag{7–59}$$

The impact of nonbiodegradable influent VSS in Eq. (7–58) on the observed yield depends on the wastewater characteristics and the type of pretreatment. The effluent substrate concentration is generally very low compared to S_o, and the term $X_{o,i}/(S_o - S)$ can be approximated by $X_{o,i}/S_o$, which is the g nbVSS/g BOD in the influent. For municipal wastewater $X_{o,i}/S_o$ values range from 0.10 to 0.30 g nbVSS/g BOD with primary treatment and 0.30 to 0.50 g nbVSS/g BOD without primary treatment. Sludge production is also addressed in Chap. 8.

Oxygen Requirements

The oxygen required for the biodegradation of carbonaceous material is determined from a mass balance using the bCOD concentration of the wastewater treated and the amount of biomass wasted from the system per day. If all of the bCOD were oxidized to CO_2 and H_2O, the oxygen demand would equal the bCOD concentration, but bacteria only oxidize a portion of the bCOD to provide energy and use a portion of the bCOD for cell growth. Oxygen is also consumed for endogenous respiration, and the amount will depend on the system SRT. For a given SRT, a mass balance on the system can be done where the bCOD removal equals the oxygen used plus the biomass VSS remaining (in terms of an oxygen equivalent), as given by Eq. (7–7). Thus, for a suspended growth process, the oxygen used is

$$\text{Oxygen used} = \text{bCOD removed} - \text{COD of waste sludge} \tag{7–60}$$

$$R_o = Q(S_o - S) - 1.42P_{X,\text{bio}} \tag{7–61}$$

where R_o = oxygen required, kg/d

$P_{X,\text{bio}}$ = biomass as VSS wasted per day, kg/d

It is important to note that $P_{X,\text{bio}}$ includes active biomass and cell debris derived from cell growth and is thus the sum of terms A and B in Eq. (7–54).

EXAMPLE 7–6 **Design of a Complete-mix Suspended Growth Process** A complete-mix suspended growth activated sludge process with recycle is used to treat municipal wastewater after primary sedimentation. The characteristics of the primary effluent are: flow = 1000 m^3/d, bsCOD = 192 g/m^3, nbVSS = 30 g/m^3, and inert inorganics = 10 g/m^3. The aeration tank MLVSS = 2500 g/m^3. Using these data and the kinetic coefficients given below, design a system with a 6-d SRT and determine the following:

1. What is the effluent bsCOD concentration?
2. What value of τ should be used so that the MLVSS concentration is 2500 g/m^3?
3. What is the daily sludge production in kg/d as VSS and TSS?
4. What is the fraction of biomass in the MLVSS?
5. What is the observed solids yield in g VSS/g bsCOD and g TSS/g bsCOD?
6. What is the oxygen requirement in kg/d?

Kinetic coefficients:

k = 12.5 g COD/g VSS·d　　K_s = 10 g COD/m^3

Y = 0.40 g VSS/g COD used　　f_d = 0.15 g VSS/g VSS

b = 0.10 g VSS/g VSS·d　　Biomass VSS/TSS = 0.85

Solution

1. Determine the effluent bsCOD concentration using Eq. (7–46).

$$S = \frac{K_s[1 + b(\text{SRT})]}{\text{SRT}(Yk - b) - 1}$$

$$= \frac{(10\text{ g bsCOD/m}^3)[1 + (0.10\text{ g VSS/g VSS·d})(6\text{ d})]}{(6\text{ d})[(0.40\text{ g VSS/g COD})(12.5\text{ g COD/g VSS·d}) - (0.10\text{ g VSS/g VSS·d})] - 1}$$

$$= 0.56\text{ g bsCOD/m}^3$$

2. Determine τ for 2500 g/m^3 MLVSS concentration.

Solve for τ in Eq. (7–53).

$$X_T = Y(S_o - S)\text{SRT}/[1 + b(\text{SRT})](\tau) + (f_d)(b)(X)\text{SRT} + (X_{o,i})\text{SRT}/\tau$$

$$\begin{aligned}2500\text{ g VSS/m}^3 = {} & (0.40\text{ g VSS/g COD})[(192 - 0.56)\text{ g COD/m}^3](6\text{ d})/\\ & [(1 + 0.10\text{ g VSS/g VSS·d}(6\text{ d})(\tau)]\\ & +(0.15\text{ g VSS/g VSS})(0.10\text{ g VSS/g VSS·d})(X)(6\text{ d})\\ & +30\text{ g bsCOD/m}^3(6\text{ d}/\tau)\end{aligned}$$

$$2500 = 287.2/\tau + 0.09(X) + 180/\tau$$

The biomass concentration X is determined using Eq. (7–42).

$$\begin{aligned} X &= [Y(S_o - S)]\text{SRT}/[1 + b(\text{SRT})](\tau) \\ &= \frac{(0.40\text{ g VSS/gCOD})[(192 - 0.56)\text{ g COD/m}^3](6\text{ d})}{[1 + (0.10\text{ g VSS/g COD})(6\text{ d})](\tau)} \\ &= (287.2\text{ g/m}^3\cdot\text{d})/\tau \end{aligned}$$

Substituting for X in the above expression yields:

$$2500 = 287.2/\tau + 180/\tau + 25.8/\tau = 493/\tau$$

and solving the above expression for τ yields

$$\tau = 0.197\text{ d}$$

Aeration tank volume $= \tau(Q) = 0.197\text{ d }(1000\text{ m}^3/\text{d}) = 197\text{ m}^3$

3. Determine the total sludge production as kg VSS/d using Eq. (7–48).

$$\begin{aligned} P_{X_T,\text{VSS}} &= X_T(V)/(\text{SRT}) \\ &= (2500\text{ g VSS/m}^3)(197\text{ m}^3)(1\text{ kg}/10^3\text{g})/6\text{d} = 82.1\text{ kg VSS/d} \end{aligned}$$

4. Determine the total sludge production as kg TSS/d using Eq. (7–55) and the assumed coefficients.

$$P_{X_T,\text{TSS}} = \frac{QY(S_o - S)}{1 + (b)\text{SRT}}\left(\frac{1}{0.85}\right) + \frac{(f_d)(b)YQ(S_o - S)\text{SRT}}{1 + (b)\text{SRT}}\left(\frac{1}{0.85}\right) + QX_{o,i} + Q(\text{TSS}_\text{o} - \text{VSS}_\text{o})$$

$$= \frac{(1000\text{ m}^3/\text{d})(0.40\text{ g VSS/g COD})[(192 - 0.56)\text{ g COD/m}^3]}{[1 + (0.10\text{ g VSS/g VSS}\cdot\text{d})(6\text{ d})](0.85)}$$

$$+ \frac{(0.15)(0.10)(1000\text{ m}^3/\text{d})(0.40)[(192 - 0.56)\text{ g COD/m}^3](6\text{ d})}{[1 + (0.10\text{ g VSS/g VSS}\cdot\text{d})(6\text{ d})](0.85)}$$

$$+ (1000\text{ m}^3/\text{d})(30\text{ g/m}^3) + (1000\text{ m}^3/\text{d})(10\text{ g/m}^3)$$

$$= (56.3 + 5.1 + 30 + 10)(10^3\text{ g/d}) = 101.4 \times 10^3\text{ g/d} = 101.4\text{ kg/d}$$

5. Determine the biomass fraction from the values for X and X_T.

$$X = (287.2\text{ g/m}^3\cdot\text{d})/\tau = (287.2\text{ g/m}^3\cdot\text{d})/0.197\text{ d} = 1458\text{ g VSS/m}^3$$

Biomass fraction $= X/X_T = 1458/2500 = 0.58$

6. Calculate the observed solids yield, g VSS/g bsCOD removed and g TSS/g bsCOD removed.

Solids wasted/d $= P_{X_T} = 82.1$ kg VSS/d and 101.4 kg TSS/d

$$\begin{aligned} \text{bsCOD removed/d} &= Q(S_o - S) \\ &= (1000\text{ m}^3/\text{d})[(192 - 0.56)\text{ g COD/m}^3](1\text{ kg}/10^3\text{ g}) \\ &= 191{,}440\text{ g COD/d} = 191.4\text{ kg/d} \end{aligned}$$

As VSS, $Y_\text{obs} = 82.1/191.4 = 0.43$ g VSS/g bsCOD

As TSS, $Y_\text{obs} = 101.4/191.4 = 0.53$ g TSS/g bsCOD

7. Determine the oxygen required using Eq. (7–61).

$$R_o = Q(S_o - S) - 1.42\, P_{X,\text{bio}}$$

$$\begin{aligned} P_{X,\text{bio}} &= P_{X_T,\text{VSS}} - P_{\text{nbVSS}} \\ &= 82.1\,\text{kg/d} - (1000\,\text{m}^3/\text{d})(30\,\text{g VSS/m}^3)(1\,\text{kg}/10^3\,\text{g}) = 52.2\,\text{kg/d} \end{aligned}$$

$$\begin{aligned} R_o &= (1000\,\text{m}^3/\text{d})[(192 - 0.56)\text{g COD/m}^3](1\,\text{kg}/10^3\,\text{g}) - 1.42(52.2\,\text{kg VSS/d}) \\ &= 117.7\,\text{kg O}_2/\text{d} \end{aligned}$$

Comment The same approach can be used to treat wastewater with particulate biodegradable COD by assuming it is equal to bsCOD. For complete-mix suspended growth designs if the SRT is 3 d or more, essentially all of the degradable particulate COD will be converted to bsCOD.

Design and Operating Parameters

In the mass balance for the complete mix reactor presented above, the SRT was introduced as the fundamental process parameter that affects the treatment efficiency and general performance for the activated sludge process. Two other activated sludge process parameters used for the design and operation of the activated sludge process, the food to microorganism ratio and the volumetric loading rate, are introduced below.

Food to Microorganism (F/M) Ratio. The F/M ratio is defined as the rate of BOD or COD applied per unit volume of mixed liquor:

$$\text{F/M} = \frac{\text{total applied substrate rate}}{\text{total microbial biomass}} = \frac{QS_o}{VX} \tag{7–62}$$

and

$$\text{F/M} = \frac{S_o}{\tau X} \tag{7–63}$$

where F/M = food to biomass ratio, g BOD or bsCOD/g VSS·d
Q = influent wastewater flowrate, m^3/d
S_o = influent BOD or bsCOD concentration, g/m^3
V = aeration tank volume, m^3
X = mixed liquor biomass concentration in the aeration tank, g/m^3
τ = hydraulic retention time of aeration tank, V/Q, d

The F/M ratio is useful for understanding the effect of transient loads on a system. The higher the specific BOD loading rate (g BOD/g VSS·d), the faster is the substrate utilization rate and thus the reactor would have a higher substrate concentration.

F/M ratio and SRT. The F/M ratio can be related to the system SRT by noting that there is a given substrate removal efficiency at a given F/M. The process removal efficiency E as percent BOD or bsCOD removed across the activated sludge system is defined as

$$E = \frac{S_o - S}{S_o}(100) \tag{7–64}$$

Thus E/100 times F/M equals

$$\frac{E}{100}\left(\frac{\mathrm{F}}{\mathrm{M}}\right) = \frac{QS_o(S_o - S)}{VX(S_o)} = \frac{Q(S_o - S)}{VX} \tag{7–65}$$

Substituting Eq. (7–40) in Eq. (7–65) yields

$$\frac{E}{100}\left(\frac{\mathrm{F}}{\mathrm{M}}\right) = \frac{r_{su}}{X} \tag{7–66}$$

In Eq. (7–66), the term (r_{su}/X) is known as the *specific substrate utilization rate, U*. Rearranging Eq. (7–39) it can be shown that U is related to the SRT and the bacteria growth and decay coefficients.

$$U = \frac{r_{su}}{X} = \frac{\left(\dfrac{1}{\mathrm{SRT}} + b\right)}{Y} \tag{7–67}$$

where terms are as defined earlier.

Substituting Eq. (7–67) into Eq. (7–66) yields

$$\frac{1}{\mathrm{SRT}} = Y(F/M)\frac{E}{100} - b \tag{7–68}$$

From Eq. (7–68) it can be seen that systems operated with a higher F/M ratio result in a lower steady-state SRT value. For systems designed for the treatment of municipal wastewater with activated sludge SRT values in the 20- to 30-d range, the corresponding F/M values may range from 0.10 to 0.05 g BOD/g VSS·d, respectively. At SRTs in the range of 5 to 7 d, the F/M values may range from 0.3 to 0.5 g BOD/g VSS·d, respectively.

Organic Volumetric Loading Rate. The organic volumetric loading rate, defined as the amount of BOD or COD applied to the aeration tank volume per day, is

$$L_{\mathrm{org}} = \frac{(Q)(S_o)}{(V)(10^3\ \mathrm{g/1\ kg})} \tag{7–69}$$

where L_{org} = volumetric organic loading rate, kg BOD/m^3·d
Q = influent wastewater flowrate, m^3/d
S_o = influent BOD concentration, g/m^3
V = aeration tank volume, m^3

Process Performance and Stability

The effects of the kinetics considered above on the performance and stability of the system shown on Fig. 7–14 will now be examined further. It was shown previously that 1/SRT, the net microorganism specific growth rate, and U, the specific substrate utilization rate, are related directly [see Eqs. (7–67) and (7–39)]. For a specified waste, a given biological community, and a particular set of environmental conditions, the kinetic coefficients Y, k, K_s, and b are fixed. It is important to note that domestic wastewater may have significant variability in its composition and may not always be treated as a single waste type in evaluating the kinetic coefficients. For given values of the coefficients, the steady state effluent substrate concentration from the reactor is a direct function of the SRT, as given by Eq. (7–46). Setting the SRT value fixes the values of U and m and also defines the efficiency of biological waste stabilization. Equation (7–46) for substrate is plotted on Fig. 7–14(a) for a growth-specified complete-mix system with recycle. As shown, the

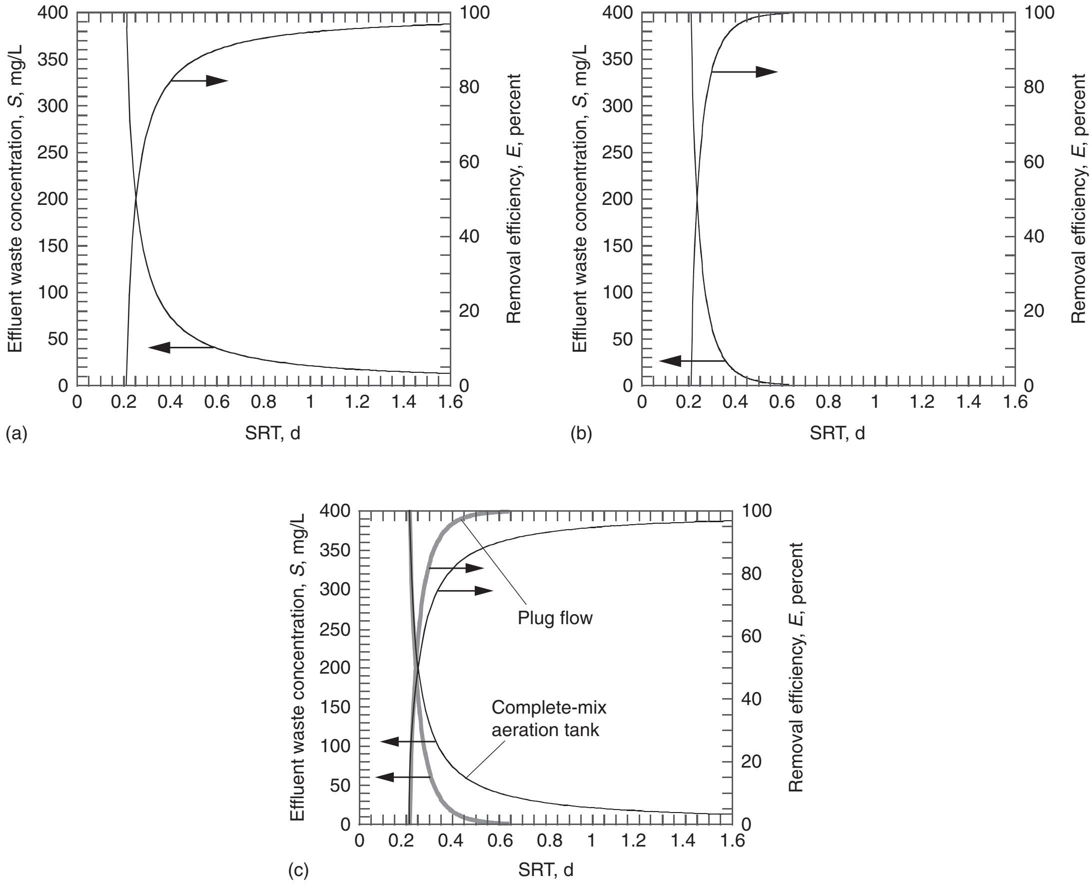


Figure 7–14

Effluent substrate concentration and removal efficiency versus SRT for: (a) complete-mix reactors with recycle, (b) plug flow reactor with recycle, and (c) plug flow and complete-mix plotted on the same graph for comparison.

treatment efficiency and the substrate concentration are related directly to the SRT, and the reactor hydraulics (i.e., complete-mix or plug-flow).

It can also be seen from Fig. 7–14(a) that there is a certain value of SRT below which waste stabilization does not occur. The critical SRT value is called the minimum solids retention residence time $\mathrm{SRT}_{\min}$. Physically, $\mathrm{SRT}_{\min}$ is the residence time at which the cells are washed out or wasted from the system faster than they can reproduce. The minimum SRT can be calculated using Eq. (7–23), in which $S = S_o$. When washout occurs, the influent concentration S_o is equal to the effluent waste concentration S.

$$\frac{1}{\mathrm{SRT}_{\min}} = \frac{YkS_o}{K_s + S_o} - b \tag{7–70}$$

In many situations encountered in waste treatment, S_o is much greater than K_s so that Eq. (7–70) can be rewritten to yield

$$\frac{1}{\mathrm{SRT}_{\min}} \approx Yk - b \tag{7–71}$$

or

$$\frac{1}{\text{SRT}_{\text{min}}} \approx \mu_m - b \tag{7–72}$$

Equations (7–71) and (7–72) can be used to determine the SRT_{min}. Typical kinetic coefficients that can be used to solve for SRT_{min} for BOD removal systems are given in Table 7–8. Biological treatment processes should not be designed with SRT values equal to SRT_{min}. To ensure adequate waste treatment, biological treatment processes are usually designed and operated with a design SRT value from 2 to 20 times SRT_{min}. In effect, the ratio of the design SRT (SRT_{des}) to SRT_{min} can be considered to be a process safety factor, SF, against system failure (Lawrence and McCarty, 1970).

$$\text{SF} = \frac{\text{SRT}_{\text{des}}}{\text{SRT}_{\text{min}}} \tag{7–73}$$

Modeling Plug-Flow Reactors

The plug-flow system with biomass recycle can be used to model certain forms of the activated sludge process. The distinguishing feature of this recycle system is that the hydraulic regime of the reactor is of a plug-flow nature. In a true plug-flow model, all the particles entering the reactor stay in the reactor an equal amount of time. Some particles may make more passes through the reactor because of recycle, but while they are in the tank, all pass through in the same amount of time.

A kinetic model of the plug-flow system is mathematically complex, but Lawrence and McCarty (1970) have made two simplifying assumptions that lead to a useful kinetic model of the plug-flow reactor:

1. The concentration of microorganisms in the influent to the reactor is approximately the same as that in the effluent from the reactor. This assumption applies only if $\text{SRT}/\tau > 5$. The resulting average concentration of microorganisms in the reactor is symbolized as $\overline{X}$.
2. The rate of change of substrate concentration as the waste passes through the reactor is given by the following expression:

$$\frac{dS}{dt} = -\frac{kS\overline{X}}{K_s + S} \tag{7–74}$$

Integrating Eq. (7–74) over the retention time of the wastewater in the aeration tank, substituting Eq. (7–42) for $\overline{X}$, and simplifying, the following expression is obtained:

$$\frac{1}{\text{SRT}} = \frac{Yk(S_o - S)}{(S_o - S) + (1 + R)K_s \ln(S_i/S)} - b \tag{7–75}$$

where S_o = influent concentration
S = effluent concentration
S_i = influent concentration to reactor after dilution with recycle flow

$$= \frac{S_o + RS}{1 + R}$$

R = clarifier return sludge recycle ratio (return flowrate divided by the influent wastewater flowrate)

Other terms are as defined previously.

The effluent substrate concentration in Eq. (7–75) is a function of both the influent concentration and SRT, whereas for a complete-mix system [see Eq. (7–46)], the effluent substrate concentration was only a function of the SRT. A version of Eq. (7–75) in which Eq. (7–42) is not substituted for $\overline{X}$ is shown in Chap. 8 in the design of sequencing batch reactors.

The true plug-flow recycle system is theoretically more efficient in the stabilization of most soluble wastes than in the continuous-flow stirred-tank recycle system. A graphical representation is shown on Fig. 7–14(b). In actual practice, a true plug-flow regime is essentially impossible to obtain because of longitudinal dispersion caused by aeration and mixing. By dividing the aeration tank into a series of reactors, the process approaches plug-flow kinetics with improved treatment efficiency compared to a complete-mix process. Because of the greater dilution with the influent wastewater, the complete-mix system can handle shock loads better than staged reactors in series. Reactor selection is discussed further in Chap. 8.

7–7 SUBSTRATE REMOVAL IN ATTACHED GROWTH TREATMENT PROCESS

In an attached growth treatment process, a biofilm consisting of microorganisms, particulate material, and extracellular polymers is attached and covers the support packing material, which may be plastic, rock, or other material (see Fig. 7–15). The growth and substrate utilization kinetics described for the suspended growth process were related to the dissolved substrate concentration in the bulk liquid. For attached growth processes, substrate is consumed within a biofilm.

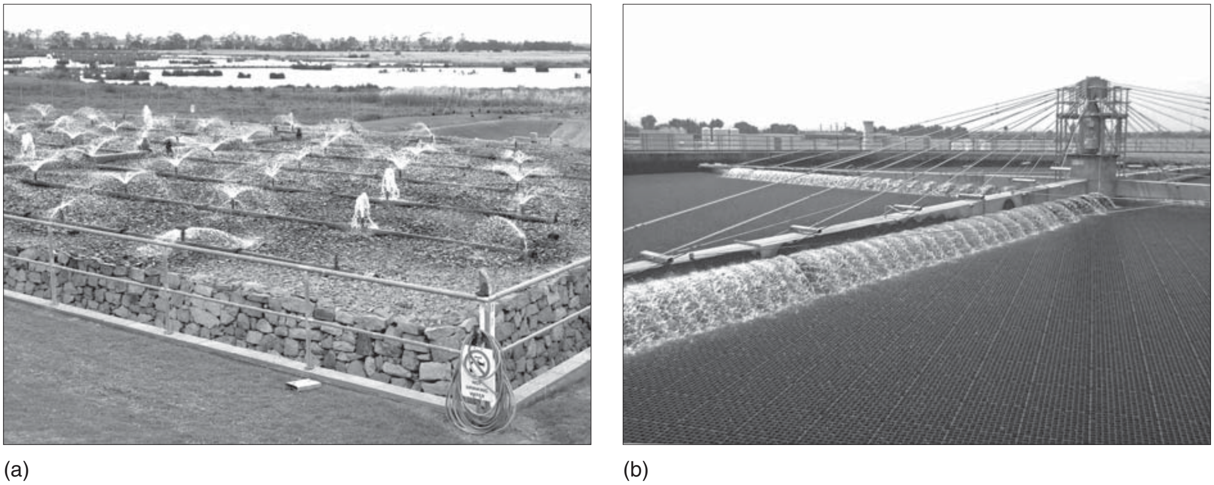
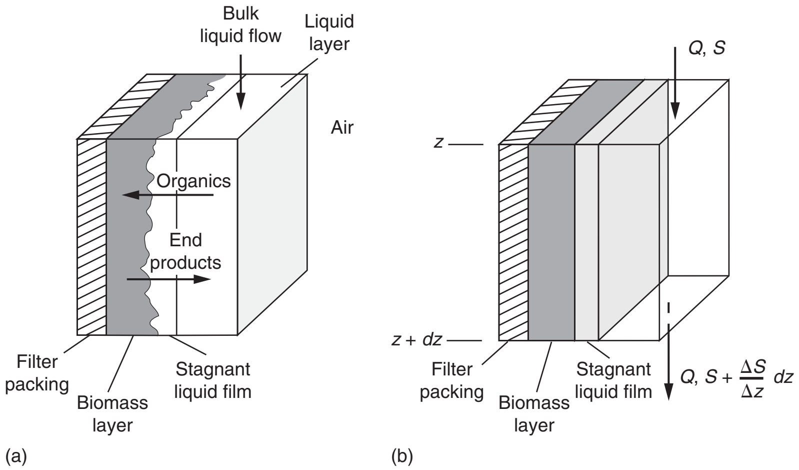

(a) (b)

Figure 7–15

Typical packing for trickling filters: (a) rock with fixed spray nozzles and (b) plastic media in tower trickling filter.

Figure 7–16

Schematic representation of the cross section of a biological slime layer in a tricling filter: (a) pictorial and (b) idealized.



Biofilm Characteristics

Depending on the growth conditions and the hydrodynamics of the system, the biofilm thickness may range from 100 μm to 10 mm (WEF, 2000). A stagnant liquid layer (diffusion layer) separates the biofilm from the bulk liquid that is flowing over the surface of the biofilm or is mixed outside of the fixed film [see Fig. 7–16(a)]. Substrates, oxygen, and nutrients diffuse across the stagnant liquid layer to the biofilm, and products of biodegradation from the biofilm enter the bulk liquid after diffusion across the stagnant film.

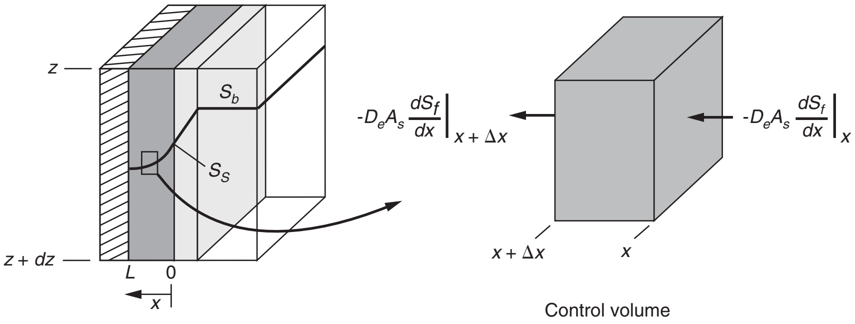
The substrate concentration at the surface of the biofilm, S_s as shown on Fig. 7–17, decreases with biofilm depth as the substrate is consumed and diffuses into the biofilm layers. As a result, the process is said to be *diffusion limited*. The substrate and oxygen concentrations within the film are lower than the bulk liquid concentration and change with biofilm depth and the substrate utilization rate. The overall substrate utilization rate is less than would be predicted based on the bulk liquid substrate concentration.

Biomass Characterization

The total amount of substrate used per unit of biofilm cross-sectional area must diffuse across the stagnant layer. This rate of mass transfer is termed the surface flux and is expressed as mass per unit area per unit time (g/m^2·d). The biofilm layer is not simply a planar surface as depicted on Fig. 7–16(b) (Costerton et al., 1995). The biofilm layers are in fact very complex nonuniform structures with uneven protrusions much like peaks and

Figure 7–17

Definition sketch for the analysis of substrate concentration in the biofilm.



valleys, and are believed to have vertical and horizontal pores through which liquid flows. The biomass can be very dense in biofilms and may also vary in density and depth. Biofilm VSS concentrations may range from 40 to 100 g/L. Uniform growth across the support packing also does not occur because of periodic sloughing, as well as the hydrodynamics and media configuration (Hinton and Stensel, 1991).

Mechanistic Models

Mechanistic models have been developed by a number of investigators to describe mass transfer and biological substrate utilization kinetics in biofilms (Williamson and McCarty, 1976; Rittman and McCarty, 1980; Kissel et al., 1984; Saez and Rittman, 1992; Suidan and Wang, 1985; Wanner and Gujer, 1986; and Rittman and McCarty, 2001) and provide useful tools for the evaluation of biofilm processes. However, because of the complexity of attached growth reactors and the inability to define accurately the physical parameters and model coefficients, empirical relationships, based on observed performance, are used for design. The empirical relationships used for design are presented and illustrated in Chap. 9. Fundamental concepts of mass transfer and substrate utilization that can be used to model the behavior of substrate removal in attached growth processes are introduced here.

Substrate Flux in Biofilms

The substrate flux across the stagnant layer to the biofilm, a function of the substrate diffusion coefficient and concentration, is given by Eq. (7–76). The negative sign is used because the substrate concentration is decreasing along the stagnant layer and substrate is removed from the bulk liquid.

$$r_{sf} = -D_w \frac{dS}{dx} = -D_w \frac{(S_b - S_s)}{L} \tag{7–76}$$

where r_{sf} = rate of substrate surface flux, g/m^2·d
D_w = diffusion coefficient of substrate in water, m^2/d
dS/dx = substrate concentration gradient, g/m^3·m
S_b = bulk liquid substrate concentration, g/m^3
S_s = substrate concentration at outer layer of biofilm, g/m^3
L = effective thickness of stagnant film, m

The thickness of the stagnant layer will vary with the fluid properties and fluid velocity. Higher velocities result in thinner films with greater substrate flux rates (Grady et al., 1999).

Mass transfer within the biofilm is described by Fick's law for diffusion (see Sec. 1–9 in Chap. 1) in an aqueous solution, with a modification to the diffusivity constant to account for the effect of the biofilm structure on the effective diffusion.

$$r_{bf} = -D_e \frac{dS_f}{dx} \tag{7–77}$$

where r_{bf} = rate of substrate flux in biofilm due to mass transfer, g/m^2·d
D_e = effective diffusivity coefficient in biofilm, m^2/d
dS_f/dx = substrate concentration gradient, g/m^3·m

The substrate utilization rate within the biofilm at any point can be defined as a saturation-type reaction (Eq. 7–12) for the substrate concentration (S_f) at that location:

$$r_{su} = \frac{kS_fX}{K_s + S_f} \tag{7–78}$$

where r_{su} = rate of substrate utilization in biofilm, g/m^2·d
S_f = substrate concentration at a point in the biofilm, g/m^3

Substrate Mass Balance for Biofilm

A substrate mass balance around a differential element (dx) shown on Fig. 7–17 within the biofilm yields

$$\begin{array}{c} \text{Rate of substrate} \\ \text{accumulation within} \\ \text{differential element} \end{array} = \begin{array}{c} \text{rate of substrate} \\ \text{flow into} \\ \text{differential element} \end{array} - \begin{array}{c} \text{rate of substrate} \\ \text{flow out of} \\ \text{differential element} \end{array} - \begin{array}{c} \text{rate of substrate} \\ \text{utilization in} \\ \text{differential element} \end{array} \tag{7–79}$$

For steady-state conditions, the mass balance is

Accumulation = inflow – outflow + generation − utilization

$$0 = -D_e A_s \frac{dS_f}{dx}\bigg|_{x} + D_e A_s \frac{dS_f}{dx}\bigg|_{x+\Delta x} - \Delta x A_s\left(\frac{kS_fX}{K_s + S_f}\right) \tag{7–80}$$

where A_s = biofilm area normal to the substrate flux, m^2

Δx = width of differential section, m

Dividing both sides by A_s and dx, and taking the limit as Δx approaches zero yields the following general equation for the change in substrate concentration within the biofilm:

$$D\frac{d^2S_f}{dx^2} - X\left(\frac{kS_f}{K_s + S_f}\right) = 0 \tag{7–81}$$

Solutions to the above equation require two boundary conditions. The first boundary condition is that the substrate flux at the biofilm surface equals the substrate flux through the stagnant liquid film, as given by Eq. (7–76). The second boundary condition is that there is no flux at the packing surface.

$$\frac{dS_f}{dx}\bigg|_{x=L} = 0 \tag{7–82}$$

Solutions for Eq. (7–81) vary, depending on (1) whether a deep biofilm exists such that the biofilm substrate concentration approaches zero toward the support surface, (2) whether a shallow film exists such that S_f is a finite value throughout the film, and (3) the relative concentration of S_f compared to K_s. Solution approaches are provided in a number of references, including Williamson and McCarty (1976), Grady et al. (1999), and Rittman and McCarty (2001).

Substrate Flux Limitations

An important implication of diffusion-limited processes is the relationship between the bulk liquid electron donor and electron acceptor concentrations. An assumption in the mechanistic models used is that either the electron donor or electron acceptor (i.e., oxygen or nitrate) is limiting. The substrate limitation may be due to reaction rates within the biofilm or to bulk liquid concentrations and diffusion rates across the stagnant layer. These are referred to by Williamson and McCarty (1976) as substrate and surface flux limitations, respectively. There are situations where the substrate limitation may switch between electron donor and electron acceptor with depth in the biofilm. For the situation where the substrate limitation can switch, numerical analysis techniques must be used to evaluate the biofilm behavior. A simple method that can be used to evaluate whether a surface flux limitation exists has been proposed by Williamson and McCarty (1976). The proposed method can also be used to assess the relative electron acceptor bulk liquid substrate concentrations needed to sustain electron donor utilization within the biofilm.

The effect of surface flux substrate limitation described by Williamson and McCarty (1976) is summarized in the following two equations:

$$\nu_d + \nu_a + \text{growth requirements} \rightarrow \text{end products} + \text{cells} \tag{7–83}$$

$$S_{ba} < \frac{D_{wd}\nu_a mw_a}{D_{wa}\nu_d mw_d} S_{bd} \tag{7–84}$$

where ν_d = molar stoichiometric reaction coefficient for electron donor, mole
ν_a = molar stoichiometric reaction coefficient for electron acceptor, mole
S_{ba} = bulk liquid electron acceptor substrate concentration, mg/L
S_{bd} = bulk liquid electron donor substrate concentration, mg/L
D_{wd} = diffusivity coefficient of electron donor in water, cm^2/d
D_{wa} = diffusivity coefficient of electron acceptor in water, cm^2/d
mw_a = molecular weight of electron acceptor, g
mw_d = molecular weight of electron donor, g

Nitrification rates in fixed-film systems are often limited by the bulk liquid DO concentration. In the following example, Eqs. (7–83) and (7–84) are used to illustrate this important issue regarding fixed-film process applications.

EXAMPLE 7–7 **Oxygen Limitation for Nitrification in a Biofilm** For bulk liquid NH_4-N concentrations of 1.0, 2.0, and 3.0 mg/L, respectively, what bulk liquid DO concentration must be present so that the nitrification rate in the biofilm is not limited due to the surface flux rate of oxygen? Assume the following conditions apply:

Electron donor = NH_4-N, $mw_d = 14$

Electron acceptor = oxygen, $mw_a = 32$

NH_4-N diffusivity coefficient at 20°C = $D_{wd} = 1.6\ \text{cm}^2/\text{d}$

Oxygen diffusivity coefficient at 20°C = $D_{wa} = 2.6\ \text{cm}^2/\text{d}$

Solution

1. Determine the stoichiometric coefficients from the reaction stoichiometry.

$$NH_4^+ + 2O_2 \rightarrow NO_3^- + 2H^+ + H_2O$$

$$\nu_d = 1.0$$

$$\nu_a = 2.0$$

2. Determine the DO concentration where oxygen is flux-limited using Eq. (7–84).

$$S_{ba} < \frac{D_{wd}\nu_a mw_a}{D_{wa}\nu_d mw_d} S_{bd} < \frac{(1.6\ \text{m}^2/\text{d})(2.0)(32\ \text{g/mole})}{(2.6\ \text{m}^2/\text{d})(1.0)(14\ \text{g/mole})} S_{bd} = 2.8\ S_{bd}$$

Thus, if S_{ba} is equal to 2.8 (S_{bd}), the nitrification rate is not hindered by the oxygen flux rate through the stagnant layer. Bulk liquid DO concentrations necessary to prevent an oxygen flux limitation for nitrification are summarized in the following table.

Bulk liquid NH_4-N concentration, g/m^3	Bulk liquid DO concentration, g/m^3
1.0	2.8
2.0	5.6
3.0	8.4

Comment For low bulk liquid NH_4-N concentrations, which result in lower nitrification rates in the biofilm, lower DO concentrations can be tolerated.

7–8 AEROBIC OXIDATION

Dating back to the early 1900s, the primary purpose of biological wastewater treatment has been to (1) remove organic constituents and compounds to prevent excessive DO depletion in receiving waters from municipal or industrial point discharges, (2) remove colloidal and suspended solids to avoid the accumulation of solids and the creation of nuisance conditions in receiving waters, and (3) reduce the concentration of pathogenic organisms released to receiving waters. The U.S. EPA secondary treatment regulatory standards, set in 1972 and still in effect, were focused mainly on the removal of BOD and TSS, and require 85 percent removal of each (see Table 1–2 in Chap. 1). Most treatment applications involve the removal of organic constituents and compounds. Because a wide range of constituents and compounds exist in wastewater, the organic content is quantified in terms of biodegradable soluble COD (bsCOD) or BOD. Additional information on the characterization of the organic constituents in wastewater is presented in Sec. 8–2 in Chap. 8.

Process Description

The removal of BOD can be accomplished in a number of aerobic suspended growth or attached (fixed film) growth treatment processes as illustrated on Figs. 7–3 and 7–4, respectively, and described in detail in Chaps. 8 and 9. Both require sufficient contact time between the wastewater and heterotrophic microorganisms and sufficient oxygen and nutrients. During the initial biological uptake of the organic material, more than half of it is oxidized, and the remainder is assimilated as new biomass, which may be further oxidized by endogenous respiration. For both suspended and attached growth processes, the excess biomass produced each day is removed and processed to maintain proper operation and performance. The biomass is separated from the treated effluent by gravity separation, and more recent designs using membrane separation are finding applications.

Microbiology

A wide variety of microorganisms are found in aerobic suspended and attached growth treatment processes used for the removal of organic material. Aerobic heterotrophic bacteria found in these processes are able to produce extracellular biopolymers that result in the formation of biological flocs (or biofilms for attached growth processes) that can be separated from the treated liquid by gravity settling with relatively low concentrations of free bacteria and suspended solids.

Protozoa also play an important role in aerobic biological treatment processes. By consuming free bacteria and colloidal particulates, protozoa aid effluent clarification. Protozoa require a longer SRT than aerobic heterotrophic bacteria, prefer dissolved oxygen concentrations above 1.0 mg/L, and are sensitive to toxic materials. Thus, their presence is a good indicator of a trouble-free stable process operation. Because of their size, protozoa can easily be observed with a light microscope at a magnification factor of 100 to 200. Rotifers can also be found in activated sludge and in biofilms, as well as nematodes and other multicellular microorganisms. These organisms occur at longer biomass retention times, and their importance has not been well defined.

Aerobic attached growth processes, depending on the biofilm thickness, generally have a much more complex microbial ecology than activated sludge. Biofilms contain bacteria, fungi, protozoan, rotifers, and possibly annelid worms, flatworms, and nematodes (WEF, 2000). The nature of biofilms and their microbial composition is discussed in more detail in Chap. 9.

Process Operation Issues

Aerobic suspended growth processes are not without operational issues. Two important operational issues are sludge bulking and foaming as discussed below.

Sludge Bulking. A principal concern in activated sludge processes with clarifiers for liquid-solids separation is to maintain a good settling sludge. However, depending on the activated sludge tank configuration, environmental factors, and operating conditions, a poor settling sludge or *bulking sludge* can develop. The term bulking is derived from the observation that the volume of a given mass of sludge increases when poor settling occurs. For this reason, the sludge volume index (SVI) is used as an indicator of sludge settling properties. The SVI is defined as the volume (mL) occupied per g of settled sludge after 30 min of settling in a 1 to 2 liter graduated container. The onset of a bulking condition and potential settling problems in gravity solids separation correlates with SVI values greater than 150 mL/g. Other forms of the SVI tests are discussed in Chap. 8. In the extreme, bulking sludge can result in high effluent suspended solids concentrations and poor treatment performance. Many bulking conditions have been associated with a wide range of filamentous type bacteria, which grow as a string of attached single cells to great lengths outside the activated sludge floc (Jenkins et al., 2004).

Foaming. Another nuisance condition in the activated sludge system is foaming, which has been related to the development of bacteria with hydrophobic cell surfaces that attach to air bubbles (see Fig. 7–18). One type results in a thick brown viscous foam layer on the activated sludge surface with bubbles that collapse inward and has been referred to as *Nocardia* foaming because of the name given to the type of bacteria observed under the microscope. However, most of the foam-causing organisms observed in activated sludge

Figure 7–18

Examples of foam caused by *Gordonia amarae* accumulated on the surface of activated sludge aeration tanks.

(a) (b)

are contained in the phylum *Actinobacteria* and include the nocardia type or *Gordonia amarae* under the Mycolata morphytes and *Candidatus Microthrix parvicella* (Seviour et al., 2008). The organisms can be found at high concentrations in the foam above the activated sludge liquid. The above types of nuisance organisms along with others are considered in Chap. 8.

Stoichiometry of Aerobic Biological Oxidation

The stoichiometry for aerobic oxidation was discussed previously but is repeated here for completeness. In aerobic oxidation, the conversion of organic matter is carried out by mixed bacterial cultures in general accordance with the stoichiometry shown below.

$$\underset{\substack{\text{Organic}\\\text{matter}}}{\text{COHNS}} + O_2 + \text{nutrients} \xrightarrow{\text{bacteria}} CO_2 + NH_3 + \underset{\text{new cells}}{C_5H_7NO_2} + \begin{matrix}\text{other}\\\text{end}\\\text{products}\end{matrix} \tag{7–85}$$

Endogenous respiration:

$$C_5H_7NO_2 + O_2 \xrightarrow{\text{bacteria}} 5CO_2 + 2H_2O + NH_3 + \text{energy} \tag{7–86}$$

In Eq. (7–85), COHNS is used to represent the organic matter in wastewater, which serves as the electron donor while the oxygen serves as the electron acceptor. Although the endogenous respiration reaction [Eq. (7–86)] is shown as resulting in relatively simple end products and energy, stable organic end products are also formed. If all of the cells (i.e., the electron donor) were oxidized completely, the UBOD or COD of the cells is equal to 1.42 times the concentration of cells as VSS [see Eq. (7–5)]. At longer SRT values, a greater portion of the cells will be oxidized.

Using the half reactions given in Table 7–6, the stoichiometry for the aerobic oxidation of acetate (the electron donor) can be represented as given below, assuming ammonia will serve as the nitrogen source for cell tissue, oxygen is the electron acceptor, and f_s for the reaction is 0.59 (see Example 7–4).

$$\begin{aligned}0.125CH_3COO^- + 0.0295NH_4^+ + 0.103O_2 \rightarrow\ & 0.0295C_5H_7O_2N + 0.0955H_2O\\ & + 0.0955HCO_3^- + 0.007CO_2\end{aligned} \tag{7–87}$$

Growth Kinetics

The form of the rate expressions for substrate utilization and biomass growth for the heterotrophic oxidation of organic substrates, based on the stoichiometry given above, were presented previously but are repeated below for ease of reference.

$$r_{su} = \frac{kXS}{K_s + S} \tag{7–12}$$

$$r_x = Yr_{su} - bX \tag{7–21}$$

$$= Y\frac{kXS}{K_s + S} - bX \tag{7–22}$$

These expressions, as noted previously, are similar to the saturation equation proposed by Monod (1942) for growth and the Michaelis-Menten equation for substrate utilization (Bailey and Ollis, 1986). Typical k and K_s values at 20°C vary from 8 to 12.0 g COD/g VSS·d and 10 to 40 g bsCOD/m^3, respectively. As noted in Sec. 7–3, the K_s value can vary depending on the nature and complexity of the bsCOD components. For easily biodegradable single substrates, K_s values of less than 1.0 mg bsCOD/L have been measured (Bielefeldt and Stensel, 1999).

Applying the above expressions for substrate utilization and biomass growth leads to the development of a series of design parameters including the solids retention time (SRT), the *food to microorganism ratio* (F/M), and the *specific utilization rate* (U). These design parameters are applied to the design of a variety of activated sludge processes in Chap. 8. With the exception of some difficult-to-degrade constituents in industrial wastewaters, the kinetics for aerobic oxidation of organic substrates seldom control the SRT design value for the activated sludge process. For good floc formation, sufficient time is needed for the biomass in the activated sludge aeration tank to develop extracellular polymers and a floc structure. More optimal flocculation and TSS removal in clarification occur typically at SRT values greater than 2.5 to 3.0 d at 20°C and 3 to 5 d at 10°C. However, some wastewater-treatment plants in warmer climates operate at SRT values varying from less than 1 to 1.5 d. High purity oxygen systems routinely operate at low SRT values. Excessively long SRTs (>20 d) may lead to floc deterioration with the development of small pinpoint floc particles that produce a more turbid effluent. However, even with pinpoint floc, effluent suspended solids concentrations of less than 30 mg/L are generally achieved. The SRT may be varied in treatment plant operations to find the most optimal settling condition.

Environmental Factors

For carbonaceous removal, pH in the range of 6.0 to 9.0 is tolerable, while optimal performance occurs near a neutral pH. A reactor DO concentration of 2.0 mg/L is used commonly, and at concentrations above 0.50 mg/L there is little effect of the DO concentration on the degradation rate. Where industrial wastewaters are discharged to municipal collection systems, care must be taken to assure that sufficient nutrients (N and P) are available for the amount of bsCOD to be treated. Heterotrophic bacteria responsible for BOD removal can tolerate higher concentrations of toxic substances as compared to the bacteria and archaea responsible for ammonia oxidation or the production of methane.

7–9 BIOLOGICAL OXIDATION OF INORGANIC NITROGEN

The need to oxidize ammonia (NH_4-N) and nitrite (NO_2-N) in wastewater treatment arises from water quality concerns over (1) the effect of ammonia on receiving water with respect to DO concentrations and fish toxicity, (2) the need to provide nitrogen removal to control eutrophication, and (3) the need to provide nitrogen control for water-reuse applications including groundwater recharge. For reference, the current (2001) drinking water maximum contaminant level (MCL) for nitrate is 45 mg/L as nitrate or 10 mg/L as nitrogen. The total concentration of organic and ammonia nitrogen in municipal wastewaters is typically in the range from 25 to 45 mg/L as nitrogen based on a flowrate of 380 L/capita·d (100 gal/capita·d). In many parts of the world with limited water supplies, total nitrogen concentrations in excess of 200 mg/L as N have been measured in domestic wastewater.

Nitrification is the term used to describe the two-step biological process in which one type of autotrophic bacteria oxidize ammonia to nitrite (NO_2-N) and a second type of autotrophic bacteria oxidize nitrite to nitrate (NO_3-N). Nitritation is the term used to describe a biological process with oxidation of NH_4-N to only NO_2-N. Nitification occurs in most aerobic suspended growth and fixed film biological process designs used to accomplish ammonia or nitrogen removal, but nitritation processes are also used in which the growth of autotrophic bacteria that oxidize nitrite to nitrate is prevented or inhibited.

Nitritation is a key factor in processes that reduce or eliminate carbon requirements for nitrogen removal. These include the SHARON Process (Single Reactor System for High Activity Ammonia Removal Over Nitrite) (Hellinga et al., 1998) and the ANAMMOX Process (Anaerobic Ammonia Oxidation) (Mulder et al., 1995), which are introduced and discussed in detail in Chap. 15 under sidestream treatment technologies. Nitritation may also occur in activated sludge "simultaneous nitrification-denitrification" (SNdN) systems or biological fixed film systems treating influent wastewater under limited dissolved oxygen conditions.

For many decades since Winogradsky's (1890) isolation of *Nitrosomonas*, the common believe was that autotrophic aerobic bacteria are responsible for ammonia and nitrite oxidation. However, Strous et al. (1999a) reported a novel bacteria capable of oxidizing ammonia with nitrite under anaerobic conditions (Anammox Process), and Konneke et al. (2005) showed that ammonia oxidation is not limited to that by bacteria and can be done by organisms of the domain Archaea. The first *ammonia-oxidizing archaea* (AOA) isolate was obtained from a marine aquarium tank and named *Candidatus Nitrosopumilus maritimus* (Konneke et al., 2005). Since then, the presence of AOA in activated sludge wastewater treatment systems has been identified by Park et al. (2006), Wells et al. (2009), Zhang et al. (2011), and Limpiyakorn et al. (2011). It has been reported that AOA were dominant over AOB in a membrane bioreactor system (MBR) at high mixed liquor concentrations (>8,000 mg/L) and low DO concentration (<0.20 mg/L), but that this dominance decreased following operation during months with colder temperatures Giraldo et al. (2011a, 2011b).

Process Description

As with BOD removal, nitrification can be accomplished in both suspended growth and attached growth biological processes. For suspended growth processes, a more common approach is to achieve nitrification along with BOD removal in the same single-sludge process, consisting of an aeration tank, clarifier, and sludge recycle system [see Fig. 7–19(a)]. In cases where there is a significant potential that the wastewater may contain toxic and inhibitory substances, a two-sludge suspended growth system may be considered [see Fig. 7–19(b)]. The two-sludge system consists of two aeration tanks and two clarifiers in series with the first aeration tank/clarifier unit operated at a short SRT for BOD removal. The BOD and toxic substances are removed in the first unit, so that nitrification can proceed unhindered in the second. A portion of influent wastewater usually has to be bypassed to the second sludge system to provide a sufficient amount of solids for efficient solids flocculation and clarification. A two-sludge suspended growth system has also been used in biological nitrogen removal systems (Boehnke et al., 1997, WERF, 2010) as an approach to sustainable environmental engineering by reducing aeration energy needs and increasing sludge wasting to anaerobic digesters to enhance biomethane production [see Fig. 7–19(c)]. The first sludge system is operated with a short SRT and high BOD loading rate. The second sludge system is operated at a longer SRT to promote ammonia oxidation. The bacteria responsible for nitrification grow much more slowly than heterotrophic bacteria, so that systems designed for nitrification generally have much longer hydraulic and solids retention times than those for systems designed only for BOD removal.

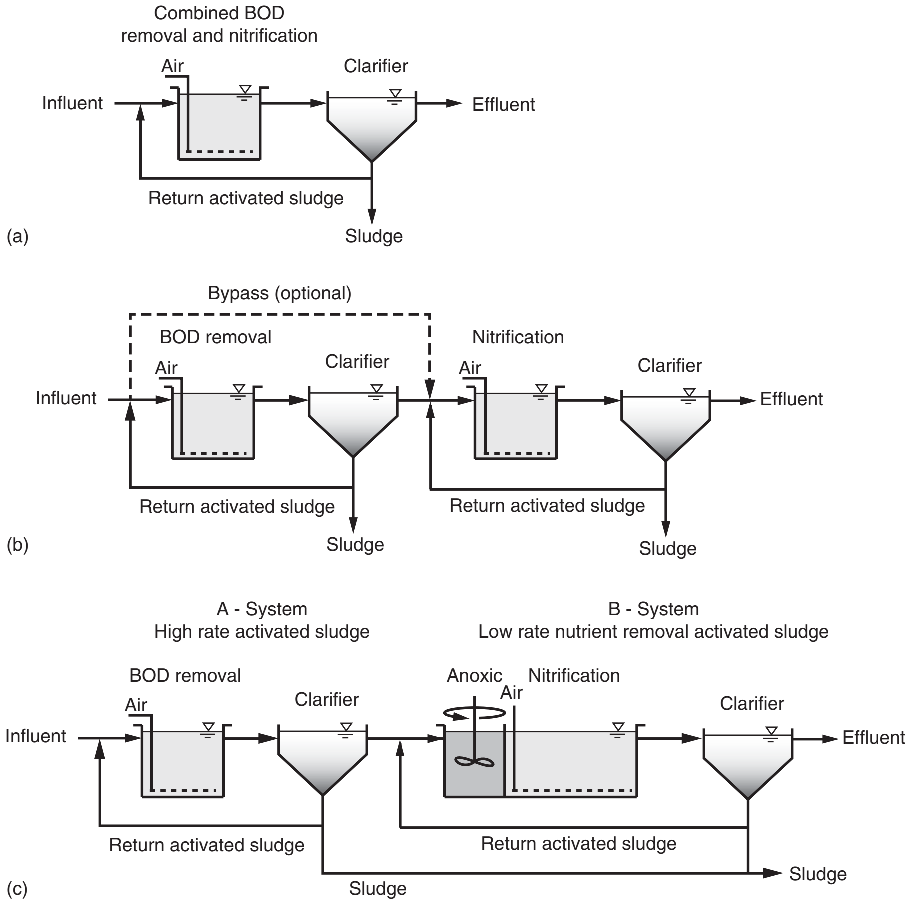
In attached growth systems used for nitrification, most of the BOD must be removed before nitrifying organisms can be established. The heterotrophic bacteria have a higher biomass yield and thus can dominate the surface area of fixed-film systems over nitrifying bacteria. Nitrification is accomplished in an attached growth reactor after BOD removal or in a separate attached growth system designed specifically for nitrification. The design of attached growth biological systems is described in Chap. 9.

Microbiology

Ammonia-oxidizing bacteria (AOB) and *nitrite-oxidizing bacteria* (NOB) are aerobic chemoautotrophs because they use CO_2 for their carbon source and require dissolved oxygen to oxidize an inorganic compound (NH_4-N or NO_2-N) to obtain cell energy. The

Figure 7–19

Process configuration used for biological nitrification: (a) single-sludge suspended growth system, (b) two-sludge suspended growth system, and (c) AB Process (A-system with high rate activated sludge, B-system with low rate nutrient removal activated sludge) (Boehnke et al., 1997).



phylogenetic distribution of AOB and NOB is based on differences in 16S ribosomal RNA (rRNA) sequences (Rowan et al., 2003) and are classified under the *α-Proteobacteria* and *β-Proteobacteria*. The AOB under the *α-Proteobacteria* are found in marine environments and thus not important for wastewater treatment applications. The AOB in the *β-Proteobacteria* are divided into two genera: the *Nitrosomonas* and the *Nitrosospira* (Purkhold et al., 2000 and Koops and Pommerening-Roser, 2001).

Distribution of AOB and NOB. The phylogenetic distribution of AOB within these genera is summarized in Table 7–9. There are five sub clusters under *Nitrosomonas* and two of these, *N. marina* and *N. cryotolerans*, are obligate halophilic bacteria and thus of less importance in wastewater treatment. The remaining three are *europaea/eutropha*, *N. communis*, *N. oligotropha*.

The NOB phylogeny, summarized in Table 7–10, has more diversity with four genera in three Proteobacteria groups. *Nitrobacter* are within *α-Proteobacteria, Nitrococcus* within *γ-Proteobacteria*, and *Nitrospina* and *Nitrospira* within the *δ-Proteobacteria*. Of these, the *Nitrospira marina,* the *Nitrospina gracilis,* and *Nitrococcus mobilis* are obligate halophilic bacteria and are less important to wastewater treatment. While *Nitrobacter* generally occur as free cells, *Nitrospira* are more commonly observed attached to flocs or biofilms in their natural environments.

Table 7–9

Phylogeny of Ammonia-Oxidizing Bacteria [Adapted from Koops and Pommerening-Roser (2001) and Ward et al. (2011)]

Subclass of Proteobacteria	Genus	Sub Clusters	Species
β-Proteobacteria	*Nitrosomonas*	*europaea-mobilis*	*Nitrosomonas europea*
			Nitrosomonas eutropha
			Nitrosomonas halophilar
			Nitrosococcus mobilis
		communis	*Nitrosomonas communis*
			Nitrosomonas sp. I
			Nitrosomonas sp. II
			Nitrosomonas nitrosa
		oligotrophia	*Nitrosomonas ureae*
			Nitrosomonas oligotropha
		marina	*Nitrosomonas marina*
			Nitrosomonas sp.III
			Nitrosomonas aestuarii
			Nitrosomonas cryotolerans
	Nitrosospira	*nitrosospira*	*Nitrosolobus multiformis*
			Nitrosovibrio tenuis
			Nitrosospira sp. I

Molecular Tools. Molecular tools have been helpful for the evaluation of nitrifying cultures in suspended growth and fixed film nitrification systems. Using oligonucleotide probes for ammonia-oxidizing bacteria, Wagner et al. (1995) showed that *Nitrosomonas* was common in activated sludge systems. A review by Geets et al. (2006) of molecular biology studies of nitrifying populations in wastewater treatment facilities found that the genus *Nitrosomonas* often dominates the AOB population with a few studies showing domination by the Nitrosospira-related strains. A review of population studies of NOB in

Table 7–10

Phylogeny of Nitrite-Oxidizing Bacteria [Adapted from Koops and Pommerening-Roser (2001)]

Subclass of Proteobacteria	Genus	Species
α-Proteobacteria	Nitrobacter	*Nitrobacter alkalicus*
		Nitrobacter winogradskyi
		Nitrobacter vulgaris
		Nitrobacter hamburgenis
γ-Proteobacteria	Nitrococcus	*Nitrococcus mobilis*
	Nitrospina	*Nitrospina gracilis*
δ-Proteobacteria	Nitrospira	*Nitrospira moscoviensis*
		Nitrospira marina

wastewater facilities (Kim et al., 2006) found that, contrary to the commonly accepted idea that Nitrobacter were the major nitrite-oxidizers, the genus Nitrospira was the more common dominant population. Teske et al. (1994) also found that *Nitrococcus* was quite prevalent in fixed film nitrification systems.

Diverse populations are also found for activated sludge and fixed film systems. Siripong and Rittman (2007) found that coexistence of AOB *Nitrosomonas* and *Nitrosospira* genera and NOB *Nitrobacter* and *Nitrospira* genera was found in studies of seven activated sludge facilities in the Metropolitan Water Reclamation District of Greater Chicago (MWRDGC). Diversity of AOB in an aerated-anoxic process was also noted by Park et al., (2002) in which they observed both *Nitrosomonas* and *Nitrosospira*. Varying growth conditions in a nitrification system may encourage the development of diverse populations as reported for NOB by Daims et al. (2001) for a sequencing batch reactor (SBR) operation. Higher nitrite concentrations during the initial aerobic cycle were thought to favor *Nitrobacter* growth but later at lower nitrite concentrations *Nitrospira* growth was favored. Coskuner and Curtis (2002) also reported on the co-existence of *Nitrospira* and *Nitrobacter* in a full scale activated sludge plant. Dytczak et al. (2008) suggested that different growth conditions in an aerobic-only SBR versus an anoxic-aerobic SBR selected for different AOB populations with different nitrification kinetics with higher rates seen for the anoxic-aerobic system.

Dominant Nitrifying Bacteria. The nitrifying bacteria that dominate a biological suspended growth or fixed film process are assumed to be the result of selective pressures, such as NH_4-N or NO_2-N concentration, DO concentration, pH, salinity, and temperature in the bioreactor. In terms of the nitrogen and DO concentration, the bacteria are divided into r-and K-strategists (Andrews and Harris, 1986). The r-strategists grow slower than the K-strategists at low substrate concentration, but grow faster than the K-strategists at higher substrate concentration. The nitrifying bacteria with high substrate affinity at low concentration (K-strategists) are *Nitrosospira* and *Nitrospira* for AOB and NOB, respectively (Schramm et al., 1999), while the AOB *Nitrosomonas Europea* and NOB *Nitrobacter* spp. grow faster at high concentration (r-strategists). Kim et al. (2006) found that low NO_2-N concentrations favored growth of *Nitrospira* and high NO_2-N concentrations favored *Nitrobacter*, which had a specific NO_2-N utilization rate five times greater than that for *Nitrospira*. Park and Noguera (2004) showed a population shift for AOB growing at high DO concentration (8.5 mg/L) versus growth at low DO concentration (0.12 to 0.24 mg/L) in laboratory reactors seeded with a wastewater facility activated sludge. However, the population change was not to a genera associated with low DO growth, but to bacteria under a *Nitrosomonas Europea* linkage.

Stoichiometry of Biological Nitrification

The energy-yielding two-step oxidation of ammonia to nitrate is as follows:
Nitroso-bacteria (Nitritation):

$$2NH_4^+ + 3O_2 \rightarrow 2NO_2^- + 4H^+ + 2H_2O \tag{7–88}$$

Nitro-bacteria:

$$2NO_2^- + O_2 \rightarrow 2NO_3^- \tag{7–89}$$

Total oxidation reaction:

$$NH_4^+ + 2O_2 \rightarrow NO_3^- + 2H^+ + H_2O \tag{7–90}$$

Based on the above total oxidation reaction, the oxygen required for the complete oxidation of ammonia is comprised of 4.57 g O_2/g NH_4-N oxidized, 3.43 g O_2/g NH_4-N for oxidation of NH_4-N to NO_2-N, and 1.14 g O_2/g NO_2-N for NO_2-N oxidation to NO_3-N. Neglecting cell tissue, the amount of alkalinity required to carry out the reaction given in Eq. (7–88) can be estimated by writing Eq. (7–90) as follows:

$$NH_4^+ + 2HCO_3^- + 2O_2 \rightarrow NO_3^- + 2CO_2 + 3H_2O \tag{7–91}$$

In the above equation, 2 moles of alkalinity are consumed per mole of NH_4-N oxidized, which is equal to 7.14 g of alkalinity as $CaCO_3$ consumed per g of NH_4-N oxidized [2 × (50 g $CaCO_3$/eq)/14]. The actual amount of oxygen and alkalinity consumed per unit of ammonia removed in nitrification systems is less than that predicted by the above stoichiometric equations because some of the ammonia removed is incorporated into the biomass produced during nitrification.

The estimated cell yields from nitrification vary widely. Biomass growth yields for AOB are 0.33 g VSS/g NH_4-N based on energetics calculations (Rittman and McCarty, 2001) and from 0.10 to 0.15 from experimental results (U.S. EPA, 1993, Haug and McCarty, 1972 and Fang et al., 2009). Biomass growth yields for NOB are 0.08 g VSS/g NH_4-N based on energetics calculations (Rittman and McCarty, 2001) and from 0.04 to 0.07 from experimental results (U.S. EPA, 1993, Haug and McCarty, 1972 and Fang et al., 2009).

Assuming a synthesis yield value of 0.12 g VSS per g NH_4-N oxidized to NO_2-N and the addition of alkalinity for NH_4-N oxidation, the following stoichiometric equation can be written for nitritation.

$$\begin{aligned} &NH_4(HCO_3) + 0.9852Na\,(HCO_3) + 0.07425CO_2 + 1.4035O_2 \rightarrow \\ &\qquad 0.01485C_5H_7NO_2 + 0.9852NaNO_2 + 2.9406H_2O + 1.9852CO_2 \end{aligned} \tag{7–92}$$

The amount of oxygen needed for the nitritation reaction shown in Eq (7–92) is 3.21 g O_2 per g NH_4-N oxidized to NO_2-N, which can be compared to a value of 3.43 [Eq. (7–88)] for nitritation without considering nitrogen used for cell synthesis.

To complete nitrification and assuming a synthesis yield coefficient of 0.04 g VSS/g NO_2-N the overall nitrification reaction can be written as follows:

$$\begin{aligned} &NH_4(HCO_3) + 0.9852Na\,(HCO_3) + 0.0991CO_2 + 1.8675O_2 \rightarrow \\ &\qquad 0.01982C_5H_7NO_2 + 0.9852NaNO_3 + 2.9232H_2O + 1.9852CO_2 \end{aligned} \tag{7–93}$$

From the above equation, for each g of NH_4-N converted to NO_3-N, 4.25 g of O_2 are utilized, 0.16 g of biomass is formed, and 7.09 g of alkalinity as $CaCO_3$ is consumed. The oxygen required to oxidize 1.0 g of NH_4-N to NO_3-N (4.27 g) is less than the theoretical value of 4.57 g computed using Eq. (7–88), which did not consider the portion of ammonia used for cell synthesis. Similarly, the alkalinity required for nitrification in Eq. (7–93) (7.09 g/g) is less than the value of 7.14 g/g. It should be recognized that the coefficient values in Eq. (7–93) are dependent upon the assumed biomass yield values. A lower net nitrifier biomass yield would have resulted in a value higher than 4.25 g O_2 per g of NH_4-N converted to NO_3-N. Werzernak and Gannon (1967) found a total oxygen consumption for nitrification of 4.33 g O_2/g N with 3.22 g O_2/g N used for NH_4-N oxidation and 1.11 g O_2/g N used for NO_2-N oxidation. These values compare to 3.21 g O_2/g N used for NH_4-N oxidation and 1.06 g O_2/g N used for NO_2-N oxidation for the above biomass synthesis assumptions and stoichiometric equations shown.

The wastewater nitrogen concentration, BOD concentration, alkalinity, temperature, and potential for toxic compounds are major issues in the design of biological nitrification processes. Nitrifying bacteria need CO_2 and phosphorus for cell growth, as well as trace elements. With such a low cell yield, the CO_2 in air is adequate, and phosphorus is seldom a concern unless metal salts are added to primary treatment or other upstream processes for phosphorus removal. Trace element concentrations that have been found to stimulate the growth of nitrifying bacteria in pure culture work are: Ca = 0.50, Cu = 0.01, Mg = 0.03, Mo = 0.001, Ni = 0.10, and Zn = 1.0 mg/L (Poduska, 1973).

Nitification Kinetics

The Monod growth kinetics model is commonly used to describe NH_4-N and NO_2-N oxidation kinetics. The specific growth rate of AOB and NOB is a function of the concentration of the respective nitrogen species being oxidized, DO concentration, and endogenous decay rate as shown in Eqs. (7–94) and (7–95). For completely mixed activated sludge systems at temperatures below 25°C, the process has normally been modeled by considering only the AOB, as the NOB are able to use nitrite much faster and therefore minimal NO_2-N is present. However, at temperatures above 28°C or at low DO concentrations (below 0.50 mg/L), the kinetics of both groups must be considered as the effect of higher temperature and low DO favors AOB over NOB such that the NH_4-N concentration may be lower than the NO_2-N concentration. The difference between AOB and NOB kinetics at high temperatures is an underlying principle of the SHARON® process for nitrogen removal, which is described in Sec. 15–9 in Chap. 15. Concentrations of NO_2-N may also be significant in batch fed systems or for aerobic reactors in series, and thus it is appropriate to model both AOB and NOB activity (Chandran and Smets, 2000 and Wett et al., 2011).

Modeling both species is also useful for predicting transient operating conditions. During the initiation of nitrification, NO_2-N concentrations will be greater than NH_4-N concentrations, as the growth of NOB cannot occur until the AOB generate nitrite. Under transient conditions, NO_2-N concentrations of 5 to 20 mg/L are possible.

The Monod models given below account for the effects of the reactor nitrogen and DO concentrations and endogenous decay on the specific growth rate of AOB or NOB.

$$\mu_{\mathrm{AOB}} = \mu_{\mathrm{max,AOB}}\left(\frac{S_{\mathrm{NH}}}{S_{\mathrm{NH}} + K_{\mathrm{NH}}}\right)\left(\frac{S_o}{S_o + K_{\mathrm{o,AOB}}}\right) - b_{\mathrm{AOB}} \tag{7–94}$$

$$\mu_{\mathrm{NOB}} = \mu_{\mathrm{max,NOB}}\left(\frac{S_{\mathrm{NO}}}{S_{\mathrm{NO}} + K_{\mathrm{NO}}}\right)\left(\frac{S_o}{S_o + K_{\mathrm{o,NOB}}}\right) - b_{\mathrm{NOB}} \tag{7–95}$$

Where μ_{AOB} = specific growth rate of ammonia-oxidizing bacteria, g VSS/g VSS·d
μ_{NOB} = specific growth rate of nitrite-oxidizing bacteria, g VSS/g VSS·d
$\mu_{\mathrm{max,AOB}}$ = maximum specific growth rate of ammonia-oxidizing bacteria, g VSS/g VSS·d
$\mu_{\mathrm{max,AOB}}$ = maximum specific growth rate of nitrite-oxidizing bacteria, g VSS/g VSS·d
b_{AOB} = specific endogenous decay rate of ammonia-oxidizing bacteria, g VSS lost/g VSS·d
b_{NOB} = specific endogenous decay rate of nitrite-oxidizing bacteria, g VSS lost/g VSS·d
S_{NH} = NH_4-N concentration, mg/L
K_{NH} = half-velocity coefficient for NH_4-N, mg/L
S_o = DO concentration, mg/L
$K_{o,\mathrm{AOB}}$ = half-velocity coefficent for DO for AOB, mg/L
S_{NO} = NO_2-N concentration, mg/L
K_{NO} = half-velocity coefficient for NO_2-N, mg/L
$K_{o,\mathrm{NOB}}$ = half-velocity coefficent for DO for NOB, mg/L

The maximum specific growth rate and specific endogenous decay rate coefficients are known to be a function of temperature and are modeled according to Eq. (1–44) in which the value for any temperature is related to the value at 20ºC and the temperature correction coefficient, θ.

$$\mu_{\max,T} = \mu_{\max,20}(\theta^{T-20}) \tag{7–96}$$

$\mu_{\max,T}$ = maximum specific growth rate coefficient at temperature T, ºC
$\mu_{\max,20}$ = maximum specific growth rate coefficient at 20ºC
θ = temperature coefficient

$$b_T = b_{20}(\theta^{T-20}) \tag{7–97}$$

Where b_T = endogenous decay coefficient at temperature T, ºC
b_{20} = endogenous decay coefficient at 20ºC

For a completely mixed activated sludge system (CMAS) operated at steady state conditions, the specific growth rate is the inverse of the SRT.

$$\mu_{\text{AOB}} = \frac{1}{\text{SRT}} \tag{7–98}$$

By combining Eqs (7–46) and (7–98), the effluent NH_4-N concentration can be determined using Eqs (7–99) and (7–100) at steady state operation for a CMAS as a function of the SRT and DO concentration and kinetic coefficients.

$$S_{\text{NH}} = \frac{K_{\text{NH}}[1 + b_{\text{AOB}}(\text{SRT})]}{\text{SRT}(\mu_{\max,\text{AOB,DO}} - b_{\text{AOB}}) - 1.0} \tag{7–99}$$

Where SRT = solids retention time, d
$\mu_{\max,\text{AOB,DO}} = \mu_{\max,\text{AOB}}$ corrected for DO concentration, g/g·d

$$\mu_{\max,\text{AOB,DO}} = \frac{(\mu_{\max,\text{AOB}})(S_o)}{(S_o + K_{o,\text{AOB}})} \tag{7–100}$$

A similar set of equations can be used to describe the NO_2-N concentrations for a CMAS reactor at steady state conditions.

The ammonia oxidation rate (r_{NH}, g/m^3·d) is a function of nitrification kinetics and reactor NH_4-N, DO and AOB concentrations as described by the general expression in Eq. (7–101). For a steady state operating condition, AOB concentration (X_{AOB}) is determined by applying Eq. (7–42) for oxidation of substrate as NH_4-N.

$$r_{\text{NH}} = \left(\frac{\mu_{\max,\text{AOB}}}{Y_{\text{AOB}}}\right)\left(\frac{S_{\text{NH}}}{S_{\text{NH}} + K_{\text{NH}}}\right)\left(\frac{S_o}{S_o + K_{o,\text{AOB}}}\right)X_{\text{AOB}} \tag{7–101}$$

$$X_{\text{AOB}} = \frac{Q(Y_{\text{AOB}})(N_{OX})\text{SRT}}{V[1 + b_{\text{AOB}}(\text{SRT})]} \tag{7–102}$$

Where: X_{AOB} = AOB concentration, g/m^3
Q = average daily influent flowrate, m^3/d
N_{OX} = NH_4-N oxidized by AOB from influent, g/m^3
V = volume of reactor containing AOB, m^3

The NH_4-N oxidation rate for AOB as a function of the biomass growth conditions [system SRT, average amount of NH_4-N oxidized daily (QN_{Ox}), and reactor volume] and reactor NH_4-N and DO concentrations is obtained by combining Eqs. (7–101) and (7–102).

$$r_{\mathrm{NH}} = \mu_{\max,\mathrm{AOB}}\left(\frac{S_{\mathrm{NH}}}{S_{\mathrm{NH}} + K_{\mathrm{NH}}}\right)\left(\frac{S_o}{S_o + K_{o,\mathrm{AOB}}}\right)\left\{\frac{Q(N_{ox})\mathrm{SRT}}{V[1 + b_{\mathrm{AOB}}(\mathrm{SRT})]}\right\} \tag{7–103}$$

Equations (7–99) and (7–103) can be used for two different types of tests to obtain nitrification kinetics. In the first case, the effluent NH_4-N concentration is measured at different operating steady-state SRT values to fit Eq. (7–99) to obtain the maximum specific growth rate, $\mu_{\max,\mathrm{AOB}}$ at a given temperature. Equation (7–103) is used with batch test data to obtain kinetic parameter values. Sufficient operating data on feed flowrate and NH_4-N oxidized at a constant SRT are needed before collecting test data to quantify biomass growth conditions. During the batch tests r_{NH} versus S_{NH} is measured with time and used in model fitting to obtain $\mu_{\max,\mathrm{AOB},}$ K_{NH}, $K_{\mathrm{o,AOB}}$, or $\mathrm{b}_{\mathrm{AOB.}}$ It should be noted that in both cases the value used for the endogenous decay rate, $\mathrm{b}_{\mathrm{AOB}}$, affects the calculated $\mu_{\max,\mathrm{AOB}}$ value.

AOB Kinetics

Values reported for $\mu_{\max,\mathrm{AOB}}$ have varied widely, ranging from 0.33 to 1.0 g/g•d at 20ºC, (Sedlak, 1991 and Randall et al., 1992). The values used for the specific decay rate coefficient ($\mathrm{b}_{\mathrm{AOB}}$) in nitrification kinetic studies have a major affect on the $\mu_{\max,\mathrm{AOB}}$ value. The selection of lower specific decay rate coefficient values results in higher $\mu_{\max,\mathrm{AOB}}$ values and vice versa. Thus, it is difficult to compare nitrification kinetic values between studies without an accurate value for $\mathrm{b}_{\mathrm{AOB}}$ (Dold et al., 2005). In early work on nitrification kinetics, the specific endogenous decay rate was believed to be very low and generally ignored, with recommended 20ºC $\mu_{\max,\mathrm{AOB}}$ values of 0.65 g/g•d (U.S. EPA, 1993) and 0.76 g/g•d (Downing et al., 1964). When considering the 20ºC endogenous decay coefficient value of 0.17 g/d•d, determined in a separate test method, a $\mu_{\max,\mathrm{AOB}}$ value of 0.90 g/g•d was recommended in a Water Environment Research Foundation (WERF) study (Melcer et al., 2003). The values of the temperature coefficient, θ, for these findings were 1.072 and 1.029 for $\mu_{\max,\mathrm{AOB}}$ and b, respectively.

Effect of Operating Conditions. The activated sludge operating conditions may affect the type of AOB selected and observed nitrification kinetics. Wett et al. (2011) claimed that two types of AOB, the "*r*" (or *u*) strategist and "*K*" strategist, may be present in the same activated sludge system with different proportions depending on the operation conditions. The *r*-strategist have higher specific growth rates and are favored by growth with higher NH_4-N concentrations. The *K*-strategist are the opposite, being more competitive at growth at lower NH_4-N concentrations and have lower half-velocity values (K_{NH}) then the *r* strategist. The effect of an increasing recycle ratio on decreasing $\mu_{\max}$ in a full scale activated sludge system with recycle from the aerobic nitrification tank back to an upstream anoxic tank was accounted for by modeling with the two types of AOB populations. Based on the modeling results, the fraction of *K*-strategist AOBs increased as the internal recycle ratio increased. Values for $\mu_{\max}$ and K_{NH} at 20°C used in the modeling for *K*- and *r*-strategists were 0.75 g/g•d and 0.30 g/m^3 and 0.95 g/g•d and 0.70 g/m^3, respectively. Based on nitrification pilot plant results from treating municipal wastewater, Munz et al. (2010) claimed that initial growth with a 20-d SRT favored *K*-strategist AOB over *r*-strategists at growth at an 8-d SRT. The 20ºC $\mu_{\max}$ values were 0.49 and 0.72 g/g•d, respectively, and the specific endogenous decay coefficient value of 0.10 g/g•d was applied in both cases.

Half-Velocity Constant. The half-velocity coefficient, K_{NH}, is an important kinetic parameter in the application of the Monod model to nitrification reactor design. Lower K_{NH} values predict lower effluent NH_4-N concentrations for a given aerobic SRT. The K_{NH}

values are affected by the system growth conditions and nitrifying bacteria selection of *r* versus *K*-strategist. Examples of widely varying literature values are 0.14 to 5.0 g/m^3 (Sin et al., 2008a) and 0.60 to 3.6 g/m^3 (U.S. EPA, 1993). Based on full scale plant performance K_{NH} values between 0.30 and 0.70 g/m^3 may be expected. In more recent models, no effect of temperature is assumed on the K_{NH} value (Henze et al., 1998 and Henze et al., 2008).

Specific Endogenous Decay. The AOB 20ºC specific endogenous decay coefficient value in aerobic activated sludge of 0.17 g/g·d for AOB reported by Melcer et al. (2003) is in general agreement with 0.17 g/g·d by Copp and Murphy (1995) and 0.15 g/g·d reported by Manser et al. (2006). Salem et al. (2006) obtained a AOB decay coefficient of 0.20 g/g·d at 23.5ºC from studies at a full-scale activated sludge facility, which is in close agreement the specific endogenous decay coefficient value of 0.17 g/d·d and temperature θ value by Melcer et al. (2003). During the anoxic phase of anoxic-aerobic activated systems, specific endogenous decay coefficients for AOB have been reported to be about 50 percent of that in the aerobic zone (Salem et al., 2006; Lee and Oleszkiewicz, 2003; Nowak et al., 1994; and Siegrist et al., 1999).

NOB Kinetics

In many applications the nitrification process has been designed adequately considering only AOB kinetics. For fully acclimated complete-mix activated sludge nitrification systems, at temperatures below 25°C with sufficient DO present, the NO_2-N concentration may be less than 0.10 mg/L as compared to NH_4-N concentrations in the range of 0.50 to 1.0 mg/L. However, there are cases where it is important to apply a two-step kinetic model for NH_4-N oxidation to NO_2-N by AOB and NO_2-N oxidation to NO_3-N by NOB. During the initial start up of nitrification, NO_2-N concentrations will be greater than NH_4-N concentrations, as the NOB growth cannot occur until the AOB generate nitrite. Under transient conditions, NO_2-N concentrations of 5 to 20 mg/L are possible at low NH_4-N concentrations. Under low DO concentration operation applications NO_2-N concentrations can be higher than NH_4-N concentrations, which can only be predicted by the two-step kinetic model. NOB kinetics are also important in batch or staged activated sludge processes, in which NO_2-N concentrations can be elevated during the initial reaction time or initial stages. At temperatures above 27ºC the NOB kinetics become less favorable compared to AOB, and use of the two-step kinetic model has led to controlling nitritation in the SHARON® process used in sidestream treatment of anaerobic digestion centrate or filtrate, which is presented in Chap.15.

Effect of Operating Conditions. Compared to AOB there have been limited studies on NOB kinetics, and in most cases the kinetics have been developed from model fitting of pilot plant or bench-scale results. Nitrite oxidation kinetic coefficients summarized by Sin et al. (2008a) show a wide variation that is related to the systems tested, model structure, and model fitting methods. The maximum specific growth rates for NOB at 20ºC are frequently reported to be higher than that for AOB, with values of 1.33 g/d·d (Wett and Rauch, 2003), 1.8 g/g·d (Sin et al., 2008a), 1.0 g/g·d (Kaelin et al., 2009) and 0.70 g/g·d (Jones et al., 2007). The μ_{max} temperature correction θ value for NOB has been given as 1.06 (Wett and Rauch, 2003, and Jones et al., 2007). As noted above the NOB kinetics can vary depending on reactor operating conditions that favor *r*- or *K*-strategist NOBs (Kim et al., 2006).

Half-Velocity Constant. There is less work on half-velocity coefficients for nitrite oxidation by NOB. In a comparative study, investigators found that the NOB substrate half-velocity coefficient was slightly higher than that for AOB (Manser et al., 2005). However in a summary of half velocity coefficients for AOB and NOB prepared by Sin et al. (2008a), lower K_{NO} values were reported for NOB, ranging from 0.05 to 0.30 g/m^3 for four sources. Much lower half-velocity coefficient values for NOB versus that for AOB were also reported by Cuidad et al. (2006).

Specific Endogenous Decay. The specific endogenous decay coefficient value found for NOB has been reported to be the same as that for AOB (Copp and Murphy, 1995; Manser et al., 2006; and Salem et al., 2006). In addition, the same temperature correction as that used for AOB has been applied to NOB (Wett and Rauch, 2003; Kaelin et al., 2009).

Environmental Factors

Nitrification and nitritation rates are affected by a number of environmental factors including dissolved oxygen concentration, pH, toxicity, metals, and unionized ammonia.

Dissolved Oxygen Concentration. Nitrification rates are affected by the liquid DO concentration in activated sludge (attached growth effects are described in Chap. 9). In contrast to what has been observed for aerobic heterotrophic bacteria degradation of organic compounds, nitrification rates increase up to DO concentrations of 3 to 4 mg/L. To account for the effects of DO, the double Monod growth rate model as shown in Eqs. (7–94) and (7–95) includes the effect of both substrate and DO concentration, with values for the DO half-velocity constant, K_o, used to describe the impact of DO on the process kinetics. Substrate oxidation rates by heterotrophic bacteria are generally not affected until the liquid DO concentration is below 0.20 mg/L.

Dissolved Oxygen Inhibition. Observed NOB NO_2 oxidation rates are inhibited more at lower DO concentrations than that for AOB NH_4-N oxidation rates, with elevated NO_2-N concentrations found when DO concentrations are depressed (Picioreanu et al., 1997; Garrido et al., 1997; Peng and Zhu 2006; Contreras, et al. 2008). Values of K_o for AOB have been reported over a wide range, from 0.10 to 1.0 g/m^3 (Cuidad et al., 2006, and Sin et al., 2008a) but respective NOB K_o values are greater. The ratio of the NOB to the AOB DO half-velocity coefficient values for studies with both AOB and NOB were 2.36 (Guisasola et al., 2005) and 1.4 (Ciudad et al., 2006). Ratios of 2.5 (Wett and Rauch, 2003), 2.0 (Jones et al., 2007) and 3.0 (Sin et al., 2008b) were used in nitrification modeling applications. Absolute values of oxygen half-velocity coefficients depend on the operating and activated sludge floc conditions and possible diffusion limitations in addition to the bacterial affinity for DO.

Absolute values of oxygen half-velocity coefficients depend on more than just the affinity for DO by the AOB or NOB. Stenstrom and Song (1991) have shown experimentally that the effect of DO on nitrification is affected by the activated sludge floc size and density, and total oxygen demand of the mixed liquor. Nitrifying bacteria are distributed within a floc containing heterotrophic bacteria and other solids, with floc diameters ranging from 100 to 400 μm. Oxygen from the bulk liquid diffuses into floc particles, and bacteria deeper within the floc are exposed to lower DO concentrations. At higher organic loading rates, there is a greater substrate concentration in the mixed liquor, which causes a higher oxygen consumption rate within the floc. Therefore, a higher bulk liquid DO concentration is needed to maintain the same internal floc DO concentration and subsequent nitrification rate.

With regard to floc size, lower K_o values have been shown for membrane bioreactor (MBR) systems with smaller floc size versus activated sludge/clarifier systems with large floc sizes (Manser et al., 2005 and Daebel et al., 2007). Blackburne et al. (2007) varied the floc size of a *Nitrobacter* culture by sonication and found K_o values from approximately 0.4 g/m^3 for small and medium size flocs, to 1.7 g/m^3 for large flocs.

Hydrogen-Ion Concentration (pH). Optimal nitrification rates occur at pH values in the 7.5 to 8.0 range. Ammonia oxidation rates decline significantly at pH values below 7.0. The decrease in the ammonia oxidation rate at lower pH may be due to the reduction in free ammonia (NH_3) concentration, as it has been reported that NH_3-N may be the true substrate for AOB (Suzuki et al., 1974). At pH values near 5.8 to 6.0, the ammonia oxidation rates may be 10 to 20 percent of the rate at pH 7.0 (U.S. EPA, 1993). Using the data from this source, the following relationship can be used to describe the relative nitrification rate at a given pH below 7.0 to that at pH 7.2.

$$\text{Relative nitrification rate} = \frac{NR_{pH}}{NR_{7.2}} = (0.0004017)e^{1.0946pH} \tag{7–104}$$

where: NR_{pH} = nitrification rate at pH
$NR_{7.2}$ = nitrification rate at pH = 7.2

Most observed pH effects on nitrification rates have been under transient conditions with cultures starting near a neutral pH. Long term acclimation at a lower pH could result in the selection of a different AOB species with much different rates at lower pH than the initial culture. Nitrification activity in a suspended growth reactor at pH values as low as 4.3 has been reported (Ward et al., 2011). The AOB community was dominated by *Nitrosomonas oligotropha* instead of *Nitrosomonas europaea* or *Nitrosomonas communis*, more common in activated sludge reactors.

To minimize possible inhibitory effects of low pH on nitrification rates, it is common to try to maintain an operational alkalinity value of 50–60 mg/L as $CaCO_3$, with a corresponding pH of 6.8 or greater. Alkalinity addition may be required for nitrification activated sludge treatment systems receiving wastewaters with low alkalinity or high ammonia concentrations. Alkalinity may be added in the form of lime, soda ash, sodium bicarbonate, or magnesium hydroxide depending on costs and chemical handling issues.

Toxicity. The AOB are the most sensitive nitrifying bacteria and their activity can be affected by a wide range of organic and inorganic compounds at concentrations well below those concentrations that would affect aerobic heterotrophic bacteria. In many cases, ammonia oxidation rates are inhibited, even though bacteria continue to grow and oxidize ammonia, but at significantly reduced rates. In some cases, toxicity may be sufficient to kill the nitrifying bacteria.

The AOB have been shown to be good indicators of the presence of organic toxic compounds at low concentrations (Blum and Speece, 1991). A comprehensive list of organic compounds that cause nitrification toxicity can be found in the Nutrient Control Design Manual (U.S. EPA, 2010). Compounds that are toxic include solvent organic chemicals, amines, proteins, tannins, phenolic compounds, alcohols, cyanates, ethers, carbamates, and benzene. Because of the numerous compounds that can inhibit nitrification, it is difficult to pinpoint the source of nitrification toxicity for wastewater plants with inhibition, and extensive sampling of the collection system is normally needed to find the source.

Metals. Metals are also of concern for nitrifiers. Skinner and Walker (1961) have shown complete inhibition of ammonia oxidation at 0.25 mg/L nickel, 0.25 mg/L chromium, and 0.10 mg/L copper.

Free Ammonia and Nitrous Acid Inhibition. Both AOB and NOB activity can be inhibited in systems that have high ammonia or nitrite concentrations, which can occur in aerobic digesters and in the treatment of high ammonia strength wastewaters, such as from anaerobic digester centrate return and animal feedlots. The compounds causing inhibition are un-ionized ammonia (NH_3-N) or free ammonia, and unionized nitrous acid (HNO_2). Their concentrations are a function of the reactor pH, temperature, and for NH_3-N, the total ammonia concentration or TAN (NH_3-N + NH_4-N), and for HNO_2, the NO_2-N concentration. The NH_3-N and HNO_2 concentrations can be estimated according the following equations (Anthonisen et al., 1976):

$$NH_3\text{-N} = \frac{\text{TAN}(10^{\text{pH}})}{(1/K_a) + 10^{\text{pH}}} \tag{7–105}$$

and

$$\frac{1}{K_a} = \exp[6334/(273 + T)] \tag{7–106}$$

where TAN = total NH_3-N + NH_4-N concentration, g/m^3
T = temperature, °C
K_a = ionization constant for ammonium

$$HNO_2\text{-N} = \frac{NO_2\text{-N}}{(K_n)(10^{\text{pH}})} \tag{7–107}$$

and

$$K_n = \exp[-2300/(273 + T)] \tag{7–108}$$

where HNO_2-N = free nitrous acid concentration as N, g/m^3
K_n = ionization constant for nitrous acid

At higher pH and temperature values, a greater fraction of the NH_3-N plus NH_4-N shifts to NH_3-N; at lower pH and temperature values a greater fraction of NO_2-N shifts to HNO_2-N.

Inhibition to AOB by NH_3-N was reported at 7.0 g/m^3 by Anthonisen et al. (1976). Complete loss of AOB activity at a 20 g/m^3 NH_3-N was found by Abeling and Seyfried (1992), but Wong-Chong and Loehr (1975) reported stable AOB activity at 50 g/m^3 NH_3-N following long term acclimation. Effects of AOB inhibition can occur in aerobic digestion of waste activated sludge due to NH_4-N accumulation under conditions of high intermittent feeding or limited DO concentrations. During solids digestion, protein from the activated sludge biomass is released and deamination produces NH_4-N and alkalinity. If the ammonia oxidation rate is not adequate relative to the NH_4-N production rate, the pH and NH_4-N increase and the NH_3-N concentration may increase to levels that inhibit AOB activity and cause further NH_4-N accumulation and inhibition. Possible TAN concentrations (NH_4 + NH_3) that may inhibit AOB at pH values from 7.0 to 8.5 and at temperatures from 20°C to 35°C are summarized in Table 7–11.

Inhibition to AOB by HNO_2, associated with high NO_2-N concentration, was reported at 0.065 to 0.83 g/m^3 by Anthonisen et al. (1976). For example at 20°C, the respective inhibitory NO_2-N concentration is estimated at 80 to 990 mg/L at pH 6.5. At a pH of 6.8, temperature of 30°C, and NO_2-N concentration of 200 g/m^3 (free nitrous acid equals 0.063 g/m^3), Silva et al. (2011) found a 37 percent reduction in AOB activity and 67 percent reduction in NOB activity showing that inhibition of NOB can also occur at high NO_2-N concentrations.

Table 7–11

Total ammonia-N (TAN) concentrations corresponding to AOB inhibition by 7.0 g/m^3 free ammonia-N at various pH and temperature conditions

	Temperature, °C			
pH	20	25	30	35
7.0	1712	1198	846	597
7.5	541	384	272	189
8.0	171	126	91	60
8.5	54	45	34	19

The NOB are more sensitive to NH_3-N inhibition than HNO_2. Inhibitory NH_3-N concentrations have been reported at 0.10 to 1.0 g/m^3 by Turk and Mavinic (1986), at 6.6 to 8.9 g/m^3 by Mauret et al. (1996), and at 3.5 g/m^3 by Wong-Chong and Loehr (1975) for unacclimated bacteria and 40 g/m^3 after an acclimation phase. Assuming NH_3-N inhibition at 8.9 g/m^3, NOB would be inhibited at a NH_4-N concentration of 488 g/m^3 at a pH of 7.5 and temperature of 25°C, which compares to 384 g/m^3 in Table 7–11 for AOB.

Ammonia Oxidizing Archaea. Limited information is available on the kinetics of ammonia oxidizing archaea (AOA), as most of the current focus has been to detect their occurrence in activated sludge treatment facilities. The only isolated ammonia-oxidizing archaea is *Nitrosopumilus maritimus* (Konneke et al., 2005). Originally classified under the Crenarchaeota phylum in the domain Archaea, *N. maritimus* has been reclassified under a new phylum named Thaumarchaeota. *N. maritimus* only grow within a narrow temperature range of 20 to 30°C and at pH between 7.0 and 7.8 (Ward et al., 2011). Unidentified AOA have been found in many activated sludge facilities used to treat domestic wastewater using specific polymerase chain reaction (PCR) primers that target the archaeal *amo*A gene (Park et al., 2006, Wells et al., 2009, Zhang et al., 2009, Limpiyakorn et al., 2011, and Giraldo et al., 2011a).

The maximum specific growth rate of *N. maritimus* at 28ºC was found to be 0.78 g/g·d by Konneke et al. (2005), but no such kinetic rates are currently found for AOA in WWTP activated sludge processes. A distinguishing characteristic with regard to kinetics of AOA is a very low half-velocity coefficient for both DO and ammonia. *N. maritimus* appears to be a *K*-strategist organism with a reported NH_4-N half-velocity coefficient (K_{NH}) of only 0.002 g/m^3 (Ward et al., 2011). Similarly with regard to DO, Giraldo et al. (2011a) found a DO half-velocity coefficient (K_o) of 0.01 g/m^3 for a mixed liquor from an MBR process operated at low DO with AOA accounting for 85 percent of the ammonia-oxidizing microorganisms based on *q*PCR measurements. Under winter temperature operation the population shifted to more AOB and the observed K_o increased to 0.25 g/m^3 (Giraldo et al., 2011b). Sonthiphand et al. (2011) also observed the advantages given to AOA at low NH_4-N concentrations. In enrichment reactors seeded with WWTP activated sludge the AOA population was abundant when the reactor NH_4-N concentration was in the range of 0.06 g/m^3, but it decreased by over 50 percent when the reactor NH_4-N concentration was 0.25 to 0.55 g/m^3.

7–10 DENITRIFICATION

The biological reduction of nitrate or nitrite to nitrogen gas is termed *denitrification*. Biological denitrification is an integral part of nitrification and denitrification processes for biological nitrogen removal, which with few exceptions is the method of choice for nitrogen removal for domestic and industrial wastewater treatment. Biological nitrogen removal is important in wastewater treatment where there are concerns for eutrophication and where groundwater must be protected against elevated NO_3-N concentrations when wastewater-treatment plant effluent is used for groundwater recharge and other reclaimed water applications. Without biological denitrification, 10 to 30 percent nitrogen removal may be found in domestic wastewater treatment due to biomass synthesis from BOD removal. The amount is a function of the influent BOD:N ratio, SRT, and the amount of nitrogen in recycle streams from solids dewatering after anaerobic or aerobic digestion.

Two modes of nitrate removal can occur in biological processes, and these are termed assimilatory or dissimilatory nitrate reduction (see Fig. 7–20). Dissimilatory denitrification is responsible for biological denitrification for enhanced nitrogen removal and involves nitrate/nitrite serving as the ultimate electron acceptor in lieu of oxygen in the bacteria cell respiratory electron transport chain for the oxidation of a variety of organic and inorganic substrates. Assimilatory nitrate reduction is independent of DO concentration and involves the reduction of nitrate to NH_4-N for use in cell synthesis when NH_4-N is not available.

In dissimilatory biological denitrification by heterotrophic bacteria, nitrate reduction proceeds through a series of intermediate products, nitrite (NO_2^-), nitric oxide (NO), and nitrous oxide (N_2O) to nitrogen gas (N_2) as shown in Eq. (7–109). NO and N_2O are nitrogen gases and the emission of N_2O is of greatest concern because it is one of the most potent greenhouse gases (see Sec. 7–12).

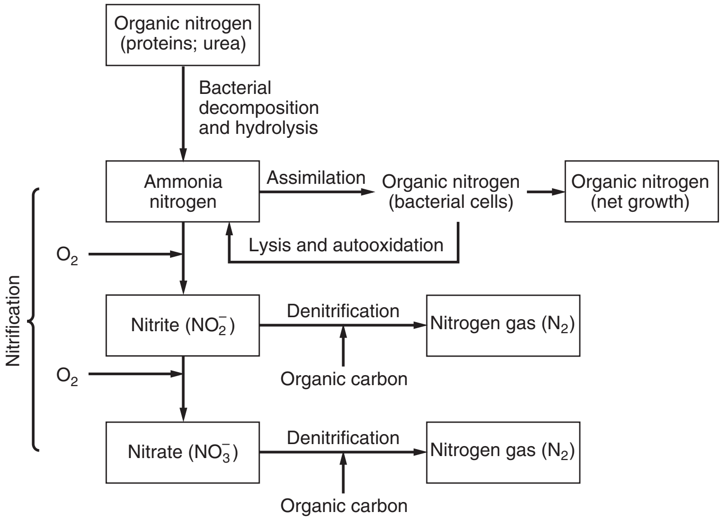
$$NO_3^- \rightarrow NO_2^- \rightarrow NO \rightarrow N_2O \rightarrow N_2 \qquad (7\text{–}109)$$

Process Description

Two basic flow diagrams for activated sludge denitrification and the conditions that drive the denitrification reaction rates are illustrated on Fig. 7–21. The first flow diagram [see Fig. 7–21(a)] is for the Modified Ludzak-Ettinger (MLE) process (U.S. EPA, 1993), the most common process used for biological nitrogen removal in municipal wastewater treatment. The process consists of an anoxic tank followed by the aeration tank where nitrification occurs. Nitrate produced in the aeration tank is recycled back to the anoxic tank.

Figure 7–20

Nitrogen transformations in biological treatment process. (Adapted from Sedlak, 1991.)



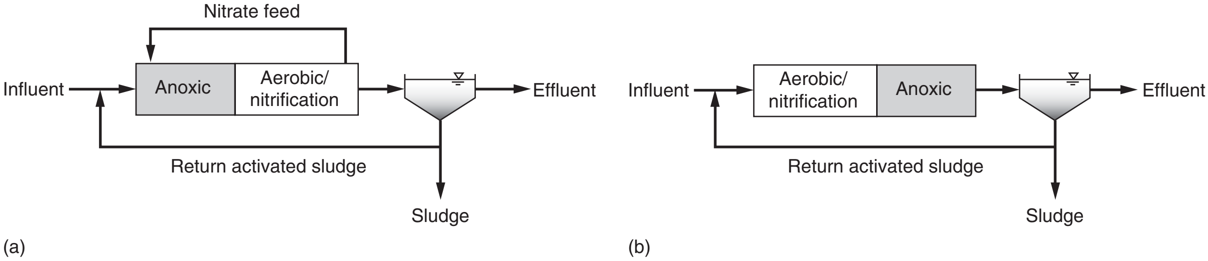


Figure 7–21

Types of denitrification processes and the reactors used for their implementation: (a) substrate driven (preanoxic denitrification) and (b) endogenous driven (postanoxic denitrification).

Because the organic substrate in the influent wastewater provides the electron donor for oxidation-reduction reactions using nitrate, the process is termed *substrate denitrification*. Further, because the anoxic process precedes the aeration tank, the process is known as *preanoxic denitrification*.

In the second process shown on Fig. 7–21(b), denitrification occurs after nitrification and the electron donor source is from endogenous decay. The process illustrated on Fig. 7–21(b) is generally termed *postanoxic denitrification* as wastewater influent BOD removal has occurred first and is not available to drive the nitrate reduction reaction. When a postanoxic denitrification process depends solely on endogenous respiration for energy, it has a much slower rate of reaction than for the preanoxic processes using wastewater BOD. Often an exogenous carbon source such as methanol or acetate is added to postanoxic processes to provide sufficient BOD for nitrate reduction and to increase the denitrification rate. Postanoxic processes include both suspended and attached growth systems. In one attached growth granular-medium filtration process, both nitrate reduction and effluent suspended solids removal occur in the same reactor.

The denitrification preanoxic and postanoxic processes described employ heterotrophic bacteria for nitrate reduction, but other pathways for biological nitrogen removal exist. Denitrification can also be accomplished by heterotrophic and autotrophic nitrifying bacteria. Autotrophic bacteria can also accomplish denitrification under anaerobic conditions, including the autotrophic AOB and the "Anammox" bacteria, which are described in the following section. Denitrifying bacteria in wastewater treatment are presented in the Microbiology section below with the exception of the anammox bacteria.

Microbiology

Bacteria capable of denitrification are both heterotrophic and autotrophic. Denitrification can be done by a wider range of heterotrophic organisms including the following genera: *Achromobacter*, *Acinetobacter*, *Agrobacterium*, *Alcaligenes*, *Arthrobacter*, *Bacillus*, *Chromobacterium*, *Corynebacterium*, *Flavobacterium*, *Halobacterium, Hypomicrobium*, *Methanomonas, Moraxella*, *Neisseria*, *Paracoccus*, *Propionibacterium*, *Pseudomonas*, *Rhizobium*, *Rhodopseudomonas*, *Spirillum*, and *Vibrio* (Payne, 1981 and Gayle, 1989). Most of these bacteria are facultative aerobic organisms with the ability to use oxygen as well as nitrate or nitrite, and some can also carry out fermentation in the absence of nitrate or oxygen.

Use of Supplemental Carbon. Methanol (CH_3OH) and other compounds such as waste glycerin and other commercially available supplemental carbon products have been added to many denitrification systems as a supplemental carbon to increase the amount of nitrate plus nitrite (NO_x) removed and/or to accelerate denitrification rates. As a single carbon compound, methanol degradation is done by more specialized bacteria, which are not the same as those growing on the influent wastewater organic substrates. The most commonly found methanol-utilizing bacteria are of the genus *Hyphomicrobium* (Timmermans and Van Haute, 1983; Sperl and Hoare, 1971) and possibly *Paracoccus denitrificans* (Van Verseveld and Stouthamer, 1978). In population characterization studies of a methanol-fed nitrification-denitrification system by stable isotope probing, Baytshtok et al. (2008) identified a *Methyloversatilis universalis* strain in addition to *Hyphomicrobium zavarzinili*. A unique feature of the *Methyloversatilis universalis* strain was that it could also grow on ethanol, which is not possible for the *Hyphomicrobium zavarzinili*.

Denitrification by Heterotrophic Bacteria. Denitrification can occur under aerobic conditions by heterotrophic nitrifying bacteria (Robertson and Kuenen, 1990, and Patureau et al. 1994), so that simultaneous nitrification and denitrification exist with the conversion of ammonia to gaseous nitrogen products. The heterotrophic bacteria, *Paracoccus pantotropha*, have been studied extensively for simultaneous ammonia oxidation and nitrate reduction. The oxidation of ammonia by heterotrophic bacteria requires energy, which can be obtained by nitrate or nitrite reduction using acetate by *P. pantotropha* under aerobic conditions. The conditions required for this form of denitrification are not considered practical in biological nitrification systems (van Loosdrecht and Jetten, 1998), and their presence was not found by Littleton et al. (2003) under DO condition most favorable for them in a full-scale oxidation ditch system.

Autotrophic Denitrification. A number of autotrophic bacteria have found to be capable of nitrate or nitrite reduction and oxidize a variety of electron donors including zero valance iron and Fe(II) by *Paracoccus ferrooxidans*, *Paracoccus denitrificans*, *P. pantotrophus,* and *P. versutus* (Kumaraswamy et al., 2006 Kielemoes et al., 2000), reduced sulfur compounds by *Thiobacillus denitrificans* (Bock et al., 1995) and ammonia by *Nitrosomonas eutropha*, *Nitrosomonas europaea*, and *Nitrosolobus multiformis* (Poth and Focht 1985, Bock et al., 1995; Zart and Bock 1998; Schmidt et al., 2003). Ammonia oxidation by *Nitrosomonas* under anoxic conditions has been shown to be slow and of little practical significance in activated sludge treatment (Littleton et al., 2003).

Stoichiometry of Biological Denitrification and Denitritation

Biological denitrification involves the biological oxidation of soluble organic substrates in wastewater treatment using nitrate and/or nitrite as the electron acceptor instead of oxygen. In the absence of DO or under limited DO concentrations, the nitrite and nitrate reductase enzymes in the electron transport respiratory chain are induced, which results in the transfer of hydrogen and electrons to nitrite or nitrate as the terminal electron acceptors. Oxidation-reduction reaction stoichiometry for three different electron donors for nitrate reduction is shown as follows. The term $C_{10}H_{19}O_3N$ is often used to represent the biodegradable organic matter in wastewater (U.S. EPA, 1993).

Wastewater:

$$C_{10}H_{19}O_3N + 10NO_3^- \rightarrow 5N_2 + 10CO_2 + 3H_2O + NH_3 + 10OH^- \tag{7–110}$$

Methanol:

$$5CH_3OH + 6NO_3^- \rightarrow 3N_2 + 5CO_2 + 7H_2O + 6OH^- \tag{7–111}$$

Acetic acid:

$$5CH_3COOH + 8NO_3^- \rightarrow 4N_2 + 10CO_2 + 6H_2O + 8OH^- \tag{7–112}$$

It should be noted that acetic acid or acetate is not used commonly because of cost. In all the above heterotrophic denitrification reactions, one equivalent of alkalinity is produced per equivalent of NO_3-N reduced, which equates to 3.57 g of alkalinity (as $CaCO_3$) production per g of nitrate nitrogen reduced. The same amount of alkalinity is produced for NO_2-N reduction. Recall from nitrification that 7.14 g alkalinity (as $CaCO_3$) was consumed per g of NH_4-N oxidized, so that by denitrification about one-half of the amount destroyed by nitrification can be recovered.

Denitritation refers to biological nitrite reduction. Using acetic acid as an example, the oxidation-reduction reaction for nitrite as the electron acceptor is as follows:

$$3CH_3COOH + 8NO_2^- \rightarrow 4N_2 + 6CO_2 + 2H_2O + 8OH^- \tag{7–113}$$

Comparing Eq. (7–112) to Eq. (7–113), denitritation requires 60 percent of the acetate required for denitrification. However, when applying the bioenergetics analysis (Sec. 7–4) to account for COD used for biomass synthesis with the energy yielding oxidation reactions, the COD or acetate used with nitrite reduction is about 67 percent of that used with nitrate reduction, due to the fact that theoretically more energy is produced with nitrite versus nitrate reduction. The following stoichiometric equations developed from bioenergetics can be used to compare the effect of nitrate and nitrite electron acceptors along with ammonia as the nitrogen source on biomass growth:

Nitrate reduction with acetate:

$$\begin{aligned} NO_3^- + H^+ + 0.33\,NH_4^+ + 1.45\,CH_3COO^- \\ \rightarrow 0.5\,N_2 + 0.33\,C_5H_7O_2N + 1.60\,H_2O + 1.12\,HCO_3^- + 0.12\,CO_2 \end{aligned} \tag{7–114}$$

Nitrite reduction with acetate:

$$\begin{aligned} NO_2^- + H^+ + 0.24\,NH_4^+ + 0.98\,CH_3COO^- \\ \rightarrow 0.5\,N_2 + 0.24\,C_5H_7O_2N + 1.24\,H_2O + 0.74\,HCO_3^- + 0.008CO_2 \end{aligned} \tag{7–115}$$

Organic Substrate Requirements for Denitrification and Denitritation

A sufficient amount of organic substrate must be available for depletion of the nitrite/nitrate (NOx) fed to an anoxic zone by recycle and upstream flows. Thus, an important design parameter for nitrogen removal is the amount of bsCOD or BOD needed to provide a sufficient amount of electron donor for NOx reduction. In biological nitrogen removal processes, the electron donor is typically one or more of the following four sources: (1) the bsCOD in the influent wastewater, (2) the bsCOD produced during biological hydrolysis of particulates and colloids, (3) the bsCOD produced during endogenous decay, and (4) an exogenous source such as methanol or acetate. The latter has been added in separate treatment units, such as polishing filters, after nitrification where almost no bsCOD remains. As a general rule for domestic wastewater treatment in a preanoxic aerobic (MLE) process, Barth et al. (1968) estimated that 4 g of BOD is needed per g of NO_3-N reduced. However, the actual value will depend on the system operating conditions and the type of electron donor used for denitrification.

When using an exogenous carbon source it is important to estimate the substrate dose as a function of the NOx to be removed. The dose can be quantified in terms of the g COD added per g of NOx removed and is referred to as the COD/N ratio. The COD/N ratio is determined by a COD balance, similar to what was done to predict the amount of oxygen per unit of bsCOD removed in Example 7–1 (Sec. 7–3). The oxygen equivalent of nitrate and nitrite must be known to perform a COD balance for substrate and NOx removal in a denitrification reactor.

From oxidation-reduction half reactions, the oxygen equivalent of using nitrate or nitrite as electron acceptors can be determined. The half reactions per mole e-transferred from Table 7–6 are as follows:

For oxygen:

$$0.25O_2 + H^+ + e^- \rightarrow 0.5H_2O \tag{7–116}$$

For nitrate:

$$0.20NO_3 + 1.2H^+ + e^- \rightarrow 0.1N_2 + 0.6H_2O \tag{7–117}$$

For nitrite:

$$0.333NO_2^- + 1.333H^+ + e^- \rightarrow 0.6667H_2O + 0.1667N_2 \tag{7–118}$$

Comparing the above half reactions for oxygen [Eq. (7–116)] and nitrate [Eq. (7–117)], 0.25 mole of oxygen is equivalent to 0.2 mole of nitrate for one electron transfer in oxidation-reduction. Thus, the oxygen equivalent of nitrate is (0.25 × 32 g O_2/mole) divided by the nitrate gram equivalent (0.20 × 14 g N/mole) and equals 2.86 g O_2/g NO_3-N. The oxygen equivalent is a useful design factor when calculating the total oxygen required for nitrification-denitrification biological treatment systems. Similarly, for nitrite as the electron acceptor, the oxygen equivalent of nitrite is 1.71 g O_2/g NO_2-N.

It has been shown in Example 7–1 and Eq. (7–6) that the COD removed is either oxidized or accounted for in cell growth. A similar expression can be applied to biological removal of soluble COD:

$$\text{bsCOD}_r = \text{bsCOD}_{\text{cell}} + \text{bsCOD}_{\text{ox}} \tag{7–119}$$

For cell synthesis, the $\text{bsCOD}_{\text{cell}}$ is calculated from the net biomass yield and the ratio of 1.42 g O_2/g VSS.

$$\text{bsCOD}_{\text{cell}} = 1.42Y_n\text{bsCOD}_r \tag{7–120}$$

where Y_n = net biomass yield, g VSS/g bsCOD_r

and

$$Y_n = \frac{Y}{1 + b(\text{SRT})} \tag{7–121}$$

Thus,

$$\text{bsCOD}_r = \text{bsCOD}_{\text{ox}} + 1.42Y_n\text{bsCOD}_r \tag{7–122}$$

Rearranging yields

$$\text{bsCOD}_{\text{ox}} = (1 - 1.42Y_n)\text{bsCOD}_r \tag{7–123}$$

In Eq. (7–123), bsCOD_{ox} is the COD oxidized and is equal to the oxygen equivalent of the NO_3-N used for bsCOD oxidation. Hence,

$$\text{bsCOD}_{ox} = 2.86\,NO_x \tag{7–124}$$

where 2.86 = O_2 equivalent of NO_3-N, g O_2/g NO_3-N
NO_x = NO_3-N reduced, g/d

Substituting Eq. (7–124) in Eq. (7–123) yields

$$2.86\ \mathrm{NO_3} = (1 - 1.42\ Y_n)\ \mathrm{bsCOD}_r \tag{7–125}$$

or

$$\frac{\mathrm{bsCOD}}{\mathrm{NO_3\text{-}N}} = \frac{2.86}{1 - 1.42Y_N} \tag{7–126}$$

$$\text{Thus, g COD/g NO}_3\text{-N} = \frac{2.86}{1 - 1.42Y_N} \tag{7–127}$$

$$\text{and the g COD/g NO}_2\text{-N} = \frac{1.71}{1 - 1.42Y_N} \tag{7–128}$$

If oxygen is entering an anoxic zone via recycle streams, electron donor would be used also for oxygen consumption with following ratio:

$$\text{g COD/g O}_2 = \frac{1.0}{1 - 1.42Y_N} \tag{7–129}$$

The g COD/g NO_3-N in Eq. (7–127) is the same as the substrate "consumptive ratio" defined by McCarty et al. (1969) for using exogenous carbon sources for denitrification. They showed that exogenous substrates with lower biomass synthesis yields had lower consumptive ratios and thus required less substrate addition for a given amount of NO_3-N to be removed. The consumptive ratio for methanol was about 70 percent of that for glucose, due to methanol having a lower synthesis yield, allowing a greater fraction of the COD added to be used for nitrate reduction. Required methanol consumptive ratios are 4.0 to 5.0 g COD/g NO_3-N, based on reported biomass yields ranging from 0.20 to 0.30 g VSS/g COD used (Stensel et al., 1973, Christensen et al., 1994, Purtschert and Gujer, 1999, Sobieszuk et al., 2006, Dold et al., 2008, and Baytshtok et al., 2008). Methanol has a COD of 1.5 g COD/g CH_3OH, so the requirement based on methanol is 2.7 to 3.3 g CH_3OH/g NO_3-N. Field applications in the range of 3.5 to 3.8 are common and account for the yield, DO entering the anoxic zone, and methanol in the anoxic zone effluent. The use of exogenous carbon sources is discussed further in denitrification system designs in Section 8–6.

Denitrification Kinetics

Denitrification rates have often been expressed in terms of a specific denitrification rate (SDNR) based on g NO_3-N/g MLVSS·d, which has been found to vary as a function of the type and amount of biodegradable COD available and the temperature in the anoxic reactor. For preanoxic zones treating domestic wastewater the SDNR may range from 0.04 to 0.25 g NO_3-N/g MLVSS·d. The SDNR in postanoxic tanks, where the substrate for denitrification is provided by endogenous decay, ranges from 0.01 to 0.03 g NO_3-N/g MLSS·d or it can be 5 to 10 times higher if a supplemental substrate is added to the postanoxic tank. Supplemental substrates used to enhance denitrification include methanol, acetate, ethanol, glycerin, corn syrup, and various brewery, distillery, and food processing wastes. For all these cases the soluble organic substrate utilization rate by denitrifying bacteria determines the nitrate and nitrite reduction rates.

Soluble Substrate Utilization Rate. Whether for preanoxic or postanoxic reactors, the soluble substrate utilization rate can be described by modifying Eq. (7–24) to account for the effect of having a zero DO concentration in the anoxic zone and the effect of the nitrate concentration on the substrate utilization kinetics. Note that the effect of the nitrate concentration on substrate utilization rates is only at concentrations below 0.10 to 0.20 g/m³.

$$r_{\text{su}} = \left[\frac{\mu_{H,\max} S_S}{Y_{\text{H}}(K_s + S_S)}\right]\left(\frac{S_{\text{NO}}}{K_{\text{NO}} + S_{\text{NO}}}\right)(\eta)X_H \tag{7–130}$$

Where r_{su} = substrate utilization rate in the reactor, g/m³·d
Y_{H} = heterotrophic bacteria synthesis yield, g VSS/g COD used
$\mu_{H,\max}$ = maximum specific growth rate of denitrifying heterotrophs, g VSS/g VSS·d
S_S = soluble degradable substrate concentration *in the reactor*, g COD/m³
K_S = substrate utilization half-velocity coefficient, g COD/m³
S_{NO} = NO_3-N concentration, g/m³
K_{NO} = nitrate half-velocity coefficient, g/m³
η = fraction of substrate utilization rate with nitrate as the electronic acceptor versus oxygen
X_H = heterotrophic bacteria concentration, g/m³

Additional equations and mass balances are required to determine the reactor soluble degradable COD concentration as a function of the soluble substrate in the reactor influent, the soluble substrate produced from hydrolysis of particulate and colloidal matter, and from endogenous decay.

Nitrate Consumption Rate. The reactor nitrate consumption rate (r_{NO}) can be determined as a function of the substrate utilization rate by applying Eq. (7–127).

$$\text{g COD/g NO}_3\text{-N} = \frac{r_{\text{su}}}{r_{\text{NO}}} = \frac{2.86}{1 - 1.42Y_N}$$

Substituting for r_{su} from Eq. (7–130) gives the reactor nitrate consumption rate:

$$r_{\text{NO}} = \left(\frac{1 - 1.42Y_{\text{H}}}{2.86}\right)\left[\frac{\mu_{\text{H,max}} S_S}{Y_{\text{H}}(K_s + S_S)}\right]\left(\frac{S_{\text{NO}}}{K_{\text{NO}} + S_{\text{NO}}}\right)(\eta)X_{\text{H}} \tag{7–131}$$

where:

r_{NO} = nitrate consumption rate, g/m³·d
Other terms as defined previously.

The term η in Eqs. (7–130) and (7–131) is applied to account for the fact that a lower equivalent oxygen consumption rate occurs when the electron acceptor is nitrate or nitrite instead of oxygen. The term is also used to account for possible changes in kinetic rates using NO_x electron acceptors and for the fact that not all of the heterotrophic bacteria can use nitrate/nitrite. The value for η has been found to vary from 0.20 to 0.80 for preanoxic denitrification reactors fed domestic wastewaters (Stensel and Horne, 2000). The activated sludge configuration, the system SRT, and the fraction of influent BOD removed with nitrate appear to affect the η value. For anoxic/aerobic processes with substantial substrate and nitrate utilization in the preanoxic zone an η value of 0.80 is often appropriate.

The nitrate consumption rate in Eq. (7–131) can be applied to anoxic reactor applications in which the substrate used is derived from the influent wastewater, endogenous decay, or from supplemental carbon addition. When supplemental carbon addition is used, the heterotrophic bacteria concentration (X_H) may have to be based on growth using only the supplemental substrate, and not from growth on the supplemental substrate and the BOD removed from the influent wastewater. This growth situation occurs when methanol is added as the supplemental carbon, which results in the growth of only specialized methylotrophic bacteria from methanol utilization. Also when applying the substrate utilization kinetic model to a system with both influent and supplemental carbon, Eq. (7–131) has to be applied separately and cumulatively to account for both substrate types. Appropriate coefficients must be used for $\mu_{H,\max}$, Y_{H}, K_S, and η for the different substrates and bacteria populations.

For separate postanoxic suspended growth or attached growth processes after the activated sludge BOD removal and nitrification process, the biomass is developed under mainly anoxic conditions and with a selected single organic substrate. In this case, the η term in Eq. (7–130) is not necessary because the biomass consists of mainly denitrifying bacteria. The biokinetic equations presented previously can then be used with the appropriate kinetic coefficient values to design a postanoxic complete-mix suspended growth process. The kinetic coefficient values for growth using methanol have been developed at 10 and 20°C in laboratory studies (Randall et al., 1992). The kinetics for methanol utilization are such that the SRT values required for a denitrification suspended growth process are in the same range as SRT values for aerobic systems designed for BOD utilization only, about 3 to 6 d.

Effect of Dissolved Oxygen Concentration. Dissolved oxygen can inhibit nitrate reduction by repressing the nitrate reduction enzyme. In activated sludge flocs and biofilms, denitrification can proceed in the presence of low bulk liquid DO concentrations. A dissolved oxygen concentration of 0.2 mg/L and above has been reported to inhibit denitrification for a *Pseudomonas* culture (Skerman and MacRae, 1957; Terai and Mori, 1975) and by Dawson and Murphy (1972) for activated sludge treating domestic wastewater. Nelson and Knowles (1978) reported that denitrification ceased in a highly dispersed growth at a DO concentration of 0.13 mg/L. The effect of DO concentration on nitrate utilization rates is accounted for by adding a correction factor to Eq. (7–131) to account for DO inhibition as follows:

$$r_{\mathrm{NO}} = \left(\frac{1 - 1.42Y_{\mathrm{H}}}{2.86}\right)\left[\frac{\mu_{H,\max}S_S}{Y_{\mathrm{H}}(K_s + S_S)}\right]\left(\frac{S_{\mathrm{NO}}}{K_{\mathrm{NO}} + S_{\mathrm{NO}}}\right)\left(\frac{S_o}{K_o' + S_o}\right)(\eta)X_H \tag{7–133}$$

where K_o' = DO inhibition coefficient for nitrate reduction, g/m^3

Other terms are as defined previously.

The value of K_o' is system-specific. Values in the range from 0.1 to 0.2 mg/L have been proposed for K_o' (Barker and Dold, 1997). Assuming a K_o value of 0.1 mg/L the rate of substrate utilization with nitrate as the electron acceptor at DO concentrations of 0.10, 0.20, and 0.50 mg/L would be at 50, 33, and 17 percent of the maximum rate, respectively.

Effect of Simultaneous Nitrification-Denitrification. In activated sludge systems, the issue of DO concentration is confounded by the fact that the measured bulk liquid DO concentration does not represent the actual DO concentration within the activated sludge floc. Under low DO concentration conditions, denitrification can occur in the floc interior, while nitrification is occurring at the floc exterior. Also in activated sludge tanks operated at low DO concentrations, both aerobic and anaerobic zones exist depending on mixing conditions and distance from the aeration point, so that nitrification and denitrification occur in the same tank. Under these conditions, nitrogen utilization that occurs in a single aeration tank is referred to as *simultaneous nitrification and denitrification*. Although both nitrification and denitrification are occurring at reduced rates as indicated by the DO effects described for both processes, if a sufficient SRT and τ exist, the overall nitrogen removal can be significant. Rittman and Langeland (1985) reported greater than

90 percent nitrogen removal by nitrification and denitrification in an activated sludge system used to treat municipal wastewater at DO concentrations below 0.50 mg/L and with values of τ greater than 25.

Environmental Factors

Alkalinity is produced in denitrification reactions and the pH is generally elevated instead of being depressed, as in nitrification reactions. In contrast to nitrifying organisms, there has been less concern about pH influences on denitrification rates. No significant effect on the denitrification rate has been reported for pH between 7.0 and 8.0, while Dawson and Murphy (1972) reported a decrease in the denitrification rate as the pH was decreased from 7.0 to 6.0 in batch unacclimated tests. Biomass acclimation to external carbon sources must also be considered.

7–11 ANAEROBIC AMMONIUM OXIDATION

The oxidation of ammonia by lithotrophic bacteria using nitrate or nitrite as an electron acceptor was proposed as being thermodynamically possible in a 1977 paper by E. Broda, which claimed "two kinds of lithotrophs missing in nature" (Kuenen, 2008). At the same time, based on the analysis of nitrogen balances in a highly stratified anoxic fjord, an unexplained ammonia loss was found (Ward et al., 2011). From the summary of the Gibbs free energy for ammonia oxidation by various electron acceptors given in Table 7–12, bacterial growth with nitrate or nitrite as electron acceptors should be comparable to that with oxygen. Finally in 1995, Mulder et al. (1995) observed anaerobic ammonia oxidation with nitrogen gas production in a laboratory fluidized-sand bed reactor and assigned the name Anammox (anaerobic ammonium oxidation) to the biological process. Tests by Van de *Graaf* et al. (1995) with ^{15}N labeled ammonium and biological inhibitors confirmed that the ammonia oxidation was accomplished with nitrite under anaerobic conditions and was biologically mediated. Strous et al. (1999a) were able to identify the responsible organisms as an autotrophic bacterium under the order Planctomycetales through density gradient centrifugation of an anammox enrichment to characterize its 16S rRNA gene phylogeny. Researchers have since found numerous occurrences of the anammox bacteria in wastewater treatment and in marine and fresh water sediments (Kuenen, 2008, Van Hulle et al., 2010).

Process Description

Anammox has been defined as a "deammonification" process involving two steps: the partial nitritation of ammonia and subsequently the anaerobic oxidation of ammonia and nitrite to nitrogen gas. The anammox process requires that aerobic nitritation to NO_2-N be accomplished for about 55 percent of the NH_4-N in the waste stream before the anaerobic oxidation

Table 7–12

Comparison of Gibbs Energy for inorganic nitrogen oxidation

Inorganic *N* Oxidation Reaction	ΔG°, kJ/mole
$NH_4^+ + 1.5O_2 \rightarrow NO_2^- + 2H^+$	−275
$NO_2^- + 0.5O_2 \rightarrow NO_3^-$	−74
$NH_4^+ + NO_2^- \rightarrow N_2 + H_2O$	−375
$5NH_4^+ + 3NO_3^- \rightarrow 4N_2 + 9H_2O + 2H^+$	−297

Adapted from Schmidt et al. (2003) and Jetten et al. (1999).

of NH_4-N with NO_2-N as the electron acceptor and N_2 production. The process provides a means for biological nitrogen removal without any organic carbon consumption, in contrast to the classic biological aerobic nitrification/anoxic denitrification process. In addition, because it is an autotrophic bacteria, no organic carbon is needed for cell growth. Because only a portion of the ammonia in the deammonification process feed stream is oxidized to nitrite, much less aeration energy is needed compared to the conventional biological nitrogen removal process in which most of the available ammonia is oxidized to nitrate.

The deammonification process has been demonstrated successfully for high temperature (30–35ºC) and high ammonia waste streams in both fixed film and suspended growth biological reactors (Schmidt et al., 2003). A unique feature of the anammox bacteria that has been used to an advantage in suspended growth anammox processes is that they are able to form a compact granular, rapid settling biomass (Innerebner et al., 2007). The anammox process technology has not yet been developed for the reliable removal of nitrogen in low or ambient temperature and dilute ammonia concentrations in wastewater treatment, but the technology has been applied full-scale for nitrogen removal from anaerobic digester dewatering recycle streams. Anammox process designs for side stream treatment are presented in Chap. 15

Microbiology

Investigations on anammox bacteria have resulted in identifying nine species within five genera as members of the bacteria order Planctomycetales (Ward et al., 2011). They are preceded by the name Candidatus, which is used when a species or genus is well characterized but has not been studied in pure culture. These are "Candidatus *Kuenenia*," "*Candidatus Brocadia*," "*Candidatus Scalindua*," "*Candidatus Jettenia*," and "*Candidatus Anammoxogloubus.*" The species found in wastewater are *Kuenenia stuttgartiensis*, *Anammoxoblobus propionicus*, *Jettenia asiatica, Brocadia anammoxidans, Brocadia fulgida, Scalindua wagneri,* and *Scalindua brodae. Scalindua sorokinii* and *Scalindua arabica* were found in marine environments (Van Hulle et al., 2010). *"*Candidatus *Kuenenia"* and "*Candidatus Brocadia"* are the most commonly found organisms in enrichments from WWTPs (Kuenen, 2008). None of these bacteria have been identified in pure cultures but have been characterized by molecular methods (Strous et al., 2002). In contrast to other prokaryotes, the Planctomcyetes typically contain membrane-bound sub-cellular compartments. Jetten et al. (2001) identified such an organelle in *B. anammoxidans*, which contained hydroxylamine oxidoreductase and named it anammoxosome. Anammox bacteria enrichments develop a deep red color, which is thought to be related to hemes present in the hydroxylamine oxidoreductase enzyme, which was referred to as cytochrome P460 (Jetten et al., 1999).

Anammox bacteria are found in dense granular flocs in stable operating systems. Strous et al. (1999b) reported the need to have 10^{10} to 10^{11} cells/mL for an active anammox system. A possible explanation for this is that hydrazine produced in the anammox reaction may be lost by diffusion to the bulk liquid in smaller flocs and thus limit anammox activity and growth. The addition of hydroxylamine or hydrazine can stimulate anammox bacteria activity (Van Hulle et al., 2010). Seeding from a previous anammox system is necessary to avoid an extensive time period for the start of a new anammox systems.

Anammox Stoichiometry

The energy yielding reaction in the anammox process is the oxidation of NH_4-N by NO_2-N:

$$NH_4^+ + NO_2^- \rightarrow N_2 + H_2O \tag{7–134}$$

A proposed metabolic model by Van de Graaf et al. (1997) consists of (1) the reduction of nitrite to hydroxylamine (NH_2OH), (2) the condensation of hydroxylamine with ammonium to hydrazine (N_2H_4), and (3) the oxidation of hydrazine to nitrogen gas. During the conversion of ammonia some nitrate is formed from nitrite, which may provide the reducing power for fixation of carbon dioxide (Schmidt et al., 2002). The anabolic reaction is as follows (van Niftrik et al., 2004):

$$CO_2 + 2NO_2^- + H_2O \rightarrow CH_2O + 2NO_3^- \tag{7–135}$$

The following overall reaction, accounting for cell synthesis, has been suggested by Strous et al. (1999b).

$$\begin{aligned} &1.0NH_4^+ + 1.32NO_2^- + 0.066HCO_3^- + 0.13H^+ \\ &\quad \rightarrow 1.02N_2 + 0.26NO_3^- + 0.066CH_2O_{0.5}N_{0.021} + 2.03H_2O \end{aligned} \tag{7–136}$$

The amount of NO_3-N produced accounts for 11.2 percent of the NH_4-N and NO_2-N metabolized. Acid is decreased in the biological reactor due to the anammox reactions as shown by the 0.13 moles of H^+ consumed and the removal of nitrous acid (HNO_2^-) with a smaller molar amount of nitric acid (HNO_3^-) production. As demonstrated in Example 7–8, the amount of alkalinity consumed for the anammox process is less than that for a nitritation and heterotrophic denitritation process.

EXAMPLE 7–8 **Alkalinity Consumption in the Anammox Process** Compare the net amount of alkalinity consumed by the anammox process from the biological conversion of 200 g NH_4-N/m^3 to that for the same amount of NH_4-N conversion to nitrogen gas via biological aerobic nitritation and heterotrophic denitritation with carbon addition.

1. Alkalinity consumed by aerobic NH_4-N oxidation to NO_2-N = 7.14 g as $CaCO_3$/g NH_4-N
2. Alkalinity produced by heterotrophic denitritation = 3.57 g as $CaCO_3$/g NO_x reduced to N_2.
3. Use Eq. (7–136) to determine the amount of NO_2-N produced by nitritation and the alkalinity changes caused by acid consumption in the anammox process

Solution

1. Determine the net alkalinity consumption by nitritation and heterotrophic denitritation for processing 200 g/m^3 NH_4-N.

   Net alkalinity consumed =

   Alkalinity consumed for nitritation – alkalinity produced by denitritation

   $= [(7.14-3.57) \text{ g as } CaCO_3/\text{g N}](200 \text{ g/m}^3) = 714.0 \text{ g/m}^3 \text{ as } CaCO_3$

2. Determine the net alkalinity change by the anammox process.
   a. Determine the alkalinity consumed by NO_2-N production by nitritation.
      Determine the fraction of NH_4-N processed via NO_2-N production using Eq. (7–136)

$$\text{Fraction } NO_2\text{-N} = \frac{1.32 \text{ mole } NO_2}{1.0 \text{ mole } NH_4 + 1.32 \text{ mole } NO_2} = 0.57$$

NO_2-N produced = $0.57(200 \text{ g/m}^3) = 113.8 \text{ g/m}^3$

Alkalinity used for NO_2-N production:

$= (7.14 \text{ g as } CaCO_3/\text{g N})(113.8 \text{ g/m}^3) = 812.5 \text{ g/m}^3 \text{ as } CaCO_3$

b. Determine the net acid consumed in eq/mole N processed in the anammox reaction in Eq. (7–136). Note the computed value is equivalent to alkalinity production

i. Net mole H^+ consumed/mole N processed =

$$\frac{0.13 \text{ mole } H^+ - 0.066 \text{ mole } HCO_3^-}{2.32 \text{ mole N}} = 0.0276 \text{ mole H consumed/mole N}$$

Equivalent of H/mole H = 1.0

ii. Net change in acid due to nitrous acid depletion and nitric acid production in mole/mole N processed

$$= \frac{1.32 \text{ mole } HNO_2 - 0.26 \text{ mole } HNO_3}{2.32 \text{ mole N}} = 0.457 \text{ mole acid/mole N}$$

Equivalent of acid/mole acid = 1.0

iii. Total acid consumption in eq/mole N processed =

$$= [0.0276 \text{ mole H/mole N} + 0.457 \text{ mole acid/mole N}](1.0 \text{ eq/mole})$$

$$= 0.4844 \text{ eq alkalinity production/mole N processed}$$

Alkalinity produced/g N =

$$\left(\frac{0.4846 \text{ eq}}{\text{mole N}}\right)\left(\frac{50 \text{ g as } CaCO_3}{\text{eq}}\right) = \frac{24.23 \text{ g as } CaCO_3}{\text{mole N}}$$

$$\text{mole N processed} = \frac{200 \text{ g N/m}^3}{14 \text{ g N/mole N}} = 14.29 \text{ mole N/m}^3$$

Alkalinity produced in g/m^3 =

$$= \left(\frac{24.23 \text{ g as } CaCO_3}{\text{mole N}}\right)\left(\frac{14.29 \text{ mole N}}{\text{m}^3}\right) = 346.2 \text{ g as } CaCO_3$$

Net alkalinity consumed = (812.5 − 346.2) g/m^3 = 466.3 g/m^3 as $CaCO_3$

This compares to 714.0 g/m^3 as $CaCO_3$ for the nitritation and heterotrophic denitritation process.

Anammox Seeding. Anammox bacteria are relatively slow-growing bacteria, and an initial seed source with the typical dense granular anammox floc greatly accelerates the start up of a new anammox process application (Strous et al., 1999a). Strous et al. (1999a) found that a critical bacteria concentration of 10^{10} to 10^{11} cells/mL was necessary to realize good anammox activity. A possible reason is that for small aggregates the hydrazine produced diffuses quickly to the bulk liquid. The addition of hydroxylamine or hydrazine was found to accelerate anammox bacteria activity (Van Hulle et al., 2010).

Alternative Anammox Substrates. Anammox bacteria are capable of nitrate reduction with the oxidation of formate, acetate, and propionate as electron donors (Guven et al., 1995; Kartal et al., 2007a). These organic acids are not used in cell growth. The ability to reduce nitrate was found with five anammox species: *B. anammoxidans*, *B. fulgida,* *A. propionicus*, *K. stuttgartiensis*, and *Scalindua sp*. (Ward et al., 2011). Whereas heterotrophic denitrification bacteria reduce nitrate to nitrogen gas via nitrite, nitric oxide, and nitrous oxide as shown in Sec. 7–10, through ^{15}N labeling studies, it has been found that the anammox bacteria reduce nitrate to nitrite and then to ammonia (Kartal et al., 2007a). Both ammonia and nitrite serve as substrates for the anammox reaction in the reduction of nitrate to nitrogen gas. With propionate addition, it has been found that *A. propionicus* can outcompete the growth of other anammox bacteria and heterotrophic denitrifiers (Kartal et al., 2007b). It also appears the nitrate reduction with propionate occurs simultaneously with the anammox reaction. Of the anammox species listed above, *B. fulgida* is more competitive for acetate. The long-term growth of *B. fulgida* in a granular sludge reactor at a lower temperature than most previous anammox applications (18°C) and the ability to remove acetate and nitrate at a COD/N ratio of 0.50 have been demonstrated (Winkler et al., 2012). Higher ratios may encourage the growth of heterotrophic denitrifiers.

Growth Kinetics

Most of the information on anammox bacteria kinetics has been obtained from studies and reactor operations at 30 to 35°C. At this temperature, favorable growth rates have been observed, which also coincides with the temperatures in the treatment of anaerobic digester reject water with the anammox process for biological nitrogen removal. Anammox growth between 4 and 43°C has been claimed by Ward at al. (2011), with lower temperature growth based on observations of anammox activity in Arctic environments. Sustained growth at 15°C (Ward et al., 2011) and 18°C (Winkler et al., 2012) has been demonstrated in laboratory reactors. In batch tests using anammox bacteria from a 30°C reactor, Strous at al. (1997a) found maximum ammonium oxidation rates between 30 and 35°C, but the rates were less than 5.0 percent of that at temperatures below 20°C. Rates would be higher with temperature acclimation.

The biokinetics of anammox bacteria are compared to that for ammonia oxidation by AOB in Table 7–13. At 30°C the anammox bacteria maximum specific growth rate is less than 10 percent of that for AOB. Because of the reduced growth rate the SRT for ammonia oxidation by anammox must be about 10 times longer than that for aerobic ammonium oxidation. However, a maximum specific growth rate that is almost twice as high as that shown in Table 7–13 for anammox bacteria has been reported by van der Star et al. (2008). The ability to achieve a long SRT for the anammox process can be aided by the dense granular floc that is developed. The anammox bacteria have a very high affinity for ammonia and nitrite as indicated by their very low half-velocity coefficients in the table. The biomass yield for anammox bacteria is in the same range as that for AOB and much lower

Table 7–13

Comparison of biokinetics of ammonia oxidizing bacteria (AOB) and anammox bacteria

Parameter	Units	AOB[a] at 20°C	Anammox at 30–35°C	Anammox references
μ_{max}	g VSS/g VSS·d	0.90	0.06–0.07	Jetten et al. (2001), Schmid et al. (2003), Strous et al. (1998)
K_{NH_4}	g/m^3	0.50	<0.10 0.07	Strous et al. (1999) Jetten et al. (2001)
K_{NO_2}	g/m^3		<0.10	Strous et al. (1999)
Yield	g VSS/g NH_4-N	0.12	0.07–0.13	Schmid et al. (2003), Strous et al. (1999)

[a] Typical values from Section 7–8.

than for heterotrophic bacteria, as is typical for autotrophic bacteria with their energy needs for CO_2 fixation.

Environmental Factors

Important environmental factors affecting anammox bacteria activity are pH, and nitrite and DO concentration. Growth at pH 6.7 to 8.3 is feasible, with pH 8.0 reported to be optimal (Strous et al., 1999a).

Nitrite Inhibition. Nitrite, a key substrate in the anammox process, inhibits anammox bacteria activity at elevated concentrations. The inhibitory nitrite concentrations are affected by the reactor operational mode. Nitrite-N concentrations at 60 g/m^3 were not inhibitory in an anammox SBR operation with ammonium and nitrite in the feed stream (Strous et al., 1999) and van der Star et al. (2007) indicated stable growth at nitrite-N concentrations fed from 40 to 80 g/m^3 in a full-scale anammox reactor. However, in an intermittently aerated reactor in which nitrite was produced in the aerobic period, Wett et al., (2007) found inhibition at nitrite-N concentrations as low as 5.0 g/m^3 and an irreversible inhibition at 50 g/m^3. Sustained nitrite-N concentrations above 5.0 g/m^3 resulted in loss of anammox activity and were considered a long term toxic concentration (Wett et al., 2010). The addition of hydrazine at 3.0 g/m^3 was able to restore anammox activity in a previously inhibited reactor (Strous et al., 1999a).

Dissolved Oxygen Inhibition. Anammox bacteria are inhibitated by DO at very low concentrations as they are strict anaerobic bacteria. The effective inhibitory DO concentration is a function of the granular anammox floc thickness or fixed film thickness as AOB can consume oxygen in the outer floc or biofilm. Strous et al. (1997b) found that DO inhibition was reversible and thus suggested that an intermittently aerated reactor for nitritation and deammonification was possible.

7–12 GREENHOUSE GAS FROM BIOLOGICAL NITROGEN TRANSFORMATIONS

Certain biological nitrogen transformation processes produce nitric oxide (NO) and nitrous oxide (N_2O). The impact of nitrous oxide, one of the most potent greenhouse gases, is estimated to be 300 times that of carbon dioxide. It also has the greatest impact on ozone depletion, which is expected to continue as such through the 21st century (Ravishankara et al., 2009). Nitric oxide is not considered a greenhouse gas but can be oxidized rapidly in air to nitrogen dioxide ($NO_{2,g}$), which is a toxic air pollutant and a contributor to smog.

Source of Nitrous Oxide Emissions

Agriculture is the major source of nitrous oxide emissions, and wastewater treatment sources have been estimated as the sixth largest contributor, accounting for about three percent of the total global emissions in 2006. The wastewater contribution is expected to rise between now and 2020 (Law et al., 2012). Based on the results of monitoring studies at twelve WWTPs in the United States, it was found that nitrous oxide emissions varied from 0.01 to 1.8 percent of the influent nitrogen, but other investigators have found emissions as high as 15 percent (Ahn et al., 2010a). Nitrous oxide emissions were higher in the aerated zones than the anoxic zones of anoxic/aerobic biological nitrogen removal processes, and it varied with diurnal nitrogen loading rates (Ahn et al., 2010b). Higher gas stripping from aeration versus the nonaerated mixed anoxic zones must also be considered in comparing nitrous oxide emissions in biological nitrogen removal processes.

Nitrous Oxide Production Pathways

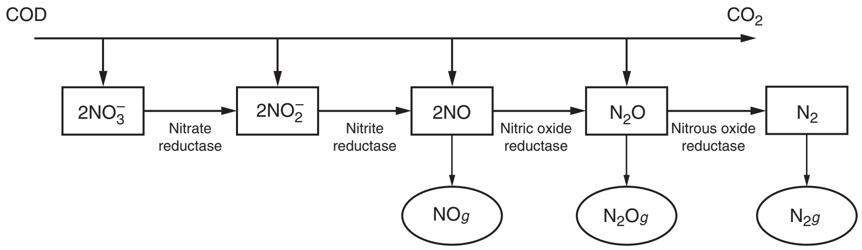
Nitrous oxide is produced during biological nitritation by AOB oxidation of ammonia to nitrite and during biological denitrification by heterotrophic bacteria oxidizing organic substrates with nitrate or nitrite as the initial electron acceptor. There is no evidence for direct nitrous oxide production by NOB. The anammox bacteria and AOB capable of deammonification are also not expected to produce nitrous oxide, as their degradation pathway under anaerobic conditions involves a direct reaction between ammonia and nitrite to produce dinitrogen gas. Presently, there is insufficient information to determine if the AOA have the necessary genes and pathways for nitrous oxide production (Ward et al., 2011).

Nitrous Oxide Production from Heterotrophic Denitrification. The pathways for nitrous oxide production during heterotrophic denitrification and aerobic ammonia oxidation are summarized on Fig. 7–22. The heterotrophic denitrification pathway is the same as that given in Eq. (7–109). The COD is oxidized under anoxic conditions with electron transfer to nitrate, nitrite, nitric oxide, and finally nitrous oxide to produce dinitrogen gas. Under steady operating conditions, the nitrous oxide reduction rate is almost four times faster than the nitrate and nitrite reduction rates, and thus, little if any nitrous oxide would be present (Wicht, 1996). But under transient conditions nitrous oxide can accumulate because the induction of nitrous oxide reductase is slower than that for the upstream reductase enzymes (Holtan-Hartwig et al., 2000).

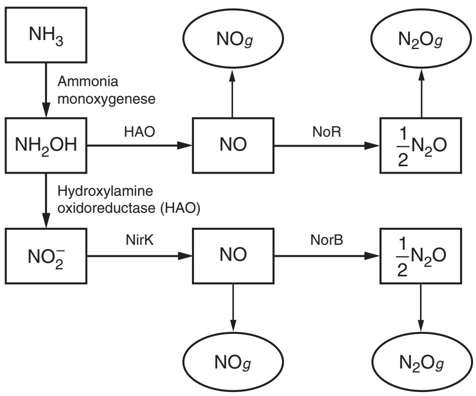
Dissolved oxygen affects nitrous oxide production by inhibiting the synthesis and activity of nitrous oxide reductase, and its activity has been found to stop immediately when the denitrifying bacteria moved from an anaerobic to aerobic environment. Nitrite reductase activity continues at a lower rate under the same transition, so that nitrous oxide emissions will occur (Law et al., 2012). The effect of DO is not the same for all denitrifying bacteria.

Figure 7–22

Pathways for nitrus oxide (N_2O) production in biological nitrogen transformation.



(a) Heterotrophic denitrification



(b) Ammonia oxidation

Lu and Chandran (2010) found that bacteria grown with methanol substrate were much less sensitive to transient DO changes than bacteria grown with ethanol. Insufficient carbon for complete denitrification can also cause more nitrous oxide production (Rassamee et al., 2011).

Nitrous Oxide Production from Ammonia Oxidation. Biological ammonia oxidation to nitrite plays a major role in nitrous oxide production. As illustrated on Fig. 7–22(b), two pathways have been proposed for the production of nitrous oxide by AOB: (1) hydroxylamine (NH_2OH) oxidation and (2) denitrification of nitrite (Ward et al., 2011 and Law et al., 2012).

In the upper pathway, aerobic oxidation of ammonia is started with the membrane bound ammonia mono-oxygenase enzyme (AMO) to produce NH_2OH. The subsequent step of converting NH_2OH to NO_2^- is catalyzed by hydroxylamine oxidoreductase (HAO), which involves a nitroxyl radical (NOH) and conversion of NOH to NO_2^-. Nitric oxide can be formed from HAO activity and the NOH intermediate as shown on Fig. 7–22. The NO thus formed can then be reduced to N_2O. There are also other enzymes, especially in AOB, which can convert NO to N_2O, which for the sake of brevity could be collectively termed NO reductases.

In the second pathway involving nitrite reduction, the AOB contain gene homologues that encode for the production of nitrite reductase (NirK) and nitric oxide reductase (NorB), respectively. There is no gene to express nitrous oxide reductase, and thus nitrous oxide is the end product of nitrite reduction by AOB (Yu et al., 2010). AOB can use hydroxylamine, hydrogen, and ammonia as electron acceptors for nitrite and nitric oxide reduction (Poth and Focht, 1985, Bock et al., 1995, and Ritchie and Nicholas, 1972). Nitrous oxide production by AOB can occur at both low DO (Poth and Focht, 1985) and high DO concentrations (Beaumont et al., 2004), depending on the level of gene expression for *nirK* and *norB*. A key factor to inducing enzymes for more nitrous oxide production is higher specific AOB ammonium oxidation rates, which occurs with an increase in the reactor ammonium concentration and DO concentration (Yu et al., 2010).

An exponential increase in nitrous oxide production with increasing specific ammonium oxidation rates has been shown by Law et al. (2012). Such conditions can occur in biological nutrient removal processes, with a preanoxic contact tank before aerobic nitrification or transient operating conditions that caused spikes in ammonium concentration (Yu et al., 2010; Kampschreur et al., 2008). Higher nitrite concentrations during nitrification can also cause more nitrous oxide emission (Kim et al., 2010; Yang et al., 2009; Gustavsson and Jansen, 2011; Ahn et al., 2011). Nitrification processes that have less variable conditions in ammonia and DO concentration and low nitrite concentration are more favorable for minimizing nitrous oxide emissions. In anoxic/aerobic activated sludge systems for biological nitrogen removal, the nitrous oxide emissions are greater from the nitrification step than from the preanoxic zone due to conditions that increase the ammonia concentrations (Chandran et al., 2011).

Nitrous Oxide Production from Anammox Deammonification. Nitrous oxide emissions have been observed in applications of the anammox deammonification process. Although anammox bacteria by themselves cannot produce N_2O, they can produce NO, which might be reduced to N_2O by other bacteria in the mixed population, including AOB and heterotrophic denitrifying bacteria. Nitrous oxide production has been related to a carbon limitation by having only endogenous respiration to provide carbon for heterotrophic denitrification (Schneider et al., 2011). In another study, a recommendation to avoid excessive nitrous oxide emissions was to operate the anammox process with a minimal reactor nitrite concentration (Weissenbacher et al., 2010).

7–13 ENHANCED BIOLOGICAL PHOSPHORUS REMOVAL

Phosphorus removal from wastewater treated effluent is done to control eutrophication because phosphorus is a limiting nutrient in most freshwater systems. Treatment plant effluent discharge permit limits for phosphorus removal have ranged from 0.10 to 1.0 mg/L of phosphorus, and in some locations as low as 0.05 mg/L, depending on plant location and potential impact on receiving waters. Phosphorus can be removed by chemical treatment, biological phosphorus removal, or a combination of both. Chemical treatment using alum or iron salts, followed by tertiary filtration or membrane separation, is the technology used most commonly for achieving low effluent phosphorus concentrations (see Sec. 6–4 in Chap. 6).

Enhanced biological phosphorus removal involves the incorporation of phosphorus in the biomass produced in the treatment system and subsequently the removal of the biomass in sludge wasting. Biomass produced by cell growth by ordinary heterotrophic bacteria from BOD removal may contain about 0.015 g P/g VSS, such that only 10 to 20 percent phosphorus removal can be expected by this pathway in domestic wastewater treatment. However, since the late 1970s full-scale plant design configurations that select for phosphorus storing bacteria commonly called *phosphorus accumulating organisms* (PAOs) have been used to provide over 80 percent biological phosphorus removal. These processes have been referred to as *enhanced biological phosphorus removal* (EBPR). The principal advantages of enhanced biological phosphorus removal are reduced chemical costs and less sludge production as compared to chemical precipitation. A review of the discovery and implementation of EBPR processes is provided by Barnard (1998). In addition EBPR captures influent phosphorus in a form that is suitable for phosphorus recovery in wastewater treatment.

Process Description

Phosphorus accumulating organisms (PAOs) are encouraged to grow and consume phosphorus in systems that use a reactor configuration that provides PAOs with a competitive advantage over other bacteria. The reactor configuration utilized for phosphorus removal is comprised of an anaerobic tank having a hydraulic retention time, τ, of 0.50 to 1.0 h that is placed ahead of the activated sludge aeration tank (see Fig. 7–23). The contents of the anaerobic tank are mixed to provide contact with the return activated sludge and influent wastewater. Anaerobic contact tanks have been placed in front of many different types of suspended growth processes (see detailed discussion in Sec. 8–6 in Chap. 8), with aerobic SRT values ranging from 3 to 40 d.

The key to the PAO competitive advantage in the anaerobic contact zone is that they are able to transport and consume influent readily biodegradable COD (rbCOD) in the form of volatile fatty acids (e.g., acetic acid and propionic acid) by using energy made available from their stored phosphorus as polyphosphates. They are also able to use that energy along with cellular glycogen to direct the substrate uptake to internal carbohydrate storage products for oxidation later in the aerobic zone. The other heterotrophic bacteria in the anaerobic zone are not able to consume the rbCOD because they need an electron acceptor, such as oxygen, nitrate, or nitrite, for oxidation-reduction reactions to provide energy for substrate utilization.

The anaerobic zone in the anaerobic/aerobic treatment process is termed a “selector,” because it provides conditions that favor the uptake of rbCOD and proliferation of the PAOs over other heterotrophic bacteria. Because the PAOs prefer low-molecular-weight fermentation product substrates, the preferred food source would not be available without the anaerobic zone that provides for the fermentation of the influent rbCOD to acetate.

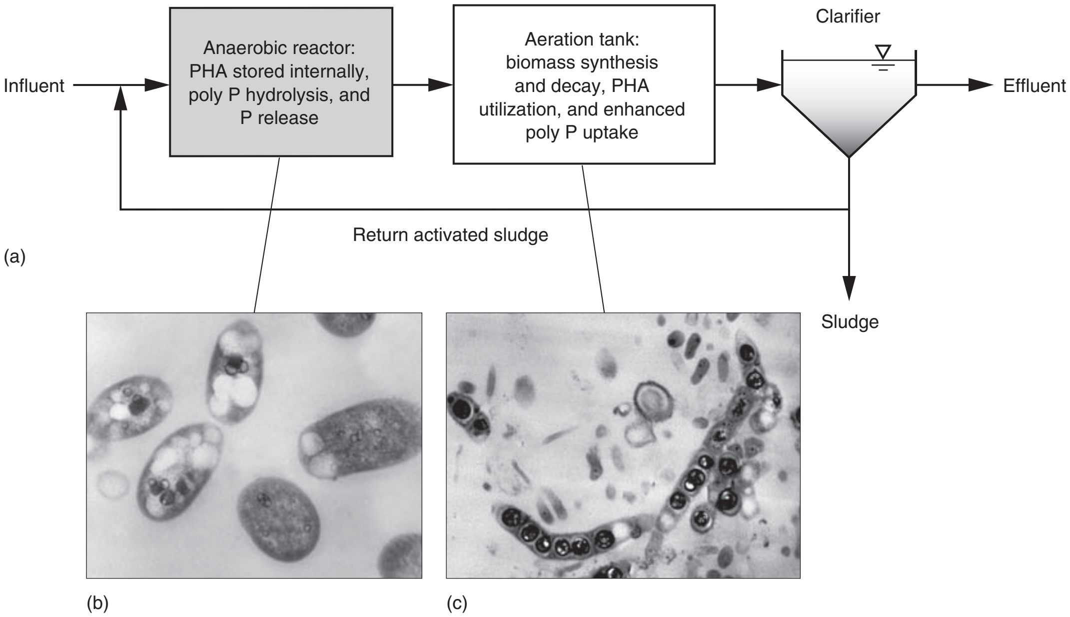


Figure 7–23

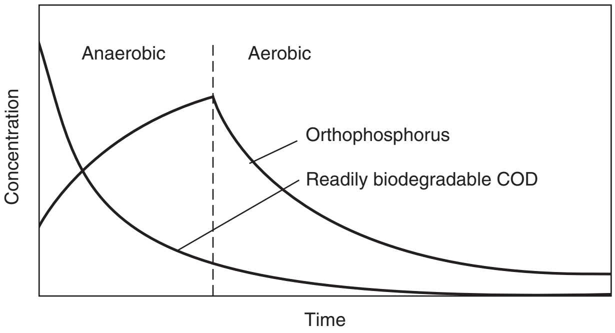
Enhanced biological phosphorus removal: (a) typical reactor configuration. Photos below flow diagram are of (b) transmission electron microscope image of polyhydroxybutyrate storage and (c) polyphosphate storage granules.

Other aerobic heterotrophic bacteria have no such mechanism for acetate uptake, and they are starved while the PAOs assimilate COD in the anaerobic zone. It should also be noted that the PAOs form very dense floc that settles well, which is an added benefit. In some facilities, the anaerobic/aerobic process sequence has been used as a "selector" to produce a sludge that settles well, even though EBPR was not required.

The biological process steps for EBPR are summarized below and indicated in the anaerobic and aerobic zones in Fig. 7–23. An example of the change in orthophosphorus and rbCOD concentrations in a batch anaerobic/aerobic operation after wastewater feeding is given on Fig. 7–24. The anaerobic zone soluble orthophosphorus concentration is often two to three times higher than the influent soluble orthophosphorus concentration and is a good indicator of an active EBPR system. The phosphorus release rate is normally faster than the subsequent phosphorus uptake rate.

Figure 7–24

Fate of rbCOD and phosphorus in enhanced biological phosphorus removal reactor. (Adapted from Sedlak, 1991.)



Processes Occurring in the Anaerobic Zone

Acetic and propionic acid are consumed by PAOs in the anaerobic zone. These VFAs may enter the anaerobic zone in the influent wastewater and are also produced by other facultative bacteria through fermentation of influent rbCOD, which, as defined earlier, is dissolved degradable organic material that can be assimilated easily by the biomass. Most of the volatile fatty acids available are in the form of acetate. Depending on the anaerobic zone hydraulic retention time, some colloidal and particulate COD is also hydrolyzed and converted to acetic acid/propionic acid, but the amount is generally much less than that from the rbCOD.

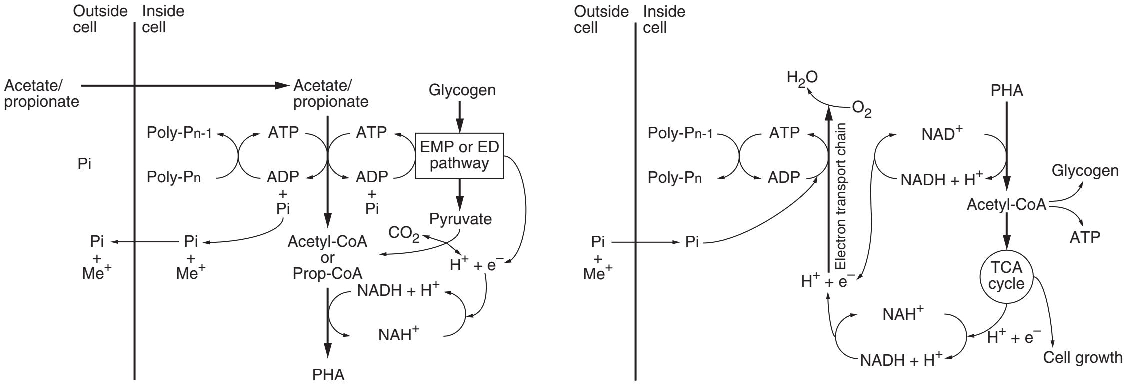
Using energy available from stored polyphosphates, the PAOs assimilate acetate and produce intracellular poly-β-hydroxyalkanoate (PHA) storage products. The typical PHAs are poly-β-hydroxybutyrate (PHB) and polyhydroxyvalerate (PHV). Some glycogen contained in the cell is also used for PHA storage. Concurrent with the acetic acid/propionic acid uptake is the release of orthophosphate (O-PO_4), as well as magnesium, potassium, and calcium cations. The PHA content in the PAOs increases while their polyphosphate content decreases.

Processes Occurring in a Downstream Aerobic or Anoxic Zone

The principal processes occurring downstream are

1. Stored PHA is metabolized, providing energy from oxidation and carbon for new cell growth.
2. Some glycogen is produced from PHA metabolism.
3. The energy released from PHA oxidation is used to form polyphosphate bonds in cell storage so that soluble orthophosphate (O-PO_4) is removed from solution and incorporated into polyphosphates within the bacterial cell. Cell growth also occurs due to PHA utilization and the new biomass with high polyphosphate storage accounts for phosphorus removal.
4. As a portion of the biomass is wasted, stored phosphorus is removed from the biotreatment reactor for ultimate disposal with the waste sludge.
5. These processes can also occur in a downstream anoxic zone prior to anaerobic zone, as there are PAO species that can use nitrate and nitrite as the electron acceptor for substrate oxidation.

Biochemical models have been presented to describe the interactions between acetate uptake, polyphosphates, and PHA storage in the anaerobic and aerobic zones and a representation of these is given on Fig. 7–25 (Comeau et al., 1986, Wentzel et al., 1991, Smolders et al., 1995, and Mino et al., 1998). In the anaerobic zone, energy is required for transport of acetate across the cell and for conversion of it to acetyl coenzyme A (acetyl-CoA). As the cell uses energy, adenosine diphosphate (ATP) is converted to adenosine disphosphate (ADP). The ADP is regenerated to ATP from hydrolysis of the stored polyphosphates with release of phosphorus and a metal cation for charge balance, typically potassium or magnesium. The PHA is produced from the acetyl-CoA and the reducing power provided by degradation of glycogen through the Embden-Meyerhof or Entner-Doudoroff pathways. The glycolysis also provides some ATP for the production of acetyl-CoA. In the aerobic zone the PAOs use the stored PHA as an energy and carbon source for cell growth. Metabolism of the PHA results in cell growth, glycogen production by gluconeogenisis, and energy production through electron transport phosphorylation with



(a) Anaerobic

(b) Aerobic

Figure 7–25

Biochemical model for phosphorus release and removal by PAOs: (a) under anaerobic conditions and (b) under aerobic conditions.

oxygen or nitrate/nitrite as the final electron acceptor. With the energy produced, polyphosphate synthesis occurs with uptake of orthophosphate and the metal cations from solution.

Microbiology

Isolation of PAOs has not been successful, but through the use of molecular techniques Bond et al. (1995) identified PAOs in the *Rhodocyclus* group in the *Betaproteobacteria* named them "Candidatus *Accumulibacter Phosphatis*," which have subsequently been divided into Accumulibacter Type I and Type II (Oehmen et al., 2010). Type I can use nitrate or nitrite as electron acceptors and are referred to as denitrifying PAOs (DPAOs) (Nielsen et al., 2010). The complete genome of *A. Phosphatis* has been sequenced to provide valuable information about the gene makeup and metabolism (Martin et al., 2006). A *Tetrasphaera*-related PAO within the *Actinobacteria* has also been found in domestic WWTPs and with greater abundance in industrial WWTPs (Kong et al., 2005). *Tetrasphaera* do not store PHA, may prefer amino acids, and can use nitrate but not nitrite in addition to oxygen as an electron acceptor. However, information on their metabolism is limited.

A competitor to PAOs for volatile fatty acid (VFA) uptake under anaerobic conditions is referred to as a glycogen accumulating organism (GAO). This type of organism was originally observed as tetrad forming cells by Cech and Hartman (1990) and referred to as "*G*" bacteria because of their growth and glycogen storage with glucose feed. The term GAO is from Mino et al. (1995) based on the organism's storage of glycogen under aerobic conditions and consumption of glycogen under anaerobic conditions to provide energy for VFA uptake and production of PHA in the anaerobic zone of an EBPR system. The GAOs do not have stored polyphosphates for energy during VFA uptake and do not provide EBPR. One indicator of having a competitive GAO population in an EBPR system is a reduced phosphorus release to acetate uptake ratio in the anaerobic zone to well below the typical value of 0.50 g P/g acetate for PAOs (Gu et al., 2008).

The GAOs have been identified as a phenotype belonging to *Gammaproteobacteria* and were named "Candidatus *Competibacter Phosphatis*" by Crocetti et al. (2002). These and many GAOs form tetrads, which are large cells in groupings of four. *Defluviicoccus vanus* is another tetrad-forming GAO found within the *Alphaproteobacteria* (Wong et al., 2004), but these are not considered as common in EBPR WWTPs as the *Competibacter*. All of the GAOs identified so far have been shown capable of using nitrate as an electron acceptor in addition to oxygen, but only *Competibacter* Type I can use nitrite as well (Nielsen et al., 2010).

Factors that affect the competition between PAOs and GAOs include the feed acetate and propionate composition, pH, temperature, and SRT. *Accumulibacter* are able to use acetate and propionate at similar rates and outcompete *Competibacter* with only propionate present due to a greatly reduced uptake of propionate versus acetate by *Competibacter* (Oehmen et al., 2006). However, the *Alphaproteobacteria* GAOs consume propionate faster than acetate and thus can compete with the PAOs. A strategy of alternating VFA feed between acetate and propionate almost completely eliminated GAOs, as the *Accumulibacter* could compete against the *Competibacter* under propionate feed and against the *Alphaproteobacteria* under acetate feeding (Lu et al., 2006).

Higher pH values above 7.0 favor PAO growth over GAO, with a pH value above 7.5 being most favorable for PAOs (Lopez-Vazquez et al., 2009a). A critical pH of 7.25 was reported by Filipe et al. (2001), with GAO growth greatly favored below that value due to a faster anaerobic VFA consumption rate at lower pH. Zhang et al. (2005) found that EBPR performance was greatly decreased when the pH was reduced from 7.0 to 6.5.

Effective EBPR performance has been observed at temperatures as low as 5°C (Brdjanovic et al., 1998). At temperatures of 10°C and less, PAOs are greatly favored over GAOs, regardless of pH (Lopez-Vazquez et al., 2009a). Based on a metabolic model for PAOs and GAOs, Lopez-Vazquez et al. (2009b) found that at temperatures between 20 and 30°C, GAO tend to dominate over PAOs unless a high pH is present (≥7.5) or a favorable acetate to propionate feed ratio (75:25 or 50:50, respectively) is used. At temperatures below 15°C and above 30°C, the GAOs have much lower growth rates and require much higher aerobic tank SRTs than the PAOs (Lopez-Vazquez et al., 2009a). A dominant GAO population was changed to a dominant PAO population by Whang and Park (2006) at 30°C by reducing the aerobic SRT from 6 to 1.8 d.

Other Process Considerations for EBPR

If significant amounts of dissolved oxygen or nitrate enter the anaerobic zone, the VFAs can be depleted before it is taken up by the PAOs, and treatment performance will be hindered due to less growth of PAOs. Biological phosphorus removal is not used in systems that are designed with nitrification without including a means for denitrification to minimize the amount of nitrate in the return sludge flow to the anaerobic zone. These processes are described in Sec. 8–7 in Chap. 8.

Phosphorus release by PAOs in the EBPR system or in waste sludge handling can have a negative impact on the WWTP phosphorus removal efficiency. When the PAO-containing mixed liquor, whether in the EBPR process or in waste sludge processing, is held under anaerobic conditions, phosphorus release will occur. Release of ortho phosphate (O-PO_4) is possible even without acetate addition as the bacteria use the stored polyphosphate for an energy source. The release of O-PO_4 can also occur after extended contact time in the anaerobic or anoxic zones of an EBPR system in the absence of VFA. However, because the release was not associated with acetate or propionate uptake with PHA storage for later oxidation, phosphorus may not be taken up in the aerobic zone. The release of O-PO_4

under these conditions is termed secondary release (Barnard, 1984), which can lead to a lower phosphorus removal efficiency for the biological process. Recycle return flow from waste sludge handling processes used in EBPR systems may contain high levels of phosphorus if the waste sludge handling involves anaerobic holding times, such as that from gravity thickeners and dewatering following anaerobic digestion. In addition, phosphorus is released due to biomass destruction in aerobic digestion. Uncontrolled struvite precipitation in sidestream piping system is also a common problem at EBPR plants that use digesters. The chemistry of struvite formation is considered in Chap. 6. The management of these return streams, including phosphorus recovery processes, is addressed in Sec. 15–4. in Chap. 15.

Stoichiometry of Enhanced Biological Phosphorus Removal

Based on the description of the phosphorus removal mechanism, the PAOs remove two other components from the influent wastewater to support their growth and phosphorus removal via waste sludge from an EBPR system: (1) metal cations and (2) VFA as acetate or propionate. The recommended molar ratio of P/Mg/K has been given as 1.0/0.33/0.33 respectively, by Pattarkine and Randall (1999). With consideration for calcium uptake, the recommended molar ratio of P/Mg/K/Ca of 1/0.28/0.26/0.09, respectively, was given by Sedlak (1991). Based on the latter molar ratio, 0.63 g of other inorganics is added to the waste sludge weight per g of phosphorus as P removed by PAOs. An increase in sludge volume is minimally affected due to the dense floc formed by PAOs. The VSS/TSS ratio for an enriched PAO system can be as low as 60 to 65 percent compared to 85 percent for ordinary heterotrophs. Most municipal wastewaters have sufficient amounts of the metal cations needed by PAOs, but care must be taken to assure sufficient amounts in industrial applications or laboratory experiments.

Acetate or propionate uptake in the anaerobic zone is critical in determining the amount of PAOs that can be produced and, thus, the amount of phosphorus that can be removed by this pathway. The amount of phosphorus removed by biological storage can be estimated from the amount of bsCOD that is available in the wastewater influent by assuming an rbCOD fraction that can be converted to acetate in the short anaerobic hydraulic retention time τ. Methods for determining the amount of rbCOD in the influent will be described in Sec. 8–2 in Chap. 8, which deals with wastewater characterization.

The following assumptions are used to evaluate the stoichiometry of biological phosphorus removal: (1) approximately 1.0 g acetate COD/g rbCOD fermented will be produced as most of the COD fermented will be converted to VFAs due to the low cell yield of the fermentation process, (2) a cell yield of 0.45 g VSS/g acetate COD consumed by PAOs, and (3) a cell phosphorus content 0.20 to 0.30 g P/g VSS at 20°C (Panswad et al., 2003). Using these assumptions, about 7 to 11 g of rbCOD will be required to remove 1 g of phosphorus by the EBPR mechanism. Actual values in EBPR systems may range from 8 to 20 g P/g rbCOD, depending on the relative GAO and PAO populations and the fraction of acetate in the influent rbCOD. Other bCOD removal in the activated sludge system will result in additional phosphorus removal by normal cell synthesis.

Better performance for biological phosphorus removal systems is achieved when rbCOD or acetate is available at a steady pace. Periods of starvation or low rbCOD concentrations result in changes in the intracellular storage reserves of glycogen, PHA, and polyphosphates and rapidly lead to decreased phosphorus removal efficiency (Stephens and Stensel, 1998). The amount of phosphorus that can be removed from a wastewater is illustrated in Example 7–9.

EXAMPLE 7–9 **Estimating the Amount of Phosphorus Removal** Given the following influent wastewater characteristics and the corresponding biological process information, estimate the effluent phosphorus concentration. Assume that 90 percent of the rbCOD is fermented to acetate in the EBPR anaerobic contact zone. and that the PAO specific endogenous decay rate is 0.08 g/g-d. No nitrification has occurred in the system, and the amount of DO entering the anaerobic zone is negligible.

Influent	Concentration, g/m^3
COD	300
bCOD	200
rbCOD	50
Phosphorus, as P	6.0

1. Conversion of rbCOD to acetate in the anaerobic zone = 90 percent
2. Bacteria synthesis yield, $Y = 0.45$ g VSS/g COD
3. Endogenous decay coefficient, $b = 0.08$ g VSS/g VSS·d
4. SRT = 5 d
5. Phosphorus content of PAOs = 0.30 g P/g VSS
6. Phosphorus content of other bacteria = 0.02 g P/g VSS
7. Clarifier effluent VSS concentration = 8 g/m^3

Solution

1. Determine phosphorus removed by PAOs.
   a. Acetate COD production = 0.90(50 g/m^3 rbCOD) = 45 g/m^3 COD
   b. Determine PAO biomass produced, normalized to flow, using Eq. (7–54) and neglecting cell debris.

$$\text{PAO biomass produced} = \left[\frac{Y}{1 + b(\text{SRT})}\right]\text{bsCOD}$$

$$= \left\{\frac{(0.45\text{ g VSS/g COD})}{[1 + (0.08\text{ g/g·d})(5\text{d})]}\right\}(45\text{ g bsCOD/m}^3) = 14.5\text{ g VSS/m}^3$$

   c. Determine the phosphorus removed by PAO biomass.

   P removed = (0.30 g P/g VSS)(14.5 g VSS/m^3) = 4.4 g/m^3

2. Determine phosphorus removed by other heterotrophs for cell synthesis from the conversion of bCOD.
   a. Determine the bCOD removed by other heterotrophic bacteria.

   bCOD removed = 200 – 45 g/m^3 = 155 g/m^3

$$\text{Other biomass produced} = \left[\frac{Y}{1 + b(\text{SRT})}\right]\text{bCOD}$$

$$= \left\{\frac{(0.45\text{ g VSS/g COD})}{[1 + (0.08\text{ g/g·d})(5\text{d})]}\right\}155\text{ g bCOD/m}^3 = 49.8\text{ g VSS/m}^3$$

   b. Determine the phosphorus removed.

   P removed = 0.02 g P/g VSS = 0.02(49.8) = 1.0 g/m^3

3. Determine total phosphorus removed into biomass and effluent soluble phosphorus concentration.

   Total P removed = 4.4 + 1.0 = 5.4 g/m^3

   Effluent soluble concentration = 6.0 – 5.4 = 0.60 g/m^3

4. Estimate P content of effluent VSS.

Average P content of effluent VSS

$$= \frac{(0.30\,\text{g P/g VSS})(14.5\,\text{g/m}^3) + (0.02\,\text{g P/g VSS})(49.8\,\text{g/m}^3)}{[(14.5 + 49.8)\,\text{g/m}^3)]}$$

$$= 0.083\,\text{g P/g VSS}$$

Phosphorus in effluent VSS $= 0.083(8\ \text{g/m}^3) = 0.67\ \text{g/m}^3$

Total effluent P concentration $= 0.60 + 0.67 = 1.27\ \text{g/m}^3$

Growth Kinetics

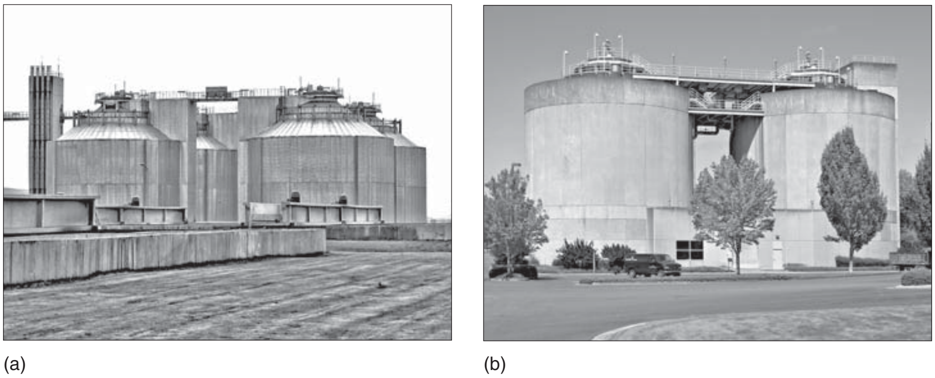
Biological phosphorus growth kinetics are within the same order of magnitude as other heterotrophic bacteria. Mamais and Jenkins (1992) showed that enhanced biological phosphorus removal could be maintained in anaerobic/aerobic systems at aerobic SRTs greater than 2.5 d at 20°C. A maximum specific growth rate at 20°C is given as 0.95 g/g•d (Barker and Dold, 1997).

Environmental Factors

System performance is not affected by DO as long as the aerobic zone DO concentration is above 1.0 mg/L. The effect of pH is primarily related to PAO versus GAO population distribution as given above. Denitrifying PAOs can use nitrite as an electron acceptor in addition to nitrate in the anoxic zone for oxidation of intracellular PHA. However, nitrite concentrations greater than 2.0 g/m^3 are inhibitory to phosphate uptake under both anoxic and aerobic conditions, with a greater effect under aerobic conditions. At 6.0 g/m^3, aerobic uptake of phosphate by PAOs is severely limited (Saito et al., 2004).

7–14 ANAEROBIC FERMENTATION AND OXIDATION

Anaerobic fermentation and oxidation processes are used primarily for the treatment of waste sludge (see Fig. 7–26) and high-strength organic wastes. In warm climates, anaerobic fermentation has been used as a pretreatment step for conventional biological treatment. Applications for dilute waste streams have also been demonstrated. A major


(a)

(b)

Figure 7–26

Views of anaerobic digesters: (a) Ankara, Turkey, and (b) Tigard, OR.

advantage of anaerobic fermentation and oxidation processes are lower biomass yields and energy production in the form of methane from the biological conversion of organic substrates. Although most fermentation processes are operated in the mesophilic temperature range (30 to 35°C), there is increased interest in thermophilic fermentation alone or before mesophilic fermentation for municipal sludge processing. The latter is termed temperature phased anaerobic digestion (TPAD) and is typically designed with a sludge SRT of 3 to 7 d in the first thermophilic phase at 50 to 60°C and 7 to 15 d in the final mesophilic phase (Han and Dague, 1997). Thermophilic anaerobic digestion processes, considered in Chap. 13, are used to accomplish high pathogen kill to produce *Class A biosolids* (defined in Chap. 14) that can be used in the United States for unrestricted reuse applications.

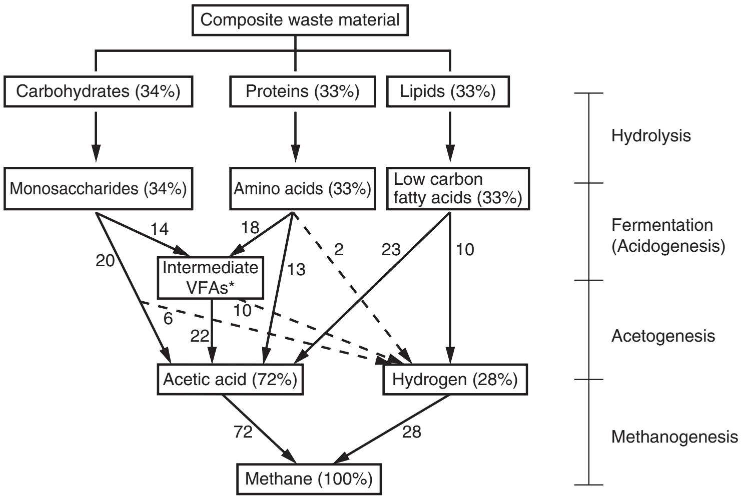
Anaerobic treatment for high-strength industrial wastewaters has been shown to provide a very cost-effective alternative to aerobic processes with savings in energy, nutrient addition, and reactor volume. Because the effluent quality is not as good as that obtained with aerobic treatment, anaerobic treatment is commonly used as a pretreatment step prior to discharge to a municipal collection system or is followed by an aerobic process. Suspended and attached growth anaerobic treatment process designs for liquid streams are presented in Chap. 10, and anaerobic digester designs for sludge treatment are presented in Chap. 13.

Process Description

Three basic steps are involved in the overall anaerobic oxidation of a waste: (1) hydrolysis, (2) acidogenesis (also known as fermentation or anaerobic oxidation), and (3) methanogenesis. The three steps are illustrated schematically on Fig. 7–27, which shows the fate of solids through hydrolysis, volatile fatty acids (VFAs) and hydrogen production to methane. An intermediate step, termed acetogenesis, occurs for some of the VFAs produced from acidogenesis. The starting point on the schematic for a particular application depends on the nature of the waste to be processed. Sometimes the process is intentionally stopped midway such as when primary solids are fermented in gravity thickeners, and the supernatant, which is rich in VFAs, is used for EBPR.

Figure 7–27

Fate of biodegradable COD in anaerobic processing of waste solids. (Adapted from Jerris and McCarty, 1963, 1981, and Batstone et al., 2006.)



* Propionate, butyrate, valerate

Hydrolysis. The first basic step, in which particulate material is converted to soluble compounds that can then be hydrolyzed further to simple monomers that are used by bacteria that perform fermentation, is termed *hydrolysis*. For some high strength soluble industrial wastewaters, fermentation may be the first step in the anaerobic process. Hydrolysis is carried out with extracellular enzymes produced by a variety of facultative and obligate anaerobes (Confer and Logan, 1998; Song et al., 2005). Lipids are broken down to long chain fatty acids (LCFAs) by lipases produced by bacteria that include *Butyrivibrio* sp., *Clostridium* sp., and *Anaerovibrio lipolytica.* Peptide and amino acid are due to bacteria exhibiting extracellular protease activity including *Clostridium proteolyticum*, *Eubacterium* sp., and *Peptococcus anaerobicus (*McInerney, 1988).

Acidogenesis. The second basic step, which is done by bacteria, is *acidogenesis* (also termed *fermentation*) and results in the production of VFAs, CO_2, and hydrogen as shown on Fig. 7–27. In the fermentation process, substrates serve as both the electron donors and acceptors. The principal fermentation products from the sugars and amino acids are acetate, propionate, butyrate, CO_2, and hydrogen. Fermentation of the LCFAs results in the production of acetate, CO_2, and hydrogen. A larger fraction of the LCFA COD is converted to hydrogen than that for the sugars and amino acids.

Acetogenesis. Acetogenesis refers to further fermentation by bacteria to convert intermediate products of acidogenesis (propionate and butyrate) to also produce acetate, CO_2, and hydrogen. Thus, the final products of fermentation are acetate, hydrogen, and CO_2, which are the precursors of methane formation. The free energy change associated with the conversion of propionate and butyrate to acetate and hydrogen requires that hydrogen be at low concentrations in the system ($H_2 < 10^{-4}$ atm), or the reaction will not proceed (McCarty and Smith, 1986). Most of the hydrogen produced comes from the oxidation of LCFAs and intermediate VFAs to acetic acid and is referred to as anaerobic oxidation.

Methanogenesis. The third basic step, *methanogenesis*, is carried out by a group of *Archaea* organisms known collectively as methanogens. Two groups of methanogenic organisms are involved in methane production. One group, termed *aceticlastic methanogens*, split acetate into methane and carbon dioxide. The second group, termed hydrogen-utilizing methanogens or *hydrogenotrophic methanogens*, use hydrogen as the electron donor and CO_2 as the electron acceptor to produce methane. Bacteria within anaerobic processes, termed *acetogens*, are also able to use CO_2 to oxidize hydrogen and form acetic acid. However, the acetic acid will be converted to methane, so the impact of this reaction is minor. As shown on Fig. 7–27, about 72 percent of the methane produced in anaerobic digestion is from acetate formation. The composition of the gas produced from a stable fermentation and methanogenesis operation typically contains about 65 percent methane and 35 percent CO_2. A higher lipid fraction in the waste results in a higher methane fraction in the digester gas (Li et al., 2002).

Microbiology

The group of nonmethanogenic microorganisms responsible for hydrolysis and fermentation consists of a diverse group of facultative and obligate anaerobic bacteria. Organisms isolated from anaerobic digesters include *Clostridium* spp., *Peptococcus anaerobus*, *Bifidobacterium* spp., *Desulphovibrio* spp., *Corynebacterium* spp., *Lactobacillus*, *Actinomyces*, *Staphylococcus*, and *Escherichia coli*. Other physiological groups present include those producing proteolytic, lipolytic, ureolytic, or cellulytic enzymes.

The microorganisms responsible for methane production, classified as archaea, are strict obligate anaerobes. Many of the methanogenic organisms identified in anaerobic digesters are similar to those found in the stomachs of ruminant animals and in organic sediments taken from lakes and rivers. The hydrogenotrophic methanogens are found within four orders of the Archaea: *Methanobacteriales*, *Methanococcales*, *Methanomicrobiales,* and *Methanopyrales* (Madigan et al., 2012). They obtain their energy from the oxidation of hydrogen and also use CO_2 for their carbon sources, resulting in low biomass synthesis yields.

$$4H_2 + CO_2 \rightarrow CH_4 + 2H_2O \tag{7–137}$$

The archaea that produce methane from acetate are referred to as aceticlastic methanogens and are in the order *Methanosarcinales*. The acetate is cleaved, and the methyl carbon ends up in methane and the carboxyl carbon in CO_2.

$$CH_3COOH \rightarrow CH_4 + CO_2 \tag{7–138}$$

Only two genera within the order *Methanosarcinales* are able to use acetate to produce methane and carbon dioxide: *Methanosarcina* and *Methanosaeta* (formerly named *Methanothrix*) (Madigan et al., 2012). These organisms differ in both their morphology and kinetic characteristics. The *Methanosarcina* have a coccoid shape and appear in grape-like clumps, while the *Methanosaeta* have long rod to filament shapes (Lange and Ahring, 2001). The *Methanosarcina* have a high maximum specific growth rate ($\mu_{\max}$) and high half-saturation coefficient (K_S) and the *Methanosaeta* have a low $\mu_{\max}$ and low $K_{S\text{ values.}}$ The *Methanosaeta* are dominant in anaerobic digesters due to the long SRT and low acetate concentration. However, *Methanosarcina* handles increases in acetate more efficiently and thus promotes more stable digestion. Laboratory anaerobic digester investigations and digester modeling studies showed that a *Methanosarcina* population could be established with daily batch feeding versus frequent feeding intervals (Conklin et al., 2006 and Straub et al., 2006). Values for K_s for acetate utilization by enrichments of *Methanosaeta* and *Methanosarcina* are 90 g/m^3 and 320 g/m^3, respectively. Representative $\mu_{\max}$ values at 35°C are 0.16 g/g·d and 0.80 g/g·d for *Methanosaeta* and *Methanosarcina*, respectively (Conklin et al., 2006).

The acetate-utilizing methanogens were also observed in thermophilic reactors (van Lier, 1996; Zinder and Koch, 1984; and Ahring, 1994). Some species of *Methanosarcina* were inhibited by temperature at 65°C, while others were not, but no inhibition of *Methanosaeta* was shown. For hydrogen-utilizing methanogens at temperatures above 60°C, *Methanobacterium* was found to be very abundant.

Other anaerobic reactions by methanogenic archaea are formate degradation (Eq. 7–139) by organisms in the orders *Methanobacteriales*, *Methanomicrobiales*, and *Methanococcales*, methanol degradation Eq. (7–140) by organisms in the orders *Methanobacteriales* and *Methanosarcinales*, and methylamines degradation (Eq. 7–141) by organisms in the order *Methanosarcinales* (Madigan et al., 2012).

$$4HCOO^- + 4H^+ \rightarrow CH_4 + 3CO_2 + 2H_2O \tag{7–139}$$

$$4CH_3OH \rightarrow 3CH_4 + CO_2 + 2H_2O \tag{7–140}$$

$$4(CH_3)_3N + H_2O \rightarrow 9CH_4 + 3CO_2 + 6H_2O + 4NH_3 \tag{7–141}$$

Syntrophic Relationships in Fermentation. The methanogens and the acidogens form a syntrophic (mutually beneficial) relationship in which the methanogens convert fermentation end products such as hydrogen, formate, and acetate to methane and carbon dioxide. Because the methanogens are able to maintain an extremely low partial pressure of H_2, the equilibrium of the fermentation reactions is shifted toward the formation of more oxidized end products (e.g., formate and acetate). The utilization of the hydrogen produced by the acidogens and other anaerobes by the methanogens is termed *interspecies hydrogen transfer*. In effect, the methanogenic organisms serve as a hydrogen sink that allows the fermentation reactions to proceed. If process upsets occur and the methanogenic organisms do not utilize the hydrogen produced fast enough, the propionate and butyrate fermentation will be slowed with the accumulation of volatile fatty acids in the anaerobic reactor and a possible reduction in pH.

Nuisance Organisms. Nuisance organisms in anaerobic operations are the sulfate-reducing bacteria, which can be a problem when the wastewater contains significant concentrations of sulfate. These organisms can reduce sulfate to sulfide, which can be toxic to methanogenic bacteria at high enough concentrations. Where high sulfide concentrations occur, one solution is to add iron at controlled amounts to form iron sulfide precipitate. Sulfate-reducing bacteria, obligate anaerobes of the domain Bacteria, are morphologically diverse but share the common characteristic of being able to use sulfate as an electron acceptor and are divided into one of two groups, depending on whether they produce fatty acids or use acetate. Group I sulfate reducers can use a diverse array of organic compounds as their electron donor, oxidizing them to acetate and reducing sulfate to sulfide. A common genus found in anaerobic biochemical operations is *Desulfovibrio*. Group II sulfate reducers oxidize fatty acids, particularly acetate, to carbon dioxide, while reducing sulfate to sulfide. A bacteria commonly found in this group is in the genus *Desulfobacter*.

Stoichiometry of Anaerobic Fermentation and Oxidation

A COD balance can be used to account for the changes in COD during fermentation and oxidation. Instead of oxygen accounting for the change in COD, the COD loss in the anaerobic reactor is accounted for by the methane production. By stoichiometry the COD equivalent of methane can be determined. The COD of methane is the amount of oxygen needed to oxidize methane to carbon dioxide and water.

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O \tag{7–142}$$

From the above, the COD per mole of methane is 2(32 g O_2/mole) = 64 g O_2/mole CH_4. The volume of methane per mole at standard conditions (0°C and 1 atm) is 22.414 L, so the CH_4 equivalent of COD converted under anaerobic conditions is 22.414/64 = 0.35 L CH_4/g COD.

EXAMPLE 7–10 **Prediction of Methane Gas Production** An anaerobic reactor, operated at 35°C, is used to process a wastewater stream with a flow of 3000 m^3/d and a bCOD concentration of 5000 g/m^3. At 95 percent bCOD removal and a net biomass synthesis yield of 0.04 g VSS/g COD used, what is the amount of methane produced in m^3/d?

Solution

1. Prepare a steady-state mass balance for COD to determine the amount of the influent COD converted to methane.
   a. The required steady-state mass balance is

$$0 = \begin{array}{c}\text{Influent}\\ \text{COD}\end{array} - \begin{array}{c}\text{portion of}\\ \text{influent COD}\\ \text{in effluent}\end{array} - \begin{array}{c}\text{influent COD}\\ \text{converted to}\\ \text{cell tissue}\end{array} - \begin{array}{c}\text{influent COD}\\ \text{converted to}\\ \text{methane}\end{array}$$

$$COD_{in} = COD_{eff} + COD_{VSS} + COD_{methane}$$

b. Determine the values of the individual mass balance terms

$$\text{COD}_{\text{in}} = (5000\ \text{g/m}^3)(3000\ \text{m}^3/\text{d}) = 15{,}000{,}000\ \text{g/d}$$

$$\text{COD}_{\text{eff}} = (1 - 0.95)(5000\ \text{g/m}^3)(3000\ \text{m}^3/\text{d}) = 750{,}000\ \text{g/d}$$

$$\begin{aligned}\text{COD}_{\text{VSS}} &= (1.42\ \text{g COD/g VSS})(0.04\ \text{g VSS/g COD})(0.95)(15{,}000{,}000\ \text{g/d})\\ &= 809{,}400\ \text{g/d}\end{aligned}$$

c. Solve for the COD converted to methane

$$\text{COD}_{\text{methane}} = 15{,}000{,}000 - 750{,}000 - 809{,}400 = 13{,}440{,}600\ \text{g/d}$$

2. Determine the amount of methane produced at 35°C.

a. Determine the volume of gas occupied by 1 mole of gas at 35°C

$$V = \frac{nRT}{P}$$

$$V = \frac{(1\ \text{mole})(0.082057\ \text{atm}\cdot\text{L/mole}\cdot\text{K})[(273.15)K]}{1.0\ \text{atm}}$$

$$= 25.29\text{L}$$

b. The CH_4 equivalent of COD converted under anaerobic conditions is

$$(25.29\ \text{L/mole})/(64\ \text{g COD/mole CH}_4) = 0.40\ \text{L CH}_4/\text{g COD}.$$

c. Methane production

$$\begin{aligned}\text{CH}_4\ \text{production} &= (13{,}440{,}600\ \text{g COD/d})(0.40\ \text{L CH}_4/\text{g COD})(1\ \text{m}^3/10^3\ \text{L})\\ &= 5376\ \text{m}^3/\text{d}\end{aligned}$$

$$\begin{aligned}\text{At 65 percent methane the total gas flowrate} &= (5376\ \text{m}^3/\text{d})/0.65\\ &= 8271\ \text{m}^3/\text{d}\end{aligned}$$

Comment It is important to determine the volume occupied by the gas at the actual operating temperature.

Process Kinetics

Because of the relatively low free energy change for anaerobic reactions, growth yield coefficients are considerably lower than the corresponding values for aerobic oxidation. Typical synthesis yield and endogenous decay coefficients for fermentation and methanogenic anaerobic reactions are $Y = 0.06$ and 0.03 g VSS/g COD and $b = 0.02$ and 0.008 g VSS/g VSS·d, respectively.

In anaerobic processes two rate-limiting concepts are important: (1) the hydrolysis conversion rate and (2) the soluble substrate utilization rate for fermentation and methanogenesis. The flow of COD through acidogenesis and to methane production starts with hydrolysis of colloidal and solid particles.

Hydrolysis Conversion Limitations. For anaerobic digestion processes used for municipal waste sludges solids, hydrolysis rates and the reactor SRT determine the total amount of solids converted. A first order hydrolysis rate of 0.33 d^{-1} may be applied to the biodegradable portion of combined waste primary and waste activated sludge from domestic wastewater treatment to estimate the solids destruction efficiency of a mesophilic digester (35°C) as a function of the SRT. About 25 percent of the waste primary solids and waste activated sludge solids for domestic wastewater treatment are estimated as nonbiodegradable (Moen et al., 2004). An SRT greater than 30 d is needed to approach full conversion of degradable solids. The soluble substrate utilization kinetics by the bacteria responsible for fermentation and methanogenesis are of great concern to maintain a stable anaerobic process.

Production Versus Utilization. Production kinetics for VFAs are faster than the corresponding utilization and methane production kinetics for methanogens, but at steady state operation a sufficient methanogenic population is established to maintain a stable digester operation with a low VFA concentration (<200 g/m^3) and pH ≥ 7.0. Unstable digester operation can develop under transient loading conditions in which the VFA production rate exceeds the methanogenic VFA utilization rate. Subsequently the VFA concentration (typically acetate and propionate) increases, which may cause the pH to drop, depending on the amount of alkalinity available to buffer the organic acid concentration increase.

Effects of pH. The methanogenic VFA utilization kinetics decrease at lower pH, so the effect of transient loads that cause a pH drop can be compounded to lead to further accumulation of VFA and further decline in methanogenic activity. If this reduced utilization continues butyric acid also accumulates and the reactor operation is greatly inhibited. Butyric acid accumulation causes a sour odor and has led to the term of a "sour" or stuck digester to describe this extreme digestion upset condition. Methanogenic inhibition can occur when acetate concentration has exceeded 3000 g/m^3, even though there was sufficient alkalinity to maintain pH above 7.0 (Stallman et al., 2012). Unstable conditions can also be caused by anaerobic reactor upset by transient temperature changes and sufficient concentration of inhibiting substances.

Transient Capacity. Anaerobic reactors have an inherent capacity for a finite transient increase in feed COD without succumbing to an unstable condition. The excess capacity is related to the maximum acetate utilization capacity ($V_{\max}$) for the acetoclastic methanogens (Conklin et al., 2008). The $V_{\max}$ value varies with systems and is a function of the SRT, average COD loading rate history, and methanogenic acetoclastic population selected. Batch biochemical methane production (BMP) tests are done to determine $V_{\max}$ for a given system. The test involves spiking acetate to the anaerobic reactor sludge in closed serum bottles, followed by incubation at the reactor temperature and measuring CH_4 production rates over time. An acetate capacity number (ACN) is then determined as the ratio of $V_{\max}$ to the average estimated acetate production rate based on the reactor operating history. About 70 percent of the COD removed in the system is assumed to be converted to acetate prior to methane production.

$$\text{ACN} = \frac{V_{\max}}{V_{\text{plt}}} \tag{7–143}$$

where ACN = acetate capacity number,
$V_{\max}$ = maximum acetate utilization rate for the anaerobic reactor solids, g acetate COD/m^3·d
V_{plt} = average daily acetate utilization rate for the anaerobic reactor, g acetate COD/m^3·d

The ACN concept can be applied to anaerobic reactor operations to determine acceptable transient loading rates that can be used without causing an unstable digester condition as shown in the following example. Details on the BMP test procedure can be found in Conklin et al. (2008). An application of the ACN concept to evaluate an acceptable transient loading in an anaerobic reactor is illustrated in Example 7–11.

EXAMPLE 7–11 **Estimating the Acceptable Transient Load for Codigestion** Given the following anaerobic sludge digestion operating information and results from a BMP test for a high strength food waste to be added to the digester for codigestion, determine what volume of food waste can be added as a transient load without causing digester instability.

Parameter	Unit	Value
Digester average feed rate	m^3/d	1000
Digester SRT	d	20
Digester average feed COD	g/m^3	85,000
Digester average CH_4 production rate	m^3/d	16,000
BMP average acetate V_{max} at standard conditions	mL CH_4/mL·d	0.5
Codigest COD concentration	g/m^3	800,000
Codigest biodegradability fraction	%	90

1. Portion of methane produced from acetate utilization for anaerobic digester and codigest waste = 70 percent
2. Gas production at standard conditions = 0.35 m^3 CH_4/kg COD

Solution

1. Determine the digester ACN.
   a. Digester average acetate COD utilization rate, V_{plt},

   V_{plt} = kg acetate COD used/m^3·d

   Acetate COD used per day =

$$0.70(16{,}000 \text{ m}^3/\text{d CH}_4)\left(\frac{273}{273 + 35}\right)\left[\frac{1}{(0.35 \text{ m}^3 \text{ CH}_4/\text{kg COD})}\right] = 28{,}363 \text{ kg COD/d}$$

   b. Digester volume = $Q(\tau)$, τ = SRT = 20 d

   Volume = (1000 m^3/d)(20 d) = 20,000 m^3

   V_{plt} = (28,363 kg COD/d)/(20,000 m^3) = 1.41 kg acetate COD/m^3·d

   c. BMP test digester sludge acetate utilization rate, V_{max}, g COD/L·d

$$V_{max} = \left(\frac{0.65 \text{ m}^3 \text{ CH}_4}{\text{m}^3 \cdot \text{d}}\right)\left[\frac{1}{(0.35 \text{ m}^3 \text{ CH}_4/\text{kg COD})}\right] = 1.86 \text{ kg acetate COD/m}^3\cdot\text{d}$$

   d. ACN = V_{max}/V_{plt} = 1.86/1.41 = 1.32

   Thus, the digester methanogens have capacity for 32% more acetate.

2. Determine the codigest volume that can be added.
   a. Additional acetate COD loading = 0.32(28,363 kg COD/d)

   = 11,954 kg acetate COD/d

   b. Acetate COD available in codigest feed

   = (800,000 g COD/m^3)(0.90 g degrad./g COD)(0.70 g acetate COD/g COD)

   = 504,000 g acetate COD/m^3

c. Codigest volume

$$= \frac{(11{,}954 \text{ kg acetate COD/d})(10^3 \text{ g/1 kg})}{(504{,}000 \text{ g acetate COD/m}^3)} = 23.72 \text{ m}^3\text{/d}$$

Comment For this high strength codigest feed, the feed volume to the digester is increased by only $(23.72 \text{ m}^3\text{/d})100/(1000 \text{ m}^3\text{/d}) = 2.4$ percent, which has a minimal impact on the digester SRT and methanogenic population. In this problem, it was also assumed that there is no acclimation time for degradation of codigest feed.

Soluble Substrate Utilization Limitations. For soluble feeds or after solids hydrolysis, the rate-limiting step is the conversion of VFAs by the methanogenic organisms and not the fermentation of soluble substrates by the fermenting bacteria. Thus, methanogenic growth kinetics are of most interest in anaerobic process designs. Appropriate system SRTs are selected based on kinetics and treatment goals. At 20, 25, and 35°C, the washout or SRT_{min} values for methanogenesis are 7.8, 5.9, and 3.2 d, respectively (Lawrence and McCarty, 1970). Thus, with a factor of safety of 5, design SRT values would be about 40, 30, and 15 d, respectively, for a suspended growth process. Safety factors higher than 5 have been used to provide a more stable process (Parker and Owen, 1986).

Environmental Factors

Anaerobic processes are sensitive to pH , nutrient limitations (especially industrial wastewaters) and inhibitory substances. A pH value near neutral is preferred, and below 6.8 the methanogenic activity is inhibited. Because of the high CO_2 content in the gases developed in anaerobic processes (30 to 35 percent CO_2), a high alkalinity is needed to assure pH near neutrality. An alkalinity concentration in the range of 3000 to 5000 mg/L as $CaCO_3$ is often found. For sludge digestion sufficient alkalinity is produced by the breakdown of protein and amino acids to produce NH_3, which combines with CO_2 and H_2O to form alkalinity as $NH_4(HCO_3)$. For industrial wastewater applications, especially for waste containing mainly carbohydrates, it is necessary to add alkalinity for pH control. Substances inhibitory to anaerobic processes (e.g., NH_3, H_2S, and various other inorganic and organic compounds) are considered in Chap. 10.

7–15 BIOLOGICAL REMOVAL OF TOXIC AND RECALCITRANT ORGANIC COMPOUNDS

Most of the organic compounds in domestic wastewater and some in industrial wastewaters are of natural origin and can be degraded by common bacteria in aerobic or anaerobic processes. However, currently there are over 70,000 synthetic organic chemicals, termed *xenobiotic compounds*, in general use (Schwarzenbach et al., 2003). Unfortunately, some of these organic compounds pose unique problems in wastewater treatment, due to their resistance to biodegradation and potential toxicity to the environment and human health. Organic compounds that are difficult to treat in conventional biological treatment processes are termed *refractory*. In addition, there are naturally occurring substances, such as those found in petroleum products that are of similar concern. Examples of petroleum compounds and synthetic organic compounds found in different types of wastewater are reported in Table 7–14.

Table 7–14

Examples of toxic and recalcitrant organic compounds found in wastewater[a]

Type of waste	Types of organic compounds
Petroleum	Alkanes, alkenes, polyaromatic hydrocarbons, monocyclic aromatics-benzene, toluene, ethylbenzene, xylenes, naphthenes
Non-halogenated solvents	Alcohols, ketones, esters, ethers, aromatic and aliphatic hydrocarbons, glycols, amines
Halogenated solvents	Chlorinated methanes-methylene chloride, chloroform, carbon tetrachloride, chlorinated ethenes-tetrachloroethene, trichloroethene, chlorinated ethanes-trichlorethane, chlorinated benzenes
Insecticides, herbicides, fungicides	Organochloride compounds, organophosphate cmpds, carbamate esters, phenyl ethers, creosotes, chlorinated phenols
Munitions and explosives	Nitroaromatics-trinitrotolune, nitramines, nitrate esters
Industrial intermediates	Phthalate esters, benzene, phenol, chlorobenzenes, chlorophenols, xylenes,
Transformer and hydraulic fluids	Polychlorinated biphenyls
Production products	Dioxin, furans

[a] Adapted from Watts (1997).

Development of Biological Treatment Methods

Since the early 1970s, information and knowledge related to the biodegradation of toxic and refractory compounds has increased significantly, based on work with specific industrial wastewaters (i.e., petrochemical, textile, pesticide, pulp and paper, and pharmaceutical industries). In addition, since the 1980s, significant progress has also been made on the biodegradation of organic substances found at hazardous waste sites. Work in both of these fields has expanded knowledge on the capabilities and limitations of biodegradation. With a few exceptions most organic compounds can be biodegraded eventually, but in some cases the rates may be slow, unique environmental conditions may be required (i.e., redox potential, pH, temperature), fungi may be needed instead of prokaryotes, or specific bacteria capable of degrading the xenobiotic compounds may be needed. For example, anaerobic degradation of polychlorinated biphenyls (PCB) occurred using bacteria seed from sediment in the Hudson River where PCB had accumulated over decades, but after 1.5 years of exposure in a laboratory anaerobic digester used to treat municipal wastewater plant sludge, bacteria could not be developed to degrade PCB (Ballapragada et al., 1998).

Importance of Specific Microorganisms. The ability to degrade toxic and recalcitrant compounds will depend primarily on the presence of appropriate microorganism(s) and acclimation time. In some cases, special seed sources are needed to provide the necessary microorganisms. Once the critical microorganism is present, long-term exposure to the organic compound may be needed to induce and sustain the enzymes and bacteria required for degradation. Acclimation times can vary from hours to weeks depending on the microorganism population and organic compound. Melcer et al. (1994) found that a period of 3 weeks was required before complete removal of dichlorobenzene (DCB) occurred in a municipal activated sludge plant, and that intermittent addition of DCB resulted in much lower treatment efficiencies. Without acclimation and no biodegradation, the DCB was removed from the activated sludge aeration tank by volatilization as described in Chap. 16. Strand et al. (1999) found that after 4 weeks of constant exposure to dinitrophenol in a laboratory activated sludge process, seeded from a municipal wastewater plant, dinitrophenol degradation increased from 0 to 98 percent. When dinitrophenol was not added to the process, the ability to degrade dinitrophenol was eventually lost. Thus, it appears that a relatively constant supply of toxic and recalcitrant organic compounds can lead to better biodegradation performance than intermittent additions.

Biodegradation Pathways. The three principal types of degradation pathways that have been observed are (1) the compound serves as a growth substrate; (2) the organic compound provides an electron acceptor; and (3) the organic compound is degraded by cometabolic degradation. In cometabolic degradation, the compound that is degraded is not part of the microorganism's metabolism. Degradation of the compound is brought about by a nonspecific enzyme and provides no benefit to the cell growth. Complete biodegradation of toxic and recalcitrant organic compounds to harmless end products such as CO_2 and H_2O or methane may not always occur, and instead biotransformation to a different organic compound is possible. Many toxic and recalcitrant organic compounds are degraded under anaerobic conditions, with the compound serving as a growth substrate with fermentation and ultimately methane production. Typical examples include nonhalogenated aromatic and aliphatic compounds such as phenol, toluene, alcohols, and ketones. However, most chlorinated organic compounds are not attacked easily under anaerobic conditions and do not serve as growth substrates. Fortuitously, many of these compounds also serve as electron acceptors in anaerobic oxidation-reduction reactions. Most of the work and application for anaerobic degradation of chlorinated organic compounds have been related to subsurface contamination of chlorinated solvents at hazardous waste sites (McCarty, 1999).

Examples of chlorinated compounds degraded under anaerobic conditions include tetrachloroethene (perchloroethylene, PCE), trichloroethene, carbon tetrachloride, trichlorobenzene, pentachlorophenol, chlorohydrocarbons, and PCBs. The chlorinated compound serves as the electron acceptor, and hydrogen produced from fermentation reactions provides the main electron donor. Hydrogen replaces chlorine in the molecule, and such reactions under anaerobic conditions have generally been referred to as *reductive dechlorination* or a *dehalorespiration process*. For example, dechlorination of tetrachloroethene proceeds sequentially with a loss of chlorine in each step via trichloroethene to dichloroethene to vinyl chloride and finally to ethene. A number of bacteria capable of anaerobic reductive dechlorination chlorinated ethenes have been isolated, but only the eubacterium, *Dehaloccoides ethenogenes*, has been able to dechlorinate tetrachloroethene to ethene completely. Hydrogen is the only electron donor used by *Dehaloccoides ethenogenes*, but some dehalorespiring bacteria have been able to use formate, pyruvate or acetate (Holliger et al., 1999).

As the number of chlorine molecules on the organic molecule decreases, the reactions tend to be slower and less complete. Dechlorination of tetrachloroethene, trichlorobenzene, and pentachlorophenol has been demonstrated in lab-scale anaerobic digesters (Ballapragada et al., 1998) treating municipal primary and secondary sludge. However, the reaction rates were slow with mono- and dichlorophenol and mono- I dichlorobenzenes remaining. Conversion of tetrachloroethene to vinyl chloride and ethene occurred in the digesters after one year of acclimatization and constant exposure of the chloroethenes.

Aerobic Biodegradation

With proper environmental conditions, seed source, and acclimation time, a wide range of toxic and recalcitrant organic compounds have been found to serve as growth substrates for heterotrophic bacteria. Such compounds include phenol, benzene, toluene, polyaromatic hydrocarbons, pesticides, gasoline, alcohols, ketones, methylene chloride, vinyl chloride, munitions compounds, and chlorinated phenols. However, many chlorinated organic compounds cannot be attacked readily by aerobic heterotrophic bacteria and, thus, do not serve as growth substrates. Some of the lesser chlorinated compounds, such as dichloromethane, 1,2-dichloroethane, and vinyl chloride can be used as growth substrates by aerobic bacteria. Fortunately, a number of chlorinated organic compounds are degradable by cometabolic degradation. It should be noted that organic compounds that are saturated fully with chlorine are degraded only by anaerobic dechlorination (Stensel and Bielefeldt, 1997).

Cometabolic Degredation. Chlorinated organic compounds that have been degraded by cometabolic degradation include trichloroethene, dichloroethene, vinyl chloride, chloroform, dichloromethane, and trichloroethane. Cometabolic degradation is possible by bacteria that produce nonspecific mono-oxygenase or dioxygenase enzymes. These enzymes mediate a reaction with oxygen and hydrogen and change the structure of the chlorinated compound. Bacteria that produce oxygenase enzymes oxidize certain substrates that induce the enzyme. Oxygenase-producing bacteria include methanotrophic bacteria that oxidize methane, a number of bacteria that can oxidize phenol or toluene, a number of bacteria that can oxidize propane, and nitrifying bacteria that oxidize ammonia to nitrite.

Nonspecific Oxygenase Enzyme. The reaction of the nonspecific oxygenase enzyme with the organic chlorinated compound typically produces an intermediate compound that is degraded by other aerobic heterotrophic bacteria in the biological consortia. Various reactor designs have been developed to apply this biological process for treatment of contaminated groundwater or vapor extraction gas streams (Lee et al., 2000). While such reactions are possible in municipal and industrial biological wastewater treatment processes, a large amount of the chlorinated organic compounds that may be present are more likely lost from the process by volatilization during aeration, because of their high volatility and the minimal potential for cometabolic bacteria to be present.

Abiotic Losses

Due to concerns about environmental and health effects of toxic and recalcitrant compounds, it is important to understand their fate and transport in biological treatment processes. For many toxic and recalcitrant organic compounds entering biological wastewater treatment processes, nonbiological or abiotic losses may be more significant than biodegradation. Abiotic losses include adsorption of the compound to the mixed-liquor solids in the reactor with subsequent transport out of the system by the waste sludge and volatilization with release of the compound to the surrounding atmosphere.

Losses Due to Adsorption. For certain compounds, removal by partitioning (i.e., adsorbing) onto the biomass can be more significant than biodegradation or volatilization. To describe solids partitioning, the Freundlich Isotherm model (see Sec. 11–7 in Chap. 11) is modified to a general linear equilibrium relationship ($n = 1$) for adsorption to solids at relatively low liquid organic concentrations:

$$q = K_p S \tag{7–144}$$

where q = g organic adsorbed/g adsorbent
K_p = partition coefficient, L/g
S = concentration of organic compound in liquid, g/L

Adsorption of organic compounds in biological treatment processes has been observed to be relatively fast (Melcer et al., 1994), so that Eq. (7–144) can be used to describe the distribution of the compound between the solid biomass and liquid phases as a function of the partition coefficient K_p for the compound. The value for K_p depends on the hydrophobic nature of the compound and the adsorption characteristics of the solids. Solids with high carbon content and greater surface area result in higher K_p values. It is common to consider the partition coefficient as a function of the carbon content of the solids instead of the solids weight. In this case the partition coefficient is given as K_{oc}:

$$K_{oc} = \frac{K_p}{f_{oc}} \tag{7–145}$$

where K_{oc} = partition coefficient relative to solids carbon concentration, L/kg
f_{oc} = fraction of carbon in solids, g carbon/g solids

More hydrophobic compounds are expected to partition more to the solid phase and thus there had been interest to relate the partition coefficient of a compound for solids in wastewater treatment to the octanol-water partition coefficient (Dobbs et al., 1989), which are available for many compounds.

$$\log K_{oc} = \text{A} \log K_{ow} + B \tag{7–146}$$

where K_{ow} = octanol/water partition coefficient, dimensionless
A, B = empirical coefficients, dimensionless

Values for coefficients in Eq. (146) can be found in Schwarzenbach et al. (2003) for many compounds found in wastewater treatment. Octanol-water partition coefficient values have been developed for many organic compounds (Schwarzenbach et al., 2003; LaGrega et al., 2001) and values for some selected compounds are given in Table 16–12 in Chap. 16. The test to determine K_{ow} involves measuring the concentration of the organic compound in an octanol/water mixture after quiescent separation of the octanol layer above the water layer. Greater amounts of more hydrophobic compounds will be found in the octanol layer, and these compounds will have greater K_p values in solids/water mixtures. Ranges of K_p values for various types of compounds are shown in Table 7–15. Based on the information given in Table 7–15, benzopyrene and PCBs are more likely to be found on solids than in the liquid than compounds like benzene and trichloroethene, as their K_p values are greater by a factor of about 150. Such high partition coefficients result in very low liquid concentrations, which minimize the amount of the organic compound lost by biodegradation and volatilization.

Table 7–15

Comparison of selected estimated partition coefficients (K_p) values for different types of organic compounds

Organic compound	K_p, (L/g)
Benzene	0.23
Dinitrotoluene	0.29
Dieldrin	0.81
Phenanthrene	5.33
Pentachlorophenol	10.96
Polychlorinated biphenyl	43.87
Benzopyrene	45.15

Using the equilibrium partition coefficient, the amount of organic compound removed by sludge wasting can be estimated from the ratio of the mass of organic compound adsorbed to the mass of solids wasted per day:

$$q = \frac{r_{ad}}{r_{X,w}} \tag{7–147}$$

where q = g organic compound adsorbed/g solids
r_{ad} = rate of organic compound absorbed daily, g/d
$r_{X,w}$ = rate of solids wasted daily, g/d

By substituting Eq. (7–144) for q and solving for r_{ad}, the amount of compound lost daily due to adsorption is

$$r_{ad} = r_{X,w}K_pS \tag{7–148}$$

At steady state, the amount of solids wasted daily is related to the average SRT value for the activated sludge system as given by Eq. (7–48):

$$r_{X,w} = \frac{X_TV}{\text{SRT}} \tag{7–149}$$

Substituting Eq. (7–149) into Eq. (7–148) yields the following expression for the mass loss rate due to adsorption:

$$r_{ad} = \frac{X_TVK_pS}{\text{SRT}} \tag{7–150}$$

Losses Due to Volatilization The removal of volatile organic compounds due to aeration (volatilization), discussed in detail in Sec. 16–4 in Chap. 16, is reviewed here briefly. The rate of loss due to volatilization can be modeled as follows:

$$r_{sv} = -K_La_s(S) \tag{7–151}$$

where r_{sv} = loss due to volatilization, mg/L·d
K_La_s = gas-liquid mass transfer coefficient, K_La, of organic compound, d^{-1}
S = concentration of organic compound in liquid, mg/L

The above equation is satisfactory to represent the stripping loss by aeration if a surface aerator is used, in which case the air content of the volatile compound is considered negligible. However, for diffused aeration the volatile compound would be captured by the rising bubbles with concentration changes with depth. For this case the following mass transfer expression applies (Bielefeldt and Stensel, 1999).

$$r_{sv} = Q_gS_{g,\text{VOC}} = Q_g(H)S_{L,\text{VOC}}\left\{1 - \exp\left[\frac{(\alpha K_La_{,\text{VOC}}V)}{Q_g(H)}\right]\right\} \tag{7–152}$$

where Q_g = gas flowrate through reactor, m^3/d
$S_{g,\text{VOC}}$ = VOC content in the gas leaving the reactor, g/m^3
H = Henry's constant of the VOC at the reactor temperature, $\text{L}_{\text{water}}/\text{L}_{\text{air}}$
$S_{L,\text{VOC}}$ = liquid concentration of the VOC, g/m^3
$K_La_{,\text{VOC}}$ = VOC mass transfer coefficient, d^{-1}
α = ratio of mass transfer in reactor mixed liquor to that in clean water
V = reactor volume, m^3

The gas-liquid mass transfer coefficient for the VOC can be estimated from the mass transfer coefficient for oxygen for the aeration system and the ratio of the VOC and oxygen diffusion coefficients:

$$K_{LA,\mathrm{VOC}} = \left(\frac{D_{\mathrm{VOC}}}{D_{O_2}}\right)^n \tag{7–153}$$

where D = the diffusion coefficient, $\mathrm{m^2/s}$

n = coefficient from 0.50 to 1.0 as function of aeration system

Modeling Biotic and Abiotic Losses

A number of models have been developed and evaluated that account for the fate of recalcitrant organic compounds in biological treatment processes (Melcer et al., 1995; Melcer et al., 1994; Monteith et al., 1995; Parker et al., 1993; Grady et al., 1997; and Lee et al., 1998a). In general, the models contain basic mechanisms and mass balances that account for the mass rate of the organic compound entering the treatment process, and leaving in the liquid effluent by biodegradation, by volatilization, and by adsorption on waste solids.

In the following discussion, all of these mechanisms as discussed above are combined into a general model to describe the fate of specific compounds in a biological treatment process. The following steady-state mass balance (i.e., accumulation = 0) across a complete-mix activated sludge process can be prepared to predict the fate of an organic compound subject to biotic and abiotic processes.

$$0 = \begin{array}{c}\text{organic}\\ \text{constituent}\\ \text{in influent}\end{array} - \begin{array}{c}\text{loss of organic}\\ \text{constituent due}\\ \text{to biodegradation}\end{array} - \begin{array}{c}\text{loss of organic}\\ \text{constituent due}\\ \text{to sorption}\end{array} - \begin{array}{c}\text{loss of organic}\\ \text{constituent due}\\ \text{to volatilization}\end{array} - \begin{array}{c}\text{loss of organic}\\ \text{constituent}\\ \text{in effluent}\end{array}$$

$$QS_o = r_{su} + r_{ad} + r_{sv} + QS \tag{7–154}$$

where QS_o = mass of compound in wastewater influent, g/d

r_{su} = biodegradation rate, g/d

r_{ad} = solids adsorption rate, g/d

r_{sv} = volatilization rate, g/d

QS = mass of compound in wastewater effluent, g/d

Substituting the appropriate reaction terms for each component of the mass balance yields the following expression:

$$QS_o = \left(\frac{1}{Y}\right)\frac{\mu_m(S)}{(K_s + S)}(X_s)(V) + \frac{X_T V K_p S}{\mathrm{SRT}} + K_L a_s S V + QS \tag{7–155}$$

The fate of the compound in the influent wastewater as a function of the solids concentration, liquid concentration, hydraulic retention time, τ, SRT, and rate terms is obtained by dividing Eq. (7–155) by Q:

$$S_o = \left(\frac{1}{Y}\right)\frac{\mu_m(S)}{(K_s + S)}(X_s)\tau + K_p S X_T\left(\frac{\tau}{\mathrm{SRT}}\right) + K_L a_s S(\tau) + S \tag{7–156}$$

Note in Eqs. (7–155) and (7–156), X_s is the biomass concentration capable of degrading the specific organic compound, and X_T is the total MLVSS concentration that includes all the biomass grown on various substrates plus the nonbiodegradable VSS. The value of X_s can be calculated as a function of the amount of substrate that is biodegraded and kinetic coefficients, and the system τ, and SRT. The following expression is used to calculate X_s

at steady state for a complete-mixed reactor with consideration for losses of the substrate by volatilization and adsorption:

$$X_s = \frac{Y[(S_o - S) - K_p S X_T(\tau/\text{SRT}) - K_L a_s S(\tau)]}{b(\tau) + (\tau/\text{SRT})} \tag{7–157}$$

For most cases where biodegradation is occurring and the loss is not overwhelming due to volatilization and/or solids adsorption, the liquid constituent concentration, based on Eq. (7–46), for a complete-mix reactor at steady state is given by

$$S = \frac{K_s[1 + b(\text{SRT})]}{\text{SRT}(\mu_m - b) - 1} \tag{7–158}$$

For the other exceptional cases, Eq. (7–156) and (7–157) can be solved simultaneously.

The approach outlined above can be used to estimate the fate of an organic compound in a complete-mix activated sludge reactor assuming steady-state conditions, a constant input of the organic constituent, and a fully acclimated culture. Predicting the fate of an organic constituent in an activated sludge treatment process is illustrated in Example 7–12.

EXAMPLE 7–12 Predicting the Fate of Benzene in an Activated Sludge Treatment Process

A complete-mix activated sludge system is used to treat domestic wastewater, but receives a wastewater discharge containing benzene. Given the following information on the activated sludge process design and biotic and abiotic rate information for benzene, what is the effluent soluble benzene concentration and the relative amounts of benzene lost through biodegradation, sorption to solids, volatilization, and in the liquid effluent?

1. Influent benzene concentration, $S_o = 2.0\ \text{g/m}^3$
2. System SRT $= 6.0$ d
3. Aeration tank retention time, $\tau = 0.25$ d
4. MLVSS concentration, $X_T = 2500\ \text{g/m}^3$
5. $K_p = 0.234 \times 10^{-3}\ \text{m}^3/\text{g}$
6. $K_L a_s = 3/\text{h} = 72/\text{d}$
7. $\mu_m = 2.0$ g VSS/g VSS·d
8. $K_s = 0.50\ \text{g/m}^3$
9. $b = 0.10$ VSS/g VSS·d
10. $Y = 0.60$ g VSS/g benzene

Solution

1. Determine the liquid benzene concentration using Eq. (7–158).

$$S = \frac{(0.5\ \text{g/m}^3)[1 + (0.10\ \text{g VSS/g VSS·d})(6.0\ \text{d})]}{(6.0\ \text{d})[(2.0 - 0.10)\ \text{g VSS/g VSS·d}] - 1} = 0.077\ \text{g/m}^3$$

2. Determine concentration of biomass degrading benzene *Xs* from Eq. (7–157):

$$X_s = \frac{Y[(S_o - S) - K_p S X_T(\tau/\text{SRT}) - K_L a_s S(\tau)]}{b(\tau) + (\tau/\text{SRT})}$$

$$X_s = \frac{\left\{\begin{matrix}(0.60\ \text{g/g})[(2.0 - 0.077)\text{g/m}^3] - (0.234 \times 10^{-3}\ \text{m}^3/\text{g})(0.077\ \text{g/m}^3) \times \\ (2500\ \text{g/m}^3)(0.25\text{d}/6.0\text{d}) - (72/\text{d})(0.077\ \text{g/m}^3)(0.25\text{d})\end{matrix}\right\}}{(0.10\ \text{g/g·d})(0.25\text{d}) + (0.25\text{d}/6.0\text{d})}$$

$$= 4.83\ \text{g/m}^3$$

3. Determine the loss of benzene due to biodegradation [1st term in Eq. (7–156)]:

$$\left(\frac{1}{Y}\right)\frac{\mu_m(S)}{(K_s + S)}(X_s)\tau = \left[\frac{1}{(0.6\text{ g/g})}\right]\left[\frac{(2.0/\text{d})(0.077\text{ g/m}^3)}{(0.5 + 0.077)\text{ g/m}^3}\right](4.83\text{ g/m}^3)(0.25\text{ d})$$

$$= 0.537\text{ g/m}^3$$

4. Determine the loss of benzene due to sorption [2nd term in Eq. (7–156)]:

$$\frac{K_p S X_T \tau}{\text{SRT}} = \frac{(0.234 \times 10^{-3}\text{ m}^3/\text{g})(0.077\text{ g/m}^3)(2500\text{ g/m}^3)(0.25\text{ d})}{6.0\text{ d}}$$

$$= 0.0019\text{ g/m}^3$$

5. Determine the loss of benzene due to volatilization [3rd term in Eq. (7–156)]:

$$(K_L a_s)(S)(\tau) = (72/\text{d})(0.077\text{ g/m}^3)(0.25\text{ d}) = 1.386\text{ g/m}^3$$

6. Summarize the losses due to the various mechanisms.

Pathway	Influent fate, g/m³	Fraction of total
Effluent	0.077	0.039
Biodegradation	0.537	0.268
Sorption	0.002	0.001
Volatilization	1.386	0.692
Total	2.002[a]	1.000

[a] 0.002 is due to round-off error.

a. For benzene, which is a volatile organic compound with a low solids partition coefficient, 69.2 percent of the influent benzene is transferred to the atmosphere, 26.8 percent is biodegraded, 3.9 percent remains dissolved in the liquid effluent, and 0.1 percent is sorbed onto the solids leaving the process.

b. The effect of the aeration system benzene K_La value on the relative amounts biodegraded or stripped to the atmosphere is illustrated in the following summary table:

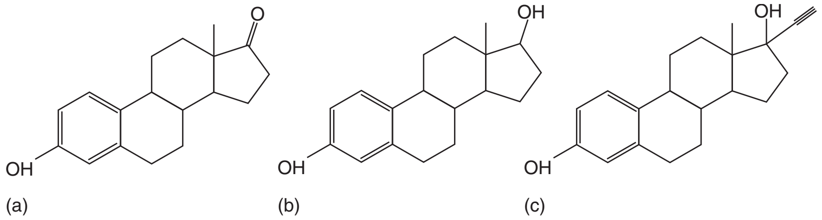
		K_La, h⁻¹			
Parameter	Unit	1.5	3.0	4.0	6.3
Effluent,	g/m³	0.077	0.077	0.077	0.052
Effluent	%	3.9	3.9	3.9	2.6
Biodegraded	%	61.4	26.8	3.7	0.0
Volatilized	%	34.6	69.2	92.3	97.3

7–16 BIOLOGICAL REMOVAL OF TRACE ORGANIC COMPOUNDS

Trace organic compounds (TrOCs) are present in domestic wastewater plant influent at ng/L to ug/L concentrations and include natural hormones, pharmaceuticals excreted by humans, and personnel care products such as shampoo, toothpaste, and fragrance substances. Pharmaceuticals include analgesics, anticonvulsants, beta blockers, antibiotics,

Figure 7–28

Chemical structure of estrogen compounds common in WWTPs: (a) E1 – Estrone, $C_{18}H_{22}O_2$, (b) E2 – 17β-estradiol, $C_{18}H_{24}O_2$, and (c) EE2 – 17α-ethinylestradiol, $C_{20}H_{24}O_2$.



and x-ray media. Some of these TrOCs along with other xenobiotic compounds contribute to endocrine disrupting compound (EDCs) activities in WWTP effluents. Compounds that stimulate estrogenic activity include natural hormones from humans and animals, the synthetic hormone 17α-Ethinylestradiol (EE2), Dieldrin, Methoxyclor, bisphenol A, phthalates, nonylphenol and octylphenol from the breakdown of detergents, polychlorobiphenyls (PCB), polyaromatic hydrocarbons (PAH), and dioxins (Combalbert and Hernandez-Raquet, 2010). From field studies on native fish populations the occurrence of intersex fish has been found in surface waters receiving effluent from WWTPs (Harshbarger et al., 2000; Hashimoto et al., 2000). The relative activity of some of the estrogen compounds found in wastewater, relative to 17β-Estradiol (E2) arbitrarily fixed at 100, is 246 for EE2, 2.5 for Estrone (E1) (see Fig. 7–28), 1.55 for genistein, 0.66 for bisphenol A, and 0.32 for nonylphenol (Pillon et al., 2005). The structure of E1, E2, and EE2 is illustrated on Fig. 7–28.

Removal of Trace Organic Compounds

The TrOCs in general are not volatile so that the main removal mechanisms in WWTPs are via biodegradation and adsorption to waste solids. The liquid-solids partition coefficient for estrogen compounds is relatively low, with Log Koc values at less than 3.5. As a concequence of the low K_{OC} values, most of the estrogen removal in secondary wastewater treatment is by biodegradation. Because of the low concentration of TrOCs in the influent wastewater, they can not support significant biogrowth, so that the TrOC biodegradation is done mainly by bacteria growing on other substrates. Cometabolic degradation of estrogens by ammonia oxidizing bacteria has been posed based on results of batch experiments at high ammonia and nitrite concentrations (Gaulke et al. 2008), but the main mechanism is by heterotrophic bacteria (Combalbert and Hernandez-Raquet, 2010). An example of results from full scale field studies showing the range of degradation ability in domestic wastewater treatment for a number of compounds of interest is presented in Table 7–16 (Stensel, 2011).

Steady-State Fate Model

A steady state model, similar to that given in Sec. 7–14 to describe the fate of xenobiotics, is presented for TrOCs to illustrate the key terms in TrOC removal in activated sludge processes. Because the TrOC compound concentrations are very low, a psuedo first-order biodegradation model has been used and is illustrated here for an estrogen (E) compound. An additional set of equations are also needed for estrogen and some other compounds that may enter the WWTP as conjugated compounds. Most of the estrogen excreted by humans is in urine as a conjugated compound. Deconjugation to free estrogen can occur in the wastewater collection and conveyance systems and in the WWTP. The rate of estrogen degradation is as follows:

$$r_{UE} = K_b X_{H,E}(E)V \tag{7–159}$$

Table 7–16

Example of TrOC removal in MBR activated sludge treatment at 30-d SRT (Stensel, 2011)

Compound	Average Influent Concentration, ng/L	Source
Highly biodegradable (>90 percent removal)		
E1, E2	30	Natural human hormones
EE2	110	Synthetic hormone
Acetaminophen	67,290	Analgesic
Naproxen	21,560	Analgesic
Tricolsan	1100	Antibacterial
Ibuprofen	13,490	Analgesic
Caffeine	50,680	Stimulant
Atenolol	3750	Beta blocker (blood pressure)
Moderately biodegradable (60–85 percent removal)		
Bisphenol A	290	Plasticize
Erythromycin	120	Antibiotic for bacteria
Trimethoprim	180	Antibiotic for bacteria
Oxybenzone	30	Ingredient in sunscreen
Poorly biodegradable (20–60 percent removal)		
Propranolol	31	Beta blocker (blood pressure)
Fluoxetine	40	Antidepressant
Gemfibrozil	3420	Antilipidemic (Cholesterol)
Sulfamethoxazol	1000	Antibiotic for bacteria
Metoprolol	390	Beta blocker (blood pressure)
Iopromide	3190	X-ray contrast
None or very poor biodegradation (unchanged or increase from deconjugation)		
Pentoxifylline	5	Improve blood flow
Dilcofenac	90	Analgesic
Dilantin	50	Anticonvulsant
Carbamazepine	250	Anticonvulsant

$$X_{H,E} = \eta_E X_T \tag{7–160}$$

where r_{UE} = estrogen removal rate by biodegradation, ng/d
$X_{H,E}$ = heterotrophic bacteria concentration capable of degrading estrogen, g/m^3
η_E = fraction of mixed liquor solids that are estrogen degraders
K_b = first order degradation rate coefficient, m^3/g·d
E = reactor soluble estrogen concentration, ng/m^3
V = reactor volume, m^3

Considering the production of free estrogen by deconjugation and removal of estrogen in waste solids by adsorption, the following steady state equations apply.

$$QE_o = k_b(\eta_E X_T)(E)V - K_c(\eta_c X_T)E_c V + \frac{K_{p,E}X_T(E)V}{(10^6)\text{SRT}} + (Q - Q_w)E \tag{7–161}$$

$$QE_{o,c} = K_c(\eta_c X_T)(E_c)V + \frac{K_{p,Ec}X(E_c)V}{(10^6)\text{SRT}} + (Q - Q_w)E_c \tag{7–162}$$

where E_o = influent estrogen concentration, ng/m^3
$E_{o,c}$ = influent conjugated estrogen concentration, ng/m^3
E_c = reactor soluble conjugated estrogen concentration, ng/m^3
η_c = fraction of mixed liquor solids that are estrogen-deconjugating bacteria
K_c = first order deconjugation coefficient, m^3/g·d
$K_{p,E}$ = liquid-solids partition coefficients for estrogen L/kg
$K_{p,EC}$ = liquid-solids partition coefficients for conjugated estrogen, L/kg
X_T = mixed liquor suspended solids concentration, g/m^3
Q = daily influent flowrate, m^3/d
Q_w = daily waste sludge flowrate, m^3/d

Until research finds a biomarker to identify the estrogen degrading- and estrogen deconjugating-bacteria, the kinetic values are normalized to the total mixed liquor concentration. First order degradation rate coefficients for EE2, range from 5-20 L/g MLSS·d. E1 and E2 degradation rates are about 5 times faster (Gaulke et al., 2009).

7–17 BIOLOGICAL REMOVAL OF HEAVY METALS

Metal removal in biological treatment processes is mainly by adsorption and complexation of the metals with the microorganisms. In addition, processes that result in transformations and precipitation of metals are possible. Microorganisms combine with metals and adsorb them to cell surfaces because of interactions between the metal ions and the negatively charged microbial surfaces. Metals may also be complexed by carboxyl groups found in microbial polysaccharides and other polymers or absorbed by protein materials in the biological cell. The removal of metals in biological processes has been found to fit adsorption characteristics as defined by the Freundlich isotherm model (see Sec. 11–7 in Chap. 11) (Mullen et al., 1989; Kunz et al., 1976). A significant amount of soluble metal removal has been observed in biological processes, with removals ranging from 50 to 98 percent depending on the initial metal concentration, the biological reactor solids concentrations, and system SRT. In anaerobic processes the reduction of sulfate to hydrogen sulfide can promote the precipitation of metal sulfides. A classic example is the addition of ferric or ferrous chloride to anaerobic digesters to remove sulfide toxicity by the formation of iron sulfide precipitates. The precipitation of heavy metals by hydrogen sulfide is discussed in Sec. 6–5 in Chap. 6.

PROBLEMS AND DISCUSSION TOPICS

7–1 Prepare a recipe for an inorganic medium to be used in a laboratory chemostat to grow 500, 1000, or 1200 mg VSS/d (value to be selected by instructor) of bacteria biomass, assuming that the chemical formula for the biomass can be described as $C_5H_7NO_2$. Determine the concentration of essential inorganic compounds as reported in Table 7–3 for a feed rate of 1 L/d. Assume that phosphorus is added as KH_2PO_4, sulfur as Na_2SO_4, nitrogen as NH_4Cl, and other cations added are associated with chloride.

7–2 Protein is a major component of bacterial enzymes. List the key cell components involved and the major steps that lead to protein production.

7–3 From the literature (e.g., *J. Appl. Environ. Microbiol.*) identify the key physiological, metabolic characteristics, and phylogenetic classification of a bacteria that may have a role in biological wastewater treatment or toxic degradation. Cite a minimum of 3 references.

7–4 From the literature, describe an application using molecular biology (e.g., molecular probes or other methods) techniques that can be related to biological wastewater treatment. Cite a minimum of 3 references.

7–5 A 1-L sample contains 22, 26, or 32 g (value to be selected by instructor) of casein ($C_8H_{12}O_3N_2$). If 18 g of bacterial cell tissue ($C_5H_7NO_2$) is synthesized per 50 g of casein consumed, determine the amount of oxygen required to complete the oxidation of casein to end products and cell tissue. The end products of the oxidation are carbon dioxide (CO_2), ammonia (NH_3), and water. Assume that the nitrogen not incorporated in cell-tissue production will be converted to ammonia.

7–6 A complete-mix suspended growth reactor, without a clarifier and recycle, is used to treat a wastewater flow containing only soluble organic substances. The influent BOD and COD are as follows:

		Wastewater		
Influent	Unit	1	2	3
BOD	mg/L	200	180	220
COD	mg/L	450	450	480

If the effluent dissolved BOD concentration is 2.5 mg/L, and the effluent volatile suspended solids concentration is 100 mg/L, determine (wastewater to be selected by the instructor): (a) the observed yield in terms of g VSS/g BOD, g VSS/g COD, and g TSS/g BOD, (b) the effluent total sCOD concentration including nonbiodegradable dissolved COD, and (c) the fraction of the influent BOD that is oxidized to CO_2 and H_2O. Assume the biodegradable COD/BOD ratio is 1.6 and 1.42 g O_2 eq/g biomass.

7–7 An aerobic complete-mix reactor (no recycle) with a volume of 1000 L receives a 500 L/d wastewater flow and has an effluent soluble COD concentration of 10 mg/L. For one of the wastewaters with the characteristics given below (to be selected by instructor), determine (a) the τ value for the reactor in days, (b) the oxygen used per day in (g/d), (c) the effluent volatile suspended solids concentration (assume biomass oxygen equivalent of 1.42 g O_2/g VSS), and (d) the observed yield in g VSS/g bsCOD removed.

		Wastewater		
Item	Unit	1	2	3
Influent sCOD	mg/L	1000	1800	600
Reactor oxygen uptake rate	mg/L·h	10	15	8

7–8 Using the half-reaction free energy values given in Table 7–6, calculate and compare the biomass yields (g VSS/g COD_r) for the degradation of methanol, carbohydrate mixture, or ethanol (constituent to be selected by instructor) with oxygen and nitrate as the electron acceptors. Assume ammonia is available for cell synthesis needs and 1.42 g O_2 eq/g biomass.

7–9 Using the half-reaction free energy values given in Table 7–6, calculate and compare the biomass yields (g VSS/g COD_r) for the degradation of methanol, or ethanol (constituent to be selected by instructor) with nitrate and nitrite as the electron acceptors. Assume ammonia is available for cell synthesis needs and 1.42 g O_2 eq/g biomass. Using the results, compare the amount of methanol (or ethanol) needed for denitrification of NO_3-N versus NO_2-N in g COD/g N.

7–10 Nitrate and sulfate are both available in an anaerobic laboratory chemostat with both nitrate-reducing and sulfate-reducing bacteria present. The chemostat is fed continuously a solution containing the electron acceptors in equal amounts, glucose, and a nutrient media. Which biological populations will remain after long-term operation? Explain.

7–11 For Example 7–3, use the half reactions to write a balanced equation of acetate oxidation by methanogenic bacteria.

7–12 For the synthesis yield values given in Table 7–7 for organic compound degradation, what are the respective f_e and f_s values?

7–13 Compare the end products of organic compound degradation under the following conditions, and discuss how the bacterial synthesis yields are affected by them: aerobic (oxygen as acceptor), fermentation (organic compound as electron acceptor), and methanogenesis (CO_2 as electron acceptor).

7–14 If bacterial cells are of the coccus type with a diameter of 1.0, 1.3, or 1.5 μm and are 80 percent water with 90 percent of the dry weight as organic, determine (cell diameter to be selected by the instructor): (a) the volume and organic mass of one cell and (b) the number of cells present in one liter of a biomass suspension with a concentration of 100 mg VSS/L.

7–15 For aerobic bacteria with an assumed generation time of 20, 30, or 60 min (time to be selected by instructor), how many bacteria would be present after 12 h, if 20 cells are present at time zero? Using the bacteria volume and mass from Problem 7–13 for a 1-μm diameter bacteria, what would be the dry weight of the bacteria after 12 h in mg volatile suspended solids?

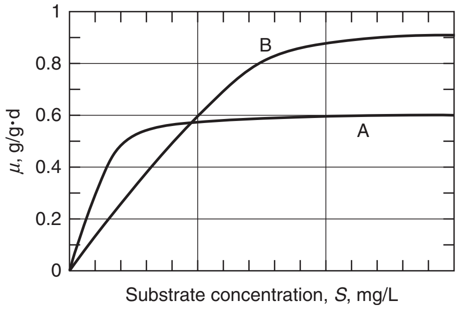
7–16 Consider a batch reaction with nitrifying bacteria in a chemostat. The initial concentration of nitrifying bacteria is 10 mg/L, and the initial substrate concentration is 50 mg NH_4-N/L. The NH_4-N is oxidized to NO_2-N, and the cell yield is 0.12 g VSS/g NH_4-N oxidized. The chemostat DO concentration is maintained at 3.0 mg/L. Other kinetic coefficients related to substrate utilization and growth are one of the following to be selected by instructor:

		Wastewater		
Coefficient	Unit	1	2	3
μ_{max}	g VSS/g VSS·d	0.60	0.75	0.60
K_n	mg/L	0.50	0.50	0.75
K_o	mg/L	0.50	0.50	0.50
b	g VSS/g VSS·d	0.08	0.08	0.04

What is the NH_4-N and biomass concentration at 0.50 d?

Plot the substrate and biomass concentration versus time up through 23 h. (Hint: one solution approach is to use a spreadsheet to solve for the biomass and substrate concentration at small time increments. Use time increments of 0.25 h).

7–17 Curves A and B represent the Monod kinetics for two different bacteria capable of degrading the same substrate. You are to operate a laboratory continuous flow CMAS reactor without recycle that is inoculated with bacteria A and B. In the first experiment (I) a high SRT is used (10 d or greater) and in the second (II) a very low SRT is used (about 1.1 d). Which bacteria will be dominant in experiments I and II? Explain why.



7–18 A complete-mix aerobic reactor without solids recycle is used to treat a wastewater containing 100 mg/L phenol (C_6H_6O) at 20°C. Using the following kinetic coefficients (coefficient set 1, 2, or 3 to be selected by instructor) determine (a) the minimal hydraulic retention time t in days at which the biomass can be washed out faster than they can grow, (b) the minimum τ value at 10°C, assuming the temperature-activity coefficient θ is 1.07 for k and 1.04 for b, (c) the effluent phenol and biomass concentration at a τ value of 4.0 d at 20°C, and (d) the amount of oxygen required in kg/d for a τ value of 4.0 d assuming a flowrate of 100 m^3/d. Plot the phenol and biomass concentration and the amount of oxygen required versus τ in days, for τ from 3.3 to 15 d at 20°C.

		Wastewater		
Coefficient	Unit	1	2	3
k	g phenol/g VSS·d	0.90	0.80	0.90
K_s	mg phenol/L	0.20	0.15	0.18
Y	g VSS/g phenol	0.45	0.45	0.40
b	g VSS/g VSS·d	0.10	0.08	0.06

7–19 Laboratory test reactors have been operated at different SRT values at steady state to obtain biological kinetic coefficients for a wastewater with soluble constituents only. The reactors are complete-mix and aerated with clarifiers and solids recycle. The τ value in all cases is 0.167 d and the SRT values were varied for the five tests. The influent and effluent soluble COD and reactor MLVSS concentrations are summarized as follows:

Test no.	SRT, d	S_o, mg COD/L	S, mg COD/L	X, mg VSS/L
1	3.1	400	10.0	3950
2	2.1	400	14.3	2865
3	1.6	400	21.0	2100
4	0.8	400	49.5	1050
5	0.6	400	101.6	660

From these results determine the values for the biokinetic coefficients, k, K_s, μ_m, Y, and b. (Note: calculate the solids production at each SRT.)

7–20 The following data were obtained using four bench-scale continuous-flow activated sludge units to treat a food-processing waste. Using these data, determine Y and b.

	Parameter		
Unit	X, g MLVSS/L	r_g, g MLVSS/L·d	U, g BOD/g MLVSS·d
1	18.81	0.88	0.17
2	7.35	1.19	0.41
3	7.65	1.42	0.40
4	2.89	1.56	1.09

7–21 Using the data given below for three different complete-mix activated sludge reactors, determine (reactor to be selected by instructor): (a) the system SRT, (b) how much oxygen is required in kg/d if the effluent soluble COD concentration = 5 mg/L, and (c) the oxygen uptake rate, expressed in mg/L·h, at steady state in the aeration tank. Assume 1.42 g COD/g VSS.

		Reactor		
Item	Unit	1	2	3
Aeration tank MLVSS	mg/L	3000	3000	3000
Aeration tank volume	m^3	1000	1000	1000
Influent flowrate	m^3/d	5000	5000	5000
Waste sludge flowrate	m^3/d	59	45	65
Waste sludge VSS concentration	mg/L	8000	8000	8000
Influent soluble COD concentration	mg/L	400	400	400

7–22 A complete-mix activated sludge process with secondary clarification and sludge recycle is used to treat a dairy wastewater at a flowrate of 1000 m^3/d with a degradable influent COD of 3000 mg/L and BOD of 1875 mg/L. The MLSS concentration is 2800, 3300, or 3500 mg/L (MLSS value to be selected by instructor), MLVSS/MLSS ratio is 0.80, effluent TSS concentration is 20 mg/L, t is 24 h, recycle MLSS concentration is 10,000 mg/L, and waste sludge flowrate from the recycle line is 85.5 m^3/d. Using the given information, determine (a) the system SRT, the F/M ratio in g BOD/g MLVSS·d, and the volumetric BOD loading rate (kg/m^3·d), (b) the observed yield in terms of g TSS/g BOD and g TSS/g COD, and (c) the synthesis yield, assuming that $b = 0.10$ g VSS/g VSS·d and $f_d = 0.15$ g VSS/g VSS.

7–23 A conventional activated sludge plant is operated at SRT values of 8, 10, or 12 d (value to be selected by instructor). The reactor volume is 8000 m^3 and the MLSS concentration is 3000 mg/L. Determine (a) the sludge production rate, (b) the sludge wasting flowrate when wasting from the reactor, and (c) the sludge wasting flowrate when wasting from the recycle line. Assume that the concentration of suspended solids in the recycle is equal to 10,000 mg/L, and the solids loss in the secondary clarifier effluent is minor and can be neglected.

7–24 A complete-mix activated sludge process with a clarifier and sludge recycle receives an influent wastewater flowrate of 2000 m^3/d and influent particulate concentration of 400, 500, or 600 mg VSS/L (value to be selected by instructor) that is entirely biodegradable. The volume of the activated sludge reactor is 500 m^3. The biokinetic coefficients for particulate degradation (Eq. 7–20) are $k_p = 2.2$ g VSS/g biomass·d and $K_x = 0.15$ g VSS/g biomass. The yield and endogenous decay coefficients are 0.50 g biomass/g VSS and 0.10 g VSS/g VSS·d, respectively. Using the given information: (a) develop a steady-state mass balance for particulate removal in the activated sludge system, (b) develop equations for the aeration tank particulate and biomass concentrations as a function of SRT (assume that the effluent contains no degradable particulates, particulates only leave the system via the waste sludge, and soluble COD is negligible), (c) determine the biomass and particulate concentrations in the aeration tank at SRT values of 3, 5, and 10 d, and (d) determine the percent removal of particulates at 3, 5, and 10 d.

7–25 A completely mixed activated sludge process is operated at a 10.5-d SRT, 12°C temperature, and a 3500 mg/L MLSS concentration to produce an effluent NH_4-N concentration of 1.0 mg/L. The average sludge production rate is 753 kg TSS/d, and the oxygen consumption rate is 1225 kg/d including that for nitrification. The system aeration rate (air supply) is controlled to maintain a DO concentration at 1.0 mg/L in the aeration basin. If the oxygen demand increases, the aeration blower air output is automatically increased to maintain a DO of 1.0 mg/L. The plant operator is thinking about tinkering with the SRT and asks you what changes would be expected if the SRT is increased to 15 d. Indicate if values for the following parameters will increase (I), decrease (D) or remain the same (S) and give reasons for your selection.

a. Sludge production rate, kg/d

b. Oxygen consumption rate, kg O_2/d

c. Effluent soluble biodegradable COD concentration, mg/L

d. Aeration tank MLSS concentration, mg/L

e. Effluent NH_4-N concentration, mg/L

f. Effluent NO_2-N concentration, mg/L

7–26 The following sets of kinetic coefficients (to be selected by instructor) are given for the treatment of a municipal wastewater with an influent degradable COD of 300 mg/L and influent nbVSS concentration of 100 mg/L. Using these data and assuming the effluent degradable COD concentration is negligible compared to the amount of COD removed, prepare plots of (a) the observed yield (as g VSS/g COD) removed as a function of SRT and (b) the g oxygen used/g COD removed as a function of SRT. On the plot in part (a) also show the fraction of the yield from cell debris and influent nbVSS.

		Coefficient set		
Coefficient	Unit	1	2	3
Y	g VSS/g COD	0.40	0.40	0.35
b	g VSS/g VSS·d	0.10	0.08	0.12
f_d	g VSS/g VSS	0.10	0.15	0.15

7–27 Design a complete-mix activated sludge process with recycle to treat an industrial wastewater with one of the following characteristics (to be selected by instructor) at peak month conditions.

		Wastewater		
Item	Unit	1	2	3
Flowrate	m^3/d	4000	4300	4000
BOD	mg/L	800	600	1000
nbVSS	mg/L	200	200	200
TKN	mg/L	30	30	40
Total phosphorus	mg/L	8	8	6
Temperature	°C	15	15	15

The relevant biokinetic coefficients and operating conditions are

$Y = 0.45$ g VSS/g COD	SRT = 10 d
$b = 0.10$ g VSS/g VSS·d	Return sludge = 8000 mg TSS/L
$\mu_m = 2.5$ g VSS/g VSS·d	Aeration tank MLSS = 2500 mg/L
$K_s = 20$ mg COD/L	Clarifier effluent TSS = 15 mg/L
$f_d = 0.10$ g VSS/g VSS	
bCOD = 1.6 (BOD)	

Using the given information and biokinetic coefficients, determine (a) the aeration tank volume (m^3), the amount of waste solids produced/d (kg/d), the oxygen requirement (kg/d), the aeration tank oxygen uptake rate (mg/L·h), the effluent soluble BOD concentration, the return sludge recycle ratio for the following design conditions, and the MLVSS to MLSS ratio, and (b) whether supplemental nitrogen or phosphorus is required and, if so, how much in mg/L? Assume the biomass contains 12 percent nitrogen and 2 percent phosphorus on a volatile suspended solids basis. Assume no nitrification occurs.

7–28 For the same industrial wastewater application given in Problem 7–27, powdered activated carbon (PAC) is added to the influent at a dose of 50 mg/L to sorb potential toxic substances. The SRT is still held at 10 d. Determine the MLSS concentration, the MLVSS/MLSS ratio, and the total daily sludge production in kg TSS/d with the PAC addition.

7–29 A complete-mix activated sludge system receives wastewater with one of the following characteristics (wastewater to be selected by instructor):

		Wastewater		
Item	Unit	1	2	3
Flowrate	m^3/d	6000	6000	6000
Biodegradable BOD	mg/L	300	400	500
Influent nbVSS	mg/L	100	100	150

The relevant design criteria are
Flowrate = 6000 m^3/d
Biodegradable COD = 300 mg/L
Influent nbVSS = 100 mg/L

The following biokinetic coefficients can be assumed:
$Y = 0.40$ g VSS/g COD
$b = 0.10$ g VSS/g VSS·d
$f_d = 0.10$ g VSS/g VSS
$\mu_m = 5.0$ g VSS/g VSS·d
$K_s = 20$ mg COD/L

If the system aeration oxygen transfer capacity is 52 kg O_2/h, what maximum SRT can be used so that the oxygen requirements can be met by the existing oxygen transfer capacity?

7–30 The kinetics for substrate utilization can be described by a first-order relationship (Eq. 7–18) ($r_{su} = kSX$). (a) Using the given first-order kinetic relationship instead of the Michaelis-Menten relationship for substrate utilization, derive a steady-state relationship that can be used to calculate the effluent soluble substrate concentration from a complete-mix suspended growth reactor. Verify that Eq. (7–42) can be used to determine the biomass (X) concentration. (b) For the following reactor conditions and biokinetic information, determine the SRT needed to provide an effluent soluble substrate concentration of 1.0 mg/L, and the biomass concentration.

S_o = 500 mg/L COD
τ = 0.25 d
Y = 0.50 g VSS/g COD removed
b = 0.06, 0.10, or 0.12 g VSS/g VSS·d (to be selected by instructor)
$r_{su} = -kSX$, where k = 0.504 g/g·d

7–31 An aerobic digester receives thickened waste activated sludge and holds it for a number of days for further aeration and solids destruction by endogenous decay of the biomass. The biomass concentration entering the digester is defined as X_o and is 24 g VSS/L. The influent flow also contains 6 g/L of inert nonbiodegradable VSS ($X_{I,o}$) for a total influent VSS concentration of 30 g/L. The digester biomass VSS concentration is X, and the inert VSS concentration is X_I. The digester volume is V, and the hydraulic retention time (V/Q) is 20 d. A membrane is installed in the digester, and digester liquid is drawn through the membrane to provide solids thickening within the digester. The membrane effluent liquid flow is defined as Q_M, and its VSS concentration zero. The biomass VSS endogenous decay rate is given as $r_{xd} = bX$, where: r_{xd} = the rate of biomass solids destruction (g VSS/L·d, and b = specific biomass endogenous decay rate, (g VSS/g VSS·d). Values for b and f_d = 0.10 g/g·d and 0.10 g/g

a. Write the mass balance equation for X and give equation to solve for X at steady state conditions.

b. Write the mass balance equation for $X_{I,o}$ and give equation to solve for $X_{I,o}$ at steady state conditions.

c. Write an equation defining the SRT in terms of the reactor volume (V), biomass concentration (X), and waste solids flowrate.

d. What is the value for X, $X_{I,o}$, SRT and percent reduction of influent biomass and influent total VSS, if the membrane liquid removal rate is equal to 50 percent of Q ($Q_M = 0.50Q$)

7–32 Consider a biofilm treating a liquid stream containing acetate and dissolved oxygen with a stagnant liquid layer above the biofilm. (a) Using the stoichiometric relationship developed for the biological degradation of acetate in Example 7–4, determine the maximum acetate concentration in the bulk liquid that can be satisfied before the aerobic degradation in the biofilm is limited by the surface flux rate of oxygen, where the bulk liquid DO concentration is 2.0, 3.0, or 4.0 mg/L (DO value to be selected by instructor). (b) Compare these results to the results of NH_4-N oxidation in Example 7–7. Why is the bulk liquid NH_4-N concentration so much lower at a DO concentration of 2.0 mg/L?

Given:
Acetate diffusivity coefficient = 0.9 cm^2/d
Oxygen diffusivity coefficient = 2.6 cm^2/d

7–33 An activated sludge system treating domestic wastewater is operated at a solids retention time of 10 d with a mixed-liquor temperature of 18°C. For many weeks nitrification has occurred, with an effluent NH_4-N concentration reported at less than 1.0 mg/L. After some time, the nitrification performance declines with effluent NH_4-N concentrations exceeding 10 mg/L. As the city engineer you are requested to investigate the cause of the decline in performance and to make recommendations for actions that will get the discharge quality back in compliance. Describe possible causes for the decline in nitrification efficiency and how you would evaluate the problem.

7–34 Using the data in Table 7–13, what SRT is needed to achieve a steady state effluent NH_4-N concentration of 1.0 mg/L for an aerobic suspended growth nitrifying system at 20°C and for an anammox suspended growth system at 30°C? Assume the aerobic system is not limited by dissolved oxygen and the anammox system is not limited by nitrite.

7–35 Using the half-reactions from Table 7–6 calculate the oxygen equivalent of nitrite (g O_2/g NO_2-N), for biological reaction with nitrite as the electron acceptor instead of oxygen.

7–36 An anoxic suspended growth reactor is operated at an SRT of 5.0 d treating clarifier effluent from an activated sludge nitrification process. Acetate is added as the electron donor. Given the following coefficients for acetate under nitrate reduction conditions, determine (a) How much acetate is needed, in kg/d, to remove the influent NO_3-N concentration of 40.5, 20.5 or 30.5 g/m^3 (to be selected by instructor) in a treatment flowrate of 4000 m^3/d, (b) The biomass production rate in kg/d.

The reactor effluent acetate concentration at the 5.0-d SRT is 2 mg/L. The effluent NO_3-N is 0.50 mg/L. The nitrogen for biomass growth is from the influent NO_3-N.

$Y = 0.3$ g VSS/g COD removed
$b = 0.08$ g VSS/g VSS·d

Nitrogen for biomass growth = 0.12 g N/g biomass VSS

Ignore the biomass debris production ($f_d = 0$). What is the COD of acetate? i.e. g COD/g acetate. Provide a mass balance and steady state expression for the reactor acetate COD, biomass, and NO_3-N concentration.

7–37 Two complete-mix suspended growth laboratory reactors with sludge recycle fed the same synthetic wastewater are operated in parallel at the same aerobic SRT. One reactor has an anaerobic/aerobic sequence to promote enhanced biological phosphorus removal, and the other is operated only with the aerobic portion. The influent flow contains 100, 200, or 300 mg/L acetate (to be selected by instructor). The phosphorus and volatile fraction contents of the two mixed liquors are as follows: The lower VSS/TSS ratio for the biological phosphorus removal reactor accounts for both polyphosphate and associated cations in the storage products.

Reactor	g P/g VSS	g VSS/g TSS
Aerobic only	0.015	0.85
EBPR	0.250	0.65

Using the following operating conditions and coefficients, how much phosphorus is removed from the influent for each system in mg/L, and what are the aerobic reactor MLVSS and MLSS concentrations? (Note: for this problem the coefficients are assumed equal for both types of organisms, but in practice they may be different.)

$Y = 0.40$ g VSS/g COD
$b = 0.10$ g VSS/g VSS·d
SRT = 5 d
$\tau = 3$ h
$f_d = 0.10$ g VSS/g VSS

7–38 A laboratory reactor is to be operated to study operating conditions that affect biological phosphorus removal. The influent phosphorus concentration will be 10, 20, or 30 mg/L (as selected by instructor). What minimum concentrations of magnesium, potassium, and calcium should be in the influent liquid?

7–39 For the enhanced biological phosphorus removal process shown in Figure 7–23, indicate the effect of the changes listed below (one at a time from a base case design) on the effluent soluble phosphorus concentration. Will it increase (I), decrease (D) or remain the same (S)? Give a reason for each selection. Note the base case is operated at a low SRT with no nitrification but sufficient for PAO growth.

a. The total system SRT is increased and nitrification occurs.

b. The fraction of rbCOD in the influent biodegradable COD increases from 20 percent of bCOD to 35 percent of bCOD.

c. It is summer and the activated sludge temperature is 25°C and the pH drops from 7.5 to 6.8 because the city budget for purchasing alkalinity was depleted due to an unexpected increase in alkalinity costs.

d. The aeration tank DO concentration drops long term from a normal level of 2.0 mg/L to between 0.30 and 0.50 mg/L due to an equipment failure.

7–40 An anaerobic treatment process is used to treat a flowrate of 500 m^3/d with an influent soluble COD concentration of 2000, 5000, or 9000 mg/L (value to be selected by instructor). The net biomass yield is 0.04 g VSS/g COD removed and 95 percent soluble COD removal occurs at a temperature of 30°C. Assuming the gas contains 65 percent methane, calculate the total gas flow in m^3/d. What is the energy value of the gas produced in kJ/d? (The heat value of methane is 50.1 kJ/g at 30°C.)

7–41 A professor claims that the effect of processing food waste in an anaerobic digester and using the methane for fuel has a lower greenhouse gas effect than if the food waste is composted and used for a beneficial use for application on agriculture land. Do you agree or disagree? Explain the basis for your position.

7–42 Based on a review of the literature (cite a minimum of two references) explain the importance of the syntrophic relationship between methanogens and acid fermenters in an anaerobic process. What is the effect (increase, decrease, or remain the same) on the gas production rate, percent methane in the gas, volatile fatty acid concentration, and pH if an upset occurs to create an imbalance between fermenters and methanogens?

7–43 Modify Eqs. (7–156), (7–157), and (7–158) based on using first-order kinetics for the substrate removal rate versus the Monod growth kinetic model, where: $r_{su} = kSX$. The first-order model is often used to describe the biodegradation kinetics of a number of priority pollutants.

7–44 Assume a complete-mix reactor is to be used to treat a wastewater containing a priority pollutant with the following characteristics and other easily degradable organic compounds. The priority pollutant is not very volatile so that losses due to stripping can be ignored. Using the following information determine (a) the fate of the compound in terms of biodegradation losses, and removal in the system effluent and waste sludge, and (b) the values computed in part (a) if the value for μ_m is 3 times higher.

Design data and coefficients:

System SRT (to be selected by instructor) 5, 10, or 15 d
Reactor MLVSS = 2000 mg/L
Reactor $\tau = 0.25$ d
Compound characteristics and biokinetic coefficients:
Influent concentration = 5.0 mg/L
$K_p = 15 \times 10^{-3}\ m^3/g$
$\mu_m = 2.0$ g VSS/g VSS·d
$Ks = 0.4\ g/m^3$
$Y = 0.6$ g VSS/g compound
$b = 0.08$ g VSS/g VSS·d

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8

Suspended Growth Biological Treatment Processes

WORKING TERMINOLOGY

Term	Definition
Activated sludge process	Biological treatment process that involves the conversion of organic matter and/or other constituents in the wastewater to gases and cell tissue by a large mass of aerobic microorganisms maintained in suspension by mixing and aeration. The microorganisms form flocculent particles that are separated from the process effluent in a sedimentation tank (clarifier) and are returned subsequently to the aeration process or wasted.

Term	Definition
Aerobic (oxic) processes	Biological treatment processes that occur in the presence of free dissolved oxygen; oxygen is consumed by aerobic microorganisms in oxidation/reduction reactions to produce energy for cell growth and cell maintenance.
Anaerobic processes	Biological treatment processes that occur in the absence of oxygen.
Anoxic process	Biological treatment process that occurs in the absence of free dissolved oxygen where nitrate and nitrite are used as the main electron acceptors in biological oxidation/reduction reactions; denitrification is an example of an anoxic process.
Biomass	The total mass of solids in a reactor consisting mainly of organic matter and microorganisms.
Biological nutrient removal (BNR)	The term applied to the removal of nitrogen and phosphorus in biological treatment processes.
Denitrification	The biological process by which nitrate or nitrite is reduced to nitrogen and other gaseous end products.
Enhanced biological phosphorus removal (EBPR)	Removal of phosphorus by extraordinary storage in bacteria selected in anaerobic/aerobic process configuration and subsequent solids separation.
Hindered settling	Settling which occurs when the activated sludge flocs interfere with each other as they settle.
Facultative processes	Biological treatment processes in which the organisms can function in the presence or absence of molecular oxygen.
Fermentation	The conversion of organic matter to volatile fatty acids in the absence of oxygen, nitrate, and nitrite.
Membrane bioreactor (MBR)	A process that combines a suspended growth process with a membrane separation system within the process aeration tank; membrane separation is accomplished by either microfiltration or ultrafiltration.
Membrane flux	The rate of flow across a membrane per unit of surface area, $L/m^2 \cdot h$.
Mixed liquor suspended solids (MLSS)	The biomass contained in a treatment reactor used to bring about treatment of the organic material in wastewater.
Nitrification	The two-step biological process by which nitrogen (mostly in the form of ammonia) is converted to nitrite and then to nitrate.
Nocardioform foam	A thick layer of brown, biological foam caused by a filamentous bacteria that forms on the top of aeration tanks and secondary clarifiers.
Nonbiodegradable volatile suspended solids (nbVSS)	These are suspended solids contained in influent wastewater to activated sludge processes that are organic but not biodegradable. They impact sludge production.
Phosphorus accumulating organisms (PAOs)	Heterotrophic bacteria selected in EBPR processes that have the ability for high intracellular phosphorus storage.
Readily biodegradable COD (rbCOD)	Dissolved biodegradable organic substrates which are removed by bacteria much faster than colloidal or particulate degradable COD. The rbCOD impacts spatial oxygen demand, EBPR removal efficiency, and denitrification rates.
Sequencing batch reactor (SBR)	An SBR is a batch fill and draw activated sludge treatment process. It involves a treatment sequence of fill, react, settling, supernatant decanting, and idle. Activated sludge aeration and liquid solids separation occurs in the same tank.
Simulation models	Mathematical models, based on a set of equations, used to assess the effects of kinetics and changes in the wastewater characterizes on process performance.
Simultaneous nitrification and denitrification (SNdN)	Nitrogen removal occurs in same activated sludge floc or in a biofilm due to nitrification in aerobic outer layer and denitrification in interior due to the lack of dissolved oxygen and presence of nitrate or nitrite.

Term	Definition
Sludge production	The amount of solids produced during the biological processing of wastewater including influent nonbiodegradable solids and the biomass resulting from the conversion of organic.
Sludge yield	The amount of solids produced relative to the amount of BOD or COD removed during the biological processing of wastewater.
Solids flux analysis	A method used to determine the area required for hindered settling based on an analysis of the solids (mass) flux.
Solids retention time (SRT)	The average period of time in which solids remain in a suspended growth process (also called sludge age).
Staged process	Processes which occur with more than one independent reactor or compartment in series.
Surface overflow rate	The hydraulic flowrate applied relative to the clarifier surface area ($m^3/m^2 \cdot d$).
Suspended growth processes	Biological treatment processes in which microorganisms responsible for the conversion of organic matter or other constituents in the wastewater to gases and cell tissue are maintained in suspension within the liquid.
Volumetric organic loading rate	The amount of BOD or COD applied to the aeration tank volume per day (e.g., kg BOD or COD/$m^3 \cdot d$).

The theory of biological wastewater treatment is presented and discussed in detail in Chap. 7. Biological treatment processes, as noted in Chap. 7, may be classified as aerobic and anaerobic suspended growth, attached growth, and various combinations thereof. The focus of this chapter is on suspended growth treatment processes as exemplified by the activated sludge process for BOD and nitrification and for nitrogen and phosphorus removal. Attached growth and combined processes are discussed in Chap. 9, and suspended and attached growth anaerobic processes are considered in Chap. 10. Included in this chapter are (1) introduction to the activated sludge process. (2) wastewater characterization, (3) fundamentals of process selection, design, and control, (4) selector types and design considerations, (5) use of simulation models for activated sludge process design considerations, (6) processes for BOD removal and nitrification, (7) processes for biological nitrogen removal, (8) processes for enhanced biological phosphorus removal, (9) aeration tank design for activated sludge processes, (10) analysis of liquid separation for activated sludge processes with clarifiers, (11) design considerations for secondarily clarifiers, and (12) solids separation for membrane bioreactors. Aerated lagoons, non-aerated lagoons, and stabilization ponds are not covered in this text, as they are used mainly for small rural communities where sufficient land is available and discharge requirements may not be as stringent as in urban areas. Detailed design information on aerated lagoons and stabilization ponds may be found in the 4th edition of this textbook (Tchobanoglous et al., 2003). Additional sources of information may be found in Crites and Tchobanoglous (1998) and Reed et al. (1995).

8–1 INTRODUCTION TO THE ACTIVATED SLUDGE PROCESS

To provide a basis for the process designs presented in the subsequent sections of this chapter, it will be useful to consider (1) a brief summary of the historical development of the activated sludge process, (2) a description of the basic process, (3) a brief review of the evolution of the activated sludge process, and (4) an overview of recent process developments.

Historical Development of Activated Sludge Process

The activated sludge process is now used routinely for the biological treatment of municipal and industrial wastewaters. The antecedents of the activated sludge process date back to the early 1880s in England, to the work of Dr. Angus Smith, who investigated the aeration of wastewater in tanks, which hastened the oxidation of the organic matter. The aeration of wastewater was studied subsequently by a number of investigators, and in 1910 Black and Phelps reported that a considerable reduction in putrescibility could be secured by forcing air into wastewater in basins. In experiments with aerated wastewater, conducted at the Lawrence Experiment Station during 1912 and 1913, Clark and Gage found that growths of organisms could be cultivated in bottles and in tanks partially filled with roofing slate spaced about 25 mm (1 in.) apart and that these growths greatly increased the degree of purification obtained (Clark and Adams, 1914).

The results of the work at the Lawrence Experiment Station, with respect to the treatment of wastewater, were so striking that knowledge of them led Dr. G. J. Fowler of the University of Manchester, England to suggest that experiments along similar lines be conducted at the Manchester Sewage Works where Ardern and Lockett carried out valuable research on the subject. During the course of their experiments, Ardern and Lockett found that the sludge played an important part in the results obtained by aeration, as announced in their paper of May 3, 1914 (Ardern and Lockett, 1914). The process was named *activated sludge* by Ardern and Lockett because it involved the production of an activated mass of microorganisms capable of aerobic stabilization of organic material in wastewater (Metcalf & Eddy, 1935).

Basic Process Description

By definition, the basic activated sludge treatment process, as illustrated on Figs. 8–1(a) and (b), consists of the following three basic components: (1) a reactor in which the microorganisms responsible for treatment are kept in suspension and aerated; (2) liquid-solids separation unit, usually in a sedimentation tank; and (3) a recycle system for returning solids removed from the liquid-solids separation unit back to the reactor. Numerous process configurations have evolved employing these components. An important feature of the activated sludge process is the formation of flocculent settleable solids that can be removed by gravity settling in sedimentation tanks. In most cases, the activated sludge process is employed in conjunction with physical and chemical processes that are used for the preliminary and primary treatment of wastewater (discussed in Chap. 5), and post treatment, including disinfection (Chap. 12), and possibly filtration (Chap. 11).

Historically, most activated sludge plants have been used to treat wastewaters that have been pretreated by primary sedimentation, as shown on Figs. 8–1(a) and (b). Primary sedimentation is most efficient at removing settleable solids, whereas the biological processes are essential for removing soluble, colloidal, and particulate (suspended) organic substances; for nitrification and denitrification; and for biological phosphorus removal. For applications such as treating wastewater from smaller-sized communities, primary treatment is often not used as more emphasis is placed on simpler and less operator-intensive treatment methods. Primary treatment is omitted frequently in areas of the world that have hot climates, where odor problems from primary tanks and primary sludge can be significant. For these applications, various modifications of conventional activated sludge processes are used, including sequencing batch reactors, oxidation ditch systems, and membrane bioreactors.

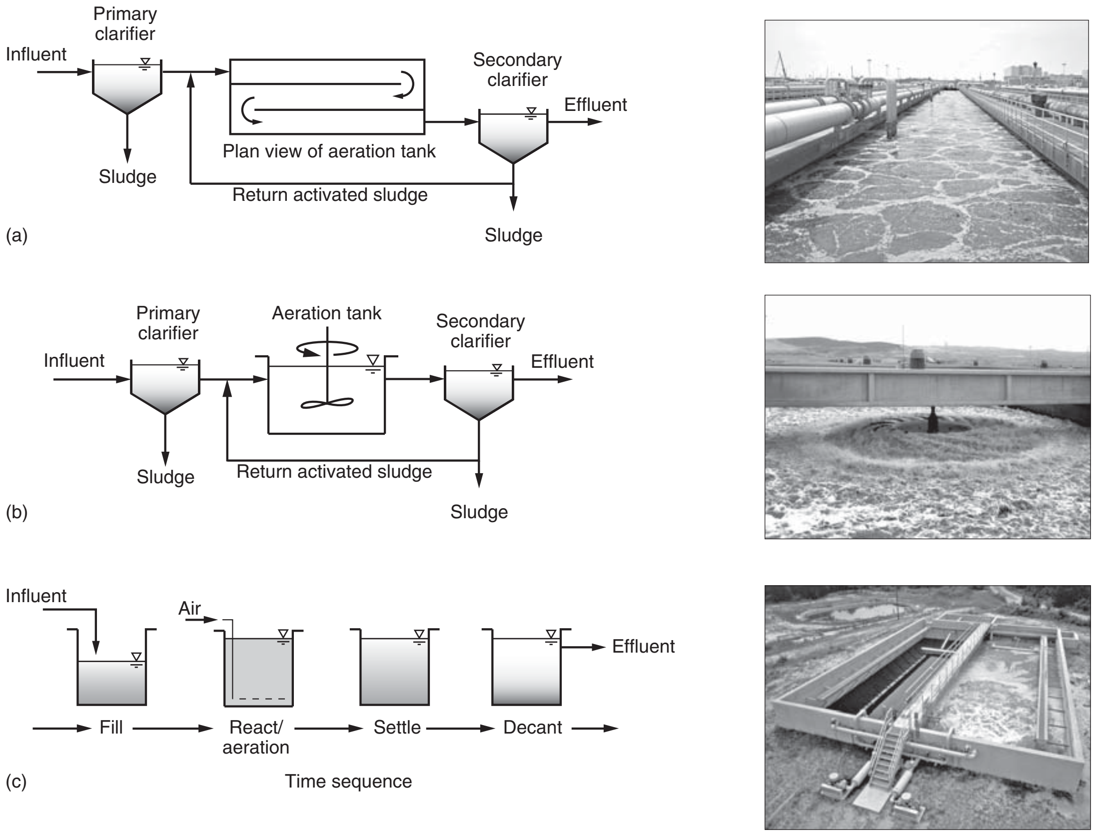


Figure 8–1

Typical activated sludge processes with different types of reactors: (a) schematic flow diagram of plug-flow process and view of plug-flow reactor, (b) schematic flow diagram of complete-mix process and view of complete-mix activated sludge reactor, and (c) schematic diagram of sequencing batch reactor process and view of sequencing batch reactor.

Evolution of the Conventional Activated Sludge Process

Prior to the 1980s, the principal objective of activated sludge process designs was aimed mainly at achieving a "secondary treatment" standard of 85 percent BOD and TSS removal. Since then, more emphasis has been placed on meeting more stringent discharge limits as well as the removal of nutrients (nitrogen and phosphorus). Thus, a number of activated sludge processes and design configurations have evolved in response to (1) the need for higher-quality effluents from wastewater treatment plants; (2) the need to remove nutrients; (3) increased discoveries and understanding of microbial processes and fundamentals; (4) technological advances in equipment, materials, electronics, and process control; and (5) the continual need to reduce capital and operating and energy costs for municipalities and industries. Many activated sludge processes used today and expected to be used in the future may incorporate nitrification, biological nitrogen removal, and/or biological phosphorus removal. Typically, reactors in series, operated under aerobic, anoxic, and anaerobic conditions are used. The general types of activated sludge processes used (i.e., plug flow, complete mix, and sequencing batch reactor), illustrated on Fig. 8–1, are considered in the following discussion.

Plug-Flow Process Configurations. Since the process came into common use in the early 1920s and up until the late 1970s, the type of activated sludge process used most commonly was the one in which a plug-flow reactor with large length to width ratios (typically $>$ 10:1) was used [see Fig. 8–1(a)]. In considering the evolution of the activated sludge process, it is important to note that the discharge of industrial wastes to domestic wastewater collection systems increased in the late 1960s. The use of a plug-flow process became problematic when industrial wastes were introduced because of the toxic effects of some of the discharges.

Complete-Mix Process Configurations. The complete-mix reactor was developed, in part, because the larger volume allowed for greater dilution and thus mitigated the effects of toxic discharges. The more common type of activated sludge process in the 1970s and early 1980s tended to be single-stage, complete-mix activated sludge (CMAS) processes [see Fig. 8–1(b)], as advanced by McKinney (1962). For some nitrification applications, two-stage systems (each stage consisting of an aeration tank and clarifier) were used with the first stage designed for BOD removal, followed by a second stage for nitrification.

Comparing Plug Flow and Complete Mix Process Configurations. In comparing the plug-flow [see Fig. 8–1(a)] and complete-mix activated sludge (CMAS) [see Fig. 8–1(b)] processes, the mixing regimes and tank geometry are quite different. In the CMAS process, the mixing of the tank contents is sufficient so that ideally the concentrations of the mixed-liquor constituents, soluble substances (i.e., COD, BOD, NH_4-N), and colloidal and suspended solids do not vary with location in the aeration basin. The plug-flow process involves relatively long, narrow aeration basins, so that the concentration of soluble substances and colloidal and suspended solids varies along the reactor length. Although process configurations employing long, narrow tanks are commonly referred to as plug-flow processes, in reality, true plug flow does not exist. Depending on the type of aeration system, back mixing of the mixed liquor can occur and, depending on the layout of the reactor and the system reaction kinetics, nominal plug flow may be described more appropriately by the series of complete-mix reactors as discussed in Chap. 4.

Sequencing Batch Process Configuration. With the development of simple inexpensive program logic controllers (PLCs) and the availability of level sensors and automatically operated valves, the sequencing batch reactor (SBR) process [see Fig. 8–1(c)] became used more widely by the late 1970s, especially for smaller communities and industrial installations with intermittent flows. In recent years, however, SBRs are being used for larger cities. The SBR is a fill-and-draw type of reactor system involving a single complete-mix reactor in which all steps of the activated sludge process occur. Mixed liquor remains in the reactor during all cycles, thereby eliminating the need for separate sedimentation tanks.

Other Activated Sludge Processes. Other activated sludge processes that have found application, with their dates of major interest in parentheses, include the oxidation ditch (1950s), contact stabilization (1950s), Krause process (1960s), pure oxygen activated sludge (1970s), Orbal process (1970s), deep shaft aeration (1970s), and sequencing batch reactor process (1980).

Development of Selectors. Activated sludge process designs before and until the late 1970s generally involved the configurations shown on Figs. 8–1(a) and (b). These designs very often suffered from solids settling problems in the secondary clarifiers due to the proliferation of filamentous-type bacteria. In the early 1980s researchers and practicing engineers advanced the concept of a "biological selector" in activated sludge design, which was first introduced in a patent by Davidson (1957) to select for good settling "floc-forming" activated sludge over filamentous bacteria. Selectors are smaller single or multi-staged aerated reactors in front of the main activated sludge treatment aeration basin. The selector concept is also inherent in designs with single or multi-staged anoxic or anaerobic reactors before the main aeration tank to select for conditions for denitrification of nitrate/nitrite or for phosphorus-storing bacteria. The anoxic or anaerobic reactors also serve as selectors resulting in the development of good settling activated sludge. Selectors are considered in detail in Sec. 8–4.

Membrane Bioreactor Process Configuration. A membrane bioreactor (MBR) is an activated sludge system with membranes located at the end of the activated sludge basin(s) for liquid-solids separation in lieu of using secondary clarifiers (see Fig. 8–2). In the integrated MBR system shown on Fig. 8–2 the key component is the microfiltration or ultrafiltration membrane that is immersed directly into the activated sludge reactor. The membranes are mounted in modules (sometimes called cassettes) that can be lowered into the bioreactor. The modules are comprised of the membranes, support structure for the membranes, feed inlet and outlet connections, and an overall support structure. The membranes are subjected to a vacuum (less than 50 kPa) that

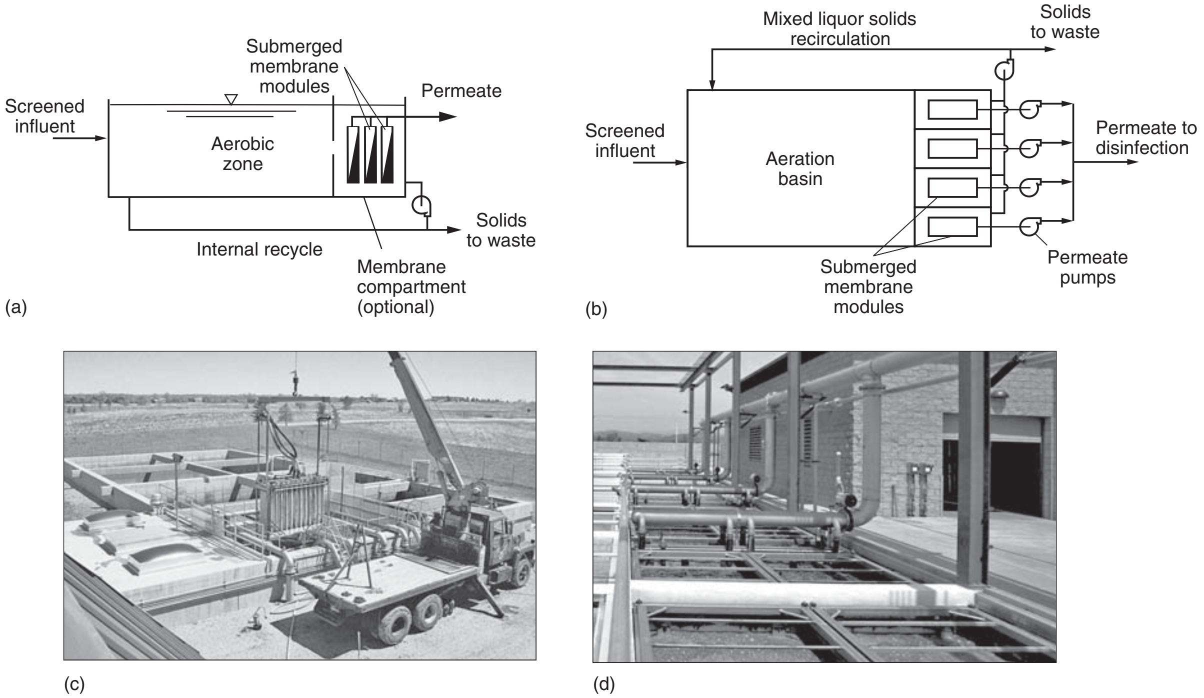


Figure 8–2

Membrane bioreactor (MBR). A multi-staged activated sludge system with membranes for liquid-solids separation: (a) section through MBR with separate compartment for the membranes, (b) plan view of MBR, (c) view of membrane cassettes being placed in separate compartment, and (d) view of separate membrane compartment.

draws water (permeate) through the membrane while retaining solids in the reactor. To minimize the accumulation of solids and fouling on the exterior of the membranes, compressed air is introduced through a distribution manifold at the base of the membrane module. As the air bubbles rise to the surface, scouring of the membrane surface occurs; the air also provides oxygen to maintain aerobic conditions and solids suspension within the reactor.

Implementation of MBR Process. In 1969, U.S. Patent 3,472,765 was issued to William E. Budd and Robert W. Okey of Dorr-Oliver for a process that integrated complete-mix activated sludge process with membrane technology. Membrane separation for activated sludge treatment was first demonstrated at Pikes Peak, CO, in 1974 by Dorr Oliver, but it was not economically feasible for widespread use until membrane materials and manufacturing methods improved. In addition, the initial designs employed cross-flow membrane separation units located outside the activated sludge tanks, which had high energy requirements for pumping mixed liquor across the membrane to control fouling. Placement of the membrane separation unit in the activated sludge reactor and using coarse bubble aeration, in the late 1980s, was less energy intensive and led to future MBR applications (Yamamoto et al., 1989). The lower energy MBR system was first commercialized with flat plate membranes by Kubota for wastewater treatment in Japan in 1990. About three years later an MBR system using Zenon's hollow fiber ZeeWeed® system was installed at Stoney Creek, Ontario Canada. The first MBR installation for biological wastewater treatment in the U.S. was in 1998 at the Arapohoe County Lone Tree Creek WWTP in Colorado. The use of ultrafiltration (UF) and microfiltration (MF) membranes in MBR systems for activated sludge treatment was well accepted in the late 1990s and early 2000s.

MBR Process Advantages and Disadvantages. The use of membrane liquid-solids separation provides many advantages compared to activated sludge processes with gravity clarifiers. These are (1) a much smaller area requirement (less than 50%) due to operation with a higher mixed liquor concentration (typically 8000 to 12,000) and the reduced space for membrane separation compared to clarifiers, (2) simpler process operation with no concerns about the effect of filamentous activated sludge, (3) a reclaimed water quality effluent due to complete suspended solids capture across the membrane separation, and (4) a lower disinfection dose requirement due to the low turbidity effluent. Disadvantages are an increased energy cost, the need for future membrane replacement, and the cleaning and operational demands for membrane fouling control.

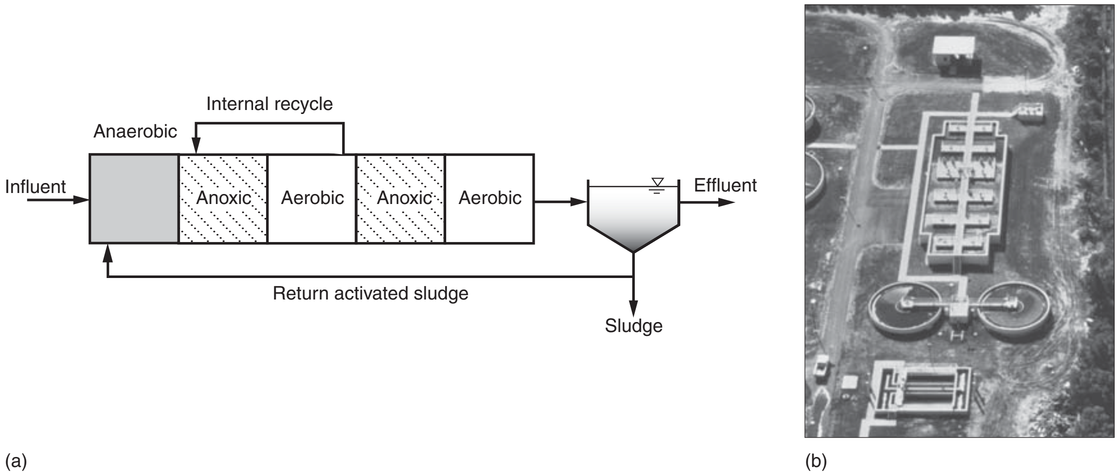
MBR Process Applications. The most frequent applications for MBRs have been for domestic wastewater treatment and water reuse, ranging in flows from small housing developments and apartments to large centralized wastewater treatment facilities. The largest facility, as of 2008, is the King County, Washington Brightwater wastewater treatment plant, at an average design flowrate of 117,000 m^3/d (Judd, 2008a). MBR designs have also been used for industrial wastewater treatment applications, including food and beverage processing, chemical plants, automotive plants, dairy wastewater, oil refinery wastewater, landfill leachate and pharmaceuticals, and also in anaerobic treatment processes (Yang et al., 2006).

Process applications of MBRs are described along with similar applications using activated sludge and clarifiers for nitrification, nitrogen removal and enhanced biological phosphorus removal in Secs. 8–7 and 8–8. Design considerations associated with the use of gravity clarifiers or membranes for activated sludge liquid-solids separation are discussed in Secs. 8–10, 8–11, and 8–12. Membrane systems used in advanced wastewater applications including membrane materials, membrane designs, and operating conditions are discussed in Sec. 11–6 in Chap. 11.

Nutrient Removal Processes

Over the past 10 years, achieving higher nutrient (nitrogen and phosphorus) removals has gained importance in the implementation of the activated sludge treatment process. As a result, a number of biological nutrient-removal configurations have been developed including those using activated sludge with secondary clarifiers or membranes for liquid-solids separation. Nearly all of the various activated sludge process modifications, whether with secondary clarifiers or membranes, are based on the same fundamental principles of biological treatment as described previously in Chap. 7. Processes used in full-scale operation are described in Secs. 8–6, 8–7, and 8–8; design examples for the processes most commonly used are also included.

A key element in the new nutrient removal designs is the use of internal recycle from the aeration or anoxic tanks to upstream reactors (see Fig. 8–3), in contrast to recycle in the past being only from the secondary clarifier underflow to the head of the activated sludge aeration tank as shown on Figs. 8–1(a) and (b). The process efficiency benefits of using reactors in series as well as staged reactors has also been recognized and implemented in full-scale designs. However, because the design and operation of activated sludge nutrient removal processes has become more complex, computer modeling is an increasingly important tool to incorporate the large number of components and reactions necessary to evaluate activated sludge performance in nutrient removal application. The use of simulation models for suspended growth systems is discussed in Sec. 8–5.



(a)

(b)

Figure 8–3

Modified Bardenpho process with stage reactors for biological nitrogen removal and enhanced biological phosphorus removal: (a) schematic diagram of staged process and (b) view of a Modified Bardenpho treatment plant in Palmetto, FL; the first of its type in the United States built in 1979. (From H. D. Stensel.) (Coordinates 27.5256 N 82.5959 W, view at altitude 360 m, since then, an oxidation ditch has been built alongside.)

8–2 WASTEWATER CHARACTERIZATION

Activated sludge process design requires determining (1) the influent characteristics of the wastewater, (2) the aeration tank volume, (3) the sludge production rate, (4) the oxygen supply rate needed, and (5) the effluent concentration of important parameters. To design an activated sludge treatment process properly, characterization of the wastewater is perhaps the most critical step in the process. For biological nutrient-removal processes, wastewater characterization is essential for predicting performance. Wastewater characterization is an important element in the evaluation of existing facilities for optimizing performance and available treatment capacity. Flowrate characterization is also important including diurnal, seasonal, and wet-weather flowrate variations (see Chap. 3). Without comprehensive wastewater characterization, facilities may either be under- or overdesigned, resulting in inadequate or inefficient treatment.

Key Wastewater Constituents for Process Design

Wastewater characteristics of importance in the design of the activated sludge process can be grouped into the following categories: (1) carbonaceous constituents, (2) nitrogenous compounds, (3) phosphorus compounds, (4) total and volatile suspended solids (TSS and VSS), and (5) alkalinity. Typical wastewater constituents quantified for use in desktop designs of wastewater treatment processes are reported in Table 8–1. Desktop designs, based on assuming steady- state operating conditions, are useful to determine reasonable values for the key activated sludge design parameters listed above. However for biological nutrient removal (BNR) processes with multiple reactors in series (including anaerobic, anoxic, and aerobic zones) and internal recycle streams, and/or for process analyses under variable flowrate and load conditions, simulation models with differential equations that can be used to account for all forms of the constituents of concern in the wastewater and biological reactors are most useful. When simulation models are used, the number of wastewater constituents evaluated must be increased. The constituents that must be considered are summarized in Table 8–2 along with the nomenclature used in this textbook and common to the International Water Association (IWA) activated sludge process simulation models (Henze et al., 1995). The following letters represent the state of the wastewater constituent: S is soluble, C is colloidal, X is particulate, and T is total of the individual constituents ($S + C + X$).

Table 8–1

Example of typical domestic wastewater characterization parameters and typical values

Component	Concentration, mg/L[a]
COD	508
sCOD	177
BOD	200
TSS	195
VSS	150
TKN	35
NH_4-N	20
NO_3-N	0
Total phosphorus	5.6
Alkalinity	200 (as $CaCO_3$)

[a] Typical medium strength wastewater, from Table 3–18.

Table 8–2

Definition of terms used to characterize important wastewater constituents used for the analysis and design of biological wastewater treatment processes

Constituent[a,b]	Symbol[c]	Definition
BOD		
BOD		Total 5-d biochemical oxygen demand
sBOD		Soluble 5-d biochemical oxygen demand
UBOD		Ultimate biochemical oxygen demand
COD		
TCOD	COD_T	Total chemical oxygen demand
bCOD		Biodegradable chemical oxygen demand
pCOD		Particulate chemical oxygen demand
sCOD		Soluble chemical oxygen demand
nbCOD		Nonbiodegradable chemical oxygen demand
rbCOD	S_S	Readily biodegradable chemical oxygen demand
bsCOD		Biodegradable soluble chemical oxygen demand
b_{COL}COD	X_{COL}	Biodegradable colloidal chemical oxygen demand
sbCOD	X_S	Slowly biodegradable chemical oxygen demand
bpCOD	X_{SP}	Biodegradable particulate chemical oxygen demand
nbpCOD	X_I	Nonbiodegradable particulate chemical oxygen demand
nbsCOD	S_I	Nonbiodegradable soluble chemical oxygen demand
Nitrogen		
TKN		Total Kjeldahl nitrogen
bTKN		Biodegradable total Kjeldahl nitrogen
sTKN		Soluble (filtered) total Kjeldahl nitrogen
ON		Organic nitrogen
NH_4-N	S_{NH4}	Ammonia nitrogen
bON		Biodegradable organic nitrogen
nbON		Nonbiodegradable organic nitrogen
pON		Particulate organic nitrogen
bpON	X_{NS}	Biodegradable particulate organic nitrogen
nbpON	X_{NI}	Nonbiodegradable particulate organic nitrogen
sON		Soluble organic nitrogen
bsON	S_{NS}	Biodegradable soluble organic nitrogen
nbsON		Nonbiodegradable soluble organic nitrogen
TP		Total phosphorus
PO^4	S_{PO4}	Orthophosphate
bpP	X_P	Biodegradable particulate phosphorus
nbpP	X_{PI}	Nonbiodegradable particulate phosphorus
bsP	S_P	Biodegradable soluble phosphorus
nbsP	S_{PI}	Nonbiodegradable soluble phosphorus
Suspended solids		
TSS		Total suspended solids
VSS		Volatile suspended solids
nbVSS		Nonbiodegradable volatile suspended solids
iTSS		Inert total suspended solids

[a] Note: b = biodegradable; i = inert; n = non; p = particulate; s = soluble.

[b] Measured constituent values, based on the terminology given in this table, will vary depending on the technique used to fractionate a particular constituent.

[c] Commonly used symbol for constituents in IWA activated sludge models.

The subscripts S and I are used to indicate whether the constituent component is biodegradable (S) or nonbiodegradable/inert (I). Other subscripts are used to indicate the specific constituent under S and *X*. For the simulation models the carbonaceous material is quantified in terms of COD. The terms presented in Table 8–2 are introduced, discussed, and applied in the following paragraphs. In the text of this chapter, the units of expression for constituent concentrations are given in mg/L. In the examples, however, constituent concentrations are expressed as g/m^3 (which is equivalent to mg/L) for ease of use in process computations, thus eliminating one unit conversion step.

Carbonaceous Constituents. Carbonaceous constituents measured by BOD or COD analyses are critical to the activated sludge process design. Higher concentrations of degradable COD or BOD result in (1) a larger aeration tank volume, (2) greater required oxygen transfer rates, and (3) greater rates of sludge production. While BOD is the common parameter used to characterize carbonaceous constituents in wastewater, COD is the biodegradable carbonaceous parameter used in most comprehensive computer simulation design models. In these models, a COD mass balance is used to account for the fate of carbonaceous COD material between the amount oxidized, the amount in the effluent, and the amount in waste solids as biomass or nondegraded influent VSS. The various forms of the COD in wastewater are shown on Fig. 8–4 and defined in Table 8–2. The measurement methods and the relative amounts of different forms of COD are illustrated on Fig. 8–5.

COD Fractions. Unlike BOD, some portion of the COD is not biodegradable, so the COD is divided into *biodegradable* and *nonbiodegradable* concentrations. The next level of interest is how much of the COD in each of these categories is *dissolved* (*soluble)*, and how much is *particulate*, comprised of colloidal and suspended solids. The *nonbiodegradable soluble COD* (nbsCOD) will be found in the activated sludge effluent, and nonbiodegradable particulates will contribute to the total sludge production.

Using the nomenclature in Table 8–2, the total COD can be presented as the sum of the wastewater characterization constituents.

$$\text{TCOD} = \text{rbCOD} + \text{sbCOD} + \text{nbsCOD} + \text{nbpCOD} \tag{8–1}$$

$$\text{COD}_T = S_S + X_S + S_I + X_I \tag{8–2}$$

$$X_S = X_{\text{COL}} + X_{\text{SP}} \tag{8–3}$$

Figure 8–4

Fractionation of COD and wastewater. Information on the COD fractions is used in computer simulation models for activated sludge processes.

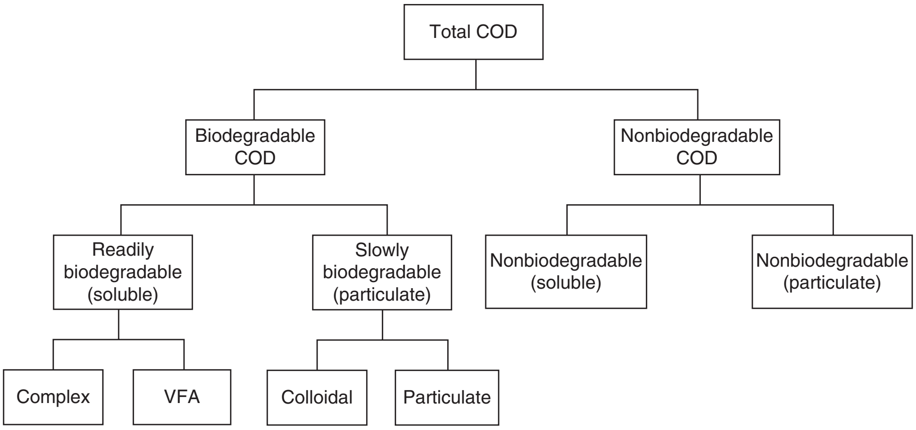
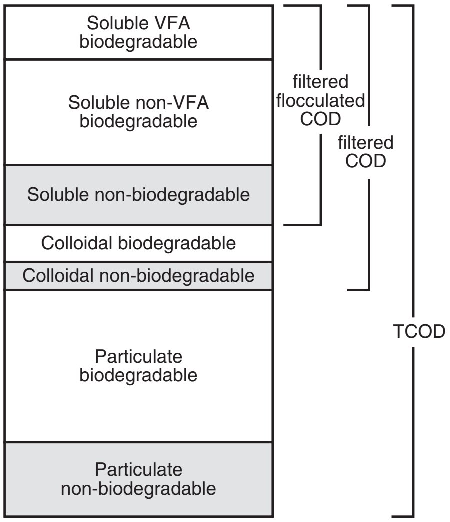


Figure 8–5

Schematic of COD components and separation methods used to obtain components.



Understanding the fractions of biodegradable COD that are measured as soluble readily biodegradable COD (rbCOD), and slowly biodegradable particulate is extremely important for activated sludge process design. The rbCOD portion is assimilated quickly by the biomass, while the particulate and colloidal COD must first be dissolved by extracellular enzymes and are thus assimilated at much slower rates. The rbCOD fraction of the COD has a direct effect on the activated sludge biological kinetics and process performance. Process applications where the rbCOD concentration affects the process design and performance are summarized in Table 8–3.

For conventional plug-flow or staged aerobic activated sludge reactors, a greater oxygen transfer rate will be required toward the front of the aeration tank where there is a greater influent rbCOD concentration. The rbCOD concentration has a significant effect on the denitrification rate in preanoxic zones in biological nitrogen-removal processes,

Table 8–3

Biological processes affected by readily biodegradable COD (rbCOD) concentration in influent wastewater

Process	Effect of rbCOD
Activated sludge aeration	For plug flow or staged aeration zones, there will be a higher oxygen demand toward the front of the tank with higher fraction of rbCOD in the influent COD.
Biological nitrogen removal	For preanoxic tank, there will be a higher denitrification rate with a higher fraction of rbCOD in the influent COD. Can result in smaller anoxic tank volume.
Enhanced biological phosphorus removal	Greater influent rbCOD concentration results in a greater amount of enhanced biological phosphorus removal.
Activated sludge selector	Higher fraction of rbCOD in influent COD provides more COD for floc-forming bacteria in selector. Can have a greater impact on improving sludge volume index (SVI).

where it will be consumed before the aeration zone. The greater the amount of rbCOD, the faster will be the nitrate reduction rate. For enhanced biological phosphorus removal (EBPR), the rbCOD can be converted rapidly to acetate via fermentation in the anaerobic zone for uptake by the phosphorus-storing bacteria. The rbCOD concentration in the influent wastewater must be known to predict more accurately the performance of enhanced biological phosphorus removal.

A further step in the characterization of the influent COD is illustrated on Fig. 8–4. The rbCOD consists of volatile fatty acids and complex soluble COD in the influent wastewater that can be fermented to volatile fatty acids (VFAs). Wastewaters that are more septic, for example, from collection systems in warm climates with minimal slopes, will contain higher concentrations of VFAs. The performance of EBPR processes is improved for wastewaters with higher influent VFA concentrations.

bCOD/BOD Ratio. BOD test data are necessary to obtain the total biodegradable COD (bCOD). Grady et al. (1999) noted that the bCOD/BOD ratio is greater than the ultimate BOD to BOD ratio (UBOD/BOD), because not all of the bCOD is oxidized in the BOD test. Some of the bCOD is converted into biomass, which can still remain as cell debris and active cells at the end of the long incubation time for the UBOD determination. For domestic wastewater with a measured UBOD/BOD ratio of 1.5, the bCOD/BOD ratio may be 1.6 to 1.7, depending on the biomass yield and cell debris fraction. The bCOD/BOD can be estimated using the following equation, which is based on the fact that the bCOD consumed in the BOD test equals the oxygen consumed (UBOD) plus the oxygen equivalent of the remaining cell debris [bCOD = UBOD + $1.42(f_d)(Y_H)$bCOD] after long-term incubation:

$$\frac{\text{bCOD}}{\text{BOD}} = \frac{\text{UBOD/BOD}}{1.0 - 1.42 f_d (Y_H)} \tag{8–4}$$

where f_d = fraction of cell mass remaining as cell debris, g/g
Y_H = synthesis yield coefficient for heterotrophic bacteria, g VSS/g COD used

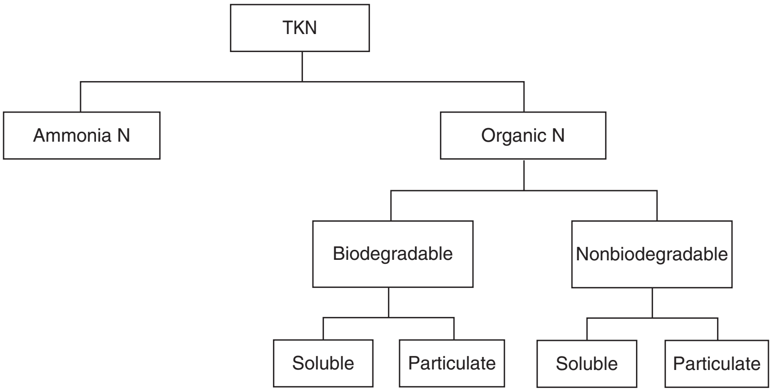
For example, using values typical of domestic wastewater (UBOD/BOD = 1.5, $f_d = 0.15$, $Y_H = 0.40$), the bCOD/BOD ratio is 1.64.

Because the *nonbiodegradable particulate COD* (nbpCOD) is organic material, it will also contribute to the VSS concentration of the wastewater and mixed liquor in the activated sludge process, and is referred to here as the *nonbiodegradable volatile suspended solids* (nbVSS). The influent wastewater will also contain nonvolatile influent suspended solids that add to the MLSS concentration in the activated sludge process. These solids are influent *inert* TSS (iTSS) and can be quantified by the difference in influent wastewater TSS and VSS concentrations.

Nitrogenous Constituents. The composition of nitrogen in wastewater is illustrated on Fig. 8–6. The total Kjeldahl nitrogen (TKN) is a measure of the sum of the ammonia and organic nitrogen. About 60 to 70 percent of the influent TKN concentration will be as NH_4-N, which is readily available for bacterial synthesis and nitrification. Organic nitrogen is present in both soluble and particulate forms, and some portion of each of these is nonbiodegradable. The particulate degradable organic nitrogen will be removed more slowly than the soluble degradable organic nitrogen because a hydrolysis reaction is necessary first. The nonbiodegradable organic nitrogen is assumed to be 6 to 7 percent of the nonbiodegradable VSS as COD in the influent wastewater (Melcer et al., 2003). The particulate nonbiodegradable nitrogen will be captured in the activated sludge floc and exit in the waste sludge, but

Figure 8–6

Fractionation of nitrogen in wastewater. Information on the nitrogen fractions is used in the detailed design of nitrification and denitrification processes.



the soluble nonbiodegradable nitrogen will be found in the secondary clarifier effluent. The soluble nonbiodegradable organic nitrogen contributes to the effluent total nitrogen concentration and typically ranges from 1 to 2 mg/L as N (Parkin and McCarty, 1981; Urgun-Demirtus et al., 2008). Some soluble nonbiodegradable organic nitrogen (0.1 to 0.3 mg/L at SRTs from 8 to 15 d) can be produced from endogenous respiration (Czerwionka et al., 2012).

Alkalinity. The concentration of alkalinity is an important wastewater characteristic that affects the performance of biological nitrification processes. Adequate alkalinity is needed to achieve complete nitrification. In some cases where a wastewater sample is not available, the total alkalinity of the wastewater may be estimated from information on the alkalinity of the potable water plus the alkalinity contributed through domestic use (see Table 3–16).

Measurement Methods for Wastewater Characterization

Special procedures are used to quantify the rbCOD, nbVSS, and soluble organic nitrogen (sON) and nonbiodegradable organic nitrogen (nbON) concentrations in wastewaters. Some of the methods and techniques used to quantify these constituents are discussed below.

Readily Biodegradable COD. The rbCOD concentration can be determined by a batch oxygen uptake rate test as described by Ekama et al. (1986) or a relatively simple chemical-physical test procedure. The latter is the most commonly used method and involves determining what is referred to as a *flocculation-filtration COD* (ffCOD) concentration in an influent wastewater sample. Separation techniques for characterization of wastewater COD are illustrated on Fig. 8–5.

The ffCOD method is based on a procedure presented by Mamais et al. (1993) in an attempt to separate the colloidal and particulate COD from the true soluble COD in the wastewater sample. The ffCOD test is applied to both the wastewater sample and a secondary effluent sample or a settled supernatant sample after sufficient contact and aeration of the wastewater sample with activated sludge. The soluble COD measured in the secondary effluent sample is the nonbiodegradable soluble COD (nbsCOD) as the rbCOD would be removed by the activated sludge process. The floc/filtration method is used widely at wastewater-treatment facilities because of its simplicity. As long as the selected procedure is compatible with the design models used to evaluate the activated sludge process, a useful design approach is possible.

The procedure is based on the assumption that suspended solids and colloidal material can be captured and removed effectively by flocculation with a zinc hydroxide precipitate to leave only truly dissolved organic material after filtration. The steps in the method for each sample are as follows: (1) 1 mL of a 100 g/L $ZnSO_4$ solution is added to 100 mL of sample with vigorous mixing for 1 min, (2) the pH is raised to about 10.5 using 6M (molar) NaOH, with 5 to 10 min of gentle mixing for floc formation, (3) the sample is settled for 10 to 20 min and the supernatant is withdrawn and filtered using a 0.45-μm membrane filter, and (4) the filtrate is analyzed for COD concentration. The difference in COD concentration between the wastewater and activated sludge treated sample is the rbCOD.

Nonbiodegradable Volatile Suspended Solids. The concentration of nonbiodegradable volatile suspended solids (nbVSS) in wastewater can be estimated from analyses for COD, sCOD, BOD, sBOD, and VSS concentration, and by assuming a constant COD/VSS ratio for both biodegradable and nonbiodegradable VSS:

$$\text{nbVSS} = \left[1 - \left(\frac{\text{bpCOD}}{\text{pCOD}}\right)\right]\text{VSS} \qquad (8\text{–}5)$$

$$\frac{\text{bpCOD}}{\text{pCOD}} = \frac{(\text{bCOD/BOD})(\text{BOD} - \text{sBOD})}{\text{COD} - \text{sCOD}} \qquad (8\text{–}6)$$

where bpCOD = concentration of biodegradable particulate COD, mg/L
pCOD = concentration of particulate COD, mg/L
sCOD = concentration of soluble COD, mg/L

Care must be taken in sample handling and analyses to obtain reliable nbVSS concentration data. A sufficient number of composite samples must be obtained to assure that the results are representative of the wastewater characteristics. Samples must be well mixed when taken for analyses, and for small sample volumes, the pipettes must have wide openings at the tip to better capture solids. When small sample volumes are used, for example, with the HACH COD analysis, treating the sample first in a high-speed blender is often done. The filtration pore size to obtain soluble samples for COD and BOD analyses is the same as that used for the TSS/VSS filtration.

A simplified approach that is used more often in lieu of the above procedure is based on the assumption that the g COD per g VSS is the same for the biodegradable and nonbiodegradable VSS. The nbVSS is then determined using the following equations.

$$\text{nbpCOD} = \text{TCOD} - \text{bCOD} - \text{nbsCODe} \qquad (8\text{–}7)$$

$$\text{VSS}_{\text{COD}} = \frac{\text{TCOD} - \text{sCOD}}{\text{VSS}} \qquad (8\text{–}8)$$

$$\text{nbVSS} = \frac{\text{nbpCOD}}{\text{VSS}_{\text{COD}}} \qquad (8\text{–}9)$$

where nbsCODe = filtered COD in activated sludge effluent, mg/L
VSS_{COD} = g COD/ g VSS

Nitrogen Compounds. For the nitrogen compounds, the soluble organic nitrogen concentration is of interest from the standpoint of its effect on the effluent total nitrogen concentration. The fractionation of nitrogen in wastewater is illustrated on Fig. 8–6. A filtered sample from the plant effluent or from a bench-scale treatability reactor can be used to determine the total effluent soluble organic nitrogen concentration by the difference between the TKN concentration of the filtered sample and the effluent NH_4-N concentration. The nonbiodegradable soluble organic nitrogen (nbsON) cannot be determined directly, but from a practical standpoint and considering the low concentration of effluent soluble organic nitrogen, the total effluent soluble organic nitrogen concentration from activated sludge treatment with SRTs between 5 and 10 d usually provides a sufficient estimate.

The *nonbiodegradable particulate organic nitrogen* (nbpON) can be estimated by an analysis of the influent VSS for organic nitrogen and the estimated amount of nbVSS. The fraction of nitrogen in the VSS is as follows.

$$f_N = \frac{(\text{TKN} - \text{sON} - \text{NH}_4\text{-N})}{\text{VSS}} \tag{8–10}$$

$$\text{nbpON} = f_N\,(\text{nbVSS}) \tag{8–11}$$

where f_N = fraction of organic nitrogen in VSS, g N/g VS.
TKN = total TKN concentration, mg/L.
sON = soluble (i.e., filtered) organic nitrogen concentration, mg/L
nbpON = nonbiodegradable particulate organic nitrogen concentration, mg/L
Other terms as defined previously.

Summary Tabulation. In summary, the wastewater COD and nitrogen components can be tabulated as follows:

$$\text{TCOD} = \text{bCOD} + \text{nbCOD} \tag{8–12}$$

$$\text{bCOD} \approx 1.6(\text{BOD}) \tag{8–13}$$

$$\text{nbCOD} = \text{nbsCOD} + \text{npbCOD} \tag{8–14}$$

$$\text{bCOD} = \text{sbCOD} + \text{rbCOD} \tag{8–15}$$

$$\text{TKN} = \text{NH}_4 - \text{N} + \text{ON} \tag{8–16}$$

$$\text{ON} = \text{bON} + \text{nbON} \tag{8–17}$$

$$\text{nbON} = \text{nbsON} + \text{nbpON} \tag{8–18}$$

where the terms are as defined in Table 8–2.

The application of the above equations in determining the characteristics of a wastewater is illustrated in Example 8–1.

Example 8–1 **Wastewater Characterization Evaluation** Given the following wastewater characterization results, determine concentrations for the following:

1. bCOD (biodegradable COD)
2. nbpCOD (nonbiodegradable particulate COD)
3. sbCOD (slowly biodegradable COD)
4. nbVSS (nonbiodegradable VSS)
5. iTSS (inert TSS)
6. nbpON (nonbiodegradable particulate organic nitrogen)
7. Total degradable TKN

Influent wastewater characteristics:

Constituent	Concentration, mg/L
BOD	200
TCOD	420
sCOD	170
rbCOD	80
TSS	220
VSS	200
TKN	40
NH_4-N	26
Alkalinity	200 (as $CaCO_3$)

Activated sludge effluent:

Constituent	Concentration, mg/L
sCODe	30.0
sON	1.2

Solution

1. Determine biodegradable (bCOD) using Eq. (8–13).

$$\text{bCOD} \approx 1.6(\text{BOD})$$

$$= 1.6(200 \text{ mg/L}) = 320 \text{ mg/L}$$

2. Determine the nbpCOD.
   a. Determine the nbCOD using Eq. (8–12).

$$\text{nbCOD} = \text{TCOD} - \text{bCOD}$$

$$\text{nbCOD} = (420 - 320) \text{ mg/L} = 100 \text{ mg/L}$$

   b. Determine the nbpCOD using Eq. (8–14).

$$\text{nbpCOD} = \text{nbCOD} - \text{sCOD}_e$$

$$= (100 - 30) \text{ mg/L} = 70 \text{ mg/L}$$

3. Determine the sbCOD using Eq. (8–15).

$$\text{sbCOD} = \text{bCOD} - \text{rbCOD}$$

$$= (320 - 80) \text{ mg/L} = 240 \text{ mg/L}$$

4. Determine the nbVSS.
   a. Determine the VSS_{COD} ratio using Eq. (8–8).

$$\text{VSS}_{\text{COD}} = \frac{\text{TCOD} - \text{sCOD}}{\text{VSS}}$$

$$\text{VSS}_{\text{COD}} = \frac{420 - 170}{200} = 1.25 \text{ gCOD/gVSS}$$

b. Determine the nbVSS using Eq. (8–9).

$$\text{nbVSS} = \frac{\text{nbpCOD}}{\text{VSS}_{\text{COD}}}$$

$$\text{nbVSS} = \frac{70}{1.25} = 56 \text{ mg/L}$$

5. Determine the inert TSS.

$$\text{iTSS} = \text{TSS} - \text{VSS} = (220 - 200) \text{ mg/L} = 20 \text{ mg/L}$$

6. Determine the nbpON.

a. Determine the organic N content of VSS using Eq. (8–10).

$$f_N = \frac{(\text{TKN} - \text{sON} - \text{NH}_4\text{-N})}{\text{VSS}}$$

$$f_N = \frac{(40 - 1.2 - 26)\text{mg/L}}{200 \text{ mg/L}} = 0.064$$

b. Determine the nbpON using Eq. (8–11).

$$\text{nbpON} = f_N(\text{nbVSS})$$

$$\text{nbpON} = 0.064(90 \text{ mg/L}) = 5.8 \text{ mg/L}$$

7. Determine total degradable TKN.

$$\begin{aligned} \text{bTKN} &= \text{TKN} - \text{nbpON} - \text{nbsON} \\ &= (40 - 5.8 - 1.2) \text{ mg/L} \\ &= 33.0 \text{ mg/L} \end{aligned}$$

Recycle Flows and Loadings

The impact of recycle flows must also be quantified and included in defining the influent wastewater characteristics to the activated sludge process. The possible sources of recycle flows include digester supernatant flows (if settling and decanting are practiced in the digestion operation), recycle of centrate or filtrate from solids dewatering equipment, backwash water from effluent filtration processes, and water from odor-control scrubbers. Depending on the source, a significant BOD, TSS, and NH_4-N load may be added to the influent wastewater. The levels of BOD and TSS concentrations possible for various solids processing unit operations are given in Table 15–1 in Chap. 15.

Compared to untreated wastewater or primary clarifier effluent, the BOD/VSS ratio is often much lower for recycle streams. In addition, a significant NH_4-N load can be returned to the influent wastewater from anaerobic digestion-related processes. Concentrations of NH_4-N in the range of 1000 to 2000 mg/L are possible in centrate or filtrate from the dewatering of anaerobically digested solids. Thus, the ammonia load from a return flow of about one-half percent of the influent flow can increase the influent TKN load to the activated sludge process by 10 to 20 percent. In all cases, a mass balance for flow and important constituents, such as BOD, TSS/VSS, nitrogen compounds, and phosphorus should be done to account for all contributing flows and loads to the activated sludge process. The separate treatment of return flows is considered in Chap. 15.

8–3 FUNDAMENTALS OF PROCESS SELECTION, DESIGN, AND CONTROL

The purpose of this section is to introduce (1) overall considerations in treatment process implementation, (2) important factors in process selection and design, (3) process control issues, (4) operational problems associated with the activated sludge secondary clarifier process, and (5) operational problems associated with the MBR process. The information presented in this section is applied to the analysis and design of alternative activated sludge processes in the remainder of this chapter. Many of the equations presented in this chapter were derived previously in Chap. 7 and are summarized in this section for convenient reference.

Overall Considerations in Treatment Process Implementation

The selection of an activated sludge treatment process is always based on a review a number of local factors that will govern the final choice. The principal factors that must be considered are summarized in Table 8–4. The relative importance of the factors presented in Table 8–4 is site specific. Current and future treatment needs are typically driven by regulatory requirements with regard to the impact of the point discharge to surface or groundwater or reclaimed water quality. Wastewater characteristics were considered previously in Sec. 8–2. Flowrates and their variation are discussed in Chap. 3. Local environmental conditions, space constraints, and costs are site-specific. Energy considerations are discussed in Chap. 17. Important factors in the selection of a specific activated sludge process are discussed in this section.

Important Factors in Process Selection and Design

In the selection and design of the activated sludge process, consideration must be given to (1) the type of activated sludge process and reactor configuration, (2) applicable kinetic relationships, (3) solids retention time and loading, (4) sludge production rate, (5) oxygen demand rate and transfer, (6) nutrient requirements, (7) other chemical requirements, (8) activated sludge settling characteristics, (9) liquid-solids separation of mixed liquor, and (10) effluent characteristics.

Selection of Activated Sludge Process and Reactor Configuration. The many different types of activated sludge processes that can be selected are dependent on treatment needs required to meet effluent discharge limits. In general, the reactor types employed are plug flow, complete-mix, and batch (e.g., sequencing batch reactor). Regardless of the type of reactor or reactor combination used, a critical element in the performance of the various activated sludge processes that employ secondary clarification for liquids-solids separation is the settleability of the mixed liquor suspended solids (MLSS). In turn, the settleability of the MLSS depends on the nature of the microorganisms that comprise the MLSS. Occasionally, a proliferation of filamentous bacteria can occur. When a proliferation of filamentous bacteria occurs, the MLSS biological flocs do not settle well, which can result in a high solids levels in the secondary clarifiers and the loss of solids in the final clarifier overflow. The term *bulking sludge* is used to describe the poor settling sludge. Bulking sludge is considered further in Sec. 8–4 along with another type of bulking, known as *viscous bulking*.

Table 8–4

General considerations for the selection of the type of suspended growth reactor

Factor	Description
Treatment needs	Treatment requirements and process selection can be categorized according to effluent discharge water quality needs, which may range from secondary treatment for BOD removal, nitrification to achieve low effluent ammonia concentration, anoxic-aerobic processes to provide nitrogen removal, and anaerobic-anoxic-aerobic processes to provide nitrogen and phosphorus removal.
Future treatment needs	Potential future treatment needs can have an impact on present process selection. For example, if water reuse is anticipated in the future, the process selection should favor designs that can easily accommodate nitrogen removal and effluent filtration.
Sludge settleability	Activated sludge *selector* designs can be used that control filamentous bacteria growth that leads to poor sludge settling and thickening in secondary clarifiers. Some selector designs are inherent in nitrogen and phosphorus removal processes.
Effect of reaction kinetics	Both completely-mixed and plug flow reactor configurations with similar volumes have been commonly used for BOD removal designs as both require a minimum SRT to provide acceptable sludge settling properties. Staged-reactors or plug flow designs can exploit reaction kinetic advantages for nitrification or preanoxic tanks to result in less volume than that for a single completely-mixed tank. Such designs require that the aeration equipment provides a high enough oxygen transfer rate in the first stage or at the front of a plug-flow tank to meet the oxygen demand for BOD removal and nitrification. The aeration equipment design must account for different oxygen demand rates along the length of the aeration tank. The oxygen demand is less variable and lower in completely-mixed tanks.
Wastewater characteristics	Wastewater characteristics are affected by contributions from domestic and industrial sources and inflow/infiltration flows. Large variations in wastewater concentrations due to wet weather or seasonal loads can affect process selection. Wastewater alkalinity and pH are also important for nitrification and enhance biological phosphorus removal processes.
Local environmental conditions	Temperature is an important environmental condition that affects treatment performance and lower rates occur at lower temperatures. The size of the facility and plant staffing are also important and smaller plants with less staffing favor processes that are simpler to operate and are more robust to influent wastewater variations. Concerns for aesthetics for facilities with close neighbors can affect process selection.
Toxic or inhibitory substances	Industrial pretreatment standards and enforcement provide substantial protection against biological process upsets from toxic or inhibitory substances disposed into the collection system. If potential exists for shock industrial toxic loads, completely-mixed activated sludge processes with greater design safety factors are considered.
Space	Space limitations for new or existing plant retrofits often limits the candidate processes that can be considered. Membrane bioreactors, integrated fixed film activated sludge, and biological aerated filter processes are good candidates for limited space.
Cost	Construction and operating costs are very important considerations in selecting the type and size of biological reactors. Because the associated settling facilities are an integral part of the activated sludge process, the selection of the reactor and the solids separation facilities must be considered as a unit.

Prior to the 1970s, filamentous bulking was considered an inevitable consequence of activated sludge treatment, but work by Chudoba et al. (1973) with staged versus complete-mix activated sludge reactors led to the concept that reactor configuration designs, now termed selectors, could be used to control filamentous bulking and improve sludge-settling characteristics. Because of their impact on the operation of the activated sludge process, the use of selectors is now a common design element of the activated sludge process. Selector types and designs are considered in detail in Sec. 8–4.

Kinetic Relationships. As developed in Chap. 7, kinetic relationships are used to determine biomass growth and substrate utilization rates, and to define process performance. The derivation of important kinetic relationships may be found in Chap. 7 and their application is demonstrated for various designs in this chapter.

Selection of Solids Retention Time and Loading Criteria. Certain design and operating parameters distinguish one activated sludge process from another. The common parameters used are the solids retention time (SRT), the food to biomass (F/M) ratio (also known as food to microorganism ratio), and the volumetric organic loading rate. While the SRT is the basic design and operating parameter, the F/M ratio and volumetric loading rate values are useful for comparison to historical data and typical observed operating conditions. The F/M and volumetric loading parameters are described in Chap. 7.

Solids Retention Time. The SRT, in effect, represents the average period of time during which the sludge has remained in the system. As presented previously in Chap. 7, SRT is the most critical parameter for activated sludge design and operation as SRT affects the treatment process performance, aeration tank volume, sludge production, and oxygen requirements. For BOD removal, SRT values generally range from 3 to 5 d, depending on the mixed-liquor temperature. At 18 to 25°C an SRT value close to 3 d is desired where only BOD removal is required and to discourage nitrification and eliminate the associated oxygen demand. To limit nitrification, some activated sludge plants have been operated at SRT values of 1 d or less. At 10°C, SRT values of 5 to 6 d are common for BOD removal only. Temperature and other factors that affect SRT in various treatment applications are summarized in Table 8–5.

Table 8–5

Typical minimum SRT ranges for activated sludge treatment[a]

Treatment goal	SRT range, d	Factors affecting SRT
Removal of soluble BOD in domestic wastewater	1–2	Temperature
Conversion of particulate organics in domestic wastewater	2–5	Temperature
Develop flocculent biomass for treating domestic wastewater	2–3	Temperature
Provide complete nitrification	3–18	Temperature/inhibitory substances
Biological phosphorus removal	2–4	Temperature
Aerobic digestion of waste activated sludge	20–40	Temperature
Degradation of xenobiotic compounds	5–50	Temperature/specific bacteria/ compounds

[a] SRT is based on aerobic volume.

SRT Values for Nitrification. Because nitrification performance is temperature-dependent, the design SRT for nitrification must be selected with caution as variable nitrification growth rates have been observed at different sites, presumably due to the presence of inhibitory substances (Barker and Dold, 1997; Fillos et al., 2000). For desktop nitrification designs, which are based on constant influent flow and TKN concentration, a safety factor is used to increase the SRT above that calculated from nitrification kinetics and the required effluent NH_4-N concentration. A factor of safety is used for two reasons: (1) to allow flexibility for operational variations in controlling the SRT, and (2) to provide for additional nitrifying bacteria to handle peak TKN loadings. The influent TKN concentration and mass loading can vary throughout the day (a peak to average TKN loading of 1.3 to 1.5 is not unusual, depending on plant size) and can also be affected by return flows from digested and dewatered biosolids processing. By increasing the design SRT, the inventory of nitrifying bacteria is increased to meet the NH_4-N concentration at the peak load so that the effluent NH_4-N concentration requirement is achieved.

Typical SRT Factors of Safety for Nitrification. Typically, the value of the factor of safety is equal to the peak/average TKN load. Because use of the peak/average TKN load is conservative, the NH_4-N concentration during the normal loading period will be lower with the net effect of a composite effluent NH_4-N concentration that is somewhat lower than the design goal. Dynamic simulation models can be used to optimize the design SRT value to meet target effluent NH_4-N concentrations, subject to changing influent flow and TKN concentrations (Barker and Dold, 1997). The steady-state solution approach described in Sec. 8–6 has resulted in reasonable designs, and can provide a starting point for using simulation models to analyze and design activated sludge nitrification processes.

Sludge Production. The design of the sludge-handling and disposal/reuse facility depends on the prediction of sludge production for the activated sludge process. If the sludge-handling facilities are undersized, treatment process performance may be compromised. Sludge will accumulate in the activated sludge process if it cannot be processed fast enough by an undersized sludge-handling facility. Eventually, the sludge inventory capacity of the activated sludge system will be exceeded and excess solids will exit in the secondary clarifier effluent, potentially violating TSS discharge limits. The sludge production relative to the amount of BOD removed also affects the aeration tank size. Two methods are used to determine sludge production as a function of SRT. The first method is based on an estimate of an observed sludge production yield from published data for domestic water, and the second is based on wastewater characterization information with consideration to the various sources of sludge production.

Sludge Production Based on Observed Yield. The use of observed yield is often satisfactory for determining an initial activated sludge process design and for estimating sludge production rate. The quantity of sludge produced daily (and thus wasted daily at steady state) can be estimated using Eq. (8–19). For a given wastewater, the Y_{obs} value will vary depending on whether the substrate is defined as BOD, bCOD, or COD.

$$P_{X,\text{VSS}} = Y_{\text{obs}}(Q)(S_o - S)(1\ \text{kg}/10^3\ \text{g}) \tag{8–19}$$

where $P_{X,\text{VSS}}$ = net waste activated sludge produced each day, kg VSS/d
Y_{obs} = observed yield, g VSS/g substrate removal
Q = influent flowrate, m^3/d
S_o = influent substrate concentration, mg/L
S = effluent substrate concentration, mg/L

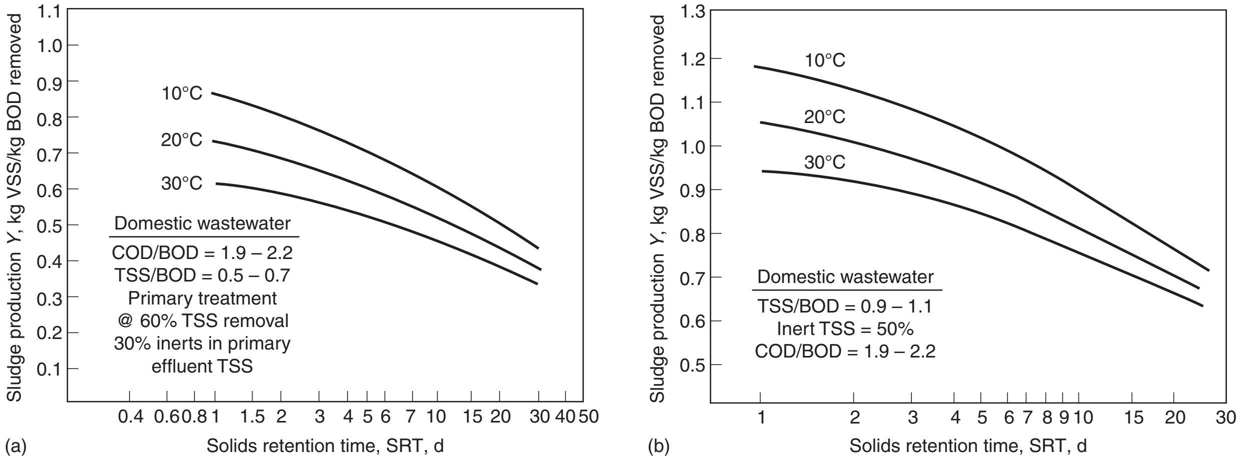


Figure 8–7

Net solids production as a function of solids retention time (SRT) and temperature: (a) with primary treatment and (b) without primary treatment.

Observed volatile suspended solids yield values, based on BOD, are illustrated on Fig. 8–7. The observed yield decreases as the SRT is increased due to biomass loss by more endogenous respiration. The yield is lower with increasing temperature as a result of a higher endogenous respiration rate at higher temperature. The yield is higher when no primary treatment is used as more nbVSS remains in the influent wastewater. The temperature correction value θ for endogenous respiration [see Eq. (2–25)] is 1.04 between 20 and 30°C, and 1.12 between 10 and 20°C. A θ value of 1.04 has been adopted in this text for the temperature effect on endogenous decay.

Sludge Production Based on Wastewater Characteristics. With sufficient wastewater characterization, a more accurate prediction of sludge production can be made. The following equation, based on Eq. (7–54) in Chap. 7, accounts for the heterotrophic biomass growth, cell debris from endogenous decay, nitrifying bacteria biomass, and nonbiodegradable volatile suspended solids and can be used to estimate sludge production. The subscripts H and n are used to distinguish the synthesis yield and decay coefficients between heterotrophic bacteria and nitrifying organisms.

$$P_{X,\mathrm{VSS}} = \underbrace{\frac{QY_{\mathrm{H}}(S_o - S)(1\ \mathrm{kg}/10^3\ \mathrm{g})}{1 + b_{\mathrm{H}}(\mathrm{SRT})}}_{\substack{\text{(A)}\\ \text{Heterotrophic}\\ \text{Biomass}}} + \underbrace{\frac{(f_d)(b_{\mathrm{H}})QY_{\mathrm{H}}(S_o - S)\mathrm{SRT}(1\ \mathrm{kg}/10^3\ \mathrm{g})}{1 + b_{\mathrm{H}}(\mathrm{SRT})}}_{\substack{\text{(B)}\\ \text{Cell}\\ \text{debris}}}$$

$$+ \underbrace{\frac{QY_n(\mathrm{NO}_x)(1\ \mathrm{kg}/10^3\ \mathrm{g})}{1 + b_n(\mathrm{SRT})}}_{\substack{\text{(C)}\\ \text{Nitrifying bacteria}\\ \text{biomass}}} + \underbrace{Q(\mathrm{nbVSS})(1\ \mathrm{kg}/10^3\ \mathrm{g})}_{\substack{\text{(D)}\\ \text{Nonbiodegradable}\\ \text{VSS in influent}}} \tag{8–20}$$

where NO_x = concentration of $\mathrm{NH_4}$-N in the influent flow that is nitrified, mg/L
b_n = endogenous decay coefficient for nitrifying organisms, g VSS/g VSS·d
Other terms as defined previously.

The total mass of dry solids wasted/day includes TSS and not just VSS. The TSS includes the VSS plus inorganic solids. Inorganic solids in the influent wastewater ($\text{TSS}_o - \text{VSS}_o$) contribute to inorganic solids and are an additional solids production term that must be added to Eq. (8–20). The biomass terms in Eq. (8–20) (A, B, and C) contain inorganic solids and the VSS fraction of the total biomass is about 0.85, based on the cell composition given in Table 7–4. Thus, Eq. (8–20) is modified as follows to calculate the solids production in terms of TSS:

$$P_{X,\text{TSS}} = \frac{A}{0.85} + \frac{B}{0.85} + \frac{C}{0.85} + D + \underbrace{Q(\text{TSS}_o - \text{VSS}_o)}_{\substack{\text{(E)} \\ \text{Influent Inert TSS}}} \tag{8–21}$$

where TSS_o = influent wastewater TSS concentration, mg/L
VSS_o = influent wastewater VSS concentration, mg/L

The daily mass of solids in the aeration tanks is determined from the SRT. The daily sludge production can be computed by Eqs. (7–56) and (7–57).

$$(X_{\text{VSS}})(V) = (P_{X,\text{VSS}})\ \text{SRT} \tag{7–56}$$

$$(X_{\text{TSS}})(V) = (P_{X,\text{TSS}})\ \text{SRT} \tag{7–57}$$

By selecting an appropriate MLSS concentration, the aeration volume can be determined using Eq. (7–57). Concentrations of MLSS selected range from 1200 to 4000 mg/L, but must be compatible with the sludge settling characteristics and clarifier design as discussed later in Sec. 8–10.

Oxygen Requirements. The oxygen required for the biodegradation of carbonaceous material is for the amount of bCOD oxidized to provide energy during bCOD consumption for cell synthesis plus the oxygen consumed during the endogenous respiration by the biomass produced. The calculation of the oxygen required for a completely mixed activated sludge system [see Fig. 8–8(a)] involves a simple mass balance on the bCOD removal across the system in which the bCOD removed is equal to the oxygen used plus the

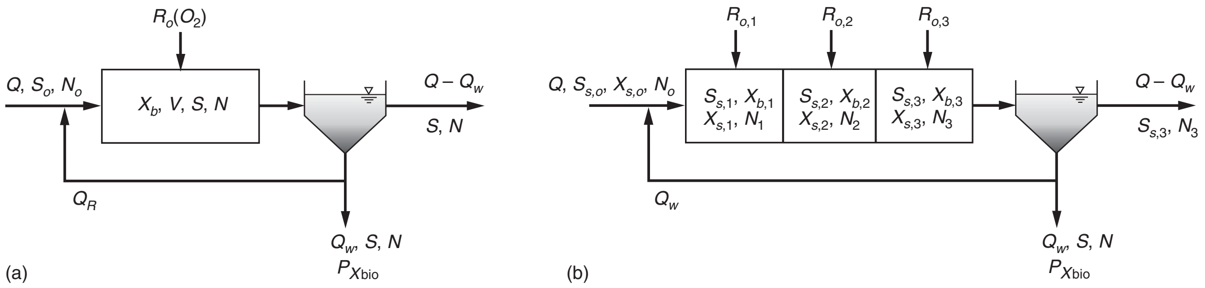


Figure 8–8

Schematic showing components used in oxygen demand rate analysis for completely-mixed activated sludge and staged activated sludge processes: (a) single stage reactor and (b) reactors in series.

bCOD in the excess biomass produced. The biomass produced ($P_{x,\text{bio}}$) is component A plus B in Eq. 8–20 and thus the total oxygen required, R_o, is as follows as shown in Chap. 7:

$$R_o = Q(S_o - S) - 1.42P_{x,\text{bio}} \tag{7–61}$$

$$R_o = Q(S_o - S) - 1.42\left[\frac{QY_{\text{H}}(S_o - S)}{1 + b_{\text{H}}(\text{SRT})} + \frac{f_d\,(b_{\text{H}})QY_{\text{H}}(S_o - S)\text{SRT}}{1 + b_{\text{H}}(\text{SRT})}\right] \tag{8–22}$$

As an approximation, for BOD removal only, the oxygen requirement will vary from 0.90 to 1.3 kg O_2/kg BOD removed for SRTs from 5 to 20 d, respectively (WEF, 2010).

Oxygen Required for Nitrification. When nitrification is included in the process, the total oxygen requirements will include the oxygen required for removal of carbonaceous material plus the oxygen required for ammonia and nitrite oxidation to nitrate (see Sec. 7–9 in Chap. 7) as follows:

$$R_o = Q(S_o - S) - 1.42\,P_{x,\text{bio}} + 4.57\,Q(\text{NO}_x) \tag{8–23}$$

where R_o = total oxygen required, g/d
$P_{X,\text{bio}}$ = biomass as VSS wasted, g/d [parts A, B, and C of Eq. (8–20)]
NO_x = Amount of NO_3-N produced from nitrification of NH_4-N, g/m^3
Other terms as defined previously.

As shown in Eq. (8–23), NO_x is the amount of TKN oxidized to nitrate. A nitrogen mass balance for the system that accounts for the influent TKN, nitrogen removed for biomass synthesis, and unoxidized effluent nitrogen is done to determine NO_x. Unless a careful wastewater characterization study is done to determine the nonbiodegradable particulate and soluble nitrogen (nbpON and nbsON), these components are ignored. Ignoring these terms results in predicting a slightly higher NO_x concentration (5 to 15 percent) and a more conservative oxygen requirement estimate using Eq. (8–23). The nitrogen mass balance, based on the assumption that biomass (which can be represented by the formula $C_5H_7NO_2$) contains 0.12 g N/g biomass, is as follows:

$$\begin{array}{ccccccc} \text{Nitrogen Oxidized} & = & \text{nitrogen in influent} & - & \text{nitrogen in effluent} & - & \text{nitrogen in cell mass} \\ Q(\text{NO}_x) & = & Q(\text{TKN}_o) & - & QN_e & - & 0.12\text{P}_{x,\text{bio}} \end{array}$$

$$\text{NO}_x = \text{TKN} - N_e - 0.12P_{x,\text{bio}}/Q \tag{8–24}$$

where NO_x = nitrogen oxidized, mg/L
TKN_o = influent TKN concentration, mg/L
N_e = effluent NH_4-N concentration, mg/L
Other terms as defined previously.

Equation (8–24) can be solved for the NO_x concentration by estimating the effluent NH_4-N concentration from the nitrification process design.

Oxygen Required for Staged Systems. For activated sludge systems with aerobic reactors in series, the calculation for the oxygen requirement in each reactor is different than above and more complex. The oxygen requirement is highest in the first stage of activated sludge systems with reactors in series [see Fig. 8–8(b)] and decreases in subsequent stages. The amount needed in each stage is a function of (1) the rate of soluble and particulate bCOD removed, (2) the rate of NH_4-N oxidized, and (3) the rate of oxygen used for endogenous respiration. For example, the R_o for reactor 2 shown on Fig. 8–8(b) is as follows:

$$\begin{aligned} R_{o,2} &= (Q + Q_R)(1 - Y_H)[(S_{s,1} - S_{s,2}) + (X_{s,1} - X_{s,2})] \\ &\quad + (Q + Q_R)\,4.57\,(NO_2 - NO_1) + 1.42 b_H (X_{b,2}) V_2 \end{aligned} \tag{8–25}$$

where $R_{o,2}$ = oxygen demand rate in reactor 2, g/d
S_s = soluble bCOD concentration, g/m^3
X_s = particulate bCOD concentration, g/m^3
Y_H = synthesis yield, g biomass COD/g bCOD removed
NO = NO_3-N concentration, g/m^3
X_b = biomass concentration, g VSS/m^3
V = reactor volume, m^3
Q_R = return activated sludge recycle flowrate, m^3/d

The oxygen demand for each stage cannot be determined without first calculating the soluble and particulate bCOD, biomass and the NO_3-N concentration in each stage (assumes that no appreciable NO_2-N is present). These values can be obtained from mass balance equations for each of these constituents, but simulation models, discussed in Sec. 8–4, provide a more efficient solution method.

As an estimate for the design of a 3-stage activated sludge system treating domestic wastewater, Eq.(8–23) can be used to estimate the total oxygen demand rate for the system. Then a distribution of that demand as 60, 25, and 15 percent for reactors 1, 2, and 3, respectively, may be used.

The oxygen demand rate for an activated sludge aeration basin will vary during the day due to diurnal loading changes. The hourly rate can be 1.3 to 1.8 times the rate based on the average daily loading depending on the magnitude of the diurnal changes in influent BOD and TKN concentrations and flowrate.

Nutrient Requirements. If a biological system is to function properly, nutrients must be available in adequate amounts. As discussed in Chaps. 2 and 7, the principal nutrients are nitrogen and phosphorus. Using the formula $C_5H_7NO_2$, for the composition of cell biomass, about 12.4 percent by weight of nitrogen will be required. The phosphorus requirement is 1.5 to 2.0 percent by weight of the cell biomass. These are typical values, not fixed quantities, because it has been shown that the percentage distribution of nitrogen and phosphorus in cell tissue varies with the system SRT and environmental conditions. The amount of nutrients required can be estimated based on the daily biomass production rate [terms A, B, and C in Eq. (8–21)]. It should be noted that nutrient limitations can occur when the concentrations of inorganic nitrogen as N and orthophosphate as P are less than 0.1 mg/L (de Barbadillo et al., 2006). As a general rule, for SRT values greater than 7 d, about 5 g nitrogen and 1 g phosphorus will be required per 100 g of BOD to provide an excess of nutrients.

Other Chemical Requirements. In addition to the nutrient requirements, alkalinity is a major chemical requirement needed for nitrification. The amount of alkalinity required for nitrification, taking into account cell growth, is about 7.14 g $CaCO_3$/g NH_4-N [see Eq. (7–91) in Chap. 7]. In addition to the alkalinity required for nitrification, additional alkalinity must be available to maintain the pH in the range from 6.8 to 7.4. Typically the amount of residual alkalinity required to maintain pH near a neutral point (i.e., pH $\sim$ 7) is between 70 and 80 mg/L as $CaCO_3$, based on the equilibrium relationship between the gas phase CO_2, bicarbonate alkalinity, and pH.

Liquid-Solids Separation of Mixed Liquor. Liquid-solids separation of the activated sludge mixed liquor and return of activated sludge to the activated sludge treatment reactors is critical to the process function and performance. The two methods used for liquid-solids separation are gravity settling in secondary clarifiers and membrane separation. Both methods provide solids thickening for return activated sludge flows and sludge wasting, but differ in the quality of the secondary treatment effluent TSS concentration. With well designed biological nutrient removal facilities secondary clarifier effluent TSS concentrations may range from 4 to 10 mg/L. The effluent from a membrane separation system is termed *permeate* and has no measureable TSS concentration after the liquid passes through the membrane separation unit which have membrane pore sizes of about 0.02 or 0.40 μm, depending on the choice of membranes. Each method has issues with regard to the effect of activated sludge mixed liquor characteristics on performance and design. Specific information on the design of secondary clarifiers and membrane separation is presented in Secs. 8–10 and 8–12, respectively.

Effluent Characteristics. The major parameters of interest that determine the effluent quality from biological treatment processes consist of organic compounds, suspended solids, and nutrients as indicated in Table 8–6. The biodegradable soluble organic concentration is minimal in effluents from systems with complete nitrification and SRT values above 5.0 d, and is likely within the detection limit of the BOD test, which is about 2.0 mg/L. Most of the BOD is in particulate form associated with biomass contained in the effluent VSS concentration.

Table 8–6

Effluent characteristics for biological wastewater treatment processes

Category	Characteristics	Key constituents
Soluble COD	Biodegradable	Remaining influent bsCOD
		Metabolic intermediates
		bsCOD from cell lysis
	Nonbiodegradable	Influent nbsCOD
		Metabolic products
		Nonbiodegradable colloids
Particulate COD	Biodegradable	Biomass VSS
		Non-captured influent VSS
	Nonbiodegradable	Biomass cell debris
		Non-captured influent nbVSS
Nitrogen	Inorganic	NH_4-N, NO_3-N, NO_2-N
	Organic	Dissolved organic N
		Particulate organic N in VSS
Phosphorus	Inorganic	PO_4-P
	Organic	Dissolved organic P
		Particulate organic P in VSS

With a proper secondary clarifier design and good settling sludge, the effluent suspended solids will typically be in the range of 4 to 10 mg/L. Assuming a sBOD of 2.0 mg/L, a VSS/TSS ratio of 0.85, and an effluent TSS of 6 mg/L, the final effluent BOD concentration, BODe, can be estimated as follows:

$$\mathrm{BOD_e} = \mathrm{sBOD} + \left(\frac{0.60 \text{ g BOD}}{\text{g UBOD}}\right)\left(\frac{1.42 \text{ g UBOD}}{\text{g VSS}}\right)\left(\frac{0.85 \text{ g VSS}}{\text{g TSS}}\right)(\text{TSS, mg/L}) \tag{8–26}$$

$$\mathrm{BOD_e} = 2 \text{ mg/L} + (0.60)(1.42)(0.85)(6 \text{ mg/L})$$

$$\mathrm{BOD_e} = 10.2 \text{ mg/L}$$

Because the MBR processes produce an effluent with non-detectable TSS, the BOD concentration is also minimal.

Effluent nitrogen contains inorganic and organic forms. The effluent dissolved organic nitrogen (DON) may range from 0.5 to 2.0 mg/L (Urgun-Demirtas et al., 2007), and thus can be a very significant fraction of the effluent total nitrogen concentration for system confronted with permits with stringent effluent TN concentrations (typically below 3.0 mg/L).

Process Control

To maintain high levels of treatment performance with the activated sludge process under a wide range of operating conditions, special attention must be given to process control. The principal approaches to process control are (1) maintaining a target SRT, (2) maintaining target dissolved oxygen levels in the aeration tanks, and (3) regulating the return activated sludge (RAS) flowrate. The waste activated sludge (WAS) rate is selected to meet the SRT. The SRT is the most common control parameter, but in some cases the WAS rate is adjusted to meet a target MLSS concentration. Thus, higher WAS rates result in a lower SRT and vice versa. Return activated sludge is important in maintaining the MLSS concentration and controlling the sludge blanket level in the secondary clarifier. Aeration tank oxygen uptake rates (OURs) are useful for understanding the process operating conditions and oxygen transfer requirements, and in some cases, have been used in process control algorithms. Routine microscopic observations are important for monitoring the microbial characteristics and for early detection of changes that might negatively impact sludge settling and process performance.

SRT Control. To maintain a given SRT, the excess activated sludge produced each day must be wasted. The most common practice for both activated sludge/secondary clarifier and MBR systems is to waste sludge from the return sludge line because RAS is more concentrated and requires smaller waste sludge pumps than if wasting was done by removing mixed liquor from the aeration tank. In some activated sludge/secondary clarifier applications, such as for smaller flow systems with minimal operating staff, WAS is taken directly from the aeration tank. Though this method requires a larger wasting volume, advantages are that the waste solids concentration is more uniform and the SRT control can be based on volumetric wasting without the need for measuring suspended solids in the aeration tank, effluent and return sludge. The waste sludge may be discharged to various types of thickening processes alone or combined with sludge removed from primary clarification tanks. The actual amount of liquid that must be pumped to achieve SRT process control depends on the method used and the location from which the wasting is to be accomplished.

Wasting from Return Line. For example, if SRT is used for process control and wasting is from the return sludge line, the wasting rate can be computed by modifying the terms of Eq. (7–31) (see Table 8–10).

$$\text{SRT} = \frac{VX}{(Q_W X_R + Q_e X_e)} \tag{8–27}$$

where V = volume of the reactor, m^3
X = aeration tank solids concentration, mg/L
Q_W = waste sludge flowrate from the return sludge line, m^3/d
X_R = concentration of sludge in the return sludge line, mg/L
Q_e = effluent flowrate from the secondary clarifier, m^3/d
X_e = effluent TSS concentration, mg/L

The daily manual wasting flowrate out of the RAS line is then

$$Q_W = \frac{VX}{X_R(\text{SRT})} - \frac{Q_e X_e}{X_R} \tag{8–28}$$

If it is assumed that the concentration of solids in the effluent from the settling tank is low, then Eq. (8–28) reduces to

$$\text{SRT} \approx \frac{VX}{Q_W X_R} \tag{8–29}$$

and

$$Q_W \approx \frac{VX}{X_R(\text{SRT})} \tag{8–30}$$

At higher operating SRT values, the effect of solids loss in the effluent flow on the SRT is less significant. To determine the waste flowrate using Eq. (8–30), the solids concentration in both the aeration tank and the return line must be measured.

Wasting from Aeration Tank. If wasting is done from the aeration tank and the solids in the settled effluent are again neglected, then the rate of pumping can be estimated using the following relationship:

$$\text{SRT} = \frac{V}{Q_W} \tag{8–31}$$

or

$$Q_W = \frac{V}{\text{SRT}} \tag{8–32}$$

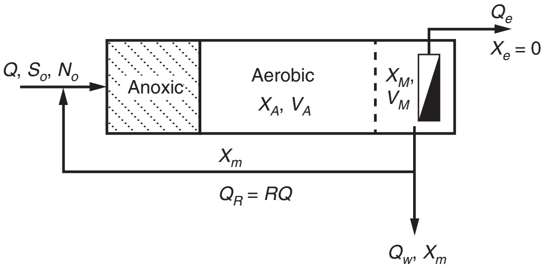
where Q_W = waste sludge flowrate from the aeration tank, m^3/d

Thus, the process may be controlled by daily wasting of a quantity of flow equal to the volume of the aeration tank divided by the SRT.

Wasting from MBR. The WAS flowrate needed to maintain a target aerobic SRT in an MBR system can be set strictly on the system aerobic volumes and return sludge recycle ratio. Such systems typically have an aerobic tank before the aerobic membrane

Figure 8–9

Schematic showing waste activated sludge flow from an anoxic/aerobic MBR.



liquid-solids separation tank (see Fig. 8–9). In some cases the aerobic tank prior to the membrane separation tank is referred to as a preaeration tank. The aerobic SRT is by definition:

$$\mathrm{SRT} = \frac{X_A V_A + X_M V_M}{Q_W X_M} \tag{8–33}$$

where V_A = volume of preaeration tank(s) before membrane tank, m^3
V_M = volume of aerated membrane separation tank, m^3
X_A = solids concentration in preaeration tank, mg/L
X_M = solids concentration in membrane tank, mg/L

The solids concentration is thicker in the membrane tank due to the removal of permeate. The preaeration and membrane tank solids concentration can be related by mass balance, and ignoring the relatively small increase in solids concentration in the flow through the tank due to the removal of influent solids and bsCOD.

$$(RQ)X_M + Q(0) = (Q + RQ)X_A \tag{8–34}$$

and thus

$$X_A = \left(\frac{R}{1 + R}\right)X_M \tag{8–35}$$

substituting for X_A in Eq. (8–33) yields

$$\mathrm{SRT} = \frac{\left(\frac{R}{1 + R}\right)V_A + V_M}{Q_W} \tag{8–36}$$

and

$$Q_W = \frac{\left(\frac{R}{1 + R}\right)V_A + V_M}{\mathrm{SRT}} \tag{8–37}$$

Using Eq. (8–37), the daily WAS flowrate from an MBR system can be set for a given return activated sludge recycle ratio and SRT.

It should be noted that the SRT can be treated as an average value for a system operation and need not be maintained with an exact wasting rate each day. The SRT can be averaged over a time period equal to the target SRT value.

Dissolved Oxygen Control. Aeration equipment must be designed to supply the oxygen demand rate of the microorganisms in the activated sludge system under a wide range of flowrates and loads, while at the same time assuring that some minimum DO concentration is met. In DO control systems, the output of the aerator is adjusted to meet changing demands based on DO concentration measurements in the aeration tank. For example, when the influent BOD and ammonia load drops, the aeration tank DO concentration rises. The control system should be designed to sense DO concentration changes and makes appropriate adjustments to change the aeration rate to lower the DO to its target range so that energy is not wasted at lower loads. At higher loads the DO concentration will start to drop and the aeration output must be increased to keep the desired elevated DO concentration.

When aeration occurs at a lower DO concentration in the aeration basin, less energy is needed because of the higher driving force between the saturated DO concentration and the aeration basin DO concentration. However, if the DO concentration is too low, filamentous organisms may predominate and the settleability and quality of the activated sludge may be poor. In general, the dissolved oxygen concentration in the aeration tank should be maintained at about 1.5 to 2.0 mg/L in all areas of the aeration tank. A minimum DO concentration of about 0.7 mg/L is required to initiate nitrification. Operation at DO concentrations below 1.0 mg/L can save energy and is done in some designs to provide simultaneous nitrification and denitrification. However, the aerobic biological reaction rates are lower, requiring greater tank volume. Higher DO concentrations (2.0 to 3.0 mg/L) provide a small additional increase in nitrification rates. Values of DO above 4 mg/L result in little or no improvement in performance, but do increase aeration costs significantly, and can potentially result in the growth of foaming organisms.

Return Activated Sludge Control. The purpose of the return of activated sludge is to maintain the necessary concentration of activated sludge in the aeration tank and to keep the sludge blanket in the secondary clarifier at a low safe depth. Sufficient return sludge pumping capacity is needed, along with sufficient clarifier depth (3.7 to 6.5 m), to maintain the blanket below the effluent weirs. Return sludge pumping rates of 50 to 75 percent of the average design wastewater flowrate are typical, and the design average capacity is typically 100 to 150 percent of the average design flowrate (recycle ratio of 1.0 to 1.5). The use of variable speed pump drives allows operation between 50 to 150 percent of the influent flowrate.

Impact of SVI on RAS. Higher return activated sludge (RAS) recycle ratios (4.0 to 6.0) are used in MBR systems (Fig. 8–9) to take advantage of the fact that the system can operate at higher MLSS concentrations (8000–12,000 mg/L) than for activated sludge/secondary clarifier systems. Clarifier thickener capabilities typically limit the return sludge concentrations to between 6000 and 12,000 mg/L.

When the activated sludge has good settling properties (see Fig. 8–10) thickening occurs readily in the clarifier and a wide range of RAS ratios are possible, while keeping the clarifier sludge blanket below 0.15 to 0.30 m. Sludges that settle well typically have a sludge volume index (SVI) value equal to or less than 120, although the exact value will vary from plant to plant. For poorer settling sludge, higher recycle rates may be needed. The SVI test is considered in greater detail in Sec. 8–10 which deals with the liquid-solids separation for activated sludge with clarifiers.

Relationship between RAS and MLSS. The relationship between the RAS ratio and aeration tank MLSS concentration can be evaluated by mass balance. A certain MLSS

Figure 8–10

Field test for determining the sludge volume index (SVI).

concentration is necessary for a given aeration tank volume and SRT as shown by Eq. (7–57). The appropriate boundaries for two mass-balance analyses are illustrated on Fig. 8–11. Assuming the sludge-blanket level in the settling tank remains constant and that the solids in the effluent from the settling tank are negligible, the mass balance around the settling tank shown on Fig. 8–11(a) is as follows:

Accumulation = inflow − outflow

$$0 = X(Q + Q_R) - Q_R X_R - Q_W X_R - Q_e X_e \tag{8–38}$$

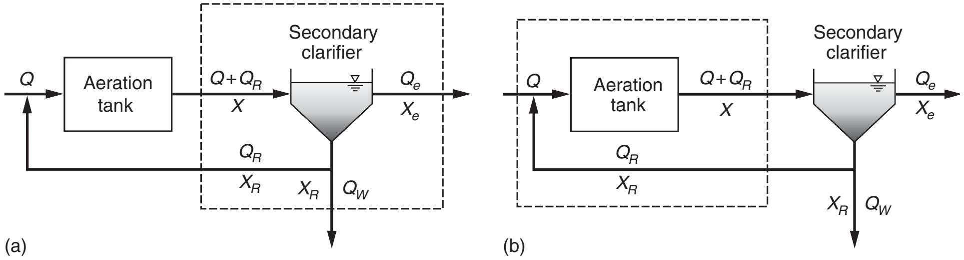
where X = mixed-liquor suspended solids, mg/L
Q = secondary influent flowrate, m^3/s
Q_R = return sludge flowrate, m^3/s
X_R = return activated sludge suspended solids, mg/L
Q_W = waste activated sludge flowrate, m^3/s
Q_e = effluent flowrate, m^3/s
X_e = effluent suspended solids, mg/L

Assuming X_e is negligible and that $Q_W X_R$ is related to the SRT [Eq. (8–28)], solving Eq. (8–38) for Q_R yields

$$Q_R = \frac{[XQ - (XV/\text{SRT})]}{X_R - X} \tag{8–39}$$

Figure 8–11

Definition sketch for suspended solids mass balances for return sludge control: (a) secondary clarifier mass balance and (b) aeration tank mass balance.



The recycle ratio ($Q_R/Q = R$) is then

$$R = \frac{1 - (\tau/\text{SRT})}{(X_R/X) - 1} \tag{8–40}$$

The required RAS pumping rate can also be estimated by performing a mass balance around the aeration tank [see Fig. 8–11(b)]. The solids entering the tank will equal the solids leaving the tank if new cell growth can be considered negligible. At SRT above 8–10 d, this is a reasonable assumption. Solids enter the aeration tank in the return sludge and in the influent to the secondary process. However, if the influent solids are negligible compared to the MLSS, the mass balance around the aeration tank results in the following expression:

Accumulation = inflow − outflow

$$0 = X_R Q_R - X(Q + Q_R) \tag{8–41}$$

Solving for the return activated sludge ratio R yields X

$$Q_R/Q = R = \frac{X}{X_R - X} \tag{8–42}$$

Thus, based on Eqs. (8–40) and Eq. (8–42) for a given RAS solids concentration (X_R), a certain RAS ratio is needed to meet the required aeration tank solids concentration (X) for the target SRT. If the solids do not settle and thicken well in the clarifier, X_R is lower and a higher RAS ratio is needed. The SVI sludge settleability test can be used to approximate X_R for adjusting the RAS ratio:

$$X_R = \frac{1}{\text{SVI}}\left(\frac{1\text{ g}}{1\text{ mL}}\right)\left[\frac{(10^3\text{ mg/1 g})}{(1\text{ L}/10^3\text{ mL})}\right] = \frac{10^6}{\text{SVI}} \tag{8–43}$$

where X_R = estimated RAS concentration, mg/L

Sludge Blanket Level. Sludge blanket depth is a useful operational parameter to assess changes in the settling characteristics. Optimal depths usually range between 0.3 and 0.6 m (1 and 2 ft). The sludge blanket method of control requires considerable operator attention because of the diurnal flow and sludge production variations and changes in the settling characteristics of the sludge. The most common method of determining the blanket level is the use of a long tube core sampler.

Oxygen Uptake Rates. Microorganisms in the activated sludge process use oxygen as they consume the substrate. The rate at which they use oxygen, known as the oxygen uptake rate (OUR), is a measure of the biological activity due to the loading on the aeration tank. Values for the OUR are obtained by performing a series of DO measurements over a period of time with mixing but no aeration, and the measured results are reported conventionally as mg O_2/L·min or mg O_2/L·h. Oxygen uptake is most valuable for plant operations when combined with VSS data. The combination of OUR with MLVSS yields a value termed the specific oxygen uptake rate (SOUR) or respiration rate and is reported as mg O_2/g MLVSS·h. Changes in SOUR values may also be used to assess the presence of toxic or inhibitory substances in the influent wastewater or load changes.

Microscopic Observations. Routine microscopic observations provide valuable monitoring information about the condition of the microbial population in the activated sludge process. Specific information gathered includes changes in floc size and density, the status of filamentous organism growth, the presence of *Nocardioform* bacteria, and the

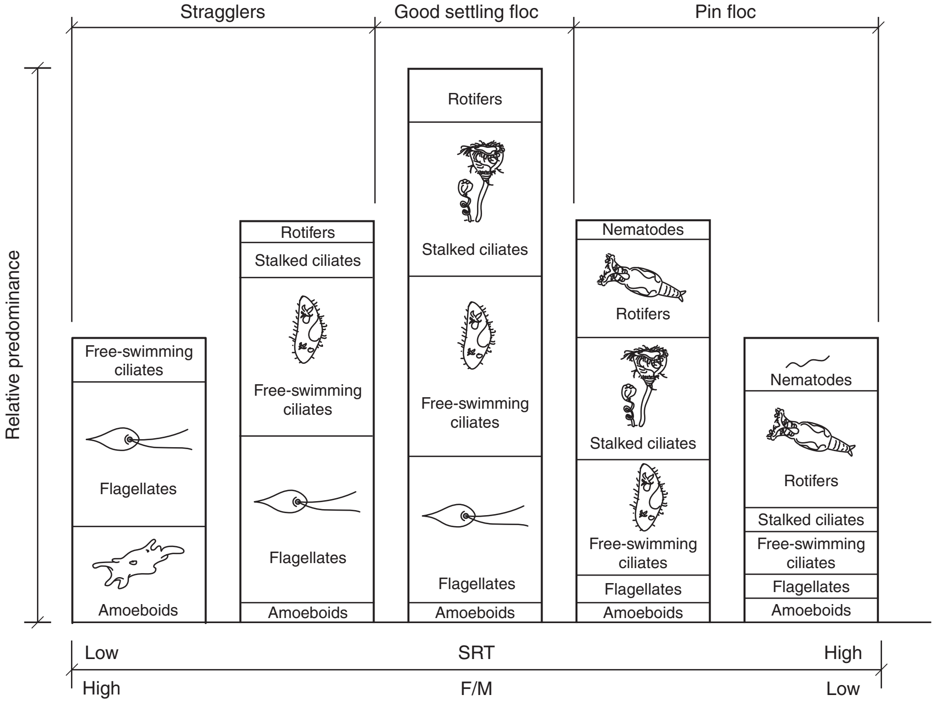


Figure 8–12

Relative predominance of microorganisms versus solids retention time (SRT) and F/M ratio. (Adapted in part from WEF, 1996.)

type and abundance of higher life-forms such as protozoans and rotifers. Changes in these characteristics can provide an indication of changes in the wastewater characteristics or of an operational problem. Examples of the changes in predominance of microorganisms versus F/M ratio and SRT are shown on Fig. 8–12. A decrease in the protozoan population may be indicative of DO limitations, operation at a lower SRT, or inhibitory substances in the wastewater. At very high SRT, small-diameter *pin floc* may develop to lead to a higher effluent suspended solids concentration. Early detection of filamentous or *Nocardioform* growth, as discussed below, will allow time for corrective action to be taken to minimize potential problems associated with excessive growth of these organisms.

Operational Problems in Activated Sludge Systems with Secondary Clarifiers

The most common problems encountered in the operation of an activated sludge/secondary clarifier plant are filamentous bulking sludge, viscous bulking sludge, *Nocardioform foaming,* and rising sludge. Because few plants have escaped these problems, it is appropriate to discuss their nature and possible plant operations responses.

Bulking Sludge. The issue of bulking sludge, which was introduced in Sec. 7–8, is always of concern for activated sludge/secondary clarifier systems. In extreme bulking sludge conditions, the sludge blanket cannot be contained and large quantities of MLSS are carried along with the clarifier effluent, potentially resulting in violation of permit requirements, inadequate disinfection, and clogging of effluent filters. The two principal types of sludge bulking problems are related to *filamentous bulking*, caused by the growth of filamentous organisms, and *viscous bulking*, caused by the presence of an excessive amount of extracellular biopolymer. The occurrence of both types of bulking, along with other forms of filamentous bulking, is considered in the following discussion. However,

Table 8–7

Filamentous bacteria found in activated sludge and associated process conditions[a]

Filament type identified	Cause of filament growth
Sphaerotilus natans, Halsicomenobacter hydrossis, Microthrix parvicella, type 1701.	Low dissolved oxygen concentration
M. parvicella, types 0041, 0092, 0675, 1851	Low F/M
H. hydrossis, Nocardia spp., Nostocoida limicola, S. natans, Thiothrix spp., types 021N, 0914	Complete mix reactor conditions
Beggiatoa, Thiothrix spp., types 021N, 0914	Septic wastewater/sulfide available
S. natans, Thiothrix spp., type 021N, possible *H. hydrossis*, types 0041, 0675	Nutrient deficiency
Fungi	Low pH

[a] From Eikelboom (1975).

before discussing the occurrence of bulking, it will be helpful to consider the general characteristics of filamentous bacteria.

Characteristics of Filamentous Bacteria. In filamentous growth, bacteria form filaments of single-cell organisms that attach end-to-end, and the filaments normally protrude out of the sludge floc. This structure, in contrast to the preferred dense floc with good settling properties, has an increased surface area to mass ratio, which results in poor settling. Methods have been developed for the identification and classification of filamentous bacteria found commonly in activated sludge systems (Eikelboom, 2000). The classification system is based on morphology (size and shape of cells, length and shape of filaments), staining responses, and cell inclusions. Common filamentous organisms are summarized in Table 8–7, along with the operating conditions that favor their growth. Identifying the specific type of filamentous organism which may help identify an operating or design condition that encourages their growth (Jenkins et al., 2004). Examples of good and poor settling with floc containing filamentous bacteria are illustrated on Fig. 8–13.

Occurrence of Filamentous Sludge Bulking. Many types of filamentous bacteria exist, but the types that occur most frequently are related to wastewater characteristics, reactor design limitations, and operational issues. Individual items associated with each of these categories are identified in Table 8–8. Activated sludge reactor operating conditions (low DO, low F/M, and complete-mix operation) clearly have an effect on the development of filamentous populations. One of the kinetic features of filamentous organisms that relates to these conditions is that they are very competitive at low substrate concentrations whether it be organic substrates, DO, or nutrients. Thus, lightly loaded complete-mix activated sludge systems or low DO (<0.5 mg/L) operating conditions provide an environment more favorable to filamentous bacteria than to the desired floc-forming bacteria.

Occurrence of Viscous Sludge Bulking. Another type of bulking that can occur, known as *viscous bulking*, is caused by the presence of an excessive amount of extracellular biopolymer, which results in a sludge with a slimy, jellylike consistency (Wanner, 1994). As the biopolymers are hydrophilic, the activated sludge is highly water-retentive. The resultant sludge has a low density with low settling velocities and poor compaction. Viscous bulking is usually found with nutrient-limited systems or in a very high F/M loading condition with wastewater having a high amount of rbCOD.

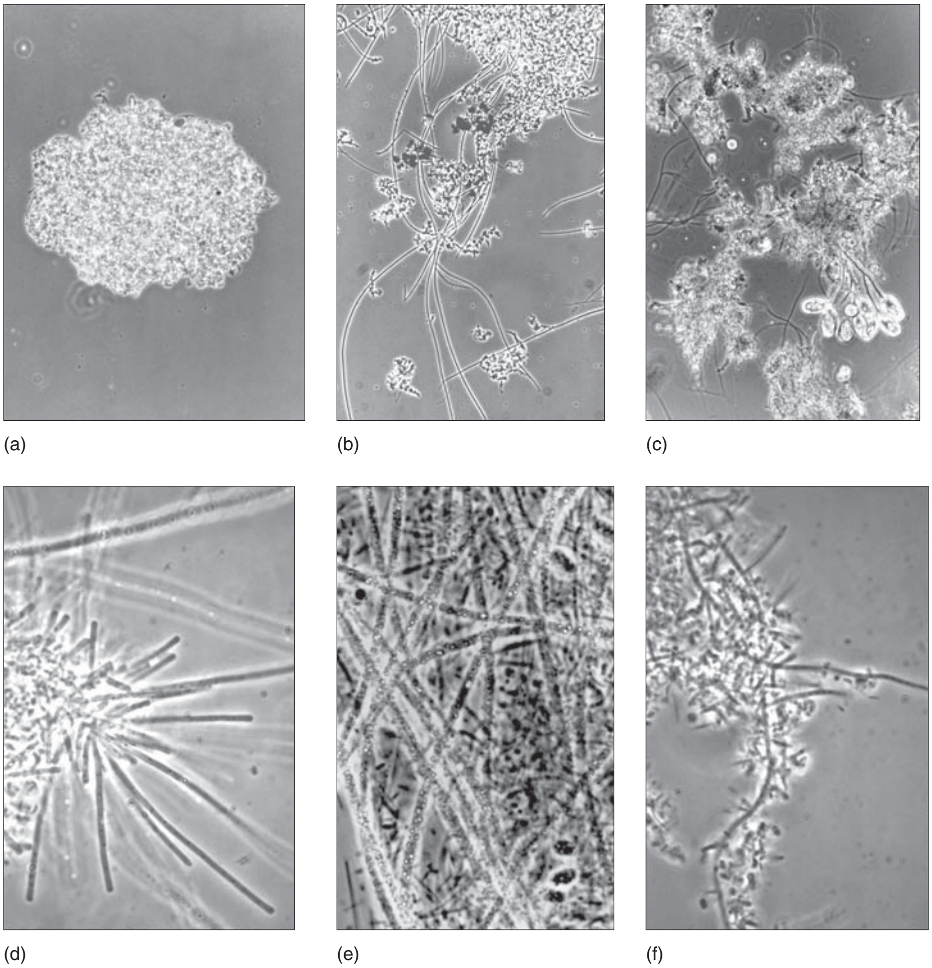


Figure 8–13

Examples of good and poor settling floc particles: (a) nonfilamentous good settling floc, (b) floc particles bridged by filamentous microorganisms, (c) floc particles with limited filamentous microorganisms in secondary form, (d) filaments extending from floc causing poor settling, (e) *Thiothrix* filaments with sulfur granules, and (f) type 1701 filamentous microorganism observed under low-dissolved oxygen conditions. (Courtesy of Dr. David Jenkins, University of California, Berkeley.)

Occurrence of Beggiatoa and Thiothrix Bulking. Filamentous bacteria such as *Beggiatoa* and *Thiothrix* grow well on hydrogen sulfide and reduced substrates, respectively, that would be found in septic wastewaters (Wanner, 1994). When the influent wastewater contains fermentation products such as volatile fatty acids and reduced sulfur compounds (sulfides and thiosulfate), *Thiothrix* can proliferate. Prechlorination of the wastewaters has been done in some cases to prevent their growth. Besides causing bulking problems in activated sludge systems, *Beggiatoa* and *Thiothrix* can create problems in fixed-film systems, including trickling filters and rotating biological contactors.

Control of Bulking Sludges. In the control of bulking, where a number of variables are possible causes, a checklist of items to investigate is valuable. The following items are recommended: (1) wastewater characteristics, (2) dissolved oxygen content, (3) process loading, and (4) internal plant overloading. One of the first steps to be taken when sludge settling characteristics change is to view the mixed liquor under the microscope to determine what type of microbial growth changes or floc structure changes can be related to the development of bulking sludge. A reasonable quality phase-contrast microscope with magnification up to 1000 times (oil immersion) is necessary to view the filamentous bacteria structure and size.

Table 8–8

Treatment plant designs and operational factors that can affect sludge bulking

Factor	Description
Wastewater characteristics	Variations in flowrate
	Variations in composition
	pH
	Temperature
	Septicity
	Nutrient content
	Nature of waste components
Design limitations	Limited air supply
	Poor mixing
	Short circuiting (aeration tanks and clarifiers)
	Clarifier design (sludge collection and removal)
	Limited return sludge pumping capacity
Operational issues	Low dissolved oxygen
	Insufficient nutrients
	Low F/M
	Insufficient soluble BOD

Wastewater Characteristics. The nature of the components found in wastewater or the absence of certain components, such as trace elements, can lead to the development of a bulked sludge (Wood and Tchobanoglous, 1975). If it is known that industrial wastes are being introduced into the system either intermittently or continuously, the quantity of nitrogen and phosphorus in the wastewater should be checked first, because limitations of both or either are known to favor bulking. Nutrient deficiency is a classic problem in the treatment of industrial wastewaters containing high levels of carbonaceous BOD. Highly septic influent wastewaters with high sulfide content can encourage the growth of *Beggiatoa* and *Thiothrix* filamentous bacteria. Wide fluctuations in pH are also known to be detrimental in plants of conventional design. Variations in organic waste loads due to batch-type operations can also lead to DO concentration and bulking problems and should be checked.

Dissolved Oxygen Concentration. Limited dissolved oxygen has often been associated with bulking sludge conditions. If the problem is due to limited oxygen, it can usually be confirmed by operating the aeration equipment at full capacity or by decreasing the system SRT, if possible, to reduce the oxygen demand. The aeration equipment should have adequate capacity to maintain at least 2 mg/L of dissolved oxygen in the aeration tank under normal loading conditions. If 2 mg/L of oxygen cannot be maintained, installation of improvements to the existing aeration system may be required.

Low DO filaments can be generated in anaerobic and anoxic selector zones under periods of low organic loads in high DO in the influent wastewater. A combination of wet weather infiltration/inflow and headworks designs that create cascading flow and aeration can cause this problem.

Internal Plant Overloading. To avoid internal plant overloading, recycle loads should be controlled so they are not returned to the plant flow during times of peak hydraulic and organic loading. Examples of recycle loads are centrate or filtrate from sludge dewatering operations and supernatant from sludge digesters. As a minimum, flow and/or load equalization can be used to mitigate the impacts associated with return flows. The treatment of return flows is considered in Chap. 15.

Temporary Control Measures. In an emergency situation or while the aforementioned factors are being investigated, chlorine and hydrogen peroxide may be used to provide temporary help. Chlorination of return sludge has been practiced quite extensively as a means of controlling bulking. A typical design for a low (5 to 10 h) τ system uses 0.002 to 0.008 kg of chlorine per kg MLSS/d (Jenkins et al., 2004). Although chlorination is effective in controlling bulking caused by filamentous growths, it is ineffective when bulking is due to light floc containing bound water. Chlorination normally results in the production of a turbid effluent until such time as the sludge is free of the filamentous forms. Chlorination of a nitrifying sludge can often result in the loss of nitrification or reduce efficiency due to their slower growth rate compared to heterotrophic bacteria. The use of chlorine also raises issues about the formation of trihalomethanes and other chloroorganic compounds with potential health and environmental effects. Hydrogen peroxide has also been used in the control of filamentous organisms in bulking sludge. Dosage of hydrogen peroxide and treatment time depend on the extent of the filamentous development.

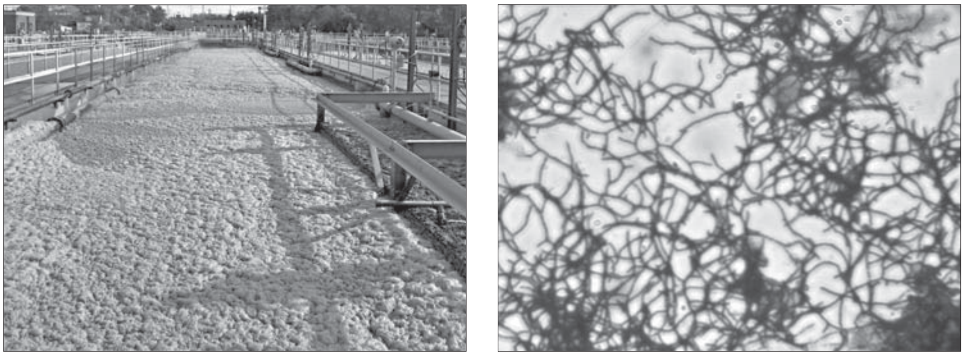
Nocardioform Foam. Two bacteria genera, the *Nocardioforms* type named *Gordonia amarae* and *Candidatus* "*Microthrix parvicella*" are associated with extensive foaming in activated sludge processes. These organisms have hydrophobic cell surfaces and attach to air bubbles, where they stabilize the bubbles to cause foam. The organisms can be found at high concentrations in the foam above the mixed liquor. Both types of bacteria can be identified microscopically.

Characteristics of Nocardioforms. They have a short filamentous structure, which is typically contained within the floc particles. *Microthrix parvicella*, has thin filaments extending from the floc particles. Foaming on an activated sludge basin and a microscopic view of *Nocardioform* foam are shown on Fig. 8–14. The foam is thick, has a brown color, and can build up in thickness of 0.5 to 1 m.

Occurrence of Nocardioform Foam. *Nocardioform* foaming can occur with both diffused and mechanical aeration, but is more pronounced with fine bubble diffused aeration and with higher air flowrates. *Nocardioform* foaming is also a common occurrence in anoxic/aerobic BNR processes. The presence of *Nocardioforms and Microthrix* has also been

Figure 8–14

Nocardioform foam: (a) example of foam on an aeration tank and (b) microscopic observation of gram stain Nocardioform foam filaments.



(a) (b)

associated with fats and edible oils in wastewater. Problems of *Nocardioform* foaming in the activated sludge can also lead to foaming in anaerobic and aerobic digesters that receive the waste-activated sludge. Both *Nocardioform* and *M. parvicella* growth is common where surface scum is trapped in either the aeration basin or secondary clarifiers. Aeration basins that are baffled with flow from one cell to the next occurring under the baffles, instead of over the top, encourage such organism growth and foam collection. By accumulating in the reactor, foam causing organisms gain a competitive advantage of having a longer effective SRT than the other activated sludge organisms.

Control of *Nocardioform* Foam. Methods that can be used to control *Nocardioform* foam include (1) avoiding trapping foam in the secondary treatment process, (2) surface wasting of activated sludge, (3) avoiding the recycle of skimmings into the secondary treatment process, (4) using chlorine spray on the surface of the foam, and (5) reducing the oil and grease content from discharges to the collection system from restaurants, truck stops, and meatpacking facilities. The addition of a small concentration of cationic polymer has been used with some success for controlling *Nocardioforms* foaming (Shao et al., 1997).

Of the above methods, the first three have been found to be most effective. As noted above, submerged weirs can be used in place of underflow baffles to limit the accumulation of foam causing bacteria. Coupled with surface wasting, foam causing bacteria lose the competitive advantage as they are removed from the reactor. The surface wasting strategy for foam control was recommended by Barnard in 1998 and successful surface foam control was reported for a full-scale BNR plant evaluation at the Groos WWTP in Norway (Ydstebo et al., 2000). Parker et al. (2011) reported on the application of this foam control method for a number of activated sludge WWTPs and termed the in-tank design as a *classifying selector*. Biological nitrogen removal processes for New York City have included both surface wasting as a routine sludge wasting method, and chlorine spray hoods across the tank width towards the end of the first pass of a 4-pass plug flow system (Mahoney et al., 2007). The chlorine, which is typically applied at a concentration of 2000 to 3000 mg/L, is effective at killing the foam causing bacteria on the activated sludge surface.

Rising Sludge. Occasionally, sludge that has good settling characteristics will be observed to rise or float to the surface after a relatively short settling period. The most common cause of this phenomenon is denitrification, in which nitrites and nitrates in the wastewater are converted to nitrogen gas. As nitrogen gas is formed in the sludge layer, much of it is trapped in the sludge mass. If enough gas is formed, the sludge mass becomes buoyant and rises or floats to the surface. Rising sludge can be differentiated from bulking sludge by noting the presence of small gas bubbles attached to the floating solids and the presence of more floating sludge on the secondary clarifier surface. Rising sludge is common in short SRT systems, where the temperature encourages the initiation of nitrification, and the mixed liquor is very active due to the low sludge age.

Controlling Rising Sludge. Rising sludge problems may be overcome by (1) increasing the return activated sludge withdrawal rate from the clarifier to reduce the detention time of the sludge in the clarifier, (2) decreasing the rate of flow of aeration liquor into the offending clarifier if the sludge depth cannot be reduced by increasing the return activated sludge withdrawal rate, (3) where possible, increasing the speed of the sludge-collecting mechanism in the settling tanks, and (4) decreasing the SRT to bring the activated sludge out of nitrification. For warm climates where it is very difficult to operate at a low enough SRT to limit nitrification, an anoxic/aerobic process will eliminate the problem of rising sludge and provide a more stable operation.

Operational Problems with MBR Systems

The principal operational problems with MBR treatment processes are foaming and fouling. Each of these is considered briefly below and in Sec. 8–12 which deals with liquid solids separation by membranes.

Foaming. Similar to activated sludge and secondary clarifier systems, *Nocardioform* foaming can occur in MBR systems operated with fine pore diffused aeration. The procedures described for dealing with foaming issues in activated sludge and secondary clarifier systems also apply to MBR systems.

Fouling. MBR systems must be operated in a preventative maintenance mode to avoid operating problems from fouled membranes. The WWTP capacity can be compromised due to the lower flux associated with fouled membrane. Membrane fouling is prevented by employing the cleaning and operating procedures provided by the membrane supplier, maintaining the upstream fine screening equipment, and operating the system within acceptable SRT and mixed liquor concentration limits. Improper screening would allow the accumulation of hair and fibrous material in the membranes, which cannot be removed by the normal membrane cleaning program. A lower SRT of about 8.0 d is normally recommended to prevent excessive fouling due to the release of microbial substances from a younger activated sludge. Excessively long SRTs may result in a higher amount of free bacteria and floc fines to increase fouling rates.

Concentrations of MLSS in the range 8000 to 14,000 mg/L are normally within acceptable operating ranges. Very high MLSS concentrations require a much lower flux to maintain a balance between the amount of solids directed to the membrane surface versus the solids removal rate by the air scour. If excessive MLSS concentrations (>18,000 mg/L) exist under operation at normal design flux values, the membranes can become what is termed "sludged up" and special cleaning methods may be needed to regain the expected operating flux.

Certain wastewater substances must be prevented from entering the treatment facility or MBR system to maintain proper membrane operation. Cooking oils and grease can collect on membrane surfaces and lead to excessive fouling that can only be removed by special membrane cleaning methods.

8–4 SELECTOR TYPES AND DESIGN CONSIDERATION

A selector is a small tank (30 to 60 min contact time) or a series of tanks located before the process aeration tank in which the incoming wastewater is mixed with return sludge under aerobic, anoxic, or anaerobic conditions. The purpose of including a selector as part of the activated sludge process is to create a condition that favors the growth of floc-forming bacteria and suppress the growth of filamentous bacteria that cause sludge bulking as described in the previous section. The use of selector designs in activated sludge is, as noted previously, more common because of the many other advantages, such as nitrogen and phosphorus removal, in addition to improved sludge settling. By improving sludge settling, the activated sludge treatment capacity may be increased, as higher MLSS concentrations are usually possible. The hydraulic capacity of the secondary clarifiers is also increased. The cause of sludge bulking and the types and design considerations for the selectors used for the control of filamentous bacteria are described in this section.

Selector Types and Design Considerations

The concept of a selector involves the use of a specific bioreactor design that favors the growth of floc-forming bacteria instead of filamentous bacteria to provide an activated sludge with better settling and thickening properties. Various types of anaerobic, aerobic and anoxic selectors are shown on Figs. 8–15(a), 8–15(b), and 8–15(c), respectively.

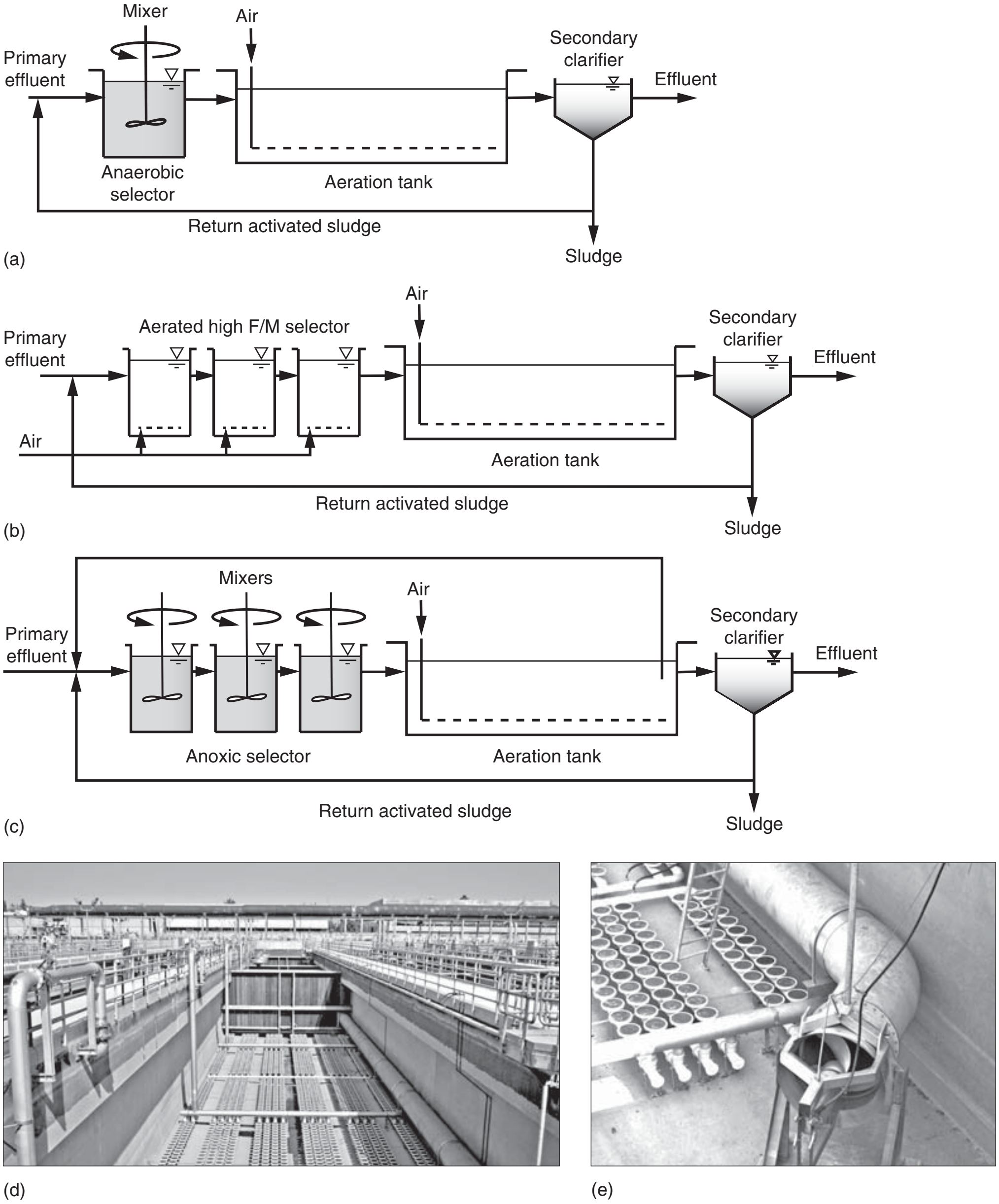


Figure 8–15

Typical selector configurations: (a) anaerobic/aerobic, (b) high F/M, (c) anoxic selector, (d) view of plug-flow reactor (taken from end of aeration tank) with anoxic selectors. The pipe in the bottom right is used to return mixed liquor suspended solids to anoxic selectors as shown in (c) above and (e) axial-flow pump at end of aeration than used to used to pump the return mixed liquor to the anoxic selectors.

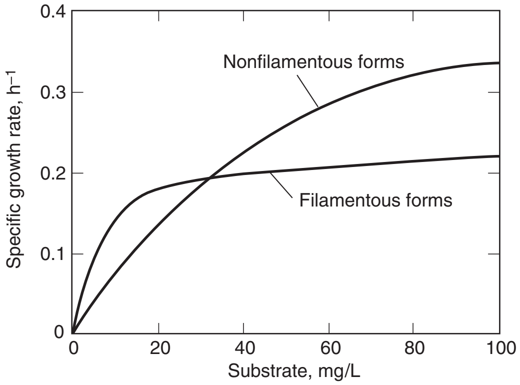


Figure 8–16

Illustration of kinetic base selector model with higher specific growth rate for filamentous bacteria at low substrate concentrations.

The selector reactor precedes the activated sludge aeration tank and may be designed as a separate reaction stage for a complete-mix reactor or as individual compartments in a plug-flow system. Sequencing batch reactors may also be operated to employ the selector concept. The goal in the selector is to have most of the influent rbCOD consumed by the floc-forming bacteria, instead of the filamentous bacteria. Selector designs, as described below, are based on either kinetic or metabolic mechanisms (Albertson, 1987; Jenkins et al., 2004; and Wanner, 1994). The kinetics-based selector designs are called *high F/M selectors*, and the metabolic-based selectors are either anoxic or anaerobic processes.

Kinetics-Based Selector. Selector designs based on a biokinetic model provide for reactor substrate concentrations that result in faster substrate uptake by the floc-forming bacteria. While filamentous bacteria are more efficient for substrate utilization at low substrate concentrations, the floc-forming bacteria have higher growth rates at high soluble substrate concentrations as illustrated on Fig. 8–16. An alternative diffusion mechanism to the kinetic mechanism has been posed and yields the same apparent result on the effect of rbCOD substrate concentration on population selection (Martins et al., 2003). In the diffusion case it is hypothesized that filamentous bacteria selection is related to substrate gradient in the floc and not to differences in biokinetics between filamentous and floc-forming bacteria. At low substrate concentrations the filaments gain an advantage based on their morphology as they can extend from the floc and have easier access to substrate in the bulk liquid. A series of reactors at relatively low τ values (min) is used to provide high soluble substrate concentrations, in contrast to feeding influent wastewater to aeration tanks with τ values on the order of hours.

In kinetic or high F/M selectors, a typical design as shown in Table 8–9 uses at least three reactors in series with the first two of equal volume and the third at twice the volume. The F/M ratio is calculated for the first reactor using the volume and MLSS concentration at that reactor and the influent wastewater flowrate and COD concentration. The F/M

Table 8–9

Design loadings for 3-staged aerobic and anoxic selectors

	Aerobic		Anoxic
Selector zone	F/M, g COD/g MLSS•d	O_2 transfer rate, $gO_2/g•h$	F/M, g COD/g MLSS•d
1	12	15–35	6
2	6	15–35	3
3	3	15–35	1.5

value shown for the second reactor includes the volume of the first and second reactor and the applied loading as the product of the influent flowrate and COD concentration. The amount of oxygen required is about 15 to 25 percent of the soluble COD removal rate (Jenkins et al., 2004) and an oxygen transfer supply rate of 15–20 mg O_2/g mass·h at DO concentrations less than 0.20 mg/L and 30–35 mg O_2/g mass·h at a DO concentration > 1.0 mg/L (Albertson, 1991). The use of selector reactors in series is also advantageous for maximizing rbCOD removal under varying influent flow and loading conditions.

The ability of the high F/M selectors to perform successfully at low or zero DO concentration is due to the fact that a portion of the influent rbCOD is taken up into cell storage due to the elevated substrate concentration (Dionisi et al., 2002). If the loading to the first reactor is too high (F/M $>$ 8 g BOD/g MLSS·d), a viscous, nonfilamentous-type bulking can develop (Albertson, 1987).

A sequencing batch reactor (SBR) can also act as a very effective high F/M selector, depending on the wastewater strength and feeding strategy. For high-strength wastewaters with a relatively large fraction of the SBR volume occupied by the influent wastewater, a high initial F/M ratio can occur. The subsequent reaction by the batch process is equal to that for a plug-flow reactor.

Metabolic-Based Selector. With biological nutrient-removal processes, improved sludge-settling characteristics and minimal filamentous bacteria growth has often been observed. The anoxic or anaerobic metabolic conditions used in these processes favor growth of the floc-forming bacteria. The filamentous bacteria cannot use nitrate or nitrite for an electron acceptor as fast as floc-forming bacteria, thus yielding a significant advantage to denitrifying floc-forming bacteria. Similarly, the filamentous bacteria do not store polyphosphates and thus cannot consume acetate in the anaerobic contact zone in enhanced biological phosphorus removal designs, giving an advantage for substrate uptake and growth to the phosphorus accumulating organisms (PAOs). In some wastewater-treatment facilities (Seattle South Plant and San Francisco, for example), an anaerobic selector has been used for SVI control in low SRT activated sludge systems designed for BOD removal, even though phosphorus removal is not required.

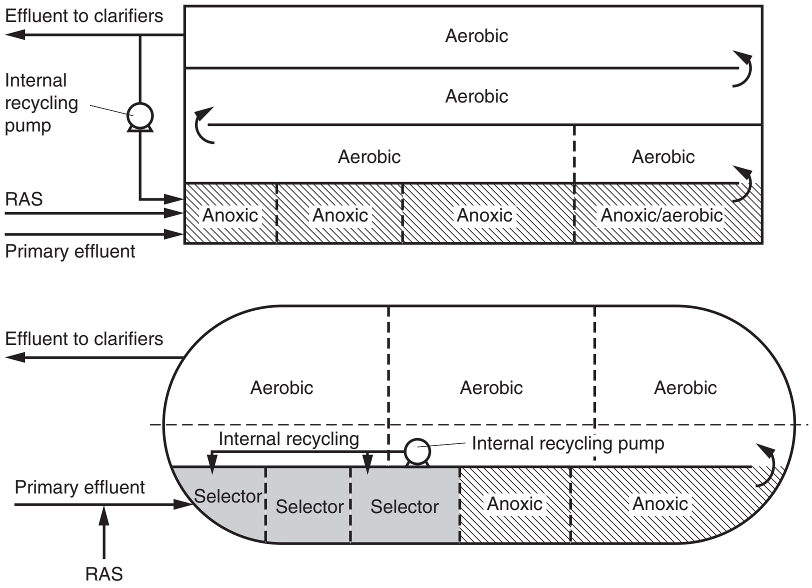
Where nitrification is used and phosphorus removal is not required, staged high F/M gradient or the single-stage anoxic selectors have been used. For the high F/M anoxic or anaerobic selectors, the resultant mixed-liquor SVI may be in the range of 65 to 90 mL/g, and for single-tank anoxic selectors, SVI values in the range of 100 to 120 mL/g are more commonly obtained. Examples of wastewater plant retrofits to anoxic selectors are shown on Fig. 8–17. The use of multiple stages in the aerobic zone in addition to the anoxic zone has been shown to improve sludge settleability with resulting lower SVI values (Albertson 1991, Kruit et al., 2002, and Xin et al., 2008).

Poor Settling Even With Use of Selector

The use of bioselectors has resulted in improved sludge settling characteristics (Albertson 1991, and Parker et al., 2004), but poor settling has also been found on occasion in spite of a bioselector installation. Possible reasons are (1) an inadequate selector design in terms of volume, staging and/or mixing, (2) inadequate aeration or SRT in the downstream aerobic treatment zone, (3) septic influent wastewater with a high sulfide content, and (4) the persistence of certain filamentous organisms that are less impacted by the bioselector design. These include *Microthrix parvicella*, which may be able to take up long chain fatty acids in anoxic zones. *M. parvicella* can be controlled by reducing the aerobic basin SRT, using staged anoxic and aerobic zones, maintaining DO $>$ 2.0 mg/L in the aerobic zones, and eliminating foam-trapping areas (Jenkins et al., 2004).

Figure 8–17

Examples of full-scale wastewater treatment plant modifications to anoxic/aerobic treatment for nitrogen removal with staged anoxic selector zones.



8–5 ACTIVATED SLUDGE PROCESS DESIGN CONSIDERATIONS

Activated sludge design involves performing mass balances on key constituents and the application of fundamental kinetic relationships. These calculations can be done using desktop calculations, spreadsheets and/or computer simulation models. It is common today to use steady-state spreadsheet calculations to aid in the understanding of the process to be designed and to provide a basic starting point in the design to be used in computer simulation models for dynamic process analysis and design optimization. The purpose of this section is to provide a summary of steady-state design approaches and an introduction computer simulation modeling, including common model parameters and reactions and the universal matrix model format that is used to describe reaction components, stoichiometry, and kinetics.

Steady-State Design Approach

The fundamental principles of wastewater characterization, biological treatment and process analysis were presented in Chap. 7 and in Secs. 8–2 and 8–3. The activated sludge system SRT has been presented as a fundamental process parameter that can be used to determine the effluent substrate concentration, sludge wasting rate, and total oxygen demand rate. The principal relationships between SRT and the basic design and performance parameters presented in the previous sections are summarized in Table 8–10, along with equations to determine the aeration tank volume as a function of SRT and MLSS concentration. These equations are appropriate for determining process design and effluent concentrations for steady-state operating conditions at constant flow and influent wastewater constituent concentrations.

Impact of Diurnal Variations in Wastewater Characteristics. As reported in Chap. 3, wastewater flowrates and concentrations are not constant, but vary diurnally. When operating at a given SRT, the effluent substrate concentration will be higher

Table 8–10

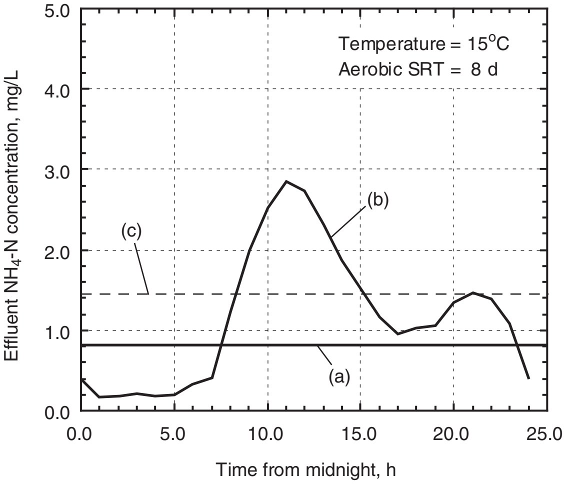
Summary of equations used in the analysis of suspended growth processes

Application	Equation	Eq. No.
Temperature	$k_T = k_{20}\theta^{(T-20)}$	1–44
Rate of sCOD Utilization	$r_{SU} = \dfrac{kXS}{K_S + S}$	7–12
	$\mu_{\max} = Yk$	7–16
Rate of NH_4-N Oxidation	$r_{NH_4} = \left(\dfrac{\mu_{\max,AOB}}{Y_{AOB}}\right)\left(\dfrac{S_{NH_4}}{S_{NH_4} + K_{NH_4}}\right)\left(\dfrac{S_o}{S_o + K_{o,\,AOB}}\right)X_{AOB}$	7–101
Rate of NO_3-N Utilization	$r_{NO_3} = \left(\dfrac{1 - 1.42Y_H}{2.86}\right)\left[\dfrac{\mu_{H,\,\max}S_s}{Y_H(K_s + S_s)}\right]\left(\dfrac{S_{NO_2}}{K_{NO_2} + S_{NO_2}}\right)\left(\dfrac{K'_o}{K'_o + S_o}\right)(\eta)X$	7–133
Specific Growth Rate and SRT	$\mu_{AOB} = \mu_{\max,\,AOB}\left(\dfrac{S_{NH_4}}{S_{NH_4} + K_{NH_4}}\right)\left(\dfrac{S_o}{S_o + K_{o,\,AOB}}\right) - b_{AOB}$	7–94
	$\text{SRT} = \dfrac{1}{\mu_{AOB}}$	7–98
	$\text{SF} = \text{SRT}_{des}/\text{SRT}_{min}$	7–73
Biomass Production, Heterotrophs (VSS)	$P_{X,\,bio} = \dfrac{QY_H(S_o - S)}{1 + b_H(\text{SRT})} + \dfrac{(f_d)(b_H)QY_H(S_o - S)\text{SRT}}{1 + b_H(\text{SRT})}$	8–20 (A+B)
Sludge Production ($P_{x,VSS}$)	$P_{x,VSS} = P_{X,bio} + \dfrac{QY_n(NO_X)}{1 + b_n(SRT)} + Q(\text{nbVSS})$	8–20
Sludge Production ($P_{x,TSS}$)	$P_{x,TSS} = \dfrac{P_{X,bio}}{0.85} + \dfrac{QY_n(NO_X)}{0.85[1 + b_n(\text{SRT})]} + Q(\text{nbVSS}) + Q(\text{TSS}_o - VSS_o)$	8–21
Reactor Mass and Volume	$Mass = X_{VSS}(V) = (P_{X,VSS})\text{SRT}$	7–56
	$Mass = X_{TSS}(V) = (P_{X,TSS})\text{SRT}$	7–57
SRT	$\text{SRT} = \dfrac{VX}{(Q - Q_w)X_e + Q_wX_R}$	8–27
	$\text{SRT} = \dfrac{V}{Q_w}$	8–31
	$\text{SRT} = \dfrac{\left(\dfrac{R}{1 + R}\right)V_A + V_M}{Q_w}$	8–36
CMAS Effluent bsCOD	$S = \dfrac{K_s[1 + b_H(\text{SRT})]}{\text{SRT}(Y_Hk - b_H) - 1}$	7–46
CMAS Biomass	$X = \left(\dfrac{\text{SRT}}{\tau}\right)\left[\dfrac{Y_H(S_o - S)}{1 + b_H(\text{SRT})}\right]$	7–42
CMAS Oxygen Req'd	$R_o = Q(S_o - S) - 1.42P_{x,bio} + 4.57Q(NO_X)$	8–23
Ammonia oxidized	$NO_X = \text{TKN} - N_e - 0.12P_{x,\,bio}/Q$	8–24
Stage Reactor Oxygen Req'd	$R_{o,2} = (Q_1)(1 - Y_H)[(S_{s,1} - S_{s,2}) + (X_{s,1} - X_{s,2})] + (Q_1)4.57(NO_2 - NO_1) + 1.42b_H(X_{b,2})V_2$	8–25
Food to Mass Ratio	$F/M = \dfrac{QS_o}{VX}$	7–62
Organic Loading	$L_{org} = \dfrac{(Q)(S_o)}{(V)}$	7–69

All terms as defined previously.

Figure 8–18

Comparison of effluent NH_4-N concentrations from nitrification activated sludge systems operated at the same temperature and SRT with the same average influent flowrate and TKN loading: (a) effluent NH_4-N concentration with constant influent conditions, (b) effluent NH_4-N concentration versus time with diurnal variations in influent flowrate and concentration, and (c) value of the 24-h composite NH_4-N concentration with diurnal variations in influent flowrate and concentration.



at the higher loading periods and lower at the lower loading periods. An example of the effect of diurnal load variations on effluent NH_4-N concentrations for a nitrification activated sludge process operated with an 8-d aerobic SRT and 15°C temperature is illustrated on Fig. 8–18. The average influent BOD and TKN concentrations used for calculations in this example are 220 mg/L and 35 mg/L, respectively. Assuming constant flow and loadings, the steady state effluent NH_4-N concentration, calculated using Eqs. (7–94) and (7–98), is 0.82 mg/L. The diurnal load was based on the diurnal flowrate and BOD variations for domestic wastewater shown on Fig. 3–11. Using the same kinetic parameters and coefficient values in a dynamic simulation model, the effluent NH_4-N concentration will vary from 0.2 to 2.8 mg/L over the 24 h period, and the flow-weighted composite effluent NH_4-N concentration is 1.45 mg/L. To assure that the daily composite effluent sample substrate concentration is at or below the desired design value, the design SRT that yields the target steady state effluent substrate concentration is multiplied by a design safety factor (typically 1.3 to 1.5 for domestic wastewater treatment) so that a sufficient biomass is available to handle the higher loadings.

Impact of Using Staged Reactors. Use of steady-state equations is satisfactory for the design of completely mixed activated sludge (CMAS) processes, but is not used as easily to determine substrate concentrations and oxygen demand rates in each stage of activated sludge process designs with multiple reactors in series. Staged reactors are common for biological nutrient removal and nitrification process designs. As indicated on Fig. 8–8(b), the bsCOD and pbCOD concentrations vary from stage to stage and accordingly their degradation rates vary. The necessary calculations of process behavior under varying load conditions for single-stage CMAS systems or CSTRs in series, including aerobic, anoxic, and anaerobic conditions, can be accomplished more readily with computer simulation models.

Use of Simulation Models

Simulation models can be used to account for the kinetics and changes in constituent concentrations in each mixed tank of an activated sludge process. Computer modeling provides the tool to incorporate the large number of components and reactions to evaluate activated sludge performance under both dynamic and steady-state conditions, and to easily design multiple-staged processes as well as single-stage complete-mix processes.

Historical Development of Activated Sludge Model. The development of the first comprehensive activated sludge model (ASM1) by the International Water Association (IWA) task group (Henze et al., 1987) was a major step forward in the use of computer simulation models for activated sludge process analysis and design. Limited to carbon oxidation, nitrification and denitrification, the ASM1 model was followed by the ASM2 and ASM2d models, which included fermentation, enhanced biological phosphorus removal, and chemical phosphorus removal (Henze et al., 1995, Barker and Dold, 1997). Further model structure modifications have been made in ASM3 (Gujer et al., 1999, and Henze et al., 2000). Commercial software packages that include these models are used commonly by wastewater process engineers for the process design of different activated sludge system configurations.

General Model Characteristics. The components included in activated sludge simulation models are termed *state variables.* The principal components are summarized in Table 8–11 along with brief descriptions of the types of reactions causing their production or depletion. Activated sludge process models used today are much more complex and now include BOD removal, nitrification, denitrification, and phosphorus removal. The reactions involved in these processes are carried out by different types of bacteria that include a mixture of heterotrophic bacteria that are phosphorus-storing and non-phosphorus-storing, and among these, bacteria that can and cannot use nitrate as an electron acceptor, as well as autotrophic nitrifying bacteria. The impact of wastewater components such as fermentable soluble COD, acetate, biodegradable particulate COD, inert particulate and soluble COD, inorganic nitrogen, and soluble phosphorus on oxygen consumption and sludge production rates are now included.

Important Model Features. The models are based on growth as opposed to substrate utilization. Monod specific growth rate kinetics are used to model the growth of autotrophic or heterotrophic bacteria. Substrate, oxygen, and nutrient utilization rates are related to the growth rates by stoichiometric factors. Another important feature of the models is that COD is used as the common measure of organic substrate and biomass so that a COD balance exists for substrate utilization, biomass growth, and oxygen consumption. The models also use the lysis-regrowth model for endogenous respiration instead of a net endogenous decay coefficient as presented in this chapter for activated sludge design. In the lysis-regrowth model, endogenous decay results in the release of biomass particulate material, some of which is biodegradable and is hydrolyzed to provide a source of rbCOD. Another portion remains as cell debris, similar to what has been incorporated for endogenous respiration in the activated sludge design model presented in this chapter. To compute the same amount of sludge production as the endogenous respiration model, the decay coefficient value in the lysis-regrowth model is higher.

Inclusion of Nitrite. The ASM1, ASM2, ASM2d, and ASM3 models do not include nitrite as a state variable, because most of the ammonia oxidized in many activated sludge processes is complete to nitrate with very little nitrite present. However, nitrite is included in many simulation models today because it may represent an important fraction of ammonia oxidation under certain process conditions, including higher temperatures (>25°C), the initial stages of a multi-staged nitrification system, and low dissolved oxygen concentration.

Matrix Model Format. A long list of complex equations would be needed to describe the various reactions in an activated sludge process involving numerous components such as organic substrates (soluble and particulate), inorganic substrates (ammonia, nitrate, and

Table 8–11

Key process components in ASM2d and types of reactions affecting their reactor concentration

		Reactions or input	
Model component	**Symbol**	**Production or input**	**Depletion**
Dissolved O_2	S_{O_2}	• Influent wastewater • Aeration	• Consumption by X_H, X_{AUT}, X_{PAO}
rbCOD	S_F	• Influent wastewater • Hydrolysis of X_S	• Biodegradation by X_H • Fermentation by X_H
Acetate	S_A	• Influent WWT • Fermentation of S_F	• Uptake by X_{PAO} • Biodegradation by X_H
Ammonia	S_{NH_4}	• Influent wastewater • Hydrolysis of organic N • Hydrolysis of cell decay products	• Oxidation by autotrophic bacteria (X_{AUT}) • Synthesis uptake by X_H, X_{AUT}, X_{PAO}
Nitrate	S_{NO_3}	• Oxidation of S_{NH_4} by X_{AUT}	• Synthesis
Phosphorus	S_{PO_4}	• Influent wastewater • Hydrolysis of organics	• Synthesis uptake by X_H, X_{AUT}, X_{PAO} • Anoxic and aerobic uptake by X_{PAO}
Alkalinity	S_{ALK}	• Influent wastewater • During biological reduction of S_{NO_3}	• During S_{NH4} oxidation by X_{AUT}
Biodegradable particulate COD	X_I	• Influent wastewater • Cell decay	
Slowly biodegradable COD	X_S	• Influent wastewater • Cell decay	• Hydrolysis by X_H
Ordinary heterotrophs	X_H	• Growth from S_F, S_A	• Cell decay
Phosphorus accumulating heterotrophs	X_{PAO}	• Growth from using X_{PHA}	• Cell decay
Stored PHA	X_{PHA}	• Production in anaerobic zone by X_{PAO} from S_A	• Biodegradation by X_{PAO} in aerobic and anoxic zones
Stored polyphosphate	X_{PP}	• During oxidation of X_{PHA}	• Release in anaerobic conditions by X_{PAO}
Ammonia oxidizing bacteria	X_{AUT}	• Growth during S_{NH_4} oxidation	• Cell decay

phosphorus), dissolved oxygen, and various heterotrophic and autotrophic bacteria. Instead of presenting the model in terms of numerous equations, a more general matrix model approach has been adopted. Process reactions and the stoichiometric factors that link the components to the various reactions are presented in a matrix model format. The advantage of the matrix format is that a relatively simple, concise format can be used to describe the process. The purpose of this section is to provide a basic introduction to the matrix model approach showing the components, reactions, and stoichiometric coefficients for the activated sludge process model and how the matrix format can be used to describe the process. In addition, the approach can also be used to illustrate how the matrix model can be interpreted to describe a complete set of equations for a given process component. For example, the Activated Sludge Model No. 2 (ASM2d) (Henze et al., 1995) is used to describe basic features of a comprehensive activated sludge model.

Model Matrix Format, Components, and Reactions

A convenient matrix format is used to describe the model without having to present the large number of equations involved. The ASM2d model includes 19 components and accounts for 21 process reactions. Some of the component and reaction terms are described here to illustrate the basic model format.

Process Reactions and Stoichiometric Coefficients. Fourteen key components related to the biological processes are described in Table 8–11. An example of some of the process reactions and the corresponding stoichiometric rate coefficients are given in Tables 8–12 and 8–13, respectively. The stoichiometric coefficients are used to relate

Table 8–12

Example of process rate equations selected from ASM2d

i[a]	*i*Process	Process rate equations, r_i
Hydrolysis Processes		
1	Aerobic Hydrolysis	$K_h\left(\frac{S_{O_2}}{K_{O_2}+S_{O_2}}\right)\left(\frac{X_S/X_H}{K_X+X_S/X_H}\right)X_H$
2	Anoxic Hydrolysis	$K_h\left(\frac{K'_{O_2}}{K'_{O_2}+S_{O_2}}\right)\left(\frac{S_{NO_3}}{K_{NO_3}+S_{NO_3}}\right)\left(\frac{X_S/X_H}{K_X+X_S/X_H}\right)(\eta_{NO_3})X_H$
Heterotrophic Organisms, X_H		
4	Aerobic Growth on S_F	$\mu_H\left(\frac{S_{O_2}}{K_{O_2}+S_{O_2}}\right)\left(\frac{S_F}{K_F+S_F}\right)\left(\frac{S_F}{S_A+S_F}\right)(\text{Growth}_{\text{Lim}})(X_H)$
5	Aerobic Growth on S_A	$\mu_H\left(\frac{S_{O_2}}{K_{O_2}+S_{O_2}}\right)\left(\frac{S_A}{K_A+S_A}\right)\left(\frac{S_A}{S_A+S_F}\right)(\text{Growth}_{\text{Lim}})(X_H)$
6	Anoxic Growth on S_F	$\mu_H(\eta_{NO_3})\left(\frac{K'_{O_2}}{K'_{O_2}+S_{O_2}}\right)\left(\frac{S_{NO_3}}{K_{NO_3}+S_{NO_3}}\right)\left(\frac{S_F}{K_F+S_F}\right)\left(\frac{S_F}{S_A+S_F}\right)(\text{Growth}_{\text{Lim}})(X_H)$
7	Anoxic Growth on S_A	$\mu_H(\eta_{NO_3})\left(\frac{K'_{O_2}}{K'_{O_1}+S_{O_2}}\right)\left(\frac{S_{NO_3}}{K_{NO_3}+S_{NO_3}}\right)\left(\frac{S_A}{K_A+S_A}\right)\left(\frac{S_A}{S_A+S_F}\right)(\text{Growth}_{\text{Lim}})(X_H)$
8	$\text{Growth}_{\text{Lim}}$ Equation	$\left(\frac{S_{NH_4}}{K_{NH_4}+S_{NH_4}}\right)\left(\frac{S_{PO_4}}{K_{PO_4}+S_{PO_4}}\right)\left(\frac{S_{ALK}}{K_{ALK}+S_{ALK}}\right)$
9	Cell Lysis	$b_H(X_H)$
Ammonia Oxidizing Bacteria, X_{AUT}		
18	Aerobic Growth on S_{NH_4}	$\mu_{AUT}\left(\frac{S_{O_2}}{K_{O_2}+S_{O_2}}\right)\left(\frac{S_{NH_4}}{K_{NH_4}+S_{NH_4}}\right)\left(\frac{S_{PO_4}}{K_{PO_4}+S_{PO_4}}\right)\left(\frac{S_{ALK}}{K_{ALK}+S_{ALK}}\right)(X_{AUT})$
19	Cell Lysis	$b_{AUT}(X_{AUT})$

[a] *i* = ASM2d equation number.

Table 8–13
Stoichiometric matrix for selected components in ASM2d

		Component									
i	***i*Process**	S_F	S_A	S_I	S_{NH_4}	S_{O2}	S_{NO_3}	X_S	X_I	X_H	X_{AUT}
Hydrolysis processes											
1	Aerobic hydrolysis	1						–1			
2	Anoxic hydrolysis	1						–1			
Heterotrophic organisms, X_H											
4	Aerobic growth on S_F	$-\frac{1}{Y_H}$			$\frac{-i_N}{X_H}$	$1-\frac{1}{Y_H}$				1	
5	Aerobic growth on S_A		$-\frac{1}{Y_H}$		$\frac{-i_N}{X_H}$	$1-\frac{1}{Y_H}$				1	
6	Anoxic growth on S_F	$-\frac{1}{Y_H}$			$\frac{-i_N}{X_H}$		$-\frac{(1-Y_H)}{2.86\,Y_H}$			1	
7	Anoxic growth on S_A		$-\frac{1}{Y_H}$		$\frac{-i_N}{X_H}$		$-\frac{(1-Y_H)}{2.86\,Y_H}$			1	
9	Cell lysis							$1-f_{XI}$	f_{XI}	–1	
Phosphorus accumulating organisms											
Ammonia oxidizing bacteria, X_{AUT}											
18	Aerobic growth on S_{NH_4}				$-\frac{i_N}{X_{AUT}}-\frac{1}{Y_{AUT}}$	$\frac{4.57-Y_{AUT}}{Y_{AUT}}$					1
19	Cell lysis							$1-f_{XI}$	f_{XI}		–1

Simultaneous precipitation of phosphorus with ferric hydroxide.

changes in model components to the growth rate. The process reactions in ASM2d are divided into the following five groups dealing with:

- Hydrolysis processes
- Heterotrophic organisms (including aerobic oxidation, denitrification, fermentation, and cell lysis)
- Phosphorus accumulation organisms
- Ammonia oxidizing organisms
- Simultaneous precipitation of phosphorus with ferric hydroxide

For simplicity, the process equations related to enhanced biological phosphorus removal including a description of PHA and phosphorus storage under anaerobic, anoxic, and aerobic conditions; phosphorus accumulating organism (X_{PAO}) growth under anoxic and aerobic conditions; and X_{PAO} cell lysis and release of phosphorus and PHA are not shown in Table 8–12.

For example, as given in Table 8–12, the change in the heterotrophic bacteria (X_H) concentration is related to process reactions 4, 5, 6, 7, and 9. The term η represents the fraction of heterotrophic bacteria that can use nitrate in place of dissolved oxygen.

The corresponding stoichiometric rate coefficients for these biomass growth reactions are 1, 1, 1, 1 as given in Table 8–13. For cell decay, the death-lysis model is used, which is given by process reactions 9 and 19 for heterotrophic and autotrophic bacteria, respectively (see Table 8–12). The cell debris material produced during biomass decay is indicated by X_I and the production of degradable particulate substrate (X_s) is indicated by a stoichiometric coefficient ($1 - f_{XI}$). The hydrolysis of the particulate organic substrates under either aerobic or anoxic conditions is described by processes 1 and 2, respectively.

Rate Expressions. The rate expression for each component for a specific reaction is the product of the stoichiometric coefficient given in Table 8–13 and the corresponding process reaction given in Table 8–12. An equation for each component (e.g., S_F, S_A, etc.) can be written by summing the product of the stoichiometric coefficients given in Table 8–13 for that term times the corresponding rate expressions given in Table 8–12. The summation rate equation is:,

$$R_C = \sum_{i=1}^{n} C_i(j_i) \tag{8–44}$$

where R_C = rate of change for the process component C (e.g., S_F, S_A, X_I, X_S, etc.)
C_i = stoichiometric coefficient for component C, reaction i
j_i = rate reaction i

For example, for component S_A under aerobic growth (reaction 5, Table 8–12), the stoichiometric coefficient from Table 8–13 is $-(1/Y_H)$. Thus, under aerobic conditions, the rate of change of S_A is given by the following expression:

$$R_{S_A} = -\frac{1}{Y_H}\mu_H\left(\frac{S_{O_2}}{K_{O_2} + S_{O_2}}\right)\left(\frac{S_A}{K_A + S_A}\right)\left(\frac{S_A}{S_A + S_F}\right)(\text{Growth}_{\text{Lim}})(X_H) \tag{8–45}$$

$$\text{where Growth}_{\text{Lim}} = \left(\frac{S_{NH_4}}{K_{NH_4} + S_{NH_4}}\right)\left(\frac{S_{PO_4}}{K_{PO_4} + S_{PO_4}}\right)\left(\frac{S_{ALK}}{K_{ALK} + S_{ALK}}\right)$$

The stoichiometric coefficients for oxygen consumption are described as follows: for heterotrophic growth, the term $(1 - Y_H)$ is the g O_2 used/g COD removed. The term $(1 - Y_H)$ is divided by Y_H (g cell COD/g COD used) to obtain the stoichiometric coefficient as g O_2/g cell COD produced to fit to the matrix format. The stoichiometric term for autotrophic growth contains the factor 4.57. The term is required because ammonia, the substrate for the nitrifying bacteria, is expressed as nitrogen in the matrix S_{NH}, and oxygen is expressed as COD. The oxygen equivalent for ammonia is 4.57 g O_2/g NH_4-N. The amount in the numerator is lower by Y_A, which accounts for the ammonia used in cell synthesis. Application of Eqs. (8–44) and (8–45) is demonstrated in Example 8–2. It is intended to illustrate how the model matrix is interpreted to describe the rate of concentration change of a model component.

Example 8–2 **Apply ASM2d Model Matrix** Use the ASM2d model matrix to describe the concentration of the readily biodegradable COD component, S_F, due to nonphosphorus accumulating heterotrophic bacteria in reactor number 2 of the staged reactor shown on Fig 8-8(b). Substitute $S_F + S_A$ for S_s in figure for the rbCOD components. For simplicity in demonstrating the use of the matrix information the PAO reactions are not included here.

Solution

1. Develop mass balance equation for S_F and S_A for reactor 2.
   a. The general word statement is

      Rate of change = rate in − rate out + rate of production + rate of depletion

   b. Write the symbolic representation.
      i. The incoming and outgoing mass of S_F with the flow is $(Q + Q_R)\,S_{F,1}$, and $(Q + Q_R)\,S_{F,2}$ expressed in g/d, respectively.
      ii. The rate of production of S_F due to aerobic and anoxic hydrolysis, given by reactions 1 and 2 in Table 8–12, can be represented as R_1 and R_2, respectively.
      iii. The rate of depletion of S_F due to aerobic and anoxic growth, given by reactions 4 and 6 in Table 8–12, can be represented as R_4 and R_6, respectively.
      iv. Substituting the terms given above in the mass balance equation yields the following expression

$$V_2\frac{dS_{F2}}{dt} = (Q + Q_R)S_{F,1} - (Q + Q_R)S_{F,2} + R_1V_2 + R_2V_2 + R_6V_2$$

2. Write the mass balance equation with the appropriate rate expressions including the process reactions from Table 8–12 and the corresponding stoichiometric coefficients from Table 8–13. For example, the rate expression for aerobic hydrolysis, R_1, for component S_F (rbCOD) is:

$$R_1 = 1 \times K_h\left(\frac{S_{\mathrm{O_2}}}{K_{\mathrm{O_2}} + S_{\mathrm{O_2}}}\right)\left[\frac{X_{\mathrm{S},2}/X_{\mathrm{H},2}}{K_X + (X_{\mathrm{S},2}/X_{\mathrm{H},2})}\right]X_{\mathrm{H},2}$$

   Thus, the mass balance S_A for the reactor 2 can be expressed as:

$$\begin{aligned}
V_2\frac{dS_{\mathrm{F},2}}{dt} &= (Q + Q_R)S_{\mathrm{F},1} - (Q + Q_R)S_{\mathrm{F},2} \\
&\quad + K_h\left(\frac{S_{\mathrm{O_2},2}}{K_{\mathrm{O_2}} + S_{\mathrm{O_2},2}}\right)\left[\frac{X_{\mathrm{S},2}/X_{\mathrm{H},2}}{K_X + (X_{\mathrm{S},2}/X_{\mathrm{H},2})}\right]X_{\mathrm{H},2}(V_2) \\
&\quad + K_h\left(\frac{K'_{\mathrm{O_2}}}{K'_{\mathrm{O_2}} + S_{\mathrm{O_2}}}\right)\left(\frac{S_{\mathrm{NO_3}}}{K_{\mathrm{NO_3}} + S_{\mathrm{NO_3}}}\right)\left[\frac{X_{\mathrm{S},2}/X_{\mathrm{H},2}}{K_X + (X_{\mathrm{S},2}/X_{\mathrm{H},2})}\right](\eta_{\mathrm{NO_3}})X_{\mathrm{H},2}(V_2) \\
&\quad + \left(\frac{-1}{Y_H}\right)\mu_H\left(\frac{S_{\mathrm{O_2},2}}{K_{\mathrm{O_2}} + S_{\mathrm{O_2},2}}\right)\left(\frac{S_{\mathrm{F},2}}{K_{\mathrm{F}} + S_{\mathrm{F},2}}\right)\left(\frac{S_{\mathrm{F},2}}{S_{\mathrm{A},2} + S_{\mathrm{F},2}}\right)(\mathrm{Growth}_{\mathrm{Lim},2})(X_{\mathrm{H},2})(V_2) \\
&\quad + \left(\frac{-1}{Y_{\mathrm{H}}}\right)\mu_H(\eta_{\mathrm{NO_3}})\left(\frac{K'_{\mathrm{O_2}}}{K'_{\mathrm{O_2}} + S_{\mathrm{O_2},2}}\right)\left(\frac{S_{\mathrm{NO_2},2}}{K_{\mathrm{NO_3}} + S_{\mathrm{NO_3},2}}\right)\left(\frac{S_{\mathrm{F},2}}{K_{\mathrm{F}} + S_{\mathrm{F},2}}\right)\left(\frac{S_{\mathrm{F},2}}{S_{\mathrm{A},2} + S_{\mathrm{F},2}}\right) \\
&\quad (\mathrm{Growth}_{\mathrm{Lim},2})(X_{\mathrm{H},2})(V_2)
\end{aligned}$$

Comment The solution for S_{F} depends on the concentrations of (1) non-PAO heterotrophic bacteria, (2) slowly biodegradable COD, $\mathrm{X_s}$, (3) dissolved oxygen, (4) nitrate-nitrogen, (5) acetate COD, S_{A}, (6) ammonia-nitrogen, (7) phosphate, and (8) alkalinity. Similar sets of equations can be provided for these components using the matrix information shown in Table 8–13.

Other Simulation Model Applications

In addition to process design and analysis, models may be used as (1) a research tool to evaluate biological processes and to better understand important parameters that affect a certain type of performance, and (2) as a means to evaluate the treatment capacity of a given facility. For item 2, the availability of accurate and representative wastewater characterization data is critical where dynamic simulations are used assess the effect of variable flow and concentrations with time.

Evaluation of Plant Capacity. To evaluate the capacity of an existing plant, the model is calibrated using wastewater characterization and plant performance data. Calibrations based only on comparing the model predictions to the plant effluent concentration are not necessarily valid. Low effluent concentrations normally exist for all soluble degradable components, and thus the ability of the model to predict the plant performance is clouded by analytical accuracy and practical considerations. Intermediate soluble substrate concentrations from aerobic stages and/or anoxic and anaerobic stages provide a more reliable indication of the ability to describe the kinetics for the site. Oxygen uptake rate data are more meaningful for model calibration, as the data reflect different factors in the model including kinetic rates for different reactions and stoichiometric ratios for cell yield and decay. Oxygen uptake rate data for staged systems are valuable for model calibration. Nitrification rates and sludge production rates are also important parameters for model calibration.

Use of Model Default Values. Default values for 45 kinetic parameters in ASM2d have been selected and are summarized in the model report (Henze et al., 1995). Using these values does not ensure that the model can be used to predict the performance of an activated sludge process accurately, as some of the coefficient values can be different at different sites. One of the parameters found to vary the most at different sites and often adjusted during model calibration is the maximum specific growth rate of nitrifying bacteria μ_{AUT}. Variations in nitrification kinetics may be due to differences in wastewater characteristics, population selection, and inhibitors on nitrification or may also reflect other kinetic changes for which μ_{AUT} adjustments improve the overall fit.

Evaluation of Different Process Configuration. Process design engineers can take advantage of the powerful tool provided by simulation models to study various design configurations and possible operating conditions on expected system performance. Because the selection of process reaction equations and coefficient values used in different commercial software packages varies, the design engineer must be familiar with all model assumptions and model structure. Simple desktop designs, as will be presented in the next three sections, can be used to obtain reactor sizing and configuration for use in the simulation models and to also gain a sense for the expected results as a function of SRT, number of stages, DO conditions, and recycle streams. At a minimum the designer should be comfortable with the oxygen demand rates and sludge production rates predicted by the simulation models relative to alternate calculation methods.

Recent Simulation Models. The model presented in this section is consistent with ASM2d, but the latest commercial computer simulation models apply additional process models that are not included in ASM2d or ASM3. For example, ammonium oxidation to nitrite and nitrite oxidation to nitrate is not separated in the activated sludge models discussed above. In most commercial computer simulation programs, these two reactions are treated separately to describe more accurately the nitrogen removal processes, and also allow a modeling of nitritation/denitritation. Models for deammonification have also been developed and used by computer modeling programs.

8–6 PROCESSES FOR BOD REMOVAL AND NITRIFICATION

Important considerations for the design of activated sludge processes were presented in Secs. 8–3, 8–4, and 8–5. The purpose of this section is to illustrate in detail the design procedure for three common but different activated sludge processes for BOD removal and nitrification. The objective of the three different activated sludge process design examples presented in this section is to demonstrate the application of the fundamental principles discussed previously for BOD removal and nitrification processes and to provide insight into their behavior along with key design features. The material in this section is organized into the following topics: (1) overview of BOD removal and nitrification processes, (2) general process design considerations, (3), (4), (5) process designs for three different activated sludge processes, and (6) a summary of alternative processes used for BOD removal and nitrification, typical process design parameters, and process selection considerations. Biological nitrogen and phosphorus removal can be incorporated into most of the processes used for BOD removal and nitrification, but because additional design factors must be considered, nitrogen and phosphorus removal processes are covered separately in Secs. 8–7 and 8–8, respectively. Details for the selection and design of physical facilities may be found in Secs. 8–9, 8–11, and 8–12.

Overview of BOD Removal and Nitrification Processes

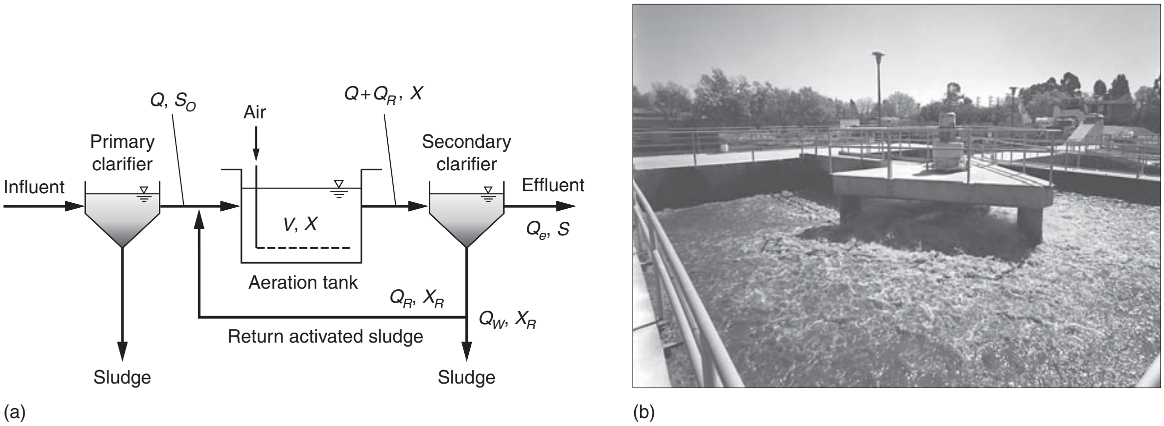
All of the considerations involved in BOD removal and nitrification process design can be applied to a BOD removal-only design by modifying the SRT and removing items that deal with nitrification. The design methodology presented is based fundamentally on using appropriate SRT values, and thus can be applied to the broad range of processes described at the end of this section. As an introduction to BOD removal and nitrification, descriptions of the following three process configurations that are used most commonly are presented below.

1. A complete-mix activated sludge process without and with nitrification
2. A sequencing batch reactor (SBR) with nitrification
3. A multi-staged nitrification process.

Each of these process configurations is considered in the following discussion.

Complete-Mix Activated Sludge Process. A typical complete-mix activated sludge (CMAS) process is shown on Fig. 8–19. Effluent from the primary sedimentation tank and recycled return activated sludge are introduced typically at several points in the reactor. Because the tank contents are mixed thoroughly, the organic load, oxygen demand, and substrate concentration are uniform throughout the entire aeration tank and the F/M ratio is low. Care should be taken to assure that the contents of CMAS reactor are well mixed and that influent feed and effluent withdrawal points are selected to prevent short-circuiting of untreated or partially treated wastewater. The complete-mix reactor is usually configured in square, rectangular, or round shapes. Tank dimensions depend mainly on the size, type, and mixing pattern of the aeration equipment.

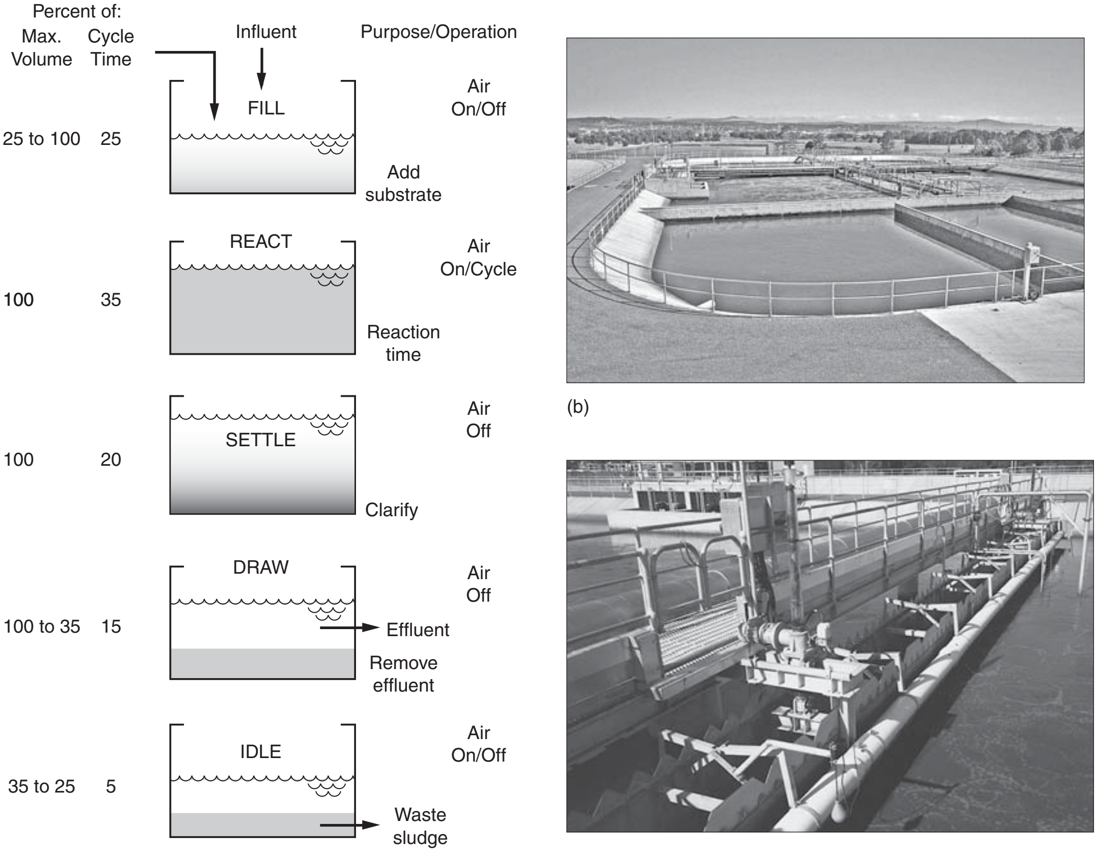
Sequencing Batch Reactor Process. The sequencing batch reactor (SBR) process utilizes a fill-and-draw reactor with complete mixing during the batch reaction step (after filling) and where the subsequent steps of aeration and clarification occur in the same tank. All SBR systems have five steps in common, carried out in following sequence: (1) fill, (2) react (aeration), (3) settle (sedimentation/clarification), (4) draw (decant), and (5) idle. Each of these steps is illustrated on Fig. 8–20 and described later in Table 8–16 on page 771. For continuous-flow applications, at least two SBR tanks must be provided so that one tank receives flow



(a) (b)

Figure 8–19

Complete-mix activated sludge process: (a) schematic diagram and (b) view of a typical complete-mix reactor.



(a) (b) (c)

Figure 8–20

Sequencing batch reactor (SBR) activated sludge process: (a) schematic diagram and (b) view of a typical SBR reactor and a view of movable weir used to decant supernatant after settling period in SBR reactor. Weir is located on the far side of the second dividing wall shown in (c). Photographs were taken in Australia.

while the other completes its treatment cycle. Several process modifications have been made in the times associated with each step to achieve nitrogen and phosphorus removal.

Staged Activated Sludge Process. Activated sludge processes can be designed with baffle walls to intentionally create a number of complete-mix activated sludge zones operating in series (see Fig. 8–22 on page 782). Long, narrow aeration tanks referred to as plug flow activated sludge systems may also have 3 to 4 effective stages depending on the tank layout and aeration mixing characteristics. For the same total reactor volume, a system with reactors in series can provide greater treatment efficiency than a single complete-mix reactor, or provide a greater treatment capacity. As a consequence, staged activated sludge process configurations are used for aerobic nitrification and anoxic zone designs at many full-scale installations.

General Process Design Considerations

The fundamental principles of wastewater characterization, biological treatment, and process analysis were presented in Chap. 7 and in Secs. 8–2 and 8–3. For wastewater characterization, bCOD, TKN, rbCOD and nbVSS are of greatest significance in process design. For BOD removal and nitrification processes, the wastewater bCOD and TKN concentration, and activated sludge temperature and aerobic SRT are important for determining the oxygen demand rate. The rbCOD, TKN, and NH_4-N concentrations are important for evaluating the oxygen demand profiles for plug-flow, staged, and batch-fed processes. The influent bCOD and nbVSS concentrations affect the process sludge production and aeration tank volume requirements. Design and operating issues specific to each of the three activated sludge process configurations considered in the overview are addressed separately.

Process Kinetics for BOD Removal. For general design purposes, kinetic expressions used for design are summarized in Table 8–10. Kinetic coefficients for the removal of carbonaceous material (based on bCOD) by heterotrophic bacteria and ammonia and nitrite oxidation by autotrophic bacteria are given in Table 8–14. Values for many of these coefficients vary widely in the literature. The values provided are those used most commonly and that also provide some degree of conservatism for the design. The values for the heterotrophic organism are essentially the same as the default values used in the IWA ASM2d model (Henze et al., 1995). Nitrification kinetics values are based, for the most part, on nitrification kinetics derived from a Water Environment Research Foundation study on parameters for activated sludge modeling (Melcer et al., 2003).

Process Kinetics for Nitrification. The nitrification kinetic values for K_{O_2}, K_{NH_4}, and μ_{max}, as discussed in Sec. 7–9 in Chap. 7, can vary with the nitrifying bacteria population selection due to the reactor configuration and operating conditions. The half-velocity coefficient for DO (K_{O_2}) is also site specific, as it is dependent on the activated sludge reactor MLSS concentration, mixing conditions, volumetric oxygen uptake rate, and floc size. Where industrial wastewater is present in the WWTP influent, the potential for nitrification inhibition must be considered. Sampling and bench scale testing should be considered in such cases to assure that expected nitrification rates can be obtained. The SRT and aeration tank volume requirements are related directly to nitrification μ_{max} values. Source control measures may be required to mitigate nitrification inhibition.

Complete-Mix Activated Sludge Process Design

In the computational approach used in the design of the activated sludge process, as outlined in Table 8–15, use is made of the design equations presented previously in Chap. 7

Table 8–14

Activated sludge design kinetic coefficients for BOD removal and nitrification at 20°C

Coefficient	Unit	COD oxidation[a]	NH_4 oxidation[b]	NO_2 oxidation[b]
μ_{max}	g VSS/g VSS·d	6.0	0.90	1.0
K_s, K_{NH_4}, K_{NO_2}	mg/L	8.0	0.50	0.20
Y	g VSS/g substrate oxidized	0.45	0.15	0.05
b	g VSS/g VSS·d	0.12	0.17	0.17
f_d	unitless	0.15	0.15	0.15
K_{o_2}	mg/L	0.20	0.50	0.90
θ Value				
μ_{max}	unitless	1.07	1.072	1.063
b	unitless	1.04	1.029	1.029
K_s, K_{NH4}, K_{NO2}	unitless	1.0	1.0	1.0

[a] Adapted from Henze et al. (1995); Barker and Dold (1997).

[b] Adapted from U.S. EPA (2010).

Table 8–15

Computation approach for the design of the activated sludge process

1. Obtain influent wastewater characterization data.
2. Determine the effluent requirements in terms of NH_4-N, TSS, and BOD concentrations.
3. Select an appropriate nitrification safety factor for the design SRT based on expected peak/average TKN loadings. Safety factors may vary from 1.3 to 2.0.
4. Select the minimum DO concentration for the aeration basin mixed liquor. A minimum DO concentration of 2.0 mg/L is recommended for nitrification.
5. Determine the nitrification maximum specific growth rate (μ_{max}) based on the aeration basin temperature and DO concentration, and determine K_n.
6. Determine the net specific growth rate μ and SRT at this growth rate, to meet the effluent NH_4-N concentration.
7. Obtain the design SRT by applying the safety factor to Step 6.
8. Determine the biomass production.
9. Perform a nitrogen balance to determine NO_X, the concentration of NH_4-N oxidized.
10. Calculate the VSS mass and TSS mass for the aeration basin.
11. Select a design MLSS concentration and determine the aeration basin volume and hydraulic residence time.
12. Determine the overall sludge production and observed yield.
13. Calculate the oxygen demand.
14. Design the aeration oxygen transfer system.
15. Determine if alkalinity addition is needed.
16. Design the secondary clarifier.
17. Summarize the final effluent quality.
18. Prepare a design summary table.

and in Secs. 8–2 and 8–3, as summarized in Table 8–10. The application of the design approach is presented in Example 8–3. The key elements of process analysis are selection of the design SRT, selection of kinetic and stoichiometric coefficients, and application of appropriate mass balances.

Example 8–3 **Complete-mix Activated Sludge Process Design for BOD Removal Only and for BOD Removal with Nitrification** Prepare a process design for a complete-mix activated sludge (CMAS) system to treat 22,700 m^3/d of primary effluent (including recycle flows) to (a) meet a BOD_e concentration less than 30 g/m^3 and (b) accomplish BOD removal and nitrification with an effluent NH_4-N concentration of 0.50 g/m^3 and BOD_e and $TSS_e \leq 15\ g/m^3$. Compare the two design conditions in a summary table. The aeration basin mixed-liquor temperature is 12°C.

The following wastewater characteristics and design conditions apply:

Wastewater characteristics:

Constituent	Concentration, mg/L
BOD	140
sBOD	70
COD	300
sCOD	132
rbCOD	80
TSS	70
VSS	60
TKN	35
NH_4-N	25
TP	6
Alkalinity	140 as $CaCO_3$
bCOD/BOD ratio	1.6

Note: g/m^3 = mg/L.

Design conditions and assumptions:

1. Fine bubble membrane diffusers with an aeration clean water O_2 transfer efficiency = 35%
2. Liquid depth for the aeration basin = 4.9 m
3. The point of air release for the ceramic diffusers is 0.5 m above the tank bottom
4. DO in aeration basin = 2.0 g/m^3
5. Site elevation is 500 m (pressure = 95.6 kPa)
6. Aeration α factor = 0.50 for BOD removal only and 0.65 for nitrification; β = 0.95 for both conditions, and diffuser fouling factor F = 0.90
7. Use kinetic coefficients given in Table 8–14
8. SRT for BOD removal = 5 d

9. Design MLSS X_{TSS} concentration = 3000 g/m^3; values of 2000 to 3000 g/m^3 can be considered
10. Peak to average TKN loading rate ratio = 1.5

Solution, Part A—BOD removal without nitrification

1. Develop the wastewater characteristics needed for design.
   a. Find bCOD.

$$\text{bCOD} = 1.6\ (\text{BOD}) = 1.6\ (140\ \text{g/m}^3) = 224\ \text{g/m}^3$$

   b. Find nbCOD using Eq. (8–12).

$$\text{nbCOD} = \text{COD} - \text{bCOD} = (300 - 224)\ \text{g/m}^3 = 76\ \text{g/m}^3$$

   c. Find effluent nonbiodegradable sCOD (nbsCOD_e).

$$\text{nbsCOD}_e = \text{sCOD} - 1.6\ \text{sBOD}$$
$$= (132\ \text{g/m}^3) - (1.6)(70\ \text{g/m}^3) = 20\ \text{g/m}^3$$

   d. Find nbVSS using Eq. (8–7, 8–8 and 8–9).

$$\text{nbpCOD} = \text{TCOD} - \text{bCOD} - \text{nbsCODe}$$

$$\text{nbpCOD} = (300 - 224 - 20)\ \text{g/m}^3 = 56\ \text{g/m}^3$$

$$\text{VSS}_{\text{COD}} = \frac{\text{TCOD} - \text{sCOD}}{\text{VSS}}$$

$$\text{VSS}_{\text{COD}} = \frac{(300 - 132)\text{g/m}^3}{60\ \text{g/m}^3} = 2.8\ \text{g COD/g VSS}$$

$$\text{nbVSS} = \frac{\text{nbpCOD}}{\text{VSS}_{\text{COD}}}$$

$$\text{nbVSS} = \frac{56\ \text{g COD/m}^3}{2.8\ \text{g COD/g VSS}} = 20.0\ \text{g nbVSS/m}^3$$

   e. Find the iTSS.

$$\text{iTSS} = \text{TSS} - \text{VSS}$$
$$= (70 - 60)\ \text{g/m}^3 = 10\ \text{g/m}^3$$

2. Design suspended growth system for BOD removal only.
   a. Determine biomass production using Eq. (8–20) in Table 8–10.

$$P_{X,\text{Bio}} = \frac{QY_H(S_o - S)}{1 + b_H(\text{SRT})} + \frac{(f_d)(b_H)QY_H(S_o - S)\text{SRT}}{1 + b_H(\text{SRT})}$$

Define input data for above equation.

Q = 22,500 m^3/d

S_o = 224 g bCOD/m^3 (see Step 1)

From Table 8–14.

$Y_H = 0.45$ g VSS/g bCOD

$b_{H,20} = 0.12$ g/g·d

$f_d = 0.15$

Determine S from Eq. (7–46) in Table 8–10. Note: Yk = μ_{max}

$$S = \frac{K_s[1 + b_H(\text{SRT})]}{\text{SRT}(\mu_{max} - b_H) - 1}$$

Use μ_{max}, b, and K_s at 20°C from Table 8–14

$\mu_{m,T} = \mu_m \theta^{(T-20)}$ from Eq. (1–44) in Table 8–1

$\mu_{m,12°C} = 6.0\text{ g/g·d }(1.07)^{12-20} = 3.5\text{ g/g·d}$

$b_{H,T} = b_{H,20}\,\theta^{(T-20)}$ from Eq. (1–44)

$b_{H,12°C} = (0.12\text{ g/g·d})(1.04)^{12-20} = 0.088\text{ g/g·d}$

$$S = \frac{(8.0\text{ g/m}^3)[1 + (0.088\text{ g/g·d})(5\text{ d})]}{(5\text{ d})(3.5 - 0.088)\text{ g/g·d} - 1} = 0.7\text{ g bCOD/m}^3$$

b. Substitute the above values in the expression given above and solve for $P_{X,VSS}$.

$$P_{X,\text{Bio}} = \frac{(22{,}700\text{ m}^3/\text{d})(0.45\text{ g/g})[(224 - 0.7)\text{ g/m}^3](1\text{ kg}/10^3\text{ g})}{[1 + (0.088\text{ g/g·d})(5\text{ d})]}$$

$$+ \frac{(0.15\text{ g/g})(0.088\text{ g/g·d})(22{,}700\text{ m}^3/\text{d})(0.45\text{ g/g})[(224 - 0.7)\text{ g/m}^3](5\text{ d})(1\text{ kg}/10^3\text{ g})}{[1 + (0.088\text{ g/g·d})(5\text{ d})]}$$

$P_{X,\text{Bio}} = (1584.0 + 104.5)\text{ kg/d} = 1688.5\text{ kg VSS/d}$

3. Determine the mass in terms of VSS and TSS in the aeration basin. The mass of VSS and TSS can be determined using Eqs. (8–20), (8–21) and (7–57) given in Table 8–10.

Mass = P_X(SRT)

a. Determine $P_{X,VSS}$ and $P_{X,TSS}$ using Eqs. (8–20) including parts A, B, and D. Part C = 0 because there is no nitrification.

$P_{X,VSS} = P_{X,\text{bio}} + Q(\text{nbVSS})$

$$\begin{aligned} P_{X,VSS} &= 1688.5\text{ kg/d} + Q(\text{nbVSS}) \\ &= 1688.5\text{ kg/d} + (22{,}700\text{ m}^3/\text{d})(20\text{ g/m}^3)(1\text{ kg}/10^3\text{ g}) \\ &= (1688.5 + 454.0)\text{ kg/d} = 2142.5\text{ kg/d} \end{aligned}$$

From Eq. (8–21), $P_{X,TSS}$ is

$$\begin{aligned} P_{X,TSS} &= [(1688.5\text{ kg/d})/0.85] + (454.0\text{ kg/d}) + Q(\text{TSS}_o - \text{VSS}_o) \\ &= 1986.5\text{ kg/d} + 454.0\text{ kg/d} + (22{,}700\text{ m}^3/\text{d})(10\text{ g/m}^3)(1\text{ kg}/10^3\text{ g}) \\ &= 2667.5\text{ kg/d} \end{aligned}$$

b. Calculate the mass of VSS and TSS in the aeration basin.
   i. Mass of MLVSS using Eq. (7–57) in Table 8–10

$$\begin{aligned}(X_{\text{VSS}})(V) &= (P_{X,\text{VSS}})\,\text{SRT}\\ &= (2142.5\,\text{kg/d})(5\text{ d}) = 10{,}712\,\text{kg}\end{aligned}$$

   ii. Mass of MLSS using Eq. (7–57) in Table 8–10

$$\begin{aligned}(X_{\text{TSS}})(V) &= (P_{\text{X,TSS}})\,\text{SRT}\\ &= (2667.5\,\text{kg/d})(5\text{ d}) = 13{,}337\,\text{kg}\end{aligned}$$

4. Select a design MLSS mass concentration and determine the aeration tank volume and detention time using the TSS mass computed in Step 3b.
   a. Determine the aeration tank volume using the relationship from Step 3b.

$$(X_{\text{TSS}})(V) = 13{,}337\,\text{kg}$$

At $X_{\text{TSS}} = 3000\text{ g/m}^3$

$$V = \frac{(13{,}337\,\text{kg})(10^3\text{ g/1 kg})}{(3000\,\text{g/m}^3)} = 4445.7\,\text{m}^3$$

   b. Determine the aeration tank detention time.
   Use 3 basins at 1480 m^3 each so that one of the basins can be taken offline for a short period of time when maintenance of the aeration system is necessary.

$$\tau = \frac{V}{Q} = \frac{(4445.7\,\text{m}^3)(24\text{ h/d})}{(22{,}700\,\text{m}^3\text{/d})} = 4.7\,\text{h}$$

   c. Determine MLVSS.

$$\text{Fraction VSS} = \frac{10{,}712\,\text{kg VSS}}{13{,}337\,\text{kg TSS}} = 0.80$$

$$\text{MLVSS} = 0.80(3000\text{ g/m}^3) = 2400\text{ g/m}^3$$

5. Determine F/M and BOD volumetric loading.
   a. Determine F/M using Eq. (7–62) in Table 8–10.

$$\begin{aligned}\text{F/M} &= \frac{QS_o}{XV} = \frac{\text{kg BOD}}{\text{kg MLVSS}\cdot\text{d}}\\ &= \frac{(22{,}700\,\text{m}^3\text{/d})(140\,\text{g/m}^3)}{(2400\,\text{g/m}^3)(4446\text{ m}^3)} = 0.30\,\text{g/g}\cdot\text{d} = 0.30\,\text{kg/kg}\cdot\text{d}\end{aligned}$$

   b. Determine volumetric BOD loading using Eq. (7–69) in Table 8–10.

$$\begin{aligned}\text{BOD loading} &= \frac{QS_o}{V} = \frac{\text{kg BOD}}{\text{m}^3\cdot\text{d}}\\ &= \frac{(22{,}700\,\text{m}^3\text{/d})(140\,\text{g/m}^3)}{(4446\,\text{m}^3)(10^3\,\text{g/1 kg})} = 0.71\text{ kg/m}^3\cdot\text{d}\end{aligned}$$

6. Determine the observed yield based on TSS and VSS.
   a. Observed yield based on TSS.

Observed yield = g TSS/g bCOD = kg TSS/kg bCOD

$P_{X,\,\mathrm{TSS}} = 2667.6\ \mathrm{kg/d}$

$$\begin{aligned}\text{bCOD removed} &= Q(S_o - S)\\ &= (22{,}700\ \mathrm{m^3/d})[(224 - 0.7)\ \mathrm{g/m^3}](1\ \mathrm{kg}/10^3\ \mathrm{g})\\ &= 5068.9\ \mathrm{kg/d}\end{aligned}$$

$$\begin{aligned}Y_{\mathrm{obs,TSS}} &= \frac{(2667.6\ \mathrm{kg/d})}{(5068.9\ \mathrm{kg/d})} = \frac{0.53\ \mathrm{kg\ TSS}}{\mathrm{kg\ bCOD}} = \frac{0.53\ \mathrm{g\ TSS}}{\mathrm{g\ bCOD}}\\ &= \left(\frac{0.53\ \mathrm{g\ TSS}}{\mathrm{g\ bCOD}}\right)\left(\frac{1.6\ \mathrm{g\ bCOD}}{\mathrm{g\ BOD}}\right) = 0.84\ \mathrm{g\ TSS/g\ BOD}\end{aligned}$$

b. Observed yield based on VSS.

$Y_{\mathrm{obs,VSS}}$: VSS/TSS = 0.80 (see Step 4c)

$$\begin{aligned}&= \left(\frac{0.53\ \mathrm{g\ TSS}}{\mathrm{g\ bCOD}}\right)\left(\frac{0.8\ \mathrm{g\ VSS}}{\mathrm{g\ TSS}}\right)\\ &= 0.42\ \mathrm{g\ VSS/g\ bCOD}\\ &= \left(\frac{0.42\ \mathrm{g\ VSS}}{\mathrm{g\ bCOD}}\right)\left(\frac{1.6\ \mathrm{g\ bCOD}}{\mathrm{g\ BOD}}\right)\\ &= 0.64\ \mathrm{g\ VSS/g\ BOD}\end{aligned}$$

7. Calculate the O_2 demand using Eq. (8–23) in Table 8–10.

$$\begin{aligned}R_o &= Q(S_o - S) - 1.42\,P_{X,\mathrm{bio}} + 4.57(Q)\mathrm{NO}_x\\ &= (22{,}700\ \mathrm{m^3/d})[(224 - 0.7)\ \mathrm{g/m^3}](1\ \mathrm{kg}/10^3\ \mathrm{g}) - 1.42(1688.5\ \mathrm{kg/d})\end{aligned}$$

$$\begin{aligned}R_o &= 5068.9\ \mathrm{kg/d} - 2397.7\ \mathrm{kg/d}\\ &= 2671.2\ \mathrm{kg/d} = 111.3\ \mathrm{kg\ O_2/h}\end{aligned}$$

8. Fine bubble aeration design – determine air flowrate at average design flowrate. Based on Eq. (5–55) the aeration oxygen transfer rate under actual conditions in the aeration tank is related to the oxygen transfer rate at standard conditions.

$$\mathrm{SOTR} = \left(\frac{\mathrm{OTR}_f}{\alpha F}\right)\left[\frac{C^*_{\infty 20}}{\beta(C_{\mathrm{st}}/C^*_{\mathrm{s20}})(P_b/P_s)(C^*_{\infty 20}) - C}\right][(1.024)^{20-T}]$$

Where SOTR = standard oxygen transfer rate at site, kg/h
OTR_f = actual oxygen transfer rate at site, kg/h
α = relative transfer rate to clean water
β = relative DO saturation to clean water (0.95 to 0.98)
F = diffuser fouling factor
$\mathrm{C}^*_{\mathrm{st}}$ = saturated DO at sea level and operating temperature, mg/L
$\mathrm{C}^*_{\mathrm{s20}}$ = saturated DO value at sea level and 20°C, mg/L
$\mathrm{C}^*_{\infty,20}$ = saturated DO value at sea level and 20°C for diffused aeration, mg/L. It is higher than C_{st} as it is affected by oxygen transfer from bubbles under pressure in water column.

$C^*_{\infty,20}$ can be estimated by the following equation (U.S. EPA, 1989):

$$C^*_{\infty,20} = C^*_{S20}\left[1 + d_e\left(\frac{D_f}{P_a}\right)\right]$$

P_a = standard pressure at sea level, (760 mm) (10.33 m)
P_b = pressure at the plant site based on elevation, m
D_f = depth of diffusers in basin, m or ft
C = operating DO in basin, mg/L
T = aeration basin temperature, °C
d_e = mid-depth correction factor; may vary from 0.25–0.45 (0.40)

i. From Table E–1 (Appendix E), $C_{s,20} = 9.09$ mg/L and $C_{12} = 10.78$ mg/L
ii. Determine the relative pressure at elevation 500 m to correct the DO concentration for altitude.
From Appendix B

$$\frac{P_b}{P_a} = \exp\left[-\frac{gM(z_b - z_a)}{RT}\right]$$

$$= \exp\left\{-\frac{(9.81\ \text{m/s}^2)(28.97\ \text{kg/kg-mole})[(500 - 0)\ \text{m}]}{(8314\ \text{kg}\cdot\text{m}^2/\text{s}^2\cdot\text{kg-mole}\cdot\text{K})[(273.15 + 12)\ K]}\right\} = 0.94$$

a. Determine the oxygen concentration at 20°C ($C_{sat,20}$) accounting for the gas release at the diffuser depth.
Tank liquid depth = 4.9 m
Diffuser depth, D_f = 4.9 m − 0.5 m = 4.4 m

$$C^*_{\infty 20} = C^*_{s20}\left[1 + d_e\left(\frac{D_f}{P_a}\right)\right]$$

$$C^*_{\infty 20} = 9.09\left[1 + 0.40\left(\frac{4.4\ \text{m}}{10.33\ \text{m}}\right)\right] = 10.64$$

b. Determine the SOTR using $\alpha = 0.50$, $\beta = 0.95$, and diffuser fouling factor $F = 0.9$.

$$\text{SOTR} = \left(\frac{\text{OTR}_f}{\alpha F}\right)\left\{\frac{C^*_{\infty 20}}{\left[\beta\frac{C^*_{\text{st}}}{C^*_{\text{s20}}}\left(\frac{P_b}{P_a}\right)(C^*_{\infty 20}) - \text{C}_\text{L}\right]}\right\}(1.024^{20-T})$$

$$\text{SOTR} = \left(\frac{111.3\ \text{kg/h}}{0.50(0.90)}\right)\left\{\frac{10.64}{\left[0.95\left(\frac{10.78}{9.09}\right)(0.94)(10.64) - 2.0\right]}\right\}(1.024^{20-12})$$

$$= 343.3\ \text{kg/h}$$

c. Determine the air flowrate.

$$\text{Air flowrate, m}^3/\text{min} = \frac{(\text{SOTR kg/h})}{[(E)(60\ \text{min/h})(\text{kg O}_2/\text{m}^3\ \text{air})]}$$

Using the data given in Appendix B, the density of air at 12°C and a pressure of 95.2 kPa (0.94 × 101.325 kPa) is 1.1633 kg/m^3. The corresponding amount of oxygen by weight is 0.270 (0.2318 × 1.1633 kg/m^3). Thus, the required air flowrate is

$$\text{Air flowrate, m}^3/\text{min} = \frac{(343.3\ \text{kg/h})}{[(0.35)(60\ \text{min/h})(0.270\ \text{kg O}_2/\text{m}^3\ \text{air})]}$$
$$= 60.5\ \text{m}^3/\text{min}$$

Note: To continue the facilities design for secondary clarifiers for the BOD removal process, move to Part C, Step 21. For nitrification design, continue to Step 9.

Solution, Part B—BOD removal and nitrification

9. Perform the nitrification design following the same steps as for BOD removal except the design SRT must first be determined. Determine the specific growth rate μ_n for the ammonia oxidizing bacteria using Eq. (7–94) in Table 8–10. The nitrification rate will control the design because the nitrifying organisms grow more slowly than the heterotrophic organisms that remove organic carbon.

$$\mu_{\text{AOB}} = \mu_{\text{max,AOB}}\left[\frac{S_{\text{NH}_4}}{S_{\text{NH}_4} + K_{\text{NH}_4}}\right]\left[\frac{S_o}{S_o + K_{o,\text{AOB}}}\right] - b_{\text{AOB}}$$

Select values for $\mu_{\text{max,AOB}}$, b_{AOB}, K_{NH_4} and $K_{o,\text{AOB}}$ from Table 8–14 at 20°C. These are 0.90 g/g·d, 0.17 g/g·d, 0.50 g/m^3, and 0.50 g/m^3, respectively. Use temperature correction, θ values, from Table 8–14.

a. Find $\mu_{\text{max,AOB}}$ at T = 12°C.

$$\mu_{\text{max,AOB,12°C}} = (0.90\ \text{g/g·d})(1.072)^{12-20} = 0.516\ \text{g/g·d}$$

b. Find b_{AOB}, at T = 12°C.

$$b_{\text{AOB,12°C}} = (0.17\ \text{g/g·d})(1.029)^{12-20} = 0.135\ \text{g/g·d}$$

c. Substitute the above and given values in Eq. (7–94) and solve for μ_{AOB}.
$S_{\text{NH}_4} = 0.50$ g/m^3, DO = 2.0 g/m^3, $K_{o,\text{AOB}} = 0.50$ g/m^3

$$\mu_{\text{AOB}} = \left\{\frac{(0.516\ \text{g/g·d})(0.50\ \text{g/m}^3)}{[(0.50 + 0.50)\ \text{g/m}^3]}\right\}\left\{\frac{(2.0\ \text{g/m}^3)}{[(2.0 + 0.50)\ \text{g/m}^3]}\right\} - (0.135\ \text{g/g·d})$$
$$= 0.0714\ \text{g/g·d}$$

10. Determine the theoretical and design SRT.

a. Find theoretical SRT using Eq. (7–98) in Table 8–10.

$$\text{SRT} = \frac{1}{\mu_{\text{AOB}}} = \frac{1}{(0.0714\ \text{g/g·d})} = 14.0\ \text{d}$$

b. Determine the design SRT using Eq. (7–73).

Design SRT = (SF)(theoretical SRT)

SF = Peak to average TKN load = 1.5

Design SRT = 1.5 (14.0 d) = 21.0 d

11. Determine biomass production using Eq. (8–20) with parts A, B, and C in Table 8–10.

$$P_{X,\text{bio,VSS}} = \frac{QY_{\text{H}}(S_o - S)}{1 + b_{\text{H}}(\text{SRT})} + \frac{(f_d)(b_{\text{H}})QY_{\text{H}}(S_o - S)\text{SRT}}{1 + b_{\text{H}}(\text{SRT})} + \frac{QY_n(\text{NO}_X)}{1 + b_{\text{AOB}}(\text{SRT})}$$

a. Define input data for the above equation.

$Q = 22{,}700\ \text{m}^3/\text{d}$

$Y_\text{H} = 0.45\ \text{VSS/g bCOD}$

$S_o = 224\ \text{g bCOD/m}^3$ (Step 1)

$b_\text{H} = 0.088\ \text{g/g·d}$ (Step 2a)

$\mu_m = 3.5\ \text{g/g·d}$ (Step 2a)

Determine S from Eq. (7–46) in Table 8–10.

$$S = \frac{K_s[1 + b_\text{H}(\text{SRT})]}{[\text{SRT}(\mu_m - b_\text{H}) - 1]}$$

$$S = \frac{(8\ \text{g/m}^3)[1 + (0.088\ \text{g/g·d})(21.0\ \text{d})]}{(21.0\ \text{d})(3.5 - 0.088\ \text{g/g·d}) - 1} = 0.32\ \text{g bCOD/m}^3$$

$Y_n = 0.15\ \text{g VSS/g NO}_x$ (Table 8–14)

$b_{\text{AOB},12°\text{C}} = 0.135\ \text{g/g·d}$ (Step 9b)

A value for NO_x is needed in part C of Eq. (8–20). As NO_x is contained in Eq. (8–20) to obtain $P_{X,\text{Bio}}$, one solution approach is to assume the NO_x concentration in Eq. (8–20) as a certain percentage of the influent TKN. An iterative solution can then be used to solve for NO_x, after calculating $P_{X,\text{Bio}}$. A value for $\text{NO}_x = 80\%$ (TKN) is a reasonable approximation for the BOD to TKN ratio for this problem. The error would be small as the nitrifier VSS yield is a normally a small fraction of total MLVSS concentration.

$\text{NO}_x = 0.80(35\ \text{g/m}^3) = 28\ \text{g/m}^3$

b. Substitute the above values in the expression given above and solve for $P_{X,\text{bio,VSS}}$.

$$P_{X,\text{bio,VSS}} = \frac{(22{,}700\ \text{m}^3/\text{d})(0.45\ \text{g/g})[(224 - 0.32)\ \text{g/m}^3](1\ \text{kg}/10^3\ \text{g})}{[1 + (0.088\ \text{g/g·d})(21.0\ \text{d})]}$$
$$+ \frac{(0.15\ \text{g/g})(0.088\ \text{g/g·d})(0.45\ \text{g/g})(22{,}700\ \text{m}^3/\text{d})[(224 - 0.32)\ \text{g/m}^3](21.0\ \text{d})(1\ \text{kg}/10^3\ \text{g})}{[1 + (0.088\ \text{g/g·d})\ (21.0\ \text{d})]}$$
$$+ \frac{(22{,}700\ \text{m}^3/\text{d})(0.15\ \text{g/g})(28\ \text{g/m}^3)(1\ \text{kg}/10^3\ \text{g})}{[1 + (0.135\ \text{g/g·d})(21.0\ \text{d})]}$$

$$\begin{aligned} P_{X,\text{bio,VSS}} &= 802.3\ \text{kg/d} + 222.4\ \text{kg/d} + 24.9\ \text{kg/d} \\ &= 1049.6\ \text{kg VSS/d} \end{aligned}$$

12. Determine the amount of nitrogen oxidized to nitrate (NO_x). The amount of nitrogen oxidized to nitrate can be found by performing a nitrogen balance using Eq. (8–24) in Table 8–10.

$$\begin{aligned} \text{NO}_x &= \text{TKN} - N_e - 0.12 P_{X,\text{bio}}/Q \\ &= 35.0\ \text{g/m}^3 - 0.50\ \text{g/m}^3 \\ &\quad - (0.12\ \text{g N/g VSS})(1049.6\ \text{kg VSS/d})(10^3\ \text{g/kg})/(22{,}700\ \text{m}^3/\text{d}) \\ &= (35.0 - 0.50 - 5.6)\ \text{g/m}^3 = 28.9\ \text{g/m}^3 \end{aligned}$$

The computed value is close to the 28.0 g/m^3 assumed. Substituting 28.9 for NO_x in the above calculation yields a NO_x value of 28.9 g/m^3 again, so one iteration was satisfactory. The increase in $P_{X,\text{bio,VSS}}$ for Part C in Eq. (8–20) is from 24.9 kg/d to 25.7 kg/d, so $P_{X,\text{bio,VSS}}$ is now 1050.4 kg VSS/d

13. Determine the concentration and mass of VSS and TSS in the aeration basin. The mass of VSS and TSS can be determined using Eqs. (8–20), (8–21) and (7–57) given in Table 8–10.

$$\text{Mass} = P_X(\text{SRT})$$

a. Determine $P_{X,\text{VSS}}$ using Eq. (8–20) including parts A, B, C and D by adding the contribution from nbVSS to the above.

$$P_{X,\text{VSS}} = P_{x,\text{bio}} + Q(\text{nbVSS})$$

$$\begin{aligned} P_{X,\text{VSS}} &= 1050.4\,\text{kg/d} + Q(\text{nbVSS}) \\ &= 1050.4\,kg/\text{d} + (22{,}700\,\text{m}^3/\text{d})(20\,\text{g/m}^3)(1\,\text{kg}/10^3\,\text{g}) \\ &= (1050.4 + 454.0)\,\text{kg/d} = 1504.4\,\text{kg/d} \end{aligned}$$

From Eq. (8–21), $P_{X,\text{TSS}}$ is

$$\begin{aligned} P_{X,\text{TSS}} &= [(1050.4\,\text{kg/d})/0.85] + (454.0\,\text{kg/d}) + Q(\text{TSS}_o - VSS_o) \\ &= 1235.8\,\text{kg/d} + 454.0\,\text{kg/d} + (22{,}700\,\text{m}^3/\text{d})(10\,g/\text{m}^3)(1\,\text{kg}/10^3\,\text{g}) \\ &= 1916.8\,\text{kg/d} \end{aligned}$$

b. Calculate the mass of VSS and TSS in the aeration basin.
   i. Mass of MLVSS using Eq. (7–57) in Table 8–10

$$\begin{aligned} (X_{\text{VSS}})(V) &= (P_{X,\text{VSS}})\,\text{SRT} \\ &= (1504.4\,\text{kg/d})(21.0\,\text{d}) = 31{,}592.4\,\text{kg} \end{aligned}$$

   ii. Mass of MLSS using Eq. (7–57) in Table 8–10

$$\begin{aligned} (X_{\text{TSS}})(V) &= (P_{X,\text{TSS}})\,\text{SRT} \\ &= (1916.8\,\text{kg/d})(21.0\,\text{d}) = 40{,}252.8\,\text{kg} \end{aligned}$$

14. Select a design MLSS mass concentration and determine the aeration tank volume and detention time using the TSS mass computed in Step 13b.
a. Determine the aeration tank volume using the relationship from Step 13b.

$$(X_{\text{TSS}})(V) = 40{,}252.8\,\text{kg}$$

At $X_{\text{TSS}} = 3000\ \text{g/m}^3$

$$V = \frac{(40{,}252.8\,\text{kg})(10^3\,\text{g}/1\,\text{kg})}{(3000\,\text{g/m}^3)} = 13{,}418\,\text{m}^3$$

b. Determine the aeration tank detention time.
Use 3 basins at 4470 m^3 each so that one of the basins can be taken offline for a short period of time when maintenance of the aeration system is necessary.

$$\tau = \frac{V}{Q} = \frac{(13{,}410\,\text{m}^3)(24\ \text{h/d})}{(22{,}700\,\text{m}^3/\text{d})} = 14.2\,\text{h}$$

c. Determine MLVSS.

$$\text{Fraction VSS} = \frac{31{,}592 \text{ kg VSS}}{40{,}253 \text{ kg TSS}} = 0.79$$

$$\text{MLVSS} = (0.79)\ 3000 \text{ g/m}^3 = 2370 \text{ g/m}^3$$

15. Determine F/M and BOD volumetric loading.

a. Determine F/M using Eq. (7–62) in Table 8–10.

$$\text{F/M} = \frac{QS_o}{XV} = \frac{\text{kg BOD}}{\text{kg MLVSS·d}}$$

$$= \frac{(22{,}700 \text{ m}^3\text{/d})(140 \text{ g/m}^3)}{(2370 \text{ g/m}^3)(13{,}410 \text{ m}^3)} = 0.10 \text{ g/g·d} = 0.10 \text{ kg/kg·d}$$

b. Determine volumetric BOD loading using Eq. (7–69) in Table 8–10.

$$\text{BOD loading} = \frac{QS_o}{V} = \frac{\text{kg BOD}}{\text{m}^3\text{·d}}$$

$$= \frac{(22{,}700 \text{ m}^3\text{/d})(140 \text{ g/m}^3)}{(13{,}410 \text{ m}^3)(10^3 \text{ g/1 kg})} = 0.24 \text{ kg/m}^3\text{·d}$$

16. Determine the observed yield based on TSS and VSS.

a. Observed yield or net yield based on TSS.

Observed yield = g TSS/g bCOD = kg TSS/kg bCOD

$P_{\text{X,TSS}} = 1917$ kg/d

$$\text{bCOD removed} = Q(S_o - S)$$

$$= (22{,}700 \text{ m}^3\text{/d})[(224 - 0.32) \text{ g/m}^3](1 \text{ kg}/10^3 \text{ g})$$

$$= 5078 \text{ kg/d}$$

$$Y_{\text{obs,TSS}} = \frac{(1917 \text{ kg/d})}{(5078 \text{ kg/d})} = \frac{0.38 \text{ kg TSS}}{\text{kg bCOD}} = \frac{0.38 \text{ g TSS}}{\text{g bCOD}}$$

$$= \left(\frac{0.38 \text{ g TSS}}{\text{g bCOD}}\right)\left(\frac{1.6 \text{ g bCOD}}{\text{g BOD}}\right) = 0.61 \text{ g TSS/g BOD}$$

b. Observed yield based on VSS.

$Y_{\text{obs,VSS}}$:VSS/TSS = 0.79(see Step 14c)

$$= \left(\frac{0.38 \text{ g TSS}}{\text{g bCOD}}\right)\left(\frac{0.79 \text{ g VSS}}{\text{g TSS}}\right)$$

$$= 0.30 \text{ g VSS/g bCOD}$$

$$= \left(\frac{0.30 \text{ g VSS}}{\text{g bCOD}}\right)\left(\frac{1.6 \text{ g bCOD}}{\text{g BOD}}\right)$$

$$= 0.48 \text{ g VSS/g BOD}$$

17. Calculate the O_2 demand using Eq. (8–23) in Table 8–10.

$$R_o = Q(S_o - S) - 1.42\,P_{X,\text{bio}} + 4.57(Q)\text{NO}_x$$

$$P_{X,\text{bio,VSS}} = 802.3\text{ kg/d} + 222.4\text{ kg/d (Step 11} - \text{does not include nitrifying bacteria)}$$

$$= 1024.7\text{ kg VSS/d}$$

$$R_o = (22{,}700\text{ m}^3\text{/d})[(224 - 0.32)\text{ g/m}^3](1\text{ kg/}10^3\text{ g}) - 1.42(1024.7\text{ kg/d})$$

$$+\,4.57(22{,}700\text{ m}^3\text{/d})(28.9\text{ g/m}^3)(1\text{ kg/}10^3\text{ g})$$

$$R_o = 5077.5 - 1455.1 + 2998.1 = 6620\text{ kg O}_2\text{/d} = 275.9\text{ kg O}_2\text{/h}$$

18. Fine bubble aeration design—determine air flowrate at average design flowrate (see procedure for Step 8).
   a. Determine the SOTR using the values given in the problem statement: $\alpha = 0.65$, $\beta = 0.95$, and $F = 0.9$.

$$\text{SOTR} = \left(\frac{\text{OTR}_\text{f}}{\alpha F}\right)\left\{\frac{\text{C}^*_{\infty 20}}{\left[\beta\dfrac{\text{C}^*_{\text{st}}}{\text{C}^*_{\text{s}20}}\left(\dfrac{P_b}{P_b}\right)(\text{C}^*_{\infty 20}) - \text{C}\right]}\right\}(1.024^{20-T})$$

$$\text{SOTR} = \left(\frac{275.9\text{ kg/h}}{0.65(0.90)}\right)\left\{\frac{10.64}{\left[0.95\left(\dfrac{10.78}{9.09}\right)(0.94)(10.64) - 2.0\right]}\right\}(1.024^{20-12}) = 654.6\text{ kg/h}$$

   b. Determine the air flowrate.

$$\text{Air flowrate, m}^3\text{/min} = \frac{(654.6\text{ kg/h})}{[(0.35)(60\text{ min/h})(0.270\text{kg O}_2\text{/m}^3\text{ air})]}$$

$$= 115.5\text{ m}^3\text{/min}$$

19. Check alkalinity.
   a. Prepare an alkalinity (Alk) balance:

Alkalinity to maintain pH ~ 7 = Influent Alk −Alk used + Alk to be added
Influent alkalinity: 140 g/m^3 as $CaCO_3$
Amount of nitrogen converted to nitrate: NO_x = 28.9 g/m^3 (see Step 12)
Alkalinity used for nitrification = (7.14 g $CaCO_3$/g NH_4-N)(28.9 g/m^3)
= 206.3 g/m^3 used as $CaCO_3$

   b. Substitute known values and solve for alkalinity needed.
   Residual alkalinity concentration needed to maintain pH in the range of 6.8–7.0 ≅ 70 g/m^{3}as $CaCO_3$

$$70\text{ g/m}^3 = \text{Influent alk} - \text{alk used} + \text{alk to be added}$$

$$70\text{ g/m}^3 = 140\text{ g/m}^3 - 206.3\text{ g/m}^3 + \text{alk to be added}$$

$$\text{Alkalinity added} = 136.3\text{ g/m}^3\text{ as CaCO}_3$$

$$= (22{,}700\text{ m}^3\text{/d})(136.3\text{ g/m}^3)(1\text{ kg/}10^3\text{ g})$$

$$= 3094\text{ kg/d as CaCO}_3$$

c. Determine the alkalinity needed as sodium bicarbonate.
Sodium bicarbonate may be preferred over lime for alkalinity addition due to ease of handling and fewer scaling problems as compared to lime. The amount of $NaHCO_3$ needed is as follows:

Equivalent weight of $NaHCO_3 = 84$ g/equivalent

$$NaHCO_3 \text{ added} = \frac{(3094\,\text{kg/d CaCO}_3)(84\,\text{g NaHCO}_3/\text{eq})}{(50\,\text{g CaCO}_3/\text{equivalent})}$$
$$= 5197\,\text{kg/d NaHCO}_3$$

20. Estimate effluent BOD using Eq. (8–26).

$$\text{BOD} = \text{sBOD}_e + \left(\frac{0.85\,\text{g BOD}}{1.42\,\text{g VSS}}\right)\left(\frac{0.85\,\text{g VSS}}{\text{g TSS}}\right)(\text{TSS, g/m}^3)$$

Assume $\text{sBOD}_e = 3.0\ \text{g/m}^3$
$\text{TSS} = 10\ \text{g/m}^3$

$$\text{BOD} = 3.0\,\text{g/m}^3 + (0.85)(0.85)(10\,\text{g/m}^3)$$
$$= 10.2\,\text{g/m}^3$$

Solution, Part C—Secondary clarifier sizing

21. Secondary clarifier design (for both BOD removal and BOD removal and nitrification).
a. Define return sludge recycle ratio [see Fig. 8–11(b)]:

$Q_R X_R = (Q + Q_R)X$ (assume waste sludge mass is insignificant)

$Q_R = RAS$ flowrate, m^3/d

X_R = return sludge mass concentration, g/m^3

RAS recycle ratio $= Q_R/Q = R$

$RX_R = (1 + R)X$

$$R = \frac{X}{X_r - X}$$

b. Determine size of clarifier.
Assume $X_R = 8000\ \text{g/m}^3$ (moderate settling/thickening sludge; per Sec. 8–3 range is 4000 to 12,000 mg/L).

$$R = \frac{(3000\,\text{g/m}^3)}{[(8000 - 3000)\,\text{g/m}^3]} = 0.60$$

Assume a hydraulic application rate of 24 $m^3/m^2{\cdot}d$ at average flowrate for the secondary clarifier (see Table 8–34); the range is 16 to 28 $m^3/m^2{\cdot}d$

$$\text{Area} = \frac{(22{,}700\,\text{m}^3/\text{d})}{(24\,\text{m}^3/\text{m}^2\cdot\text{d})} = 946\,\text{m}^2$$

Use 3 clarifiers (1 for each aeration tank)

Area/clarifier = 315 m^2

Clarifier diameter = 20 m

c. Check solids loading.

$$\text{Solids loading} = \frac{(Q + Q_r)(\text{MLSS})}{A} = \frac{(1 + R)Q(\text{MLSS})}{A}$$

where A = area of clarifier, $m^2 = (\pi/4)(20\text{ m})^2 \times 3 = 942\text{ m}^2$

$$\text{Solids loading} = \frac{(1 + 0.6)(22{,}700\text{ m}^3/\text{d})(3000\text{ g/m}^3)(1\text{ kg}/10^3\text{ g})}{(942\text{ m}^2)(24\text{ h/d})}$$

$$= 4.9\text{ kg MLSS/m}^2\cdot\text{h}$$

(within acceptable range of solids loading of 4 to 6 kg/m^2·d given in Table 8–34 on page 890)

22. Prepare design summary.

Design parameter	Unit	BOD removal only (Part A)	BOD removal and nitrification (Part B)
Average flow	m^3/d	22,700	22,700
Average BOD load	kg/d	3178	3178
Average TKN load	kg/d	795	795
Aerobic SRT	d	5.0	21.0
Aeration tanks	number	3	3
Aeration tank volume, ea	m^3	1480	4470
Hydraulic detention time,	h	4.7	14.2
MLSS	g/m^3 (mg/L)	3000	3000
MLVSS	g/m^3 (mg/L)	2400	2370
F/M	g/g·d	0.30	0.10
BOD loading	kg BOD/m^3·d	0.71	0.24
Sludge production	kg/d	2667	1917
Observed yield	kg VSS/kg BOD	0.64	0.48
	kg TSS/kg BOD	0.84	0.61
Oxygen required	kg/h	111.3	275.9
Air flowrate at average wastewater flowrate	m^3/min	60.5	115.5
RAS ratio	Unitless	0.60	0.60
Clarifier hydraulic application rate	m^3/m^2·d	24	24
Clarifiers	number	3	3

(continued)

(*Continued*)

Design parameter	Unit	BOD removal only (Part A)	BOD removal and nitrification (Part B)
	Diameter, m	20	20
Alkalinity addition as $Na(HCO_3)$	kg/d	—	5197
Effluent BOD	g/m^3 (mg/L)	<30	10.2
TSSe	g/m^3 (mg/L)	<30	10
Effluent NH_4-N	g/m^3 (mg/L)	—	≤0.5

Comment The effluent NH_4-N concentration for the BOD removal only design would be a little less than the 28.9 mg/L of NH_4-N oxidized in the nitrification system due to the shorter SRT. In this example, the design procedure is described for an average wastewater flowrate. In actual design, computations must also be made for peak flow and load conditions.

In Example 8–3 the NH_4-N oxidation was assumed to be oxidized completely to NO_3-N and the presence of NO_2-N due to the two step nitrification process of NH_4-N oxidation by AOB and NO_2-N oxidation to NO_3-N by NOB was ignored. The NO_2-N concentration is generally less than 0.30 mg/L in systems operated at typical DO concentrations at or near 2.0 mg/L and temperatures below 25°C, so that ignoring the NO_2-N concentration does not pose a significant error in calculating the oxygen demand or effluent total nitrogen concentration for most applications. However, if the activated sludge system designed in Example 8–3 is operated at a much lower DO concentration, an elevated NO_2-N concentration is possible. The effect of a lower DO concentration on the activated sludge NO_2-N and NH_4-N concentrations is illustrated in Example 8–4.

EXAMPLE 8–4 **Effect of DO Concentration on NO_2-N and NH_4-N Concentrations for the Complete-mix Activated Sludge System Designed in Example 8–3** Using the information from Example 8–3 determine: (a) the NO_2-N effluent concentration and (b) the effluent NO_2-N and NH_4-N concentrations for the complete-mix activated sludge system, if the DO concentration is 0.40 mg/L instead of 2.0 mg/L?

Design conditions and assumptions
1. SRT = 20.6 d
2. Temperature = 12°C

Solution, Part A—NO_2-N effluent concentration

1. Use Eq. (7–46) in Table 8–10 to determine the effluent NO_2-N concentration at DO = 2.0 mg/L. Note: $Yk = \mu$

$$NO_2\text{-N} = \frac{K_{NO_2}[1 + b_{NOB}(SRT)]}{SRT(\mu_{NOB} - b_{NOB}) - 1}$$

2. Determine the effective μ_{NOB} accounting for $\mu_{\mathrm{max,NOB}}$ and the effect of DO.

$$\mu_{\mathrm{NOB}} = \mu_{\mathrm{max,NOB}}\left(\frac{S_{O_2}}{S_{O_2} + K_{O_2,\mathrm{NOB}}}\right)$$

select values for $\mu_{\mathrm{max,NOB}}$, $K_{\mathrm{NO_2}}$, $K_{O_2,\mathrm{NOB}}$, and b_{NOB} from Table 8–14. The selected values are: 1.0 g/g·d, 0.20 g/m^3, 0.90 g/m^3, and 0.17 g/g·d, respectively. Use the temperature correction values from Table 8–14 to correct the coefficients for 12°C. Find $\mu_{\mathrm{max,NOB}}$ at T = 12°C

$$\mu_{\mathrm{max,NOB,12}} = (\mu_{\mathrm{max,NOB,20}})\theta^{(\mathrm{T}-20)}$$

$$\mu_{\mathrm{max,NOB,12}} = (1.0\ \mathrm{g/g{\cdot}d})1.963^{12-20} = 0.61\ \mathrm{g/g{\cdot}d}$$

$$\mu_{\mathrm{NOB,12}} = (0.61\ \mathrm{g/g{\cdot}d})\left[\frac{(2.0\ \mathrm{g/m^3})}{(2.0\ \mathrm{g/m^3} + 0.90\ \mathrm{g/m^3})}\right] = 0.42\ \mathrm{g/g{\cdot}d}$$

3. Find b_{NOB} at T = 12°C.

$$b_{\mathrm{NOB,12}} = (\mathrm{b}_{20})\theta^{(\mathrm{T}-20)}$$
$$b_{\mathrm{NOB,12}} = (0.17\ \mathrm{g/g{\cdot}d})1.029^{(12-20)} = 0.135\ \mathrm{g/g{\cdot}d}$$

4. Substitute the above coefficients in Eq. (7–46) and solve for the effluent NO_2-N concentration.

$$\mathrm{NO_2\text{-}N} = \frac{(0.20\ \mathrm{g/m^3})[1 + (0.135\ \mathrm{g/g{\cdot}d})(20.6\ \mathrm{d})]}{[(20.6\ \mathrm{d})(0.42\ \mathrm{g/g{\cdot}d} - 0.135\ \mathrm{g/g{\cdot}d})] - 1} = 0.16\ \mathrm{mg/L}$$

Solution, Part B—NO_2-N and NH_4-N effluent concentrations

1. Solve for the NO_2-N concentration at DO = 0.40 g/m^3
   a. Determine the effective μ_{NOB} at DO = 0.40 g/m^3 and T=12°C and substitute value into Eq. (7–46).

$$\mu_{\mathrm{NOB,12}} = (0.61\ \mathrm{g/g{\cdot}d})\left[\frac{(0.40\ \mathrm{g/m^3})}{(0.40\ \mathrm{g/m^3} + 0.90\ \mathrm{g/m^3})}\right] = 0.188\ \mathrm{g/g{\cdot}d}$$

$$\mathrm{NO_2\text{-}N} = \frac{(0.20\ \mathrm{g/m^3})[1 + (0.135\ \mathrm{g/g{\cdot}d})(20.6\ \mathrm{d})]}{[(20.6\ \mathrm{d})(0.188\ \mathrm{g/g{\cdot}d} - 0.135\ \mathrm{g/g{\cdot}d})] - 1} = 8.20\ \mathrm{g/m^3}$$

2. Solve for the NH_4-N concentration at DO = 0.40 g/m^3.
   a. Determine μ_{AOB} at DO = 0.40 g/m^3 and T = 12°C and substitute this value and other coefficient values from Example 8–3 into Eq. (7–46) to obtain the effluent NH_4-N concentration. From Example Problem 8–3, $b_{\mathrm{AOB,12}}$ = 0.135 g/gal and $K_{\mathrm{NH_4}}$ = 0.50 g/m^3. From Table 8–14, K_{O_2} = 0.50 g/m^3. The effective μ_{AOB} accounting for $\mu_{\mathrm{max,AOB}}$ and the effect of DO:

$$\mu_{\mathrm{AOB,12}} = \mu_{\mathrm{max,AOB,12}}\left(\frac{S_{O_2}}{S_{O_2} + K_{O_2,\mathrm{AOB}}}\right)$$

$$\mu_{\mathrm{AOB,12}} = (0.52\ \mathrm{g/g{\cdot}d})\left[\frac{(0.40\ \mathrm{g/m^3})}{(0.40\ \mathrm{g/m^3} + 0.50\ \mathrm{g/m^3})}\right] = 0.231\ \mathrm{g/g{\cdot}d}$$

$$\mathrm{NH_4\text{-}N} = \frac{0.50\ \mathrm{g/m^3}[1 + (0.135\ \mathrm{g/g{\cdot}d})(20.6\ \mathrm{d})]}{20.6\ \mathrm{d}(0.231\ \mathrm{g/g{\cdot}d} - 0.135\ \mathrm{g/g{\cdot}d}) - 1} = 1.90\ \mathrm{g/m^3}$$

Summary for effluent NH_4-N and NO_2-N concentrations at a DO of 2.0 mg/L and 0.40 g/m^3 at a 20.6-d SRT:

	DO concentration, mg/L	
Parameter	2.00	0.40
NH_4-N	0.50	1.90
NO_2-N	0.16	8.20

Comment At a low DO concentration, a much higher NO_2-N concentration is predicted, and thus the oxidation of NH_4-N to NO_3-N is incomplete. Note that it is likely that at the low DO concentration some denitrification of NO_2-N would occur within the heterotrophic floc, so that its concentration in the activated sludge reactor would be lower than predicted by assuming only an aerobic condition.

Sequencing Batch Reactor Process Design

Treatment in an SBR system is accomplished over a series of time steps in a single reactor compared to what is done spatially in a flow-through activated sludge system. A unique feature of the SBR system is that there is no need for a return activated sludge (RAS) system, because both aeration and settling occur in the same tank. The SBR process can

Table 8–16

Description of operational steps for the sequencing batch reactor (SBR)

Operational step	Description
Fill	During the fill period, raw wastewater or primary effluent is added to the reactor. In the fill step the liquid level in the reactor typically rises from 75 percent (at the end of the idle period) to 100 percent of the maximum liquid volume. When two tanks are used, the time for the fill process may last about 50 percent of the full cycle time. During fill, the reactor may be mixed only or mixed and aerated to promote biological reactions with the influent wastewater. A mix only of at least 50 percent promotes filamentous growth control and improved settling and thickening.
React	During the react period, the biomass consumes the substrate under controlled environmental conditions, which can be aeration only or cyclic aeration and mixing to promote biological nitrification and denitrification for nitrogen removal.
Settle	Solids are allowed to separate from the liquid under quiescent conditions, resulting in a clarified supernatant that can be discharged as effluent.
Decant	Clarified effluent is removed during settled supernatant removal during the decant period. Many types of decanting mechanisms can be used with the most popular being floating or adjustable weirs.
Idle	An idle period is used in a multi-tank system to provide time for one reactor to complete its fill phase before switching to another unit. It also provides capacity for handling higher flows that may occur as a function of wet weather or seasonal loadings.

also be modified to operate in a continuous-flow mode, as discussed later in this chapter. Issues specific to the SBR process include (1) sludge wasting, (2) the application of process kinetics, and (3) selection of key operating conditions including the fraction of the tank contents to be removed during decanting and the settle, decant, and aeration times.

Sludge Wasting in SBRs. Sludge wasting is another important step in the SBR operation that is needed for SRT control and greatly affects performance. Wasting is not included as one of the five basic process steps because there is no set time period within the cycle dedicated to wasting. The amount and frequency of sludge wasting is determined by performance requirements, as with a conventional continuous-flow system. In an SBR operation, sludge wasting usually occurs during the react phase so that a uniform discharge of solids (including fine material and large floc particles) occurs. SBR tanks should be designed with the ability for surface wasting of mixed liquor as discussed in Sec. 8–3 as an effective means for foam control by suppressing the growth of *M. Parvicella* and bacteria causing nocardioform foam.

Application of Process Kinetics. During the react period, batch kinetics apply. The substrate concentration is much higher initially than would be present in a CMAS system, and the substrate concentration decreases with time as it is consumed by the biomass. The change in substrate concentration with time can be determined by starting with the substrate mass balance in Chap. 7 [Eq. (7–43)] for a continuous-flow complete-mix reactor:

$$\frac{dS}{dt}V = QS_o - QS - r_{su}V \tag{7–43}$$

where $$r_{su} = \frac{\mu_m XS}{Y(K_s + S)} \tag{7–15}$$

Other terms as defined previously.

Because $Q = 0$ for the batch reaction, the substrate concentration is

$$\frac{dS}{dt} = -\frac{\mu_m XS}{Y(K_s + S)} \tag{8–46}$$

Integration of Eq. (8–59) with respect to time yields.

$$K_s \ln\frac{S_o}{S_t} + (S_o - S_t) = X\left(\frac{\mu_m}{Y}\right)t \tag{8–47}$$

where S_o = initial substrate concentration at $t = 0$, mg/L
t = time, *d*
S_t = substrate concentration at time t, mg/L

The same kinetic expression applies for nitrification where $X = X_n$, the nitrifying bacteria concentration, $S = N$, the NH_4-N concentration, and the Monod model kinetic coefficients are substituted:

$$K_{NH_4} \ln\frac{N_o}{N_t} + (N_o - N_t) = X_n\left(\frac{\mu_{\max,AOB}}{Y_n}\right)t \tag{8–48}$$

where N_o = NH_4-N concentration at $t = 0$, mg/L
N_t = NH_4-N concentration at time t, mg/L
X_n = nitrifying bacteria concentration, mg/L

The maximum specific growth rate for nitrifying bacteria is affected by the DO concentration [Eq. (7–94)], so this effect is included in Eq. (8–49) as follows:

$$K_{\mathrm{NH_4}} \ln\frac{N_o}{N_t} + (N_o - N_t) = X_n\left(\frac{\mu_{\mathrm{max,AOB}}}{Y_{\mathrm{AOB}}}\right)\left(\frac{S_o}{K_{\mathrm{O,AOB}} + S_o}\right)t \qquad (8\text{–}49)$$

The above batch kinetic equations can be used to determine if the react period aeration time selected for SBR design is sufficient to provide the desired amount of degradation. An overall mass balance can be done first, assuming a certain amount of substrate is removed, to determine the heterotrophic and nitrifier biomass concentrations (X and X_n) for use in the equations. The time needed for dissolved BOD removal is relatively short (less than 1 h) due to the batch kinetics for treating domestic wastewater, resulting in a relatively low initial dissolved BOD concentration. For nitrification, SBR aerobic react times may range from 2.0 to 4.0 h. It should be noted that the SRTs for the SBR and continuous-flow activated sludge processes are not comparable. At the same SRT, the SBR may be expected to be more efficient because of its batch kinetics, but the biomass is only under aeration for a fraction of time, due to the time needed for settling, decanting, and filling without aeration so that the effective aerobic SRT is lower.

Because the substrate concentration changes with time, the substrate utilization and oxygen demand rates change, progressing from high to low levels. The aeration system should be designed to reflect the changing requirements in oxygen demand. Additional descriptive material for the SBR process is provided in a later part of this section.

Key Operating Conditions. Because of the many design variables involved in an SBR design, an iterative approach is necessary in which key reactor operating conditions are first assumed. A set of different operating conditions can be evaluated by use of a spreadsheet analysis to determine the most optimal choice. The key operating conditions that must be selected are (1) the fraction of the tank contents removed during decanting and (2) the settle, decant, and aeration times. Because the fill volume equals the decant volume, the fraction of decant volume equals the fraction of the SBR tank volume used for the fill volume per cycle and is defined as the *fill fraction* or fill volume to total volume ratio. The fill volume fraction is a key SBR design parameter that is used to determine the required liquid volume of an SBR tank. The allowable fraction is based on having a sufficient supernatant liquid above the settled solids during decanting. The fill fraction must be lower for operations with higher MLSS concentration and higher SVI values as a greater depth would be needed to contain the settled solids. The fraction of the total tank depth needed to contain the settled volume can be estimated as follows:

$$F(V_T) = (\mathrm{MLSS,\ mg/L})\left(\frac{\mathrm{SVI,\ mL/g}}{10^6}\right) \qquad (8\text{–}50)$$

where $F(V_T)$ = fraction of SBR full tank volume occupied by settled solids

Allowable fill volume fractions are shown on Fig. 8–21 as a function of SVI and MLSS concentration at the full tank volume. The fill fractions are based on having 0.2 to 0.5 m of supernatant depth below the decanter. Depending on the decanter design and the expected fluid disturbances during decant, a more conservative lower fill volume fraction value than that shown for a given MLSS concentration and SVI would be used. The fill volume

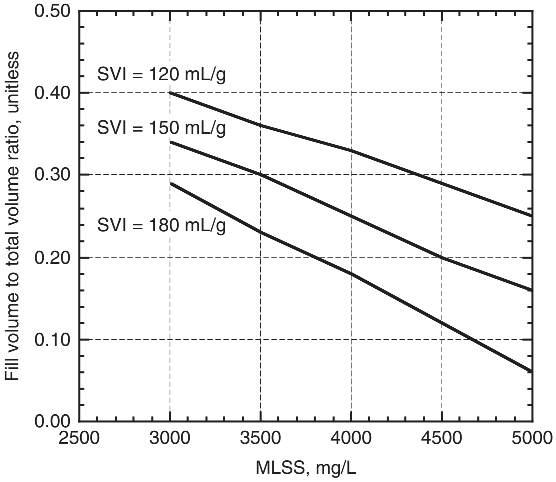


Figure 8–21

Recommended fill/decant fraction for sequencing batch reactor tank as a function of design MLSS concentration at full volume and SVI.

fraction is normally 0.25 or less to allow for a supernatant with low suspended solids after settling. The fill volume per cycle, based on the number of cycles per day and daily flowrate, is used in combination with the fill volume fraction to determine the volume of the SBR tank. The design procedure for the SBR system is presented in Table 8–17 and illustrated in Example 8–5.

Table 8–17

Computation approach for the design of a sequencing batch reactor

Item	Description
1.	Obtain influent wastewater characterization data, define effluent requirements, and define safety factors.
2.	Select the number of SBR tanks.
3.	Select the react/aeration, settling, and decant times. Determine the fill time and total time per cycle. Determine the number of cycles per day.
4.	From the total number of cycles per day, determine the fill volume per cycle
5.	Select the MLSS concentration and determine the fill volume fraction relative to the total tank volume. Determine the decant depth. Using the computed depths, determine the SBR tank volume.
6.	Determine the SRT for the SBR process design developed.
7.	Determine the amount of TKN added that is nitrified.
8.	Calculate the nitrifier biomass concentration and determine if the aeration time selected is sufficient for the nitrification efficiency needed.
9.	Adjust the design as needed–additional iterations may be done.
10.	Determine the decant pumping rate.
11.	Determine the oxygen required and average transfer rate.
12.	Determine the amount of sludge production.
13.	Calculate the F/M and BOD volumetric loading.
14.	Evaluate alkalinity needs.
15.	Prepare design summary.

EXAMPLE 8–5 **Sequencing Batch Reactor Process Design** Prepare a process design for a sequencing batch reactor process to treat a domestic wastewater with a flowrate of 7570 m^3/d with the following wastewater characteristics. The reactor mixed-liquor concentration at full volume is 3500 g/m^3 and the temperature is 12°C. The required effluent NH_4-N concentration = 1.0 g/m^3. Primary treatment is not used.

Wastewater characteristics:

Constituent	Concentration, mg/L
BOD	220
sBOD	80
COD	485
sCOD	160
rbCOD	80
TSS	240
VSS	220
TKN	35
NH_4-N	25
TP	6
Alkalinity	200 as $CaCO_3$
bCOD/BOD ratio	1.6

Note: g/m^3 = mg/L.

Design conditions and assumptions:
1. Use 2 tanks
2. Total liquid depth when full = 6 m
3. Decant depth = 20 percent of tank depth
4. SVI = 150 mL/g
5. Ammonia oxidized (NO_X) = 80 percent of TKN
6. Use kinetic coefficients in Table 8–14
7. bCOD = 1.6(BOD)

Solution

1. Develop wastewater characteristics needed for process design.
   a. Determine bCOD.

$$\text{bCOD} = 1.6(220\ g/m^3) = 352\ g/m^3$$

   b. Determine nbVSS concentration using Eq. (8–7), (8–8), (8–9).

$$\text{bsCOD} = 1.6(\text{sBOD})$$

$$\text{bsCOD} = 1.6(80\ g/m^3) = 128\ g/m^3$$

$$\text{nbsCOD}_e = \text{sCOD} - \text{bsCOD} = (160 - 128)\ g/m^3 = 32\ g/m^3$$

$$\text{nbpCOD} = \text{COD} - \text{bCOD} - \text{nbsCOD} = (485 - 352 - 32)\ g/m^3 = 101\ g/m^3$$

$$\text{VSS}_{\text{COD}} = \frac{\text{TCOD} - \text{sCOD}}{\text{VSS}}$$

$$\text{VSS}_{\text{COD}} = \frac{(485 - 160)\ \text{g/m}^3}{220\ \text{g/m}^3} = 1.48\ \text{g COD/g VSS}$$

$$\text{nbVSS} = \frac{\text{nbpCOD}}{\text{VSS}_{\text{COD}}} = \frac{101\ \text{g/m}^3}{1.48\ \text{g COD/g VSS}} = 68.2\ \text{g/m}^3$$

c. Determine iTSS.

$$\begin{aligned}\text{iTSS} &= \text{TSS}_o - \text{VSS}_o\\ &= (240 - 220)\ \text{g/m}^3 = 20\ \text{g/m}^3\end{aligned}$$

2. Determine SBR operating cycle.
The total cycle time (T_c) consists of fill (t_F), react/aerate (t_A), settle (t_S), and decant (t_D). An idle time (t_I) can also be added. Thus, the total cycle time $T_C = t_f + t_A + t_S + t_D + t_I$. At least 2 tanks are needed so that when one tank is in the fill period (t_F), the following periods are occurring in the other tank: aeration t_A, settling t_S, and decant t_D cycles. No idle time is included in this example.

$t_F = t_A + t_S + t_D$

Select period times:

Assume: $t_A = 2.0$ h
$t_S = 0.50$ h
$t_D = 0.50$ h
$t_I = 0$

Then $t_F = 2.0 + 0.50 + 0.50 = 3.0$ h for each tank (Note: Some aeration may also be done in the fill period, but if aeration occurs over more than 50% of the fill time, filamentous bulking may be encouraged.)

Total cycle time $T_C = t_f + t_A + t_S + t_D = 6.0$ h

$$\text{Number of cycles/tank}\cdot\text{d} = \frac{(24\ \text{h/d})}{(6\ \text{h/cycle})} = 4$$

$$\begin{aligned}\text{Total number of cycles/d} &= (2\ \text{tanks})\left[\frac{(4\ \text{cycles/d})}{\text{tank}}\right]\\ &= 8\ \text{cycles/d}\end{aligned}$$

$$\text{Fill volume/cycle} = \frac{(7570\ \text{m}^3/\text{d})}{(8\ \text{cycles/d})} = 946.3\ \text{m}^3/\text{fill}$$

3. Determine the tank volume and overall hydraulic retention time τ.
Full liquid depth = 6.0 m
Decant depth = 0.2 (6.0 m) = 1.2 m

$$\text{V}_\text{T} = \frac{\text{V}_\text{F}/\text{tank}}{0.2} = \frac{(946.3\ \text{m}^3/\text{tank})}{0.2} = 4732\ \text{m}^3/\text{tank}$$

$$\text{Overall}\ \tau = \frac{2\ \text{tanks}(4732\ \text{m}^3/\text{tank})(24\ \text{h/d})}{(7570\ \text{m}^3/\text{d})} = 30.0\ \text{h}$$

4. Determine the SRT.

a. Use Eqs. (8–20), (8–21), and (7–57) in Table 8–10 to obtain a relationship that can be used to solve for ($P_{X,\,\text{TSS}}$) SRT.

$$(P_{X,\text{TSS}})\text{SRT} = \frac{QY_\text{H}(S_o - S)SRT}{[1 + b_\text{H}(\text{SRT})](0.85)}$$

$$+ Q(\text{nbVSS})\text{SRT} + \frac{QY_n(\text{NO}_x)\text{SRT}}{[1 + b_n(\text{SRT})](0.85)}$$

$$+ \frac{(f_d)(b_\text{H})Q(Y_\text{H})(S_o - S)\text{SRT}^2}{[1 + b_\text{H}(S\text{RT})](0.85)} + Q(\text{TSS}_o - \text{VSS}_o)\text{SRT}$$

$$\begin{aligned}(P_{X,\text{TSS}})\text{SRT} &= (V)(X_\text{MLSS}) = (4732\text{ m}^3)(3500\text{ g/m}^3)\\ &= 16{,}562{,}000\text{ g}\end{aligned}$$

b. Develop input data to solve the above relationship for SRT:

nbVSS = 68.2 g/m^3 (from Step 1b)

Assume $S_o \approx S_o - S$

S_o = bCOD = 352 g/m^3 (Step 1a)
Q = (7570 m^3/d)/2 tanks = 3785 m^3/tank·d
iTSS$_o$ = TSS$_o$ − VSS$_o$ = 20 g/m^3 (Step 1c)
NO$_x$ = (0.80)(35 g TKN/m^3) = 28 g/m^3

Kinetic coefficients from Table 8–14,

Y = 0.45 g VSS/g bCOD
$b_{12°\text{C}} = 0.12\text{ g/g·d}(1.04)^{12-20} = 0.088\text{ g/g·d}$
Y_n = 0.15 g VSS/g NO$_x$

The nitrifier endogenous decay rate is higher during aeration (0.17 g/g·d at 20°C) and reduced during non-aerobic periods (0.07 at 20°C per Chap. 7, Sec. 7–9). Thus, a weighted average endogenous decay rate is determined.
Aerobic:
$b_{n,12°\text{C}} = 0.17\text{ g/g·d}(1.029)^{12-20} = 0.135\text{ g/g·d}$
Anoxic:
$b_{n,12°\text{C}} = 0.07\text{ (g/g·d)}(1.029)^{12-20} = 0.056\text{ g/g·d}$
Average:
$b_{n,12°\text{C}} = 0.135\text{ g/g·}(t_A/T_C) + 0.056\,(1 - t_A/T_C)\text{ g/g·d}$

$$\frac{t_A}{T_C} = \frac{2}{6} = 0.33$$

Average $b_{n,12°\text{C}}$ = (0.135 g/g·d)(0.33) + (0.056 g/g·d)(0.67) = 0.082 g/g·d

f_d = 0.15 g/g

Substituting values and calculations for above equation yields

$$16{,}562{,}000\text{ g} = \frac{(3785\text{ m}^3\text{/d})(0.45\text{ g VSS/g bCOD})(352\text{ g/m}^3)(\text{SRT})}{[1 + (0.088\text{ g/g·d})(\text{SRT})](0.85)}$$

$$+ (3785\text{ m}^3\text{/d})(68\text{ g/m}^3)(\text{SRT})$$

$$+ \frac{(3785\text{ m}^3\text{/d})(0.15\text{ g/g·d})(28\text{ g/m}^3)(\text{SRT})}{[1 + (0.082\text{ g/g·d})(\text{SRT})](0.85)}$$

$$+ \frac{(0.15\text{ g/g})(0.088\text{ g/g·d})(0.45\text{ g VSS/g bCOD})(3785\text{ m}^3\text{/d})(352\text{ g/m}^3)\text{SRT}^2}{[1 + (0.088\text{ g/g·d})(\text{SRT})](0.85)}$$

$$+ (3785\text{ m}^3\text{/d})(20\text{ g/m}^3)(\text{SRT})$$

Solve for SRT (use spreadsheet with solver or solve by successive trials)
SRT = 26.5 d

5. Determine MLVSS concentration.
   a. Solve Eq. (7–56) in Table 8–10 (SRT = 26.5 d) ($S_o \approx S_o - S$).

$$(P_{X,\mathrm{VSS}})\mathrm{SRT} = V_T(X_{\mathrm{MLVSS}})$$

$$(P_{X,\mathrm{VSS}})\mathrm{SRT} = \frac{Q(Y_{\mathrm{H}})(S_o - S)\mathrm{SRT}}{1 + b_{\mathrm{H}}(\mathrm{SRT})} + Q(\mathrm{nbVSS})\mathrm{SRT}$$

$$+ \frac{QY_n(\mathrm{NO}_x)\mathrm{SRT}}{1 + b_n(\mathrm{SRT})} + \frac{(f_d)(b_{\mathrm{H}})(Q)(Y_{\mathrm{H}})(S_o - S)\mathrm{SRT}^2}{1 + b_{\mathrm{H}}(\mathrm{SRT})}$$

$$= \frac{(3785\ \mathrm{m^3/d})(0.45\ \mathrm{g\ VSS/g\ bCOD})(352\ \mathrm{g/m^3})(26.5\ \mathrm{d})}{[1 + (0.088\ \mathrm{g/g{\cdot}d})(26.5\ \mathrm{d})]}$$

$$+ (3785\ \mathrm{m^3/d})(68\ \mathrm{g/m^3})(26.5\ \mathrm{d})$$

$$+ \frac{(3785\ \mathrm{m^3/d})(0.15\ \mathrm{g\ VSS/g\ NO}_x)(28\ \mathrm{g/m^3})(26.5\ \mathrm{d})}{[1 + (0.082\ \mathrm{g/g{\cdot}d})(26.5\ \mathrm{d})]}$$

$$+ \frac{(0.15\ \mathrm{g/g})(0.088\ \mathrm{g/g{\cdot}d})(0.45\ \mathrm{g\ VSS/g\ COD})(3785\ \mathrm{m^3/d})(352\ \mathrm{g/m^3})(26.5\ \mathrm{d})^2}{[1 + (0.088\ \mathrm{g/g{\cdot}d})(26.5\ \mathrm{d})]}$$

$$= 13{,}399{,}320\ \mathrm{m^3{\cdot}g/m^3} = V_T(X_{\mathrm{MLVSS}})$$

$$\mathrm{V_T} = 4732\ \mathrm{m^3}\ \text{(Step 3)}$$

$$\mathrm{V_T(X_{MLVSS})} = (4732\ \mathrm{m^3})(\mathrm{X_{MLVSS}})$$

$$13{,}399{,}320\ (\mathrm{m^3{\cdot}g/m^3}) = (4732\ \mathrm{m^3})(X_{\mathrm{MLVSS}})$$

$$\mathrm{X_{MLVSS}} = 2832\ \mathrm{g/m^3}$$

   b. Determine the fraction of MLVSS.

$$\frac{X_{\mathrm{MLVSS}}}{X_{\mathrm{MLSS}}} = \frac{(2832\ \mathrm{g/m^3})}{(3500\ \mathrm{g/m^3})} = 0.81$$

6. Determine amount of $\mathrm{NH_4}$-N oxidized (NO_x).
   Nitrogen balance (Eq. 8–24) in Table 8–10

$$\mathrm{NO}_x = \mathrm{TKN}_o - \mathrm{N}_e - 0.12\,P_{X,\mathrm{bio}}/Q$$

$$P_{X,\mathrm{bio}} = [\text{Items } A + B + C \text{ in Eq. } (8\text{–}20)]$$

$$P_{X,\mathrm{bio}} = \frac{QY_{\mathrm{H}}(S_o - S)}{1 + b_{\mathrm{H}}(\mathrm{SRT})} + \frac{QY_n(NO_x)}{1 + b_n(\mathrm{SRT})} + \frac{(f_d)(b_{\mathrm{H}})QY_{\mathrm{H}}(S_o - S)\mathrm{SRT}}{1 + b_{\mathrm{H}}(\mathrm{SRT})}$$

$$= \frac{(3785\ \mathrm{m^3/d})(0.45\ \mathrm{g\ VSS/g\ bCOD})(352\ \mathrm{g/m^3})}{[1 + (0.088\ \mathrm{g/g{\cdot}d})(26.5\ \mathrm{d})]}$$

$$+ \frac{(3785\ m^3/\mathrm{d})(0.15\ \mathrm{g\ VSS/g\ NO_x})(28\ \mathrm{g/m^3})}{[1 + (0.082\ \mathrm{g/g{\cdot}d})(26.5\ \mathrm{d})]}$$

$$+ \frac{(0.15\ \mathrm{g/g})(0.088\ \mathrm{g/g{\cdot}d})(3785\ \mathrm{m^3/d})(0.45\ \mathrm{g\ VSS/g\ COD})(352\ \mathrm{g/m^3})(26.5\ \mathrm{d})}{[1 + (0.088\ \mathrm{g/g{\cdot}d})(26.5\ \mathrm{d})]}$$

$$= 247{,}995\ \mathrm{g/d} = 248.0\ \mathrm{kg/d}$$

$$\mathrm{NO}_x = 35.0 - 1.0 - \frac{(0.12)(248.0\ \mathrm{kg/d})(10^3\ \mathrm{g/1\ kg})}{(3785\ \mathrm{m^3/d})}$$

$$= (35.0 - 1.0 - 7.9)\ \mathrm{g/m^3}$$

$\mathrm{NO}_x = 26.1\ \mathrm{g/m^3}$ Note: iteration by substituting 26.1 in place of the assumed 28 $\mathrm{g/m^3}$ for NO_x in item B above results in $\mathrm{NO}_x = 26.1\ \mathrm{g/m^3}$.

7. Check the degree of nitrification to determine whether NH_4-N will be removed to a level of 1.0 g/m^3 in a 2-h aeration period.
   a. Determine the amount of oxidizable N available. (Assumes all organic nitrogen is converted to NH_4-N)

$$NO_x = 26.1\ g/m^3 = NH_4\text{-N in feed flow that can be oxidized}$$

Oxidizable NH_4-N added/cycle:

$$\begin{aligned} V_F(NO_x) &= 946.3\ m^3/\text{cycle}\ (26.1\ g/m^3) \\ &= 24{,}698\ g/\text{fill} \end{aligned}$$

NH_4-N remaining before fill $= V_s(N_e)$

$$N_e = 1.0\ g/m^3\ NH_4\text{-N}$$

$$\begin{aligned} V_s(N_e) &= N_e(V - V_F) \\ &= (1.0\ g/m^3)[(4732 - 946.3)\ m^3] \\ &= 3785.7\ g \end{aligned}$$

Total oxidizable N at beginning of cycle = (24,698 + 3785.7) g = 28,483.7 g

$$\text{Initial concentration} = N_o = \frac{28{,}483.7\ g}{V_T} = \frac{28{,}483.7\ g}{4732\ m^3} = 6.0\ g/m^3$$

   b. Determine the reaction time.
   Using Eq. (8–53), the react time (aeration) after fill to achieve the desired NH_4-N concentration can be calculated. First, the nitrifier concentration must be determined.

$$K_{NH_4} \ln\left(\frac{N_o}{N_t}\right) + (N_o - N_t) = X_n\left(\frac{\mu_{\max,AOB}}{Y_n}\right)\left(\frac{S_o}{K_{o,AOB} + S_o}\right)t$$

      i. Nitrifier concentration.

$$\begin{aligned} X_n &= \frac{Q(Y_n)(NO_x)SRT}{[1 + b_n(SRT)]V} \\ &= \frac{(3785\ m^3/d)(0.15\ g\ VSS/g\ NH_4\text{-N})(26.1\ g/m^3)(26.5\ d)}{[1 + (0.082\ g/g\cdot d)(26.5\ d)](4732\ m^3)} \\ &= 26.1\ g/m^3 \end{aligned}$$

Use kinetic coefficients from Table 8–14.

$$\mu_{m,12°C} = 0.90\ g/g\cdot d\,(1.072)^{12-20} = 0.52\ g/g\cdot d$$

$$K_{NH_4,12°C} = 0.50\ g/m^3$$

$$K_o = 0.50\ g/m^3$$

      ii. Determine the time for reaction.

Solve for t for $N_o = 6.0\ g/m^3$, $N_e = 1.0\ g/m^3$

$$0.50\ln\left[\frac{(6.0\,\text{g/m}^3)}{(1.0\,\text{g/m}^3)}\right] + [(6.0 - 1.0)\,\text{g/m}^3] = (26.1\,\text{g/m}^3)\left[\frac{(0.52\,\text{g/g·d})}{(0.15\,\text{g/g})}\right]\left(\frac{2.0}{0.5 + 2.0}\right)t$$

$$t = 0.08\,\text{d} = 1.95\,\text{h}$$

c. Determine the aeration time.
Required aeration time = 1.95 h
Aeration time selected was 2.0 h; therefore, aeration time is satisfactory.

8. Determine the decant pumping rate.
Decant volume = fill volume

$$V_F = 946.3\,\text{m}^3$$

Decant time = 30 min

$$\text{Pumping rate} = \frac{946.3\,\text{m}^3}{30\,\text{min}} = 31.5\ \text{m}^3/\text{min}$$

9. Determine total oxygen required/tank using Eq. (8–23).

$$\begin{aligned} R_o &= Q(S_o - S) - 1.42P_{X,\text{bio}} + 4.57Q(\text{NO}_x) \\ &= (3785\,\text{m}^3/\text{d})(352\,\text{g/m}^3)(1\,\text{kg}/10^3\,\text{g}) - 1.42(248.0\,\text{kg/d}) \\ &\quad + 4.57(26.1\,\text{g/m}^3)(3785\,\text{m}^3/\text{d})(1\,\text{kg}/10^3\,\text{g}) \end{aligned}$$

$$R_o = (1332 - 352.1 + 451.5)\,\text{kg/d} = 2136\,\text{kg/d}$$

Number of cycles/d = 4
Oxygen required per cycle

$$= \frac{(2136\ \text{kg/d})}{(4\ \text{cycles/d})} = 534\ \text{kg O}_2/\text{cycle}$$

Aeration time/cycle = 2 h

$$\text{Average oxygen transfer rate} = \frac{(534\,\text{kg O}_2/\text{cycle})}{(2\ \text{h/cycle})} = 267\ \text{kg O}_2/\text{h}$$

Note: The oxygen demand will be much higher at the beginning of the aeration period, so the aeration system oxygen transfer capacity must be higher than the average oxygen transfer rate. The oxygen transfer rate should be multiplied by a factor of 2.0 to 3.0 to provide sufficient oxygen transfer at the beginning of the cycle and to handle peak loads.

10. Determine sludge production using Eq. (7–57). (MLSS = X_{TSS})

$$\begin{aligned} P_{X,\text{TSS}} &= \frac{(V)(\text{MLSS})}{\text{SRT}} \\ &= \frac{(2\,\text{tanks})(4732\,\text{m}^3/\text{tank})(3500\,\text{g/m}^3)(1\,\text{kg}/10^3\,\text{g})}{26.5\,\text{d}} \\ &= 1250\,\text{kg/d} \end{aligned}$$

$$\begin{aligned}\text{bCOD removed} &= (7570\,\text{m}^3/\text{d})(352\,\text{g/m}^3)(1\,\text{kg}/10^3\,\text{g})\\ &= 2664\,\text{kg/d}\end{aligned}$$

$$\text{BOD removed} = \frac{(2664\,\text{kg bCOD/d})}{(1.6\,\text{kg bCOD/kg BOD})} = 1665\,\text{kg/d}$$

$$\text{Observed yield, g TSS/g BOD} = \frac{(1250\text{ kg TSS/d})}{(1665\text{ kg BOD/d})} = \frac{0.75\,\text{g TSS}}{\text{g BOD}}$$

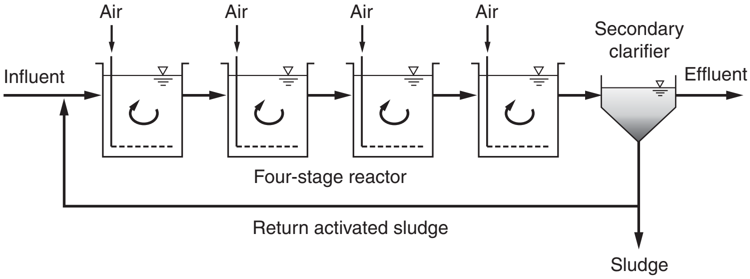
$$\begin{aligned}\text{Observed yield, g VSS/g BOD} &= \left(\frac{0.75\,\text{g TSS}}{\text{g BOD}}\right)\left(\frac{0.81\,\text{g VSS}}{\text{g TSS}}\right)\\ &= \frac{0.61\,\text{g VSS}}{\text{g BOD}}\end{aligned}$$

thus a higher yield results.

Design parameter	Unit	Value
Average flowrate	m^3/d	7570
Average BOD load	kg/d	1665
Average TKN load	kg/d	265
Number of tanks	Number	2
Fill time	h	3.0
React time	h	2.0
Total aeration time	h	2.0
Settle time	h	0.5
Cycle time	h	6.0
Total SRT	d	26.5
Tank volume	m^3	4732
Fill volume/cycle	m^3	946.3
Fill volume/tank volume	Ratio	0.2
Decant depth	m	1.2
Tank depth	m	6.0
MLSS	g/m^3	3500
MLVSS	g/m^3	2832
Decant pumping rate	m^3/min	31.5
Sludge production	kg TSS/d	1250
Observed yield	kg VSS/kg BOD	0.61
	kg TSS/kg BOD	0.75
Average oxygen required per tank/cycle	kg/d	534
Average O_2 transfer rate	kg/h	267

Figure 8–22

Schematic diagram of a staged activated sludge process.



Staged Activated Sludge Process Design

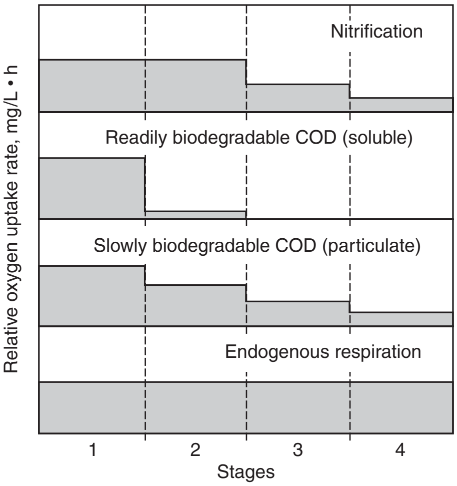
Important process variables specific to the staged activated sludge process design are the (1) oxygen demand, (2) oxygen demand distribution, and (3) NH_4-N concentration.

Oxygen Demand in Staged Processes. The oxygen demand varies in staged complete-mix reactor designs and can be high enough in the first stage to challenge the volumetric oxygen transfer capability of aeration equipment. With high-density fine bubble aeration diffusers, such as membrane aeration panels described in Sec. 5–12 in Chap. 5, oxygen transfer rates of 100 to 150 mg/L·h are possible, with some manufacturers claiming higher rates. The changes in oxygen uptake rates (OURs) in each stage of a four-stage activated sludge process (defined as a function of oxygen needed for nitrification, rbCOD removal, particulate degradable COD, and endogenous respiration) are depicted on Fig. 8–23. Most of the rbCOD will be consumed in the first stage, and the OUR for pCOD degradation will decrease from stage to stage as a function of the degradation kinetics. The nitrification rate could be at a maximum zero-order kinetic rate in the first and second stages, due to higher NH_4-N concentrations in the early stages. Oxygen demand for endogenous respiration will be relatively constant from stage to stage.

Oxygen Demand Distribution. The oxygen demand distribution may be estimated to determine the aeration design for staged processes. The percent of the total oxygen consumption may range from 40, 30, 20, and 10 percent, respectively, for a four-stage system. One design approach that can be used to obtain an estimate of the oxygen

Figure 8–23

Changes in oxygen uptake rates for a staged activated sludge process.



demand in a staged system is to calculate the total oxygen demand as would be done for a CMAS process, and then estimate the oxygen demand distribution with consideration to the various components described above. With proper selection of the type and placement of the diffusers and by providing an air supply system with DO control in each portion of the system, the air can be provided where needed. Generally, the approach outlined above is satisfactory because during the life of the process, the oxygen demand will vary across the tank as the load changes. The effect of using a staged system as compared to a single CMAS tank for nitrification is illustrated in Example 8–6.

EXAMPLE 8–6 **Evaluation of Staged Reactors for Nitrification** Compare the steady-state nitrification performance of a four-stage activated sludge system with equal volumes per stage to that for a single-stage CMAS system with the same total volume. The hydraulic retention time is 8 h for both systems, and the same SRT is used. The CMAS system SRT is calculated for an effluent NH_4-N concentration of 0.50 g/m^3 and the amount of ammonia oxidized (NO_x) is assumed at 30.0 g/m^3. Use the kinetic coefficients given in Table 8–14, along with the following conditions:

Design conditions and assumptions:

Item	Unit	Value
Temperature	°C	16
$\mu_{max,16}$	g/g·d	0.681
K_{NH_4}	g/m^3	0.50
Y_n	g VSS/g NH_4-N	0.15
$b_{n,16}$	g/g·d	0.151
Amount of NH_4-N oxidized (NO_x)	g/m^3	30.0
Effluent NH_4-N	g/m^3	0.50
DO	g/m^3	2.0
K_o	g/m^3	0.5
RAS recycle ratio	Unitless	0.5

Note: g/m^3 = mg/L.

Solution

1. Determine the SRT value and the concentration of the nitrifying bacteria for a single-stage system, $\tau = 8\text{ h} = 0.33\text{ d}$, $N = 0.50\text{ g/m}^3$.
   a. Solve for the specific growth rate using Eq. (7–94) in Table 8–10.

$$\mu_{AOB} = \mu_{max,AOB,16}\left(\frac{S_{NH}}{S_{NH} + K_{NH}}\right)\left(\frac{S_o}{S_o + K_{o,AOB}}\right) - b_{AOB}$$

$$\mu_{AOB} = \left\{\frac{(0.681\text{ g/g·d})(0.50\text{ g/m}^3)}{[(0.50 + 0.50)\text{ g/m}^3]}\right\}\left\{\frac{(2.0\text{ g/m}^3)}{[(0.50 + 2.0)\text{ g/m}^3]}\right\} - 0.151 = 0.121\text{ g/g·d}$$

b. Solve for SRT using Eq. (7–98) in Table 8–10.

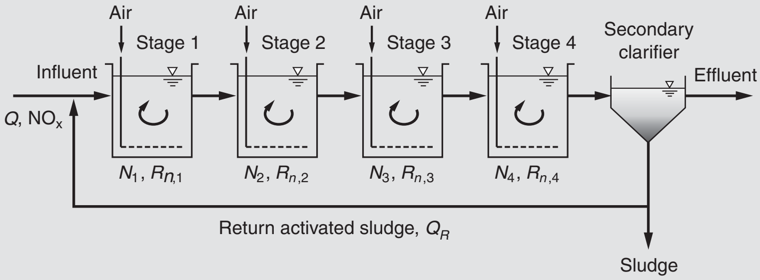
$$\text{SRT} = \frac{1}{\mu_{\text{AOB}}} = \frac{1}{0.121\ \text{g/g}\cdot\text{d}} = 8.24\ \text{d}$$

c. Solve for the concentration of nitrifying bacteria using a modified form of Eq. (7–42).

$$X_n = \frac{(\text{SRT})Y_n(\text{NO}_x)}{\tau[1 + b_n(\text{SRT})]}$$

$$= \frac{(8.24\ \text{d})(0.15\ \text{g/g})(30\ \text{g/m}^3)}{(0.33\ \text{d})[1 + (0.151\ \text{g/g}\cdot\text{d})(8.24\ \text{d})]} = 50.1\ \text{g/m}^3$$

2. Perform nitrogen mass balances for a four-stage system shown on the following figure using equal volumes per stage. The total volume of the four-stage system is equal to the volume of the CMAS system, τ/stage = 0.333 d/4 = 0.0833 d/stage.



a. For Stage 1

Accumulation = in − out + generation

$$\frac{dN_1}{dt}V = Q(\text{NO}_x) + Q_R N_4 - (Q + Q_R)N_1 - R_{n,1}V$$

The rate expression for nitrification, derived from Eq. (7–101) in Table 8–10 includes a correction for the DO concentration, and is given by

$$r_{\text{NH}_4} = \left(\frac{\mu_{\text{max,AOB}}}{Y_{\text{AOB}}}\right)\left(\frac{S_{\text{NH}_4}}{S_{\text{NH}_4} + K_{\text{NH}_4}}\right)\left(\frac{S_o}{S_o + K_{o,\text{AOB}}}\right)X_{\text{AOB}}$$

where Q = wastewater flowrate, m^3/d

NO_x = amount of available influent NH_4-N oxidized, 30 g/m^3

Q_R = recycle flowrate from stage 4, m^3/d

Q_R/Q = 0.50

N_1 = NH_4-N concentration for stage 1, g/m^3

N_4 = NH_4-N concentration for stage 4, g/m^3

$R_{n,1}$ = nitrification rate for stage 1, $\text{g/m}^3\cdot\text{d}$

X_n = nitrifying bacteria concentration, g/m^3

The nitrifying bacteria concentration is the same as that calculated for the CMAS system assuming that the same amount of NH_4-N is removed and the systems are at the same SRT.

At steady state $dN_1/dt = 0$, and

$$\text{NO}_x + Q_R/QN_4 - (1 + Q_R/Q)/N_1 - R_{n,1}V/Q = 0$$

$$\text{NO}_x + 0.5N_4 = 1.5N_1 + R_{n,1}(\tau)$$

where $\tau = 0.0833$ d, detention time of stage 1
$NO_x = 30\ g/m^3$

b. For Stage 2, use the same procedure as Stage 1.

$$V\frac{dN_2}{dt} = (Q + Q_R)N_1 - (Q + Q_R)N_2 - R_{n,2}V$$

$$1.5N_1 = 1.5N_2 + R_{n,2}(\tau)$$

c. For Stage 3

$$1.5N_2 = 1.5N_3 + R_{n,3}(\tau)$$

d. For Stage 4

$$1.5N_3 = 1.5N_4 + R_{n,4}(\tau)$$

3. $R_{n,i(i=1\text{–}4)}$ is a function of the NH_4-N concentration (N) in each stage:
For stage 1,

$$R_{n,i} = \left[\frac{(0.681\ \text{g/g·d})}{(0.15\ \text{g VSS/g NH}_4\text{-}N)}\right]\left\{\frac{N_i}{[(0.50 + N_i)\text{g/m}^3]}\right\}\left\{\frac{2.0\ \text{g/m}^3}{[(0.5 + 2.0)\ \text{g/m}^3]}\right\}(50.1\ \text{g/m}^3)$$

$$R_{n,i} = 181.96\left\{\frac{N_i}{[(0.50 + N_i)\ \text{g/m}^3]}\right\},\ \text{Where } i = 1, 2, 3, \text{or } 4 \text{ for stages } 1\text{–}4$$

4. The above equations for the four stages are solved with a spreadsheet program starting with Stage 1 either by using Solver in the Excel software or by an iterative technique. In the iterative technique the value for N_4 is assumed and N_1 is calculated. Subsequently N_2, N_3 and N_4 are calculated. Using Solver, the following effluent NH_4-N concentrations are computed for each stage for a return sludge recycle ratio of 0.50. In addition a solution is also shown for a return sludge recycle ratio of 6.0 as would be typical for an MBR system:

	NH_4-N concentration, g/m^3	
Stage	Recycle ratio = 0.50	Recycle ratio = 6.0
1	10.64	2.89
2	2.30	1.34
3	0.15	0.39
4	0.01	0.08

Comment Based on the above results, the same amount of nitrification can be achieved theoretically with the use of 4 stages for the aerobic nitrification step in a little less than 75 percent of the aeration tank volume required for a CMAS design. Thus, a staged nitrification reactor is more efficient than a CMAS reactor design and compared to the CMAS the staged system can have a lower SRT and thus lower total volume. Or for the same SRT and volume, the staged system would produce a lower average effluent NH_4-N concentration under diurnal varying ammonia loads. The above solution is also illustrative of the importance of the return activated sludge recycle ratio. The effluent NH_4-N concentration would be higher for an MBR system, which has a higher recycle ratio. The effect of the higher recycle ratio is to dilute the influent NH_4-N concentration more so that the NH_4-N concentration is lower in the first stage. Because the nitrification rate is related to the NH_4-N concentration, the rate is lower and thus the effluent concentration is higher for the same reactor volume.

Alternative Processes for BOD Removal and Nitrification

Over the past 30 years numerous activated sludge processes have been developed for the removal of organic material (BOD) and for nitrification. Some of the processes are modifications or variations of basic processes that have evolved to meet different performance objectives. Descriptions and flowsheets are presented in Table 8–18 for representative processes used for BOD removal and nitrification. The processes are grouped according to the basic reactor configuration: plug-flow, complete-mix, and sequentially operated systems. A view of a large wastewater treatment plant using plug-flow reactors is shown on Fig. 8–24 on page 792.

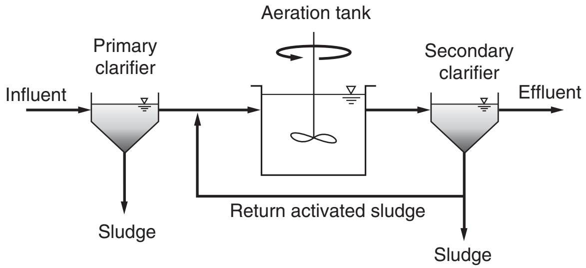
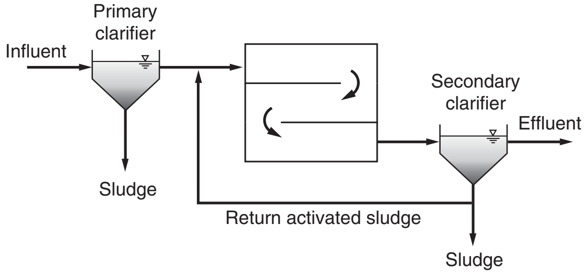
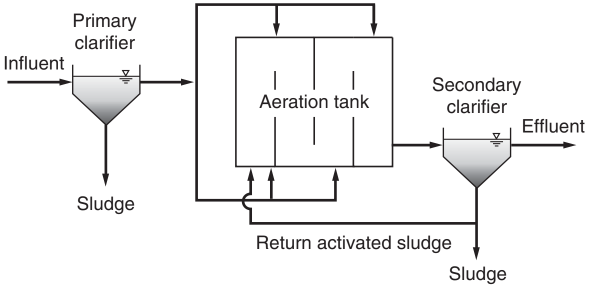
High Rate Processes. The high rate processes identified in Table 8–18 [(a) through (f)] differ in terms of their aeration configuration, aeration equipment design, solids retention time, operating mode, and ability to remove nitrogen, and some are proprietary. The contact stabilization and high-purity oxygen processes are used primarily for BOD removal only, are designed for relatively short SRTs, and require less space than other processes. These processes are particularly attractive for large municipalities where space is limited and nitrification is not needed. The conventional plug-flow, step-feed, and complete-mix processes are used for both BOD removal and nitrification and are applied over a wide range of SRTs, depending on the wastewater temperature and treatment needs. The primary motivation for using the two-sludge AB process is to reduce the amount of energy imported to the WWTP. Removal of BOD at the low SRT in the high rate first step requires less aeration energy and moves a larger amount of the influent organic material to anaerobic digestion for increased methane production.

Low Rate Processes. In contrast to the high-rate processes described above, the remaining processes in Table 8–18 [(g) through (n)] are generally operated with longer aeration times and higher SRTs, though the oxidation ditch has also been applied at lower SRTs to provide a mixing and aeration alternative to the CMAS processes using diffused aeration. Conventional extended aeration, oxidation ditch, Orbal™, and CCAS™, processes are used commonly in smaller WWTPs and employ a simpler treatment scheme by eliminating primary treatment and anaerobic digestion. An oxidation ditch and a Biolac® process are shown on Fig. 8–25 (a) and (b) respectively (on page 792) and described in Table 8–18 (h) and (k), respectively. Larger aeration tanks with longer SRTs, usually exceeding 20 d, are used.

The process approach is attractive for smaller communities where space is not an issue and less complex operation is preferred. The large aeration tank volume provides good equalization at high flow and loading occurrences, and a high-quality effluent is produced. With the exception of the conventional extended aeration process, the systems are often operated to promote denitrification in addition to nitrification. The aeration and mixing of the channel-flow processes (oxidation ditch, Orbal™, and CCAS™) require much less energy for mixing than needed for aeration so that aeration equipment design is based on meeting oxygen requirements instead of tank mixing. Less energy is required in comparison to conventional extended aeration processes in which mixing energy controls. In the past, the oxidation ditch and extended aeration processes were thought to need long SRTs to provide well-stabilized biosolids for reuse. However, with stricter regulations governing biosolids stabilization (see Chap. 14), separate aerobic digestion facilities are used to meet the requirements for reuse.

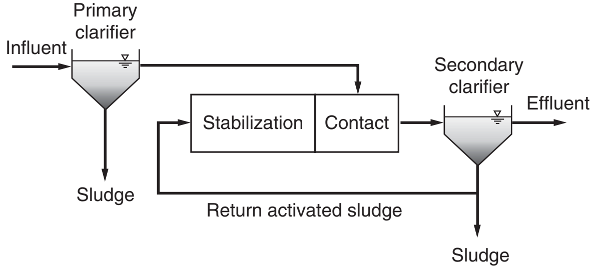
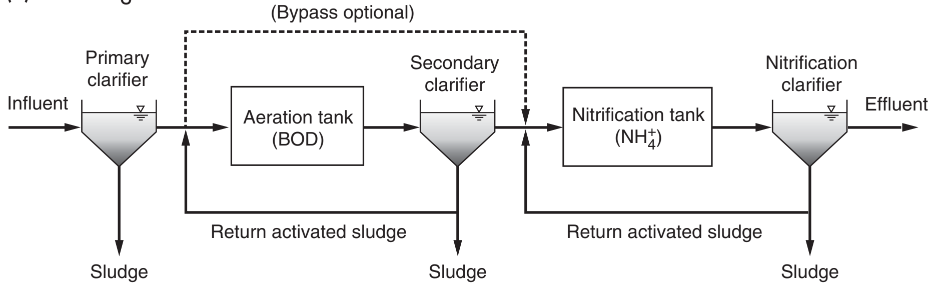
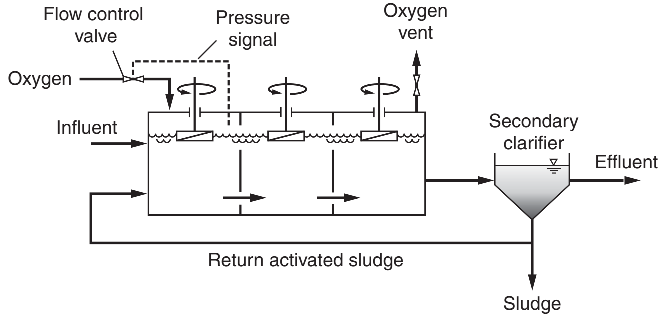
Table 8–18

Description of activated sludge processes for BOD removal and nitrification

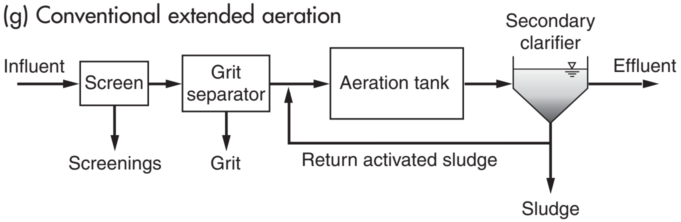
Process	Description
(a) Complete mix activated sludge (CMAS) 	The CMAS process is an application of the flow regime of a continuous-flow stirred tank reactor. Settled wastewater and recycled activated sludge are introduced typically at several points in the aeration tank. The organic load on the aeration tank, MLSS concentration, and oxygen demand are uniform throughout the tank. An advantage of the CMAS process is the dilution of shock loads that occur in the treatment of industrial wastewaters. The CMAS process is relatively simple to operate but tends to have low organic substrate concentrations (i.e., low F/M ratios) that encourage the growth of filamentous bacteria, causing sludge bulking problems.
(b) Conventional plug flow 	Conventional *plug flow* activated sludge originated from the use of narrow aeration tank widths (typically 8 to 10 m) as a result of using sidewall, coarse bubble aeration for mixing and aeration. The area provided by the limited width and the use of liquid depths in the range of 5 to 6 m (16 to 20 ft) required tank lengths typically >100 m (330 ft). Settled wastewater and return activated sludge (RAS) enter the front end of the aeration tank and typically travel through 3 to 4 channels (passes) before flow to the secondary clarifier. The oxygen demand is very high at the front of the tank and decreases with increasing tank length. The aeration system must be designed to meet the changing oxygen demand, and has been referred to as tapered aeration to reflect the change from higher aeration rates at the front of the plug flow tanks to lower aeration rates at the end. True plug flow does not exist due to back mixing caused by the aeration system. Baffle walls may be added at appropriate locations in plug flow tanks to select for a number of reactors in series with the appropriate desired volumes.
(c) Step feed 	Step feed is a modification of the conventional plug flow process in which the settled wastewater is introduced at 3 to 4 feed points in the aeration tank to equalize the F/M ratio, thus lowering peak oxygen demand. Generally, three or more parallel channels are used. Flexibility of operation is one of the important features of this process because the apportionment of the wastewater feed can be changed to suit operating conditions. The concentration of MLSS may be as high as 5000 to 9000 mg/L in the first pass, with lower concentrations in subsequent passes as more influent feed is added. The step feed process has the capability of carrying a higher solids inventory, and, thus, a higher SRT for the same volume as a conventional plug flow process. The step feed process can also be operated in the contact-stabilization mode by feeding only the last pass, and high wet weather flows can be bypassed to the last pass so that the solids load to the secondary clarifier can be minimized.

(*continued*)

Table 8–18 (*Continued*)

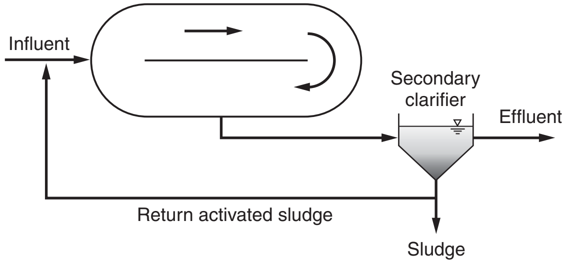
Process	Description
(d) Contact stabilization 	Contact stabilization uses two separate tanks or compartments for the treatment of the wastewater and stabilization of the activated sludge. The stabilized activated sludge is mixed with the influent (either raw or settled) wastewater in a contact zone. The contact zone detention time is relatively short (30 to 60 min), and the MLSS concentration in the contact zone is lower than that in the stabilization zone. Rapid removal of soluble BOD occurs in the contact zone and colloidal and particulate organics are captured in the activated sludge floc for degradation later in the stabilization zone. In the stabilization zone, return activated sludge (RAS) is aerated and the detention time is in the order of 1 to 2 h to maintain a sufficient SRT for sludge stabilization. Because the MLSS concentration is so much higher in the stabilization zone, the contact-stabilization process requires much less aeration volume than complete mix or conventional plug flow processes for the same SRT. The process was developed for BOD removal, and the short contact time limits the amount of soluble BOD degraded and NH_4-N oxidation. Nitrification can occur in the stabilization zone if there is sufficient volume to maintain the aerobic SRT needed for the nitrifying bacteria.
(e) Two-sludge AB Process 	The two sludge process is a two stage system using high rate activated sludge for BOD removal followed by a second stage for nitrification, which is operated at a longer SRT. A portion of the wastewater influent may be bypassed around the stage to provide BOD and suspended solids for the nitrification process to promote flocculation and solids capture in secondary clarification. Separation of the BOD removal stage from the nitrification stage reduces the aeration energy demands of the activated sludge process and directs more organic material to anaerobic digestion for increased methane production (Boehnke et al., 1997). The two stage approach also allows for removal of toxic substances in the first stage to protect the more sensitive nitrifying bacteria, but this problem is normally prevented by enforcing industrial pretreatment programs.
(f) High purity oxygen 	A staged enclosed reactor is used in the high purity oxygen activated sludge process (McWhirter, 1978). Three or four stages are generally used and the influent wastewater, RAS, and high-purity oxygen are added to the first stage. The headspace gas and mixed liquor flow concurrently from stage to stage. The oxygen partial pressure in the headspace may range from 40 to 60 percent in the first stage to 20 percent in the last stage. At high oxygen partial pressure, higher volumetric oxygen transfer rates are possible so that pure oxygen systems can have a higher MLSS concentration and operate at a shorter τ and higher volumetric organic loadings than conventional processes. The rate of oxygen addition is about 2 to 3 times greater than that by conventional aeration systems. Onsite oxygen generation equipment is needed to provide the pure oxygen gas for the process, making the process operation more complex than conventional activated sludge processes. Nitrification ability is limited with the high purity oxygen processes due to the accumulation of carbon dioxide in the gas headspace, which causes low pH in the mixed liquor (less than 6.5). Major advantages for pure oxygen systems are the reduced space requirement and greatly reduced quantities of off-gas if odor control and VOC control are required.

(g) Conventional extended aeration



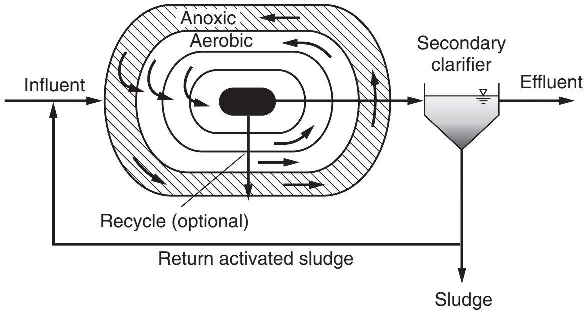
The extended aeration activated sludge process is designed with a very high SRT (20 to 30 d) and hydraulic retention times of 24 h or more. Because of the large volume needed and relatively low volumetric oxygen demand rate, the aeration equipment design is controlled by mixing needs and not oxygen demand. The process is used extensively for pre-engineered plants for small communities. Generally, primary clarification is not used. Secondary clarifiers are designed at lower hydraulic loading rates than conventional activated sludge clarifiers to better handle large flowrate variations typical of small communities. Although the biosolids are well stabilized, additional biosolids stabilization is required to permit beneficial reuse (see Chap. 14).

(h) Oxidation ditch



The oxidation ditch consists of a ring- or racetrack-shaped channel equipped with mechanical aeration and mixing devices. Screened and degritted wastewater enters the channel and is combined with the return activated sludge. The tank configuration and aeration and mixing devices promote unidirectional channel flow, so that the energy used for aeration is sufficient to provide mixing in a system with a relatively long hydraulic retention time. The aeration/mixing method used creates a velocity from 0.25 to 0.30 m/s (0.8 to 1.0 ft/s) in the channel, which is sufficient to keep the activated sludge in suspension. At these channel velocities, the mixed liquor completes a tank circulation in 5 to 15 min, and the magnitude of the channel flow is such that it can dilute the influent wastewater flowrate by a factor of 20 to 30. As a result, the process kinetics approach that of a complete mix reactor, but with plug flow along the channels. As the wastewater leaves the aeration zone, the DO concentration decreases and denitrification may occur downstream from the aeration zone. Brush-type or surface-type mechanical aerators are used for mixing and aeration (see Sec. 5–12 in Chap. 5).

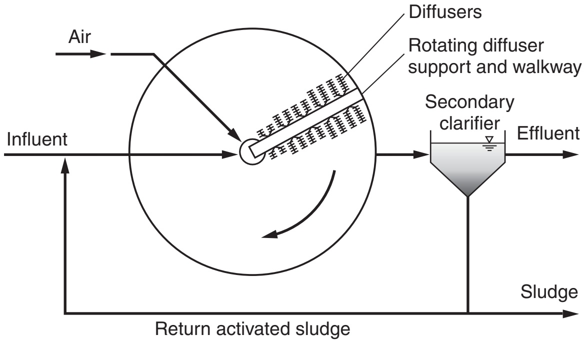
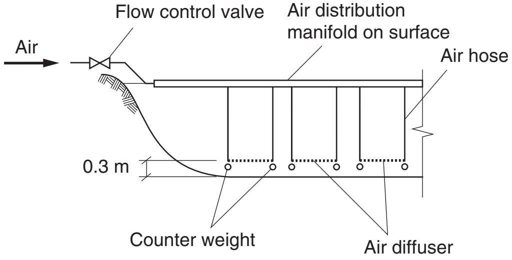
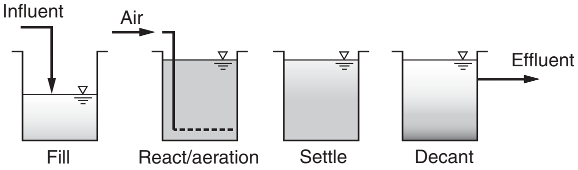
(i) Orbal®



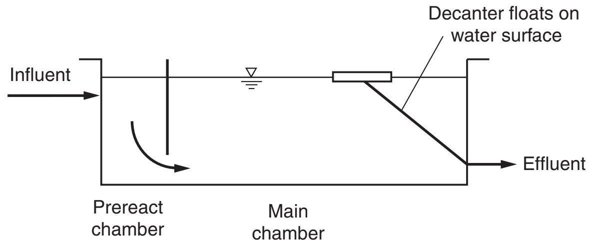
The Orbal® process is a variation of the oxidation ditch and uses a series of concentric channels within the same structure. Wastewater enters the larger outer channel and mixed liquor flows typically toward the center of the structure through at least two more channels before entering an internal clarifier or a distribution box. Disk aerators mounted on a horizontal shaft provide aeration. Channel depths range up to 4.3 m (14 ft). One version of the Orbal design (Bionutre™) limits the aeration rate in the first channel so that both nitrification and denitrification (anoxic condition) occur.

(*continued*)

Table 8–18 (*Continued*)

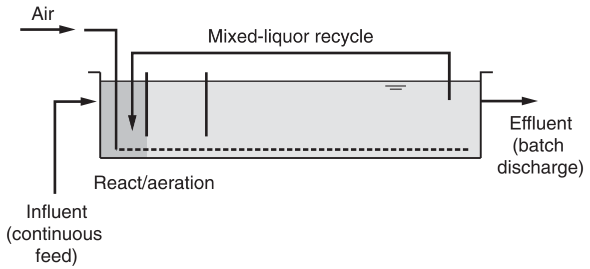
Process	Description
(j) Counter current aeration system (CCAS™)  	In the CCAS process, a unique aeration system is employed in which air diffusers are mounted at the bottom of a revolving bridge in a circular aeration tank. Because of the circulating motion of the bridge, which is moving faster than the aeration tank contents, fine bubbles are dispersed in a sweeping motion behind the traveling bridge. When the air is turned off, the movement of the diffusers creates enough mixing energy to keep the tank contents in suspension. The process is operated at a DO ranging from 0.7 to 1.0 mg/L. The low DO concentration is sufficient for nitrification at the long SRT, while allowing anoxic conditions to develop to promote denitrification. The system is normally designed with extended aeration SRTs.
(k) Biolac™ process  	Biolac is a proprietary process that combines long solids retention times with submerged aeration in earthen basins. Fine bubble membrane diffusers are attached to floating aeration chains that are moved across the basin by the air released from the diffusers. Aeration basins are typically 2.4 to 4.6 m (8 to 15 ft) deep. The process can be designed for nitrification since the SRT ranges from 40 to 70 d. The F/M ratio ranges from 0.04 to 0.1 and the MLSS range is from 1500 to 5000 g/m^3. A variation of the standard process, known as the "wave oxidation modification" allows biological nitrification and denitrification to occur simultaneously by using timers to cycle the air flowrate to each aeration chain. Either an internal or external clarifier can be used.
(l) Sequencing batch reactor (SBR)  	The SBR is a fill and draw type reactor system involving a single complete mix reactor in which all steps of the activated sludge process occur. For municipal wastewater treatment with continuous flow, at least 2 basins are used so that one basin is in the fill mode while the other goes through react, solids settling, and effluent withdrawal. An SBR goes through a number of cycles per day; a typical cycle may consist of 3-h fill, 2-h aeration, 0.5-h settle, and 0.5-h for withdrawal of supernatant. An idle step may also be included to provide flexibility at high flows. Mixed liquor remains in the reactor during all cycles, thereby eliminating the need for separate secondary sedimentation tanks. Decanting of supernatant is accomplished by either fixed or floating decanter mechanisms. SBR hydraulic retention times generally range from 18 to 30 h, based on influent flowrate and tank volume used. Aeration may be accomplished by jet aerators or coarse or fine bubble diffusers with submerged mixers (see Sec. 5–12 in Chap. 5). Separate mixing provides operating flexibility and is useful for contacting influent and mixed liquor during the fill period for anoxic or anaerobic operation period. Sludge wasting occurs normally during the aeration period.

(m) Intermittent cycle extended aeration system, ICEAS™)



The ICEAS process, developed in Australia, is another type of SBR process used for treating flowrates up to 500,000 m^3/d (130 Mgal/d). Influent wastewater is fed continuously through the same cycles of react, settle, and decant as in an SBR. The influent is fed into one side of a baffled chamber (a prereact zone) so the flow does not disturb the mixed liquor during the settling and decant period. Wastewater flows through openings at the bottom of the baffle wall and into the main react zone where BOD and nitrification occur. After aeration and settling, separated liquid is removed by an automated, time-controlled decant mechanism. Sludge wasting is also accomplished during this phase.

(n) Cyclic activated sludge system (CAAS™)



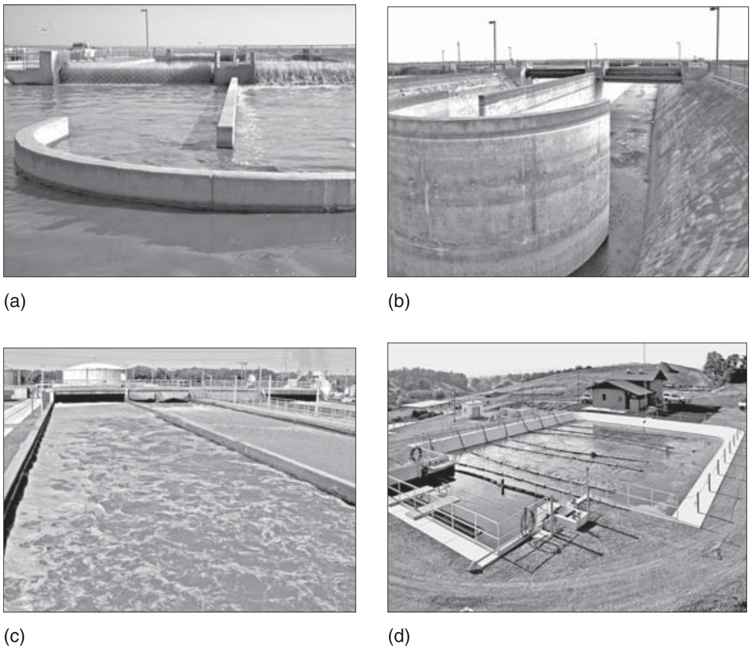
The CAAS process uses three baffled zones in an approximate volumetric proportion of 1/2/20, and mixed liquor is recycled from Zone 3 to Zone 1. Nitrification occurs because of the long SRTs used. Nitrate reduction is claimed to occur at significant levels in the sludge blanket during the settle and decant periods, as well as in the aeration period by operation at low DO concentrations. As in the ICEAS process, the influent wastewater is fed continuously while the effluent is removed on a batch basis.

Figure 8–24

View of plug-flow reactors at Owls Head Wastewater Treatment Plant, New York. (Coordinates 40.6431 N, 74.0337 W, view at altitude 900 m. (Courtesy of New York City Department of Environmental Protection.)

Figure 8–25

Views of alternative activated sludge processes: (a) oxidation ditch with brush rotors, (b) empty oxidation ditch with brush rotors and sloped sides (note vertical sides are used more commonly), (c) oxidation ditch with vertical turbine surface aerator and mixer, and (d) Biolac® process with contiguous clarifier.


(a)
(b)
(c)
(d)

Sequential Processes. Sequentially operated activated sludge processes that do not use separate tanks for liquid-solids separation are also described in Table 8–18 [(l) through (n)]. The processes include the sequencing batch reactor, batch decant reactor, and the cyclic activated sludge system. Operation is based usually on long τ and SRT values. The processes are attractive to small communities because of the simplicity of operation and relatively low cost. Sequentially operated processes are also adaptable to nitrogen removal, as discussed in Sec. 8–7.

Process Design Parameters. Typical parameters used for the design and operation of various activated sludge processes are presented in Table 8–19. Reactor configurations

Table 8–19

Typical design parameters for commonly used activated sludge processes

Process name	Type of reactor	SRT, d	F/M, kg BOD/kg MLVSS·d	Volumetric loading: lb BOD/1000 ft^3·d	Volumetric loading: kg BOD/m^3·d	MLSS, mg/L	Total τ, h
High-rate aeration (first step in AB Process)	CMAS or plug flow	0.5–2	1.5–2.0	75–150	1.2–2.4	500–1500	1–2
Contact stabilization	CMAS or plug flow	5–10	0.2–0.6	60–75	1.0–1.3	1000–3000[a] 6000–10,000[b]	0.5–1[a] 2–4[b]
High-purity oxygen	Staged	1–4	0.5–1.0	80–200	1.3–3.2	2000–4000	1–3
Conventional plug flow	Plug flow	3–15	0.2–0.4	20–40	0.3–0.7	1000–3000	4–8
Step feed	Plug flow or staged	3–15	0.2–0.4	40–60	0.7–1.0	1500–4000	3–5
Complete mix	CMAS	3–15	0.2–0.6	20–100	0.3–1.6	1500–4000	3–6
Extended aeration	CMAS or plug flow	20–40	0.04–0.1	5–15	0.1–0.3	2000–4000	20–30
Oxidation ditch	CMAS + plug flow	15–30	0.04–0.1	5–15	0.1–0.3	3000–5000	15–30
Batch decant (ICEAS, CAAS)	Plug flow	12–30	0.04–0.1	5–15	0.1–0.3	2000–5000	20–40
Sequencing batch reactor	Batch	15–30	0.04–0.1	5–15	0.1–0.3	2000–5000	15–40
Counter current aeration system (CCAS™)	Plug flow	15–30	0.04–0.1	5–10	0.1–0.3	2000–4000	15–40

[a] MLSS and detention time in contact basin.

[b] MLSS and detention time in stabilization basin.

may be completely mixed, staged, or plug flow. The oxidation ditch is a combination of complete mix and plug flow. The recirculation flowrate in the ditch provides a flowrate that is typically 20–30 times the influent flowrate so the dilution factor approaches a complete mix substrate condition. In addition, plug flow kinetics conditions are approached as the flow moves down the ditch channels away from the mixed aeration zone.

Process Selection Considerations. Selection of an activated sludge process for BOD removal and nitrification is a function of many considerations including specific site constraints, compatibility with the existing process, compatibility with existing equipment, present and future treatment needs, level of capability of the operating staff, capital costs, and operating costs. Significant features and limitations of the various activated sludge process alternatives that affect process selection in certain applications are summarized in Table 8–20.

Table 8–20

Advantages and limitations of activated sludge processes for BOD removal and nitrification

Process	Advantages	Limitations
Complete mix	Common, proven process	Susceptible to filamentous sludge bulking
	Adaptable to many types of wastewater	
	Large dilution capacity for shock and toxic loads	
	Uniform oxygen demand	
	Design is relatively uncomplicated	
	Suitable for all types of aeration equipment	
Conventional plug flow	Proven process	Design and operation for tapered aeration is more complex
	Adaptable to many operating schemes including step feed, selector design, and anoxic/aerobic processes	May be difficult to match oxygen supply to oxygen demand in first pass
		More complex operation
Step feed	Distributes load to provide more uniform oxygen demand	Flow split is not usually measured or known accurately
	Peak wet weather flows can be bypassed to the last pass to minimize high clarifier solids loading	More complicated design for process and aeration system
	Flexible operation	
	Adaptable to many operating schemes including anoxic/aerobic processes	
Contact stabilization	Requires smaller aeration volume	Has little or no nitrification capability
	Handles wet weather flows without loss of MLSS	Operation somewhat more complex
Two-sludge AB process	Requires less aeration tank volume than conventional plug flow	Requires process control to assure sufficient solids in nitrification step
	Uses less aeration energy	Requires two clarifiers
	Provides for more methane production with anaerobic digestion	Less resilient for high peak flows and loads
High purity oxygen	Requires relatively small aeration tank volume	Limited capability for nitrification
	Uses more energy than air aeration	More complex equipment to install, operate, and maintain
	Generally produces good settling sludge	*Nocardioform* foaming
	Operation and DO control are relatively uncomplicated	High peak flows can disrupt operation by washing out MLSS
Extended aeration	High quality effluent possible	Aeration energy use is high
	Relatively uncomplicated design and operation	Relatively large aeration tanks
	Capable of treating highly variable loads and intermittent toxic loads	Adaptable mostly to small plants
	Well stabilized sludge; low biosolids production	Filamentous sludge bulking is possible

(*continued*)

Table 8–20 (*Continued*)

Process	Advantages	Limitations
Oxidation ditch	Highly reliable process; simple operation	Large structure, greater space requirement
	Capable of treating shock/toxic loads without affecting effluent quality	Low F/M bulking is possible
	Economical process for small plants	Requires more aeration energy than conventional CMAS and plug flow treatment
	Uses less energy than extended aeration	Plant capacity expansion is more difficult
	Adaptable to nutrient removal	
	High quality effluent possible	
	Well stabilized sludge; low biosolids production	
Sequencing Batch Reactor	Process is simplified; final clarifiers and RAS pumping are not required	Process design and control more complicated
	Modular construction possible	High peak flows can disrupt operation
	Operation is flexible; nutrient removal can be accomplished by operational changes	Batch discharge may require equalization prior to filtration and disinfection
	Can be operated as a selector process to minimize sludge bulking potential	Higher maintenance skills required for instruments, monitoring devices, and automatic valves
	Quiescent settling enhances solids separation (low effluent TSS)	Some designs use less efficient aeration devices
	Economical for small plants	Volumetric tankage requirements can be high
Counter current aeration	High quality effluent possible	Fine screening is required to prevent diffuser fouling
	Oxygen transfer efficiencies are higher than convention aeration systems	Process is proprietary
	Well stabilized sludge; low biosolids production	Significant down-time of aeration unit for maintenance will affect plant performance
	Process design can be modified to accommodate nutrient removal	Good operator skills required

8–7 PROCESSES FOR BIOLOGICAL NITROGEN REMOVAL

Nitrogen removal is often required before discharging treated wastewater to sensitive water bodies (to prevent eutrophication), or for groundwater recharge or other reuse applications. Processes for BOD removal and nitrification, presented previously in Sec. 8–6, were based on the fundamental principles presented in Chap. 7 and Secs. 8–2 and 8–3. In this section, an additional treatment step, *biological denitrification*, is considered and information is presented that is specific to the nitrogen removal process. Nitrogen removal can be either an integral part of the biological treatment system or an add-on process to an existing treatment plant.

The purpose of this section, as in the previous section, is to illustrate in detail the design procedure for biological processes used to remove nitrogen from wastewater. The material on nitrogen removal presented in this section is organized into the following topics: (1) process development, (2) overview of biological nitrogen removal processes, (3) general process design consideration, (4), (5), (6) process designs for three different types of biological nitrogen removal processes, (7) external carbon addition, (8) summary of alternative processes used for biological nitrogen removal, and (9) process control and performance.

Process Development

The antecedents of biological nitrogen removal date back to observations in the 1940s in studies on rising sludge in secondary clarifiers due to nitrification and denitrification (Sawyer and Bradney, 1945). The first WWTP process scheme for biological nitrogen removal involved the addition of a tank whose contents were mixed, but unaerated, (post-anoxic) following an aerobic nitrification tank. This arrangement was first proposed, implemented, and discussed by Wuhrman in a series of conference proceedings from 1960 to 1962 (Wuhrman, 1964; Bishop et al., 1976). Wuhrman noted that denitrifying bacteria were abundant in activated sludge and was the first to recognize *simultaneous nitrification and denitrification* in activated sludge. He also noted that denitrification occurred as a result of a low mixed liquor DO, which was insufficient to penetrate the interior of floc particles, where nitrate reduction could occur.

Building on Wuhrman's observations on denitrification at low DO concentrations, Ludzack and Ettinger studied the effect of recycling mixed liquor from a nitrifying aerobic zone upstream to a mixed nonaerated zone so that influent BOD could be used for denitrification. This was the first preanoxic nitrogen removal process (Ludzack and Ettinger, 1962). Later Balakrishnan and Eckenfelder (1970) provided recycle of nitrate in secondary clarifier effluent to an upstream unaerated, mixed stabilization zone of a contact-stabilization process prior to the aerobic contact zone and secondary clarifier. A disadvantage of this approach was the higher hydraulic load to the secondary clarifier because of the nitrate recycle flow. Barnard (1974) provided both preanoxic and postanoxic denitrification zones in his development of the Bardenpho Process. Although the Bardenpho process could produce effluent total nitrogen concentrations below 3.0 mg/L, such low levels of nitrogen were not deemed necessary in the 1970s; thus, the post anoxic zone was not necessary and it was more common to just use the preanoxic/aerobic steps. This anoxic/aerobic sequence with recycle from the aeration zone to the preanoxic zone was termed the Modified Ludzack Ettinger process or MLE process.

In addition to this work with distinctive anoxic and aerobic zones, others investigated cyclic aerobic/mixing modes for nitrogen removal by nitrification and denitrification in single or staged reactors (Bishop et al., 1976). During this same time period, the ability to achieve high nitrogen removal in a single-tank oxidation ditch system was documented by Matsche from full-scale plant results at the Vienna Blumental wastewater treatment facility (Matsche, 1972). All of the above nitrogen removal processes were applied in single-sludge activated sludge systems, which means that only one solids-liquid separation step (secondary clarification) is used in the process. However, in the late 1960s a three sludge system was tested on a lab scale (Barth et al., 1968) and then demonstrated in a pilot plant (Heidman et al., 1975) to achieve high nitrogen removal efficiency with effluent total nitrogen concentrations of less than 3.0 mg/L. The system consisted of (1) high rate BOD removal activated sludge process with clarifier, (2) nitrification activated sludge process with clarifier, and (3) denitrification activated sludge process with clarifier. Methanol was added to the third step to provide an electron donor for denitrification. In the late 1990s and early 2000s this system was tested and installed at full scale, but only in a two-sludge system, with the nitrification and post denitrification processes with methanol addition combined in the second sludge system (Bailey et al., 1998 and Sadick et al., 2000).

Following these innovative applications for nitrogen removal in the 1960s and 1970s, a greater understanding of the nitrification and denitrification process kinetics was gained and applied in these and other design modifications in the 1980s and beyond. These processes are discussed in the following sections.

Overview of Types of Biological Nitrogen Removal Processes

All of the biological nitrogen removal processes include an aerobic zone in which biological nitrification coverts NH_4-N to NO_2-N and NO_3-N. Anoxic volume or time must also be included to provide biological denitrification to complete the objective of total nitrogen removal by both NH_4-N oxidation and NO_3-N and NO_2-N reduction to nitrogen gas. As discussed in Sec. 7–10 in Chap. 7, nitrate/nitrite reduction requires an electron donor, which can be supplied in the form of influent wastewater BOD, by endogenous respiration, or an external carbon source. The type of biological nitrogen removal process is classified in this section according to the location of the anoxic reactor relative to the nitrification step as follows:

1. Preanoxic denitrification processes
2. Postanoxic denitrification processes
3. Low DO and cyclic nitrification/denitrification processes

In addition to differences in reactor configuration for each of these process types, the primary substrate source and nitrate reduction kinetics vary. It should be noted that design and operating conditions of SBR systems can be selected to provide any of these types of denitrification processes or in various combinations including: (1) a preanoxic and postanoxic process, (2) a preanoxic and low DO process, or (3) a preanoxic and cyclic nitrification/denitrification process. The simultaneous removal of phosphorus is introduced here, but the subject of enhanced biological phosphorus removal is discussed in detail in Sec. 8–8.

Preanoxic Denitrification Processes. In the preanoxic configuration [see Fig. 8–26(a)], nitrate produced in the aerobic zone is recycled to the mixed preanoxic zone. Denitrifying bacteria in the preanoxic zone consume influent BOD to reduce the nitrate to nitrogen gas. The rate of denitrification in the preanoxic zone is affected by the rbCOD concentration in the influent wastewater, the MLSS concentration, and temperature. Typical views of the anoxic section of different activated sludge processes are shown on Fig. 8–27.

There are many advantages for using a preanoxic zone upstream of a nitrification zone. These include (1) the relative ease of retrofit to existing plants, (2) the benefits of the anoxic selector operation for control of filamentous sludge, (3) the production of alkalinity before the nitrification step, (4) the energy savings due to the use of nitrate for oxidation of the influent BOD, and (5) the ability to convert an existing biological treatment system to nitrogen removal with relatively short to moderate basin detention times. Because of these advantages a preanoxic zone should be used with activated sludge systems that are designed for nitrification, even if nitrogen removal is not required. In addition to the improved SVI and energy savings, the use of a nitrification/denitrification process has been shown to be a more economical choice as compared to nitrification alone (Rosso and Stenstrom, 2005a). Although the longer SRT values associated with nitrification and denitrification processes require increased tank volume as compared to systems designed only for BOD removal, there is only a small increase in the overall WWTP energy requirements results based on improved alpha values and aeration efficiencies and a decrease in the amount of waste sludge to handle (Leu et al., 2012).

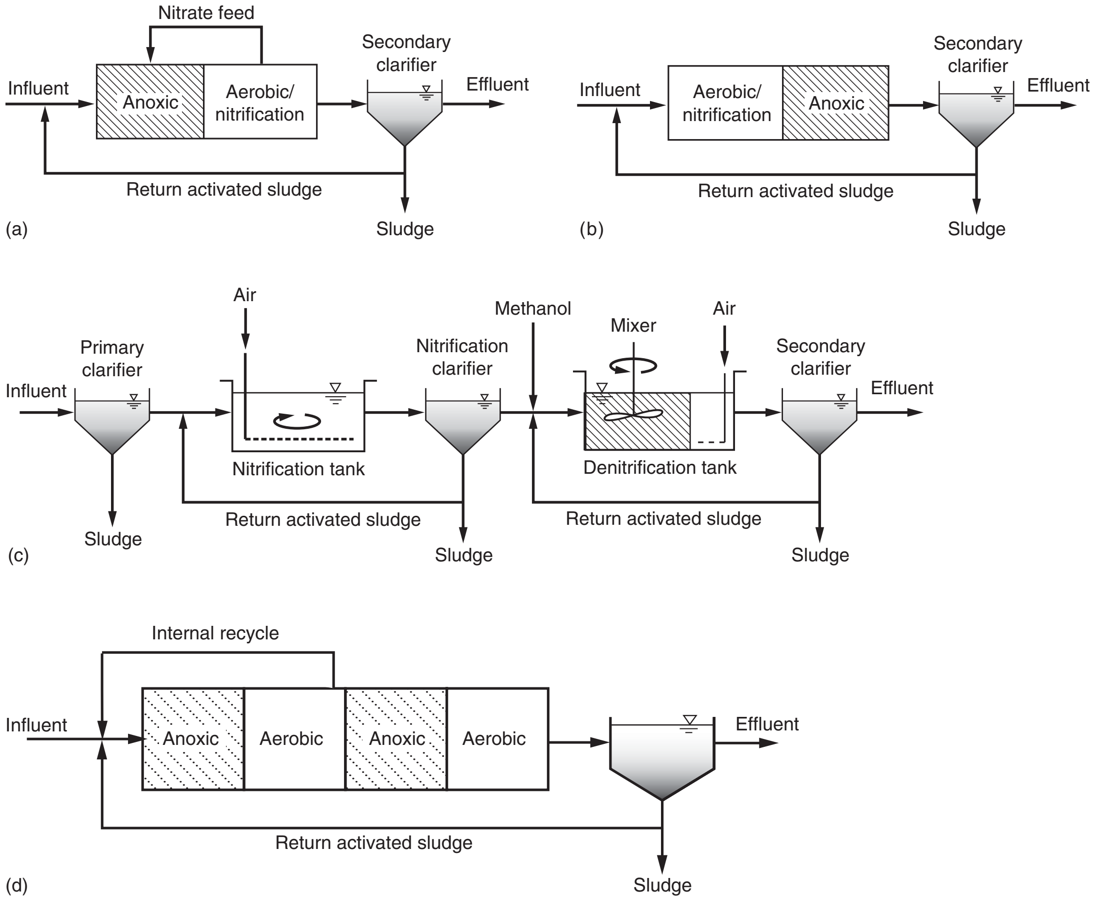


Figure 8–26

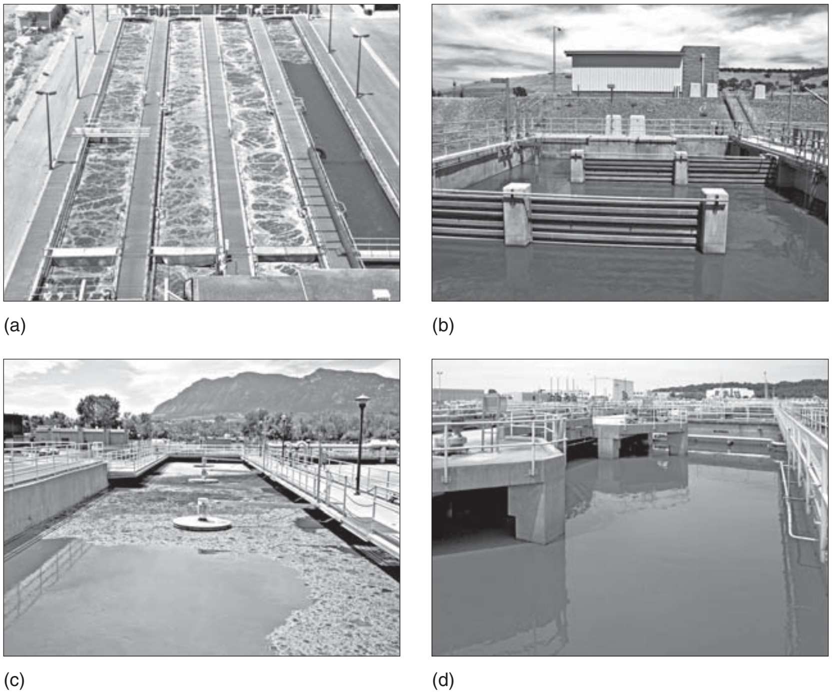
Schematic diagram of four basic biological nitrogen removal process configurations: (a) preanoxic, (b) postanoxic, (c) two-sludge nitrification-denitrification, and (d) Bardenpho process.

Postanoxic Denitrification Processes. Postanoxic designs [see Fig. 8–26(b), (c), and (d)] are commonly used as a polishing denitrification step to remove nitrate to a minimal effluent concentration. For example, in the Bardenpho process [see Fig. 8–26(d)] more than 75 percent of the nitrate is typically removed in the preanoxic zone with the reminder removed in the postanoxic zone. The postanoxic zone may be operated with or without an exogenous carbon source. Without an exogenous source, postanoxic processes depend on the endogenous respiration of the activated sludge to provide electron donor for nitrate consumption in lieu of oxygen. The denitrification rate is much slower, by a factor of 3 to 6, compared to preanoxic applications that use influent wastewater BOD for the electron donor.

Postanoxic denitrification have also been done in a two sludge system design as shown on Fig. 8–26(c) in which BOD and nitrification occurs in the first sludge system. Because of the cost of chemical addition for all of the nitrate removal and the need for a second clarifier, single-sludge systems with preanoxic zones are preferred over this suspended growth alternative for nitrogen removal. Where a two step process is used it is now more common to use fixed film denitrification processes with an exogenous carbon source for nitrate removal. Fixed film processes are considered in Chap. 9.

Figure 8–27

Views of anoxic reactors: (a) reactor used for nitrogen removal (the righthand channel without aeration is the preanoxic section, (b) baffled anoxic section of small plug-flow activated sludge reactor, (c) baffled (submerged) anoxic section with surface mixers of large plug-flow activated sludge reactor (coordinates 34.1825 N, 118.4793 W, view at altitude 250 m), and (d) postanoxic denitrification in nitrification reactors with methanol addition in deep tanks with invent mixers.

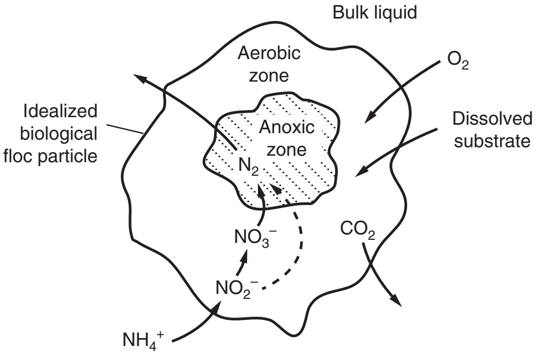


Low DO and Cyclic Nitrification/Denitrification Processes. Denitrification can be accomplished in single-reactor activated sludge systems without permanent and distinct nitrification and anoxic zones by two methods: (1) simultaneous nitrification-denitrification (SNdN) and (2) cyclic nitrification-denitrification (Cyclic NdN). These single reactor nitrification/denitrification processes rely on operational methods for aeration and mixing to accomplish nitrogen removal, instead of having distinct aerobic and anoxic zones as in the preanoxic and postanoxic processes.

Simultaneous NdN. SNdN is accomplished by operation at a low DO concentration so that the activated sludge floc contains both aerobic and anoxic zones, as illustrated in a simplified view of a biological floc on Fig. 8–28. Dissolved oxygen and dissolved substrates outside of the floc diffuse into the aerobic zone, and depending on the DO concentration and concentrations of ammonia and bCOD, oxygen may be depleted at significant rates within the floc so that the DO cannot penetrate the entire floc depth and anoxic zones occur

Figure 8–28

Diagram of an activated sludge floc showing aerobic and anoxic zones.



in the inner floc volume. Nitrite and nitrate produced by nitrification in the aerobic zone can diffuse into the inner anoxic zone along with substrate so that denitrification occurs within the floc depth. The major pathway for nitrogen removal may be NH_4-N to NO_2-N to N_2 instead of via NO_3-N to N_2 as indicated by observations of ammonia oxidizing microorganisms in limited DO aerated systems with archaea ammonia-oxidizing archaea (AOA) dominating in a low DO MBR system (Giraldo et al., 2011a and 2011b) and ammonia-oxidizing bacteria (AOB) in a lab reactor treating domestic wastewater (Peng et al., 2012).

Nitrification and denitrification rates are at less than optimal levels in simultaneous nitrification/denitrification (SNdN) processes. Only a portion of the biomass is used for each of these reactions. In addition, the nitrification rate is lower due to the low DO concentration, and the denitrification rate is lower due to substrate consumption in the aerobic portion of the floc. The impact of anoxic zones within the biological floc on decreasing nitrification rates was observed by Stenstrom and Song (1991) who reported that nitrification rates were related not only to bulk liquid DO concentration, but also to the amount of BOD present. At higher soluble BOD concentrations, higher oxygen uptake rates occurred, and lower nitrification rates were observed for the same bulk liquid DO concentration, most likely due to the fact that the aerobic zone of the activated sludge floc was decreased.

Cyclic NdN Processes. Cyclic NdN processes involve aeration and unaerated mixing process steps, either spatially or with time. An aeration step with elevated DO for nitrification is followed by an unaerated, mixed time for denitrification. Process control methods must be provided and implemented for successful nitrogen removal in cyclic NdN processes. Cyclic NdN processes have been demonstrated in: (1) oxidation ditches, (2) series compartments with alternating aerobic/anoxic operation, and (3) phased operation with tanks in series.

Reported plant effluent NO_3-N performance data for cyclic NdN processes range from 3.0 to 4.8 mg/L, with effluent total nitrogen (TN) concentrations below 8.0 mg/L (U.S. EPA, 1993). The relatively long τ values used provide sufficient dilution to minimize the effluent NH_4-N concentrations during the OFF period. A sufficiently long SRT is also needed to provide enough nitrification capacity to allow the aeration system to be operated intermittently.

Oxidation Ditch Processes. To accomplish cyclic NdN in an oxidation ditch, aerobic and anoxic zones are created along the ditch channel length with the mixed liquor recirculating. An elevated DO concentration occurs in the aeration zone of the oxidation ditch and in a portion of the volume downstream from the aeration zone so that nitrification occurs. DO is continually depleted as the flow moves down the channel until it is at or near zero and anoxic conditions then prevail for denitrification. The channel flow circulates around the ditch with circulation times typically ranging from 5 to 15 minutes with the activated sludge mixed liquor cycling through aerobic and anoxic conditions. The oxidation ditch aerator submergence or motor speed can be varied to control the size of the aerobic and anoxic zones for nitrification and denitrification.

Oxidation ditches can also employ intermittent aeration by having a directional mixer installed to maintain channel flow when the aerator is turned off. During the aeration-off period, the aeration tank operates essentially as an anoxic reactor as nitrate is used in lieu of DO for BOD removal. The operation of an oxidation ditch using intermittent aeration is shown on Fig. 8–29. Intermittent aeration systems typically are operated with SRT values in the range of 18 to 40 d and hydraulic detention times in excess of 16 h. During the anoxic reaction period [see Fig. 8–29(b)], aeration is stopped, a submerged mixer is turned on, and nitrate is used as the electron acceptor. The reactor (ditch) is operated as a complete-mix activated sludge anoxic process. During the anoxic period, DO and nitrate

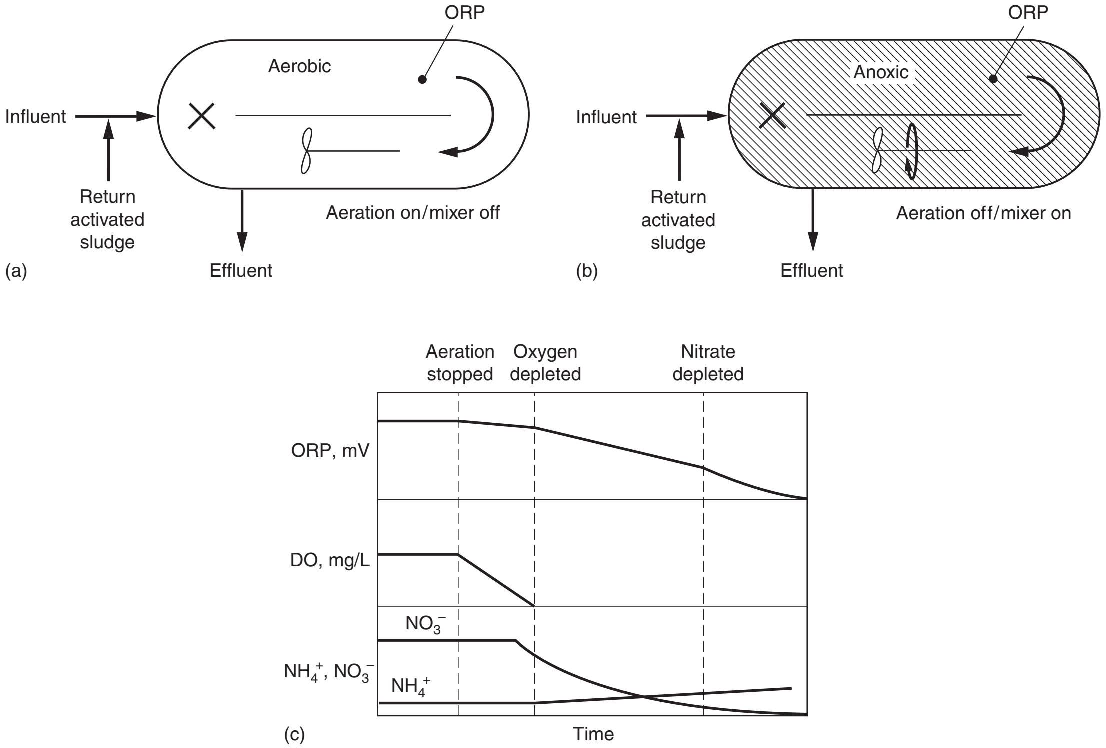


Figure 8–29

Operation of a oxidation ditch process using intermittent aeration: (a) aerobic conditions, (b) anoxic conditions, and (c) variations in ORP, DO, ammonia, and nitrate.

are depleted and the ammonia concentration increases [see Fig. 8–29(c)]. The time for the anoxic and aerobic periods is important in determining the system's treatment performance. A control method for the intermittent aeration using an oxidation-reduction probe (ORP) measurement is presented under the Process Control and Performance topic at the end of this nitrogen removal section.

Alternating Aerobic/Anoxic Processes. An alternating aerobic/anoxic operation involves the use of two compartments in series, with the influent going to the first compartment [see Fig. 8–30(a)]. At select times the first compartment is aerated and the second mixed and then it is switched with the first compartment mixed and the second aerated. Basically, the operation cycles between preanoxic and postanoxic denitrification.

Phased Operation Processes. For a phased operation, two compartments or tanks are also used but the influent feed is always fed to a mixed anoxic zone [see Fig. 8–30(b)]. In one phase tank A is fed influent and mixed and its effluent flow is to tank B which is aerated. The mixed liquor flow from tank B goes to the secondary clarifier. At selected times the flow and aeration/mixing is reversed between tank A and B. Tank B receives the influent flow and is a mixed anoxic tank due to the nitrate produced in its previous aerobic phase. In this phase the mixed liquor flow from tank B goes to tank A which is now aerated, and the effluent flow from tank A goes to the secondary clarifier. In the phased operation denitrification is done in a preanoxic denitrification process. An example of nitrogen

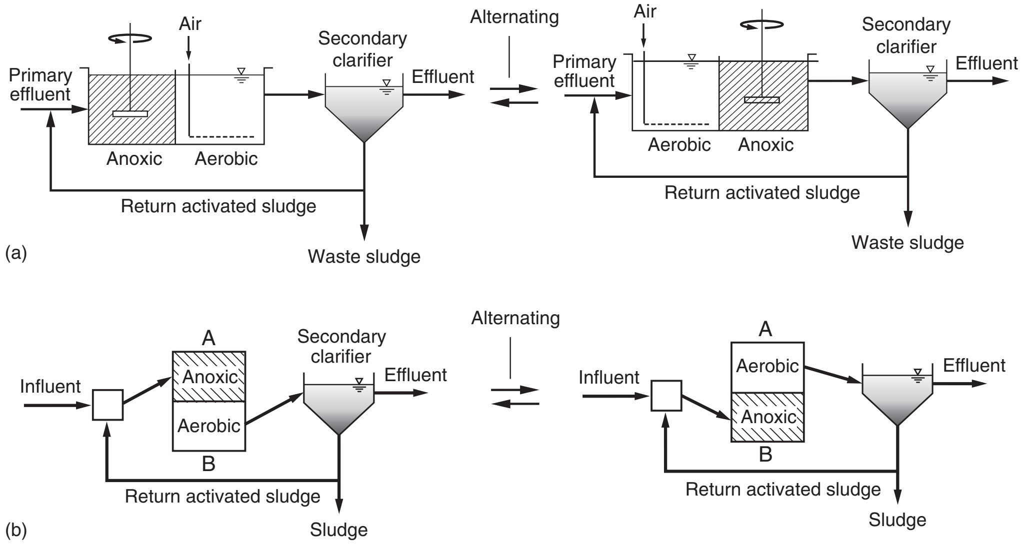


Figure 8–30

Schematic of Cyclic NdN Processes: (a) alternating or cycling process and (b) phase operation.

removal by phased operation is the BioDenitro process, which uses two oxidation ditches for the alternating anoxic and aerobic tanks (Stensel and Coleman, 2000).

Simultaneous Removal of Phosphorus. Enhanced biological phosphorus removal occurs in biological nitrogen removal processes when (1) there is excess rbCOD during the anoxic contact so that substrate is available for uptake by PAOs and (2) there is sufficient downstream aerobic time for stored carbon oxidation by the PAOs.

General Process Design Considerations

Design issues and design examples for commonly used biological nitrogen removal processes are presented below. The fundamental design concepts that are exemplified can be of use in evaluating other types of suspended growth biological nitrogen removal processes.

The main goal of the denitrification design is to determine the volume or time required for the particular type of anoxic zone. Each of the denitrification processes has unique design considerations and denitrification rates specific to that process type that are addressed in the process design for each denitrification process. Common issues related to all three types of denitrification processes discussed in this section are (1) system SRT, (2) specific denitrification rate (SDNR), (3) MLSS concentration, (4) temperature, (5) influent wastewater characteristics, (6) alkalinity, and (7) anoxic tank mixing requirements.

System SRT. The starting point for all the nitrogen removal designs is providing an adequate aerobic SRT for nitrification. The required aerobic SRT for nitrification, as discussed in Sec. 8–6, will depend on a number of factors, including the effluent NH_4-N concentration goal, flow and loading variability, DO concentration, and temperature. The required aerobic SRT is used to compute the aerobic volume and the total system SRT is approximately equal to the total volume (aerobic plus anoxic volume) to aerobic volume ratio times the aerobic SRT. Longer SRTs result in lower specific denitrification rates for postanoxic denitrification and cyclic NdN processes.

Specific Denitrification Rate (SDNR). Values of the SDNR have been used to characterize denitrification rates in different anoxic systems as well as to evaluate the effect of different external carbon sources. The SDNR in an anoxic tank is related to the amount of nitrate removed per unit time, normalized to the MLVSS concentration:

$$\text{SDNR} = \frac{\text{NO}_r}{(\text{MLVSS})(V_{\text{nox}})} \tag{8–51}$$

where SDNR = specific denitrification rate, g NO_3-N/g MLVSS·d
NO_r = nitrate-nitrogen removal rate, g/d
V_{nox} = anoxic tank volume, m^3
MLVSS = mixed liquor volatile suspended solids concentration, mg/L

The desktop design approach is based on using a specific denitrification rate (SDNR) to determine the NO_3-N removal rate in an anoxic tank volume.

$$\text{NO}_r = (V_{\text{nox}})(\text{SDNR})(\text{MLVSS}) \tag{8–52}$$

Values of SDNR observed for preanoxic tanks in full-scale installations have ranged from 0.04 to 0.42 g NO_3-N/g MLVSS·d (Burdick et al., 1982; Henze, 1991; Bradstreet and Johnson, 1994; Reardon et al., 1996; Hong et al., 1997; and Murakami and Babcock, 1998). For postanoxic denitrification without an exogenous carbon source, observed SDNRs have ranged from 0.01 to 0.03 g NO_3-N/g MLVSS·d.

MLSS Concentration. The required anoxic zone volume is less for higher MLSS concentrations. As will be shown in Sec. 8–10, the MLSS concentration that can be used is related to the activated sludge settling and thickening properties. Because the preanoxic denitrification process also serves as a selector for good settling mixed liquor, operation at higher MLSS concentrations is possible.

Temperature. As shown for nitrification, denitrification rates will also be lower for operation at lower temperatures. The lowest sustained temperature is normally selected as the worst case for the anoxic zone design. For some retrofit applications to biological nitrogen removal, it is possible that there may not be enough tankage available to provide the volume needed for both nitrification and denitrification. In that case the volume needed for nitrification is provided, and external carbon can be added to the denitrification zone to provide a faster nitrate reduction rate in the limited anoxic volume at colder temperatures. External carbon addition is addressed at the end of this section.

Influent Wastewater Characteristics. The influent wastewater characteristics are important in affecting biological denitrification rates in preanoxic, SNdN, and Cyclic NdN processes. Nitrate reduction is dependent on having sufficient electron donor so there must be a sufficient amount of influent BOD relative to the amount of nitrogen to be removed. As a rule of thumb, an influent BOD to TKN ratio of 4/1 is necessary to provide a sufficient amount of electron donor (Randall et al., 1992).

Alkalinity. Denitrifying bacteria have a wider range of pH tolerance than nitrifying bacteria so that influent alkalinity is not as critical for the denitrification design. However, it is useful to determine the amount of nitrate removal and its associated alkalinity production to determine if it can help to maintain a satisfactory alkalinity concentration and pH in the aerobic nitrification zone.

Anoxic Tank Mixing. For preanoxic, postanoxic, and some cyclic NdN systems the anoxic zones are designed as single-stage or a series of complete-mix tanks. Typical power requirements for mechanical mixing in the anoxic zone depends on the mixer design and tank geometry and can be as low as 3 kW/10^3 m^3 for slow speed mixers to 8 kW/10^3 m^3 for higher speed mixers (0.1 to 0.3 hp/10^3 ft^3).

Preanoxic Denitrification Processes

Design considerations for three biological nitrogen removal processes are discussed in this section. Following a presentation of some general background material that is specific to each of the three nitrogen removal processes, design examples for each are presented. The three design examples are: (1) preanoxic denitrification process design for the MLE process, (2) step feed nitrification and preanoxic denitrification process design, and (3) preanoxic denitrification process design for an SBR process. Similar discussion and analyses are provided for postanoxic denitrification and low DO and cyclic nitrification/denitrification processes in subsequent parts.

Preanoxic Denitrification Process Design for MLE Process. The design approach for the MLE process is summarized in Table 8–21 and in general proceeds with the following major steps: (1) determine the aerobic nitrification design, (2) determine the internal recycle rates and the amount of nitrogen to remove in the preanoxic zone, and (3) determine the anoxic tank volume and SDNR. The first step was described in Sec. 8–6. The same steps can be followed for an anoxic/aerobic design in an MBR process with necessary modifications with regard to the MLSS concentration, aeration tank configuration, and aeration design. The anoxic/aerobic design steps for an MBR system are summarized in Table 8–23 on page 807. The methods to obtain the SDNR and mass balance needed to obtain the internal recycle rate are discussed here. An example of a preanoxic denitrification process design for an MLE process with secondary clarifier and MBR system is presented following the presentation of the background material given below.

Relationship of SDNR to BOD F/M Ratio. Based on observed denitrification rates in pilot-plant and full-scale plants, empirical relationships have been developed that relate SDNR to the BOD or COD F/M ratio for the preanoxic tank (Burdick et al., 1982; U.S. EPA, 2010):

$$\mathrm{SDNR}_{20} = (\mathrm{F/M})(F_b/0.30)(\mathrm{F/M}) + 0.029 \tag{8–53}$$

$$F_b = \frac{(Y_\mathrm{H})/1 + b_\mathrm{H}(\mathrm{SRT})}{[(Y_\mathrm{H}/1) + b_\mathrm{H}(\mathrm{SRT}) + Y_\mathrm{I}]} \tag{8–54}$$

$$\frac{\mathrm{F}}{\mathrm{M}} = \frac{QS_o}{XV_\mathrm{nox}} \tag{8–55}$$

where F/M = g BOD applied/g MLVSS·d in the anoxic tank
SDNR_{20} = specific denitrification rate at 20°C, gNO_3-N/g MLVSS·d
Q = influent flowrate, m^3/d
S_o = influent BOD concentration, mg/L
X = MLVSS concentration in anoxic tank, mg/L
V_nox = anoxic tank volume, m^3
F_b = active biomass fraction of MLVSS
Y_H = heterotrophic biomass synthesis yield, 0.67 g VSS/g BOD removed
b_H = endogenous decay rate, g VSS/g VSS·d
Y_I = influent wastewater inert VSS fraction, g nbVSS/g BOD

Table 8–21

Computation approach for anoxic/aerobic process design with secondary clarification

Item	Description
1.	Establish wastewater flowrates and characteristics, including the rbCOD concentration, and effluent requirements.
2.	Follow the procedure outlined in Table 8–15 for the aerobic zone for a nitrification process with the exception that steps 13–18 are done after the anoxic reactor design.
3.	Determine the biomass concentration in the mixed liquor from the nitrification design.
4.	Determine the internal recycle (IR) ratio, using the NO_x value determined in step 9 of the nitrification design and desired effluent NO_3-N concentration.
5.	Calculate the amount of nitrate fed to the anoxic tank. The design is based on the assumption that essentially all of the nitrate fed to the anoxic zone will be reduced. A low nitrate concentration of 0.1 to 0.3 mg/L may remain, depending on the design, because the nitrate limits the denitrification reaction rate at only very low concentrations.
6.	Select the anoxic volume and configuration; single stage or multiple-staged reactors.
7.	Calculate the F/M_b based on the biomass concentration determined for the mixed liquor in the nitrification design.
8.	Use Eq. (8–57) or (8–58) and appropriate corrections for temperature and IR ratio to obtain the $SDNR_b$, also based on the biomass concentration, for the anoxic basin.
9.	Using the $SDNR_b$, biomass concentration, and anoxic volume compute the amount of nitrate removed in the anoxic basin. Compare the amount of removal required to remove all of the nitrate in the recycle streams fed to the anoxic zone.
10.	For downstream anoxic basins calculate the F/M_b after reducing the influent BOD, bCOD, and rbCOD based on the amount of nitrate removed in the upstream anoxic basin.
11.	Repeat the anoxic zone design steps as necessary to obtain a satisfactory design.
12.	Calculate the oxygen demand.
13.	Determine if alkalinity addition is needed.
14.	Design the secondary clarifier.
15.	Design the aeration oxygen transfer system.
16.	Summarize the final effluent quality.
17.	Prepare a design summary table.

The endogenous decay rate is corrected for temperature using Eq. (1–44), a b_{20} value of 0.12 g VSS/g VSS·d, and temperature coefficient θ value of 1.029. The above relationships are based on empirical observations and do not account for the effect of different rbCOD fractions in the influent wastewater bCOD. Higher SDNR values will occur in the preanoxic zone for wastewaters with a higher rbCOD/bCOD ratios.

The effect of the rbCOD/bCOD ratio, expressed as a percentage, on the SDNR is presented on Fig. 8–31. The values are generally applicable and can be used for wastewaters with different fractions of rbCOD (rbCOD/bCOD) and inert nonbiodegradable volatile solids. On Fig. 8–31, the F/M_b ratio (Eq. 8–56) and $SDNR_b$ (Eq. 8–57 or Eq. 8–58) values are based only on the active heterotrophic biomass concentration in the mixed liquor, so that the rates can be applicable to many situations regardless of the amount of nondegradable solids in the mixed liquor and the SRT. The F/M_b ratio is defined

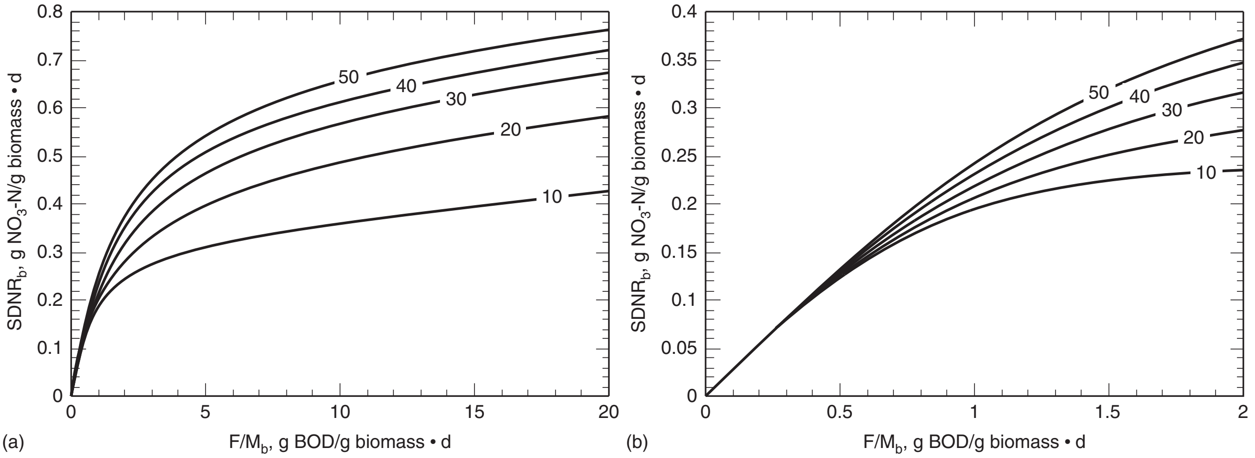


Figure 8–31

Specific denitrification rates (SDNR$_b$) based on biomass concentration at 20°C versus food to biomass (F/M$_b$) ratio for various percentages of rbCOD relative to the influent wastewater bCOD (a) for F/M$_b$ values up to 20 and (b) expanded scale for F/M$_b$ values up to 2.

as a function of the BOD loading to the anoxic volume and active heterotrophic biomass concentration, as follows:

$$\text{F/M}_\text{b} = \frac{QS_o}{(V_\text{nox})X_b} \tag{8–56}$$

where F/M$_\text{b}$ = BOD F/M ratio based on active biomass concentration, g BOD/g biomass·d
X_b = anoxic zone biomass concentration, mg/L

The curves shown on Fig. 8–31 are based on the results of model simulations of biomass, NO$_3$-N, rbCOD, and pbCOD mass balances in the anoxic tank. For lower anoxic τ values, the F/M$_\text{b}$ is higher, which resulted in greater rbCOD concentrations in the anoxic zone and thus a higher biological reaction rate and SDNR. Internal recycle rates from the aerobic zone and temperature effects were accounted for. The biokinetic coefficients used in the model simulation are model parameter default values from the ASM1 model (Grady et al., 1986) along with the rbCOD kinetics under anoxic conditions observed by Stensel and Horne (2000) from testing at different municipal wastewater-treatment facilities. Equations (8–57) and (8–58), derived from the data presented on Fig. 8–31, can be used for calculating SDNR$_\text{b}$ as a function of F/M$_\text{b}$. Values for b_0 and b_1 are given in Table 8–22.

$$\text{For F/M} > 0.50,\ \text{SDNR}_\text{b} = b_0 + b_1\,[\ln(\text{F/M}_\text{b})] \tag{8–57}$$

$$\text{For F/M} \le 0.50,\ \text{SDNR}_\text{b} = 0.24(\text{F/M}_\text{b}) \tag{8–58}$$

Table 8–22

Coefficient values for Eq. (8–57) relating SDNR$_\text{b}$ to percent rbCOD

Percent rbCOD	SDNR equation coefficients	
	b_0	b_1
10	0.186	0.078
20	0.213	0.118
30	0.235	0.141
40	0.242	0.152
50	0.270	0.162

Table 8–23

Computation approach for anoxic/aerobic process design with membrane liquid-solids separation

Item	Description
1.	Establish wastewater flowrates and characteristics, including the rbCOD concentration, and effluent requirements.
2.	Follow the procedure outlined in Table 8–15 and complete steps 1 through 10.
3.	Select the membrane design flux value and determine the installed membrane total surface area.
4.	Determine the membrane separation tank volume based on the membrane surface area needed and membrane suppliers required aerobic membrane tank volume to membrane surface area ratio.
5.	Determine the anoxic tank and preaeration tank MLSS concentrations. Use the selected membrane tank MLSS concentration and return activated sludge recycle ratio.
6.	Determine the preaeration tank volume and active biomass concentration. Determine the biomass concentration in the anoxic zone by calculating the fraction of biomass in the MLSS concentration in the nitrification design and multiplying that by the MLSS determined in Step 5 above.
7.	Using the return activated sludge ratio selected and the NO_x value determined in step 9 of the nitrification design, determine the effluent NO_3-N concentration.
8.	Calculate the amount of nitrate fed to the anoxic tank. The design is based on the assumption that essentially all of the nitrate fed to the anoxic zone will be reduced.
9.	Select the anoxic volume and configuration; single stage or multiple-staged reactors.
10.	Calculate the F/M_b based on the biomass concentration in the anoxic zone.
11.	Use Eq. (8–57) or (8–58) and appropriate corrections for temperature and IR ratio to obtain the $SDNR_b$ for the anoxic basin.
12.	Using the $SDNR_b$, biomass concentration, and anoxic volume compute the amount of nitrate removed in the anoxic basin. Compare the amount of removal required to remove all of the nitrate in the recycle streams fed to the anoxic zone.
13.	For downstream anoxic basins calculate the F/M_b after reducing the influent BOD, bCOD, and rbCOD based on the amount of nitrate removed in the upstream anoxic basin.
14.	Repeat the anoxic zone design steps as necessary to obtain a satisfactory design.
15.	Calculate the oxygen demand accounting for oxygen equivalents supplied by NO_3-N.
16.	Determine if alkalinity addition is needed.
17.	Determine the anoxic zone mixing power.
18.	Design the aeration oxygen transfer system for the aerobic tanks. First determine what fraction of the total oxygen demand is provided in the preaeration basin and membrane separation basin.
19.	Prepare a design summary table.

The design procedure using $SDNR_b$ values from Fig. 8–31 requires calculating the active biomass VSS concentration in the mixed liquor using the procedure presented previously in Example 8–2. By using the calculated biomass concentration, the effects of SRT are accounted for in the design procedure. The BOD F/M_b ratio on Fig. 8–31 is also based on the active biomass VSS concentration. The procedure is applicable to a single-stage anoxic reactor. The $SDNR_b$ for subsequent downstream staged anoxic reactors can be estimated by reducing the value for the influent BOD (S_o) by the amount removed using nitrate as the electron acceptor by 4 g BOD used per g of NO_3-N reduced (see Sec. 7–10).

The computed value should be subtracted from the influent rbCOD concentration, (assuming 1.6 g bCOD/g BOD) as that will be consumed the fastest. Thus, the rbCOD:bCOD ratio of the influent BOD to subsequent anoxic zones decreases.

At high F/M$_b$ ratios shown on Fig. 8–31, the SDNR$_b$ reaches a maximum saturated rate, as the rbCOD concentration will be very high in the anoxic reactor. The maximum saturation rate will only be experienced for high F/M$_b$ anoxic selector designs where the first anoxic zones have detention times of less than 10 to 20 min. For long anoxic detention times (on the order of 3 to 6 h), SDNR$_b$ values will be found at the lower F/M$_b$ ratios.

Temperature and Internal Recycle Corrections for SNDR. The design procedure requires correcting the SDNR values from Eq. (8–57) or (8–58) for temperature and internal recycle ratios. The temperature correction is made by using Eq. (2–25) with a θ value of 1.026. The SDNR in the preanoxic tank is affected by the internal recycle rate, often defined as an internal recycle (IR) ratio. The IR ratio is the recycle flowrate divided by the influent wastewater flowrate. At higher IR ratios, the influent rbCOD is diluted more in the anoxic reactor by mixed liquor from the aerobic reactor, resulting in a lower denitrification rate. The corrections to the SDNR for designs with internal recycle corrections greater than 1.0 are shown below. The SDNR value can be interpolated from the values listed. If the F/M is less than or equal to 1.0, no correction is required.

$$\text{IR} = 2 \qquad \text{SDNR}_{\text{adj}} = \text{SDNR}_{\text{IR1}} - 0.0166 \ln(\text{F/M}_b) - 0.078 \tag{8–59}$$

$$\text{IR} = 3\text{–}4 \qquad \text{SDNR}_{\text{adj}} = \text{SDNR}_{\text{IR1}} - 0.029 \ln(\text{F/M}_b) - 0.012 \tag{8–60}$$

where SDNR_{adj} = SDNR adjusted for the effect of internal recycle
SDNR_{IR1} = SDNR value at internal recycle ratio = 1
F/M_b = BOD F/M ratio based on anoxic zone volume and active biomass concentration, g/g·d

Nitrogen Mass Balance and Internal Recycle. A mass balance on nitrogen must be done to determine (1) how much nitrate is produced in the aeration zone, and (2) what internal recycle ratio must be used to meet the desired effluent nitrate concentration. The mass balance accounts for the nitrate produced in the aerobic zone. The nitrate production rate in the aerobic zone is based on the influent flowrate and nitrogen concentration, the amount consumed for cell synthesis, and the effluent NH_4-N and soluble organic nitrogen concentrations. As a conservative design approach, all of the influent TKN is assumed to be biodegradable and the effluent soluble organic nitrogen concentration is ignored. The nitrate produced is contained in the total flow leaving the aerobic zone, which includes internal recycle, RAS, and effluent flows. The mass balance is expressed as follows and assumes no nitrate reduction in the secondary clarifier.

$$\begin{array}{c} \text{kg/d of nitrate} \\ \text{produced in} \\ \text{aerobic zone} \end{array} = \begin{array}{c} \text{nitrate} \\ \text{in effluent} \end{array} + \begin{array}{c} \text{nitrate in} \\ \text{internal} \\ \text{recycle} \end{array} + \begin{array}{c} \text{nitrate in} \\ \text{return activated} \\ \text{sludge (RAS)} \end{array}$$

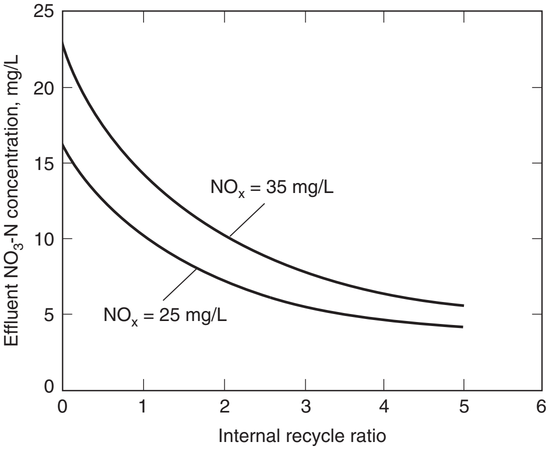
$$Q(\text{NO}_x) = N_e[Q + (\text{IR})Q + (R)Q] \tag{8–61}$$

$$\text{IR} = \frac{\text{NO}_x}{N_e} - 1.0 - R \tag{8–62}$$

where IR = internal recycle ratio (internal recycle flowrate/influent flowrate)
R = RAS recycle ratio (RAS flowrate/influent flowrate)

Figure 8–32

Effect of internal recirculation rate on effluent nitrate concentration (RAS ratio = 0.50) for an anoxic/aerobic process.



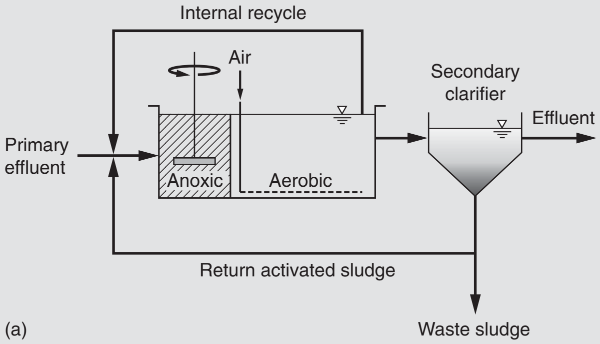
NO_x = nitrate produced in aeration zone as a concentration relative to influent flow, mg NO_3-N/L

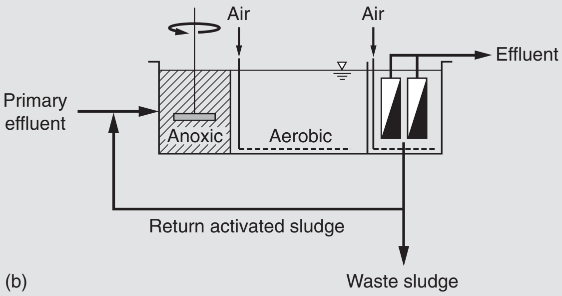
N_e = effluent NO_3-N concentration, mg/L

The effect of the IR ratio on the effluent NO_3-N concentration for a given amount of nitrate produced (NO_x) and for a RAS recycle ratio of 0.50 is illustrated on Fig. 8–32. A greater IR ratio is needed to produce the same effluent NO_3-N concentration when more NO_x is produced in the aerobic zone. To meet a standard of 10 mg TN/L or less, a design effluent NO_3-N concentration of 5 to 7 mg/L should be used. An internal recycle ratio in the range of 3 to 4 is typical, but ratios in the range of 2 to 3 are also applied for wastewaters with a lower influent wastewater TKN concentration. Recycle ratios above 4 are not used for activated sludge/secondary clarifier systems, as the incremental removal of NO_3-N is low, and more DO is recycled from the aeration zone into the anoxic zone. The higher recirculation rate also dilutes the influent rbCOD in the anoxic zone to decrease the SDNR compared to that at lower recirculation ratios. As noted previously, the amount of DO fed to the anoxic zone due to the internal recycle flow from the aerobic zone must be minimized for effective denitrification. In some designs, sections of the aerobic zone are baffled with DO control so that the DO concentration in the recycle can be controlled and minimized.

MBR Internal Recycle Ratio. The internal recycle ratio for MBR systems is typically set at 6.0 to control the MLSS concentration in the membrane separation zone. In applying Eq. (8–62) for an MBR design, the value for R is equal to 6.0 and IR is equal to zero. The effluent NO_3-N concentration (N_e) is then calculated. A concern with the higher recycle flowrate is the potential to feed excessive DO from the membrane separation tank to the preanoxic tank. High DO concentrations in recycle flows to the preanoxic tank can result in the consumption of rbCOD in the influent wastewater, thus reducing the amount of rbCOD available for NO_x reduction. A tank with a relatively short detention time could be used for holding the return activated sludge flow before it enters the preanoxic tank to allow oxygen that may be present to be consumed (e.g., deoxygenation) by the oxygen demand from endogenous decay.

EXAMPLE 8–7 **Preanoxic Denitrification Process Design for MLE Process** Design a preanoxic basin for (a) the CMAS nitrification-secondary clarifier system described in Example 8–3 to produce an effluent NH_4-N and NO_3-N concentration of 0.50 and 6.0 g/m^3, respectively and (b) design an MBR system using hollow fiber membranes to meet the NH_4-N concentration of 0.50 g/m^3 with a return activated sludge ratio of 6.0. Process schematic flow diagrams are shown below for the two systems. The design condition is based on the following information from Example 8–3, and the design steps outlined in Tables 8–22 and 8–23 are followed.





Wastewater characteristics:

Constituent	Concentration, g/m^3
BOD	140
bCOD	224
rbCOD	80
NO_x	28.9
TP	6
Alkalinity	140 as $CaCO_3$

Part A—Design for CMAS Process

1. Design conditions:

Constituent	Unit	Value
Influent flowrate	m^3/d	22,700
Temperature	°C	12.0
MLSS	g/m^3	3000
MLVSS	g/m^3	2370
Aerobic SRT	d	21.0
Aeration basin volume	m^3	13,410
Aerobic	h	14.2
Anoxic mixing energy	kW/10^3 m^3	5
RAS ratio	Unitless	0.6
R_o	kg/h	275.9

Note: g/m^3 = mg/L.

2. Assumptions:
   a. Nitrate concentration in RAS = 6 g/m^3.
   b. Use the same coefficients as the nitrification process design.
   c. Mixing energy for anoxic reactor = 5 $kW/10^3\ m^3$.

Solution

1. Determine the active biomass concentration using Eq. (7–42) and substituting V/Q for τ.

$$X_b = \left[\frac{Q(\text{SRT})}{V}\right]\left[\frac{Y_{\text{H}}(S_o - S)}{1 + b_{\text{H}}(\text{SRT})}\right]$$

where $S_o - S \approx S_o$

$$X_b = \frac{(22{,}700\ \text{m}^3/\text{d})(21.0\ \text{d})(0.45\ \text{g VSS/g COD})(224\ \text{g bCOD/m}^3)}{[1 + (0.088\ \text{g/g·d})(21.0\ \text{d})](13{,}320\ \text{m}^3)}$$
$$= 1267\ \text{g/m}^3$$

2. Determine the IR ratio using Eq. (8–62).
   Aerobic tank NO_3-N concentration = N_e = 6.0 g/m^3

$$IR = \frac{\text{NO}_x}{N_e} - 1.0 - R = \frac{(28.9\ \text{g/m}^3)}{(6\ \text{g/m}^3)} - 1.0 - 0.60 = 3.2$$

3. Determine the amount of NO_3-N fed to the anoxic tank.

$$\text{Flowrate to anoxic tank} = IR\,Q + RQ$$
$$= 3.2(22{,}700\ \text{m}^3/\text{d}) + 0.60(22{,}700\ \text{m}^3/\text{d})$$
$$= 82{,}260\ \text{m}^3/\text{d}$$

NO_x feed = $(82{,}260\ m^3/d)(6.0\ g/m^3)$ = 517,560 g/d

4. Determine the anoxic volume. At 20 percent aerobic τ, anoxic τ = 0.20(14.2h) = 2.8 h. As a first approximation, use a detention time = 2.5 h

$$\tau = \frac{2.5\ \text{h}}{(24\ \text{h/d})} = 0.104\ \text{d}$$

$$V_{\text{nox}} = \tau \times Q = 0.104\ \text{d}(22{,}700\ \text{m}^3/\text{d}) = 2361\ \text{m}^3$$

5. Determine F/M_b using Eq. (8–56).

$$\text{F/M}_\text{b} = \frac{QS_o}{V_{\text{nox}}(X_b)} = \frac{(22{,}700\ \text{m}^3/\text{d})(140\ \text{g BOD/m}^3)}{(2361\ \text{m}^3)(1267\ \text{g/m}^3)} = 1.06\ \text{g/g·d}$$

6. Determine the SDNR using Eq. (8–57).
   Fraction of rbCOD = rbCOD/bCOD = $(80\ g/m^3)/(224\ g/m^3)$ = 0.36 = 36%

$$\text{SDNR}_\text{b} = b_\text{o} + b_1[\ln(\text{F/M}_\text{b})]$$

From Table 8–22, use 30 percent rbCOD, $b_o = 0.235$, $b_1 = 0.141$

$$\text{SDNR}_\text{b} = 0.235 + 0.141[\ln(1.06)] = 0.243\ \text{g NO}_3\text{-N/g MLVSS, biomass·d}$$

Apply temperature correction using $\theta = 1.026$

$$\text{SDNR}_{12} = 0.243(1.026)^{12-20} = 0.198 \text{ g/g·d}$$

Apply recycle correction using Eq. (8–60) for $IR = 3.2$

$$\begin{aligned}\text{SDNR}_{\text{adj}} &= \text{SDNR}_{\text{IR1}} - 0.029\ln(\text{F/M}_b) - 0.012\\ &= 0.198 - 0.029\ln(1.06) - 0.012\\ &= 0.184 \text{ g/g·d}\end{aligned}$$

7. Determine the overall SDNR based on MLVSS.

   $\text{SDNR} = \text{SDNR}_b(\text{MLVSS}_b/\text{MLVSS})$

   $\text{SDNR} = 0.184 \text{ g/g·d}[(1267 \text{ g/m}^3)/(2370 \text{ g/m}^3)] = 0.10 \text{ g NO}_3\text{-N/g MLVSS·d}$

8. Determine the amount of NO_3-N that can be reduced using Eq. (8–51).
   a. Check NO_r based on $\tau = 2.5$ h.

   $$\begin{aligned}\text{NO}_r &= (V_{\text{nox}})(\text{SDNR})(\text{MLVSS, biomass})\\ &= (2361 \text{ m}^3)(0.184 \text{ g/g·d})(1267 \text{ g/m}^3) = 550{,}415 \text{ g/d}\end{aligned}$$

   Comparing 550,415 g/d versus 517,560 g/d, there is about 6% excess nitrate-removal capacity. Thus, $\tau = 2.5$ h is acceptable as the excess capacity calculated is within the range of assumptions used. If excess is greater than 20 percent, a smaller tank volume could be evaluated.

9. Go to nitrification step in design and determine net oxygen required.

   R_o (without denitrification) = 275.9 kg/h (See Step 17, Example 8–3)

   The amount of oxygen supplied by nitrate reduction is as follows:

   $$\text{Oxygen credit} = \left(\frac{2.86 \text{ g O}_2}{\text{g NO}_3\text{-N}}\right)[(28.9 - 6.0) \text{ g/m}^3]\left(\frac{22{,}700 \text{ m}^3}{\text{d}}\right)\left(\frac{1 \text{ kg}}{10^3 \text{ g}}\right)$$

   $$= 1487 \text{ kg/d} = 61.9 \text{ kg/h}$$

   $$\text{Net O}_2 \text{ required} = R_o = (275.9 - 61.9) \text{ kg/h} = 214.0 \text{ kg/h}$$

   Note the required aeration rate will decrease in proportion to a lower R_o. The oxygen required and aeration energy can be reduced by 22.4 percent.

10. Check alkalinity.
    a. Prepare an alkalinity mass balance.

    Alk to be added to maintain pH ~ 7 = Influent Alk − Alk used + Alk produced
       i. Influent alkalinity = 140 g/m^3 as $CaCO_3$
       ii. Alkalinity used = 7.14 (28.9 g NO_3-N/m^3) = 206.3 g/m^3
       iii. Alkalinity produced = 3.57 [(28.9 − 6) g/m^3] = 81.8 g/m^3
       iv. Alkalinity needed to maintain neutral pH = 70 g/m^3 as $CaCO_3$

    b. Solve the above expression for Alk to be added.

    $$\begin{aligned}\text{Alk to be added} &= (70 - 140 + 206.3 - 81.8) \text{ g/m}^3\\ &= 54.5 \text{ g/m}^3 \text{ as CaCO}_3\end{aligned}$$

    $$\begin{aligned}\text{Mass of alkalinity needed} &= (54.5 \text{ g/m}^3)(22{,}700 \text{ m}^3\text{/d})(1 \text{ kg}/10^3 \text{ g})\\ &= 1237 \text{ kg/d as CaCO}_3\end{aligned}$$

    c. Compare to alkalinity needed for nitrification only.

    For the nitrification only design, the alkalinity needed in Step 19b, Example 8–3, was 3094 kg/d as $CaCO_3$.

    $$\text{Alkalinity savings} = 3094 - 1237 = 1857 \text{ kg/d}$$

11. Determine anoxic zone mixing energy.

Mixing energy = 5 kW/10^3 m^3 (given)
Volume = 2361 m^3
Power = (2361 m^3)(5 kW/10^3 m^3) = 12 kW total

12. Prepare summary of anoxic design.

Item	Unit	Value
Effluent NO_3-N	g/m^3	6.0
Internal recycle ratio	Unitless	3.2
RAS recycle ratio	Unitless	0.6
Anoxic volume	m^3	2361
MLSS	g/m^3	3000
Overall SDNR	g NO_3-N/g MLVSS·d	0.10
Detention time	h	2.5
Reduction in oxygen demand	%	22.4
Mixing power	kW	12
Alkalinity required	kg/d as $CaCO_3$	1237

Part B—Design for MBR Process

1. Design conditions:

Item	Unit	Value
Influent flowrate	m^3/d	22,700
Temperature	°C	12.0
Membrane tank MLSS	g/m^3	12,000
Membrane tank MLVSS	g/m^3	9480[a]
Aerobic SRT	d	21.0
Anoxic mixing energy	kW/10^3 m^3	8
RAS recycle ratio	unitless	6.0
R_o	kg/h	275

[a] Same MLVSS/MLSS ratio as (a)

Note: g/m^3 = mg/L

2. Assumptions:
   a. RAS recycle ratio = 6.0
   b. Use same coefficients as the nitrification process design
   c. Mixing energy for anoxic reactor = 8 kW/10^3 m^3
   d. Membrane flux =16.1 L/m^2·h (see Fig. 8–57)
   e. Membrane tank volume to membrane area ratio = 0.025 m^3/m^2 (see Sec. 8–12)
   f. Fine bubble aeration alpha = 0.35 (see Fig. 8–40)
   g. Coarse bubble aeration alpha = 0.50 (see Fig. 8–40)
   h. Fine bubble clean water O_2 transfer efficiency = 35 percent

i. Coarse bubble clean water O_2 transfer efficiency = 10 percent
j. DO in aeration basin = 2.0 g/m^3
k. Site elevation is 500 m (pressure = 95.6 kPa)
l. Fine pore diffuser fouling factor F = 0.20

Solution

1. Determine the membrane surface area.
Flux = 16.1 L/m^2·h

$$\text{Area} = \frac{(22{,}700\ \text{m}^3/\text{d})(\text{d}/24\ \text{h})}{(16.1\ \text{L/m}^2\cdot\text{h})(1\ \text{m}^3/10^3\ \text{L})}$$
$$= 58{,}747\ \text{m}^2$$

2. Determine the membrane separation tank volume, V_m.

$$V_m = (0.025\ \text{m}^3/\text{m}^2)(58{,}747\ \text{m}^2) = 1469\ \text{m}^3$$

Therefore the hydraulic retention time for the separation tank, τ, is

$$\tau = [(1469\ \text{m}^3)/(22{,}700\ \text{m}^3/\text{d})](24\ \text{h/d}) = 1.55\ \text{h}$$

3. Establish MLSS mass balance around preanoxic tank, ignoring incremental solids production.
Membrane tank MLSS = 12,000 g/m^3
RAS flowrate = $6Q$

$$(Q)(0) + (6Q)12{,}000 = (Q + 6Q)X_{\text{NO}_x}$$

$$X_{\text{NO}_x} = \left(\frac{6}{7}\right)12{,}000 = 10{,}286\ \text{g/m}^3$$

$$X_{\text{preanox}} = X_{\text{NO}_x} = 10{,}286\ \text{g/m}^3$$

4. Determine the preaeration tank volume and active biomass concentration.
Tank volume:
From Step 13 of Example 8–3, $P_{X,\text{TSS}}$ = 1916.8 kg/d
Aerobic SRT = 20.6 d
Eq. (7–57) in Table 8–10: $(P_{X,\text{TSS}})\text{SRT} = X_{\text{TSS}}(V)$

$$X_{\text{TSS}}(V) = (X_{\text{pre}})V_{\text{pre}} + X_m(V_m) = (P_{X,\text{TSS}})\text{SRT}$$

$$[(10{,}286\ \text{g/m}^3)V_{\text{pre}} + (12{,}000\ \text{g/m}^3)(1469\ \text{m}^3)]\left(\frac{1\ \text{kg}}{10^3\ \text{g}}\right) = (1916.8\ \text{kg/d})(21.0\ \text{d})$$

$$V_{\text{pre}} = 2200\ \text{m}^3$$

Hydraulic retention time:

$$\tau = [(2200\ \text{m}^3)/(22{,}700\ \text{m}^3/\text{d})](24\ \text{h/d}) = 2.3\ \text{h}$$

Preaeration tank active biomass (X_b) concentration:
From Eq. (8–20) part A in Table 8–10,

$$P_{x_b} = \frac{Q(Y_{\text{H}})(S_o - S)}{1 + b_{\text{H}}(\text{SRT})}$$

$$= \frac{(22{,}700\ \text{m}^3/\text{d})(0.45\ \text{g VSS/g COD})(224\ \text{g bCOD/m}^3)(1\ \text{m}^3/10^3\ \text{L})}{[1 + (0.088\ \text{g/g}\cdot\text{d})(21.0\ \text{d})]}$$

$$= 803.4\ \text{kg/d}$$

$$\text{Fraction } X_b = \frac{P_{X,\text{b}}}{P_{X,\text{TSS}}} = \frac{(803.4 \text{ kg/d})}{(1916.8 \text{ kg/d})} = 0.42$$

$$X_{\text{pre},b} = V_{\text{NO}_x,b} = 0.42(10{,}286 \text{ g/m}^3) = 4320 \text{ g/m}^3$$

5. Determine the effluent NO_3-N concentration.
   Use Eq. (8–60) and $NO_x = 28.9 \text{ g/m}^3$ (see step 9 in nitrification design)

$$Q(\text{NO}_x) = N_e[Q + IRQ + RQ]$$

$$\text{NO}_x = N[1.0 + 0 + 6.0] = 28.9 \text{ g/m}^3$$

$$N_e = 4.1 \text{ g/m}^3$$

6. Determine the amount of NO_3-N fed to the anoxic tank.

$$\begin{aligned}\text{NO}_x \text{ feed} &= 6Q(N_e)\\ &= 6(22{,}700 \text{ m}^3\text{/d})(4.1 \text{ g/m}^3) = 558{,}420 \text{ g/d}\end{aligned}$$

7. Determine the size of anoxic zone.

$$\begin{aligned}\text{Assume } V_{\text{NO}_x} &= 0.20\,(V_{\text{pre}} + V_m)\\ &= 0.20(2200 + 1469) \text{ m}^3 = 734 \text{ m}^3\end{aligned}$$

8. Calculate F/M_b.

$$\text{F/M}_\text{b} = \frac{QS_o}{V_{\text{NO}_x} X_b} = \frac{(22{,}700 \text{ m}^3\text{/d})(140 \text{ g/m}^3)}{(734 \text{ m}^3)(4320 \text{ g/m}^3)} = 1.0 \text{ g/g·d}$$

9. Determine $SDNR_b$ using Eq. (8–57).
   As for activated sludge/clarifier design: $b_o = 0.235$, $b_1 = 0.141$.

$$\text{SDNR}_{20} = 0.235 + 0.141[\ln(1.0)] = 0.235 \text{ g/g·d}$$

Apply temperature correction using $\theta = 1.026$

$$\text{SDNR}_{12} = 0.235(1.026)^{12-20} = 0.191 \text{ g/g·d}$$

Apply recycle correction using Eq. (8–64)

$$\begin{aligned}\text{SDNR}_{\text{adj}} &= \text{SDNR}_{\text{IR1}} - 0.029 \ln(\text{F/M}_\text{b}) - 0.012\\ &= 0.191 - 0.029 \ln(1.0) - 0.012\\ &= 0.179 \text{ g/g·d}\end{aligned}$$

10. Determine the overall SDNR based on MLVSS.

$$\text{SDNR} = \text{SDNR}_\text{b}(\text{MLVSS}_\text{b}/\text{MLVSS})$$

$$\text{SDNR} = 0.179 \text{ g/g·d}(4320 \text{ g/m}^3/8126 \text{ g/m}^3) = 0.10 \text{ g NO}_3\text{-N/g MLVSS·d}$$

11. Determine NO_x removal in anoxic tank.

$$\begin{aligned}\text{NO}_x &= (V_{\text{no}_x})(\text{SDNR})(\text{MLVSS}, X_b)\\ &= (734 \text{ m}^3)(0.179 \text{ g/g·d})(4320 \text{ g/m}^3) = 567{,}360 \text{ g/d}\end{aligned}$$

Removal required (from above) = 558,420 g/d

Use V_{NO_x} assumed.

12. Calculate the oxygen demand with the nitrate used for influent BOD removal. R_o from nitrification design = 275.9 kg/h.

$$\text{Oxygen credit} = \left(\frac{2.86\ \text{gO}_2}{\text{g NO}_3\text{-N}}\right)(28.9 - 4.1)\ \text{g/m}^3\left(\frac{22{,}700\ \text{m}^3}{\text{d}}\right)\left(\frac{1\ \text{kg}}{10^3\ \text{g}}\right)$$
$$= 1610\ \text{kg/d} = 67\ \text{kg/h}$$

Net O_2 received = R_o = (275.9 − 67) kg/h = 208.9 kg/h
Oxygen required can be reduced by 24.3 percent.

13. Check alkalinity
    i. influent alkalinity = 140 g/m^3 as $CaCO_3$
    ii. alkalinity used = 7.14 (28.9 g NO_3/m^3) = 206.3 g/m^3
    iii. alkalinity produced = 3.57 (28.9 − 4.1) g/m^3 = 88.5 g/m^3

    Solve above expressions for alkalinity to be added

$$\text{Alkalinity to be added} = (70 - 140 + 206.3 - 88.5)\ \text{g/m}^3$$
$$= 47.8\ \text{g/m}^3 \text{ as CaCO}_3$$

$$\text{Mass of alkalinity needed} = (47.8\ \text{g/m}^3)(22{,}700\ \text{m}^3/\text{d})\left(\frac{1\ \text{kg}}{10^3\ \text{g}}\right)$$
$$= 1085\ \text{g/m}^3 \text{ as CaCO}_3$$

14. Determine anoxic zone mixing.

Mixing energy = 8 kW/10^3 m^3 (given)
Volume = 734 m^3
Power = (734 m^3)(8 kW/10^3 m^3) = 5.9 kW total

15. Design the aerator oxygen transfer system.

a. Assume same liquid depth as for nitrification design in Example 8–3.

From Example 8–3, Step 8, $C^*_{\infty 20} = 10.64$

Total oxygen demand is divided between preaeration and membrane tank

Total volume = 2200 m^3 + 1469 m^3 = 3669 m^3
Percent preaeration = 60%
Percent membrane tank = 40%

Most of the nitrification and all of the BOD removal will occur in the preaeration tank. Assume 90 percent of O_2 demand in preaeration tank. From Step 11, the total O_2 demand is 208.9 kg/h.

O_2 demand in preaeration tank: 0.90 (208.9 kg/h) = 188.0 kg/h
O_2 demand in membrane tank: 0.10 (208.9 kg/h) = 20.9 kg/h

b. Design oxygen air rate needed for preaeration tank.
From nitrification design and input assumptions,

$\alpha = 0.35$, $\beta = 0.95$, $F = 0.90$
$C^*_{\infty 20} = 10.64$ g/m^3, $P_b/P_s = 0.94$, $C^*_{s20} = 9.09$

$$\text{SOTR} = \left[\frac{\text{OTR}_f}{(\alpha)(F)}\right]\left[\frac{C^*_{\infty 20}}{(\beta)(C_{st}/C^*_{s20})(P_b/P_s)(C^*_{\infty 20}) - C}\right][(1.024)^{20-T}]$$

$$\text{SOTR} = \left[\frac{(188.0\ \text{kg/h})}{(0.35)(0.90)}\right]\left\{\frac{10.64}{\left[(0.95)\left(\frac{10.78}{9.09}\right)(0.94)(10.64) - 2.0\right]}\right\}(1.024^{20-12}) = 828.3\ \text{kg/h}$$

c. Determine the air flowrate for oxygen demand for preaeration tank

$$\text{Air flowrate, m}^3/\text{min} = \frac{(\text{SOTR kg/h})}{[(E)(60\ \text{min/h})(\text{kg O}_2/\text{m}^3\ \text{air})]}$$

Using the data given in Appendix B, the density of air at 12ºC and pressure of 95.2 kPa (0.94 × 101.325 kPa) is 1.1633 kg/m^3. The corresponding amount of oxygen by weight is 0.270 kg O$_2$/m^3 air (0.2318 × 1.1633 kg/m^3). Thus, the required air flowrate is

$$\begin{aligned}\text{Air flowrate, m}^3/\text{min} &= \frac{(828.3\ \text{kg/h})}{[(0.35)(60\ \text{min/h})(0.270\ \text{kg O}_2/\text{m}^3\ \text{air})]}\\ &= 146.1\ \text{m}^3/\text{min}\end{aligned}$$

d. Design oxygen air rate needed for oxygen demand for membrane separation tank.

From nitrification design and input assumptions,

$\alpha = 0.50$, $\beta = 0.95$, $F = 1.0$

$C^*_{\infty 20} = 10.64\ \text{g/m}^3$, $P_b/P_s = 0.94$, $C^*_{s20} = 9.09$

$$\begin{aligned}\text{SOTR} &= \left[\frac{(20.9\ \text{kg/h})}{(0.50)(1.0)}\right]\left\{\frac{10.64}{\left[(0.95)\left(\dfrac{10.78}{9.09}\right)(0.94)(10.64)\ -\ 2.0\right]}\right\}(1.024^{20-12})\\ &= 58.0\ \text{kg/h}\end{aligned}$$

e. Determine the air flowrate for oxygen demand

$$\text{Air flowrate, m}^3/\text{min} = \frac{(\text{SOTR kg/h})}{[(E)(60\ \text{min/h})(\text{kg O}_2/\text{m}^3\ \text{air})]}$$

Using the data given in Appendix B, the density of air at 12ºC and pressure of 95.2 kPa (0.94 × 101.325 kPa) is 1.1633 kg/m^3. The corresponding amount of oxygen by weight is 0.270 kg O$_2$/m^3 air (0.2318 × 1.1633 kg/m^3). Thus, the required air flowrate is

$$\begin{aligned}\text{Air flowrate, m}^3/\text{min} &= \frac{(58.0\,\text{kg/h})}{[(0.10)(60\ \text{min/h})(0.270\ \text{kg O}_2/\text{m}^3\ \text{air})]}\\ &= 35.8\ \text{m}^3/\text{min}\end{aligned}$$

16. Summary table of MBR design.

Design parameter	Unit	Value
Average flowrate	m^3/ d	22,700
Average BOD load	kg/d	3178
Average TKN load	kg/d	795
bCOD	g/m^3	224
rbCOD	%	36
NO$_x$	g/m^3	28.9

(continued)

(*Continued*)

Design parameter	Unit	Value
Temperature	°C	12.0
Total aerobic SRT	d	21.0
Preaeration Tank		
Volume	m^3	2200
Detention time	h	2.3
MLSS	g/m^3	10,286
MLVSS	g/m^3	8126
Oxygen demand	kg/h	188
Aeration rate	m^3/min	146
Membrane tank		
Volume	m^3	1469
Detention time	h	1.6
MLSS	g/m^3	12,000
MLVSS	g/m^3	9480
Oxygen demand	kg/h	21
Aeration rate (for O_2)	m^3/min	36
Membrane flux	$L/m^2 \cdot h$	16.1
Membrane area	m^2	58,750
RAS ratio	unitless	6.0
RAS flowrate	m^3/d	136,200
Anoxic tank		
Effluent NO_3-N	g/m^3	4.1
Volume	m^3	734
Detention time	h	0.8
MLSS	g/m^3	10,286
Overall SDNR	g NO_3-N/g MLVSS·d	0.10
Mixing power	kW	5.8
Alkalinity required (as $CaCO_3$)	kg/d	1085

Comment In the above design example, computations are based on an average design condition. In actual design, allowances should be provided for peak flows and loads or a safety factor should be included, as discussed in Chap. 3. Note that some process engineers have added a *swing zone* between the upstream preanoxic and aerobic nitrification zones. The swing zone provides operational flexibility and can be operated with aeration or with only mixing (anoxic) to provide additional volume where most needed for nitrification or denitrification.

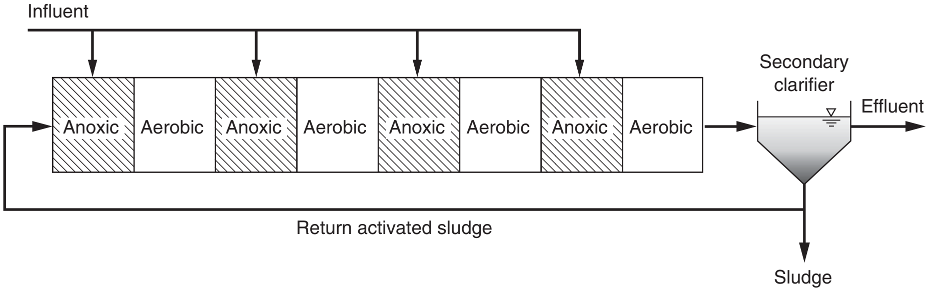


Figure 8–33

Schematic diagram of a step feed biological nitrogen removal process.

Step Feed Nitrification and Preanoxic Denitrification Process Design. Step feed for nitrogen removal is similar to the step feed process described in Sec. 8–4 for BOD and nitrification. For nitrogen removal, wastewater is introduced at several feed points (see Fig. 8–33). In most cases, where a step feed process is used for BOD removal and nitrification, it will be relatively easy to upgrade it to a step feed anoxic/aerobic biological nitrogen-removal process. For such applications the influent feed points and volumes of the individual channels in the reactor (passes) are already determined. The tank layout is generally symmetrical and the volume in each pass is equal. For a new tank design, it is possible to use a nonsymmetrical step feed design where the feed split is somewhat equal, but the volume of each pass increases as the mixed-liquor concentration decreases from the first to last pass. The tank volumes may be utilized more efficiently in the nonsymmetrical design approach by using a similar F/M ratio for each pass.

The variables involved in the design of a step feed biological nitrogen removal process for an existing basin are (1) the flow distribution between passes, (2) the relative split between anoxic and aerobic volumes, and (3) the final pass MLSS concentration. The selection of the final pass MLSS concentration is based on using an acceptable solids loading for the secondary clarifier. As will be illustrated in Example 8–8, selection of the final pass MLSS concentration, the RAS ratio, the influent flow split, and wastewater characteristics will determine the system SRT. With the known SRT value, the biomass and nitrifying bacteria concentration in the mixed liquor can be determined, which can then be used to determine the nitrification and denitrification capacity of the system. The process design procedure involves successive iterations with varying anoxic/aerobic volumes and flow splits evaluated to find the most satisfactory design. Design of a step feed BNR process is presented in Example 8–8.

EXAMPLE 8–8 **Step Feed Biological Nitrogen Removal Process Design** Determine the amount of nitrification and nitrate removal in a four-pass step feed biological nitrogen removal process (see Fig. 8–33), using the same influent flowrate, wastewater characteristics, and temperature as was used in Example 8–7, and total aerobic tank volume obtained for the nitrification solution.

1. Design conditions and assumptions:
    a. The flowrate is 22,700 m^3/d
    b. The step feed aeration tank is divided into four equal passes, with equal volumes used for the anoxic and aerobic zones
    c. The flow split to each pass is 0.10, 0.40, 0.30, and 0.20 of the influent flow, for passes 1 through 4, respectively

d. MLSS concentration in the final aerobic zone is 3000 mg/L (same as in Example 8–7)
e. RAS recycle ratio (Q_{RAS}/Q) is 0.6
f. Anoxic volume is 15% of the total reactor volume
g. Total aeration tank volume = 13,230 m^3
h. Aeration tank DO = 2.0 g/m^3
i. Temperature = 12°C
k. Effluent NH_4-N = 0.5 g/m^3

2. Wastewater characteristics:

Constituent	Concentration, g/m^3
BOD	140
rbCOD	80
bCOD	224
nbVSS	20
TKN	35
$TSS_o - VSS_o$	10

Note: g/m^3 = mg/L.

3. Kinetic coefficients:
The following kinetic coefficients, from Table 8–14, have been adjusted for temperature (Example 8–3)
a. Heterotrophs:

$$Y = 0.45 \text{ g VSS/g bCOD}$$

$$b_{12°C} = 0.088 \text{ g/g·d}$$

$$f_d = 0.15 \text{ g/g}$$

b. Nitrifiers:

$$Y = 0.15 \text{ g VSS/g } NO_x$$

$$b_{AOB,12°C} = 0.135 \text{ g/g·d}$$

$$\mu_{max,AOB,12°C} = 0.52 \text{ g/g·d}$$

$$K_{NH_4} = 0.50 \text{ g/m}^3$$

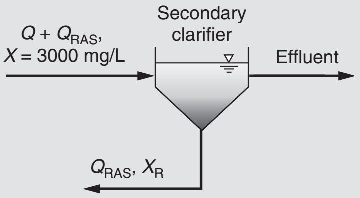
$$K_o = 0.50 \text{ g/m}^3$$

Solution

1. Determine the aeration and anoxic zone volumes. Anoxic volume is 15% of total volume per problem statement.

V = total volume = 13,230 m^3
Anoxic volume = 0.15(13,230 m^3) = 1984.5 m^3
Aerobic volume = 0.85(13,230 m^3) = 11,245.5 m^3
Anoxic volume/pass = (1984.5 m^3)/4 = 496.1 m^3
Aerobic volume/pass = (11,245.5 m^3)/4 = 2811.4 m^3

2. Determine RAS concentration.



Perform a solids balance (neglect the effluent TSS because it is not significant for the clarifier solids balance).

$Q_{\mathrm{RAS}} = 0.6Q$ (recycle ratio from problem statement)

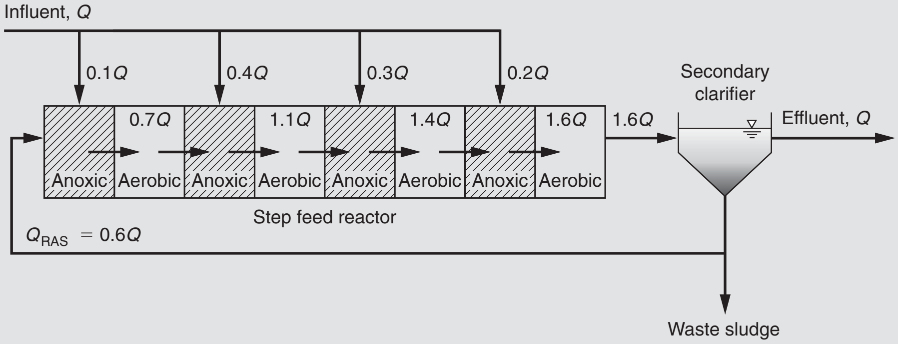
The solids balance is

$$(Q + 0.6Q)3000\ \mathrm{g/m^3} = 0.6QX_{\mathrm{R}}$$

$$X_{\mathrm{R}} = \frac{(Q + 0.6Q)(3000\ \mathrm{g/m^3})}{0.6Q}$$

$$X_{\mathrm{R}} = (1.6Q/0.6Q)(3000\ \mathrm{g/m^3}) = 8000\ \mathrm{g/m^3}$$

3. Determine MLSS concentration in each pass (see figure below).



a. Pass 1 mass balance.
   Solids in = solids out (Note: solids production for a single pass is negligible)

$$0.10Q(0) + (Q_{\mathrm{RAS}})(8000\ \mathrm{g/m^3}) = (\mathrm{RAS} + 0.1Q)X_1$$

$$0.10Q(0) + (0.6Q)(8000\ \mathrm{g/m^3}) = (0.6Q + 0.1Q)X_1$$

$$X_1 = (8000\ \mathrm{g/m^3})(0.6/0.7) = 6860\ \mathrm{g/m^3}$$

b. Pass 2 mass balance.

$$(0.7Q)X_1 + 0.4Q(0) = 1.1QX_2$$

$$X_2 = (0.7/1.1)X_1 = (0.7/1.1)6860\ \mathrm{g/m^3}$$

$$= 4365\ \mathrm{g/m^3}$$

c. Pass 3 and 4 are calculated similarly.
d. Summary of MLSS concentrations and volumes:

Pass	MLSS, g/m^3	Anoxic volume, m^3	Aerobic volume, m^3
1	6860	496	2811
2	4365	496	2811
3	3430	496	2811
4	3000	496	2811

4. Perform solids balance on system and determine aerobic SRT.
   a. Solids balance.

$$\sum X_i V_i = (P_{X,\text{TSS}})(\text{SRT})$$

$$\begin{aligned}\sum X_i V_i &= X_1V_1 + X_2V_2 + X_3V_3 + X_4V_4 \\ &= [(6860 + 4365 + 3430 + 3000)\text{ g/m}^3](2811\text{ m}^3)(1\text{ kg}/10^3\text{ g}) \\ &= 49{,}628\text{ kg}\end{aligned}$$

   b. Apply Eq. (8–20) and (7–57) in Table 8–10 assuming $\text{NO}_x \sim 0.80$ TKN.

MLSS components = heterotrophic biomass + cell debris + nitrifier biomass + nbVSS + inorganic inerts

$$\begin{aligned}(\text{SRT})(P_{X,\text{TSS}}) = {} & \frac{QY_H(S_o - S)\text{SRT}}{[1 + b_H(\text{SRT})](0.85)(10^3\text{ g/1 kg})} + \frac{f_d(b_H)QY_H(S_o - S)(\text{SRT})^2}{[1 + b_H(\text{SRT})](0.85)(10^3\text{ g/1 kg})} \\ & + \frac{QY_n(\text{NO}_x)\text{SRT}}{[1 + b_n(\text{SRT})](0.85)(10^3\text{ g/1 kg})} + \frac{Q(\text{nbVSS})\text{SRT}}{(10^3\text{ g/1 kg})} \\ & + \frac{Q(\text{TSS}_o - \text{VSS}_o)\text{SRT}}{(10^3\text{ g/1 kg})}\end{aligned}$$

$$\text{NO}_x = 0.80(35\text{ g/m}^3) = 28\text{ g/m}^3$$

   c. Solve for SRT.
   Substituting wastewater and coefficient values yields

$$\begin{aligned}49{,}628\text{ kg} = {} & \frac{(22{,}700\text{ m}^3\text{/d})(0.45\text{ g/g})(224\text{ g/m}^3)\text{SRT}}{[1 + (0.088\text{ g/g}\cdot\text{d})(\text{SRT})](0.85)(10^3\text{ g/1 kg})} \\ & + \frac{(0.15\text{ g/g})(0.088\text{ g/g}\cdot\text{d})(0.45\text{ g/g})(22{,}700\text{ m}^3\text{/d})(224\text{ g/m}^3)(\text{SRT})^2}{[1 + (0.088\text{ g/g}\cdot\text{d})(\text{SRT})](0.85)(10^3\text{ g/1 kg})} \\ & + \frac{(22{,}700\text{ m}^3\text{/d})(0.15\text{ g/g})(28\text{ g/m}^3)\text{SRT}}{[1 + (0.135\text{ g/g}\cdot\text{d})(\text{SRT})](0.85)(10^3\text{ g/1 kg})} \\ & + \frac{(22{,}700\text{ m}^3\text{/d})(20\text{ g/m}^3)\text{SRT}}{(10^3\text{ g/1 kg})} \\ & + \frac{(22{,}700\text{ m}^3\text{/d})(10\text{ g/m}^3)\text{SRT}}{(10^3\text{ g/1 kg})}\end{aligned}$$

$$49{,}628\text{ kg} = \frac{2691\text{SRT}}{1 + 0.088\text{ SRT}} + \frac{35.5\text{ SRT}^2}{1 + 0.088\text{ SRT}} + \frac{112.2\text{ SRT}}{1 + 0.135\text{ SRT}} + 454\text{ SRT} + 227\text{ SRT}$$

Solve using spreadsheet solver function or by successive iterations,

SRT = 28.1 d.

Compare to Example 8–3 with nitrification. In Example 8–3, aerobic volume = 13,230 m^3 and SRT = 20.6 d. For step feed, aerobic volume = 11,245 m^3 and SRT = 28.1 d. The lower volume and higher SRT are due to the higher MLSS concentrations in the first 3 passes.

5. Determine the composition of the MLSS and MLVSS using the above solution in Step 4 for MLSS and SRT = 28.1 d. The results of the calculation of the MLVSS components are summarized in the following table which summarizes the solids mass in the aeration tank:

Item	MLVSS, kg	Fraction of total MLVSS	MLSS, kg
Heterotrophs	18,511	0.48	21,777
Cell debris	6861	0.18	8072
Nitrifiers	559	0.04	657
nbVSS	12,748	0.33	12,748
Inert inorganics			6374
Total	38,679		49,628

a. Fraction of biomass solids (from above table).
Thus, MLVSS/MLSS = 38,679/49,628 = 0.78

Biomass = 0.48(MLVSS)

Nitrifiers = 0.01(MLVSS)

b. Nitrogen for nitrifier growth.
With biomass and cell debris calculated, the NO_x for nitrifier growth is calculated as follows:

Daily biomass + debris production = (18,511 + 6861) kg = 25,372 kg

Daily wasting = 25,372 kg/28.1 d SRT = 902.9 kg/d

$$\text{N used for synthesis} = (0.12\text{ g N/g VSS biomass})(902.9\text{ kg/d}) = 108.4\text{ kg/d}$$

Based on influent flow the N synthesis is

$$\text{N synthesis} = \frac{(108.4\text{ kg/d})(10^3\text{ g/1 kg})}{(22{,}700\text{ m}^3\text{/d})} = 4.8\text{ g/m}^3$$

$$NO_x = \text{TKN} - N_{syn} - (NH_4\text{-N})_e = (35 - 4.8 - 0.5)\text{ g/m}^3 = 29.7\text{ g/m}^3$$

c. Nitrifier mass fraction correction.

$$\text{Mass nitrifiers} = \frac{QY_n(\text{NO}_x)\text{SRT}}{[1 + b_n(\text{SRT})]}$$

$$= \frac{(22{,}700\ \text{m}^3/\text{d})(0.15\ \text{g/g})(29.7\ \text{g/m}^3)(28.1\ \text{d})}{[1 + (0.135\ \text{g/g·d})(28.1\ \text{d})](10^3\ \text{g/1 kg})} = 592.8\ \text{kg VSS}$$

Corrected MLVSS = (38,679 − 559 + 593) kg = 38,713 kg

Nitrifiers as fraction of MLVSS = 593/38,713 = 0.015 (as compared to 0.014)

d. Summary Table.
Based on the above data, prepare a summary table of biomass and nitrifier (AOB) concentrations in each pass using ratios from above.

Pass	MLSS, g/m³	MLVSS, kg/d	Nitrifiers, g VSS/m³	Biomass, g VSS/m³
1	6860	5350	80.3	2568
2	4365	3405	51.1	1634
3	3430	2675	40.1	1284
4	3000	2340	35.1	1123

6. Evaluate nitrification rate in each pass and compare to NH_4-N fed to stage.

a. Develop equation for nitrification rate in g/d. Multiply Eq. (7–101) in Table 8–10 by the reactor volume. Eq. (7–101) describes volumetric nitrification rate (g/L-d) as a function of the NH_4-N and DO concentration. The nitrification rate in each stage is:

$$R_n = \frac{\mu_{\text{max,AOB}}}{Y_{\text{AOB}}}\left(\frac{S_{\text{NH}_4}}{K_{\text{NH}_4} + S_{\text{NH}_4}}\right)\left(\frac{S_o}{K_{o,\text{AOB}} + S_o}\right)X_{\text{AOB}}V$$

As indicated in the steady state nitrogen mass balance below for each pass, the rate of oxidizable nitrogen (available nitrogen NH_4-N = NO_x) fed to the pass equals the nitrification rate plus the rate of ammonia nitrogen leaving the pass. The nitrogen entering the pass is related to the influent feed rate of available nitrogen (NO_x) to that pass and the rate of influent nitrogen from a previous pass. Before proceeding with the mass balance, prepare a flowrate summary.

b. Prepare flowrate summary and mass balances.

Pass	Flowrate from previous reactor	Influent flowrate	Total Flowrate
1	0.6 Q	0.1 Q	0.7 Q
2	0.7 Q	0.4 Q	1.1 Q
3	1.1 Q	0.3 Q	1.4 Q
4	1.4 Q	0.2 Q	1.6 Q

The balances for the 4 passes are shown for steady state:

Pass 1 (recycle NH_4-N) = last pass (pass 4) NH_4-N

In general for each pass, rate of influent N = rate of effluent N + nitrification rate

$$\text{RAS}(Q)S_{\text{NH},4} + 0.1(Q)\text{NO}_x = 0.7(Q)S_{\text{NH},1} + R_{n,1}$$

$$0.7QS_{\text{NH},1} + 0.4Q\text{NO}_x = 1.1QS_{\text{NH},2} + R_{n,2}$$

$$1.1QS_{\text{NH},2} + 0.3Q\text{NO}_x = 1.4QS_{\text{NH},3} + R_{n,3}$$

$$1.4QS_{\text{NH},3} + 0.2Q\text{NO}_x = 1.6QS_{\text{NH},4} + R_{n,4}$$

The nitrification rate (Rn) for pass 1 is shown below and the rate expression applies to the other passes also with their respective NH_4-N, DO and AOB concentrations and volumes.

$$R_{n,1} = \frac{\mu_{\max,\text{AOB}}}{Y_{\text{AOB}}}\left(\frac{S_{\text{NH}_4,1}}{K_{\text{NH}_4} + S_{\text{NH}_4,1}}\right)\left(\frac{S_{o,1}}{K_{o,\text{AOB}} + S_{o,1}}\right)X_{\text{AOB},1}V_1$$

c. Solve for nitrification rate in each stage with $\mu_{\max,\text{AOB}} = 0.52$ g/g·d, $Y_{\text{AOB}} = 0.15$ g/g, and X_{AOB} from table above. Thus, N_1, N_2, N_3, and N_4 are solved for each stage using $\text{NO}_x = 29.7\ \text{g/m}^3$ (determined earlier). The value for N_4 is assumed to solve pass 1 and then the final solution is reached by iteration until the value for N_4 equals the value used to solve pass 1. A spreadsheet program with a solver function can assist in the solution. Using the coefficients shown for the design condition, the solution is summarized below.

d. Summary of step-feed solution for NH_4-N per pass:

Pass	Nitrifiers, g/m^3	Influent flowrate, m^3/d	RAS or flow from previous pass, m^3/d	NH_4-N, g/m^3	R_n, g/d
1	80.3	2270	13,620	0.07	73,713
2	51.1	9080	15,890	0.84	249,754
3	40.1	6810	24,970	0.84	196,389
4	35.1	4540	31,780	0.54	142,076

Sufficient volume exists to provide an effluent NH_4-N concentration of 0.54 g/m^3 even though the total aerobic volume for this step feed design is less than that used for the CMAS design in Example 8–7. The volume in pass 1 is much greater than that needed to achieve an effluent NH_4-N concentration of 0.5 g/m^3. Further adjustment of step feed flow splits may be made to optimize NH_4-N removal, but it would also affect the NO_3-N removal.

7. Determine the amount of nitrate removal in the anoxic zones and effluent NO_3-N concentration.

The amount of nitrate fed to each anoxic zone for passes 2, 3, and 4 equals the nitrification rate (g/d) in the previous pass plus the nitrate not removed in the previous anoxic zones. For the first anoxic zone, the nitrate feed is equal to the effluent NO_3-N concentration times the RAS flowrate (assuming no denitrification in the secondary clarifier). For the solution, the first step is to determine the denitrification capacity for each anoxic stage by calculating the F/M_b per pass and using Eq. (8–57) and Table 8–22 to obtain the SDNR as follows:

a. Calculate F/M_b from Eq. (8–56) and BOD = 140 g/m³.

$$F/M_b = \frac{QS_o}{(V_{nox})(X_b)} \; (x_b \text{ is found from Step 5d})$$

b. Obtain $SDNR_b$ using same b_0 and b_1 values in Eq. (8–57) used in Example 8–6 for same wastewater characteristics.

c. Apply $\theta = 1.026$ for temperature correction. There is no correction for recycle flow.

$$SDNR_{12} = SDNR_{20}(1.026)^{12-20} = SDNR_{20}(0.814)$$

d. NO_3-N removed = $(SDNR_b)(V_{nox})X_b$

Sample calculation for pass 1:

$$F/M_b = \frac{(2270\ \text{m}^3/\text{d})(140\ \text{g/m}^3)}{(2568\ \text{g/m}^3)(496\ \text{m}^3)} = 0.25\ \text{g/g·d}$$

$$SDNR = 0.235 + 0.141[\ln(F/M_b)]$$

$$SDNR = 0.235 + 0.141[\ln(0.25)] = 0.04\ \text{g/g·d}$$

$$SDNR_{12} = 0.04(0.814) = 0.03$$

Pass	X_b, g/m³	Influent flowrate, m³/d	Anoxic volume, m³	F/M_b	$SDNR_{12}$, g/g·d
1	2568	2270	496	0.25	0.03
2	1634	9080	496	1.57	0.24
3	1284	6810	496	1.50	0.24
4	1123	4540	496	1.14	0.25

Using the SDNR from the above for each pass, the amount of NO_3-N removed is tabulated below.

Pass	NO_3-N removal capacity, g/g·d
1	40,987[a]
2	195,511
3	152,847
4	139,252

[a] NO_3-N removed = (0.03 g/g·d)(2568 g/m³)(496 m³) = 40,987 g/d.

8. Nitrate balance and effluent NO_3-N concentration.

a. Develop equations for nitrate mass balance.

A nitrate balance for each pass is done to determine how much nitrate remains after the anoxic reactor and the effluent nitrate nitrogen concentration. The nitrogen remaining after each reactor is

$$\begin{array}{c} NO_3\text{-N influent} \\ \text{from RAS or} \\ \text{previous pass} \end{array} + \begin{array}{c} \text{pass anoxic} \\ \text{removal} \\ \text{capacity} \end{array} = \begin{array}{c} NO_3\text{-N} \\ \text{remaining after} \\ \text{anoxic reactor} \end{array}$$

If there is a negative value for NO_3-N remaining because of excess anoxic NO_3-N removal capacity, then a value of zero is assigned [see column (3) in the following table].

The effluent NO_3-N concentration from each pass is then:

$$\begin{matrix} NO_3\text{-N remaining} \\ \text{after anoxic} \end{matrix} - \begin{matrix} NO_3\text{-N produced} \\ \text{in the pass} \end{matrix} = \begin{matrix} \text{effluent} \\ NO_3\text{-N} \end{matrix}$$

b. Prepare table to solve mass balance for each stage. The NO_3-N balance is illustrated in the following table.

Pass	Total NO_3-N to pass, g/d	Anoxic removal capacity, g/d[a]	NO_3-N remaining after anoxic, g/d[b]	NO_3-N produced (R_n) in pass, g/d[c]	Effluent NO_3-N, g/d[d]
	(1)	(2)	(3)	(4)	(5)
1	111,003	40,987	70,016	73,713	143,729
2	143,729	195,511	0	249,754	249,754
3	249,754	152,847	96,907	196,389	293,296
4	293,296	139,252	154,044	142,076	296,120
					Effluent NO_3-N = 8.15 g/m^3

[a] Source of influent nitrate: for pass 1, RAS; other passes, column (5).

[b] (1) − (2).

[c] From Step 6.

[d] (3) + (4).

c. Determine effluent NO_3-N concentration.

The nitrate fed to pass 1 from RAS is calculated as follows:

$$NO_3\text{-N to pass 1} = (Q_{RAS})N_e$$

where N_e = effluent NO_3-N concentration, g/m^3

The value for N_e is obtained using the effluent NO_3-N shown in column (5) for pass 4:

$$(Q + Q_{RAS})N_e = \text{pass 4 effluent } NO_3\text{-N (g/d)}$$

$$N_e = (296{,}120 \text{ g/d})/(Q + 0.6\,Q)$$

$$N_e = (296{,}120 \text{ g/d})/(1.6)(22{,}700 \text{ g/m}^3) = 8.15 \text{ g/m}^3$$

9. Reevaluate design.

Initially, assume an effluent concentration, e.g., 10 g/m^3, to calculate the amount of NO_3-N fed to pass 1. Using the spreadsheet, perform successive iterations until the calculated effluent NO_3-N equals the trial.

Comment Note that some excess capacity is available and is not realized in the initial passes of the symmetrical step feed process design. Different influent flow splits may be used to reduce the effluent nitrate concentration, the anoxic and aerobic volumes can be changed, the anoxic zone may be staged, and the MLSS concentration may be increased. A spreadsheet model is necessary to evaluate various design changes.

Sequencing Batch Reactor Preanoxic Denitrification Process Design. In the SBR process and other batch decant processes [see Figs. 8–1(c) and 8–20] nitrate removal can be accomplished by four methods: (1) nitrate reduction by using a mixed nonaerated fill period, (2) cycling aeration On/Off during the react period, (3) having an anoxic contact time after nitrification is complete and just before a short aeration time, settling and decant, and (4) operating at a low DO concentration to encourage SNdN during the aeration sequence. Under cyclic aeration conditions after the fill and reaction period, the SDNR is driven mainly by endogenous respiration. Denitrification during a mixed nonaerated fill period provides the most efficient means of nitrate removal and also provides a selector operation to prevent filamentous sludge bulking. Most of the nitrate produced during the previous aerobic cycle remains in the SBR tank because the decant volume is only 20 to 30 percent of the total tank volume. The mass of nitrate remaining after decant can be reduced during the fill period if sufficient BOD and time are available. The following example is used to illustrate how to estimate the amount of nitrate removed during a mixed fill period for an SBR reactor.

EXAMPLE 8–9 **Preanoxic Denitrification Process Design for SBR Process** Determine how much nitrate may be removed during a mixed, unaerated fill period in an SBR for the following design conditions. Assume the following design conditions apply.

Design conditions:

Item	Unit	Value
Number of tanks	no.	2
Flowrate/tank	m^3/d	3785
Sequence time	h	
Fill		4.0
Aerate		3.0
Settle		0.5
Decant		0.5
Fill volume fraction	V_F/V_T	0.25
Influent BOD	g/m^3	200
Influent bCOD	g/m^3	320
Influent rbCOD	g/m^3	60
Influent TKN	g/m^3	35
Effluent NH_4-N	g/m^3	0.5
SRT	d	20
Temperature	°C	16

Note: $g/m^3 = mg/L$.

Use kinetic coefficients from Table 8–14.

Solution

1. Determine tank volume.

$$\text{Cycles/day} = \frac{(24\ \text{h/d})}{(8\ \text{h/cycle})} = 3\ \text{cycles/d}$$

$$\text{Fill volume/cycle} = \frac{(3785\ \text{m}^3/\text{d})}{3\ \text{cycles}} = 1261.7\ \text{m}^3/\text{fill cycle}$$

$$V_F/V_T = 0.25$$

$$V_T = \frac{1261.7\ \text{m}^3}{0.25} = 5{,}047\ \text{m}^3$$

2. Determine nitrate produced (NO_x).
   a. Determine heterotrophic biomass produced.
      Determine heterotrophic biomass production including cell debris to estimate the nitrogen used for synthesis using Eq. (8–24) in Table 8–10. Use Eq. (8–20) in Table 8–10 to determine the heterotrophic biomass production. (Biomass produced from nitrification can be ignored, as it represents a small fraction of the biomass.)

$$P_{x,\text{bio}} = \frac{Q(Y_\text{H})(S_o - S)}{1 + b_\text{H}(\text{SRT})} + \frac{f_d(b_\text{H})Y_\text{H}(Q)(S_o - \text{S})\text{SRT}}{1 + b_\text{H}(\text{SRT})}$$

      i. Use coefficients from Table 8–14 and adjust b for temperature.

$$b_\text{H} = 0.12(1.04)^{16-20} = 0.103\ \text{g/g·d}$$

$$Y_\text{H} = 0.45\ \text{g VSS/g bCOD}$$

$$f_d = 0.15$$

      Assume $S_o - S \approx S_o$.

      ii. Substitute and solve for P_X.

$$P_x = \frac{(3785\ \text{m}^3/\text{d})(0.45\ \text{g/g})(320\ \text{g/m}^3)}{[1 + (0.103\ \text{g/g·d})(20\ \text{d})](10^3\ \text{g/1 kg})}$$

$$+ \frac{(0.15\ \text{g/g})(0.103\ \text{g/g·d})(0.45\ \text{g/g})(3785\ \text{m}^3/\text{d})(320\ \text{g/m}^3)(20\ \text{d})}{[1 + (0.103\ \text{g/g·d})(20\ \text{d})](10^3\ \text{g/1 kg})}$$

$$= (178.1 + 55.0)\ \text{kg/d} = 233.1\ \text{kg/d}$$

   b. Determine N synthesis.

$$NO_x = \text{TKN} - \text{N}_\text{syn} - (\text{NH}_4\text{-N})_e$$

$$\text{N}_\text{syn} = 0.12(P_x) = 0.12(233.1\ \text{kg/d}) = 28.0\ \text{kg/d}$$

$$\text{N}_\text{syn} = \frac{(28.0\ \text{kg/d})(10^3\ \text{g/1 kg})}{(3785\ \text{m}^3/\text{d})} = 7.4\ \text{g/m}^3$$

      Effluent NH_4-N = 0.5 g/m^3

      NO_x produced $= (35.0 - 7.4 - 0.5)\ \text{g/m}^3 = 27.1\ \text{g/m}^3$

3. Determine amount of nitrate left in SBR mixed liquor after decant.
   Assume: NO_3-N = 0 before aeration period, which means that all of the nitrate remaining in the SBR in the liquid volume after decanting is removed by denitrification during the mixing and unaerated fill period.
   a. Determine NOx produced per cycle.

$$\text{g NO}_x\text{ produced per cycle} = 27.1\text{ g/m}^3\,(1261.7\text{ m}^2\text{/fill}) = 34{,}192\text{ g NO}_x\text{/fill}$$

NO_3-N concentration at end of aeration with tank full: (V = 5047 m^3)

$$= \frac{34{,}192\text{ g}}{5047\text{ m}^3} = 6.8\text{ g/m}^3$$

Effluent NO_3-N = 6.8 g/m^3

based on all NO_3-N that is produced in the aeration period is removed by denitrification during the unaerated fill period.

   b. Determine nitrate remaining in SBR after decant (assuming no denitrification in settling and decant).

Volume remaining after decant: 0.75(5047 m^3) = 3785 m^3

NO_3-N present = 6.8 $g/m^3(V_s)$ = (6.8 g/m^3)(3785 m^3) = 25,740 g

4. Determine $SDNR_b$ in fill period.
   Active biomass concentration (from Step 2aii, biomass = 178.1 kg/d)

$$X_b = \frac{(\text{biomass})(\text{SRT})}{V_T} = \frac{(178.1\text{ kg/d})(20\text{ d})(10^3\text{ g/1 kg})}{5047\text{ m}^3}$$

$$= 705.8\text{ g/m}^3\text{ at full volume}$$

   a. Determine F/M_b ratio in fill period.

Biomass in tank = (705.8 g/m^3)(5047 m^3)(1 kg/10^3 g) = 3562 kg

BOD feed rate = $Q_F S_o$

$$Q_F = \frac{V_F}{t_F} = (1261.7\text{ m}^3\text{/4 h})(24\text{ h/d})$$

$$= 7570\text{ m}^3\text{/d}$$

$$Q_F S_o = (7570\text{ m}^3\text{/d})(200\text{ g/m}^3\text{ BOD})(1\text{kg}/10^3\text{ g}) = 1514\text{ kg/d}$$

$$\text{F/M}_b = \frac{(1514\text{ kg/d})}{3562\text{ kg}} = 0.43\text{ g/g·d}$$

   b. Determine the $SDNR_b$.

Fraction rb COD = (60 g/m^3)/(320 g/m^3) = 0.19

Use Eq. (8–57)

$$\text{SDNR}_b = b_0 + b_1[\ln(\text{F/M}_b)]$$

From Table 8–21; $b_0 = 0.213$ and $b_1 = 0.118$

At 20°C, $\text{SDNR}_b = 0.213 + 0.118\,[\text{L}_n(0.43)] = 0.112 \text{ g/g·d}$

At 16°C, $\text{SDNR}_{16} = 0.112\theta^{16-20};\ \theta = 1.026$

$= 0.102 \text{ g/g·d}$

5. Determine NO_3-N removal capacity during the fill period.

$$\begin{aligned} NO_x &= (\text{SDNR}_b)(X_b)(V_T) \text{ [Note: } (X_b)(V_T) = \text{biomass in system]} \\ &= (0.102 \text{ g/g·d})(705.8 \text{ g/m}^3)(5407 \text{ m}^3) \\ &= 390{,}410 \text{ g/d} \end{aligned}$$

Fill time = 4 h

$$NO_r \text{ at 4 h} = \frac{(390{,}410 \text{ g/d})(4 \text{ h})}{(24 \text{ h/d})} = 65{,}068 \text{ g}$$

NO_3-N available from Step 3 = 25,740 g

Therefore, all of NO_3-N can be removed in the fill period.

Note: V_F/V_T controls effluent NO_3-N concentration.

Postanoxic Denitrification Processes

A number of postanoxic denitrification processes have been introduced including the single, two, and three stage processes. The Bardenpho process is a good example of a postanoxic denitrification application after the aerobic nitrification step. At this point in the biological treatment process there is minimal rbCOD and pbCOD available as electron donors for denitrification. Thus, the electron donor that creates the demand for nitrate reduction is mainly from activated sludge endogenous respiration. Observed SDNRs have ranged from 0.01 to 0.04 g NO_3-N/g MLVSS·d under endogenous respiration (U.S. EPA, 1993; Stensel et al., 1995). Denitrification with external carbon addition is considered later in this section.

Endogenous Respiration Denitrification Rates. The equivalent endogenous oxygen utilization rate under anoxic conditions has been found to be about 50 percent of that under aerobic conditions (Randall et al., 1992; Wuhrman, 1964). Based on the cited references, the SDNR under endogenous conditions (SDNR_b) can be calculated from the endogenous decay rate as follows:

$$NO_r = \left(\frac{1.42}{2.86}\right)(b_{\text{H,anox}})X_{\text{H}} \tag{8–63}$$

$$\text{SDNR}_b = \frac{1.42(b_{\text{H,anox}})}{2.86} = 0.5(b_{\text{H,anox}}) \tag{8–64}$$

where 1.42 = g O_2/g biomass VSS

$b_{\text{H,anox}}$ = biomass endogenous decay coefficient under anoxic conditions, g VSS/g VSS biomass·d

2.86 = g O_2 equivalent/g NO_3-N.

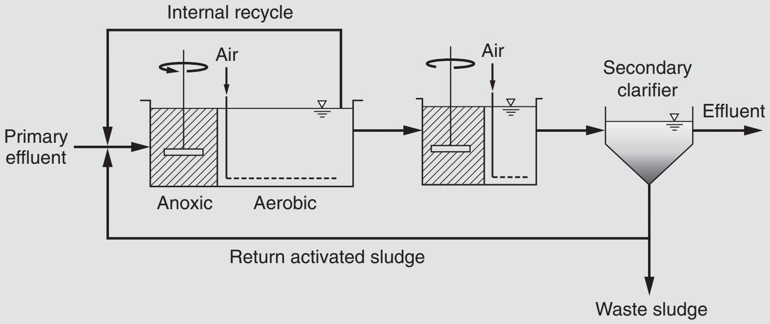
X_{H} = heterotrophic bacteria concentration, g VSS/m^3

NO_r = nitrate reduction rate, g/m^3·d

The specific endogenous decay rate under anoxic conditions is about 60 percent of that under aerobic conditions (Henze et al., 2000). As discussed previously, the $SDNR_b$ is based on the biomass concentration. As shown in Chap. 7, the fraction of biomass from the MLVSS declines as the SRT increases, so the SDNR value based on the MLVSS concentration would decrease with increasing SRT.

Ammonia Production. Ammonia is produced from endogenous decay as a result of deamination of cellular organic nitrogen released during cell lysis. About 50 percent of the biomass nitrogen is converted to ammonia during the relatively short postanoxic tank detention time, or about 0.06 g NH_4-N/g biomass VSS decayed.

EXAMPLE 8–10 **Design of Postanoxic Denitrification Tank** Add a postanoxic tank to further reduce the effluent nitrate concentration to 1.0 g/m^3 after the anoxic/aerobic process steps in Example 8–7 for the activated sludge/clarifier system as shown in the figure below. Determine (a) the postanoxic tank volume for denitrification due to endogenous decay and (b) the change in NH_4-N concentration across the postanoxic tank. Note: This configuration is the Bardenpho process as the anoxic/aerobic/anoxic tanks would be followed by a tank with short aerobic detention time (20–30 min).



Design conditions and assumptions:

1. Information from the anoxic/aerobic system in Example 8–7.

Item	Unit	Value
Flowrate	m^3/d	22,700
RAS ratio	unitless	0.60
Temperature	°C	12
MLSS	g/m^3	3000
MLVSS	g/m^3	2370
Biomass, X_H	g/m^3	1267
Anoxic Endogenous decay, $b_{H,12}$	g/g·d	0.06
Aerobic SRT	d	20.6
Aerobic basin volume	m^3	13,230
Aerobic tank NO_3-N	g/m^3	6.0

Note: g/m^3 = mg/L.

1. NH_4-N used for cell synthesis = 0.12 g NH_4-N/g VSS produced
2. NH_4-N release in endogenous decay = 0.06 g NH_4-N/g VSS
3. DO concentration = 0.0 g/m^3
4. Postanoxic effluent NO_3-N concentration = 1.0 g/m^3

Solution

1. Determine the amount of nitrate to remove in the postanoxic tank using Eq. (8–74). The rate of removal is the difference in NO_3-N concentration across the tank times the flowrate.

$$R_{NO_3} = (Q,\ \text{m}^3/\text{d})(1 + R)(6.0 - 1.0)\ \text{g/m}^3$$

$$R_{NO_3} = 22{,}700\ \text{m}^3/\text{d}\ (1 + 0.60)(6.0 - 1.0)\ \text{g/m}^3 = 181{,}600\ \text{g/d}$$

2. Determine the postanoxic tank volume.
   a. Nitrate reduction from endogenous decay using Eq. (8–63),

$$R_{NO_3} = \left(\frac{1.42}{2.86}\right)(b_{H,12})(X_H)(V_{anox})$$

$$181{,}600\ \text{g/d} = \left(\frac{1.42}{2.86}\right)(0.06\ \text{g/g·d})(1267\ \text{g/m}^3)(V_{anox})$$

$$V_{anox} = 4810\ \text{m}^3$$

$$\tau = (4810\ \text{m}^3)(24\ \text{h/d})/(22{,}700\ \text{m}^3/\text{d}) = 5.09\ \text{h}$$

3. Determine the change in NH_4-N concentration.
   a. Rate of biomass decay in anoxic tank

$$R_{VSS} = b_{H,12}(X_H)(V_{anox}) = (0.06\ \text{g/g·d})(1267\ \text{g/m}^3)(4811\ \text{m}^3)$$

$$= 365{,}732\ \text{g VSS/d}$$

   b. Ammonia production rate

$$R_{NH_4-N} = (0.06\ \text{g}\ NH_4\text{-N/g VSS})(365{,}732\ \text{g VSS/d}) = 21{,}944\ \text{g/d}$$

   c. Change in NH_4-N concentration:

$$\text{Increase} = \frac{R_{NH_4\text{-N}}}{Q(1 + R)} = \frac{(21{,}944\ \text{g}\ NH_4\text{-N/d})}{(22{,}700\ \text{m}^3/\text{d})(1 + 0.6)} = 0.60\ \text{g/m}^3$$

4. What is the observed SDNR?

$$\text{SDNR} = \frac{R_{NO_3}}{(X_{VSS})V_{anox}} = \frac{(181{,}600\ NO_3\text{-N/d})}{(2370\ \text{g/m}^3)(4811\ \text{m}^3)} = 0.016\ \text{g}\ NO_3\text{-N/g VSS·d}$$

Low DO and Cyclic Nitrification/Denitrification Processes

Low DO SNdN and cyclic NdN nitrogen removal processes are typically done in single reactor systems or single reactor compartmentalized systems with relatively long τ and SRT conditions. Thus, for design estimates of denitrification rates, the reactor can be treated as a completely mixed reactor with continuous addition of influent flow and bCOD. In contrast to the preanoxic denitrification zone, the reactor bCOD concentration and SDNR are relatively low due to the larger reactor volume. In comparison to postanoxic denitrification, the SDNR is higher.

The effect of low DO concentration on nitrification rates as well as denitrification rates must be considered. The activated sludge floc will be only partially aerobic, and denitrification occurs in the anoxic zones established within the floc particles due to oxygen depletion, with the result that simultaneous nitrification and denitrification takes place. The nitrification and denitrification rates are a function of the reaction kinetics, floc size, floc density, floc structure, rbCOD loading, and bulk liquid DO concentration. It is possible that NH_4-N is oxidized mainly to NO_2-N, with NO_2-N reduction within the anoxic portion of the floc. Because of the complex physical factors, the nitrification and denitrification rates cannot be predicted accurately with present models.

Effects of Low DO. Basic modifications to the Monod growth model, however, can be used to estimate and illustrate the effects of a low DO concentration on nitrification and denitrification rates and system performance. The effect of DO concentration on nitrification rate is described by Eq. (7–101) given in Table 8–10. The kinetic coefficient values can be very site specific as a function of the simultaneous NdN system operating conditions and ammonia-oxidizing and heterotrophic microorganism population selection. The effect of DO concentration on the design SRT is illustrated on Fig. 8–34 for a completely-mixed reactor effluent NH_4-N concentration of 1.0 mg/L at 20°C using the kinetic coefficient values for nitrification given in Table 8–14. No safety factor is used for these calculations. The actual SRT value used for design would be higher depending on the safety factor selected for a specific system. The nitrification rate at a DO concentration of 0.2 mg/L is 24 percent of the rate at 2.0 mg/L, based on the calculated SRT values of 19.9 and 4.7 d, respectively. SNdN and cyclic NdN systems are normally operated at SRT values of 20 to 30 d, so the ability to produce complete nitrification at low DO is apparent.

Nitrate Reduction Rate. The nitrate reduction rate can be related to the rate of substrate utilization as given by Eq. (7–133) in Table 8–10. In addition, nitrate reduction is also driven by substrate released by endogenous decay as shown by Eq. (8–63).

$$r_{NO_x} = NO_r = \left(\frac{1.42}{2.86}\right) b_H (X_H) \tag{8–63}$$

Figure 8–34

Effect of DO concentration on SRT required to achieve an effluent NH_4-N concentration of 1.0 mg/L at 20°C in a CMAS system based on kinetic coefficients in Table 8–14.

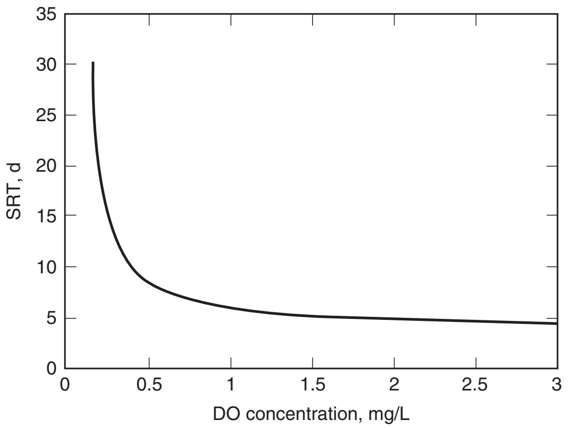
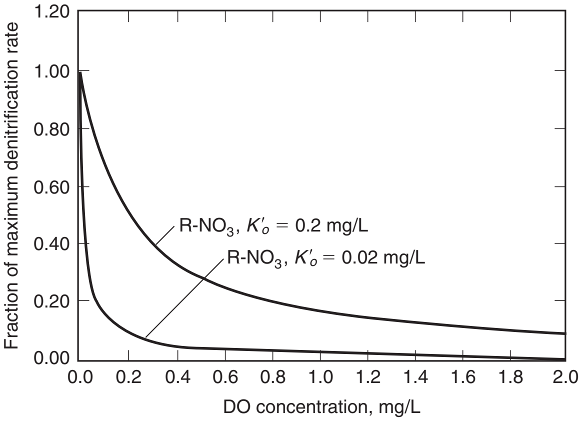


Figure 8–35

Effect of mixed liquor DO concentration on maximum denitrification rates.



Combining Eqs. (7–133) and (8–63) yields a general rate expression [(Eq. (8–65)] that can be used to describe the nitrate reduction rate in an anoxic reactor as function of bCOD, heterotrophic biomass, NO_3-N, and DO concentrations as well as various biokinetic coefficients. For NO_2-N reduction instead of NO_3-N reduction, due to ammonia oxidation to only nitrite, the 2.86 value in Eq. (8–65) is replaced by 1.71, the oxygen equivalent of NO_2-N.

$$r_{\mathrm{NO_3}} = \left(\frac{1 - 1.42Y_{\mathrm{H}}}{2.86}\right)\left[\frac{\mu_{\mathrm{H,max}} S_S}{Y_{\mathrm{H}}(K_s + S_S)}\right]\left(\frac{S_{\mathrm{NO_3}}}{K_{\mathrm{NO_3}} + S_{\mathrm{NO_3}}}\right)\left(\frac{K'_o}{K'_o + S_o}\right)(\eta)X_{\mathrm{H}} + \left(\frac{1.42}{2.86}\right)b_{\mathrm{H}}(X_{\mathrm{H}}) \tag{8–65}$$

The DO inhibition coefficient K'_o is difficult to estimate and will be site-specific, depending on the floc size and structure. The effect of DO concentration on the denitrification rate is shown on Fig. 8–35 for K'_o values of 0.02 and 0.2 mg/L. At a DO concentration of 0.2 mg/L, the denitrification rate may be 10 to 50 percent of its maximum rate. For long SRT systems with τ values in the range of 18 to 30 h, sufficient time may be available for high nitrate removal efficiency, even though the rate is somewhat inhibited by having DO present at low concentrations.

Completely-Mixed Tank SDNR. Many SNdN and cyclic NdN denitrification systems are single tank reactors with long τ and SRT values. Because of the long SRT and τ, the denitrification kinetics are related to the overall degradation of bCOD, pbCOD, and endogenous decay and are not as strongly influenced by the rbCOD fraction as for the preanoxic denitrification application with the relatively short τ. During the complete-mix anoxic period in an oxidation ditch operation or in the anoxic period of a continuously-mixed oxidation ditch, the specific denitrification rate is affected by both the endogenous respiration rate and the bCOD in the influent wastewater as continuous feeding occurs. The average specific denitrification rate, which includes these effects, can be estimated using Eq. (8–66) [based on Stensel (1981) and modified to account for the rate as a function of active heterotrophic biomass]:

$$\mathrm{SDNR_b} = \frac{0.175A_n}{Y_{\mathrm{net}}(\mathrm{SRT})} \tag{8–66}$$

where $\mathrm{SDNR_b}$ = specific denitrification rate relative to heterotrophic biomass concentration, g NO_3-N/g biomass·d

A_n = net oxygen utilization coefficient, g O_2/g bCOD removed

Y_{net} = net yield for heterotrophic biomass, g VSS/g bCOD

0.175 = based on 2.86 g O_2 equivalent/g NO_3-N and the assumption that only 50 percent of heterotrophic biomass can use nitrate in place of oxygen

Note that Eq. (8–66) can be adjusted with information from Fig. 8–33 to account for a reduced SDNR for operation with low residual DO concentration. For a complete-mix activated sludge reactor, A_n and Y_{net} are determined as follows (Stensel, 1981):

$$A_n = 1.0 - 1.42Y_{\text{H}} + \frac{1.42(b_{\text{H}})(Y_{\text{H}})\text{SRT}}{1 + b_{\text{H}}(\text{SRT})} \tag{8–67}$$

$$Y_{\text{net}} = \frac{Y_{\text{H}}}{1 + b_{\text{H}}(\text{SRT})} \tag{8–68}$$

The SDNR_{b} in Eq. (8–66) is relative to the biomass concentration that is a portion of the MLVSS concentration. Thus, the design procedure used to determine the amount of nitrate removed in the anoxic period during cyclic aerobic/anoxic operation incorporates some of the elements of the computational steps for anoxic/aerobic design described in Table 8–21. In Step 3 in Table 8–21, the heterotrophic biomass concentration has to be determined. As detailed in Step 9, the biomass concentration, SDNR_{b}, anoxic volume, and anoxic time are used to determine the amount of nitrate nitrogen removed. Use of the above equations is illustrated in Example 8–11.

EXAMPLE 8–11 **Determine Fraction of Anoxic Time Needed in Oxidation Ditch Cyclic NdN Nitrogen Removal System** Determine what fraction of the time an oxidation ditch system must be operated as an anoxic reactor for an intermittent aeration process to produce an effluent NO_3-N concentration of 7 g/m^3, based on the following design conditions.

Design conditions:

Item	Unit	Value
Oxidation ditch volume	m^3	8700
SRT	d	25
MLSS	g/m^3	3500
MLVSS	g/m^3	2500
Fraction of biomass	g biomass/g MLVSS	0.40
Temperature	°C	15
Y_{H}, b_{H}		Values given in Table 8–14
Influent flowrate	m^3/d	7570
NO_x produced in oxidation ditch, based on influent flowrate	g/m^3	27

Note: g/m^3 = mg/L.

Solution

1. Determine the SDNR.
   a. Obtain Y_H and b_H from Table 8–14 and correct b for temperature using Eq. (2–25).

$$Y_H = 0.45 \text{ g VSS/g bCOD}$$

$$b_{H,20} = 0.12 \text{ g/g·d}$$

$$b_{H,15} = b_{H,20}(1.04)^{15-20}$$

$$= 0.12(1.04)^{15-20} = 0.099 \text{ g/g·d}$$

   b. Using Eq. (8–67), determine A_n.

$$A_n = 1.0 - 1.42Y_H + \frac{1.42(b_H)(Y_H)\text{SRT}}{1 + b_H(\text{SRT})}$$

$$A_n = 1.0 - 1.42(0.45) + \frac{1.42(0.099 \text{ g/g·d})(0.45 \text{ g/g})(25 \text{ d})}{1 + (0.099 \text{ g/g·d})(25 \text{ d})}$$

$$= 0.82 \text{ g } O_2/\text{g bCOD}$$

   c. Using Eq. (8–68) determine Y_{net}.

$$Y_{net} = \frac{Y_H}{1 + b_H(\text{SRT})}$$

$$= \frac{0.45}{1 + (0.099 \text{ g/g·d})(25 \text{ d})} = 0.13 \text{ g VSS/g bCOD}$$

   d. Using Eq. (8–66), determine SDNR_b.

$$\text{SDNR}_b = \frac{0.175A_n}{(Y_{net})\text{SRT}}$$

$$\text{SDNR}_b = \frac{0.175(0.82 \text{ g } O_2/\text{g bCOD})}{(0.13 \text{ g VSS/g bCOD})(25 \text{ d})} = \frac{0.044 \text{ g } NO_3\text{-N}}{\text{g biomass·d}}$$

2. Determine biomass concentration in the mixed liquor.

$$X_b = (2500 \text{ g/m}^3 \text{ MLVSS})\left(\frac{0.40 \text{ g biomass}}{\text{g MLVSS}}\right)$$

$$= 1000 \text{ g/m}^3 \text{ biomass}$$

3. Determine amount of NO_3-N removal needed in g/d.

Nitrate removed concentration (NO_r) = (27.0 – 7.0) g/m^3 = 20 g/m^3

$$NO_r = (7570 \text{ m}^3\text{/d})(20 \text{ g/m}^3)$$

$$= 151{,}400 \text{ g/d}$$

4. Determine NO_3-N removal rate under anoxic reaction time.

$$\text{Anoxic } NO_r = (\text{SDNR}_b)(X_b)(V)$$

$$(0.044 \text{ g } NO_3\text{-N/g biomass·d})(1000 \text{ g/m}^3)(8700 \text{ m}^3)$$

$$= 382{,}800 \text{ g/d}$$

5. Determine anoxic time needed per day.

$$\text{Anoxic time} = \frac{(151{,}400 \text{ g/d})(24 \text{ h/d})}{(382{,}800 \text{ g/d})} = 9.5 \text{ h}$$

$$\text{Fraction of day} = 9.5 \text{ h}/24 \text{ h} = 0.40$$

Comment The computed value is within the range of values reported for the anoxic fraction for cyclic aeration processes as given in U.S. EPA (1993). The actual time may be less or the amount of nitrogen removal may be greater due to denitrification within the floc during low DO aeration periods and in the secondary clarifier. By knowing the approximate amount of time needed for the anoxic conditions, the aerobic SRT available for nitrification can be estimated.

Alternative Process Configurations for Biological Nitrogen Removal

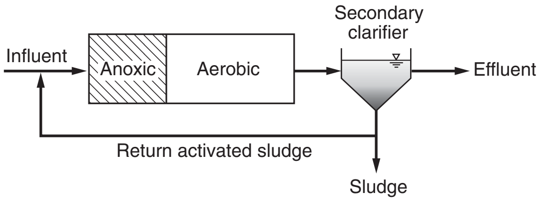
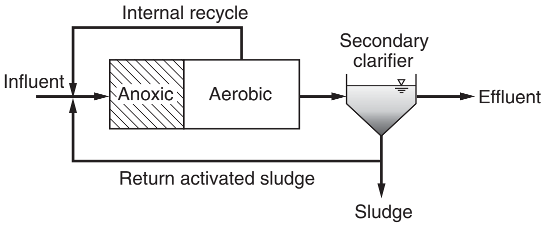
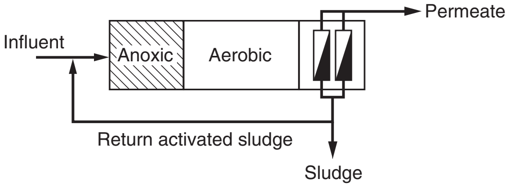
A variety of activated sludge process configurations are used to accomplish biological nitrogen removal. Representative process schematics, descriptions, and nitrogen removal capability for biological nitrogen removal processes are given in Table 8–24 and are discussed below. The advantages and disadvantages for the various processes are also discussed.

MLE Process. The MLE process is one of the most common methods used for biological nitrogen removal, and can be adapted easily to existing activated sludge facilities. The amount of nitrate removal is limited by the practical levels of internal recycle to the preanoxic zone, and the process is used more generally to achieve effluent total nitrogen concentrations between 6 and 10 mg/L. About 2–4 mg/L of the effluent total nitrogen concentration is associated with NH_4-N and dissolved and particulate organic nitrogen. With an internal recycle ratio of 4.0, the MLE process can remove about 80 percent of the NO_x produced in the nitrification process. As more water conservation is implemented and influent TKN concentrations increase, it will be more difficult to achieve an effluent TN concentration of <10 mg/L. Dissolved oxygen control should be used in the zone from which the recycle stream is taken to limit the amount of DO fed to the anoxic zone. The MLE membrane process has a high recycle flow to the preanoxic zone and if there is sufficient BOD available for DO and NO_3-N consumption, effluent TN concentrations less than 6.0 mg/L are possible.

Step Feed Process. The step feed process is also applicable for meeting effluent total nitrogen concentrations of less than 10 mg/L. However, it is theoretically possible to achieve lower effluent nitrogen total concentrations of less than 5.0 mg/L with step feed BNR. The DO concentration in the upstream aerobic zone before the step feed BNR anoxic zone must be controlled to minimize dissolved oxygen addition to the anoxic zone and consumption of rbCOD needed for NO_3-N removal. Multiple DO control points are required to optimize nitrogen removal in step feed BNR process. Influent flow splitting control and measurement are necessary to optimize the step feed reactor volume for nitrogen removal.

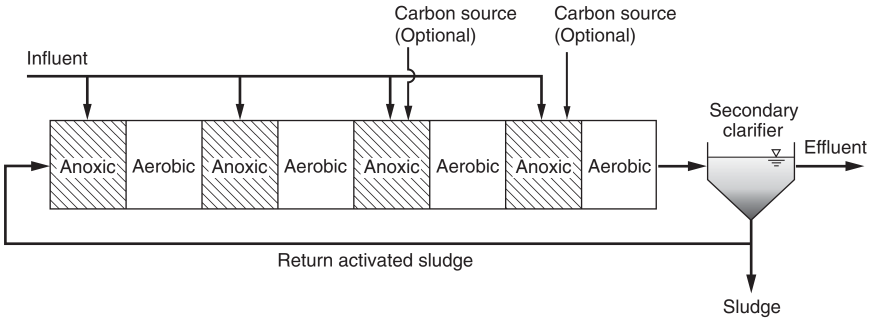
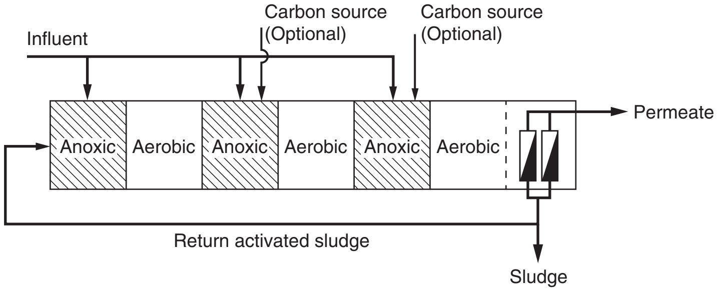
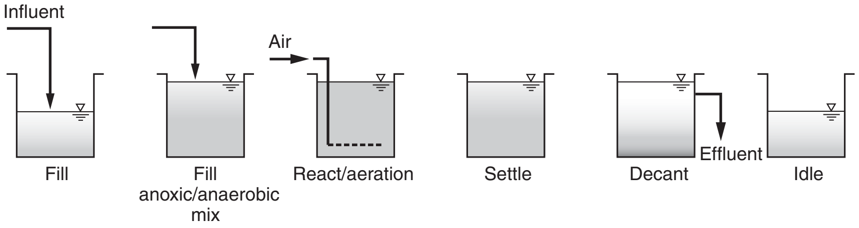
Table 8–24

Description of suspended growth processes for nitrogen removal

Process	**Description**
Preanoxic Processes	
(a) Ludzak-Ettinger 	The first concept of a preanoxic BNR was an anoxic-aerobic operating sequence by Ludzak and Ettinger (1962). The influent wastewater was fed to an anoxic zone, which was followed by an aerobic zone. The process relies on the nitrate formed in the aerobic zone being returned via the RAS to the anoxic zone. Because the only nitrate fed to the anoxic zone is that in the RAS, denitrification is limited greatly by the RAS recycle ratio. However, more recently, this process has been used with increased RAS recycle rates to prevent rising sludge in the secondary clarifiers due to denitrification.
(b) Modified Ludzak-Ettinger (MLE) 	One of the most commonly used BNR processes is the Modified Ludzak-Ettinger (MLE) process. Barnard (1973) improved on the original Ludzak-Ettinger design by providing the internal recycle to feed more nitrate to the anoxic zone directly from the aerobic zone. Both the denitrification rate and overall nitrogen removal efficiency are increased. The internal recycle flow ratio (recycle flowrate divided by influent flowrate) typically ranges from 2 to 4. With sufficient influent BOD and anoxic contact time, these recycle ratios result in an average effluent NO_3-N concentrations from 4 to 7 mg/L when treating domestic wastewater. The MLE process is very adaptable to existing activated sludge facilities and can easily meet a common effluent standard of less than 10 mg/L total nitrogen. A BOD/TKN ratio of 4:1 in the influent wastewater is usually sufficient for effective nitrate reduction by preanoxic processes. Typical anoxic tank detention times for the MLE process range from 2 to 4 h, but when the anoxic zone is divided into 3 to 4 stages in series, denitrification kinetic rates are increased and the total detention time needed may then be 50 to 70 percent of the single-tank design.
(c) MLE- Membrane Bioreactor 	Preanoxic designs are commonly used in MBR systems, which are typically operated at SRTs that result in full nitrification. The return activated sludge recycle ratio (typically 6.0) is significantly higher than that used for the activated sludge/secondary clarifier systems with the MLE process. The return activated sludge is taken directly from the aerated membrane separation tank, and thus some of the influent wastewater rbCOD will be consumed by bacteria using dissolved oxygen as an electron acceptor with less rbCOD then available for denitrification.

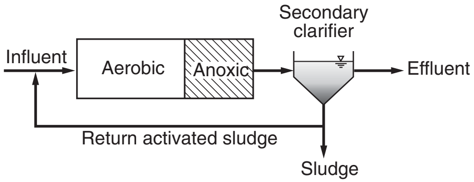
(*continued*)

Table 8–24 *(Continued)*

Process	**Description**
(d) Step feed biological nitrogen removal  	Preanoxic zones can also be used in a step feed BNR process. Because step feed BNR processes are usually adapted to existing multiple-pass full-scale tanks, symmetrical anoxic/aerobic stages are generally used. However, non-symmetrical designs with smaller initial anoxic/aerobic stages can take better advantage of the higher MLSS concentration in the early stages, due to less RAS dilution, resulting in greater treatment capacity. A possible influent flow splitting percent distribution for a 4-pass system is 15/35/30/20. The final flow portion to the last anoxic/aerobic zone is critical as the nitrate produced in the aerobic zone from that flow will not be reduced, and will, thus, determine the final effluent NO_3-N concentration. Effluent NO_3-N concentrations of less than 5 mg/L are possible. Depending on the influent TKN concentration, carbon addition, such as methanol or glycerol, may be added to reduce the anoxic zone size and to compensate for a weak influent BOD concentration.
(e) Step feed MBR  	A step feed preanoxic design can also be used with a MBR process to give a lower effluent NO_3-N concentration than for the MLE-MBR system. Effluent NO_3-N concentrations of less than 3.0 mg/L are possible. An exogenous carbon source can also be added to the final anoxic zones.
(f) Sequencing batch reactor (SBR)  	The SBR system [see Table 8–18(l)] also employs preanoxic denitrification using BOD in the influent wastewater. Mixing is used during the fill period to contact the mixed liquor with the influent wastewater. For many domestic applications, depending on the wastewater strength, sufficient BOD and fill time are available to remove almost all of the nitrate remaining in the mixed liquor after the settle and decant steps. Some nitrate removal also occurs during the non-aerated settle and decant periods. Separate mixing provides operating flexibility and is useful for anoxic operation during the aeration period, as well as anaerobic or anoxic contacting during the fill period. Mixing without aeration during the fill period is effective in improving sludge settling properties in addition to nitrogen removal. Effluent NO_3-N concentrations of less than 5 mg/L are possible.

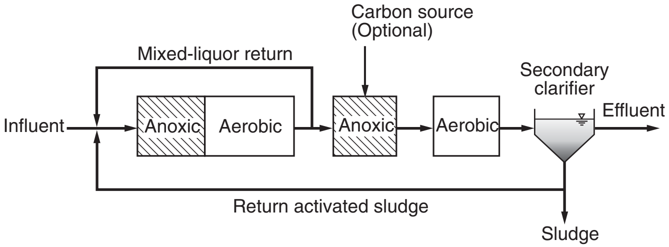
Postanoxic Processes

(g) Single-sludge



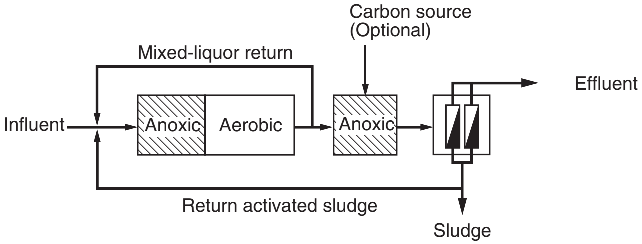
In the single-sludge process (developed by Wuhrmann, 1964), nitrogen removal was accomplished in the activated sludge process by adding a mixed anoxic tank after aerobic nitrification. To achieve high nitrate removal efficiency, a carbon source would have to be added. NH_4-N is released during endogenous respiration in postanoxic zones, which adds to the effluent total nitrogen concentration.

(h) Bardenpho (4-stage)



Both preanoxic and postanoxic denitrification are incorporated in the Bardenpho process, which was developed and applied at full-scale facilities in South Africa in the mid-1970s, before making its way to the United States in1978. The detention time of the postanoxic stage is about the same or larger than that used for the preanoxic zone. In the postanoxic zone, the NO_3-N concentration leaving the aeration zone is typically reduced from about 5 to 7 mg/L to less than 3 mg/L. During pilot plant testing with higher strength wastewaters, Barnard (1974) found that biological phosphorus removal occurred as well as nitrogen removal, hence the basis for the process name (the name comes from the first three letters of the inventor's name, Barnard, and from denitrification, and phosphorus). Carbon can be added to the postanoxic zone to provide lower effluent NO_3-N concentrations and to reduce the postanoxic tank volume. Effluent NO_3-N concentrations of less than 1.0 or 2.0 mg/L are possible.

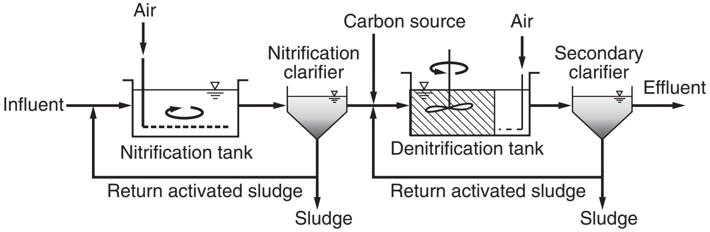
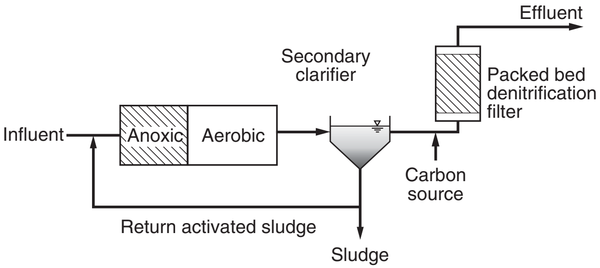
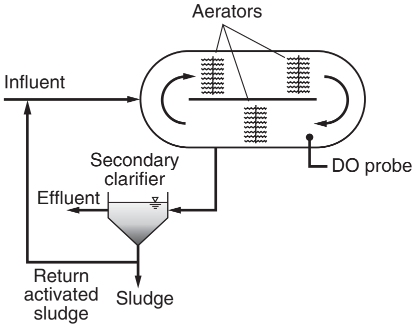
(i) Bardenpho (4-stage) MBR



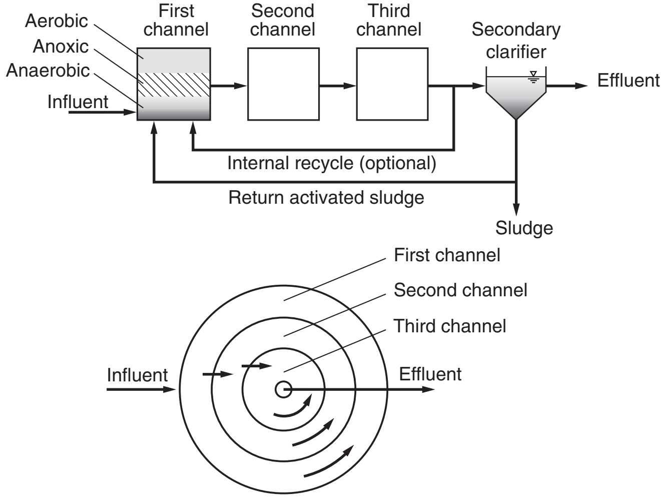
A 4-stage Bardenpho process can be incorporated in MBR systems. As with the activated sludge/secondary clarifier system, carbon addition is optional in the postanoxic zone, but is needed when a very low effluent total nitrogen concentration is required. Effluent NO_3-N concentrations of less than 1 or 2 mg/L are possible.

(*continued*)

Table 8–24 (*Continued*)

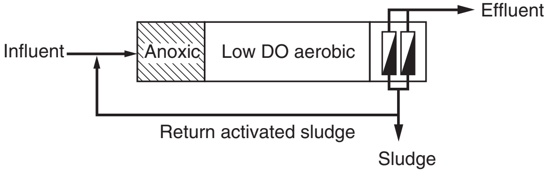
Process	Description
(j) Dual-sludge system with an external carbon source  	An approach that was most popular in the 1970s is a postanoxic design with exogenous external carbon addition, usually methanol. The anoxic zone (1 to 3 h) is mixed and a short aeration time ($<$ 30 min) follows to strip nitrogen gas bubbles from the floc and to provide aerobic conditions to improve liquid-solids separation in the clarifier.
(k) MLE-Packed bed postanoxic  	A more common approach today for postanoxic treatment is to use denitrification filters with carbon addition after the nitrification process (see Chap. 9 for postanoxic packed bed designs). An effluent NO_3-N concentration of less than 1.0 mg/L is possible.
Simultaneous nitrification/denitrification	
(l) Low DO oxidation ditch  	Oxidation ditches that have sufficient volume available are able to accommodate both nitrification and denitrification at lower rates under low DO conditions. An oxidation ditch may be used to maintain DO concentrations below 0.5 mg/L with manual or automated DO control. Where multiple aerators are used, for example with brush aerators, a low to zero DO is maintained throughout the channel. The use of variable frequency drives on the aeration equipment and the improved accuracy and reliability of low DO concentration measurements by luminescence DO probes facilitates the application of low DO control methods.

(m) Orbal™



The channels in the Orbal process are operated in series with a zero to low DO (<0.3 mg/L) concentration in the first channel, a 0.5 to 1.5 mg/L DO concentration in the second channel, and a higher DO concentration (2 to 3 mg/L) in the third channel. The first channel receives the influent wastewater and return activated sludge and generally contains about one-half the total basin volume. Volumes of the second and third channels are about one-third and one-sixth of the total basin volumes, respectively. Recycle of mixed liquor from the inner loop to the outer loop allows denitrification of nitrates resulting from nitrification in the inner channels. Variations of the process include operation with and without internal recycle flow (Bionutre™ process) from the third channel to the first channel. An oxygen supply rate of about 50 percent of the estimated design requirement has been recommended to support SNdN in the first zone.

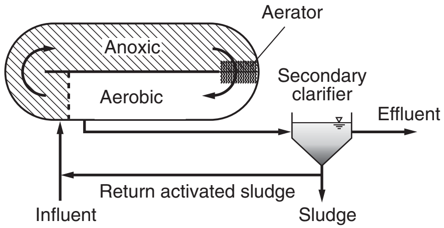
(n) Low DO MBR



MBR systems have been operated with low DO in the aerobic zone prior to the aerobic membrane separation zone. The high mixed liquor concentration results in a higher volumetric oxygen uptake, which helps to assure that anoxic zones exist within the floc at bulk liquid DO concentrations of 0.30 to 0.70 mg/L.

Cyclic NdN Processes

(o) Oxidation ditch



Depending on the aeration design and length of the oxidation ditch channel, anoxic denitrification zones can be established in oxidation ditches to accomplish biological nitrogen removal in a single tank. An aerobic zone exists after the aerator, and as the mixed liquor flows down the channel away from the aerator, the DO concentration decreases due to oxygen uptake by the biomass. At a point where the DO is depleted, an anoxic zone is created in the ditch channel and the nitrate will be used for endogenous respiration activity by the mixed liquor. Most of the readily degradable BOD had been consumed previously in the aerobic zone. Because of the large tank volumes and long SRTs used in oxidation ditch processes, sufficient capacity is available to accommodate nitrification and denitrification zones. DO control is necessary, however, to maintain a sufficient anoxic zone volume to allow for significant nitrogen removal.

(*continued*)

Table 8–24 (*Continued*)

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<tr><th>Process</th><th>Description</th></tr>
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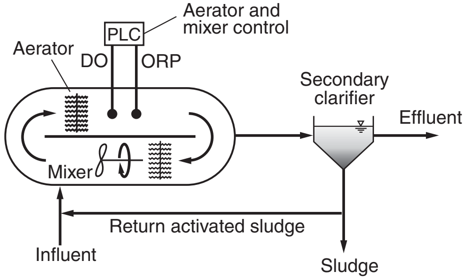
(p) dNOx™



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In the dNOx™ process, the oxidation ditch operation is switched from an aerobic to an anoxic operating condition by turning off the aeration and operating a submerged mixer to maintain channel velocity. The process depends on the use of oxidation-reduction potential (ORP) control to (1) determine when the nitrate is depleted during the anoxic operation and (2) restart aeration. At selected times, the aerators are turned off and the mixer is turned on. When the nitrate is depleted in the aeration OFF period, the ORP drops dramatically. The ORP data are interpreted by a PC, which starts the aeration. A typical operating condition for the dNOx™ process is to turn the aerators off at least twice per day, usually in the morning when the load is increasing and then in the early evening hours (Stensel and Coleman, 2000). The off-time for the nitrate depletion usually lasts 3 to 5 h depending on the plant load and amount of nitrate in the oxidation ditch. Effluent NO_3-N concentrations of less than 8 mg/L and NH_4-N concentrations ranging from 1.0 to 1.5 mg/L have been reported.

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Phased NdN Processes

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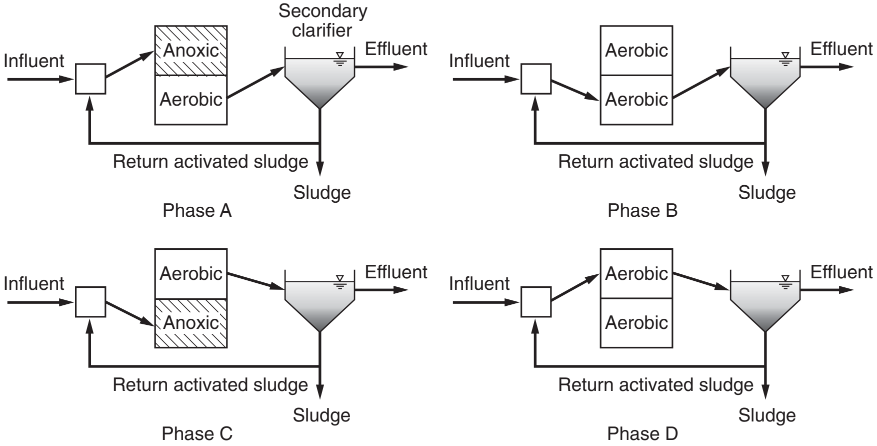
(q) BioDenitro



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The BioDenitro process has also been referred to as a phased-isolation oxidation ditch technology. The process was developed in Denmark for nitrogen removal and has been installed in over 75 full-scale facilities producing effluent total nitrogen concentrations of less than 8 mg/L (Stensel and Coleman, 2000). The technology uses at least two oxidation ditches in a series configuration in which the operating sequence of the ditches and operation of the aeration and anoxic zones is varied. Submerged mixers are installed in the ditches so that for some operating phases, the basin is only mixed and not aerated. The basin continues to receive influent wastewater and operates as a preanoxic zone. Similar to the SBR operation, nitrate is available from a previous aerobic nitrification operation. Besides denitrification in the preanoxic zones, nitrate reduction is also possible during the aerobic operation depending on the DO concentration level. A typical duration for phases A, B, C, and D are 1.5, 0.5, 1.5, and 0.5 h, respectively.

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Sequencing Batch Reactor Process. The sequencing batch reactor process provides a high degree of flexibility for nitrogen removal. Mixing during the fill period provides an opportunity for anoxic conditions for nitrate removal. During the aeration react period, the DO concentration may be cycled to provide anoxic operating periods. A modification of the conventional SBR process involves the formation of a dense granular activated sludge. This process is described in Table 8–24(n) for biological nitrogen and enhanced biological phosphorus removal.

Large Reactor Volume Process. BioDenitro™, dNO_x™, and the oxidation ditch with DO control are all processes with large reactor volumes for nitrogen removal, and represent various methods for optimizing biological nitrogen removal in oxidation ditch systems. Very low effluent total nitrogen concentrations (less than 5 mg/L) have been reported for the Bio-dNO_x ™ process. The dNO_x™ process is generally limited to effluent total nitrogen concentrations of 5 to 8 mg/L. During the aeration "off period" in the dNO_x™ process, ammonia accumulates in the oxidation ditch, resulting in higher effluent NH_4-N concentrations from the process. The effluent NH_4-N and total nitrogen concentration is dependent on the total reactor volume and influent nitrogen concentrations. Higher influent TKN concentrations can result in higher effluent ammonia concentrations.

Bardenpho Process. In full-scale applications, the Bardenpho process and other postanoxic processes with carbon addition have demonstrated the ability to achieve less than 3 mg/L total nitrogen. The second anoxic zone of the Bardenpho process has a very low denitrification rate, resulting in less efficient reactor volume utilization. The addition of an exogenous carbon to the second anoxic zone reduces the reactor volume requirements and results in less NH_4-N release that would contribute to the effluent total nitrogen concentration.

Nitrogen Removal in Anaerobic Digestion Recycle Streams.
Recycle flows from dewatering of anaerobically digested solids contain high NH_4-N concentrations (>1000 mg/L) that can increase the wastewater influent nitrogen load by 20 to 25 percent. The recycle steam is also characterized by relatively high temperature and pH. A number of sidestream treatment processes have been developed that can reduce the ammonia load to the main stream biological treatment system, provide nitrifier seed to the mainstream biological treatment system and reduce the amount of exogenous carbon needed for nitrogen removal. These include the SHARON® (single-reactor high-activity ammonia removal over nitrite) process, anammox (anaerobic ammonia oxidation) and sidestream nitrification with bioaugmentation. These processes and their designs are presented in detail in Chap. 15.

Advantages and Disadvantages of Various Processes. The advantages and limitations of the processes commonly used and their treatment capability in terms of effluent total nitrogen concentrations are summarized in Table 8–25. Advantages and disadvantages of the processes identified in Table 8–25 are considered further in the following discussion.

Process Categories. The nitrogen removal process designs can be divided into two categories; (a) processes in which the aerobic SRT and volume is based on meeting an effluent NH_4-N goal and (b) processes with long SRT values with excess aerobic nitrification capacity and a high degree of flexibility for operation with intermittent aeration or low DO concentration. The former includes the MLE, MLE-Membrane, Bardenpho, Bardenpho-MBR, step feed BNR, and dual sludge processes. The design temperature and effluent NH_4-N goal will affect the SRT needed for nitrification. The latter includes oxidation ditch, SBR and Orbal® processes. The internal recycle ratio to preanoxic zones ranges from 2 to 4 for MLE and Bardenpho processes. A higher internal recycle ratio of 6.0 is used for MLE-MBR

Table 8–25

Advantages and limitations of nitrogen removal processes

Process	Advantages	Limitations
MLE	Saves energy; BOD is removed before aerobic zone Alkalinity is produced before nitrification Higher MLSS due to selection against filamentous bacteria Very adaptable to existing activated sludge processes <10 mg/L TN is achievable	Effluent TN concentration Nitrogen removal capability is a function of internal recycle Potential *Nocardioform* foam problems DO control is required for internal recycle
MLE-Membrane	Similar to MLE in using recycled NO_3 for BOD removal, and alkalinity Smaller footprint due to high MLSS and no clarifier <6 mg/L TN is achievable High quality effluent due to complete TSS removal. Can meet Class A reclaimed water.	Nitrogen removal capability is a function of internal recycle Potential *Nocardioform* foam problems DO control is required for recycle Uses more energy than MLE Membrane fouling control needed
Step Feed BNR	Adaptable to existing plug flow activated sludge processes	Nitrogen removal capability is a function of flow distribution
	<5 mg/L TN is achievable	More complex operation than MLE; requires flow split control to optimize operation
	No internal recycle piping and pumping needed	Potential *Nocardioform* foam problems
	Has higher treatment capacity for tank volume than MLE	Requires DO control in each aeration zone
Sequencing batch reactor	Process has high flexibility	Suitable mostly for smaller flows
	Simple treatment system layout	Needs more volume for nitrogen removal than only BOD removal
	Mixed liquor solids cannot be washed out by hydraulic surges because flow equalization is provided	Redundant units are required for operational reliability
	Quiescent settling provides low effluent TSS concentration	More complex process design
	5 to 8 mg/L TN is achievable	Effluent quality depends upon reliable decanting facility May need effluent equalization of batch discharge before filtration and disinfection
BioDenitro	5 to 8 mg/L TN is achievable	Complex system to operate
	Large reactor volume is resistant to shock loads	Two oxidation ditch reactors are required; increases construction cost

(*continued*)

Table 8–25 (*Continued*)

Process	Advantages	Limitations
dNO_x™	Large reactor volume is resistant to shock loads	Nitrogen removal capability is limited by higher influent TKN concentrations
	Easy and economical to upgrade existing oxidation ditch processes	Process is susceptible to ammonia bleed-through
	Provides SVI control	Performance is affected by influent variations
	Provides energy savings	
Bardenpho (4 stage)	Capable of achieving effluent nitrogen levels less than 3 mg/L	Large reactor volumes required
	Same advantages as MLE	Second anoxic tank has low nitrogen removal efficiency without exogenous carbon addition
Oxidation ditch	Large reactor volume is resistant to load variations without affecting effluent quality significantly	Nitrogen removal capability is related to skills of operating staff and control methods if no preanoxic
	Simple plant layout and operation	
	Has good capacity for nitrogen removal; less than 5 mg/L effluent TN is possible	Large area required
	Can be effective with preanoxic zones	
Postanoxic with carbon addition	Capable of achieving effluent nitrogen levels less than 3 mg/L	Higher operating cost due to purchase of carbon source
	May be combined with effluent filtration	Carbon source feed control required
Simultaneous nitrification/denitrification	Low effluent nitrogen level possible (3 mg/L lower limit)	Large reactor volume; skilled operation also required
	Significant energy savings possible	Process control system required
	Process may be incorporated into existing facilities without new construction	
	Produces alkalinity	

applications. The MLSS concentration for activated sludge/clarifier systems with preanoxic zones range from 3000 to 4000 mg/L and for MBR systems it may range from 8000 to 14,000 mg/L. Low DO or intermittent aeration systems to promote simultaneous nitrification/denitrification can be applied to oxidation ditch, Orbal®, SBR and MLE, and MLE-Membrane processes operated at longer SRTs than in their normal application.

Advantages of Preanoxic and SNdN Over Postanoxic Processes. Preanoxic processes are normally used with and without postanoxic processes. The selection of the postanoxic suspended growth process is driven mainly by the site layout, existing reactor configuration, and equipment considerations. A major disadvantage of a postanoxic only process is the cost of providing an exogenous carbon source. The preanoxic and SNdN processes have additional important advantages over the use of only postanoxic processes. By removing nitrate before or during the nitrification step, the alkalinity produced by denitrification is made available to offset the alkalinity depleted by nitrification. Because 3.57 g of alkalinity (as $CaCO_3$) are produced per g NO_3-N oxidized, and 7.14 g alkalinity (as $CaCO_3$) are consumed per g NH_4-N oxidized, almost half of the alkalinity used for nitrification can be provided by preanoxic or SNdN processes. The recovery of alkalinity is very important for wastewaters that have low alkalinity. In some applications, alkalinity may have to be added in the form of lime or sodium hydroxide, at significant cost, to maintain an acceptable pH for the nitrification process. In addition, the preanoxic process serves a selector against filamentous bacteria growth to provide for good settling activated sludge.

Denitrification with External Carbon Addition

The use of external carbon sources for denitrification is necessary for biological nitrogen removal facilities faced with stringent effluent nitrogen limits and/or that have weak wastewaters relative to the influent BOD to TKN ratio. If the effluent total nitrogen concentration goal is less than 6 to 8 mg/L, an external carbon source is generally required. External carbon addition may be to a postanoxic zone of a Bardenpho process, the latter anoxic tanks in a step feed BNR process, an anoxic operating period towards the end of an SBR process react cycle, and an effluent denitrification filter. Seasonal wet weather operating conditions may also call for the addition of an external carbon source to improve denitrification rates with weaker wastewater. Advantages of adding external carbon to a post anoxic zone in a Bardenpho process are the ability to use a smaller anoxic tank volume and to achieve a lower effluent TN concentration due to less NH_4-N release from endogenous decay.

Sources of External Carbon. External carbon sources are primarily from industrial production, byproducts of industrial processing, and in plant fermentation of waste primary sludge. Common organic compounds that have been used include methanol, ethanol, sugar, high fructose corn syrup, acetate, glycerol, corn starch, molasses, distillery fusel oil, and commercial products such as Unicarb and MicroC^{TM} and other industrial wastes (Gu and Onnis-Hayden, 2010 and Swinarski et al., 2012). Historically, methanol has been used most commonly as the carbon source for the following reasons: (1) highest effectiveness ratio in terms of g NO_x reduced/g methanol consumed in the anoxic zone, (2) generally the lowest cost per kg of NO_x removed, (3) readily available, and (4) considerable experience with its application for denitrification. Major disadvantages of using methanol are (1) safety issues associated with its transportation and storage, (2) the need for acclimation time when first added, and (3) lower denitrification rates compared to other substrates. The use of external carbon sources such as methanol or other flammable substances requires special storage and handling designs and concern for worker safety. Issues related to storage and safety, denitrification rates, and acclimation time has led to interest in other carbon sources. Important considerations in the use of external carbon sources, discussed in this section, are carbon effectiveness, denitrification rates, carbon dose and anoxic volume, acclimation needs, and reduction end products.

External Carbon Effectiveness. The external carbon effectiveness ratio is defined as the g NO_3 reduced/g COD consumed in the anoxic zone. For substrates with a higher synthesis yield coefficient, a greater fraction of the COD removed goes to biomass and a lower fraction is oxidized, which results in a lower effectiveness ratio.

The effectiveness ratio (E_{CNO_3}) is the inverse of the consumptive ratio (C_R), defined by McCarty et al. (1969) and described in Eq. (7–127) in which the yield is the synthesis yield with consumption of the external carbon source.

$$\frac{1}{E_{CNO_3}} = C_{R,NO_3} = \frac{2.86}{1 - 1.42\,Y_H} \tag{8–69}$$

where Y_H = synthesis yield, g VSS/g COD removed

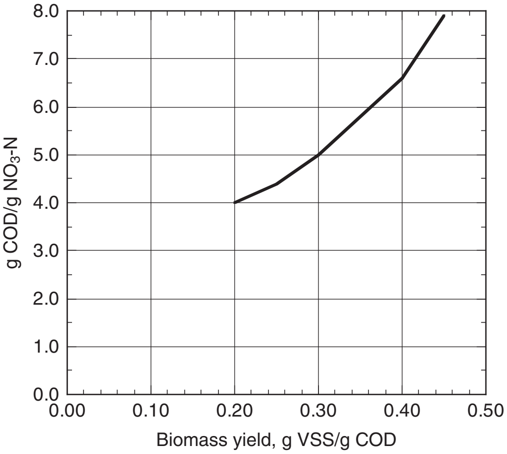


Figure 8–36

Exogenous carbon dose for nitrate removal is related to the biomass yield in g VSS produced per g COD consumed.

For the reduction of nitrite and DO the value of the numerator in Eq. (8–69) is 1.71 and 1.0, respectively. External carbon sources that result in lower synthesis yield values have a lower required C_R value, which results in a lower carbon dose for a given amount of nitrate removal. The effect of the synthesis yield on the C_R value is illustrated on Fig. 8–36. Reported synthesis yield values for methanol range from 0.20 to 0.30 (Gu and Onnis-Hayden, 2010) and thus the methanol $C_{R,\mathrm{NO_3}}$ values range from 4.0 to 5.0 g COD/g NO_3-N. Anoxic synthesis yield values for other external carbon sources may range from 0.35 to 0.40. It should be noted that the synthesis yield under anoxic conditions is 70 to 80 percent of that under aerobic respiration (Muller et al., 2003, Henze et al., 2008). Using reported synthesis yields with nitrate reduction of 0.36 (g VSS/g COD) (Christensen et al., 1994) for ethanol and 0.34 (g VSS/g COD) for glycerol (Bilyk et al., 2009), their respective C_R values from Eq. (8–69) are 5.9 and 5.5 g COD/g NO_3-N, which is higher than the corresponding values for methanol. The respective effectiveness factors for methanol, ethanol, and glycerol, based on these $C_{R,\mathrm{NO_3}}$ values are 0.22, 0.18, and 0.17 g NO_3-N/g COD consumed.

The amount of external carbon added per unit of NO_3-N removal in a postanoxic zone is higher than that predicted by the consumptive ratio. Methanol has a COD equivalent of 1.5 g COD/g CH_3OH, so the C_R based on methanol is 2.7 to 3.3 g CH_3OH/g NO_3-N removed. Methanol dosage in field application has ranged from 3.3 to 3.8 g CH_3OH/g NO_3-N removed, which is due to the net effect of the true C_R value, methanol in the anoxic zone effluent, methanol consumed by DO entering the anoxic zone, and nitrate reduction by substrate available from mixed liquor endogenous decay. For a smaller anoxic tank (lower τ) a higher denitrification rate is needed, which requires a higher concentration of the external carbon in the reactor and effluent, and thus a higher external carbon dose. The external carbon dose is related to the amount of carbon consumed to provide the demand for respiration with NO_3-N reduction and its concentration in the reactor to drive the necessary denitrification rate.

Denitrification Rates. The NO_3-N reduction rate (g/L·d) is a function of the carbon substrate, NO_3-N, and heterotrophic biomass concentrations as given by Eq. (7–133).

$$r_{\mathrm{NO_3}} = \left(\frac{1 - 1.42Y_{\mathrm{H}}}{2.86}\right)\left[\frac{\mu_{\mathrm{H,max}}S_S}{Y_{\mathrm{H}}(K_s + S_S)}\right]\left(\frac{S_{\mathrm{NO_3}}}{K_{\mathrm{NO_3}} + S_{\mathrm{NO_3}}}\right)\left(\frac{K_o'}{K_o' + S_o}\right)(\eta)X_{\mathrm{H}} \tag{7–133}$$

where $(\eta)X_{\mathrm{H}}$ = biomass concentration capable of degrading the external carbon substrate under anoxic conditions, mg VSS/L

Some external carbon substrates, such as acetate and ethanol, are readily biodegradable by most of the heterotrophic bacteria grown from BOD in the influent wastewater so that typical values of η, (0.60 to 0.80 for anoxic/aerobic systems) may be used in Eq. (7–133) with calculated values for X_H. A similar assumption cannot be used for anoxic degradation with methanol addition.

Methylotrophic bacteria, capable of degrading methanol, are more specialized bacteria capable of growing on single carbon compounds so that the value for X_H in Eq. (7–133) with methanol is related mainly to the amount of methanol fed and degraded. These bacteria can utilize other external carbon sources added to a denitrification system, but would be competing with other bacteria for the substrate. It has been shown that anoxic culture grown on methanol can contain populations also capable of ethanol degradation (Baytshtok et al., 2009). The other substrates noted in this section can be degraded by a wide range of bacteria that can also degrade substrates contained in the influent wastewater BOD.

Denitrification rates with methanol are much lower compared to those for acetate, ethanol, a commercial product MicroC™, and corn syrup, based on the maximum specific growth rates reported in Table 8–26 for bacteria consuming these substrates with nitrate reduction.

While Eq. (7–133) can be used in reactor mass balances and simulation models to analyze anoxic zone volume and external carbon needs, a desktop design approach can be done using SDNR values as was shown for the design of preanoxic tanks in Example (8–7) and Eq. (8–52).

$$\mathrm{NO}_r = (V_{\mathrm{nox}})(\mathrm{SDNR})(\mathrm{MLVSS}) \tag{8–52}$$

where NO_r = $\mathrm{NO_3}$-N removal rate, g/d

The SDNR as a function of biodegradation kinetics for the carbon source is obtained by dividing Eq. (7–133) above by the MLVSS (X_{VSS}) concentration.

$$\mathrm{SDNR} = \left(\frac{1 - 1.42Y_{\mathrm{H}}}{2.86}\right)\left[\frac{\mu_{\mathrm{H,max}}S_{\mathrm{S}}}{Y_{\mathrm{H}}(K_s + S_{\mathrm{S}})}\right]\left(\frac{S_{\mathrm{NO_3}}}{K_{\mathrm{NO_3}} + S_{\mathrm{NO_3}}}\right)\left(\frac{K'_o}{K'_o + S_o}\right)\left[\frac{(\eta)X_{\mathrm{H}}}{X_{\mathrm{VSS}}}\right] \tag{8–70}$$

where SDNR = specific denitrification rate, g $\mathrm{NO_3}$-N/g VSS·d
X_{H} = the mixed liquor biomass concentration, mg/L
η = the fraction of biomass able to degrade the external carbon with nitrate reduction
X_{VSS} = mixed liquor volatile suspended solids concentration, mg/L

SDNR values for the non-methanol carbon sources are about 2 to 2.5 times higher than that for methanol. SDNRs in anoxic activated sludge determined by Fillos et al. (2007) for

Table 8–26

Comparison of kinetics for various external carbon sources for denitrification

Substrate	$\mu_{\max}$ at 20°C, g/g·d	Arrhenius coefficient, θ	Reference
Methanol	1.12	1.12	Mokhayeri et al. (2006)
	1.3	1.1	Christensson et al. (1994)
	1.3	1.09	Dold et al. (2008)
Acetate	4.46	1.21	Mokhayeri et al. (2006)
Ethanol	3.02	1.1	Christensson et al. (1994)
MicroC™	2.05	1.02	Onnis-Hayden et al. (2011)
Corn Syrup	4.13	1.18	Mokhayeri et al. (2006)

ethanol and methanol at high substrate concentration are given by Eqs. (8–71) and (8–72). The SDNR with ethanol is about 2.1 times faster than with methanol.

$$\text{Methanol:} \quad \text{SDNR} = 0.0738(1.11)^{T-20} \tag{8–71}$$

$$\text{Ethanol:} \quad \text{SDNR} = 0.161(1.13)^{T-20} \tag{8–72}$$

Carbon Dose and Postanoxic Tank Volume. The amount of external carbon that must be added to a postanoxic tank is a function of the amount of nitrate to be removed due to consumption of the external carbon, the denitrification kinetics for the carbon selected, the reactor external carbon substrate concentration and the anoxic tank volume. The design procedure involves selecting an anoxic volume as a first step. Based on the maximum specific growth rate for anoxic methanol degradation and an endogenous decay rate of 0.05 g/g·d (Stensel et al., 1973), an anoxic zone with methanol addition needs an SRT value >1.0 d at 20°C and >2.0 d at 10°C to prevent washout. The anoxic zone SRT for other external carbon sources is 20 to 50 percent less.

The nitrate removal rate in the postanoxic zone equals the removal rate due to the external carbon consumption plus the removal rate due to endogenous decay (Eq. 8–63).

$$R_{\text{NO}_3} = \text{SDNR}(X_{\text{VSS}})(V_{\text{anox}}) + \left(\frac{1.42}{2.86}\right)(b_{\text{H,anox}})(X_{\text{H}})(V_{\text{anox}}) \tag{8–73}$$

where R_{NO_3} = NO_3-N removal rate in the anoxic tank, g/d

The removal rate equals the change in the NO_3-N concentration between the tank influent and effluent times the flowrate to the postanoxic tank.

$$R_{\text{NO}_3} = Q(1 + R)(\text{NO}_o - \text{NO}_e) \tag{8–74}$$

where: NO_o = Postanoxic tank influent NO_3-N concentration, g/m^3
NO_e = Postanoxic tank effluent NO_3-N concentration, g/m^3
R = Return activated sludge recycle ratio

For a given anoxic volume the SDNR necessary using the external carbon source is determined from Eq. (8–75).

$$\text{SDNR} = \frac{R_{\text{NO}_x} - \left(\frac{1.42}{2.86}\right)(b_{\text{H}})(X_{\text{H}})(V_{\text{anox}})}{(X_{\text{VSS}})(V_{\text{anox}})} \tag{8–75}$$

The anoxic tank external carbon substrate concentration (S_s) needed to satisfy the SDNR value in Eq. (8–75) is determined from Eq. (8–70). At higher SDNR values the anoxic tank effluent substrate concentration is higher.

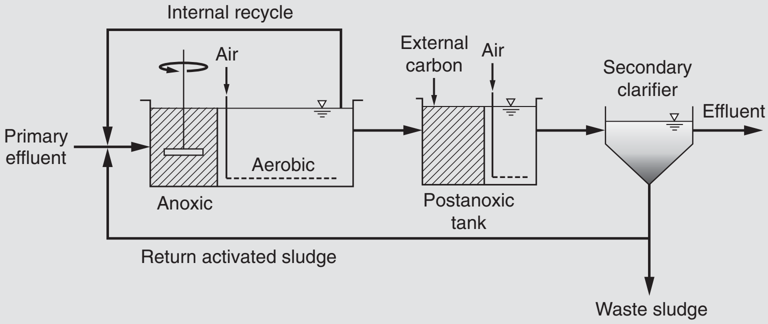
The carbon dose is a function of the carbon consumed for nitrate reduction plus the amount of carbon in the effluent needed to sustain the necessary SDNR.

$$C_{\text{D}} = \text{SDNR}(X_{\text{VSS}})(V_{\text{anox}})C_{\text{R,NO}_3} + Q(1 + R)(S_s) \tag{8–76}$$

where C_{D} = carbon dose or amount of external carbon added, g COD/d

For a smaller anoxic volume, the SDNR must be increased, which increases the effluent external carbon concentration. In addition there is less NO_3-N removal due to endogenous decay. These effects result in a higher carbon dose requirement. The application of these relationships to estimate an external carbon dose is illustrated in Example 8–12.

EXAMPLE 8–12 **Design of Postanoxic Tank with External Carbon Addition for Nitrate Removal** Using the results of the MLE design in Example 8–7 and the postanoxic design in Example 8–10, evaluate the postanoxic design use ethanol or methanol for external carbon sources. Determine (a) the amount of ethanol needed in kg/d and g/m^3 (based on the influent flowrate) for a postanoxic tank at 1/3rd the volume of that needed for nitrate reduction with only endogenous decay in Example 8–10; (b) the amount of methanol needed as a function of the postanoxic tank volume; and (c) prepare a plot of the percent NO_3-N removed by endogenous decay, the percent of the feed methanol in the effluent, and the ratio of the methanol added versus the NO_3-N removed as a function of the postanoxic tank volume and discuss the impact of increasing the volume of the postanoxic tank. Determine the change in NH_4-N concentration across the postanoxic tank for part (a). As noted in Example 8–10 the postanoxic tank is part of a Bardenpho process.



Design conditions and assumptions:

1. Information from the anoxic/aerobic system and postanoxic tank in Examples 8–7 and 8–10.

Item	Unit	Value
Flowrate	m^3/d	22,700
RAS ratio	unitless	0.60
Temperature	°C	12
MLSS	g/m^3	3000
MLVSS	g/m^3	2370
Biomass, X_H	g/m^3	1267
Endogenous decay, $b_{H,12}$	g/g·d	0.06
Aerobic SRT	d	20.6
Aerobic basin volume	m^3	13,230
Aerobic NO_3-N	g/m^3	6.0
Postanoxic basin volume	m^3	4,811

Note: g/m^3 = mg/L.

1. $\eta = 0.80$
2. NH_4-N used for cell synthesis = 0.12 g NH_4-N/g VSS produced
3. NH_4-N release in endogenous decay = 0.06 g NH_4-N/g VSS

4. Ethanol $\mu_{\max}$ at 20°C (Table 8–26) = 3.02 g/g·d
5. Ethanol $\mu_{\max}$ θ value (Table 8–26) = 1.1
6. Ethanol synthesis yield, $Y_H = 0.36$ g VSS/g COD
7. Ethanol half-velocity coefficient, K_S, = 5.0 g COD/m^3
8. Ethanol is degraded by all the mixed liquor denitrifying biomass
9. Nitrate half-velocity coefficient, $K_{NO_3} = 0.10$ g/m^3
10. DO concentration = 0.0 g/m^3
11. Methanol $\mu_{\max}$ at 20°C (Table 8–26) = 1.2 g/g·d
12. Methanol $\mu_{\max}$ θ value (Table 8–26) = 1.1
13. Methanol synthesis yield (Gu and Onnis-Hayden, 2010), $Y_H = 0.30$ g VSS/g COD
14. Methanol decay coefficient (Stensel et al., 1973) = 0.04 g/g·d
15. Methanol half-velocity constant (Torres et al., 2011) = 1.0 g/m^3
16. Postanoxic effluent NO$_3$-N concentration = 1.0 g/m^3

Solution, Part A—Postanoxic volume with ethanol dose

1. Select the anoxic volume.

$$V_{anox} = 1/3(4811\text{ m}^3) = 1604\text{ m}^3$$

$$\tau = (1604\text{ m}^3)(24\text{ h/d})/(22{,}700\text{ m}^3/\text{d}) = 1.7\text{ h}$$

2. Determine required SDNR by heterotrophic biomass utilizing ethanol with Eq. (8–73). From Example 8–10, $R_{NO_3} = 181{,}000$ g/d.

$$R_{NO_3} = \text{SDNR}(X_{VSS})(V_{anox}) + \left(\frac{1.42}{2.86}\right)(b_{H,anox})(X_H)(V_{anox})$$

$$181{,}600\text{ g/d} = \text{SDNR}(2370\text{ g/m}^3)(1604\text{ m}^3) + \frac{1.42}{2.86}(0.06\text{ g/g·d})(1267\text{ g/m}^3)(1604\text{ m}^3)$$

$$181{,}600\text{ g/d} = 3{,}801{,}480\text{ g/d(SDNR)} + 60{,}542\text{ g/d}$$

$$\text{Required SDNR} = 0.032\text{ g NO}_3\text{-N/gVSS·d}$$

3. Determine effluent ethanol concentration to provide the required SDNR by solving for S_S in Eq. (8–70) and assuming DO concentration = zero.

$$\text{SDNR} = \frac{1 - 1.42Y_H}{2.86}\left[\frac{\mu_{\max}S_S}{Y_H(K_S + S_S)}\right]\left(\frac{S_{NO_3}}{K_{NO_3} + S_{NO_3}}\right)\left[\frac{(\eta)X_H}{X_{VSS}}\right]$$

a. $\mu_{\max,12°C} = 3.02\text{ g/g·d}\,(1.1^{12-20}) = 1.41\text{ g/g·d}$

$$0.032\text{ g/g·d} = \left[\frac{1 - 1.42(0.36\text{ g/g})}{2.86}\right]\left[\frac{1.41\text{ g/g·d}(S_S)}{(0.36\text{ g/g})(5.0 + S_S)}\right]\left[\frac{(1.0\text{ g/m}^3)}{(0.1 + 1.0)\text{ g/m}^3}\right]\left[\frac{(0.80)(1267\text{ g/m}^3)}{(2370\text{ g/m}^3)}\right]$$

$$S_S = 0.70\text{ g/m}^3$$

4. Determine the ethanol dose.
   a. Determine the ethanol consumptive ratio for nitrate reduction using Eq. (8–69).

$$C_{\text{R,NO}_3} = \frac{2.86}{1 - 1.42(0.36 \text{ g VSS/g COD})} = 5.85 \text{ g COD/g NO}_3\text{-N}$$

   b. Calculate the ethanol dose using Eq. (8–76).

$$\text{Dose} = \text{SDNR}(X_{\text{VSS}})(X_{\text{anox}})C_{\text{R,NO}_3} + Q(1 + R)(S_S)$$

$$\text{Dose} = (0.032 \text{ g/g·d})(2370 \text{ g/m}^3)(1604 \text{ m}^3)(5.85 \text{ g COD/g NO}_3\text{-N}) + 22{,}700 \text{ m}^3\text{/d}(1 + 0.60)(0.70 \text{ g/m}^3)$$

$$\text{Dose} = 711{,}637 \text{ g COD/d} + 25{,}424 \text{ g COD/d}$$

$$\text{Dose} = 737{,}061 \text{ g COD/d}$$

$$\text{COD of ethanol } (CH_3CH_2OH) = 2.09 \text{ g COD/g ethanol}$$

$$\text{Dose as ethanol} = \frac{(737{,}061 \text{ g COD/d})}{(2.09 \text{ g COD/g ethanol})} = 352{,}661 \text{ g ethanol/d}$$

$$= 352.6 \text{ kg ethanol/d}$$

Dose as concentration of ethanol added to flow to anoxic tank (influent plus recycle flowrates):

$$= \frac{(352{,}661 \text{ g ethanol/d})}{[(1 + 0.6)22{,}700 \text{ m}^3\text{/d}]} = 9.7 \text{ g ethanol/m}^3 = 20.3 \text{ g COD/m}^3$$

$$\text{Dose normalized to influent flowrate} = \frac{(352{,}661 \text{ g ethanol/d})}{(22{,}700 \text{ m}^3\text{/d})}$$

$$= 15.5 \text{ g ethanol/m}^3$$

5. Determine the change in NH_4-N across the postanoxic tank. The change is due to NH_4-N release from endogenous decay and NH_4-N uptake for cell synthesis from ethanol consumption. The amount of NH_4-N released is equal to the ratio of the anoxic volume in this problem (1604 m^3) to that in Example 8–10, step 3 (4811 m^3) times the NH_4-N release (0.60 g/m^3) in Example 8–10 as the specific endogenous decay rates and biomass concentrations are the same.

$$\Delta NH_4\text{-N released} = \frac{NH_4\text{-N}(V_{\text{anox}})}{V_{\text{anox}}}$$

$$= \frac{(0.60 \text{ g/m}^3)(1604 \text{ m}^3)}{4811 \text{ m}^3}$$

$$\Delta NH_4\text{-N} = 0.20 \text{ g/m}^3$$

$$\Delta\text{biomass} = Y(\Delta\text{COD}) = (0.36 \text{ g VSS/g COD})(20.3 \text{ g COD/m}^3) = 7.3 \text{ g VSS/m}^3$$

Nitrogen uptake to biomass needed with ethanol consumption

$$= 0.12 \text{ g N/g VSS}(7.3 \text{ g VSS/m}^3) = 0.88 \text{ g/m}^3$$

$$\text{Net change in } NH_4\text{-N} = 0.20 \text{ g/m}^3 - 0.88 \text{ g/m}^3 = -0.68 \text{ g/m}^3$$

Thus, a decrease in effluent NH_4-N will occur across the anoxic zone fed ethanol. As the NH_4-N concentration is decreased, it is possible that some of the nitrogen needs will also be met with NO_3-N.

Solution, Part B—Methanol dose versus postanoxic volume

The solution procedure with methanol as the carbon source is modified from that using ethanol. It is assumed that the methanol-degrading biomass is only from growth on methanol and not from BOD in the wastewater. The solution steps require an iterative procedure (or one that can be solved with Excel Solver) to determine the postanoxic reactor methanol concentration to provide the necessary nitrate reduction rate.

1. Using the same procedure as with ethanol (Part A), determine the required methylotrophic SDNR with Eq. (8–73) for $V_{\text{anox}} = 1604\ \text{m}^3$:

$$R_{NO_3} = \text{SDNR}(X_{\text{VSS}})(V_{\text{anox}}) + \left(\frac{1.42}{2.86}\right)(b_{\text{H}})(X_{\text{H}})(V_{\text{anox}})$$

$$\begin{aligned} 181{,}600\ \text{g/d} &= \text{SDNR}(2370\ \text{g/m}^3)(1604\ \text{m}^3) \\ &\quad + \frac{1.42}{2.86}(0.06\,\text{g/g·d})(1267\ \text{g/m}^3)(1604\ \text{m}^3) \end{aligned}$$

$$181{,}600\ \text{g/d} = 3{,}801{,}480\ \text{g/d(SDNR)} + 60{,}542\ \text{g/d}$$

Required methylotrophic SDNR = 0.032 g NO_3-N/gVSS·d. The second term in the above equation is nitrogen removed by endogenous decay.

$$\text{Percent removal by endogenous decay} = \frac{60{,}542(100)}{181{,}600} = 33.3\%$$

2. Determine effluent methanol concentration to provide the required SDNR by solving for S_S in Eq. (8–70) and assuming DO concentration = zero. Note that the biomass concentration (ηX_{H}) in this case is the methylotrophic biomass or X_{M}.

$$\text{SDNR} = \frac{1 - 1.42Y_{\text{H}}}{2.86}\left[\frac{\mu_{\max} S_S}{Y_{\text{H}}(K_S + S_S)}\right]\left(\frac{S_{NO_3}}{K_{NO_3} + S_{NO_3}}\right)\left(\frac{X_{\text{M}}}{X_{\text{VSS}}}\right)$$

a. $\mu_{\max,12^\circ\text{C}} = 1.2\ \text{g/g·d}(1.1^{12-20}) = 0.56\ \text{g/g·d}$

$$\begin{aligned} 0.032\ \text{g/g·d} &= \left[\frac{1 - 1.42(0.30\ \text{g/g})}{2.86}\right]\left[\frac{(0.56\ \text{g/g·d})(S_S)}{(0.30\ \text{g/g})(1.0 + S_S)}\right] \\ &\quad \left\{\frac{(1.0\ \text{g/m}^3)}{[(0.1 + 1.0)\text{g/m}^3]}\right\}\left[\frac{(X_{\text{M}}\ \text{g/m}^3)}{(2370\ \text{g/m}^3)}\right] \end{aligned}$$

$$0.032\ \text{g/g·d} = 0.000144\left(\frac{S_S}{1.0 + S_S}\right)(X_{\text{M}})$$

b. Determine the methylotrophic biomass concentration (X_{M}) as a function of the methanol consumed. The amount of methanol consumed is equal to the methanol dose (assumes that all of the methanol added will be consumed in the anoxic tank or in the subsequent post aeration tank). Calculate the methanol dose using Eq. (8–76).

$$\text{Dose} = \text{SDNR}(X_{\text{VSS}})(V_{\text{anox}})C_{\text{R},NO_3} + Q(1 + R)(S_S)$$

The value for $[\text{SDNR}(X_{\text{VSS}})(V_{\text{anox}})]$ is determined from step 1, as it is the rate of nitrate removal minus the rate removed by endogenous decay.

$$[\text{SDNR}(X_{\text{VSS}})(V_{\text{anox}})] = 181{,}600 \text{ g/d} - 60{,}542 \text{ g/d} = 121{,}058 \text{ g NO}_3\text{-N/d}$$

$$C_{R,\text{NO}_3} = \frac{2.86}{1 - 1.42(Y_{\text{H}})},\ Y_{\text{H}} = 0.30 \text{ g VSS/g COD}$$

$$= \frac{2.86}{1 - 1.42(0.30 \text{ g VSS/g COD})} = 4.98 \text{ gCOD/g NO}_3\text{-N}$$

Thus, the methanol consumed is:

$$\text{Dose, g COD/d} = (121{,}058 \text{ g NO}_3\text{-N/d})(4.58 \text{ g COD/g NO}_3\text{-N}) + (22{,}700 \text{ m}^3\text{/d})(1.0 + 0.6)(S_S \text{ g/m}^3)$$

$$\text{Dose, g COD/d} = 554{,}445.6 + 36{,}320(S_S)$$

X_{M} is related to the biomass produced from the methanol consumed and the system SRT and τ as given by Eq. (7–42) in Table 10.

$$X_{\text{M}} = \left(\frac{\text{SRT}}{\tau}\right)\left[\frac{Y_{\text{H}}(S_o)}{1 + b_{\text{H}}(\text{SRT})}\right]$$

where S_o = methanol consumed normalized to the influent flowrate, g/m^3.

$$S_o = \frac{[554{,}445.6 + 36{,}320(S_S)]\text{g/d}}{22{,}700 \text{ m}^3\text{/d}} = [24.43 + 1.6(S_S)]\text{g/m}^3$$

The SRT and τ are based on the total Bardenpho system volume. A 0.33 h post aeration time, τ, is the assumed after the postanoxic tank. The total SRT can be estimated as proportionate to the aerobic volume and its SRT determined in example problem 8–7.

Bardenpho Tanks	Volume, m^3
Preanoxic	2361
Aerobic (SRT = 20.5 d)	13,230
Postanoxic	1604
Postaerobic	315
Total	17,510

$$\tau = \frac{(17{,}510 \text{ m}^3)}{(22{,}700 \text{ m}^3\text{/d})} = 0.77 \text{ d}$$

$$\text{SRT} = \frac{17{,}510 \text{ m}^3}{13{,}230 \text{ m}^3}(20.5 \text{ d}) = 27.1 \text{ d}$$

$$X_{\text{M}} = \left(\frac{\text{SRT}}{\tau}\right)\left[\frac{Y_{\text{H}}(S_o)}{1 + b_{\text{H}}(\text{SRT})}\right]$$

$$= \left(\frac{27.1 \text{ d}}{0.77 \text{ d}}\right)\left[\frac{(0.30 \text{ g VSS/g COD})(S_o)}{1 + (0.04 \text{ g/g·d})(27.1 \text{ d})}\right] = 5.066(S_o)$$

and from above $S_o = 24.43 + 1.6\text{SS}$, thus $X_\text{M} = 123.76 + 8.106S_S$. Using the equation for SDNR from 2a, X_M and S_S are solved.

$$0.032\text{ g/g·d} = 0.000144\left(\frac{S_S}{1.0 + S_S}\right)(X_\text{M})$$

$$S_S = 12.6\text{ g COD/m}^3 \text{ and } X_\text{M} = 236.3\text{ g VSS/m}^3$$

3. The methanol dose is now determined.

$$\text{Dose} = \text{SDNR}(X_\text{VSS})(V_\text{anox})C_{\text{R,NO}_3} + Q(1 + R)(S_S)$$

$$= (0.032\text{ g NO}_3\text{-N/g VSS·d})(2370\text{ g/m}^3)(1604\text{ m}^3)(4.98\text{ g COD/g NO}_3\text{-N})$$
$$+ (22{,}700\text{ m}^3\text{/d})(1 + 0.6)(12.6\text{ g COD/m}^3)$$

$$\text{Dose} = 605{,}803\text{ g COD/d} + 456{,}905\text{ g COD/d} = 1{,}062{,}709\text{ g COD/d}$$

The percent of methanol dose in the effluent flow from the anoxic tank:

$$\text{Percent of methanol dose in anoxic tank effluent} = \frac{100(456{,}905)}{1{,}062{,}709} = 43.0\%$$

Dose based on flowrate to the anoxic tank:

$$= \frac{(1{,}062{,}709\text{ g COD/d})}{(22{,}700\text{ m}^3)(1.0 + 0.6)} = 29.3\text{ g COD/m}^3$$

Normalized to influent flowrate:

$$= \frac{(29.3\text{ g COD/m}^3)(1.0 + 0.6)(22{,}700\text{ m}^3)}{(22{,}700\text{ m}^3)} = 46.8\text{ g COD/m}^3$$

Dose as methanol:

$$\text{Total methanol added/day} = \frac{(1{,}062{,}709\text{ g COD/d})}{(1.5\text{ g COD/g CH}_3\text{OH})(1000\text{ g/kg})} = 708.5\text{ kg CH}_3\text{OH/d}$$

$$\text{Based on flowrate to anoxic tank} = (29.3\text{ g COD/m}^3)/(1.5\text{ g COD/g CH}_3\text{OH})$$
$$= 19.5\text{ g CH}_3\text{OH/m}^3$$

$$\text{Based on influent flowrate} = (48.9\text{ g COD/m}^3)/1.5 = 32.6\text{ g CH}_3\text{OH/m}^3$$

The following solution is summarized:

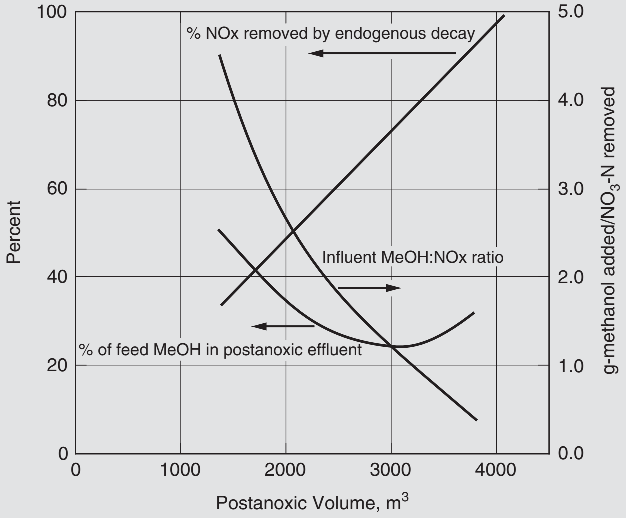
Parameter	Unit	Value
Anoxic volume	m^3	1604
Methanol dose	kg/d	708.5
Methanol dose based on influent flowrate	$\text{g CH}_3\text{OH/m}^3$	32.6
Methanol dose based on flowrate to anoxic tank	$\text{g CH}_3\text{OH/m}^3$	19.5
Percent of feed methanol in postanoxic tank effluent	percent	43.0

Solution, Part C—Prepare required plot and discuss results

1. Percent NO_3-N removed by endogenous decay:
   The percent values are obtained as illustrated in Part B, Step 1.
   Percent removal by endogenous decay for a tank volume of 1604 m^3 = 33.3%.
2. Percent of the feed methanol in the effluent is obtained as illustrated in Part B, Step 3:
   Percent of the feed methanol in the effluent for a tank volume of 1604 m^3 = 43%.
3. Ratio of the methanol added versus the NO_3 removed for a tank volume of 1604 m^3:
   The methanol added from Part B, Step 3 = 708,500 g CH_3OH /d.
   The NO_3 removed = 181,600 g/d.
   Ratio = g CH_3OH/g NO_3-N = 708,500/181,600 = 3.9.
4. Prepare summary computation table following similar calculations for different assumed anoxic volumes

Postanoxic volume, m^3	Percent of NO_3-N removed by endogenous decay	Percent of feed CH_3OH in effluent	CH_3OH/NO_3-N
1604	33.3	43.0	3.9
1800	37.4	38.5	3.4
2000	41.6	33.9	2.9
2500	52.0	25.3	2.1
3000	62.4	23.0	1.6
3500	72.7	22.2	1.2
4000	83.1	29.2	0.8

5. Prepare plot and analyze the effect of increasing the postanoxic tank volume.
   a. The required plot is given below.



b. As the postanoxic tank volume is increased, the methanol dose decreases because more nitrate removal is by endogenous decay and there is a lower effluent methanol concentration due the fact that a lower denitrification rate with the external carbon substrate was allowable. At an anoxic volume around 2000 m^3, the methanol dose is in the range of the consumptive ratio (C_{R,NO_3}) at 2.5 to 3.5 g CH_3OH/g NO_3-N removed (3.8 to 5.3 on a COD basis), but this apparent match to the C_{R,NO_3} value is the net result of methanol utilization for denitrification in the postanoxic reactor, nitrate removed by endogenous decay, and methanol in the effluent.

Comment The result of these analyses on the effect of removing nitrate in the postanoxic zone by endogenous decay only, with ethanol addition and with methanol addition illustrate the effect of the substrate utilization rate kinetics associated with a particular exogenous carbon source and the effect of the postanoxic volume size on the carbon dose needed. Because of the greater substrate utilization kinetics with ethanol, the ethanol concentration could be lower in the postanoxic reactor and less of the ethanol added was in the reactor effluent compared to methanol. This resulted in a lower dose requirement for ethanol compared to that needed with methanol.

Acclimation Needs. Applications in which an external carbon must be added on a seasonal basis or intermittently must consider if the carbon selected is readily degradable in the activated sludge mixed liquor (such as acetate) or if an acclimation time is needed to develop a sufficient biomass. An acclimation time is needed to develop a methylotrophic population as they are not present to much extent in activated sludge systems treating domestic wastewater. At least 2 weeks are needed to develop an effective methanol-degrading population, and longer times of two to three SRTs have been recommended by Nyberg (1996) to reach full population development. Glycerol requires a few days acclimation time to reach full degradation capacity (Dailey et al., 2012). If the need for an external carbon source is quite variable due to changing demand during wet weather or other local variations, a carbon source that requires little acclimation would be appropriate.

Reduction End Products. The goal in denitrification is to reduce the nitrate to nitrogen gas but in some instances a portion of the nitrate may be reduced to only nitrite. This may be due to insufficient carbon with the faster conversion of nitrate to nitrite consuming the available carbon, an insufficient detention time, or the selection of a denitrifying population that only converts nitrate to nitrite. In recent denitrification studies with acetate as the carbon source, nitrite accumulation has been observed along with nitrate reduction (Cherchi. et al., 2009). Nitrite accumulation was also observed in studies with glycerol addition (Uprety et al., 2012). Acetate is a product of glycerol fermentation (Gall et al., 2008) and thus acetate was likely produced and consumed in an anoxic reactor fed glycerol. Nitrite accumulation has been attributed to decreased nitrite reductase activity in the presence of acetate (Van Rijn, 1996), whereas Uprety et al. (2012) attribute it to a high F/M loading and substrate storage by the biomass so that limited carbon is left to drive the nitrite reduction reaction. No nitrite accumulation was observed when a second external carbon (methanol) was added with glycerol addition (Oreskovich et al., 2011).

Process Control and Performance

Process control methods for nitrogen removal systems are important with regard to the following treatment process considerations: (1) maintaining nitrification performance, (2) online instrumentation to enable the application of specific nitrogen removal processes, and (3) the use of online instrumentation to minimize effluent inorganic nitrogen concentrations and external carbon dosage. These applications involve the use of one or more of the following instruments that have been applied successfully for online measurements and control in biological nitrogen removal suspended growth systems (WERF, 2007): DO probes, suspended solids analyzers, ORP probes, ammonia analyzers, and nitrate and nitrite analyzers.

Nitrification Performance. The treatment efficiency of biological nitrogen removal processes is related to the ability to meet the desired level of nitrification. The conversion of ammonia to nitrate and nitrite, as noted previously, is related to the aerobic zone DO concentration and the ability to maintain an adequate aerobic SRT. The importance of DO and SRT control with respect to treatment kinetics was presented and discussed in Sec. 8–3. In addition, DO control is also important with respect to (1) the aeration system performance optimization and energy usage in nitrification systems, and (2) the minimization of the DO concentration in the internal recycle flows to a preanoxic zone. If surface wasting is employed to control *Nocardioform* foam, a common occurrence in MLE processes, it is of critical importance that the solids lost with the foam be accounted for in computing the SRT for process control.

Nitrogen Removal Processes with Online Analyzers. Certain types of nitrogen removal processes are dependent on the use of specific online analyzers. The low DO SNdN process depends on DO probes for aeration control at low DO concentrations. The more recently developed optical DO probes have allowed reliable, accurate, and continuous measurements of low DO concentrations for SNdN systems. Optical methods measure changes in light emitted by a luminescent chemical, which is related the solution DO concentration. The control of DO is also important in cyclic and phased operation NdN processes.

Measurements with ORP probes are important for an intermittent aeration NdN process such as the oxidation ditch shown on Fig. 8–29. In the aerobic/anoxic cycling, the aeration can be turned off at selected times and online ORP measurements can be used to determine when to start aeration again. The ORP response during an aeration-off period is shown on Fig. 8–29(c). As the DO concentrations decline, the ORP value decreases. When the NO_3-N is depleted, a dramatic decline in the ORP value occurs. The ORP decline is called the ORP knee and can be identified by calculating the ORP slope with time. The ORP values are logged onto a computer, which is programmed to turn on the aeration based on the changing slope of the ORP. The aeration-off periods are selected to occur during different times of the day; the more ideal time is when the influent BOD concentration is high so that nitrate reduction occurs at a faster rate. ORP control for nitrogen removal has also been shown for nitrogen removal in aerobic digestion (Koch et al., 1985) and in SNdN systems (Mavinic et al., 2005). It may also be useful in determining when nitrate is removed in cyclic and phased operation NdN systems.

Effluent Inorganic Nitrogen Concentration and External Carbon Dose. Continuous online measurements of ammonia and nitrate/nitrite concentrations are very useful for process optimization to achieve minimal effluent inorganic nitrogen concentrations and to control the external carbon dose in postanoxic denitrification processes.

Ion-selective electrode probes have been used for ammonia measurements; the most common method for nitrate and nitrite measurements is by UV absorbance techniques. Suspended solids filtration is often used with these methods to prevent interferences.

Online nitrate/nitrite measurements are particularly important for control of external carbon dosing in systems where the goal is to meet a minimal effluent inorganic nitrogen concentration. The amount of carbon needed varies due to diurnal variations in flow and nitrogen loadings. A high carbon dose can be used to maximize nitrate removal, but is wasteful and can also increase the effluent BOD concentration. A feed forward control scheme is used normally in which the flowrate and upstream nitrate concentration measurements before the anoxic tank are used to adjust the external carbon feed rate. In some cases, an online measurement of nitrate concentration in the anoxic zone is done to prevent over dosing by maintaining a residual low nitrate concentration.

8–8 PROCESSES FOR ENHANCED BIOLOGICAL PHOSPHORUS REMOVAL

The fundamental principles of enhanced biological phosphorus removal (EBPR) have been described in section 7–13 in Chap. 7. The process is based on contacting activated sludge mixed liquor with the influent wastewater or other feed stream containing volatile fatty acids (VFAs) in an anaerobic tank prior to entering anoxic or aerobic tanks. In contrast to the other heterotrophic organisms, the phosphorus accumulating organisms (PAOs) are able to assimilate and transform acetate and propionate to intracellular carbohydrate storage products in the anaerobic zone. These VFAs may be available in the influent feed to the anaerobic zone and/or produced by fermentation of influent rbCOD. The anaerobic zone provides a competitive advantage for PAOs and thus has been referred to as an anaerobic selector process for PAOs. A PAO-enriched activated sludge also contains a dense, well developed floc with good settling and thickening properties for the reasons discussed in Sec. 8–4 which dealt with selectors. The material on phosphorus removal presented in this section is organized into the following topics: (1) process development, (2) overview of EBPR processes, (3) general process design considerations, (4) operational factors that affect EBPR processes, (5) EBPR process design, (6) provision for chemical addition, and (7) process control and performance optimization.

Process Development

From the mid-1960s to 1971 there were many observations of high levels of phosphorus removal at municipal activated sludge treatment plants at the District of Columbia (Levin and Shapiro, 1965), San Antonio, TX (Vacker et al., 1967), Los Angeles, CA (Bergman et al., 1970), and Baltimore, MD (Milbury et al., 1971). Phosphorus removal efficiencies were greater than 80 percent, compared to typical removals of 20 to 25 percent for activated sludge treatment of domestic wastewaters due to phosphorus needs for biomass growth from BOD removal. Levin and Shapiro (1965) presented experimental results on the occurrence of excess phosphorus uptake by biological activity and termed the process *luxury uptake* of phosphorus. They further proposed the sidestream *Phostrip* process for biological release of phosphorus from return activated sludge and subsequent chemical precipitation of the phosphorus in the separated liquid from the Phostrip process. Further designs were stymied at that time due to a lack of understanding of the basic causes and mechanisms behind the so called luxury uptake of phosphorus.

Barnard (1974) was the first to clarify the need for anaerobic contacting between activated sludge and influent wastewater before aerobic degradation to accomplish what is now called EBPR instead of luxury uptake. Other modifications of the basic process include (1) combining the anaerobic/aerobic sequence with various biological nitrogen-removal designs, (2) introducing mixed liquor to the anaerobic zone from a downstream anoxic zone instead of return activated sludge after liquid-solids separation in clarifiers or with membranes, (3) adding VFAs to the anaerobic zone as either acetate or a liquid stream from a fermentation reactor processing primary clarifier sludge, and (4) using multiple-staged anaerobic and aerobic reactors. The alternating exposure to anaerobic conditions can be accomplished in the main biological treatment process, or "mainstream EBPR," or in the return activated sludge stream or "sidestream EBPR". The first "mainstream" biological phosphorus removal process in the United States, a modified Bardenpho process, that also included biological nitrogen removal was placed in operation in 1979 at Palmetto, FL, and is shown on Fig. 8–3(b).

Overview of Enhanced Biological Phosphorus Removal Processes

A variety of EBPR process designs have been developed and applied at WWTPs; the design selection has depended on the wastewater treatment process and equipment in place before converting to EBPR, the wastewater characteristics, and the treatment needs. Usually EBPR is done in mainstream processes, but sidestream processes, such as Phostrip, may be of more interest today at WWTPs where struvite recovery systems are installed. Commonly used mainstream EBPR process configurations are shown on Fig. 8–37. The three different EBPR process configurations are used for applications in which,

1. Nitrification is not required
2. Nitrification is required with high BOD/P ratios in the influent wastewater
3. Nitrification is required with low BOD/P ratios in the influent wastewater

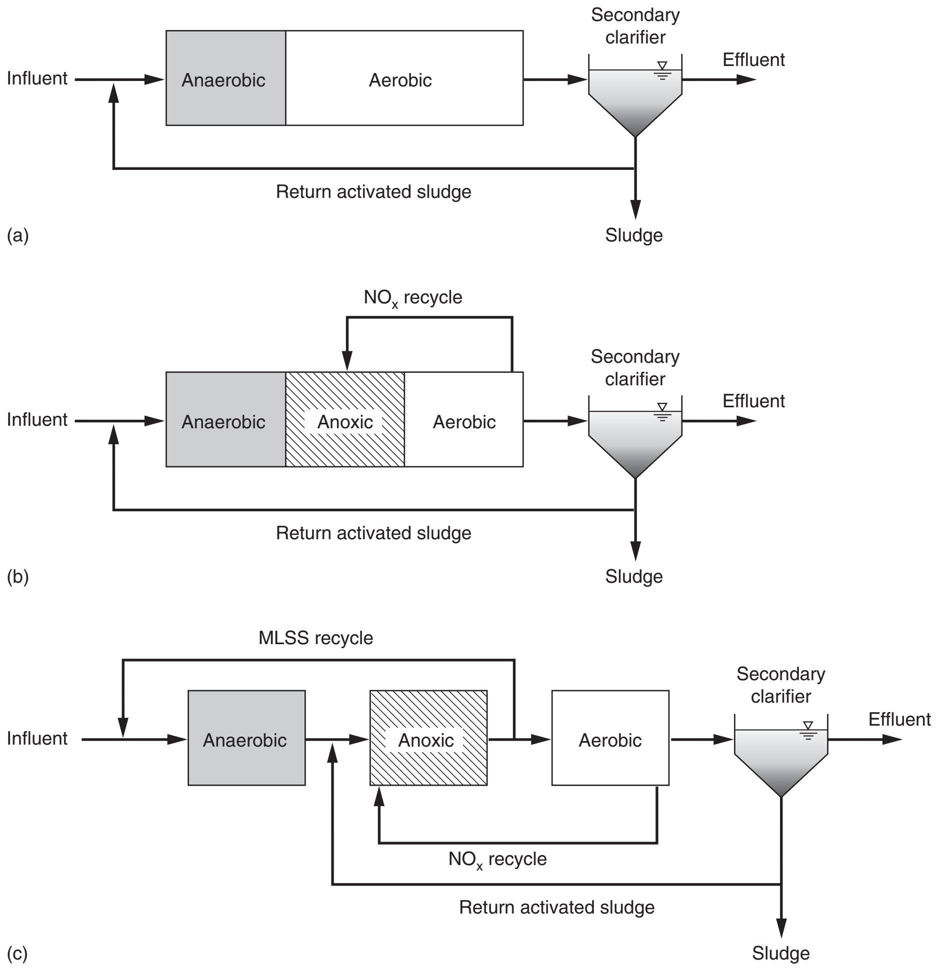
Each of these process configurations is considered in the following discussion along with a summary review of the available phosphorus removal processes and the sidestream removal of phosphorus.

Phosphorus Removal Without Nitrification. The term *Phoredox*, introduced by Barnard (1975), is used to represent any process with an anaerobic/aerobic sequence to promote EBPR. The Phoredox process, shown on Fig. 8–37(a), has also been referred to as an A/O (anaerobic/aerobic) process. As shown on Fig. 8–37(a), settled and thickened activated sludge is returned to the anaerobic reactor. Nitrification does not occur to any appreciable extent in the A/O process, as it is designed with low aerobic SRT values from 2 to 3 d at 20°C and 4 to 5 d at 10°C. These relatively low SRTs are adequate for growth and proper performance of the PAOs. Although EBPR has been done in activated sludge systems without nitrification, combined biological phosphorus and nitrogen removal processes are favored to minimize the negative impact of nitrate on EBPR performance and to reduce energy requirements.

Phosphorus Removal with Nitrification and High Influent Wastewater BOD/P Ratios. For applications where nitrification is needed to meet discharge requirements, nitrate can be present in the return activated sludge flow to the anaerobic contact zone of an EBPR process and can be detrimental to EBPR performance. Other non-PAO heterotrophic bacteria will use nitrate fed to the anaerobic zone to consume rbCOD for nitrate reduction, leaving less rbCOD available for PAOs. With less rbCOD available, there is less PAO growth and less phosphorus removal by EBPR. If no nitrate removal occurs with nitrification systems, EBPR is not possible.

Figure 8–37

Common enhanced biological phosphorus removal processes: (a) Phoredox (A/O), (b) Anaerobic/Anoxic/Aerobic (A^{2}O), (c) University of Capetown (UCT).



A very common process for nitrate removal is the Modified Ludzack-Ettinger (MLE) process as described in Sec. 8–7, and when preceded by an anaerobic contact zone for EBPR, it has been referred to as the A^{2}O (anaerobic-anoxic-aerobic) process as shown on Fig. 8–37(b). In the A^{2}O process, nitrate is removed by the anoxic/aerobic sequence shown, but the return activated sludge (RAS) recycle to the anaerobic zone still contains nitrate. If the wastewater has a high favorable influent BOD/P ratio (typically greater than 30/1) there will likely be enough rbCOD for both the nitrate removal and sufficient PAO growth. For weaker wastewaters or wastewaters with a lower BOD/P ratio, the amount of influent rbCOD consumed by heterotrophic denitrifying bacteria is such where there is not enough rbCOD left to support the amount of PAO growth needed for removal of the desired amount of influent phosphorus.

Phosphorus Removal with Nitrification and Low Influent Wastewater BOD/P Ratios. In some applications it is necessary to have very little or no nitrate in the mixed liquor recycle to the anaerobic contact zone, so that all of the rbCOD can be used by PAOs to meet the desired amount of EBPR. Avoidance of nitrate in mixed liquor returned to the anaerobic contact zone is desired for EBPR with weak wastewater, low influent BOD/P ratios and/or for a treatment goal to maximize phosphorus removal without chemical addition. The University of Capetown (UCT) process [see Fig. 8–37(c)] is a common EBPR process that has been used in anoxic/aerobic nitrogen removal processes to prevent the addition of nitrate to the anaerobic contact zone. In this process the return activated sludge is directed to an anoxic zone instead of the anaerobic contact zone. The return mixed liquor to the anaerobic contact zone is taken from the anoxic zone or effluent where the nitrate concentration is typically below 0.50 mg/L.

Sidestream Removal of Phosphorus. The Phostrip process [see Table 8–27(o)] is a sidestream EBPR process in which a portion of the return activated sludge flow is added to an anaerobic contact zone. In some cases a portion of the mainstream influent wastewater is diverted to the sidestream anaerobic contact zone to provide additional rbCOD. Growth of PAOs occurs in the anaerobic contact zone due to acetate uptake from fermentation of substrates released during cell lysis of the return activated sludge biomass and from rbCOD from wastewater, if added. Liquid-solids separation occurs, usually by gravity thickening as part of the anaerobic contact zone, and the liquid containing the released or *stripped* phosphorus is treated separately before it is returned to the mainstream process. The phosphorus-rich liquid is typically treated with lime or metal salts for phosphorus precipitation, but it could instead be added to a struvite recovery process.

The residual biomass is added back to the mainstream process with the return activated sludge flow where the PAOs take up phosphorus from the influent wastewater. The PAOs have a more efficient uptake of the influent phosphorus because they do not have to take up all of the phosphorus released in the anaerobic contact zone, as is done with the mainstream processes. In the Phostrip process, the phosphorus-removal efficiency depends less on the influent rbCOD concentrations than for other enhanced biological phosphorus removal processes. Also as noted previously, variants of the Phostrip sidestream process have gained popularity for the removal of phosphorus in the form of struvite (see Sec. 15–4 in Chap. 15).

Summary Review of EBPR Processes. A summary of the basic design features for EBPR processes, including those presented above are described in Table 8–27. They include (1) basic nitrogen removal processes that reduce the amount of nitrate fed to the EBPR process anaerobic contact zone [A^2O (b), Modified Bardenpho (d)], (2) processes to remove most all nitrates fed to the EBPR anaerobic contact zone [UCT (f), VIP (h), JHB (i,), Westbank (l)], (3) the processes in (1) and (2) using MBRs instead of an activated sludge/secondary clarifier system [(c), (e), (g), (j)], (4) processes including primary sludge fermentation for VFA production to maximize EBPR performance [VFA addition to A^2O-EBPR at Kelowna and to trickling filter-EBPR at OWASA (k), and Westbank (l)], (5) EBPR in an SBR process design (m, n), and (6) EBPR with sidestream anaerobic contacting [Phostrip (o)].

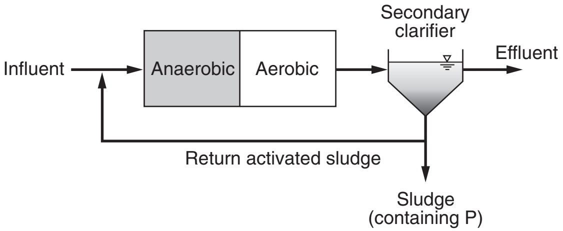
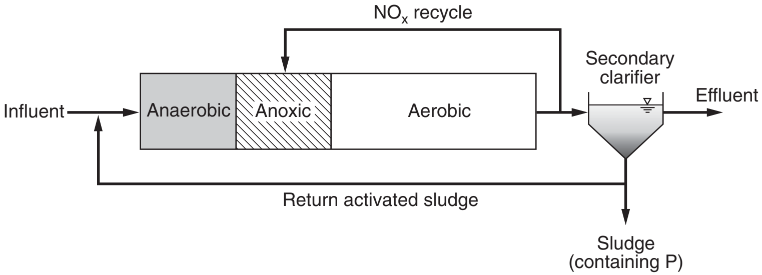
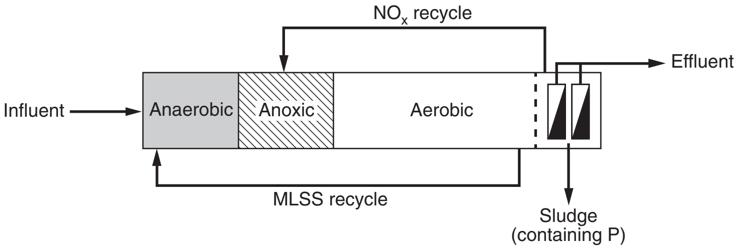
Process selection for EBPR is very site specific and a number of factors must be considered including (1) effluent phosphorus concentration needed, (2) existing process and equipment, (3) process needs for nitrification and nitrogen removal, (4) wastewater characteristics, (5) phosphorus recovery, and (6) operational requirements. Advantages and disadvantages for the most commonly used processes are given in Table 8–28 on page 871.

General Process Design Considerations

Most of the mainstream processes described for enhanced biological phosphorus removal incorporate the necessary anaerobic contacting between influent wastewater and activated

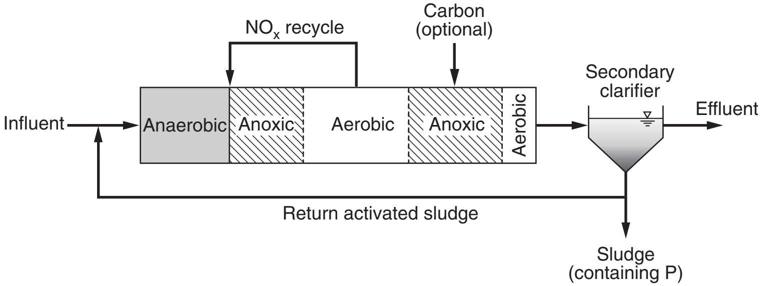
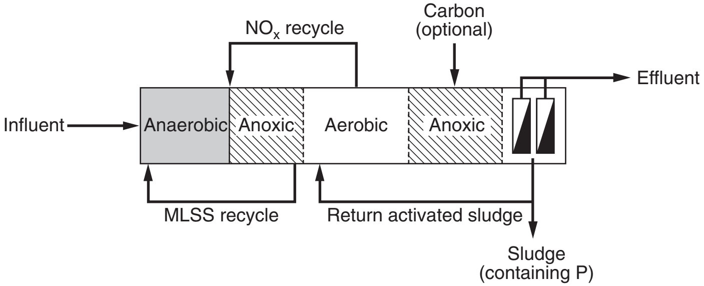
Table 8–27

Description of suspended growth processes for phosphorus removal

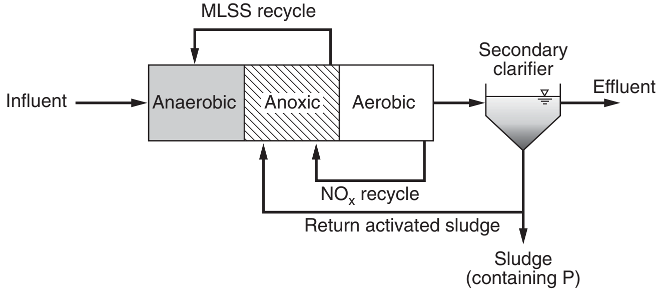
Process	**Description**
(a) Phoredox (A/O) Process 	The basic process configuration for biological phosphorous removal consists of an anaerobic zone followed by an aerobic zone. Barnard (1974) was the first to clarify the need for anaerobic contacting between activated sludge and influent wastewater before aerobic degradation to accomplish enhanced biological phosphorous removal. Barnard identified it as the Phoredox process and has also been referred to as the anaerobic/oxic (A/O) process. The SRT of the aerobic zone mixed liquor is 2 to 4 d, depending on the temperature, so that nitrification does not occur. The anaerobic contact time is 30 min to 1 h to provide the selective condition described in Sec. 7–13 for enhanced biological phosphorous removal.
(b) Anaerobic/Anoxic/Aerobic (A^2O) Process 	The A^2O process has an anoxic zone located between the anaerobic and aerobic zones and is used for EBPR systems that have nitrification. Nitrate is recycled from the aerobic zone to the anoxic zone for denitrification. The detention time in the anoxic zone is 1 to 3 h, depending on the wastewater characteristics and the amount of nitrate to remove. Use of the anoxic zone minimizes the amount of nitrate fed to the anaerobic zone in the return activated sludge.
(c) A^2O MBR Process 	A common approach for incorporating the A^2O process in an MBR is shown. Mixed liquor is recycled from the membrane separation aerobic zone to the anoxic zone for nitrate removal at a recycle ratio of about 6:1, based on the influent flowrate. Mixed liquor is recycled from the aerobic zone for contact with the influent wastewater in the anaerobic zone.

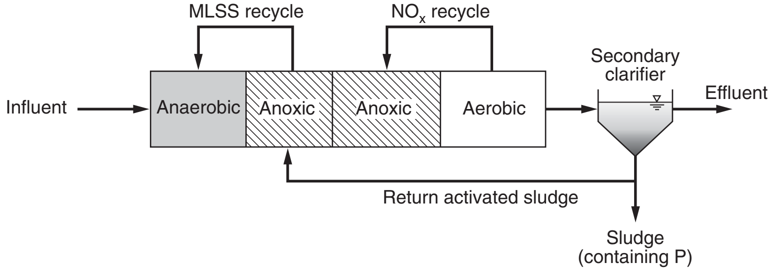
(*continued*)

Table 8–27 (*Continued*)

Process	Description
(d) Modified Bardenpho Process  	The Bardenpho process, described in Table 8–24, can be modified for combined nitrogen and phosphorus removal. The 5– stage system provides anaerobic, anoxic, and aerobic stages for phosphorus, nitrogen, and carbon removal. Mixed liquor from the first aerobic zone is recycled to the preanoxic zone. A second anoxic stage is provided for additional denitrification using nitrate produced in the first aerobic zone as the electron acceptor, and the endogenous organic carbon as the electron donor. An option is to add an exogenous carbon source to the second anoxic zone so that it has a shorter detention time and can produce lower effluent NO_3-N concentration. The final aerobic stage is used to strip residual nitrogen gas from solution and to raise the DO concentration to minimize phosphorus release in the secondary clarifier. The 5-stage process produces a lower NO_3-N concentration in the return activated sludge flow to minimize the effect of NO_3-N in the anaerobic zone.
(e) Modified Bardenpho -MBR Process  	A postanoxic zone is added to the A^2O–MBR process to create the modified Bardenpho–MBR process. Carbon addition to the second anoxic zone is also optional. Full-scale MBR plants using the modified Bardenpho–MBR process may employ three recycle streams. Mixed liquor from the aerobic membrane separation reactor is recycled to the aerobic reactor at a recycle ratio of about 6/1, based on the influent flowrate. NO_x is fed from the aerobic zone to the preanoxic zone at a rate of about 3 to 4 times the influent flowate. Mixed liquor is recycled from the preanoxic zone to the anaerobic zone at a rate equal to 1 to 2 times the influent flowrate. This recycle contains a near zero DO concentration and low NO_3-N concentration. The additional NO_3-N removal provides better EBPR performance than the A^2O-MBR process.

(f) University of Capetown (UCT) Process - Standard and Modified

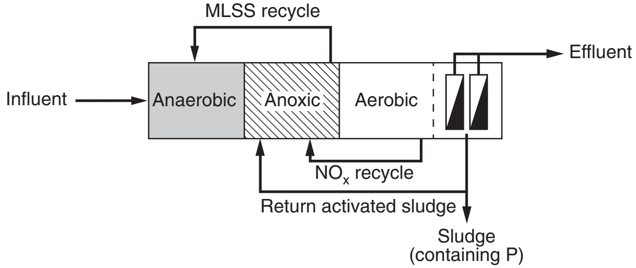




The UCT process was developed at the University of Cape Town (South Africa) to minimize the effect of nitrate entering the anaerobic contact zone in EBPR processes treating weak wastewaters. The UCT process has three recycle streams instead of two as in the A^{2}O process. The return activated sludge from the secondary clarifier is directed to the anoxic zone instead of the anaerobic zone. Similar to the A^{2}O process, internal recycle feed NO_x to the anoxic zone from the aerobic zone. The anaerobic zone receives mixed liquor from the anoxic zone instead of the return activated sludge flow so that the introduction of nitrate to the anaerobic stage is eliminated. Thus, more of the influent rbCOD is available for the PAOs in the anaerobic zone which can improve the EBPR efficiency. Because the anaerobic zone receives mixed liquor at a lower concentration than in the A^{2}O process, the anaerobic detention time must be longer than that used in the Phoredox process, and is the range of 1 to 2 h. The anaerobic recycle rate is typically 2 times the influent flowrate.

In the modified UCT process shown on the second diagram, the return activated sludge is directed to an anoxic reactor that does not receive internal nitrate recycle flow. The nitrate is reduced in this tank, and the mixed liquor from the reactor is recycled to the anaerobic tank. The second anoxic tank follows the first anoxic tank and receives internal nitrate recycle flow from the aeration tank to provide the major portion of nitrate removal for the process.

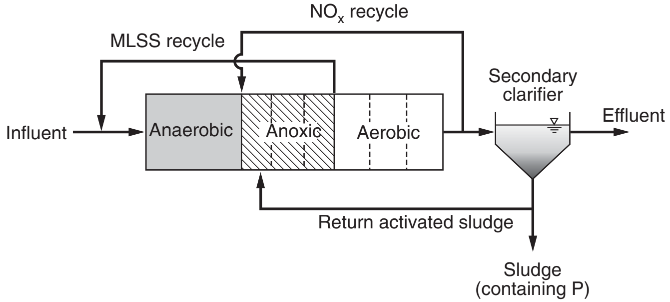
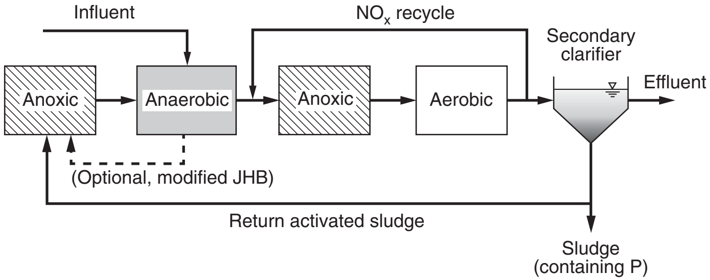
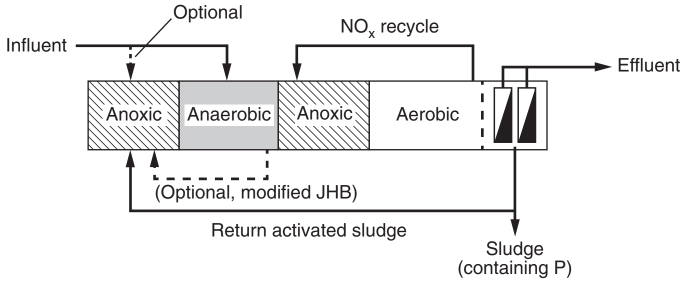
(g) Modified UCT-MBR Process



The modified UCT-MBR process also has three recycle streams, which are applied in the same manner as in the UCT process with secondary clarifiers. The return activated sludge flow from the aerobic membrane separation tank is directed to the preanoxic zone at a flowrate of about 6 times the influent flowrate. NOx produced in the aerobic zone is recycled back to the preanoxic zone at a flowrate of about 3 times the influent flowrate. The anaerobic zone receives mixed liquor with no dissolved oxygen and minimal nitrate in recycle from the preanoxic zone at a rate of 1 to 2 times the influent flowrate. By minimizing dissolved oxygen and nitrate entering the anaerobic contact zone, most of the influent rbCOD can be used by the PAOs to maximize EBPR efficiency.

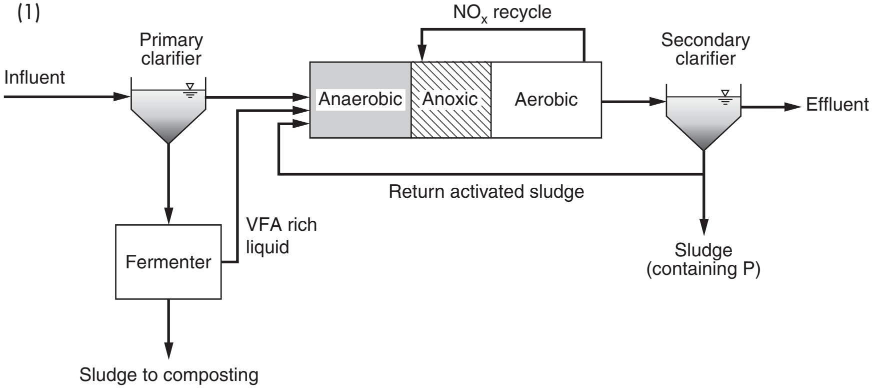
(*continued*)

Table 8–27 *(Continued)*

Process	Description
(h) Virginia Initiative Plant (VIP) Process  	The VIP process stands for the Virginia Initiative Plant (Daigger et al., 1988). The VIP process is similar to the A^2/O and UCT processes except for the increased staging and methods used for recycle systems. In the VIP process, all the zones are staged consisting of at least 2 completely mixed cells in series. The return activated sludge is discharged to the inlet of the anoxic zone along with nitrified recycle from the aerobic zone. The mixed liquor from the anoxic zone is returned to the head end of the anaerobic zone. The VIP process is also designed as a high rate system, operating with much shorter SRTs, which maximizes EBPR efficiency. The combined SRT of the anaerobic and anoxic zones is generally 1.5 to 3 d, while the anaerobic and anoxic τ values are typically 60 to 90 min each. The aeration zone is designed for nitrification.
(i) Johannesburg (JHB) Process  	The JHB process originated in Johannesburg, South Africa as an alternative to the UCT or modified UCT processes to minimize nitrate addition to the anaerobic zone to maximize EBPR for weak wastewaters. The return activated sludge is directed to an anoxic zone that has sufficient detention time to reduce the nitrate in the mixed liquor before it is fed to the anaerobic zone. The nitrate reduction is driven by the endogenous respiration rate of the mixed liquor, and the anoxic zone detention time depends on the mixed liquor concentration, temperature, and the nitrate concentration in the return sludge stream. NO_x produced in the aerobic zone is recycled to a preanoxic zone at a flowrate of 3 to 4 times the influent flowrate. Compared to the UCT process, a higher MLSS concentration and lower detention time can be maintained in the anaerobic zone (detention time of about 1 h) and two recycle streams are used instead of three.
(j) JHB-MBR Process  	The JHB process can also be Incorporated in an MBR system, also with the use of 2 recycle streams. Return activated sludge from the aerobic membrane separation zone is directed to an anoxic zone in which nitrate is reduced by the mixed liquor endogenous respiration. The high mixed liquor concentration in the MBR process results in a smaller anoxic tank volume than that required for the JHB process in an activated sludge/secondary clarifier system. An internal recycle flow from the aerobic zone to the anoxic zone is also used to remove NO_x produced in the aerobic zone.

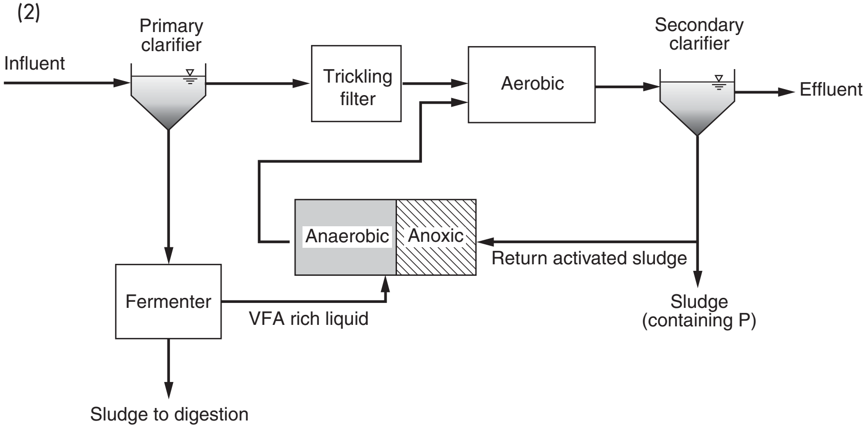
(k) EBPR with Primary Sludge Fermentation

(1)



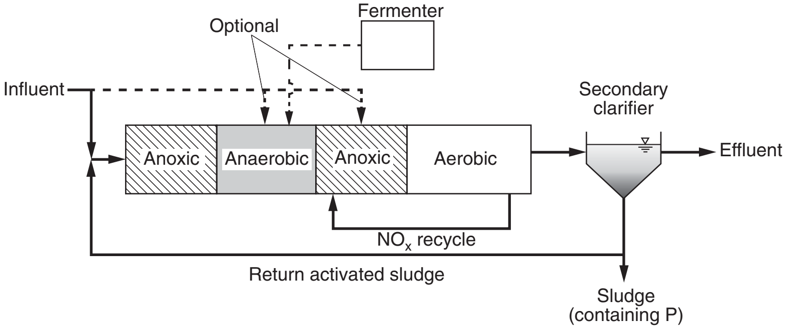
System (1) is used at the Kelowna WWTP and represents the more typical application. Fermenter liquid is added to the anaerobic zone of an EBPR process, which also receives the primary effluent. The additional VFAs in the fermenter liquid provides more substrate for PAO growth to improve phosphorus removal.

(2)



System (2) is used at the OWASA facility, which has trickling filter treatment prior to the activated sludge aeration tank. In this case the anaerobic contact zone necessary for EBPR receives the VFA-rich stream from the primary sludge fermenter in the return activated sludge flow, and is not after trickling filter effluent which has very little rbCOD.

(l) Westbank EBPR Process



The Westbank EBPR Process gets its name due to its location relative to another EBPR process used in Kelowna, British Columbia, Canada. The process anoxic/anaerobic/anoxic/aerobic zone sequence and recycle flow locations are similar to that used in the JHB process. However, the first anoxic zone receives influent wastewater in addition to the return activated sludge flow to assure complete nitrate removal. Some influent may also be added to the downstream anaerobic and anoxic zones. The subsequent anaerobic zone process is fed liquid from a primary sludge fermentation tank that is rich in volatile fatty acids (acetate and propionate) to provide substrate for PAO growth and efficient EBPR performance.

(*continued*)

Table 8–27 *(Continued)*

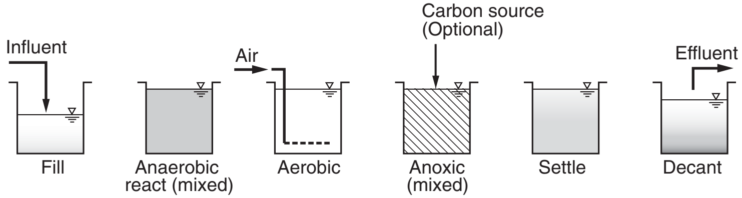
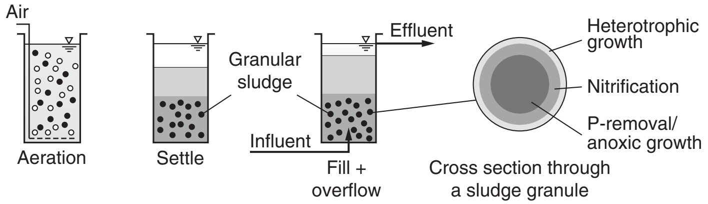
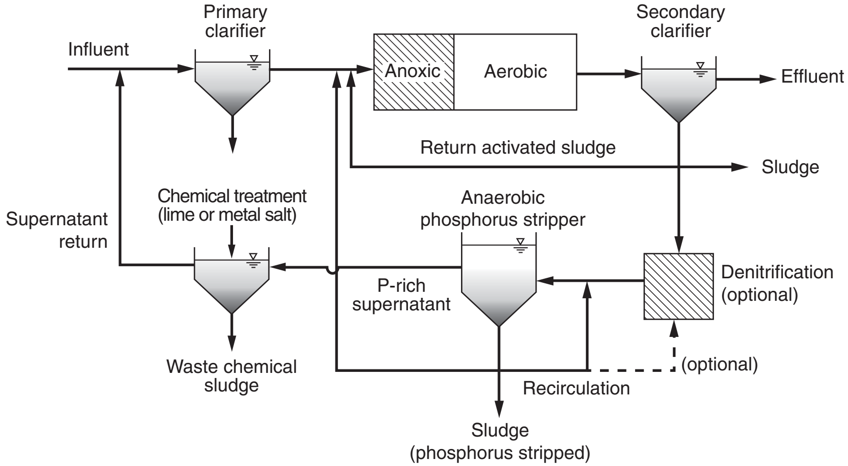
Process	Description
(m) SBR with EBPR  	If sufficient nitrate is removed during the SBR operation, an anaerobic reaction period can be developed during the SBR fill period for rbCOD fermentation and VFA uptake by PAOs. An anoxic operating period is used after a sufficient aerobic time elapses for nitrification and nitrate production. Alternatively cyclic aerobic and anoxic periods can be used during the react period. The nitrate concentration is thus minimized before settling, and little nitrate is available to compete for rbCOD in the fill and initial react period. Thus, anaerobic conditions occur in the fill and initial react period, so that rbCOD can be used by PAOs instead of by nitrate-reducing bacteria.
(n) EBPR in SBR with granular activated sludge  	An SBR system with granular activated sludge is a promising emerging technology. A selector process based on upflow velocity produces dense, granular activated sludge aggregates ($>$0.20 mm diameter) (de Kreuk and van Loosdrecht, 2007) that have the same degree of settling in 5 min as that for 30 min in an SVI test with conventional activated sludge. As a result the SBR design and operation is different from a conventional SBR. After a short settling time ($<$10 min) the influent is added upflow with the effluent being the treated upper zone liquid after settling. The fill period is anaerobic to promote EBPR. During the aeration period SNdN occurs with nitrification in the outer layer of the granules and denitrification in the inner layers. Stored carbon is oxidized by the PAOs with phosphorus uptake. The aeration efficiency is improved due to the constant liquid level and the uptake of rbCOD in the anaerobic fill results in a more uniform oxygen uptake rate during aeration. Effluent filtration may be needed for effluent TSS concentrations less than 15 mg/L. A commercial process named Nereda™ has been developed by DHV with full-scale facilities in the Netherlands.
(o) Phostrip with anoxic/aerobic activated sludge treatment  	The Phostrip process is in essence an anaerobic/aerobic EBPR process with stripping of phosphorus from the PAOs. The anaerobic condition is created by holding return activated sludge long enough in a gravity thickener (termed a *phosphorus stripper*) with residence times generally in the range of 8 to 12 h (Levin et al., 1975). Substrates released from sludge endogenous decay and lysis are fermented to VFAs for uptake by the PAOs resulting in phosphorus release. The released phosphorus is elutriated by circulation of underflow from the thickener to the thickener inlet. The thickener underflow solids are returned to the activated sludge process, where the PAOs take up phosphorus under anoxic and aerobic conditions. The overflow from the stripper tank is then treated chemically with lime or metal salts for phosphorous removal. The lime dose needed to raise the pH for phosphorus removal is a function of the wastewater alkalinity and not the amount of phosphorus present. If alum and ferric salts are used instead, the dose is proportional to the amount of phosphorus released but it can be used at minimal molar ratios for phosphorus removal. The Phostrip process can be similarly applied in systems without nitrification. For anoxic/aerobic processes the recycle may be held in an anoxic reactor for denitrification prior to its addition to the stripper tank.

Table 8–28

Advantages and limitations of phosphorus removal processes

Process	Advantages	Limitations
Phoredox (A/O)	Operation is relatively simple when compared to other processes Low influent BOD/P ratio possible Relatively short hydraulic retention time Produces good settling sludge Good phosphorus removal	Phosphorus removal declines if nitrification occurs Requires careful SRT control to prevent nitrification Difficult to operate without nitrification in warm climates
A^2O	Removes both nitrogen and phosphorus Provides alkalinity for nitrification Produces good settling sludge Operation is relatively simple Saves energy Compatible with MBRs	RAS containing nitrate is recycled to anaerobic zone thus affecting phosphorus removal capability Nitrogen removal is limited by internal recycle ratio Needs higher influent BOD/P ratio than the A/O process
UCT	Nitrate loading on anaerobic zone is reduced thus increasing phosphorus removal capability For weaker wastewater, process can achieve improved phosphorus removal Produces good settling sludge Good nitrogen removal Compatible with MBRs	More complex operation Requires additional recycle system
VIP	Nitrate loading on anaerobic zone is reduced thus increasing phosphorus removal capability Produces good settling sludge Requires lower BOD/P ratio than UCT Good nitrogen removal Compatible with MBRs	More complex operation Requires additional recycle system More equipment, baffles, and headloss required for staged design
Modified Bardenpho	Can achieve 3 to 5 mg/L TN in unfiltered effluent Produces good settling sludge Less nitrate fed to anaerobic zone than for A^2O Compatible with MBRs	Less efficient phosphorus removal due to longer SRT Requires larger tank volumes
JHB	Prevents nitrate feed to anaerobic zone Promotes high EBPR efficiency Can be used with A^2/O, Modified Bardenpho, Westbank, and Steppe processes Compatible with MBRs	Requires more tank volume Increased operational complexity
SBR with EBPR	Both nitrogen and phosphorus removal are possible Provides good settling sludge Mixed liquor solids cannot be washed out by hydraulic surges Quiescent settling may produce lower effluent TSS concentration Flexible operation Simple system layout Good plug flow kinetics for P uptake efficiency	More complex operation for N and P removal Needs larger volume than SBR for N removal only Effluent quality depends upon reliable decanting facility Design is more complex Skilled maintenance is required More suitable for smaller flowrates

(*continued*)

Table 8–28 (*Continued*)

Process	Advantages	Limitations
SBR with granular activated sludge	Both nitrogen and phosphorus removal are possible Excellent settling sludge Less space than conventional activated sludge Simple system operation Good plug flow kinetics Good aeration efficiency with constant liquid depth	Limited long-term experience Higher effluent suspended solids concentration May not be compatible with chemical addition for phosphorus removal High peak hydraulic loads not desired
Westbank	High EBPR efficiency Operational flexibility Stable EBPR performance Less susceptible to wastewater loading variations Compatible with MBRs	Needs extra tanks Increased unit process/increased operations Odor control for fermenter
Phostrip	Can be incorporated easily into existing activated sludge plants Process is flexible; phosphorus removal performance is less controlled by influent BOD/phosphorus ratio Less chemical usage than mainstream chemical precipitation process Can achieve reliable effluent orthophosphate concentrations less than 1 mg/L	Requires metal salts or lime addition for phosphorus precipitation Additional tanks needed Increased operations Increased sludge production Lime scaling may be a maintenance problem

sludge, followed by an anoxic or aerobic zone for biooxidation of stored intracellular poly-β-hydroxyalkanoates (PHA) and phosphorus uptake by the PAOs. In light of these basic requirements, process design considerations for EBPR processes include (1) treatment needs, (2) influent wastewater characteristics, (3) anaerobic contact time, (4) supplemental sources of VFAs, (5) SRT, and (6) aerobic zone and aeration design. Operational factors including (1) the effect of nitrate and oxygen addition to the anaerobic contact zone, (2) waste sludge processing, and (3) alkalinity and pH, are considered separately. The need for the provision of chemical addition is also considered separately following the presentation of the EBPR process design. The reaction kinetics for BOD removal, nitrification, and denitrification are similar to those discussed in Secs. 8–4 and 8–5. Typical parameters used in the design of EBPR processes are presented in Table 8–29.

Treatment Needs. The specific process design details are affected by effluent nitrogen concentration requirements as well as the required effluent phosphorus concentration. In addition to the process configuration selection, the treatment needs affect the design SRT, reactor staging and DO concentrations, the need for carbon addition and waste sludge handling methods.

Wastewater Characteristics. Because EBPR is affected by the influent wastewater characteristics, including rbCOD and VFAs, the measurement of these

Table 8–29

Typical design parameters for commonly used biological phosphorus removal processes

Design parameter/ process	SRT, d	MLSS, mg/L	τ, h Anaerobic zone	τ, h Anoxic zone	τ, h Aerobic zone	RAS, % of influent	Internal recycle, % of influent
A/O	2–5	3000–4000	0.5–1.5	–	1–3	25–100	
A^2/O	5–25	3000–4000	0.5–1.5	1–3	4–8	25–100	100–400
Modified Bardenpho	10–20	3000–4000	0.5–1.5	1–3 (1st stage) 2–4 (2nd stage)	4–12 (1st stage) 0.5–1 (2nd stage)	50–100	200–400
UCT	10–25	3000–4000	1–2	2–4	4–12	80–100	200–400 (anoxic) 100–300 (aerobic)
VIP	5–10	2000–4000	1–2	1–2	4–6	80–100	100–200 (anoxic) 100–300 (aerobic)
SBR	20–40	3000–4000	1.5–3	1–3	2–4		
Phostrip	5–20	1000–3000	10–12		4–10	50–100	10–20

constituents is essential to evaluate the design and performance of EBPR systems. The amount of enhanced biological phosphorus removal is related directly to the amount of acetate and propionate taken up by PAOs in the anaerobic contact zone and converted to carbon storage products that provide energy and growth in the subsequent anoxic and aerobic zones (Gerber et al., 1986). Different influent wastewater characterization parameters have been used to assess the potential for EBPR performance. The wastewater characteristics shown in Table 8–30 are arranged in order of the most direct to indirect correlation with VFA utilization in EBPR systems, to achieve an effluent soluble P concentration ≤ 0.50 mg/L. Because acetate and propionate are the substrates consumed by PAOs in the anaerobic contact zone, a VFA to P ratio is a good predictor of the amount of P that can be removed. The amount of VFA available to the PAOs in the anaerobic

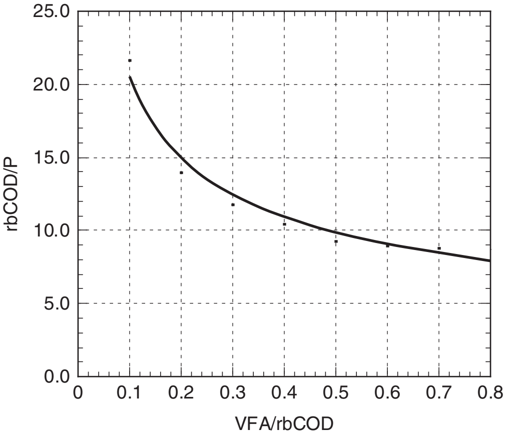
Table 8–30

Minimal influent wastewater ratios for achieving a soluble P effluent concentration of less than 0.50 mg/L in EBPR systems.

Influent substrate parameter	Value	Reference
VFA:P	8	Wentzel (1990)
rbCOD:P	18	Barnard (2006)
BOD:P	30	Sedlak (1991)
COD:P	60	U.S. EPA (2010)

Figure 8–38

Amount of rbCOD used for enhanced biological phosphorus removal is related to its volatile fatty acid content.



contact zone is greater than the influent wastewater VFA concentration due to fermentation of other rbCOD in the wastewater.

Influent rbCOD to P Ratio. Knowledge of the influent rbCOD/P ratio is very useful. A minimum value of 18 is recommended to achieve a soluble P concentration of less than 0.50 mg/L, but as the influent VFA/rbCOD ratio increases, the required rbCOD/P ratio decreases, as shown on Fig. 8–38 (Barnard, 2012). In some cases, the potential P removal must be estimated with only influent BOD or COD data. Higher BOD/P ratios for domestic wastewater generally lead to improved EBPR performance. A BOD/P ratio above 30 will generally result in high EBPR performance with effluent soluble P concentrations below 0.50 mg/L. The ratios given above are based on the assumption that the EBPR process does not have an excessively long SRT and that minimal nitrates are fed to the anaerobic contact zone. Higher values for the ratios shown in Table 8–30 will be required for EBPR processes with long SRTs (>15 to 20 d) and/or with high DO or nitrate feed to the anaerobic zone.

Diurnal Variation. The diurnal variation in wastewater strength is also an important process consideration. Because the performance of phosphorus-storing bacteria depends on the availability of fermentation substrates, it is important to know if periods of low influent wastewater strength may affect EBPR performance. For domestic wastewaters, the influent total BOD and rbCOD concentrations will vary with time over a 24-h period, with lower concentrations in the late evening and early morning hours. For smaller-sized communities, the variations are usually more pronounced and very little rbCOD may be present at certain times. During wet-weather conditions, especially in the winter, EBPR may be difficult to achieve due to cold, low strength wastewater that does not readily become anaerobic.

Extended periods of reduced rbCOD concentration will decrease EBPR performance for a number of hours after the occurrence of low substrate concentration (Stephens and Stensel, 1998). The impact of continuous acetate feeding at plants where sludge fermentation has been done to produce additional VFAs has shown the benefit of a steady supply of rbCOD for EBPR. In parallel modified Bardenpho trains at Kelowna, Canada, one train was fed fermentation liquor and the other train was used as the control. With continuous VFA addition, the effluent soluble phosphorus concentration decreased from 2.5 to 0.3 mg/L (Oldham and Stevens, 1985), and the incremental VFA/P ratio was 6.7 g/g, an amount lower than the 8 g/g ratio in Table 8–30. Based on these results, it appears that continuous acetate addition provides more efficient EBPR.

Anaerobic Contact Time. Anaerobic contact by PAOs and the availability of VFA is the most critical elements of an EBPR process design as described in Sec. 7–11. Acetate and propionate are taken up rapidly and detention times of 0.25 to 1.0 h are adequate for fermentation of rbCOD. To account for the effect of the MLVSS concentration in the anaerobic contact zone, a 1-d SRT is recommended for the anaerobic contact zone design (Grady et al., 1999). If the anaerobic contact time is too long, a *secondary release* of phosphorus can occur, which is not associated with acetate uptake (Barnard, 1984). When secondary phosphorus release occurs, the PAOs have not accumulated polyhydroxyalkanoates (PHA) for subsequent oxidation in the aerobic zone. Polyhydroxybutyrate provides energy for phosphorus uptake and storage. From SBR bench-scale studies, it has been found that secondary phosphorus release occurred for anaerobic contact times in excess of 3.0 h (Stephens and Stensel, 1998).

Supplemental Volatile Fatty Acid. The performance of EBPR systems is very site-specific and depends on the wastewater characteristics and the plant process design and operation. For wastewaters with relatively low influent rbCOD/P concentration ratios, effluent soluble phosphorus concentrations may exceed 1.0 to 2.0 mg/L, whereas effluent concentrations below 0.5 or 1.0 mg/L have been achieved with higher-strength wastewaters. The addition of VFAs to the anaerobic contact zone can result in effluent soluble P concentrations from 0.10 to 0.20 mg/L on a consistent basis, as shown for example at the Westbank BC (Rabinowitz and Barnard, 1996) and Durham, Oregon EBPR facilities (Stephens, 2004).

Additional Sources of VFAs. Additional VFAs can be obtained by importing (purchase) an acceptable exogenous carbon source or by production on site by fermentation of primary clarifier waste sludge. Primary sludge fermentation processes produce VFAs with about 50 percent acetate and 30 percent propionate. The presence of propionate is favorable for the selection of PAOs over glycogen-accumulating organisms (GAOs). Primary sludge fermentation process configurations are shown on Fig, 8–39, and key design factors are summarized in Table 8–31. The process configurations shown on Figs. 8–39(b), (c), and (d) are used at Penticton, B.C.; Kelowna, B.C.; and Kalispell, MT., respectively to achieve effluent P concentrations ranging from 0.10 to 0.20 mg/L. The primary clarifier tank (PCT) fermenter [see Fig. 8–39(a)] is the simplest design (Barnard, 1984), but there are problems associated with the high sludge blanket that must be maintained including the potential carryover of suspended solids to the secondary treatment process. The simplest and most common approach is to use primary sludge gravity thickeners for VFA production [see Fig. 8–39(c)].

Primary Sludge Fermenters. Primary sludge fermentation can occur in a separate unheated tank, clarifier sludge blanket, and gravity sludge thickener. Sludge SRTs in the range of 3 to 6 d, depending on temperature, are used to prevent methanogenic activity, which would result in consumption of the VFAs produced (Rabinowitz and Oldham, 1985). At these SRT values, the VFA production ranges from 0.1 to 0.2 g VFA/g VSS applied to the fermenters. The fermenter liquid VFA concentration can range from 150 to 300 mg/L, depending on the design and operation. Assuming an untreated wastewater influent VSS concentration of 200 mg/L with 65 percent VSS removal in primary clarification, the VFA production potential is 13 to 26 mg/L (normalized to the influent flow). The potential increase in enhanced biological phosphorus removal is 2 to 4 mg/L for this amount of VFA production. However, not all of the VFAs produced are sent to the anaerobic contact zone as a portion of the VFAs are contained in the waste sludge flow from the fermenter.

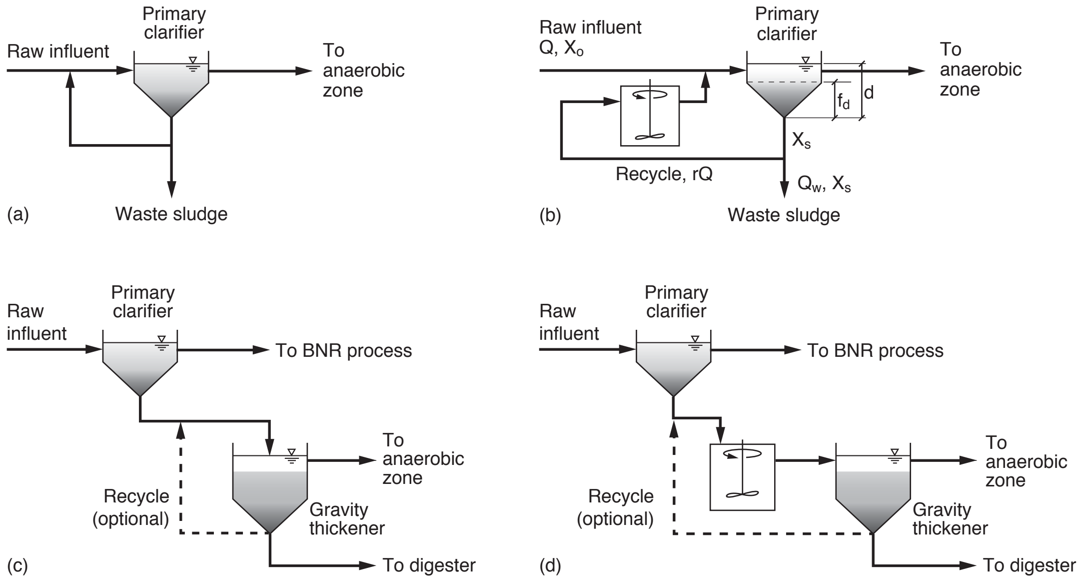


Figure 8–39

Primary sludge fermentation designs for supplying volatile fatty acids used for enhanced biological phosphorus removal: (a) primary sedimentation deep tank sludge blanket fermenter, (b) mixed fermenter tank/primary sedimentation (c) gravity thickener fermenter, and (d) mixed fermenter tank/gravity thickener.

Recycle streams around primary sludge fermenters are used to elutriate VFAs in thickened sludge processes on Fig. 8–39(a) and (b) and are optional in processes on Fig. 8–39(c) and (d). For sludge thickener fermenters, a more dilute waste primary sludge is fed to the gravity thickener, so that a higher ratio of the thickener effluent flow is in the overflow to the anaerobic contact zone, which results in feeding a higher fraction of the VFA produced to the EBPR system. Design criteria for primary sludge fermentation are reported in Table 8–31.

Estimating SRT Values and VFA Production Rates. The SRT and the VFA production rates can be estimated for primary sludge fermenters by performing mass balances on waste sludge and recycle flows and VSS concentrations, and accounting for

Table 8–31

Design criteria for primary sludge fermentation

Type of primary sludge fermentation[a]	Sludge SRT, d	Elutriation recycle ratios: Raw influent	Elutriation recycle ratios: Thickener influent	Contact tank τ, d	Thickener feed, fraction of raw influent flow	Thickener loading, kg/m²·d
(a) Activated PST[b]	2–4	0.05–0.10				
(b) Completely mixed fermenter/PST[b]	4–6	0.05–0.10		0.25–0.50		
(c) Gravity thickener fermenter	4–6		0.10–0.20		0.04–0.08	20–40
(d) Two-stage fermenter/thickener	4–6		0.30–0.50	2.0–4.0	0.02–0.04	100–150

[a] Refer to Fig. 8–39.
[b] PST – primary sedimentation tank.

sludge fermentation tank contact times. For the fermenter/PST combination system shown on Fig. 8–39(b), the sludge fermentation total SRT is determined as follows:

$$\mathrm{SRT} = \mathrm{SRT}_{\text{mixed tank}} + \mathrm{SRT}_{\text{sludge blanket}} \tag{8–77}$$

$$\mathrm{SRT} = r(T_f)t_c + \frac{f_d(d)(T_f)}{\mathrm{SOR}}$$

$$T_f = \frac{Q}{Q_w} \tag{8–78}$$

where r = primary clarifier tank recycle flow ratio relative to influent flowrate
Q = influent flowrate to the PST, $\mathrm{m^3/d}$
Q_w = PST underflow waste sludge flowrate, $\mathrm{m^3/d}$
T_f = influent VSS thickening factor, PST underflow solids concentration to influent flow solids concentration
t_c = detention time of mixed contact tank based on rQ, d
f_d = fraction of PST sludge blanket depth
d = PST depth, m.
SOR = PST surface overflow rate, $\mathrm{m^3/m^2{\cdot}d}$

Operating issues of odors, mixing, and accumulation of rags in the fermenters must also be considered in design and operation requirements.

Exogenous Carbon. Exogenous carbon has been obtained by purchasing acetate or by obtaining wastes from food processing, including corn syrup, brewery and sugar containing beverage processing, fruit and vegetable canning, and molasses waste. Care must be used in adding these wastes so as not to develop an abundant GAO population that could later affect PAO EBPR efficiency during operational changes. Glycerol and ethanol have also been evaluated in bench-scale studies for addition to anaerobic contact zones to support EBPR. They are both proven feasible, but ethanol requires a lengthy acclimation time, presumably to develop ethanol fermenters (Guerrero et al., 2012), and glycerol requires a longer than commonly used anaerobic contact to facilitate glycerol fermentation to acetate and propionate (Puig et al., 2008).

Fermentation of Activated Sludge Mixed Liquor. Fermentation of activated sludge mixed liquor has also promoted EBPR, and was the basis of the Phostrip process proposed in 1965 (Levin and Shapiro, 1965). In that process, thickener supernatant was removed for chemical treatment of the released phosphorus, and the mixed liquor is returned to the mainstream treatment for phosphorus uptake in the aerobic zone. However, Barnard et al. (2011) have reported effective EBPR removals with activated sludge fermentation without the Phostrip process. Fermentation of about 7 percent of the return activated sludge flow in anaerobic reactors at 35 to 50 h retention times has resulted in effective EBPR in a number of facilities in Denmark and England.

Solids Retention Time. Enhanced biological phosphorus-removal efficiency is affected by the activated sludge process type, the design SRT, and the influent wastewater characteristics (Randall et al., 1992). A sufficient aerobic SRT is needed for P uptake for an effective EBPR system. An SRT of ≥ 2.5 d is needed at 20°C and ≥ 4.0 d at 10°C. Biological nutrient-removal systems with longer SRTs are less efficient for EBPR than shorter SRT designs. Two adverse effects on phosphorus removal efficiency are associated with lightly loaded, long SRT processes. First, at longer SRT values more endogenous decay occurs and the amount of PAO biomass produced and wasted is less, with a proportional reduction in P removal. Second, at long SRT values the PAOs are in a more extended endogenous phase, which will deplete more of their intracellular storage products. If the intracellular glycogen is depleted, less efficient acetate uptake and PHA storage will occur in the anaerobic contact zone, thus making the overall EBPR process less efficient (Stephens and Stensel, 1998). Operation at an SRT just past that needed for nitrification appears to be the most optimal SRT for best EBPR performance (Oldham and Stevens, 1985).

Aerobic Zone and Aeration Design. The aeration zone configuration and DO concentration affects EBPR phosphorus uptake efficiency and effluent soluble P concentrations. A DO concentration ≥ 1.5 mg/L is needed in the 1st stage of a staged aerobic zone or the P removal efficiency will be compromised (Narayanan et al., 2011). Higher phosphorus removal efficiency occurs with staged aerobic zones compared to a single aerobic reactor tank due to the first-order reaction kinetics of phosphorus uptake (Petersen et al., 1998). The phosphorus uptake rate is a function of the stored PHA concentration, bulk liquid DO and PO_4 concentrations, and PAO biomass concentration.

A staged aerobic reactor has higher stored PHA and PO_4 concentrations in the upstream stages which will then have faster phosphorus uptake rates leading to much lower PO_4 concentrations in the final stage. The rate of PO_4 uptake is typically modeled with an expression such as given by Eq. (8–79):

$$R_p = q_{pp}\left(\frac{S_{O_2}}{K_{O_2} + S_{O_2}}\right)\left(\frac{S_{PO_4}}{K_{PO_4} + S_{PO_4}}\right)\left[\frac{X_{PHA}/X_{PAO}}{K_{PHA} + (X_{PHA}/X_{PAO})}\right]\left[\frac{K_{max} - (X_{pp}/X_{PAO})}{K_{ipp} + K_{max} - (X_{pp}/X_{PAO})}\right]X_{PAO} \tag{8–79}$$

where R_p = rate of PO_4 uptake, g/m^3·d
q_{pp} = maximum PO_4 uptake rate, g/m^3·d
S_{O_2} = DO concentration, g/m^3
K_{O_2}, K_{PO_4}, K_{PHA} = half-velocity coefficients, g/m^3
S_{PO_4} = PO_4 concentration, g/m^3
X_{PHA} = stored PHA concentration, g/m^3
X_{PAO} = phosphorus accumulating organisms concentration, g/m^3
X_{pp} = stored phosphorus concentration in PAOs, g/m^3
K_{max} = maximum specific phosphorus storage in PAOs, g/g
K_{ipp} = unreleasable inert phosphorus in PAOs, g/g

Operational Factors That Affect Enhanced Biological Phosphorus Removal

In addition to the influent wastewater characteristics and the other factors discussed above, there are a number of operational factors that can affect the performance of EBPR processes including the impact of nitrate and oxygen addition to the contact zone, the sludge wasting process, and the process alkalinity and pH. These factors are considered in the following discussion.

Effect of Nitrate and Oxygen Addition to Anaerobic Contact Zone. Nitrate and DO in upstream and recycle flows to the anaerobic contact zone must be avoided where possible. Filter backwash recycle flows contain DO and should be sent to the aerobic zone instead of the anaerobic or anoxic zones. Recycle streams with significant concentrations of DO and nitrate can have an adverse impact on process performance. Other heterotrophic bacteria, as noted previously, can consume rbCOD in the anaerobic contact zone using DO or NO_3-N for electron acceptors, leaving less rbCOD to support growth of PAOs for effective EBPR performance.

The amount of rbCOD consumed by NO_3-N and DO entering the anaerobic contact zone can be determined from Eqs. (7–127) and (7–129), in which the synthesis yield is used in place of the net yield. The synthesis yield given in Table 8–14 is 0.45 g VSS/g COD with oxygen as the electron acceptor. With nitrate as the electron acceptor, it is estimated at 70 percent of that with oxygen (Muller et al., 2003) or 0.32 g VSS/g COD. Thus, the ratio of rbCOD consumed to nitrate-nitrogen or oxygen made available in the anaerobic contact zone is shown as follows:

Nitrate: (Eq. 7–127)

$$\frac{\text{g rbCOD}}{\text{g NO}_3\text{-N}} = \frac{2.86}{1 - 1.42(0.32 \text{ g VSS/g COD})} = 5.2$$

Oxygen: (Eq. 7–129)

$$\frac{\text{g rbCOD}}{\text{g O}_2} = \frac{1}{1 - 1.42(0.45 \text{ g VSS/g COD})} = 2.8$$

Based on the above rbCOD/NO_3-N and rbCOD/DO ratios, the impact of DO and nitrate fed into the anaerobic contact zone on the EBPR performance can be evaluated. The rbCOD in the influent wastewater added to the anaerobic zone will most likely be removed by bacteria using oxygen and nitrate before it is available for biological phosphorus removal.

Waste Sludge Processing. Because phosphorus is contained in the sludge wasted from EBPR processes, consideration must be given to the waste sludge processing methods and the potential to recycle excessive amounts of phosphorus back to the EBPR process. Further, it has been found that anaerobic digester biosolids from waste sludge from EBPR systems exhibit poorer dewatering properties including lower cake solids and higher polymer use. The cause of the dewatering problems may be related to an increase in the ratio of monovalent to divalent cations as reported by Murthy et al. (1998). In EBPR digester sludge the ratio of monovalent to divalent cations can increase because there is a higher proportion of potassium as compared to calcium and magnesium with PAO phosphorus release under anaerobic conditions.

Impact of Anaerobic Conditions. In the review of the EBPR mechanism in Chap. 7, it was noted that phosphorus is released when the bacteria that contain stored phosphorus are subject to anaerobic conditions. Thus, anaerobic conditions in thickening and/or digestion can result in the release of significant amounts of phosphorus. The recycle stream from these processes would, in essence, increase the influent phosphorus concentration so that a greater amount of influent rbCOD is needed to produce the same effluent phosphorus concentration that would have been possible without the recycle stream. An additional problem is that solids dewatering may occur on an intermittent schedule with the release of large slugs of phosphorus in the recycle stream. Strategies for controlling the effect of recycled streams on EBPR performance include (1) select waste sludge processing methods that have minimal phosphorus release, (2) manage the return flow loadings by flow and load equalization, (3) use chemicals for phosphorus removal by precipitation and (4) use a phosphorus recovery process on anaerobic digester centrate/filtrate (struvite recovery processes are discussed in Chaps. 6 and 15).

Minimizing Phosphorus Release. Thickening of waste sludge by dissolved air flotation, gravity belt thickeners, or rotary-drum thickeners is preferred over gravity thickening of waste sludge to minimize phosphorus release. Phosphorus release occurs as biological solids are destroyed in anaerobic and aerobic digestion processes but only 20 to 40 percent of the phosphorus contained in the waste sludge has been observed in return streams. Not all of the phosphorus taken up is released and the formation of phosphorus precipitates, such as struvite and brushnite, has been credited with keeping phosphorus at lower concentrations. Phosphorus recovery technology is of interest to not only minimize the phosphorus content of return flows but to provide for reuse of a finite resource. Direct land application of liquid, digested sludge or dewatered raw sludge followed by stabilization such as composting also minimizes recycled phosphorus loads.

Alkalinity and pH. As described in Chap. 7 (Sec. 7–13) pH values greater than 7.0 can favor PAO dominance in their competition with GAOs for the acetate provided by the influent wastewater and that produced in the anaerobic contact tank. To assure that a favorable pH is available for efficient EBPR performance, an alkalinity balance, accounting for the influent alkalinity concentration, alkalinity consumed by nitrification, and alkalinity produced by denitrification, should be done. Alkalinity addition may be considered, especially for operating temperatures between 20 and 30°C, where GAOs are more competitive.

Enhanced Biological Phosphorus Removal Process Design

Key elements of process design for EBPR beyond what is involved for nitrogen removal are the anaerobic contact time, the aerobic SRT, configuration, DO concentration, and the amount of rbCOD available phosphorus uptake and removal by the PAOs. The first two items have been addressed above and in Sec. 8–7 for nitrogen removal. The importance of rbCOD and the effect of NO_3-N on phosphorus removal are illustrated in Example 8–13.

EXAMPLE 8–13 **Effect of Nitrate on Enhanced Biological Phosphorus Removal** An A^2O biological nutrient-removal process receives wastewater with the characteristics shown below. The system is operated at an 8-d SRT. The RAS recycle ratio, R, is 0.5. The anaerobic contact detention time is 0.75 h. Estimate the effluent soluble phosphorus concentration and the percent phosphorus content of the waste sludge if (a) the RAS contains 6.0 mg/L NO_3-N and (b) if a JHB EBPR process configuration is used and the RAS contains only 0.30 mg/L NO_3-N.

Design conditions and assumptions:

1. Wastewater characteristics:

Item	Unit	Value
Flowrate	m^3/d	4000
Total BOD	g/m^3	160
bCOD	g/m^3	250
rbCOD	g/m^3	75
Acetate	g/m^3	15
nbVSS	g/m^3	20
Inorganic inert matter	g/m^3	10
TKN	g/m^3	35
Phosphorus	g/m^3	6
Temperature	°C	12

Note: g/m^3 = mg/L.

2. rbCOD/ NO_3-N ratio = 5.2 g rbCOD/g NO_3-N
3. Phosphorus content of other heterotrophic biomass = 0.015 g P/g biomass
4. Nitrate oxidized (NO_x) = 28 g/m^3
5. Use coefficients from Table 8–14.

Solution, Part A—NO_3-N = 6.0 mg/L

1. Determine the rbCOD available for enhanced biological phosphorus removal.
   a. rbCOD in influent.

$$Q(\text{rbCOD}) = (4000\ \text{m}^3/\text{d})(75\ \text{g/m}^3) = 300{,}000\ \text{g/d}$$

   b. rbCOD consumed by nitrate.

   NO_3-N fed to the anaerobic contact zone

$$RQ(\text{NO}_3\text{-N}) = 0.50(4000\ \text{m}^3/\text{d})(6.0\ \text{g NO}_3\text{-N/m}^3) = 12{,}000\ \text{g NO}_3\text{-N/d}$$

$$\text{rbCOD used by NO}_3\text{-N} = (5.2\ \text{g rbCOD/g NO}_3\text{-N})(12{,}000\ \text{g NO}_3\text{-N/d})$$

$$= 62{,}400\ \text{g rbCOD/d}$$

   c. rbCOD available = (300,000 − 62,400) g rbCOD/d = 237,600 g/d
2. Determine phosphorus removal by EBPR.
   Use Fig. 8–38 to determine the rbCOD/P removal ratio.
   a. Determine the influent VFA/rbCOD ratio.

$$\frac{\text{VFA}}{\text{rbCOD}} = \frac{(15\ \text{g/m}^3)}{(75\ \text{g/m}^3)} = 0.20$$

   b. From Fig. 8–38 at VFA/rbCOD = 0.20, the rbCOD/P removal = 15.0
   c. P removal by EBPR.

$$\text{P removal} = \frac{\text{rbCOD}}{\text{rbCOD/P}}$$

   rbCOD available normalized to influent flow:

$$\text{rbCOD} = \frac{(237{,}600\ \text{g/d})}{(4000\ \text{m}^3/\text{d})} = 59.4\ \text{g/m}^3$$

$$\text{P removal} = \frac{(59.4\ \text{g rbCOD/m}^3)}{(15.0\ \text{g rbCOD/g P})} = 4.0\ \text{g/m}^3$$

3. Determine P removal by other heterotrophic bacteria for synthesis.
   a. Biomass production [Eq. (8–20), Table 10].

$$P_{x,\text{bio}} = \frac{QY_\text{H}(S_o - S)}{1 + b_\text{H}(\text{SRT})} + \frac{(f_d)(b_\text{H})QY_\text{H}(S_o - S)\text{SRT}}{1 + b_\text{H}(\text{SRT})} + \frac{Q(Y_n)(\text{NO}_x)}{1 + b_n\,(\text{SRT})}$$

   Assume $S_o - S \sim S_o$

   From Table 8–14,

   $Y_\text{H} = 0.45$ g VSS/g COD, $b_\text{H} = 0.12$ g/g·d at 20°C,
   $f_d = 0.15$ g/g, $Y_n = 0.15$ g VSS/g NO_x, $b_n = 0.17$ g/g·d at 20°C

Correct b for 12°C [Eq. 2–25, Table 8–10]

$$b_{\mathrm{H,12°C}} = b_{\mathrm{H,20}}(1.04)^{\mathrm{T-20}}$$

$$b_{\mathrm{H,12°C}} = 0.12(1.04)^{12-20} = 0.088\ \mathrm{g/g{\cdot}d}$$

$$b_{n,\mathrm{12°C}} = 0.17(1.029)^{12-20} = 0.135\ \mathrm{g/g{\cdot}d}$$

$$P_{x,\mathrm{bio}} = \frac{(4000\ \mathrm{m^3/d})(0.45\ \mathrm{g\ VSS/g\ COD})(250\ \mathrm{g\ COD/m^3})}{[1 + (0.088\ \mathrm{g/g{\cdot}d})(8\ \mathrm{d})]}$$

$$+ \frac{(0.15\ \mathrm{g/g})(0.088\ \mathrm{g/g{\cdot}d})(4000\ \mathrm{m^3/d})(0.45\ \mathrm{g\ VSS/g\ COD})(250\ \mathrm{g\ COD/m^3})(8\ \mathrm{d})}{[1 + (0.088\ \mathrm{g/g{\cdot}d})(8\ \mathrm{d})]}$$

$$+ \frac{(4000\ \mathrm{m^3/d})(0.15\ \mathrm{g\ VSS/g\ NO}_x\text{-N})(28\ \mathrm{g/m^3})}{[1 + (0.135\ \mathrm{g/g{\cdot}d})(8\ \mathrm{d})]}$$

$$P_{x,\mathrm{bio}} = 264{,}085\ \mathrm{g\ VSS/d} + 61{,}972\ \mathrm{g\ VSS/d} + 8077\ \mathrm{g\ VSS/d} = 334{,}134\ \mathrm{g\ VSS/d}$$

P removal (synthesis) = 0.015 $P_{x,\mathrm{bio}}$

P removal by synthesis = (0.015 g P/g VSS)(334,134 g VSS/d) = 5012 g P/d

$$\text{Normalized to } Q\text{, P removal} = \frac{(5012\ \mathrm{g\ P/d})}{(4000\ \mathrm{m^3/d})} = 1.2\ \mathrm{g/m^3}$$

4. Effluent P = Influent P − P_{EBPR} − $P_{\mathrm{synthesis}}$

   Effluent P = 6.0 g/m^3 − 4.0 g/m^3 − 1.2 g/m^3 = 0.80 g/m^3

5. Determine P content of waste sludge.
   a. Determine total sludge production using Eq. (8–21), Table 8–10.

$$P_{x,\mathrm{TSS}} = \frac{P_{x,\mathrm{bio}}}{0.85} + \frac{P_{x,\mathrm{AOB}}}{0.85} + Q(\mathrm{nbVSS}) + Q(\mathrm{TSS}_o - \mathrm{VSS}_o)$$

$$= \frac{(334{,}134\ \mathrm{g/d})}{0.85} + (4000\ \mathrm{m^3/d})(20\ \mathrm{g/m^3}) + (4000\ \mathrm{m^3/d})(10\ \mathrm{g/m^3})$$

$$= 433{,}099\mathrm{g/d}$$

   b. P removal = (6.0 − 0.8) g/m^3 (4000 m^3/d) = 20,800 g/d

$$\text{c. P in waste sludge, \%} = \frac{(20{,}800\ \mathrm{g/d})(100)}{(433{,}099\ \mathrm{g/d})} = 4.8\%$$

Solution, Part B—
NO_3-N = 0.3 g/m^3

1. Determine rbCOD available for enhanced biological phosphorus removal.
   a. rbCOD used by NO_3-N (proportion amount of rbCOD used for 0.30 g/m^3 to that for 6.0 g/m^3)

$$\text{rbCOD used by NO}_3\text{-N} = \frac{(0.3\ \mathrm{g/m^3})}{(6.0\ \mathrm{g/m^3})}(62{,}400\ \mathrm{g\ rbCOD/d}) = 3120\ \mathrm{g\ rbCOD/d}$$

   b. rbCOD available for enhanced biological phosphorus removal

   rbCOD available = (300,000 − 3120)g/d = 296,880 g rbCOD/d

2. Determine phosphorus removal by EBPR.
   a. rbCOD : P ratio = 15.0

   b. $$\text{rbCOD available} = \frac{(296{,}880 \text{ g rbCOD/d})}{(4000 \text{ m}^3\text{/d})} = 74.2 \text{ g rbCOD/m}^3$$

   c. P removal by EBPR

   $$P_{\text{removal}} = \frac{(74.2 \text{ g rbCOD/m}^3)}{(15.0 \text{ g rbCOD/g P})} = 4.9 \text{ g/m}^3$$

3. Estimated total P removal by EBPR + Synthesis.

   $$P_{\text{removal}} = 4.9 \text{ g/m}^3 + 1.2 \text{ g/m}^3 = 6.1 \text{ g/m}^3$$

   Note: The computed value exceeds the influent P concentration. The EBPR process will be kinetically limited for P uptake in the aerobic zone at a low P concentration that will occur. Depending on the aerobic tank design the effluent P concentration could be in the range of 0.10 to 0.30 mg/L.

Provision for Chemical Addition

Facility designs for EBPR should include provisions for phosphorus removal by chemical precipitation with alum or iron salts (See Chap. 6) in addition to that by PAOs. Chemical addition may be necessary (1) to meet strict effluent discharge phosphorus limits below that possible by EBPR, (2) to provide process reliability in the event of expected or unexpected conditions that reduce EBPR performance, and (3) for applications where the influent wastewater rbCOD/P ratio is not sufficient to provide the amount of phosphorus removal required. For the latter case, the cost of using chemical precipitation, primary sludge fermentation, return activated sludge fermentation or purchase of external carbon should be compared.

Metal Salt Addition. Extremely low effluent phosphorus concentrations of less than 0.05 mg/L have been achieved for EBPR processes with the addition of metal salts. Metal salts can be added to anaerobic, anoxic or aerobic tanks within an EBPR system, but over dosing can result in a less efficient EBPR performance. Lower EBPR removal could lead to incremental increases in the chemical dose, such that eventually most of the P removal is by chemical precipitation. If too much phosphorus is removed in the mixed liquor by chemical precipitation, less phosphorus is taken up by PAOs in the aerobic zone, which then decreases the amount of stored polyphosphates available for VFA uptake in the anaerobic zone and results in less PAO growth and removal by EBPR. To reach very low effluent P concentrations (i.e., less than 0.10 mg/L) metal salts should be added in a polishing final separate stage after the anaerobic release zone and aerobic P uptake zone. For activated sludge/secondary clarifier systems it can be added just before the clarifier step or in tertiary filtration. For an MBR system it should be added to the membrane separation zone or in a contact tank just before that zone. The addition of alum to an EBPR MBR system can result in less membrane fouling (Johannessen et al., 2005).

Chemical Addition to Primary Sedimentation. Alum or iron salts may be added to remove phosphorus in primary sedimentation prior to the EBPR system, so that minimal effluent P concentrations can be achieved by having a higher rbCOD/P ratio for the EBPR process. However, this approach has the potential of removing too much phosphorus in the primary treatment step which will result in poor performance for nitrogen and phosphorus removal in the secondary treatment step, and thus requires careful operation and control. In some situations, iron salts may be preferred over alum salts for primary treatment applications because they have the additional advantage of removing sulfide to help reduce odors.

Process Control and Performance Optimization

Process performance is affected by a number of operating conditions including (1) process SRT, (2) nitrate-removal efficiency in processes in which nitrification occurs, (3) control of dissolved oxygen and nitrate entering the anaerobic zone, (4) DO concentration in the aerobic zone, (5) phosphorus in recycle streams, (6) the amount of rbCOD and VFA available, and (7) the system effluent suspended solids concentration.

For nitrification systems, an SRT as close to the needed for nitrification will provide higher EBPR efficiency. The addition of DO and nitrate to the anaerobic contact zone must be avoided or minimized. In some cases, for activated sludge/secondary clarifier nitrification systems, it may be possible to control the secondary clarifier sludge blanket level to remove nitrate from the return activated sludge. It must be done with careful operation control and experience; if the sludge blanket is too high, secondary P release could follow the complete removal of nitrate or more solids may escape with the effluent.

Operating methods to increase available rbCOD and VFA can improve EBPR performance. Internal VFA generation by activated sludge and influent particulate fermentation has been demonstrated with improved performance at the Pinery Water, CO and Henderson, NV EBPR facilities by turning off mixers in anaerobic contact zones to allow a greater solids residence time for fermentation. The mixers are turned on 10–20 min each day (Barnard et al., 2011). Effluent soluble P concentrations of less than 0.50 and 0.10, respectively, were reported.

As discussed under Process Design Considerations, recycle streams from sludge thickening or digestion processes may contain high phosphorus concentrations. Equalization and control of the recycle flow and phosphorus load with time may help to minimize the impact of the recycled phosphorus on effluent quality. By adding the recycle streams during times of the day when the influent wastewater strength is higher may allow for the removal of recycled phosphorus in the waste sludge. Recycle streams may also be treated separately with chemical addition to minimize the phosphorus load to the liquid treatment process (see Chap. 15).

Effluent Suspended Solids. The phosphorus content in the mixed-liquor solids is greater than that from the conventional activated sludge process due to the biological phosphorus storage. The phosphorus content, on a dry solids basis, may be in the range of 3 to 6 percent (Randall et al., 1992). Thus, the total phosphorus concentration in the effluent can be affected significantly by the system effluent TSS concentration. At 3 to 6 percent phosphorus in the solids, the phosphorus contribution in an effluent having a TSS concentration of 10 mg/L would be 0.3 to 0.6 mg/L. Fortunately, the biosolids from most EBPR processes settle well with secondary clarifier effluent TSS concentrations of 10 mg/L or less. To provide very low effluent phosphorus concentrations, effluent filtration or membrane separation is needed.

8–9 AERATION TANK DESIGN FOR ACTIVATED SLUDGE PROCESSES

The selection and design of the aeration equipment and aeration tank design is of critical importance in the implementation of the activated sludge processes used with secondary clarifiers or membrane separation systems. Aeration system and aeration tank and appurtenances design considerations are considered in this section.

Aeration System

The aeration system design for the activated sludge process must be adequate to (1) satisfy the oxygen demand for the biological oxidation of the bCOD in the wastewater, (2) satisfy the oxygen demand due to the endogenous respiration of the biomass, (3) meet the oxygen demand for biological nitrification, (4) provide adequate mixing within the reactor, and (5) maintain a minimum dissolved oxygen concentration throughout the aeration tank. If the oxygen transfer efficiency of the aeration system is known or can be estimated, the actual air requirements for diffused air aeration or installed power of mechanical surface aerators may be determined. The characteristics of air diffusers and the energy requirements for mixing for diffused air and mechanical aeration systems are discussed in Sec. 5–12 in Chap. 5. An important issue considered in this section is the effect of the activated sludge process design on the oxygen transfer efficiency in mixed liquor.

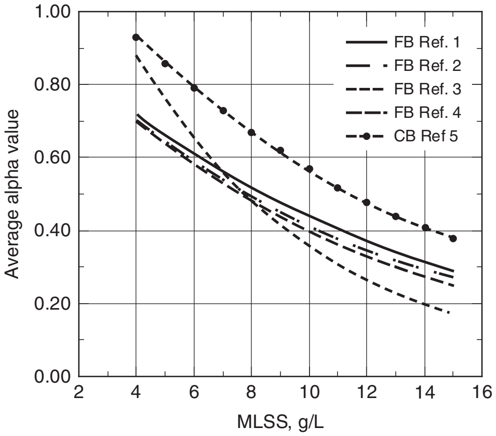
Alpha Factor for Activated Sludge Process with Clarifiers. Aeration design for activated sludge processes involves using clean water oxygen transfer performance data for aeration equipment and adjusting the oxygen transfer rates at the same diffuser air flow or mechanical aerator kW output for the effect of mixed liquor operating conditions. One of the most important factors in the mixed liquor correction is the effect that the mixed liquor has on the aeration design alpha factor. The alpha factor is defined as the ratio of the oxygen transfer rate in mixed liquor to that in clean water. In mixed liquor, the alpha is less than 1.0 due to the effect of surfactants and organic contaminants on gas-liquid mass transfer and the effects of viscosity. The surfactant and organic contaminant concentrations decrease as the SRT is increased due to biodegradation. A range of alpha values for activated sludge/secondary clarifier systems as a function of SRT has been presented by Rosso et al. (2005b) after evaluating oxygen transfer test data for 30 facilities in the United States for fine bubble aeration. Average alpha values increased from about 0.53 to 0.60 to 0.65 as the SRT was increased from 10 to 20 to 30 d.

Alpha Factor for Membrane Bioreactors. The alpha values for fine bubble aeration in MBR systems are further affected by the increased viscosity caused by the higher MLSS concentration. The effect of MLSS concentration on alpha values is shown on Fig. 8–40 for four reports with fine bubble diffused aeration and one report with coarse bubble diffused aeration. The references and correlations of alpha as a function of MLSS concentration are listed in Table 8–32. Approximate alpha values for fine bubble diffusers in MBR aeration tanks are 0.47 at an MLSS concentration of 8000 mg/L and 0.35 for an MLSS concentration of 12,000 mg/L.

Modeling Peak Oxygen Demand. Activated sludge simulation models can include the effects of diurnal changes in flowrate and influent bCOD and TKN concentrations in determining daily sustained peak oxygen demands for different reactor configurations. Experience and engineering judgment can also be used to estimate oxygen demand at peak load conditions as well as design for average oxygen demand conditions. A peaking

Figure 8–40

Effect of MLSS concentration on diffused aeration alpha values in MBR systems. References are given in Table 8–32.



factor of at least 1.5 to 2.0 times the average BOD and TKN load should be used. Aeration equipment may also be sized based on a residual dissolved oxygen (DO) of 2 mg/L in the aeration tank at the average load and 1.0 mg/L at peak load. The aeration equipment must be designed with enough flexibility to (1) meet minimum oxygen demands, (2) prevent excessive aeration and save energy, and (3) meet maximum oxygen demands. Consideration of using multiple smaller blowers to meet minimum oxygen demand is often needed so that the system is not over-aerated and energy wasted at low loads. Excessive aeration can harm biological nutrient removal process performance. The impact of the use of blowers with adjustable guide vanes and variable speed drives on the energy requirements for aeration is considered in Sec. 17–8 in Chap. 17.

Aeration Tanks and Appurtenances

After the activated sludge process and the aeration system have been selected and a preliminary design has been prepared, the next step is to design the aeration tanks and support facilities. The following design aspects are considered in the following discussion (1) aeration tanks, (2) flow distribution, and (3) froth control systems. Energy requirements for aeration tank mixing are discussed in Sec. 5–12 in Chap. 5.

Aeration Tanks. Aeration tanks usually are constructed of reinforced concrete and left open to the atmosphere. The use of rectangular or square shapes permits common-wall construction for multiple tanks. The total required volumetric capacity of the tank is

Table 8–32

Summary of relationships between alpha and MLSS concentration in MBRs

Reference on Figure 8–40	Alpha calculation	Reference
FB Ref. 1	$e^{-(0.082*MLSS)}$	Gunder and Krauth (1999)
FB Ref. 2	$e^{-(0.088*MLSS)}$	Krampe and Krauth (2003)
FB Ref. 3	$(1.6)e^{-(0.15*MLSS)}$	MBR Plant 1, Racault et al. (2010)
FB Ref. 4	$(1.0255)e^{-(0.0946*MLSS)}$	MBR Plant 2, Racault et al. (2010)
CB Ref. 5	$(1.2888)e^{-(0.0818*MLSS)}$	MBR, CB Racault et al. (2010)

Note: FB – Fine bubble diffuser, CB – Coarse bubble diffuser.

determined from the biological process design, as set forth in Secs. 8–6, 8–7, and 8–8. For plants in a capacity range of 0.22 to 0.44 m^3/s (5 to 10 Mgal/d), at least two tanks should be provided (a minimum of two tanks is preferred for smaller plants as well, for redundancy). In the range of 0.44 to 2.2 m^3/s (10 to 50 Mgal/d), four tanks are often provided to allow operational flexibility and ease of maintenance. Large plants, over 2.2 m^3/s (50 Mgal/d) in capacity, should contain six or more tanks. Some of the largest plants have from 30 to 40 tanks arranged in several groups or batteries. Although the air bubbles dispersed in the wastewater occupy perhaps 1 percent of the total volume, no allowance is made for this in tank sizing.

Use of Diffused Air Systems. If the wastewater is to be aerated with diffused air, the geometry of the tank may significantly affect the aeration efficiency and the amount of mixing obtained [see Fig. 8–41(a) and (b)]. The depth of wastewater in the tank should be between 4.5 and 7.5 m (~15 and 25 ft) to maximize the energy efficiency of diffuser systems. Freeboard from 0.3 to 0.6 m (1 to 2 ft) above the waterline should be provided. The width of the tank in relation to its depth is important if spiral-flow mixing is used in the plug-flow configuration. The width-to-depth ratio for such tanks may vary from 1.0/1 to 2.2/1, with 1.5/1 being the most common. In large plants, the channels become quite long and sometimes exceed 150 m (~500 ft) per tank. Tanks may consist of one to four channels with round-the-end flow in multiple-channel tanks. The length-to-width ratio of each channel should be at least 5/1. Where complete-mix diffused air systems are used, the length-to-width ratio may be reduced to save construction cost.

For tanks with diffusers on both sides or in a grid or panel pattern, greater widths are permissible. The important point is to restrict the width of the tank so that "dead spots" or zones of inadequate mixing are avoided. The dimensions and proportions of each independent unit should be such as to maintain adequate velocities so that deposition of solids will

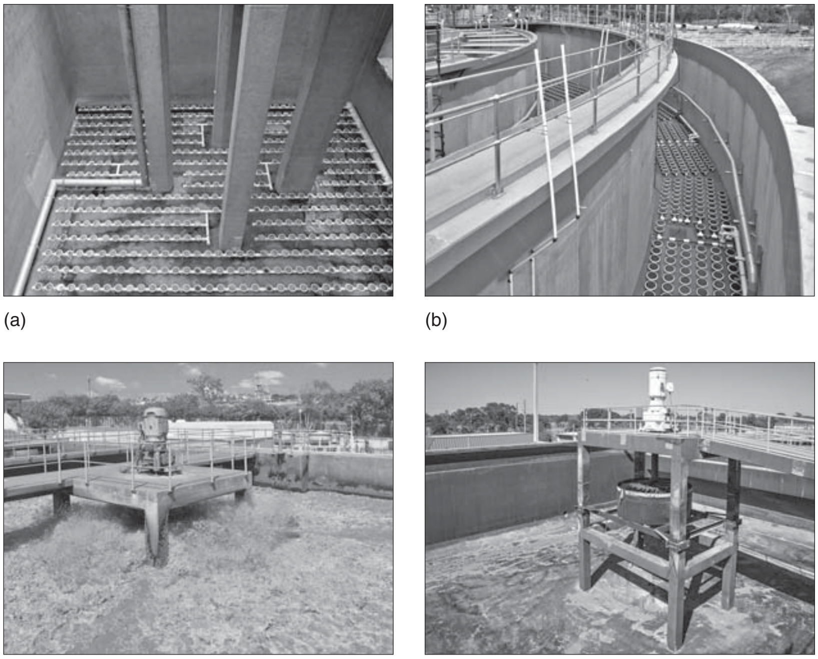

(a) (b) (c) (d)

Figure 8–41

Activated sludge aeration systems: (a) ceramic disk diffusers in a deep complete-mix nitrification tank, (b) ceramic disk diffusers placed in the outer two circular channels of an MBR activated sludge process, (c) mechanical surface aerator, and (d) view of empty reactor with mechanical aerator. Views of membrane panel aeration devices are shown on Fig. 5–67 in Chap. 5.

Table 8–33

Typical aeration tank dimensions for mechanical surface aerators

Aerator size		Tank depth		Tank width	
hp	kW	ft	m	ft	m
10	7.5	10–12	3–3.6	30–40	9–12
20	15	12–14	3.6–4.2	35–50	10.5–15
30	22.5	13–15	3.9–4.5	40–60	12–18
40	30	12–17	3.6–5.1	45–65	13.5–20
50	37.5	15–18	4.5–5.5	45–75	13.5–23
75	56	15–20	4.5–6	50–85	15–26
100	75	15–20	4.5–6	60–90	18–27

not occur. In spiral-flow tanks, triangular baffles or fillets may be placed longitudinally in the corners of the channels to eliminate dead spots and to deflect the spiral flow.

Use of Mechanical Aeration Systems. For mechanical aeration systems, the most efficient arrangement is one aerator per tank [see Fig. 8–41(c) and (d)]. Where multiple aerators are installed in the same tank for best efficiency, the length-to-width ratio of the tank should be in even multiples with the aerator centered in a square configuration to avoid interference at the hydraulic boundaries. The width and depth should be sized in accordance with the power rating of the aerator as illustrated in Table 8–33. Two-speed aerators are desirable to provide operating flexibility to cover a wide range of oxygen demand conditions. Freeboard of about 1 to 1.5 m (3.5 to 5 ft) should be provided for mechanical aeration systems.

Individual tanks should have inlet and outlet gates or valves so that they may be removed from service for inspection and repair. The common walls of multiple tanks must therefore be able to withstand the full hydrostatic pressure from either side. Aeration tanks must have adequate foundations to prevent settlement, and, in saturated soil, they must be designed to prevent flotation when the tanks are dewatered. Methods of preventing flotation include thickening the floor slab, installing hold-down piles, or installing hydrostatic pressure relief valves. Drains or sumps for aeration tanks are desirable for dewatering. In large plants where tank dewatering might be more common, it may be desirable to install mud valves in the bottoms of all tanks. The mud valves should be connected to a central dewatering pump or to a plant drain discharging to the wet well of the plant pumping station. Dewatering systems are commonly designed to empty a tank in 12 to 24 h.

Flow Distribution. For wastewater treatment plants containing multiple units of primary sedimentation basins and aeration tanks, consideration has to be given to equalizing the distribution of flow to the aeration tanks. In many designs, the wastewater from the primary sedimentation basins is collected in a common conduit or channel for transport to the aeration tanks. For efficient use of the aeration tanks, a method of splitting or controlling the flowrate to each of the individual tanks should be used. Methods commonly used are splitter boxes equipped with weirs or control valves or aeration tank influent control gates. Hydraulic balancing of the flow by equalizing the headloss from the primary sedimentation basins to the individual aeration tanks is also practiced. Flow regimes using a form of step feed particularly need a positive means of flow control. Where channels are used for aeration tank influent or effluent transport, they can be equipped with aeration devices to prevent deposition of solids [see Figs. 8–42(a) and (b)]. The air required ranges from 0.2 to 0.5 m^3/lin m·min (2 to 5 ft^3/lin ft·min) of channel.

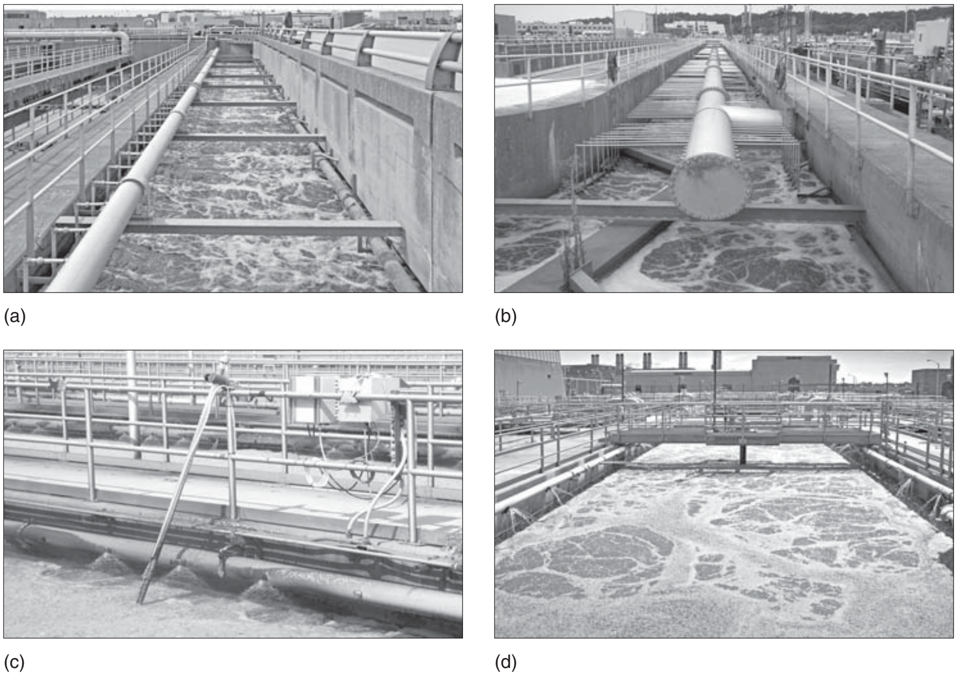

(a) (b) (c) (d)

Figure 8–42

Aeration tank appurtenances: (a) and (b) views of activated sludge aerated mixed liquor transfer channels and (c) and (d) views of foam spay nozzles used to breakup and move foam to waste collection points.

Foam Control Systems. Spray nozzles are commonly mounted above the liquid surface along sides of aeration tanks to break up foaming and to help move foam into surface wasting collection points [see Figs. 8–42(c) and (d)]. During system start up or seasonal load changes foaming may develop and sprays can help move the foam back into the mixed liquor and to keep the level of foam down. Screened or filtered plant effluent is commonly pumped through the spray nozzles. Nocardioform froth is undesirable on aeration tanks and its prevention and control has been discussed previously in Sec. 8–3.

8–10 ANALYSIS OF LIQUID-SOLIDS SEPARATION FOR ACTIVATED SLUDGE PROCESSES WITH CLARIFIERS

Liquid-solids separation is critical to the operation and successful performance of the activated sludge process. Liquid-solids separation involves two very important functions (1) gravity settling to remove over 99.5 percent of the mixed liquor TSS from the treated effluent and (2) thickening of the settled activated sludge to reduce the volume before returning it to the process for mixing and treatment with the influent wastewater. In prior sections of this chapter, activated sludge treatment flow schemes for BOD removal and nitrification, biological nitrogen removal, and enhanced biological phosphorus removal have been shown using two different methods for liquid solids separation: gravity settling in secondary clarifiers and membrane separation. Liquid-solids separation by gravity settling is considered in this section. Design considerations for secondary clarifiers are considered in the following section. Membrane separation of solids is considered in Sec. 8–12.

Solids Separation by Secondary Clarifiers

The surface overflow rate and solids loading rate are the two principal parameters used for the analysis and design of secondary clarifiers. These two parameters are both dependent on the activated sludge system design and operation, which determines the mixed liquor floc characteristics and the amount of dispersed solids not captured in the floc particles during settling. Larger and denser floc particles, usually the result of having biological selectors in the activated sludge design, result in better settling and thickening and a more efficient clarifier performance.

Surface Overflow Rate. The surface overflow rate (SOR), defined as follows, is related to the time needed to allow particle separation from the effluent liquid flow.

$$\mathrm{SOR} = \frac{Q}{A} \tag{8–80}$$

where SOR = surface overflow rate, $\mathrm{m^3/m^2 \cdot d}$
Q = influent flowrate, $\mathrm{m^3/d}$
A = clarifier surface area, $\mathrm{m^2}$

Overflow rates are based on wastewater flowrates instead of on the mixed-liquor flowrates to the clarifier, which includes the influent and recycle sludge flowrates, because the overflow rate is equivalent to an upward flow velocity. The return sludge flow is drawn off the bottom of the tank and does not contribute to the upward flow velocity. Floc and small particles with settling velocities less than the SOR will be removed with the effluent from the clarifier. Floc and small particles settling velocities greater than the SOR will be removed by gravity settling.

Selection of a surface overflow rate is influenced by the effluent requirements and the need to provide consistent process performance. Typical surface overflow rates are given in Table 8–34 and range from 16 to 33 m/d (400 to 800 gal/$\mathrm{ft^2 \cdot d}$). Because steady-state

Table 8–34
Typical design information for secondary clarifiers for the activated sludge process

Type of treatment	Overflow rate				Solids loading				Side-water depth, m[b]
	gal/$\mathrm{ft^2 \cdot d}$		$\mathrm{m^3/m^2 \cdot d}$		lb/$\mathrm{ft^2 \cdot h}$		kg/$\mathrm{m^2 \cdot h}$		
	Average	Peak	Average	Peak	Average	Peak	Average	Peak	
Settling following air activated sludge (excluding extended aeration)	400–600	1000–1200	16–28	36–56	0.8–1.2	2.0	4–6	10	4.0–5.5
Selectors, biological nutrient removal	600–800	1200–1600	24–32	40–64	1.0–1.5	2.0	5–8	10	4.0–5.5
Settling following extended aeration	200–400	600–800	8–16	24–32	0.2–1.0	1.6	1.0–5	8	4.0–5.5
Settling for effluent P concentration after chemical addition[a]									
Total $P = 2$	600–800		24–32						
Total $P = 1$[c]	400–600		16–24						
Total $P = 0.2–0.5$[d]	300–500		12–20						

[a] Adapted in part from Kang (1987), WEF (2010).

[b] m × 3.2808 = ft.

[c] Occasional chemical addition required.

[d] Continuous chemical addition required for effluent polishing.

operations seldom occur due to fluctuations in wastewater flowrate, return activated sludge flowrate, and MLSS concentrations, attention to the occurrence of peak events and use of safety factors are important design considerations. If peak flowrates are of short duration, average 24-h overflow rate values may govern; if peaks are of long duration, peak overflow rate values that prevent the solids from overflowing the tank are assumed. Transient peak flows can be handled more effectively in deep clarifiers as more volume is available to accommodate the clarifier solids inventory due to the higher solids loading rate.

While the surface overflow rate has been the historical clarifier design parameter, the solids loading rate, as discussed below, is considered by some to be the limiting parameter that affects the effluent concentration. It has been shown that with proper hydraulic design and management of solids in the sedimentation tank, the overflow rate has little or no effect on the effluent quality over a wide range of overflow rates, and the design can be based on the solids loading rates (Parker et al., 2001). Based on the evaluation of secondary clarifier performance for a number of facilities, Wahlberg (1995) also found that no effect of using surface overflow rates up to 82 m/d.

Solids Loading Rate. Initially, as noted above, floc and small particles with settling velocities greater than the SOR will settle due to gravity, forming a clear water interface. As the particles continue to settle they begin to coalesce forming a thick suspension, sometimes identified as a sludge blanket. In turn, the sludge blanket undergoes thickening in the bottom portion of the clarifier. If thickening occurs at a slower rate, the amount of solids that can be applied per unit area of the clarifier is limited. The solids loading rate is a parameter used for the design of secondary clarifiers to account for thickening limitations of the secondary clarifier and is defined in Eq. (8–81). The SLR is normally expressed in SI units of kg/m^2·h and as lb/ft^2·d for English units.

$$\text{SLR} = \frac{(Q + Q_R)\text{MLSS}(1\ \text{kg}/10^3\ \text{g})}{A} \tag{8–81}$$

where SLR = solids loading rate, kg TSS/m^2·h

Q = secondary system influent flowrate, m^3/h

Q_R = return activated sludge flowrate, m^3/h

MLSS = mixed liquor suspended solids concentration in flow entering the secondary clarifier, g/m^3

A = clarifier surface area, m^2

The SLR is related to the SOR, mixed liquor suspended solids concentration and return activated sludge recycle ratio.

$$\text{SLR} = \frac{(Q + RQ)\text{MLSS}}{A} = (1 + R)(\text{SOR})(\text{MLSS}) \tag{8–82}$$

where R = return activated sludge recycle ratio = Q_R/Q

The ability to remove particulates is related to the SOR parameter and the ability to thicken the return sludge is related to the SLR parameter. The clarifier design is most often controlled by an allowable SLR, which is related to the sludge thickening properties. If sludge thickening properties decline, the SLR and MLSS concentration must be decreased. Typical SLR values as given in Table 8–34 range from 4 to 6 kg/m^2·h (0.8 to 1.2 lb/ ft^2·d).

Assessing Sludge Thickening Characteristics

At most WWTPs the thickening characteristics of activated sludge mixed liquor are monitored, on a routine basis, using the simple sludge volume index (SVI) test, introduced previously in Sec. 7–8 in Chap. 7, and by measurement of th depth of the sludge blanket in the secondary clarifier.

Sludge Volume Index (SVI). The SVI test is used to monitor any changes in mixed liquor properties as a function of operational, influent wastewater, or seasonal temperature changes. Lower SVI values are normally associated with more rapid thickening and more efficient clarifier performance. The SVI is the volume occupied by 1 g of sludge after 30 min of settling, expressed in milliliters per gram (mL/g). The SVI is determined by placing a mixed-liquor sample in a 1- to 2-L cylinder (see Fig. 8–10) and measuring the settled volume after 30 min and the corresponding sample MLSS concentration. The numerical value is computed using the following expression:

$$\text{SVI, mL/g} = \frac{(\text{settled volume of sludge, mL/L})(10^3 \text{ mg/1 g})}{(\text{suspended solids, mg/L})} \tag{8–83}$$

For example, a mixed-liquor sample with a 3000 mg/L TSS concentration that settles to a volume of 600 mL in 30 min in a 2-L cylinder would have an SVI of 100 mL/g. A value of 100 mL/g is considered a good settling sludge (SVI values below 120 are desired). SVI values above 150 are typically associated with filamentous growth (Parker et al., 2001). A 2-L settleometer is preferred over 1-L graduated cylinders to minimize wall effects on solids thickening (Keinath and Wahlberg, 1994). Alternatively a slow-speed stirring device (~ 1 rev/min) can be used in a small-diameter test apparatus (Wahlberg et al., 1988).

Because the SVI test is empirical, it is subject to significant errors. For example, if sludge with a concentration of 10,000 mg/L did not settle at all after 30 min, the SVI value would be 100. To avoid erroneous results and to allow for a meaningful comparison of SVI results for different sludges, the diluted SVI (DSVI) test has been used (Jenkins et al., 2004). In the diluted test, the sludge sample is diluted with process effluent until the settled volume after 30 min is 250 mL/L or less. The standard SVI test is then followed with this sample.

Depth of Sludge Blanket in Clarifier. Solids thickening occurs in the lower portion of a secondary clarifier before withdrawing return activated sludge. The thickening process results in a sludge depth of some distance from the bottom of the clarifier with a distinctive change in the solids concentration at the top of the thickening zone and water column. The distance from the top of the thickening zone to the bottom of the clarifier is termed the sludge blanket depth. The sludge blanket depth is affected by (1) the clarifier solids loading and sludge thickening properties, (2) rapid changes in the solids loading rate to the clarifier, (3) the return activated sludge recycle ratio, and (4) the solids wasting strategy used by the plant operator. Activated sludge with higher SVI values results in higher sludge blanket levels due to having a slower thickening flux as discussed in Sec. 8–10. Biological nutrient processes tend to have relatively low SVIs, and average sludge blanket depths of 0.30 to 0.60 m are common.

Variations in Blanket Depth. The sludge blanket depth varies with diurnal flow changes as a result of varying solids loading rates to the clarifier. Rapid increases in influent flowrates related to storm events can result in dramatic changes in solids loads to clarifiers and increases in the sludge blanket depth. Deeper clarifiers accommodate such changes better by having more depth and volume to handle a greater solids inventory, while maintaining an adequate water column depth above the sludge blanket so that the settled solids in the sludge blanket are not carried into the effluent. The use of a lower return activated sludge recycle ratio results in a thicker return sludge concentration, which requires more thickening time and a higher sludge blanket level. Higher sludge blanket levels may also be associated with an excess sludge wasting strategy that bases the sludge wasting rates on maintaining a certain sludge blanket level in the clarifier, in lieu of the more proactive SRT control methods recommended in Sec. 8–3. For facilities with multiple activated sludge/clarifier trains, a higher sludge blanket in the clarifier of one train may be indicative of uneven flow splitting between trains.

Measurement of Blanket Depth. Sludge blanket depth measurements are an important operational tool that is used to make changes in the plant operations to control the blanket as necessary to prevent solids carry over into the clarifier effluent. Adjustments in the return activated sludge recycle ratio and sludge wasting rates provide a means to control the sludge blanket depth. The clarifier sludge blanket depth is measured several times during a 24-hour period by automatic or manual methods. Automatic measurements use online instruments that are mounted in the clarifier and employ ultrasound or light techniques. The most common manual method, used at many wastewater facilities because of its simplicity and reliability, is referred to as the *Sludge Judge*. Many variations in sludge judge designs are provided by different suppliers, but in general it is a long, clear plastic tube (19- to 32-mm diameter) that is immersed into the clarifier perpendicular to the clarifier bottom (see Fig. 8–43). The bottom section of the tube has a ball valve that allows liquid in when the tube is lowered into the clarifier and seals the sample when the tube is raised from the clarifier. The tube has markings along its length, typically at 0.3 m increments, which provides a visual means of evaluating the sludge blanket depth by the operator.

Clarifier Design Based on Solids Flux Analysis

The *solids flux* and *state point* methods of analysis are now used commonly to determine clarifier sizing. Both methods of analysis are based on a consideration the sludge-settling properties and clarifier return sludge flowrate. The solids flux method is considered in the following discussion. The state point method is considered following the presentation of the solids flux method.

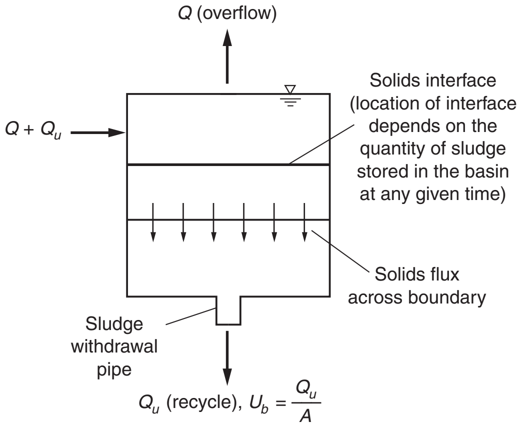
(a) (b) (c)

Figure 8–43

Views of treatment plant operators conducting sludge blanket depth measurements using a sludge judge.

Figure 8–44

Definition sketch for solids transport in a settling basin operating at steady state.



Definition of Solids Flux. Solids flux is defined as the rate of the solids mass moving downward across a unit area in the clarifier. In a settling basin that is operating at steady state, a constant flux of solids is moving downward, as shown on Fig. 8–44. Within the tank, the downward flux of solids is brought about by gravity (hindered) settling plus bulk transport due to the underflow that is being pumped out and recycled. As thickening occurs with depth, the flux due to each of these components changes. The area required for thickening of the applied mixed liquor depends on the limiting solids flux that can be transported to the bottom of the sedimentation basin. The depth of the thickening portion of the sedimentation tank must be sufficient to (1) ensure maintenance of an adequate sludge blanket depth so that unthickened solids are not recycled, and (2) temporarily store excess solids that may be applied.

Solids Flux Due to Gravity. The solids flux due to gravity is as follows:

$$\mathrm{SF_g} = C_i V_i (1\,\mathrm{kg}/10^3\,\mathrm{g}) \tag{8–84}$$

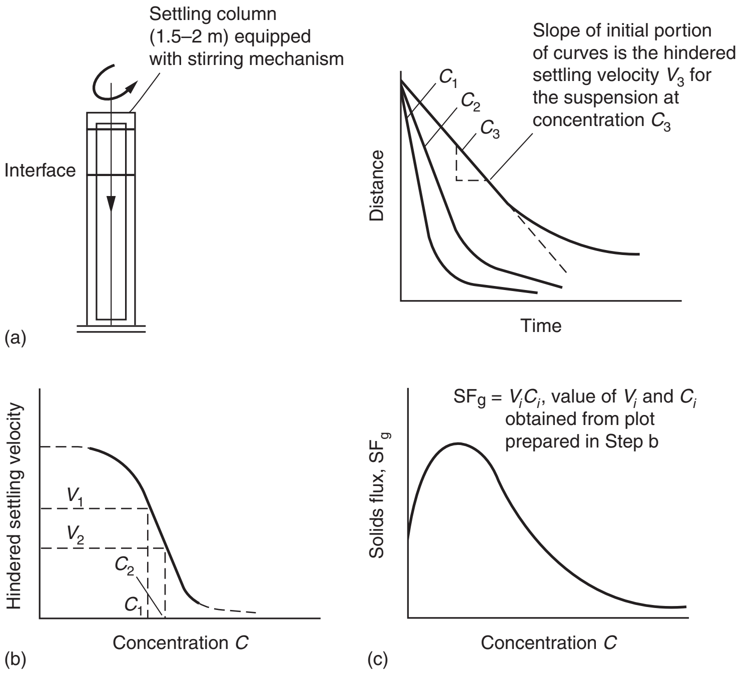
where $\mathrm{SF_g}$ = solids flux due to gravity, kg/m²·h
C_i = concentration of solids at the point in question, g/m³
V_i = initial settling velocity of the solids at concentration C_i, m/h

The flux of solids due to gravity settling depends on the concentration of solids (C_i) and the settling velocity (V_i) of the solids at that concentration. Because the solids flux due to gravity varies with the characteristics of the sludge, column settling tests are conducted to determine the relationship between the sludge concentration and the settling rate. The procedure used to develop a solids flux curve from column settling test data is illustrated on Fig. 8–45. First, a series of settling tests are done at different initial mixed liquor solids concentrations in which the initial settling velocity (V_i) is measured over a 5 to 10 min time period [see Fig. 8–45(a)]. Next, the settling velocity is plotted versus the MLSS concentration [see Fig. 8–45(b)]. The third step is to plot the solids flux as given by Eq. (8–84) versus the MLSS concentration [see Fig. 8–45(c)].

It should be noted that the initial settling velocity is also termed the *zone settling velocity* (ZSV) because an interface between a clear liquid above the sludge blanket occurs at some rate. At higher MLSS concentrations, the value for V_i will decrease. At low concentrations (below about 1000 mg/L), the movement of solids due to gravity is small, because the settling velocity of the solids is more or less independent of concentration. If the velocity remains essentially the same as the solids concentration increases, the flux due to gravity starts to increase as the solids concentration starts to increase. At higher solids concentrations, hindered settling occurs and the ZSV and gravity flux decreases. At some

Figure 8–45

Procedure for preparing a plot of solids flux due to gravity as a function of solids concentration: (a) hindered settling velocity is derived from column settling test for suspension at different concentrations, (b) plot of hindered settling velocity is obtained in step (c) versus corresponding concentration, and (d) plot of completed value of solids flux vs. corresponding concentration.



very high concentration the flux approaches zero. A typical gravity solids flux curve for an activated sludge MLSS is shown on Fig. 8–46.

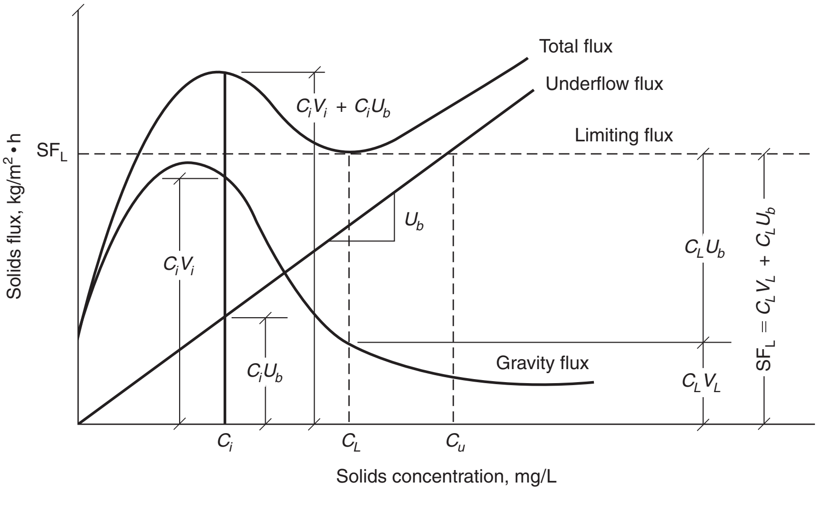
Zone Settling Velocity as a Function of MLSS and SVI. The zone settling velocity as a function of the mixed liquor concentration and SVI can be estimated using the following equation (Wilson and Lee, 1982; Wilson, 1996):

$$V_i = V_{\max} \exp[-(k/10^6)X] \qquad (8\text{–}85)$$

where V_i = settling velocity of interface, m/h
$V_{\max}$ = maximum settling velocity of interface, typically 7 m/h
K = constant, typically 600 L/mg for activated sludge mixed liquor with an SVI of 150
X = average MLSS concentration, mg/L

Figure 8–46

Definition sketch for applying the solids flux method with solids thickening data and a return sludge flowrate.



Correlations between V_i and the MLSS concentration have also been developed with SVI as an additional parameter using data from several facilities (Daigger, 1995; Wahlberg, 1995). The results of the correlations are given by the following equations in which lower zone settling velocities are computed with higher SVI values:

$$\text{In}(V_i) = 1.871 - (0.165 + 0.00159 \text{ SVI})X_T \tag{8–86}$$

$$\text{In}(V_i) = 2.082 - (0.103 + 0.00256 \text{ DSVI})X_T \tag{8–87}$$

where DSVI = diluted SVI, mL/g
X_T = MLSS concentration, g/L

The advantage of these correlations is that a solid flux analysis can be performed at a facility to estimate the clarifier capacity using easy to obtain SVI data in lieu of the extensive settling testing procedure.

Solids Flux Due to Bulk Movement. The solids flux due to the bulk movement of the suspension caused by the return sludge flow (see Fig. 8–44) is given by the following expression:

$$\text{SF}_\text{u} = C_i U_b (1\text{kg}/10^3 \text{ g}) = C_i \frac{Q_R}{A} (1 \text{ kg}/10^3 \text{ g}) \tag{8–88}$$

where SF_u = solids flux due to underflow, kg/m²·h
U_b = bulk downward velocity, m/h
Q_R = underflow flowrate, m³/h
A = cross-sectional area, m²

The solids flux due to bulk transport is a linear function of the concentration with slope equal to U_b, the underflow velocity, and is shown as the underflow flux on Fig. 8–46.

Total Solids Flux. The total solids flux SF_t, comprised of the gravity solids flux and the bulk solid flux, is given by the following expressions.

$$\text{SF}_\text{t} = \text{SF}_\text{g} + \text{SF}_\text{u} \tag{8–89}$$

$$\text{SF}_\text{t} = (C_i V_i + C_i U_b)(1 \text{ kg}/10^3 \text{ g}) \tag{8–90}$$

The total flux, as shown on Fig. 8–46, is the sum of the gravity and the underflow flux. The shape of the total flux curve follows the pattern for the gravity flux curve and decreases when hindered settling occurs as the solids get thicker. The total flux reaches a minimum value or *limiting flux* at a certain solids concentration, before increasing as the solids continue to thicken to higher concentrations. The limiting flux value is determined by drawing a horizontal line from the minimum point of the total flux curve extending to the vertical axis. The clarifier solids loading must not exceed the limiting flux or the solids will accumulate and increase the sludge blanket depth, with solids eventually spilling over into the clarifier effluent flow. The maximum allowable clarifier SLR in Eq. (8–81) is equal to the limiting solids flux.

$$\text{SLR} = \frac{(Q + Q_R)\text{MLSS}(1 \text{ kg}/10^3 \text{ g})}{A} = \text{SF}_\text{L} \tag{8–91}$$

where SF_L = limiting solid flux, kg TSS/m²·h

Using the limiting solids-flux value, the required clarifier area derived from a materials balance is given by

Based on flowrate, *Q*

$$A = \frac{(Q + Q_R)(\text{MLSS})}{\text{SF}_\text{L}}(1\ \text{kg}/10^3\ \text{g}) \qquad (8\text{–}92)$$

Based on the recycle ratio, *R*

$$A = \frac{(1 + R)(Q)(\text{MLSS})}{\text{SF}_\text{L}}(1\ \text{kg}/10^3\ \text{g}) \qquad (8\text{–}93)$$

where terms are as defined in Eqs. (8–81 and 8–82).

Return Flow Concentration. The corresponding underflow concentration (C_u) or return activated sludge concentration (X_R) is obtained by dropping a vertical line to the X axis from the intersection of the horizontal line and the underflow flux line assuming the gravity flux is negligible at the bottom of the settling basin and the solids are removed by bulk flow. The fact that the gravity flux is negligible at the bottom of the tank can be verified by performing a materials balance around the portion of the settling tank that lies below the depth where the limiting solids flux occurs and comparing the gravity settling velocity of the sludge to the velocity in the sludge withdrawal pipe.

Aeration Tank MLSS. The MLSS concentration in the aeration tank is governed by the underflow concentration which is related to the limiting solids flux and the return sludge recycle ratio. Because the sludge wasting flow is small compared to the recycle sludge flowrate, it can be ignored for the simple mass balance around the clarifier to estimate the aeration tank MLSS concentration:

$$X(Q + Q_R) = Q_R(X_R), \text{ where } X_R = C_u \qquad (8\text{–}94)$$

and

$$X = \left(\frac{R}{1 + R}\right)(X_R), \text{ where } R = \frac{Q}{Q_R} \qquad (8\text{–}95)$$

If a thicker underflow concentration is required, the slope of the underflow flux line shown on Fig. 8–46 must be reduced. The reduced slope, in turn, will lower the value of the limiting flux and increase the required settling area. In an actual design, the use of several different flowrates for the underflow should be evaluated.

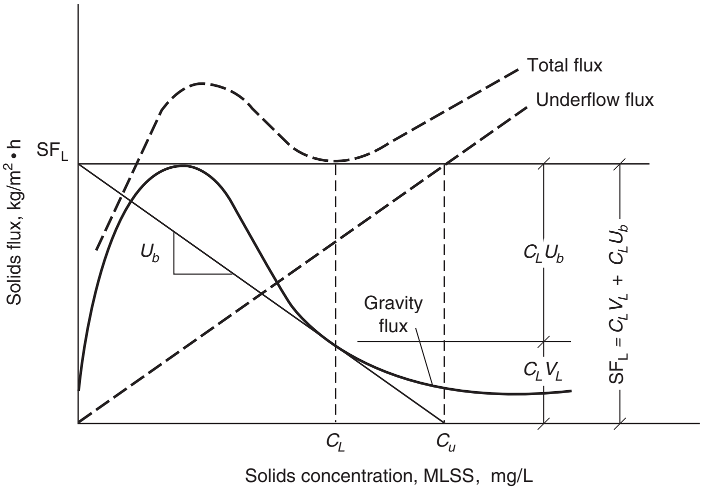
Graphical Analysis of Solid Flux. An alternative graphical method of analysis to that presented on Fig. 8–46 for determining the limiting solids flux is shown on Fig. 8–47. The graphical analysis is derived from Eq. (8–90) in which the minimal SF_t is defined where its derivative with respect to C_i equals zero:

$$\frac{\partial \text{SF}_\text{t}}{\partial C_i} = 0 = V_i - U_b \qquad (8\text{–}96)$$

Thus, as shown on Fig. 8–47, the solids concentration at the limiting solids flux is C_L and from Eq. (8–96) $C_L V_i = C_L U_b$. The value of the limiting flux on the ordinate is obtained by drawing a line tangent to the flux curve passing through the desired underflow and intersecting the ordinate. The geometric relationship of this method to that given on Fig. 8–46 is shown by the line for U_b on Fig. 8–47. The method detailed on Fig. 8–47 is especially useful where the effect of the use of various underflow concentrations on the size of the

Figure 8–47

Alternative definition sketch for determining the limiting solids flux in the solids flux method of analysis.



treatment facilities (aerator and sedimentation basin) is to be evaluated. Because the underflow velocity (return sludge flowrate) can be controlled, it is used for process control. Application of the solids flux method of analysis is illustrated in Example 8–14.

EXAMPLE 8–14 **Application of Solids Flux Analysis** Given the following settling data for a mixed liquor, derived from an anoxic/aerobic activated sludge pilot plant, estimate the maximum mixed-liquor suspended solids concentration that can be maintained in the aeration tank if the secondary clarifier surface overflow rate, Q/A, has been fixed at 24 m^3/m^2•d and the return sludge recycle rate, Q_R is equal to 75 percent of Q. The definition sketch for this problem is shown on Fig. 8–11(b). As shown, settled and thickened activated sludge from the secondary clarifier are returned to the aeration tank to maintain the desired level of mixed liquor solids. Assume that the solids wasting rate Q_w is negligible in this example.

MLSS, g/m^3	Initial settling velocity, m/h
1000	6.246
2000	3.203
3000	1.642
4000	0.842
5000	0.432
6000	0.221
7000	0.113
8000	0.058
9000	0.030
10,000	0.015
11,000	0.008
12,000	0.004
13,000	0.002
14,000	0.001

Note: g/m^3 = mg/L.

Solution

1. Set up a computation table to determine the gravity, underflow, and total solids-flux values corresponding to the given solids concentrations. Sample calculations for the required table entries are as follows.
   a. Determine the gravity solids flux as a function of the MLSS using the following relationship.

      Solids flux $= X(\text{g/m}^3)V(\text{m/h})\ (1\ \text{kg}/10^3\ \text{g})$

      For example, at $X_i = 2000\ \text{g/m}^3$,

      Solids flux $= (2000\ \text{g/m}^3)(3.203\ \text{m/h})(1\ \text{kg}/10^3\ \text{g}) = 6.41\ \text{kg/m}^2\cdot\text{h}$

   b. Determine the underflow bulk velocity.
      i. The surface loading rate on the clarifiers equals (Q/A), 24 $\text{m}^3/\text{m}^2\cdot\text{d}$ or 1.0 m/h.
      ii. The underflow velocity U_b is therefore equal to $(0.75)(24\ \text{m/h}) = 0.75\ \text{m/h}$.
   c. The underflow solids-flux is determined using the following relationship:

      $\text{SF}_\text{u} = X_i U_b (1\ \text{kg}/10^3\ \text{g})$

      where X_i = MLSS concentration, g/m^3
      U_b = bulk underflow velocity, m/h

      For example, at $X_i = 2000\ \text{g/m}^3$,

      $\text{SF}_\text{u} = (2000\ \text{g/m}^3)(0.75\ \text{m/h})(1\ \text{kg}/10^3\ \text{g}) = 1.5\ \text{kg/m}^2\cdot\text{d}$

   d. Determine the total solids-flux using Eq. (8–89).

      $\text{SF}_\text{t} = \text{SF}_\text{g} + \text{SF}_\text{u}$

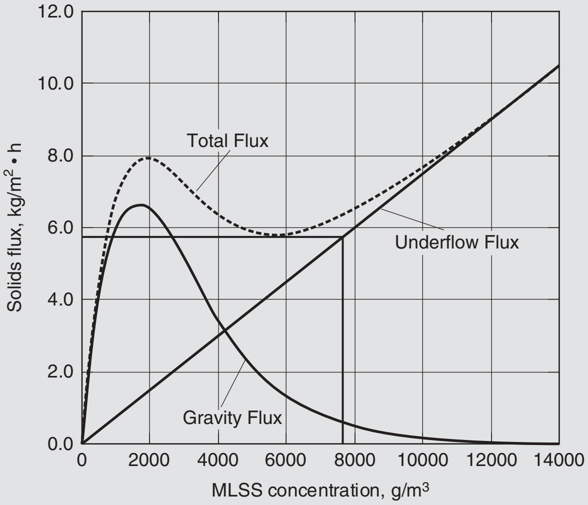
      For example, at $X_i = 2000\ \text{g/m}^3$,

      $\text{SF}_\text{t} = 6.41 + 1.50 = 7.91\ \text{kg/m}^2\cdot\text{h}$

   e. Prepare summary table of the gravity, underflow, and total solids-flux.

MLSS, g/m^3	Gravity solids flux, kg/m$^2\cdot$h	Underflow solids flux, kg/m$^2\cdot$h	Total solids flux, kg/m$^2\cdot$h
1000	6.25	0.75	7.00
2000	6.41	1.50	7.91
3000	4.93	2.25	7.18
4000	3.37	3.00	6.37
5000	2.16	3.75	5.91
6000	1.33	4.50	5.83
7000	0.79	5.25	6.04
8000	0.47	6.00	6.47
9000	0.27	6.75	7.02
10,000	0.15	7.50	7.65
11,000	0.09	8.25	8.34
12,000	0.05	9.00	9.05
13,000	0.03	9.75	9.78
14,000	0.01	10.50	10.51

2. Plot the flux curves (see following figure).



3. Determine the limiting solids flux and maximum underflow concentration.
   a. The limiting solids flux value is found by drawing a horizontal line extending to the underflow flux line, tangent to the total solids flux curve.

      From the plot in Step 2, the limiting flux at the point of tangency to the total flux curve is

      $SF_L = 5.8 \text{ kg/m}^2 \cdot \text{h}$

   b. The maximum underflow solids concentration at the intersection of the horizontal line and the underflow flux is equal to 7800 g/m^3.
4. Estimate the maximum solids concentration that can be maintained in the reactor.
   a. Write a mass balance for the system within the boundary, neglecting the rate of cell growth within the reactor. Let X_o = influent TSS to the aeration tank.

      $$QX_O + Q_R X_R = (Q + Q_R)X$$

   b. Assuming the $X_O = 0$ ($X_O << X_R$) and that $Q_R/Q = 0.75$, solve for the maximum MLSS concentration in the reactor.

      $$0.75Q(7800 \text{ g/m}^3) = (1 + 0.75)QX$$
      $$X = 3340 \text{ g/m}^3$$

Comment As shown in the above analysis, the concentration of the return solids will affect the maximum concentration of solids that can be maintained in the aeration tank. Thus, the secondary clarifier must be considered an integral part of the design of an activated sludge treatment process.

Clarifier Design Based on State Point Analysis

The state point analysis procedure extends the principles of the solids-flux analysis to provide a convenient means to assess different mixed-liquor concentrations and clarifier operating conditions relative to the limiting solids-flux operating condition (Keinath et al., 1977; Keinath, 1985).

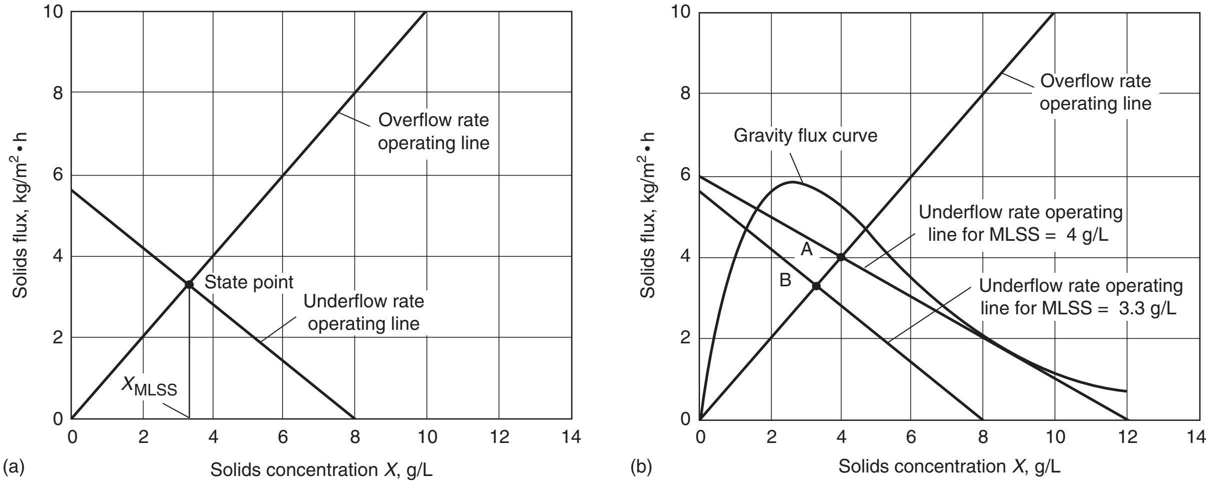


Figure 8–48

State point analysis for assessing clarifier operating conditions: (a) state point at intersection of overflow rate and underflow rate operating lines and (b) under loaded (B), and critically loaded (A) state points relative to settling flux curve.

The State Point. The state point, as shown on Fig. 8–48(a), is the intersection of the clarifier overflow solids-flux rate and underflow solids-flux rate lines. Thus, the analysis accounts for the actual mixed-liquor concentration, clarifier hydraulic application rate, return activated sludge recycle rate, and whether the combination of these operating parameters results in a condition that is within the clarifier solids-flux limitations for a sludge with specific thickening characteristics.

Overflow Solids Flux. The clarifier overflow solids flux as shown on Fig. 8–48(a) is

$$\mathrm{SF_Q} = \frac{Q(X)}{A} \tag{8–97}$$

where $\mathrm{SF_Q}$ = overflow solids-flux rate, kg/m²·d
Q = clarifier effluent flowrate, m³/d
A = clarifier cross-section area, m²
X = aeration tank MLSS concentration, g/L

The aeration tank MLSS concentration (X) at any point along the overflow solids-flux line is found by constructing a vertical line to the X axis.

Underflow Operating Line. The underflow operating line represents the negative slope of the clarifier underflow velocity as was also shown on Fig. 8–47. The intercept of the horizontal line drawn from the point of intersection of the vertical line and operating line is the total solids flux ($\mathrm{SF_X}$) to the clarifier. Evaluating the slope of the underflow flux rate line it can be shown that U_b, the bulk downward velocity, equals

$$U_b = \frac{\mathrm{SF_t} - \mathrm{SF_Q}}{0 - X_{\mathrm{MLSS}}} \tag{8–98}$$

$$U_b = \frac{[(Q + Q_R)X_{\mathrm{MLSS}}/A\,]}{-\,X_{\mathrm{MLSS}}} \tag{8–99}$$

$$U_b = -\frac{Q_R}{A} \tag{8–100}$$

The state point and underflow solids-flux line can be compared to the gravity flux curve to determine if the clarifier operation is within its solids-flux limitation [see Fig. 8–48(b)]. The underflow line at state point A is tangent to the gravity flux curve, which, as shown on Fig. 8–47, represents the limiting solids-flux condition. Thus, the clarifier is loaded critically for this underflow velocity and the MLSS concentration at the state point is equal to 4.0 g/L. If the operation is changed to obtain a higher MLSS concentration, and the underflow line crosses the lower limb of the gravity flux curve, the limiting solids flux will be exceeded and the clarifier blanket will rise to the effluent weir. At state point B on Fig. 8–48(b), a lower MLSS concentration is used and the underflow line is thus well below the lower limb of the gravity flux curve. An underloaded operation exists relative to the solids loading.

Use of the State Point Analysis. The state point analysis procedure provides a method to evaluate various clarifier overflow rates, and MLSS concentrations give a gravity flux curve that represents the activated sludge settling properties. The state point analysis technique can be used with settling tests on activated sludge mixed liquor at an existing facility to determine an optimal MLSS concentration and return sludge recycle ratio for a given influent flow condition. Application of state point analysis is illustrated in Example 8–15.

EXAMPLE 8–15 **Evaluate Secondary Clarifier Operating Conditions Using State Point Analysis** Determine acceptable operating conditions using the following solids-settling test results for one or two clarifiers in operation in an activated sludge system. The system is to be evaluated at a maximum month design flowrate of 15,070 m^3/d and the following design conditions.

Design conditions:

1. Two 20-m diameter clarifiers are to be used.
2. With both clarifiers in operation, the desired MLSS concentration is 3500 mg/L.
3. Evaluate the feasibility of operating the clarifier with underflow concentrations of 10, 12, and 14 g/L and determine the recycle ratios.
4. Determine the MLSS concentration with one clarifier in operation using an underflow solids concentration of 12 g/L.
5. Determine the solids loading to the clarifier for the 12 g/L underflow concentration and MLSS = 3500 g/m^3 for (a) two-clarifier operation and (b) one clarifier.
6. The solids settling results are:

Mixed-liquor solids concentration, g/m^3	Interfacial settling velocity, m/h
2000	2.90
3000	1.90
4000	1.30
5000	0.90
6000	0.60

(*continued*)

(*Continued*)

Mixed-liquor solids concentration, g/m^3	Interfacial settling velocity, m/h
8000	0.26
9000	0.17
10,000	0.12
12,000	0.05
16,000	0.01

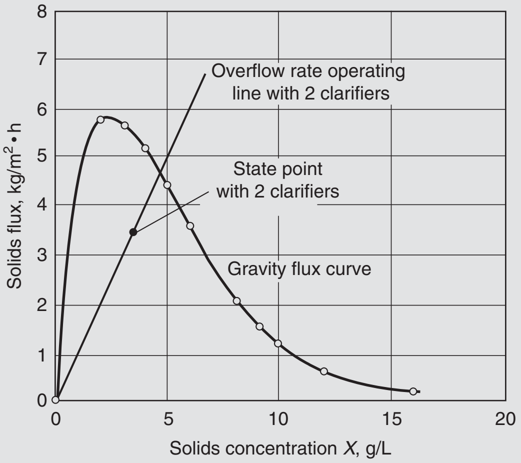
Note: g/m^3 = mg/L.

Solution

1. Develop gravity flux curve (shown below) using Eq. (8–84), $SF_g = C_iV_i$.
   a. Determine the SF_g values using the given data.

C_i, g/L	V_i, m/h	SF_g, kg/m^2·h
2.0	2.90	5.80
3.0	1.90	5.70
4.0	1.30	5.20
5.0	0.90	4.50
6.0	0.60	3.60
8.0	0.26	2.08
9.0	0.17	1.53
10.0	0.12	1.20
12.0	0.05	0.60
16.0	0.01	0.16

   b. Plot the gravity solids flux curve.



2. Add the overflow rate operating line and MLSS concentration state point at 3500 mg/L overflow solids flux rate.
   a. Determine the clarifier surface area.

$$A(\text{area/clarifier}) = \pi D^2 / 4 = \pi(20)^2 / 4 = 314\ \text{m}^2$$

$$\text{Total area (2 clarifiers)} = 2 \times 314 = 628\ \text{m}^2$$

   b. Determine the overflow rate, expressed as kg/m²·h, as a function of X using Eq. (8–97).

$$\text{SF}_\text{Q} = \frac{Q(X)}{A} = \frac{(15{,}070\ \text{m}^3/\text{d})(1\ \text{d}/24\ \text{h})(X)}{628\ \text{m}^2} = 1.0\ \text{m/h}(X)$$

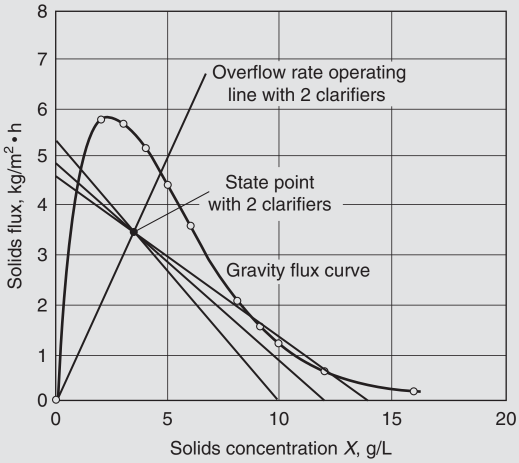
For example at $X = 5000\ \text{g/m}^3$,

$$\text{SF}_\text{o} = (1.0\ \text{m/h})(5\ \text{kg/m}^3) = 5.0\ \text{kg/m}^2\cdot\text{h}$$

   c. Plot the overflow flux line on the graph developed in Part 1.
3. Plot the state point on the overflow flux line for an underflow concentration of 3500 mg/L (3.5 kg/m³).
   a. Determine the flux at the state point.

$$\text{SF}_\text{Q} = (1.0\ \text{m/h})(3.5\ \text{kg/m}^3) = 3.5\ \text{kg/m}^2\cdot\text{h}.$$

   b. Plot the state point on the overflow flux line graph (see Plot developed in Part 1).
4. Evaluate underflow conditions at 10, 12, and 14 g/L (see following figure).
   a. Analysis for an underflow concentration of 14 g/L. Draw a line that intercepts 14 g/L on the *x* axis and passes through the state point. The line intercepts the *y* axis at 4.67 kg/m²·h; however, this flux rate is not a feasible operating condition as the line crosses above the gravity flux curve. For 10 and 12 g/L underflow concentrations, the lines cross below the gravity flux curve; therefore, both concentrations are feasible.



   b. Calculate recycle ratio at 10 g/L (10 kg/m³) underflow concentration. Determine the slope of the underflow operating rate curve from the figure. The intercept on the x axis is 5.38 kg/m²·h and the slope is negative and is equal to the recycle velocity, m/h.

$$\text{Operating curve slope} = \frac{[(5.38 - 0)\text{kg/m}^2\cdot\text{h}]}{[(0 - 10)\text{g/m}^3]} = -0.538 \text{ m/h}$$

$$\text{Underflow velocity} = -(0.538 \text{ m/h}) = 0.538 \text{ m/h}$$

$$\text{Clarifier overflow rate} = \frac{(15{,}070 \text{ m}^3/\text{d})(1\text{d}/24 \text{ h})}{628 \text{ m}^2} = 1.0 \text{ m/h}$$

$$\text{Recycle ratio} = \frac{(0.538 \text{ m/h})}{(1\text{m/h})} = 0.538$$

c. Check recycle ratio at 10 g/L underflow concentration using solids balance.

$$X_R Q_R = (Q_R + Q)X$$

$$X_R R = (1 + R)X \text{ where } R = \text{recycle ratio}$$

$$R = \left(\frac{X_R}{X} - 1\right)^{-1} = \left[\frac{(10 \text{ g/L})}{(3.5 \text{ g/L})} - 1\right]^{-1}$$

$$R = 0.538$$

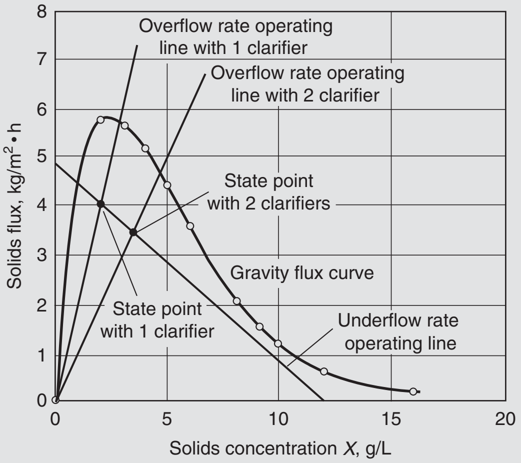
d. Calculate recycle ratio at 12 g/L (12 kg/m³) underflow concentration using the same procedure as above.

$$\text{Underflow velocity} = \frac{[(4.94 - 0) \text{ kg/m}^2\cdot\text{h}]}{[(0 - 12) \text{ g/m}^3]} = -0.4 \text{ m/h}$$

$$R = 0.41$$

5. Determine what MLSS concentration is possible with one clarifier operating and underflow solids concentration 12 g/L.

Using the gravity flux curve, draw an overflow operating rate line for one clarifier (see following figure).



For $A = 314\ \text{m}^3$ and $X = 2\ \text{g/L}$,

$$\text{SF}_Q = \frac{(15{,}070\ \text{m}^3/\text{d})(1\ \text{d}/24\ \text{h})(2\ \text{kg/m}^3)}{314\ \text{m}^2} = 4\ \text{kg/m}^2\cdot\text{h}$$

The MLSS concentration possible is the "state point" where the overflow rate operating line intersects the underflow rate operating line. In the above figure, the state point is approximately 2.1 g/L (2100 mg/L) MLSS concentration.

6. Determine clarifier solids loading.
   a. For 2 clarifiers: $A = 628\ \text{m}^2$; MLSS $= 3.5$ g/L, and $R = 0.41$

   Solids loading $= Q(1 + R)(X)/A$

   $$= \frac{(15{,}70\ \text{m}^3/\text{d})(1 + 0.41)(3.5\ \text{kg/m}^3)}{(314\ \text{m}^2)(24\ \text{h/d})} = 4.93\ \text{kg/m}^2\cdot\text{h}$$

   b. For one clarifier: $A = 314\ \text{m}^2$

   $$\text{Clarifier overflow rate} = \frac{(15{,}070\ \text{m}^3/\text{d})(1\ \text{d}/24\ \text{h})}{314\ \text{m}^2} = 2.0\ \text{m/h}$$

   Underflow velocity (from step 3d) $= 0.41$ m/h

   $R = (0.41\ \text{m/h})/(2.0\ \text{m/h}) = 0.205$

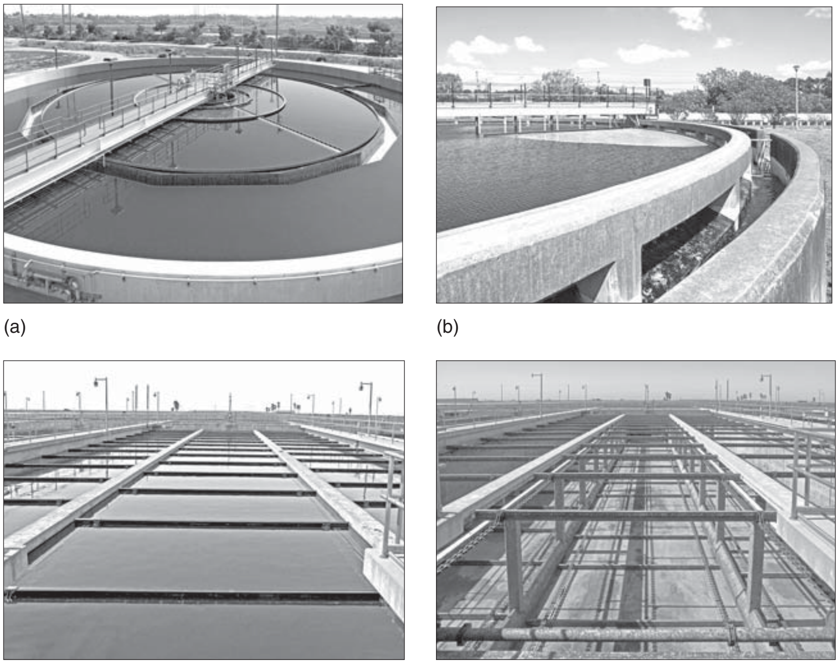
   $$\text{Solids loading} = \frac{(15{,}070\ \text{m}^3/\text{d})(1 + 0.205)(2.1\ \text{kg/m}^3)}{(314\ \text{m}^2)(24\ \text{h/d})} = 5.06\ \text{kg/m}^2\cdot\text{h}$$

8–11 DESIGN CONSIDERATIONS FOR SECONDARY CLARIFIERS

Solids separation is the final step in the production of a well-clarified, stable effluent low in BOD and TSS and, as such, represents a critical link in the operation of an activated sludge treatment process. Although much of the information presented in Chap. 5 for the design of primary sedimentation tanks is applicable, the presence of a large volume of flocculent solids in the mixed liquor requires that special consideration be given to the design of activated sludge settling tanks. As mentioned previously, these solids tend to form a sludge blanket in the bottom of the tank that will vary in thickness. The blanket may fill the entire depth of the tank and overflow the weirs at peak flowrates if the return sludge pumping capacity or the size of the settling tank is inadequate. Further, the mixed liquor, on entering the tank, has a tendency to flow as a density current, interfering with the separation of the solids and the thickening of the sludge. To cope successfully with these characteristics, the following factors must be considered in the design of secondary sedimentation tanks: (1) surface and solids loading rates, (2) tank types, (3) sidewater depth, (4) flow distribution, (5) inlet design, (6) weir placement and loading rates, and (7) scum removal. Surface and solids loading rates have been considered previously in Sec.8–10. The remaining factors are considered in this chapter.

Types of Sedimentation Tanks

The most commonly used types of activated sludge settling tanks are either circular [see Figs. 8–49(a) and (b)] or rectangular [see Figs. 8–49(c) and (d)]. Square tanks are used on occasion but they are not as effective in retaining separated solids as circular



(a) (b) (c) (d)

Figure 8–49

Typical views of secondary clarifiers: (a) circular clarifier with inboard effluent weirs, sludge collection mechanism is mounted in the center, (b) circular clarifier peripheral weir and rim drive bridge for the collection of sludge and surface skimming, (c) large rectangular clarifier shown full, and (d) tank shown on (c) empty. Note chain and flight sludge collection and skimming mechanism. Because of width of clarifier three chain and flight mechanisms are used.

or rectangular tanks. Solids accumulate in the corners of the square tanks and are frequently swept over the weirs by the agitation of the sludge collectors. Circular tanks have been constructed with diameters ranging from 3 to 60 m (10 to 200 ft), although the more common range is from 10 to 40 m (30 to 140 ft). The tank radius should preferably not exceed five to six times the sidewater depth.

Center and Rim Feed Circular Clarifiers. Two basic types of circular tanks are used for secondary sedimentation: center-feed and rim-feed (see Fig. 5–41 in Chap. 5). Both types use a revolving mechanism to transport and remove the sludge from the bottom of the clarifier. Mechanisms are of two types: those that scrape or plow the sludge to a center hopper similar to the types used in primary sedimentation tanks, and those that remove the sludge directly from the tank bottom through suction orifices that serve the entire bottom of the tank in each revolution. Of the latter, in one type the suction is maintained by reduced static head on the individual suction pipes [see Fig. 8–50(a)]. In another patented suction system, sludge is removed through a manifold either hydrostatically or by pumping. Spiral-type scrapers are also used to accelerate movement of settled solids from the tank periphery to the collection sump [see Fig. 8–50(b)].

Rectangular Clarifiers. Rectangular tanks must be proportioned to achieve proper distribution of incoming flow so that horizontal velocities are not excessive. The maximum length of rectangular tanks normally should not exceed 10 times the depth, but lengths up to 90 m (300 ft) have been used successfully in large plants. Where widths of rectangular tanks exceed 6 m (20 ft), multiple sludge collection mechanisms may be used to permit tank widths up to 24 m (80 ft). Regardless of tank shape, the sludge collector selected should be able to meet the following operational conditions: (1) the collector should have enough capacity so that when a high sludge-recirculation rate is desired, channeling of the overlying liquid through the sludge will not result, and (2) the mechanism

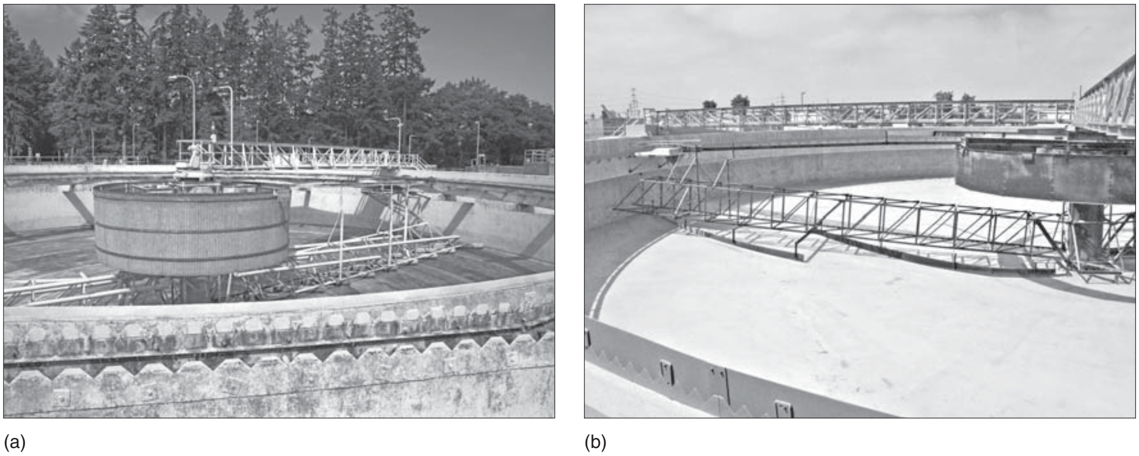


Figure 8–50

Typical circular sludge collection mechanisms: (a) suction-type and (b) spiral-type scrapers.

should be sufficiently rugged to transport and remove very dense sludges that could accumulate in the settling tank during periods of mechanical breakdown or power failure.

Two types of sludge collectors are commonly used in rectangular tanks: (1) traveling flights [see Fig. 8–51(a)] and (2) traveling bridges [see Fig. 8–51(b)]. Traveling flights are similar to those used for the removal of sludge in primary settling tanks. For very long tanks [see Fig. 8–52(a)], it is desirable to use two sets of chains and flights in tandem with a central hopper to receive the sludge to minimize the sludge transport distance [see Fig. 8–52(b)]. Sludge may be collected at the influent or effluent end of the tank. The traveling bridge, which is similar to a traveling overhead crane, travels along the sides of the sedimentation tank or on a support structure if several bridges are used. The bridge serves as the support for the sludge-removal system, which usually consists of a scraper or a suction manifold from which the sludge is pumped. The sludge is discharged to a collection trough that runs the length of the tank.

Other Types for Clarifiers. Other types of settling tanks that are used include stacked clarifiers, and tube and plate settlers (see Chap. 5). Stacked clarifiers (see Fig. 5–45 in Chap. 5) are used in installations where limited land area is available for clarifiers. Stacked clarifiers are used at the Deer Island Wastewater Treatment Plant in Boston, MA, for secondary sedimentation and were selected because of limited land area.

Clarification Tank Improvements. The efficiency of conventional or shallow clarifiers may be improved by the installation of tubes or parallel plates to establish laminar flow (see Fig. 5–25 in Chap. 5). Constructed of bundles of tubes or plates set at selected angles (usually 60°) from the horizontal, tube and plate settlers have a very short settling distance and circulation is dampened because of the small size of the tubes. Solids that collect in the tubes or on the plates tend to slide out due to gravitational forces. The major drawback in wastewater treatment is a tendency of the tubes and plates to clog because of the accumulation of biological growth, grease, and small objects that pass through coarse screens. Another drawback that can occur is if the characteristics of the MLSS change, the fixed angle of the plates or tubes may no longer be optimal.

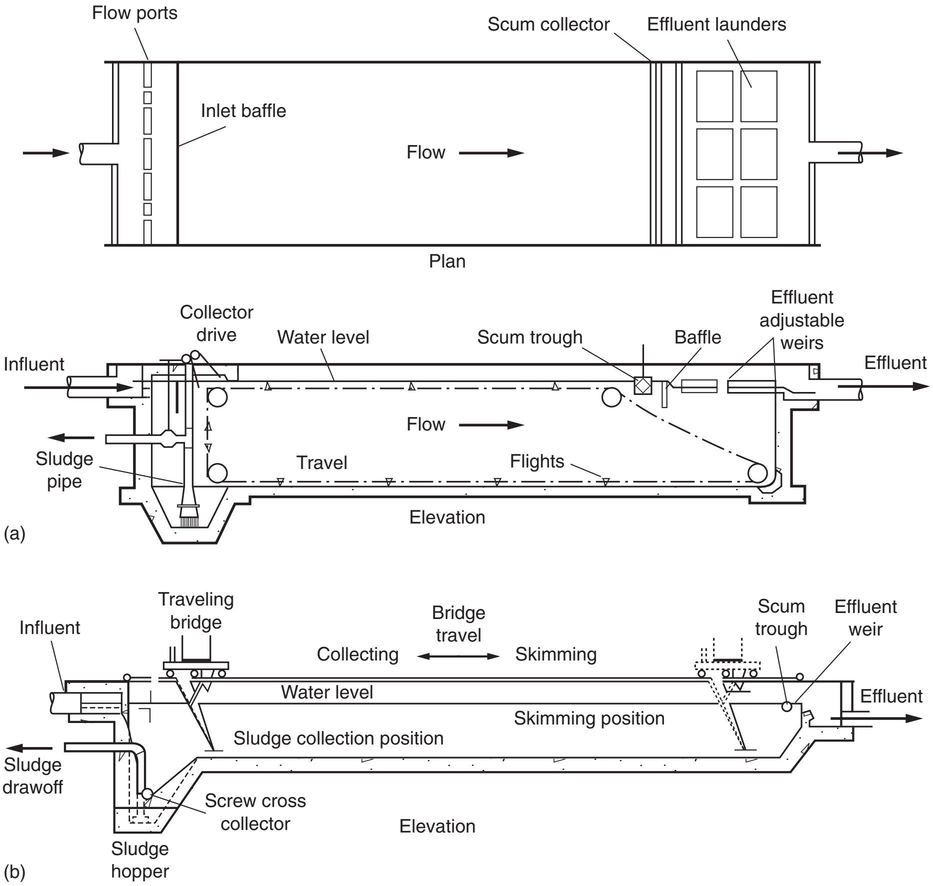


Figure 8–51

Typical rectangular sludge collection mechanisms: (a) chain and flight and (b) traveling bridge.

(a) (b)

Figure 8–52

Sludge-collection in long rectangular clarifiers: (a) chain and flight mechanisms are used to bring sludge to a central location from both ends of the clarifier where it is removed with a cross collector (coordinates 40.6430 N, 74.0343 W, view at altitude 750 m). The cross collector shown in (b) is located under central walkway dividing the clarifier. The flight mechanism on left side of divider also includes skimmers to collect floating material.

Sidewater Depth

Liquid depth in a secondary clarifier is normally measured at the sidewall in circular tanks and at the effluent end wall for rectangular tanks. The liquid depth is a factor in the effectiveness of suspended solids removal and in the concentration of the return sludge. Other factors such as inlet design, type of sludge-removal equipment, sludge blanket depth, and weir type and location also affect clarifier performance. In recent years, the trend has been toward increasing liquid depths to improve solids inventory capacity for periods of high flow variations and higher transient solids loading to the clarifier, which improves overall performance. Typical sidewater depths are presented in Table 8–34. Current practice favors a minimum sidewater depth of 4 to 5 m (13 to 16 ft) for large secondary clarifiers. Depths up to 6 m (20 ft) have been used. The cost of tank construction has to be considered in selecting a sidewater depth, especially in areas of high groundwater levels. Tanks with depths less than 3.5 m (~12 ft) often have difficulty containing the typically low-density activated sludge, and low-density sludge blankets are more easily disturbed by hydraulic fluctuations, especially morning diurnal variations. Deeper tanks therefore provide greater flexibility of operation and a larger margin of safety when changes in the activated sludge system occur.

Flow Distribution

Flow imbalance between multiple process units can cause under- or overloading of the individual units and affect overall system performance. In plants where parallel tanks of the same size are used, flow between the tanks should be equalized. In cases where the tanks are not of equal capacity, flows should be distributed in proportion to surface area. Methods of flow distribution to the secondary sedimentation tanks include weirs, flow distribution boxes, flow control valves, hydraulic distribution using hydraulic symmetry, and feed gate or inlet port control (see Fig. 8–53). Effluent weir control, although frequently used to effect flow splitting, is usually ineffective and should be used only where there are two tanks of equal size.

Tank Inlet Design

Poor distribution or jetting the influent to the clarifier can increase the formation of density currents and scouring of settled sludge, resulting in unsatisfactory tank performance. Clarifier inlets should be designed to dissipate the influent energy, distribute the flow evenly in horizontal and vertical directions, mitigate density currents, minimize sludge blanket disturbance, and promote flocculation. In circular center-feed clarifiers, a common design is to use small, solid-skirted, cylindrical baffles to dissipate the influent energy and distribute flow.

Figure 8–53

Alternative methods of flow splitting: (a) hydraulic symmetry, (b) flow measurement and feedback control, (c) hydraulic split with weirs, and (d) inlet feed gate control.

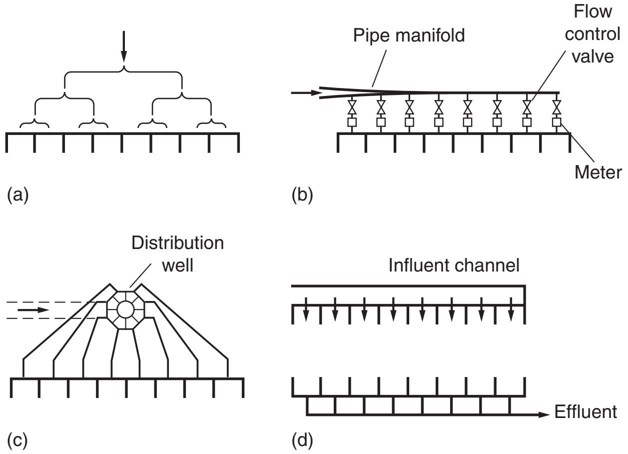
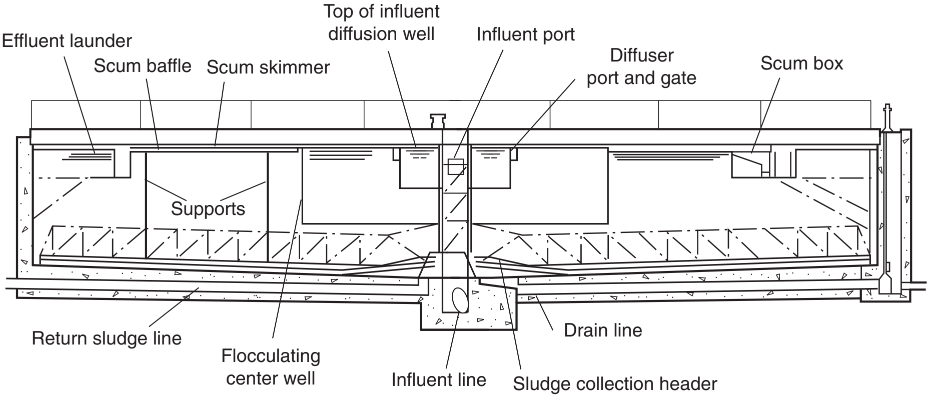


Figure 8–54

Typical secondary clarifier with a flocculating center feed well.



However, it has been found that a density current waterfall can be created in clarifiers using skirted baffles resulting in poor vertical flow distribution (Crosby and Bender, 1980). Methods to overcome these problems include the use of a large center diffusion well or a flocculating-type clarifier (see Fig. 8–54). The large center diffusion well, with a minimum diameter of 25 percent of the tank diameter, provides a greater area for dissipation of the influent energy and distribution of the incoming mixed liquor. The bottom of the feed well should end well above the sludge blanket interface to minimize turbulence and resuspension of the solids.

Flocculating center-feed clarifiers can incorporate an energy-dissipating inlet (EDI) and means to promote flocculation in the center-feed well [see Fig. 8–55(a)]. Typical flocculation feed wells have diameters of 30 to 35 percent of the tank diameter. An alternative device for dissipating energy, developed by the City of Los Angeles, is shown on Fig. 8–55(b). Operationally, the flow is discharged from a centerwell through a series of

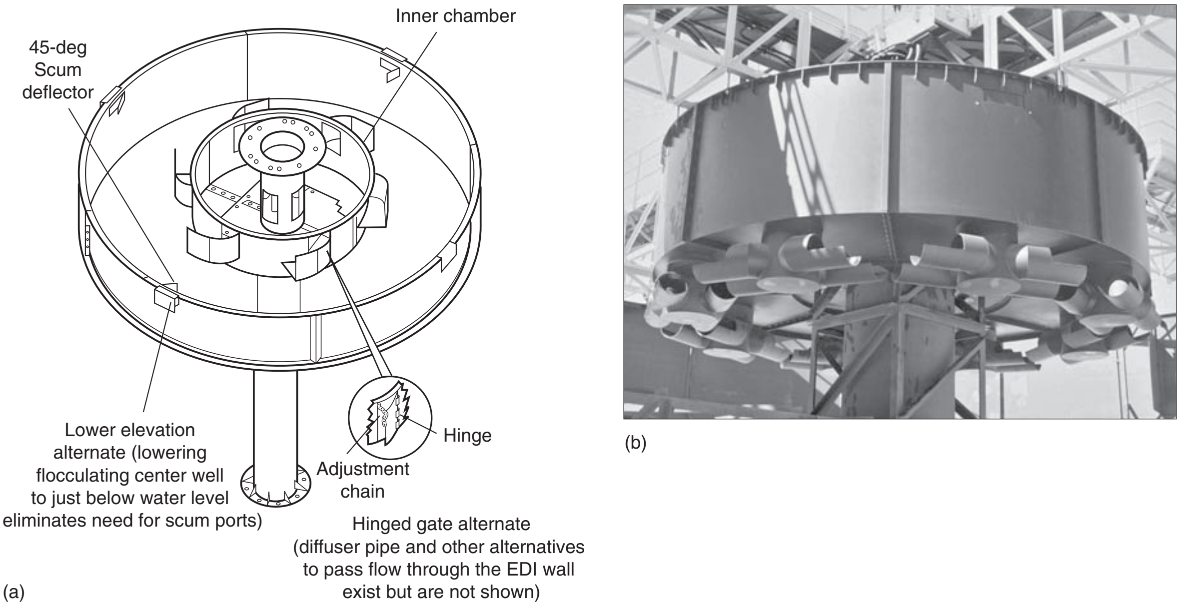


Figure 8–55

Energy-dissipating inlet devices used in circular clarifiers: (a) schematic of a center column energy-dissipating inlet and flocculating feed well (WEF, 1998) and (b) view of an energy-dissipating feed well (courtesy of the city of Los Angeles).

downward-facing discharge ports. By arranging the discharge ports so they discharge facing each other, the momentum energy is dissipated as the discharge streams impact each other. In rectangular tanks, inlet ports or baffles should be provided to achieve flow distribution. Inlet port velocities are typically 75 to 150 mm/s (15 to 30 ft/min) (WPCF, 1985).

Weir Placement and Loading

When density currents occur in a secondary clarifier, mixed liquor entering the tank flows along the tank bottom until it encounters a countercurrent pattern or an end wall. Unless density currents are considered in the design, solids may be discharged over the effluent weir. Experimental work performed by Anderson (1945) at Chicago on tanks approximately 38 m (126 ft) in diameter indicated that a circular weir trough placed at two-thirds to three-fourths of the radial distance from the center was in the optimum position to intercept well-clarified effluent. With low surface loadings and weir rates, the placement of the weirs in small tanks does not significantly affect the performance of the clarifier. Circular clarifiers are manufactured with overflow weirs located near both the center and the perimeter of the tank. If weirs are located at the tank perimeter or at end walls in rectangular tanks, a baffle should be provided to deflect the density currents toward the center of the tank and away from the effluent weir. Alternative baffle arrangements are shown on Fig. 8–56.

Weir loading rates are used commonly in the design of clarifiers, although they are less critical in clarifier design than hydraulic overflow rates. Weir loading rates used in large tanks should preferably not exceed 375 m^3/lin m·d (30,000 gal/lin ft·d) of weir at maximum flow when located away from the upturn zone of the density current, or 250 m^3/lin m·d (20,000 gal/lin ft·d) when located within the upturn zone. In small tanks, the weir loading rate should not exceed 125 m^3/lin m·d (10,000 gal/lin ft·d) at average flow or 250 m^3/lin m·d at maximum flow. The upflow velocity in the immediate vicinity of the weir should be limited to about 3.5 to 7 m/h (12 to 24 ft/h).

Scum Removal and Management

In many well-operating secondary plants, very little scum is formed in the secondary clarifiers. However, occasions arise when some floating material is present (see "Operating Problems" in Sec. 8–3), necessitating its removal.

Scum Removal by Skimming. Where primary settling tanks are not used, skimming of the final tanks is essential. Most designs in recent years provide scum removal for both circular and rectangular secondary clarifiers. Typical scum-removal equipment includes desert beach and scraper type, rotating pipe-through skimmer, and slotted pipes.

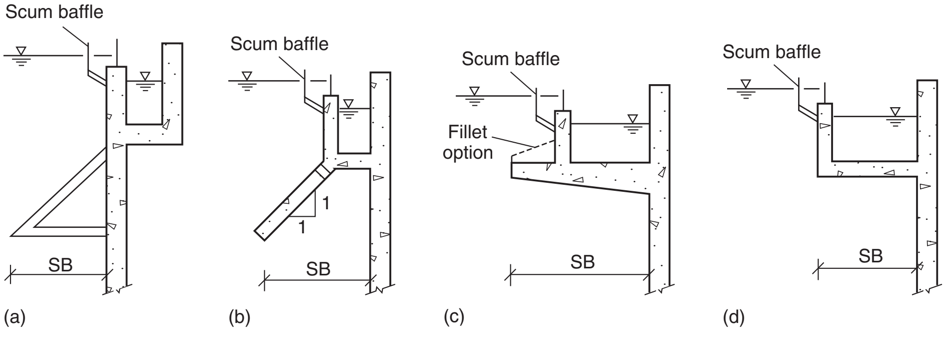


Figure 8–56

Alternative peripheral baffle arrangements: (a) Stamford, (b) unnamed, (c) McKinney (also known as the Lincoln baffle), and (d) Interior trough. (WEF, 1998.)

Note: SB will vary from 0.5 to 1.5 m, depending on the diameter of the clarifier

Scum Management. Scum should not be returned to the plant headworks because microorganisms responsible for foaming (typically *Nocardioform* such as *Gordonia amarae*) will be recycled, causing foaming problems to persist because of continuous seeding of the unwanted microorganisms. In some plants, scum is discharged to sludge-thickening facilities or is added directly to digester feed streams, as appropriate.

8–12 SOLIDS SEPARATION FOR MEMBRANE BIOREACTORS

In membrane separation, liquid-solids separation is accomplished by filtration or sieving. Water to be removed is withdrawn through a thin synthetic membrane which excludes colloidal and suspended solids because of the membrane pore sizes. In MBR systems the membranes are placed (immersed) in the activated sludge mixed liquor with a dedicated air supply system to provide a tangential flow across the membrane to prevent fouling by mixed liquor accumulation on the membrane surface. MBR technology background, applications, and advantages and disadvantages have been discussed previously in Sec. 8–1. Process configurations for biological nitrification and nitrogen removal, and enhanced biological phosphorus removal, with liquid solids separation using membranes in membrane bioreactors (MBRs) have been presented in Secs 8–6 and 8–7. The following topics are considered in this section: (1) design parameters for membrane separation in MBRs, (2) types of membranes and their characteristics, (3) membrane applications, (4) operating characteristics, (5) membrane fouling issues and (6) membrane fouling control methods.

Design Parameter

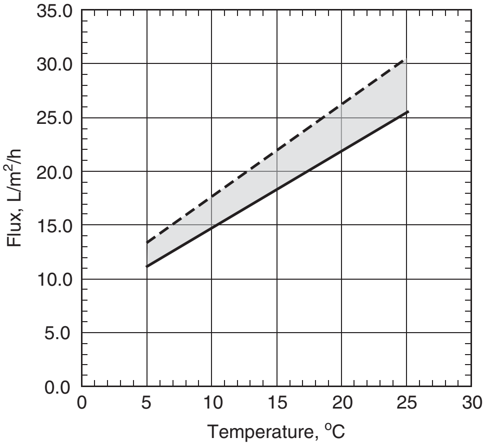
Key design and operating parameters for membrane separation are the membrane *flux* and *transmembrane pressure (TMP)*. The flux is the flowrate per unit area of membrane and is commonly expressed in $L/m^2 \cdot d$ or $gal/ft^2 \cdot d$. The higher the allowable flux, the lower is the membrane area required for a design flowrate. An acceptable pressure drop across the membrane or TMP is desired at the design flux. *Permeability* is a parameter used to reflect both flux and membrane pressure drop (TMP), and is the flux per unit of pressure driving force $[(L/m^2 \cdot h)/kPa]$. A decline in the permeability in a membrane system is typically due to membrane fouling.

The effluent from membrane separation is termed *permeate* and the remaining solids collected behind the membrane is called *retentate*. In the case of MBRs the retentate comprise solids in the return activated sludge flow and mixed liquor wasted. Recycle flowrates from the membrane separation zone are in the range of 4 to 6 times the influent flowrate to prevent excessive MLSS concentrations. Whereas solids thickening characteristics and the rate of solids and liquid applied per unit cross-section clarifier area (SOR and SLR) are key design parameters for the design of secondary clarifiers in activated sludge treatment, the liquid application rate across the membrane area (flux), pressure drop (TMP), and fouling issues apply to liquid-solids separation by membranes.

Membrane Flux. The membrane flux rate is a critical design parameter that is used to determine the required membrane surface area, membrane air scour supply requirements, and membrane tank volume. The flux is a function of the MBR MLSS concentration, temperature, TMP, and degree of membrane fouling. At a given TMP the flux is related inversely to viscosity, which increases at lower temperature and higher MLSS concentration (Trussell et al., 2007). There is, however, a tradeoff between using higher MLSS concentrations and smaller total tank volume for a given SRT, and having a greater membrane surface area.

Figure 8–57

Membrane flux decreases with decreasing temperature. The upper and lower lines represent flux values from low to high MLSS concentrations, respectively.



Although MBR systems can operate at very high MLSS concentrations (15,000 to 25,000 mg/L) (Cote et al., 1998), design MLSS concentrations in the range of 8000 to 12,000 mg/L appear to be most cost-effective when all factors are considered. A range of representative flux values is shown on Fig. 8–57 for average sustained design flows and at acceptable TMP settings. The bottom line with lower flux values applies to higher MLSS concentrations.

Because the MBR system must handle specified influent wastewater flowrates, the flux value is set for the expected controlling design condition of temperature, MLSS concentration, and acceptable TMPs, to determine the membrane available surface area. Typical operating TMPs are reported in Table 8–35 for hollow fiber and flat plate membranes. A higher pressure drop occurs for the membranes with smaller effective pore size.

In addition to the average design flux, the membrane suppliers set a peak allowable flux for 24 h or 6 h sustained peak flowrates. These higher transient flux values may be 1.5 to 2.0 times the average flux. Peak flow considerations affect the design and economics of MBR systems. In contrast to secondary clarifiers there is little economy of scale for higher flows. The required membrane area and tank volume is directly related to an average sustained flow or peak flow. Flow equalization basins are considered for handling high transient peak flowrates as an alternative to adding membrane surface area.

Membrane Properties

The types of membranes, membrane configurations, and membrane appurtenances are considered below.

Types of Membranes and Materials. Two types of membranes are used: (1) a hollow fiber membrane and (2) a flat sheet membrane or membrane plate. The membranes consist of a thin surface layer of a polymeric substance, with a high surface porosity and selected narrow pore size, over a thicker macro porous support structure, which provides structural strength and mechanical stability. The polymer materials used may be polyvinylidene difluoride (PVDF), polyethylene (PE), polyethylsulphone (PES), and polypropylene (PP), with proprietary manufacturing techniques and designs from different suppliers. The liquid flow through the membranes is termed *outside/in* with flow from the agitated mixed liquor on the surface of the membrane to the internal membrane structure from which the separated liquid is removed via a series of tubes or manifold system. The membrane pore sizes fall into either the range of microfiltration (MF) (0.01 to 0.40 μm) or ultrafiltration (UF) (0.01 to 0.10 μm). Microfiltration membranes can retain bacteria, while

Table 8–35

Design and operating characteristics of various proprietary MBR systems[a]

Manufacturer	**GE Zenon**	**Kubota**	**Mitsubishi**	**Siemens**	**Huber**
Membrane	Hollow fiber	Plate	Hollow fiber	Hollow fiber	Plate
Pore size, μm	0.04	0.4	0.04	0.04	0.04
Filtration type	UF	MF	UF	UF	UF
Configuration	Vertical	Vertical	Horizontal	Vertical	Rotating disks
Specific surface area, m^2/m^3	300	150	333	334	160
Location	In basin or separate cell	In basin or separate cell	Throughout	Separate cell	In basin or separate cell
Operational TMP, kPa	3–14	14–55	3–14	3–14	14–55
Fouling Control					
Pretreatment screening, mm	1–2	≤3	1–2	1–2	≤3
Air Scour					
Type	Coarse	Coarse	Coarse	Jet Aeration	Coarse
Air on/off, sec/sec	10/10*a*	Constant	Constant	Constant	Constant
Permeate on/off (relax), min/min	9.5/0.5	9/1	9/1	9/1	Constant
Permeate backpulse	Yes	No	Yes	Yes	No
Chlorine backpulse	1–2/wk	None	None	1–2/wk	None
Citric acid backflush	1/wk	None	None	None	None
Recovery Cleaning					
Frequency per year	2–3	2–3	3–4	3–4	As needed
Isolated membrane cell	Drained	In place	In place	Drained	In place
Application	Soak	Backwash	Backwash	Soak	Backwash
Chemical	Hypochlorite Citric acid	Hypochlorite Citric acid	Hypochlorite Citric acid	Hypochlorite Citric acid	Hypochlorite Citric acid

[a] Adapted in part from Yang et al. (2006), Babcock (2007), and Asano et al. (2007).

ultrafiltration can retain bacteria and viruses. However, permeate disinfection is normally employed for MBR applications, because the treated water is often used as reclaimed water and to also protect the effluent quality in case of a membrane failure or leak. The impact of fiber breakage is examined in Example 11–5 in Chap. 11.

Membrane Configurations. Individual membranes are contained in a *module* (also referred to as an *element*), which is a collection of membranes to be mounted as a unit. Another term to describe a membrane unit is a cassette, which is an assembly of membrane modules in a frame and with connections for permeate withdrawal and an air sparge system at opposite ends (WEF, 2006). The cassettes are standardized units with a set total membrane surface area. Membrane system manufacturers provide for air sparging at sufficient rates and locations to control fouling of the membranes by moving solids away from the membrane surface by a tangential flow while liquid is being drawn through the membrane. The distance between individual hollow fiber membranes or membrane plates is based on the flow needed across the membrane surface for fouling control and varies according to manufacturers' specifications. Hollow fiber and flat sheet membrane configurations are shown on Fig. 8–58.

The membrane specific surface relative to the cassette volume varies from 150 to 334 m^2/m^3 for different membrane designs, as reported in Table 8–35. The hollow fiber membranes have a greater specific surface area and thus require smaller membrane separation tank volumes. The additional volume needed to house the membrane cassettes varies with the membrane type and supplier designs, and may range from 3 to 10 m^3 tank volume/m^3 cassettes with the higher value used for hollow fiber membranes. On a membrane surface area basis the ratio ranges from 0.015 to 0.05 m^3 tank volume/m^2 of membrane surface area, with the smaller value applicable for hollow fiber membranes.

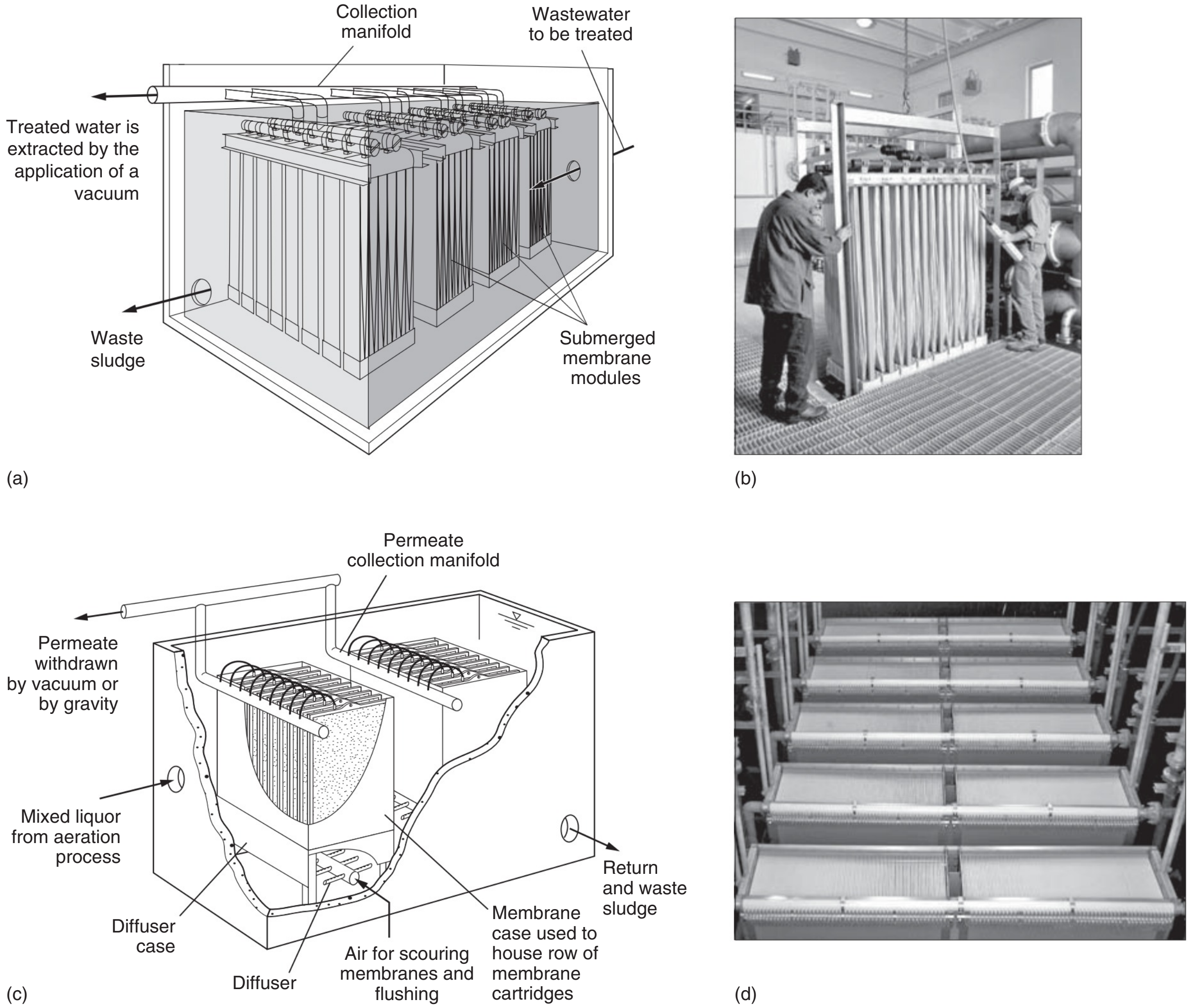


Figure 8–58

Example of membrane bioreactors: (a) schematic of placement of hollow fiber membrane bundles in an activated sludge reactor, (b) membrane bundle in position to be placed in a membrane bioreactor, (courtesy of Zenon environmental, Inc.), (c) schematic of placement of flat sheet membrane modules in an activated sludge reactor, and (d) flat sheet membrane modules in place with clear water.

Membrane Appurtenances. Because of the need to integrate the activated sludge system with the membrane flux and performance, and to provide proper membrane fouling control, the membrane suppliers provide other equipment items in addition to the membrane cassettes. Support facilities required include permeate pumps, chemical storage tanks, chemical feed pumps and all process controls for the membranes, including the motor control center for the membrane process. The membrane support equipment also includes an air-scour system and a back-pulse water-flushing system. The air-scour system consists of coarse bubble diffusers located in the aeration basin and provides continuous agitation on the outside of the membranes to minimize solids deposition. The air supply for the air-scour system is typically provided in addition to the activated sludge tank process air. Wastewater plant designers must first know the membrane system selected and design, before the final design of the activated sludge layout and configuration can be completed.

Membrane Design and Operating Characteristics

Design and operating characteristics for number of proprietary MBR systems representing most of the U.S. and world installations are summarized in Table 8–35. With the exception of the Mitsubishi designs, the membrane separation system may be contained in a separate tank or can be in a single aeration tank. The latter case is usually used for small facilities with coarse bubble aeration. When a separate tank is used for the membrane separation, coarse bubble aeration is only used in the membrane tank and fine bubble aeration is used in the preceding aerobic tank(s) for higher energy efficiency.

Membrane Usage

Currently (2012), the largest share of membrane installations utilize hollow fiber UF membranes, followed by the flat plate MF membranes (Yang et al., 2006). Hollow fiber membranes are typically about 2 m long with a 1.9 mm outer diameter, 0.8 mm inner diameter, and a spacing between membranes of about 3.0 mm. Some membranes are about 10 cm longer than the cassette height, which allows them to flex during air scour to provide further agitation against solids deposition. A typical UF cassette for a MBR system is shown on Fig. 8–58. The cassette shown on Fig. 8–58(b) is composed of hollow fiber membranes and has overall dimensions of 0.91 m wide by 2.13 m long, and approximately 2.44 m high (3 ft by 7 ft by 8 ft). Flat plate membranes [see Fig. 8–58(c) and (d)] consist of panels of 1.5 m wide by 0.55 mm high, a panel thickness of a 8 mm, and spacing between panels of 7 to 8 mm. A common application is to stack to panels, which reduces the tank area required and air flowrate for membrane fouling control (Judd, 2008b).

Membrane Fouling Issues

Membrane fouling is the reduction or loss of membrane performance due to the deposition of particulate or dissolved substance on the membrane surface or within the membrane pore structure (Koros et al., 1996). The effect of fouling is observed by an increase in TMP and thus decreased permeability for a give membrane flux. TMP is an important monitoring parameter to indicate when fouling exceeds acceptable levels in MBR systems to signal the need for special cleaning procedures to recover membrane performance. TMP values may remain stable and only slowly increase with time, but after some critical fouling point, they can increase rapidly (Gulglielmi et al., 2007). It appears this observed phenomenon is a two-step process in which gradual deposition occurs due to extracellular polymeric substances (EPS) substances, followed by cake formation and further headloss after fouling reaches a critical point. Note that fouling is different than membrane clogging or membranes becoming “sludged up,” which is the result of attempting to operate with excessive MLSS concentrations beyond the air scour hydrodynamic capacity to move mixed liquor off the membrane surface as permeate moves through it. Clogged membranes may need to be replaced or removed for special cleaning procedures.

Causes of Fouling. Fouling can be caused by physical, chemical and biological mechanisms as shown in Table 8–36. Some of the fouling agents are in the influent wastewater, such as hair and fibrous material, high alkalinity, soluble iron, and oil and grease. Oil and grease are very hydrophobic and can coat membrane material, but normally are not at levels of concern for most domestic wastewater flows. However, for smaller flows, where cooking oil is frequently used, or for systems receiving a high proportion of restaurant wastewaters without grease trap controls, it can be of great concern. Biofouling is a constant concern for MBR operations, but has been shown to be controllable in domestic wastewater treatment by operating within specified SRT and flux ranges, and maintaining the supplier's fouling control methods.

While there are conflicting literature reports on the effect of different microbial components contributing to fouling, it is generally agreed that macromolecules, colloids and soluble microbial products from bacteria growth and decay are major contributors to membrane fouling. Microbial fouling substances have in general been referred to as extracellular polymeric substances, which is a general term including autochthonous macromolecules, which include cell polymeric substances found outside bacteria cells and soluble material secreted by cells, such as proteins and polysaccharides (Judd, 2008a). Higher EPS production and increased fouling occurs at SRT values below 10 d (Trussell et al., 2006; Ke and Junxin, 2009). Less fouling was found at higher SRTs with good membrane performance at 30 and 50 day SRTs (Van den Broeck et al., 2012). MBR systems are generally designed with SRTs ≥ 10 d.

Table 8–36

Constituents that affect MBR membrane fouling and control methods[a]

Fouling type	Specific constituent	Control method
Physical	Hair, fibrous material, colloidal inorganic constituents, MLSS	Fine screening Air scour Relax operation Backpulse Chlorine
Chemical	High alkalinity,	Citric acid
	soluble iron,	Citric acid
	oil and grease	Chlorine Hypochlorite
Biological	Extracellular polymeric material	SRT control Air scour Chlorine Citric Acid
Biological	Colloidal organic matter	SRT control Backpulse Chlorine Citric Acid

[a] Refer to Table 11–25 for a broader range of fouling constituents and control methods.

Membrane Fouling Control Methods. The three principal methods used to control membrane fouling (see Table 8–35) are (1) pretreatment with fine screens, (2) operational fouling control procedures, and (3) recovery cleaning methods. Pretreatment by fine screening with sieve sizes in the range of 0.8 to 2 mm are needed to protect the membrane system from failure. If there is inadequate screening, hair and fibrous material can collect on membranes and on the air scouring diffusers to prevent proper cleaning and to cause a decline in the membrane flux capacity.

Operational Control of Fouling. Procedures to control fouling are provided by the membrane suppliers. Air scour by coarse bubble aeration systems that are designed specifically for each type of membrane by the respective supplier is critical to preventing solids accumulation on the membrane surface and fouling. Air scour rates are in the range of 3 to 12 L air/min·m^2 of membrane surface area. Some membrane systems do not require continuous air scour, but use a series of valves to alternate the air scour every 10 sec between two sets of membranes fed with a common blower (Palowski et al., 2007).

In addition to air scour, other fouling control methods are referred to as relax, backpulsing, maintenance cleaning, and recovery cleaning. Almost all of the membrane suppliers use a *relax* period in which the permeate withdrawal is stopped so that the air scour can wash solids off of the membrane surface during a period without solids application. A common relax strategy shown in Table 8–35 is 1 min of no permeate removal every 10 min. Some of the hollow fiber membrane suppliers reverse the flow through the membrane by a backpulse with permeate for about 0.50 min every 12 min to flush solids from the membrane.

Maintenance and Recovery Cleaning. Two types of cleaning regimens are used: (1) maintenance cleaning and (2) recovery cleaning. A maintenance cleaning involves back flushing the membrane with a sodium hypochlorite solution (200 mg/L) and/or citric acid (2000 mg/L) about 1 to 2 times per week. These are done in a series of repetitive alternative flushing with the chemical and permeate over a 60 to 75 min period. The combination of air scour, backflushing, and maintenance cleaning is not completely effective in controlling membrane fouling, and the pressure drop across the membrane increases with time. A recovery cleaning is done 2 to 4 times per year, or as needed, depending on the membrane type. It is required when TMP values remain elevated above the desired operating conditions. Recovery cleaning involves extensive chemical contact time, in the range of 4–6 h, and is done with chlorine (1000 mg/L) and citric acid (2000 mg/L), if needed for inorganic fouling. Some of the membrane systems have the recovery cleaning in place without draining the tank, while others drain the tank of mixed liquor and allow a soaking period with the chemical solutions.

PROBLEMS AND DISCUSSION TOPICS

8–1 Given the following laboratory test BOD and UBOD test results for one of the following wastewaters (to be selected by the instructor), determine the biodegradable COD (bCOD) concentration. Assume that values for f_d and Y_H are 0.15 g/g and 0.40 g VSS/g COD, respectively.

		Wastewater		
Test Parameter	Unit	1	2	3
BOD	mg/L	120	200	200
UBOD	mg/L	180	300	340

8–2 An influent wastewater sample is evaluated in a laboratory respirometer test to determine its readily biodegradable COD (rbCOD) concentration. The respirometer bottle is prepared by adding 500 mL of the influent sample with 500 mL of activated sludge mixed liquor. The respirometer records the accumulative oxygen consumption with time. The oxygen consumption occurs at a relatively constant rate initially (Phase A-total oxygen demand), and then the rate declines to another relatively constant rate (Phase B-oxygen demand for nitrification, slowly degradable COD, and endogenous decay), which continues for a number of hours. Finally, the oxygen consumption rate declines dramatically again and continues at a relatively constant rate (Phase C-oxygen demand for endogenous decay). The respirometer data are summarized in the following table for three different samples (to be selected by instructor). For the selected sample, determine the rbCOD concentration of the wastewater in mg/L, assuming the biomass yield for the heterotrophic bacteria is 0.45 g VSS/g COD and the oxygen equivalent of the biomass is 1.42 g COD/g VSS.

Phase	Duration of phase, h	Respirometer accumulative oxygen consumption for each phase, mg Wastewater 1	2	3
A	0.8	64	100	70
B	3.2	192	288	192
C	2.0	40	50	46

8–3 Primary effluent 24-h, flow proportioned composite samples are analyzed to determine its readily biodegradable COD concentration by using the flocculation-filtration COD (ffCOD) test method. The ffCOD is also measured for secondary effluent samples from the activated sludge treatment system, which has an aerobic SRT of 8 d. The average ffCOD values from 10 sampling days are shown in the table below. Determine the readily biodegradable concentration for one of the following wastewater plants (to be selected by the instructor).

Sample source	Unit	Wastewater 1	2	3
Primary effluent	mg/L	90	110	60
Secondary effluent	mg/L	30	20	30

8–4 For one of the following wastewater samples (to be selected by instructor) shown in the following table, with values for conventional wastewater characterization parameters, determine (a) the biodegradable COD concentration, (b) the slowly biodegradable COD concentration (c) the non-biodegradable volatile suspended solids (nbVSS) concentration, (d) the inert total suspended solids (iTSS) concentration, and (e) the average COD/VSS ratio. Assume that the bCOD/BOD ratio equals 1.6 and the activated sludge treatment effluent sCOD equals 30.0 mg/L.

Parameter	Unit	Wastewater 1	2	3
TSS	mg/L	220	170	90
VSS	mg/L	200	140	70
BOD	mg/L	200	160	120
rbCOD	mg/L	100	40	80
TCOD	mg/L	500	400	280
sCOD	mg/L	160	200	180

8–5 Given the following wastewater characteristics (to be selected by instructor), determine (a) the organic nitrogen, (b) the nonbiodegradable particulate organic nitrogen (nbpON), and (c) the biodegradable organic nitrogen (bON) concentrations.

		Wastewater		
Parameter	Unit	1	2	3
TKN	mg/L	40	45	50
NH_4-N	mg/L	25	30	35
Soluble organic nitrogen	mg/L	5.0	2.0	3.0
Nonbiodegradable soluble organic nitrogen	mg/L	1.0	1.0	1.0
VSS	mg/L	180	180	190
Nonbiodegradable VSS fraction	Percent	40	40	40

8–6 Using an observed yield value from Fig. 8–11, and Eqs. (8–19) and (7–56) (see Table 8–10), determine (a) the aeration tank volume in m^3 and (b) the amount of sludge wasted daily in kg TSS/d for an activated sludge system designed to treat a 6000 m^3/d wastewater flow with an influent BOD concentration of 120, 140, or 160 mg/L (as selected by instructor). The SRT is 6 d, the mixed-liquor temperature is 10°C, and primary treatment is used. What is the aeration tank volume and amount of daily sludge production if the SRT is increased to 12 d? Assume that the MLVSS and MLSS concentrations are 2500 mg/L and 3000 mg/L, respectively.

8–7 The following information is given for an activated sludge system design:

Parameter	Unit	Value
Flowrate	m^3/d	10,000
Influent BOD	mg/L	150
Effluent BOD	mg/L	2
τ	h	4
SRT	d	6
Synthesis yield, Y_H	g VSS/g bCOD	
Wastewater 1		0.40
Wastewater 2		0.50
Wastewater 3		0.30
Cell debris yield, f_d	g VSS/g VSS	0.15
Endogenous decay, b_H	g VSS/g VSS·d	0.08
nbVSS	mg/L	40
Temperature	°C	10

Note: Wastewater 1, 2, or 3 to be selected by instructor.

Assume no nitrification occurs due to the SRT selected and low temperature. Determine (a) the aeration tank oxygen requirements in kg/d, (b) the aeration tank oxygen uptake rate in mg/L·h, and (c) the aeration tank biomass concentration (mg/L). Assume bCOD = 1.6(BOD).

8–8 The following information is given for an activated sludge system designed with a long enough SRT to provide complete nitrification:

Parameter	Unit	Value
Flowrate	m^3/d	10,000
Influent BOD	mg/L	150
Effluent BOD	mg/L	2
Influent TKN	mg/L	35
Effluent NH_4-N	mg/L	1.0
τ	h	8.0
SRT	d	15
Temperature	°C	10
Cell debris yield, f_d	g VSS/g VSS	0.10
Synthesis yield, Y_H	g VSS/g bCOD	
Wastewater 1		0.40
Wastewater 2		0.50
Wastewater 3		0.30
Endogenous decay, b_H	g VSS/g VSS·d	0.08
Nitrifier yield, Y_n	g VSS/g NH_4-N	0.18
Nitrifier decay, b_n	g VSS/g VSS·d	0.12

Note: Wastewater 1, 2, or 3 to be selected by instructor.

Determine (a) the aeration tank oxygen requirements in kg/d, (b) the aeration tank oxygen uptake rate in mg/L·h, (c) the aeration tank biomass concentration (mg/L), and (d) the portion of the total oxygen required that is needed for nitrification.

8–9 Using Eqs. (8–20) and (8–21), compare the amount of sludge wasted daily as (a) VSS, (b) TSS, and (c) biomass for operation at an SRT of 10 and 20 d for the following wastewater and design conditions. Assume all of the influent TKN is biodegradable and complete nitrification to NO_3-N with no significant NO_2-N present (less than 0.10 mg/L). Repeat the calculation without accounting for cell debris. How much error is introduced?

Parameter	Unit	Value
Flowrate	m^3/d	15,000
Influent BOD	mg/L	200
Effluent BOD	mg/L	2
Influent TKN	mg/L	35
Effluent NH_4-N	mg/L	0.5
Heterotrophic yield, Y_H	g VSS/g bCOD	0.4
Heterotrophic decay, b_H	g VSS/g VSS·d	0.10
Cell debris yield, f_d	g VSS/g VSS	0.15
Nitrifier yield, Y_n	g VSS/g NH_4-N	0.18
Nitrifier decay, b_n	g VSS/g VSS·d	0.12
nbVSS	mg/L	
Wastewater 1		100
Wastewater 2		120
Wastewater 3		80
Temperature	°C	15

Note: Wastewater 1, 2, or 3 to be selected by instructor.

8–10 An activated sludge system consists of three completely-mixed reactors in series. The dissolved oxygen concentration in each reactor is 2.0 mg/L. Use the information given below for the biodegradable COD (bCOD) and NH_4-N concentrations in each stage for wastewaters 1 and 2. Determine the total oxygen transfer rate required for bCOD removal, nitrification, and endogenous decay. Prepare a table of the total oxygen transfer rate required and the amount for each of these 3 components in each stage in kg/h. Compare the percent of the total oxygen required for stages 1, 2, and 3. Assume that in Stage 3 the bCOD concentration is mainly as dissolved bCOD. Ignore the NH_4-N used for synthesis as this has already been accounted in determining the influent NH_4-N concentration available for nitrification. Assume that nitrification is complete to NO_3-N (i.e., NO_2-N concentration is insignificant).

		Wastewater	
Parameter	Unit	1	2
Flowrate	m^3/d	15,000	5,000
Recycle sludge ratio	unitless	1.0	0.5
Influent bCOD	mg/L	320	200
Influent available NH_4-N for nitrification	mg/L	30	35
Volume per stage	m^3	2300	500
Synthesis yield, Y_H	g VSS/g COD	0.45	0.45
Endogenous decay, b_H	g VSS/g VSS·d	0.10	0.10
STAGE 1			
bCOD	mg/L	30	50
NH_4-N	mg/L	8.0	17.0
biomass	mg VSS/L	1500	1200
STAGE 2			
bCOD	mg/L	5	8
NH_4-N	mg/L	3.0	6.0
biomass	mg VSS/L	1500	1200
STAGE 3			
bCOD	mg/L	0.5	2
NH_4-N	mg/L	0.2	2.0
biomass	mg VSS/L	1500	1200
Temperature	°C	15	

Note: Wastewater 1 or 2 to be selected by instructor.

8–11 An industrial wastewater from food processing is to be treated in an activated sludge process. The wastewater consists of soluble organics (no particulate degradable COD) with low nitrogen and phosphorus concentrations. For the summary of wastewater characteristics and design assumptions given in the following table, determine the amount of nitrogen and phosphorus that must be added to the influent flow as mg/L and in kg/d. Assume that residual effluent NH_4-N and soluble phosphorus concentrations of 0.10 mg/L are needed to prevent nutrient limitations, and no nitrification.

Parameter	Unit	Value
Flowrate	m^3/d	3000
Influent soluble bCOD	mg/L	
Wastewater 1		2000
Wastewater 2		3000
Wastewater 3		2500

(*continued*)

(*Continued*)

Parameter	Unit	Value
Effluent soluble bCOD	mg/L	5
Influent NH_4-N	mg/L	20
Influent phosphorus as P	mg/L	5
SRT	d	10
Synthesis yield, Y_H	g VSS/g bCOD	0.4
Heterotrophic decay, b_H	g VSS/g VSS·d	0.10
Cell debris yield, f_d	g VSS/g VSS	0.10

Note: Wastewater 1, 2, or 3 to be selected by instructor.

8–12 An activated sludge system is operating with the conditions described below for system 1, 2, or 3 to be selected by instructor. The conditions include the average influent flow, aeration tank volume and MLSS concentration, return activated sludge flow recycle ratio and TSS concentration, and the secondary clarifier effluent TSS concentration. Wasting is from the return activated sludge line leaving the bottom of the secondary clarifier.

(a) What should be the average daily waste sludge rate in m^3/d to maintain an SRT of 10 d?

(b) The plant operator decides to waste 1/10th of the aeration volume each day to maintain the SRT. What is the waste volume in m^3/d and what is the actual SRT?

		Activated sludge system		
Parameter	Unit	1	2	3
Flowrate	m^3/d	4000	10,000	5000
Aeration tank volume	m^3	2000	4000	5000
Aeration tank MLSS concentration	mg/L	3000	3500	3000
Clarifier effluent TSS concentration	mg/L	10.0	10.0	10.0
Return sludge recycle ratio	unitless	0.5	1.0	0.75
Return sludge TSS concentration	mg/L	9000	7000	7000

8–13 A membrane bioreactor consists of an anoxic tank followed by an aeration tank. The influent wastewater is added to the anoxic tank. The aeration tank is divided into two compartments. The first compartment is 75 percent of the total aerobic volume and is aerated with fine bubble diffusers, and the second compartment contains the membrane separation cassettes and is aerated with coarse bubble diffusers. The total aerobic volume is 4000 m^3. The return activated sludge flow is from the membrane separation compartment to the anoxic tank and is at six times the influent flowrate. The MLSS concentration in the membrane separation compartment is 12,000 mg/L. What is the daily waste sludge flowrate from the return activated sludge line, to maintain an SRT of 10 d?

8–14 A 2-L settleometer is used to perform an SVI test. The MLSS concentration for the test is 3500 mg/L and the settled sludge volume after 30 min is 840 mL. What is the sludge volume index in mL/g?

8–15 Using the ASM2d Model matrix described in Tables 8–12 and 8–13, write a mass balance equation for the second stage of a three stage aerobic activated sludge reactor for one or a combination of the following (1, 2, 3 or 4 or combination selected by instructor): (1) Heterotrophic bacteria (2) Autotrophic nitrifying bacteria (3) Slowly degradable substrate, and (4) Ammonia nitrogen.

8–16 A completely mixed activated sludge system treating domestic wastewater is operated at an SRT of 15 d at 12°C, such that complete nitrification occurs. The clarifier surface overflow rate is 1 m/h at average flow conditions, but the clarifier has a high sludge blanket with significant solids loss in the effluent. Describe (a) the specific steps you recommend to investigate the cause of the bulking sludge condition, (b) possible short-term immediate action to reduce the effluent TSS concentration, and (c) the selector alternatives that can be considered for bulking sludge control. Which one would you use and why?

8–17 Using the kinetic coefficients in Table 8–14 and assuming complete nitrification to NO_3-N (also ignore solids growth from nitrite oxidation), plot the following as a function of SRT (ranging from 3.0 to 20.0 d) for the municipal wastewater described in the table below. Assume that the MLSS concentration is 2500 mg/L and that all the TKN is biodegradable. The parameters to plot are (a) solids wasted as kg TSS/d, (b) aeration tank volume (m^3) and τ (h), (c) observed yield, as g TSS/g BOD, and g TSS/g $bCOD_r$, (d) effluent soluble bCOD concentration, (e) effluent NH_4-N concentration, and (f) oxygen requirements, kg/d.

Parameter	Unit	Value
Flowrate	m^3/d	20,000
BOD	mg/L	
Wastewater 1		220
Wastewater 2		250
Wastewater 3		180
bCOD/BOD	g/g	1.6
TSS	mg/L	220
VSS	mg/L	200
nbVSS	mg/L	
Wastewater 1		100
Wastewater 2		120
Wastewater 3		80
TKN	mg/L	40
Temperature	°C	15
Aeration tank DO	mg/L	2.0

Note: Wastewater 1, 2, or 3 to be selected by instructor.

8–18 Solve Problem 8–17 for a MBR system, assuming that the aeration tank is divided into two compartments. The volume of the second compartment for the membrane separation cassettes is 25 percent of the total aerobic volume. Assume that the MLSS concentration in the membrane separation compartment is 12,000 mg/L and the return activated sludge ratio is 6.0.

8–19 Solve Problem 8–17 using primary clarification, assuming 35 percent BOD removal, 65 percent TSS and VSS removal, 10 percent TKN removal, and 80 percent nbVSS removal.

8–20 An activated sludge system is to be designed for nitrification to achieve an effluent NH_4-N concentration of 0.50, 0.80, or 1.0 mg/L (to be selected by the instructor) at a reactor DO concentration of 2.0 mg/L and 10°C temperature. The peak/average TKN load is 1.8. Determine the design SRT. Use coefficients from Table 8–14.

8–21 A completely mixed activated sludge system has a hydraulic retention time, τ, of 8.3, 10.8, or 13.1 h (to be selected by the instructor), an aeration tank DO concentration of 2.0 mg/L, and an MLSS concentration of 3000 mg/L. For the municipal wastewater characteristics given below determine (a) the aeration tank average SRT, (b) the effluent NH_4-N concentration and (c) the nitrification safety factor if the desired average effluent NH_4-N concentration is 1.0 mg/L. Use coefficients from Table 8–14.

Parameter	Unit	Value
Temperature	°C	10
Flowrate	m^3/d	15,000
BOD removed	mg/L	130
nbVSS	mg/L	30
TSS	mg/L	70
VSS	mg/L	60
TKN	mg/L	40

8–22 A 3000 m^3/d industrial wastewater with a soluble COD concentration of 1800 mg/L is to be treated in a completely mixed activated sludge process at 15°C and 2500 mg/L MLSS concentration. Using the kinetic coefficients and assumptions provided below determine (a) the aeration tank volume in m^3, and τ in h, (b) the oxygen required in kg/d, (c) the sludge production in kg TSS/d, (d) the effluent sBOD concentration from the secondary clarifiers in mg/L, (e) the clarifier diameter in m, assuming two clarifiers, and (f) the air flowrate for fine bubble diffused air aeration. Assume very little excess NH_4-N exists after cell synthesis needs are met so that nitrification is not significant.

Parameter	Unit	Value
bCOD/BOD	g/g	1.6
μ_{max}	g VSS/g VSS·d	3.0
K_s	mg bCOD/L	60.0
Y_H	g VSS/g bCOD	0.40
b_H	g VSS/g VSS·d	0.08
f_d	g VSS/g VSS	0.15
SRT	d	
Wastewater 1		8.0
Wastewater 2		12.0
Wastewater 3		16.0
Alpha factor, α	Unitless	0.45
Fouling factor, F	Unitless	0.90
Beta factor, β	Unitless	1.0
Elevation	m	300
Effective DO saturation depth	m	2.5
Aeration tank liquid depth	m	5.0
Clean H_2O oxygen transfer efficiency	%	30

Note: Wastewater 1, 2, or 3 to be selected by instructor.

8–23 An oxidation ditch process is designed for the following wastewater using the conventional design approach of no primary treatment, a 24-h aeration tank detention time, and 3500 mg/L MLSS concentration. The lowest expected mixed-liquor temperature is 10°C and an average effluent NH_4-N concentration of 1.0 mg/L is required with a 1.5 safety factor for peak loads. Mechanical surface aerators are used to provide a DO concentration in the aeration zone of 2.0 mg/L. For one of the three wastewaters given below (to selected by instructor) determine (a) the SRT in d, (b) the sludge production in kg TSS/d, (c) the MLVSS concentration in mg/L, (d) the oxygen required in kg/d, (e) the total required aeration horsepower in kW, and (f) the ratio of the total volume provided to the necessary nitrification volume. Use coefficients from Table 8–14 and assume that all the TKN is biodegradable.

		Values		
Parameter	Unit	Wastewater 1	Wastewater 2	Wastewater 3
Flowrate	m^3/d	4000	4000	4000
BOD	mg/L	270	250	200
nbVSS	mg/L	130	120	100
TSS	mg/L	250	230	200
VSS	mg/L	240	215	180
TKN	mg/L	40	40	40
Clean H_2O oxygen transfer efficiency	kg O_2/kWh	0.9	0.9	0.9
Alpha factor, α	Unitless	0.90	0.90	0.90
Beta factor, β	Unitless	0.98	0.98	0.98
Elevation	m	500	500	500

Note: Wastewater 1, 2, or 3 to be selected by instructor.

8–24 For the same wastewater characteristics and mixed-liquor temperature as in Problem 8–23 and assuming uniform continuous flow, two sequencing batch reactor tanks are operated under the following conditions:
Fraction of tank liquid depth used for decant depth = 0.20
Aeration time = 2 h
Settle time = 1 h
Decant time = 0.5 h
Determine (a) the fill time in h, (b) the total time/cycle in h, (c) the total full volume of each tank in m^3, (d) the SBR SRT in d assuming an MLSS concentration of 3500 mg/L, and (e) the decant pumping rate in m^3/min.

8–25 A sequencing batch reactor treating the following wastewater is operated under the following conditions:
Temperature = 15°C
Aeration time/cycle = 2.0 h
Fill volume/total SBR tank liquid volume per cycle = 0.20
SRT = 20 d
DO = 2.0 mg/L
Two SBR reactor tanks are used and the full volume for each SBR tank = 3000 m^3.

		Values		
Parameter	Unit	Wastewater 1	Wastewater 2	Wastewater 3
Flowrate	m^3/d	4,800	4,800	4,800
BOD	mg/L	250	250	200
TKN	mg/L	45	40	30

Note: Wastewater 1, 2, or 3 to be selected by instructor.

Determine the effluent NH_4-N concentration in mg/L. Use kinetic coefficients values from Table 8–14.

8–26 An SBR tank has a total liquid depth of 5.5 m when full. The desired operating MLSS concentration is 3500 mg/L. If the SVI is 150, 180, or 200 mL/g (to be selected by instructor), determine the possible fill volume to total liquid volume ratio, assuming a 0.6 m clear liquid depth above the settled sludge blanket.

8–27 Repeat computations for Example 8–6 using the same design conditions and assumptions, but for an effluent NH_4-N concentration of 1.0, 2.0, and 4.0 mg/L (to be selected by instructor). How do the advantages of the staged system compare to the single-stage nitrification system as the desired effluent NH_4-N concentration is increased?

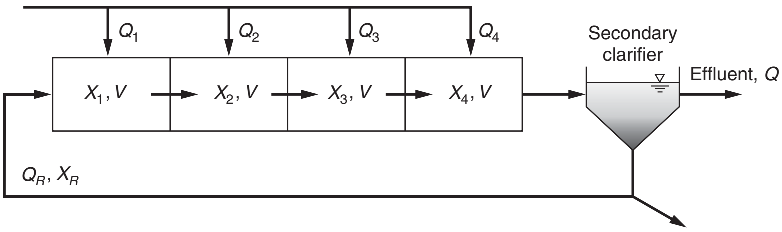
8–28 A four-stage activated sludge system with equal tank volumes of 240 m^3/stage is used to treat an industrial wastewater with a soluble BOD concentration of 300 mg/L. The influent flow is 4000 m^3/d and the RAS recycle ratio of 0.5. The active biomass concentration is 1600 mg/L. For the following biokinetic information determine: (a) the substrate concentration in each stage (mg/L as soluble bCOD), (b) the oxygen required per stage (kg/d), and (c) the percent of the total oxygen required per stage. Hint: to start the solution assume a 4th stage soluble bCOD concentration of 1.0 mg/L. (ignore nitrification).

Parameter	Unit	Value
Maximum specific substrate utilization rate, k	g COD/g VSS·d	1.2
Half velocity coefficient, K_s	mg bCOD/L	
Wastewater 1		50
Wastewater 2		75
Wastewater 3		100
Synthesis yield, Y_H	g VSS/g bCOD removed	0.35
b_H, endogenous decay	g VSS/g VSS·d	0.10

Note: Wastewater 1, 2, or 3 to be selected by instructor.

8–29 Compare the following activated sludge processes in terms of effluent quality, space requirements, complexity, energy requirements, operational requirements, and ability to handle variable flows and loads: completely mixed activated sludge, membrane bioreactor, contact stabilization, and oxidation ditch.

8–30 A four-pass step feed activated sludge system, shown in the figure below, has equal aeration tank volumes in each pass of 240 m^3. Using the design parameters given below, determine the MLVSS concentration in each tank.



Parameter	Unit	Value		
		Wastewater 1	Wastewater 2	Wastewater 3
X_R	mg VSS/L	10,000	10,000	10,000
Q_R	m^3/d	2000	4000	6000
Q	m^3/d	4000	4000	4000
Q_1	m^3/d	800	800	800
Q_2	m^3/d	1200	1200	1200
Q_3	m^3/d	1000	1000	1000
Q_4	m^3/d	1000	1000	1000

Note: Wastewater 1, 2, or 3 to be selected by instructor.

8–31 The following operating conditions can be used to describe an anoxic/aerobic process such as shown in Example 8–7 for 85 percent nitrogen removal.

Parameter	Unit	Value
Flowrate	m^3/d	1000
BOD	mg/L	200
rbCOD	mg/L	
Wastewater 1		60
Wastewater 2		95
Wastewater 3		120
Alkalinity	mg/L as $CaCO_3$	200
TKN	mg/L	35
Temperature	°C	15
MLSS	mg/L	3500
Biomass (VSS)	mg/L	1620
RAS (TSS)	mg/L	10,000
Aerobic volume	m^3	460
Aerobic SRT	days	10.0
Biomass nitrogen content	g N/g VSS	0.12
Effluent NH_4-N	mg/L	1.0

Note: Wastewater 1, 2, or 3 to be selected by instructor.

Determine (a) the internal recycle ratio and flowrate (m^3/d), (b) the anoxic tank volume and τ for a single-stage anoxic reactor, (c) the final alkalinity concentration, and (d) the oxygen required (kg/d) for the anoxic/aerobic process versus the aerobic system without the preanoxic tank. Use coefficients in Table 8–14.

8–32 Solve problem 8–31 for an anoxic/aerobic MBR with a two compartment aeration tank; a preaeration zone followed by a membrane separation zone. Use a return sludge recycle ratio from the membrane separation tank to the anoxic tank of 6.0 (i.e., calculate the resulting effluent NO_3-N concentration). Assume that the MLSS concentration in the membrane separation compartment is 10,000 mg/L. From Figure 8–52 the membrane flux is 20 L/m^2•h. Assume that the membrane compartment tank volume to membrane area ratio is 0.025 m^3/m^2.

8–33 An existing activated sludge system is operated at a minimal temperature of 10°C. The system is to be modified to an anoxic/aerobic process with the anoxic tank accounting for

10 percent of its total volume. For the following design wastewater conditions and total tank volume, determine (a) the effluent NH_4-N and NO_3-N concentrations, and (b) the minimal internal recycle ratio needed to match the nitrate removal capacity of the anoxic tank. Use necessary coefficients from Table 8–14.

Parameter	Unit	Value
Flowrate	m^3/d	8000
bCOD	mg/L	240
rbCOD	mg/L	
Wastewater 1		25
Wastewater 2		50
Wastewater 3		75
nbVSS	mg/L	60
TSS	mg/L	80
VSS	mg/L	70
TKN	mg/L	40
MLSS	mg/L	3500
Tank volume	m^3	3600
RAS ratio	Unitless	0.50
Biomass nitrogen content	g N/g VSS	0.12
Aeration tank DO	mg/L	2.0

Note: Wastewater 1, 2, or 3 to be selected by instructor.

8–34 An oxidation ditch system is operated with a single mechanical surface aerator such that one-half of the tank volume is aerobic with the DO concentration varying from 0 to 2.0 mg/L. For the information given below, determine the effluent NH_4-N (assuming 1.5 safety factor) and NO_3-N concentrations.

Parameter	Unit	Value
Ditch volume	m^3	4600
MLSS	mg/L	3500
Temperature	°C	10
BOD	mg/L	
Wastewater 1		250
Wastewater 2		220
Wastewater 3		200
nbVSS	mg/L	80
TKN	mg/L	40
TSS	mg/L	220
VSS	mg/L	210

Note: Wastewater 1, 2, or 3 to be selected by instructor.

8–35 Given the following wastewater characteristics, design an SBR system using two tanks, to produce an effluent NO_3-N and NH_4-N concentration of 6 mg/L and 1.0 mg/L, respectively. Assume that the aeration, settle, and decant times per cycle are 2.0, 1.0, and 0.5 h, respectively, and the MLSS concentration is 4000 mg/L. Determine (a) the fill volume fraction, (b) the volume of each SBR tank, (c) the decant pumping rate, and (d) the nitrification safety factor. Assume the aeration DO concentration is 2.0 mg/L and only anoxic mixing occurs during the fill period.

Parameter	Unit	Value
Flowrate	m^3/d	5000
BOD	mg/L	250
rbCOD	mg/L	50
nbVSS	mg/L	120
TKN	mg/L	
Wastewater 1		45
Wastewater 2		40
Wastewater 3		35
TSS	mg/L	220
VSS	mg/L	210
Temperature	°C	12

Note: Wastewater 1, 2, or 3 to be selected by instructor.

8–36 Exogenous carbon is to be added to a post anoxic tank in a Bardenpho process to remove 5 mg/L NO_3-N based on the influent flowrate to the tank. Compare the carbon dose needed in mg COD/L and mg/L of substrate for methanol, acetate, and ethanol. Assume a synthesis yield coefficient of 0.25, 0.40, and 0.36 g VSS/g COD removed under anoxic conditions for methanol, acetate, and ethanol, respectively. (Ignore the amount of exogenous carbon added that is contained in the effluent flow from the anoxic tank).

8–37 Acetate is to be added to a post anoxic tank in a Bardenpho process to decrease the NO_3-N concentration from an influent value of 6.0 mg/L to 0.30 mg/L in the anoxic tank effluent. For System 1, 2, or 3 below (to be selected by the instructor) determine (a) the acetate consumptive ratio in g COD/g NO_3-N removed, (b) the amount of NO_3-N reduced in mg/L (normalized to the tank influent flowrate) due to the substrate release and electron acceptor demand from endogenous decay, (c) the post anoxic tank effluent acetate concentration in mg/L, (d) the required acetate dose in mg COD/L and mg/L acetate (normalized to the influent flowrate), (e) the amount of acetate required per day in kg/d and the increase in NH_4-N concentration across the postanoxic tank.

		Values		
Parameter	Unit	System 1	System 2	System 3
Flowrate	m^3/d	5000	5000	5000
RAS recycle ratio	unitless	0.5	0.5	0.5
Influent NO_3-N	mg/L	6.0	6.0	6.0
Effluent NO_3-N	mg/L	0.3	0.3	0.3
Post anoxic volume	m^3	250	200	350
MLVSS	mg/L	3000	3000	3000
Biomass VSS	mg/L	1200	1200	1200
Temperature	°C	15	15	15
$b_H{}^a$	g VSS/g VSS·d	0.098	0.098	0.098
Acetate synthesis yield, Y_H	g VSS/g CODr	0.4	0.4	0.4

(*continued*)

(*Continued*)

Parameter	Unit	Values System 1	System 2	System 3
η	Unitless	0.80	0.80	0.80
Acetate $\mu_{max,20}{}^b$	g VSS/g VSS·d	4.46	4.46	4.46
Acetate μ_{max} temperature θ	unitless	1.21	1.21	1.21
Acetate, K_s	mg/L	5.0	5.0	5.0
NO_3-N K_{NO}	mg/L	0.1	0.1	0.1

[a] from Table 8–14.
[b] from Table 8–26.

8–38 An A^2O system for biological nitrogen and phosphorus removal is designed with a 15-d SRT. Assume an observed yield of 0.60 g TSS/g BOD removed and a net biomass yield of 0.30 g VSS/g BOD removed. The following wastewater characteristics following primary treatment have been obtained. The return activated sludge flow contains 5.0 mg/L NO_3-N and its DO concentration equals zero.

Parameter	Unit	Value
BOD	mg/L	
Wastewater 1		160
Wastewater 2		140
Wastewater 3		120
rbCOD	mg/L	
Wastewater 1		70
Wastewater 2		60
Wastewater 3		40
P	mg/L	7
TKN	mg/L	35
TSS	mg/L	82
VSS	mg/L	72
pH	units	7.2

Note: Wastewater 1, 2, or 3 to be selected by instructor.

Determine (a) the estimated effluent soluble phosphorus concentration for the biological phosphorus-removal process and (b) the phosphorus content (in percent on a dry weight basis) of the waste sludge. What process changes would you recommend to improve the amount of phosphorus removal?

8–39 An A^2O process is operated with a 12 d SRT and achieves the following effluent nutrient concentrations with internal and return activated sludge (RAS) recycle ratios of 3.0 and 0.5, respectively.
Effluent soluble P concentration = 0.50 mg/L
Effluent NO_3-N concentration = 5.0 mg/L
Assuming that the influent wastewater characteristics and SRT remain the same, determine (a) the change in effluent soluble phosphorus and NO_3-N concentrations if the internal recycle ratio is changed to 2.0, 2.5, or 2.8 (to be selected by instructor) and the RAS ratio is increased to 1.0 and (b) how much additional influent rbCOD is consumed by nitrate fed to the anaerobic zone.

8–40 Two 20-m diameter circular clarifiers are used for liquid-solids separation for an air activated sludge system. The MLSS concentration is 3000 mg/L and the return activated sludge recycle ratio is 50 percent. Assuming average solids loading rate values, within the range shown in Table 8–34, of 4, 5, or 6 kg/m^2·h (as selected by instructor), determine the allowable average influent flowrate in m^3/d and the return MLSS concentration.

8–41 State whether the effluent phosphorus concentration for an A^{2}O system will increase, decrease, or remain the same for each of the following changes in the wastewater characteristics or process operating conditions. Refer to basic process fundamentals to explain the basis for your answer.

a. The SRT is increased.

b. The influent rbCOD concentration increases.

c. The clarifier effluent suspended solids concentration increases.

d. A higher NO_3-N concentration exists in the return activated sludge recycle.

e. The influent particulate BOD concentration increases.

8–42 Given the settling data in the following table from tests with mixed liquor from an activated sludge plant, determine the percent RAS recycle rate if the clarifier overflow rate is 0.82, 1.0, or 1.2 m/h (to be selected by instructor) and the RAS MLSS concentration is 10,500 mg/L. What will the recycle rate be if the RAS MLSS concentration is 15,000 mg/L?

	Sludge liquid/solids interface depth in column test as a function of time and initial mixed-liquor concentration					
	MLSS concentration, mg/L					
Time, min	1000	2000	3000	5000	10,000	15,000
0	0	0	0	0	0	0
10	117.1	90.5	41.2	17.1	4.9	3.0
20	189.0	167.1	84.1	34.1	10.1	6.1
30	192.1	182.9	127.7	50.9	14.9	9.1
40	193.0	188.1	156.1	68.0	20.1	11.9
50	193.0	189.0	166.2	85.1	25.9	14.0
60	193.9	189.9	172.0	102.1	31.1	15.9

Note: Data in table corresponds to the distance from the top of the settling column to the sludge interface at indicated time, cm.

8–43 Two secondary clarifiers are operating with an overflow rate of 1 m/h and the activated sludge tank MLSS concentration is 4000 mg/L. Based on sludge thickening tests, the interfacial settling velocity can be described by the following relationship:

$$V_i = V_o(e^{-kX})$$

where V_i = interfacial settling velocity, m/d
X = MLSS concentration, g/L
V_o = 172 m/d
k = 0.4004 L/g

a. Plot the solids flux due to thickening as a function of the MLSS concentration in g/L.

b. On the same curve draw the overflow rate operating flux line and show the operating state point.

c. Determine the solids flux rate and recycle ratio for operating the clarifiers at an underflow concentration of 10, 11, or 12 g/L (as selected by instructor).

d. Determine the MLSS concentration possible with only one clarifier in operation using the above underflow concentration. Indicate the new overflow rate operating flux line and operating state point.

8–44 A conventional activated sludge process with an aeration tank volume of 4600 m^3 is operated with a 6-d SRT and 2500 mg/L MLSS concentration, when treating the following wastewater after primary treatment:

Parameter	Unit	Value
Flowrate	m^3/d	15,000
BOD	mg/L	150
nbVSS	mg/L	35
TSS	mg/L	80
VSS	mg/L	68
TKN	mg/L	35

The system must be upgraded to treat additional flow and to provide nitrification to produce an effluent NH_4-N concentration of 1.0 mg/L. A 12-d SRT is selected and membrane separation will replace the secondary clarifier operation, due to limited space for additional aeration tanks. Assuming the same wastewater characteristics, and using the same aeration tank volume with an MLSS concentration of 8000, 10,000, or 12,000 mg/L (to be selected by instructor), determine (a) the new wastewater treatment flowrate possible, (b) the previous and new volumetric BOD loadings (kg/m^3 d), and (c) the membrane surface area needed assuming a flux rate of 20.0 $L/m^2 \cdot h$.

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9

Attached Growth and Combined Biological Treatment Processes

WORKING TERMINOLOGY

Term	**Definition**
Attached growth aerobic process	Aerobic treatment processes in which the biomass responsible for treatment is attached to some type of medium (packing material).
Biofilm	A layer of biomass that develops on the surface of media used in attached growth biological processes.
Biological aerated filter (BAF)	An aerobic downflow or upflow submerged growth process that uses an inert media to support the growth of a biofilm for the removal of dissolved organic matter. Filtration also occurs for suspended solids removal.
Denitrification filter	An anoxic packed bed reactor that contains granular or synthetic media to support a biofilm for nitrate and nitrite reduction and to filter effluent suspended solids.
Diffusion limited substrate removal	The substrate removal rate in a biofilm is limited by the rate of diffusion across a stagnant layer between the biofilm and mixed bulk liquid and diffusion into the unmixed biofilm layer.
Fluidized-bed bioreactor (FBBR)	A submerged attached growth process that can be used for either aerobic or anaerobic applications where wastewater is fed upward to a bed of sand or activated carbon; the fluid velocity causes the bed to expand and form a biofilm that removes soluble organic matter.
Hybrid processes	Those processes that use a combination of attached-growth and suspended-growth processes.
Integrated fixed film activated sludge (IFAS)	An activated sludge process in which a portion of the biomass is maintained on fixed or suspended media that is separated from the reactor effluent by screening. Return activated sludge is required.
Moving-bed bioreactor (MBBR)	A submerged attached growth biological process in which most of the biomass is maintained on suspended media, which is separated from the reactor effluent by screening. No return activated sludge is used.
Packed bed filter	A process that makes use of biofilm microbial communities attached to fixed packing materials.
Oxygen limited substrate removal	Rate of substrate removal in a biofilm may be limited by the rate that oxygen can be supplied by diffusion from the bulk liquid. Higher bulk liquid DO concentrations than used in suspended growth processes must be provided to realize the full removal rate potential of an attached growth biofilm.
Recirculation rate	The rate at which treated wastewater is returned to the process influent to dilute the organic concentration and to provide more optimal hydraulics to maintain the viability of the microorganisms in nonsubmerged attached growth processes.

Term	Definition
Rotating biological contactor (RBC)	A fixed-film biological treatment device in which microorganisms are grown on circular plastic disks mounted on a horizontal shaft that rotates slowly while partially immersed in wastewater.
Submerged attached growth processes	Aerobic submerged fixed film processes involving three phases: a support material, the attached biofilm, and liquid to be treated.
Trickling filter	A nonsubmerged aerobic fixed-film biological reactor that uses rock or a plastic packing over which wastewater is distributed for treatment.
Trickling filter/activated-sludge process (TF/AS)	A sequential combined attached and suspended growth biological process in which the influent wastewater is treated in a highly loaded trickling filter prior to activated sludge treatment for production of a high quality effluent.
Trickling filter/solids-contact process (TF/SC)	A sequential combined attached and suspended growth biological process. The main differences between the TF/SC and TF/AS processes are in the trickling filter loading and activated sludge SRT values.

The concept of attached growth processes was introduced in Chap. 7, along with the fundamental mechanisms of mass transfer of substrate and electron acceptors into the biofilms that develop in attached growth systems. In this chapter, various aerobic attached growth processes used for wastewater treatment are introduced and discussed. Following a brief introduction to the general characteristics of attached growth processes, separate sections are devoted to (1) nonsubmerged attached growth aerobic processes, (2) sequential combined trickling filter-activated sludge processes, (3) activated sludge processes with various media for attached growth, (4) submerged attached growth aerobic processes with suspended media, (5) submerged fixed bed attached growth processes, and (6) attached growth processes used for biological denitrification.

9–1 INTRODUCTION TO ATTACHED GROWTH PROCESSES

To introduce attached growth processes, it will be helpful to review briefly the types of processes and their evolution, and to consider the importance of mass transfer on their performance.

Types of Attached Growth Processes

Attached growth aerobic processes can be grouped into five general categories as illustrated on Fig. 9–1: (1) nonsubmerged attached growth aerobic processes, (2) partially submerged attached growth aerobic processes, (3) sequential nonsubmerged attached growth-activated sludge process (4) submerged attached growth aerobic processes, and (5) activated sludge processes with biofilm carriers. Features that are common to each of these processes include:

1. Growth of the biomass on a fixed media
2. Removal of excess solids by sedimentation after solids sloughing off the fixed film media or by backwashing of the media
3. Need to provide oxygen by either air movement through the void volume in nonsubmerged processes or by air sparging into fixed or moving submerged media, or oxygenation of recycle flow in a fluidized bed reactor
4. Need to provide distribution and contact of the influent flow with the media surface area
5. Need for an underdrain or other methods of collecting the treated effluent

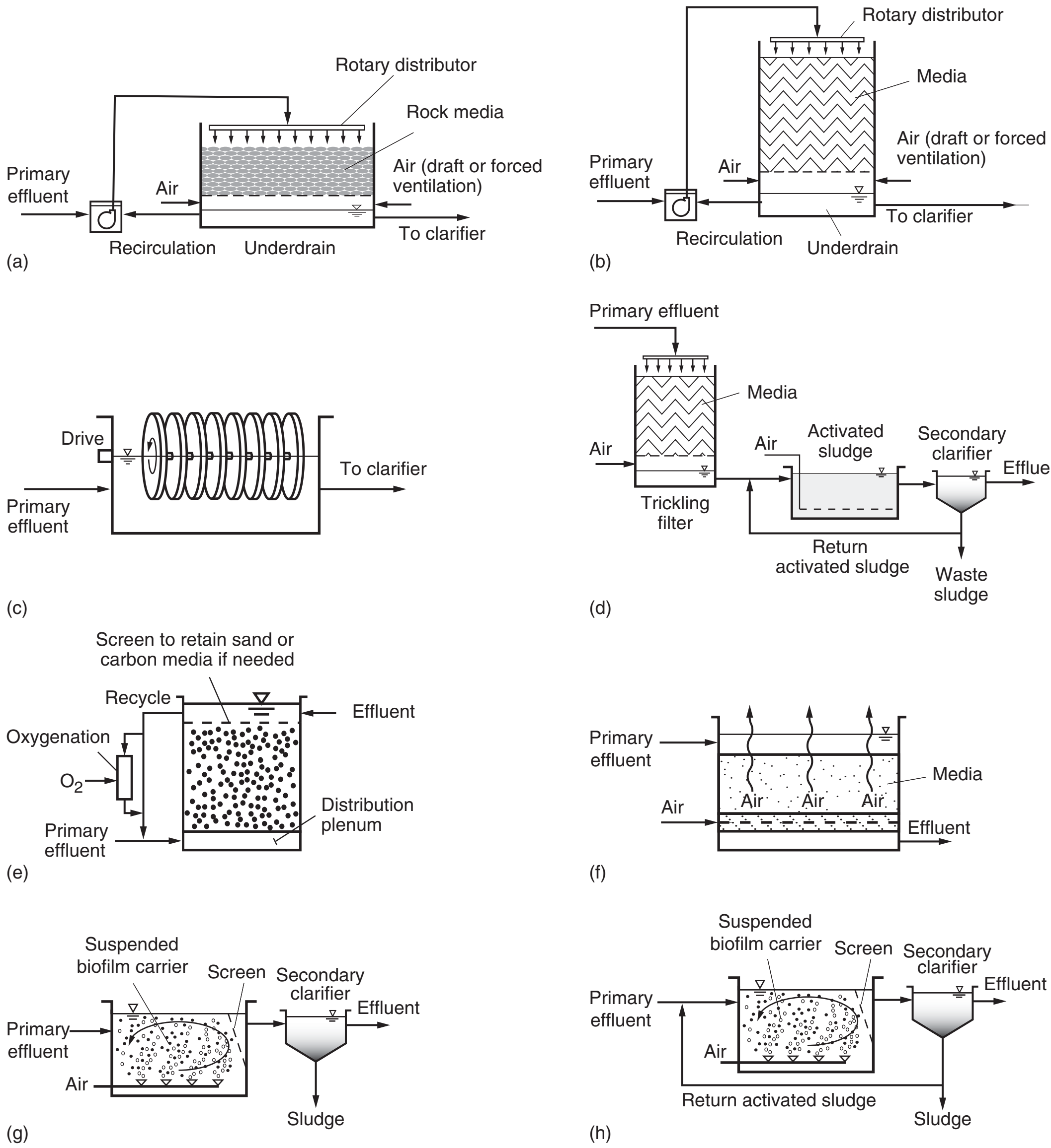


Figure 9–1

Schematics of common attached growth processes: (a) nonsubmerged attached growth shallow depth rock trickling filter, (b) nonsubmerged attached growth tower type trickling filter with plastic media, (c) nonsubmerged attached growth rotating biological contactor, (d) combined trickling filter/activated sludge process, (e) submerged upflow fixed media, (f) submerged downflow fixed media, (g) submerged moving bed attached growth bioreactor, and (h) combined submerged attached growth-activated sludge.

As the media size gets smaller, the area for biomass growth per unit of reactor volume increases and the reactor volume can become smaller. However, the oxygen transfer rate needed per unit volume increases, which cannot be supplied by forced air ventilation or natural draft air flow. Thus, a submerged media with air sparging is used in systems with media that have high surface area to volume ratios. A brief historical background is provided below for these processes.

Nonsubmerged Attached Growth Processes. The principal processes considered under this heading are the many variations of the trickling filter process [see Figs. 9–1(a) and (b)]. Trickling filters with rock packing [see Fig. 9–1(a)] have been a common, simple, and low-energy process used for secondary treatment since the late 1800s. A trickling filter is a non-submerged fixed film biological reactor using rock or plastic packing over which wastewater is distributed continuously. Treatment occurs as the liquid flows over the attached biofilm. The concept of a trickling filter grew from research at the Lawrence Experimental Station in Massachusetts, established in 1887, and the use of contact filters in England in the early 1890s (Alleman, 1982). Work at Lawrence confirmed wastewater treatment in an intermittent dosed sand filter, while treatment in watertight basins filled with broken stones with a cyclic operating mode was demonstrated in England. The bed was filled with wastewater from the top, and the wastewater was allowed to contact the packing for a short time. The bed was then drained and allowed to rest before the cycle was repeated. A typical cycle required 12 h (6 h of operation and 6 h of resting). The limitations of the contact filter included a relatively high incidence of clogging, the long rest period required, and the relatively low loading that could be used. Because of the clogging problems, larger packing was used until a rock size 50 to 100 mm (2 to 4 in.) was reached. Rotary distributor designs driven by water jets were developed at the Lawrence Experimental Station in 1894 to allow continuous uniform distribution of wastewater to the trickling filter (WEF, 2011).

In the early 1950s, Dow Chemical developed a plastic packing media, termed *Surfpak*, that enabled the use of taller filters (also known as biotowers), higher void volume, and better ventilation with less land area needs and improved process efficiency [see Fig. 9–1(b)] (Bryan, 1955). In the United States, plastic media has essentially replaced rock media in nonsubmerged attached growth processes; however, rock is still used in many parts of the world.

Partially Submerged Attached Growth Processes. In the mid-1960s, Allis Chalmers Corporation started investigating the use of metal rotating discs for wastewater treatment at the Jones Island treatment plant in Milwaukee. Soon after, they licensed a similar process termed *Bio-Disc* from a German company, which used polystyrene discs. However, in 1972 the process was passed to Autotrol Corporation, which introduced a new rotating biological contactor (RBC) media made from corrugated sheets of polyethylene. In the RBC attached growth process the packing is rotated in the wastewater treatment tank in contrast to pumping and applying the wastewater over a vertical static packing [see Fig. 9–1(c)].

The first RBC unit in the United States was for a small cheese factory in 1969 and its application became widespread in the United States through the 1970s (Alleman, 1982). However, problems with lower performance at design loadings, excess biomass accumulations on the discs due to the development of an undesirable type of biological growth, shaft breakage, and media deterioration on the disks has resulted in very few new installations over the last 20 years. Because of their relatively infrequent application, RBCs are not considered further in this chapter. The use of RBCs for the treatment of return flows is considered in Chap. 15. A complete discussion of the RBC process including design equations and examples may be found in the fourth edition of this textbook (Tchobanoglous et al., 2003). Process design guidelines are also provided in a U.S. EPA report on RBCs (U.S. EPA 1984).

Sequential Nonsubmerged Combined Attached Growth-Activated Sludge Process. Trickling filters have also been used before activated sludge in a coupled process to utilize the benefits of both processes, in terms of energy savings and effluent quality [see Fig. 9–1(d)]. The first sequential *trickling filter/activated sludge* systems were applied for the treatment of industrial wastewaters or high strength combined domestic/industrial wastewaters. The nonsubmerged attached growth process was used for partial BOD removal without solids clarification followed by activated sludge treatment for final treatment. The upstream attached growth process greatly improved the activated sludge settling characteristics with low SVIs, and the energy needed for BOD removal in the attached growth process was 20 to 40 percent of that needed per kg of BOD removal in the activated sludge process (Biesinger et al., 1980). Other combined trickling filter processes are considered in Sec. 9–3.

Another sequential trickling filter and activated sludge system is the *trickling filter/solids contact process*, which was developed in the late 1970s from research in Corvallis, OR, aimed at a higher quality effluent after trickling filter treatment of domestic wastewater (Norris et al., 1982). In this case the trickling filter was designed to remove most of the soluble BOD, and was followed by an aerated solids contact channel also receiving return activated sludge from secondary clarifiers. The primary purpose of the aerated solids contact channel was to flocculate trickling filter effluent suspended solids in the activated sludge.

Submerged Attached Growth Aerobic Processes. Beginning in the 1970s and extending into the 1980s, a new class of aerobic attached growth processes became established alternatives for biological wastewater treatment. These are upflow, downflow, and fluidized bed reactors with fixed or moving media that do not use secondary clarification or sparged aeration reactors. Work by Jeris et al. (1977) developed an upflow fluidized bed reactor that was applied for BOD removal [see Fig. 9–1(e)] and also for denitrification in an anoxic reactor. The first downflow submerged fixed media bed system [see Fig. 9–1(f)] using a fired clay media was developed from research work in Paris, France (Leglise et al., 1980). Later, upflow submerged fixed media bed systems were developed. In the late 1980s a moving bed bioreactor (MBBR), containing a mixed and aerated high density polyethylene biofilm media in a submerged horizontal flow tank [see Fig. 9–1(g)] was developed in Norway (Ødegaard, 2006).

The unique advantage of the submerged attached growth system is the small footprint required with an area requirement that is a fraction (one-fifth to one-third) of that needed for activated sludge treatment. Other advantages compared to activated sludge are the ability to handle dilute wastewaters and the avoidance of activated sludge settling concerns. Though they are more compact, their capital costs are generally higher than that for activated sludge treatment. In addition to BOD removal, submerged attached growth processes have also been used for tertiary nitrification and denitrification following suspended or attached growth nitrification.

Activated Sludge Processes With Fixed Film Carriers. The placement of material for attached growth in the aeration tank of the activated sludge process dates back to the 1940s with the Hays and Griffith processes (WEF, 2011), in which cement asbestos or wood baffles were added to activated sludge aeration tanks. Present day designs use more engineered materials and include the use of synthetic media that are suspended in the aeration tank with the mixed liquor, fixed synthetic material placed in portions of the aeration tank, and submerged RBCs. This combined attached growth/activated sludge process [see Fig. 9–1(h)] has also been referred to as a *hybrid* process or *integrated fixed-film activated sludge* (IFAS) process. The advantages claimed for these activated sludge process enhancements are as follows:

1. Increased treatment capacity
2. Greater process stability

3. Reduced sludge production
4. Enhanced sludge settleability
5. Reduced solids loadings on secondary clarifier
6. No increase in operation and maintenance costs

Mass Transfer Limitations in Attached Growth Processes

A significant process feature of attached growth processes in contrast to activated sludge treatment is the fact that the performance of biofilm processes is often diffusion limited. Substrate removal and electron donor utilization occurs within the depth of the attached growth biofilm and subsequently the overall removal rates are a function of diffusion rates and the electron donor and electron acceptor concentrations at various locations in the biofilm. By comparison, the process kinetics for the activated sludge process are generally characterized by the bulk liquid concentrations.

The diffusion-limited concept is especially important when considering the measurable bulk liquid DO concentrations on attached growth process biological reaction rates. Where a DO concentration of 2 to 3 g/m^3 is generally considered satisfactory for most suspended growth aerobic processes, such low DO concentrations can be limiting for attached growth processes. For uninhibited nitrification in the biofilm a much higher DO concentration may be required, as shown in Sec. 7–7, depending on the ammonia-N concentration.

The concept of diffusion limitations on nitrification rates and the ability to develop anaerobic layers within the biofilm may be exploited to accomplish both nitrification and denitrification in attached growth processes with positive bulk liquid DO concentrations. Investigators have shown how aerobic and anaerobic layers can be developed in the biofilm to accomplish nitrogen removal by nitrification and denitrification (Chui et al., 1996, Richter et al., 1994, and Meaney et al., 1994).

9–2 NONSUBMERGED ATTACHED GROWTH PROCESSES

Trickling filters and variants are the principal nonsubmerged attached growth processes used for aerobic biological treatment of wastewater. The objective of this section is to consider the fundamentals of the trickling process, important design considerations, and the implementation of trickling filters for BOD removal and nitrification. Topics considered in this section include (1) general process description, (2) trickling filter classification and application, (3) advantages and disadvantages of trickling filters, (4) physical facilities for trickling filters, (5) operational considerations (6) process analysis for BOD removal, (7) process analysis for BOD removal and nitrification design, and (8) process analysis for tertiary nitrification design. The first series of topics (1 through 5) deal with a general description of the characteristics and physical features of trickling filters. The focus of the last three topics (6, 7, and 8) is with the important process analysis considerations for trickling filters and their application for the removal of BOD and nitrification with or following BOD removal.

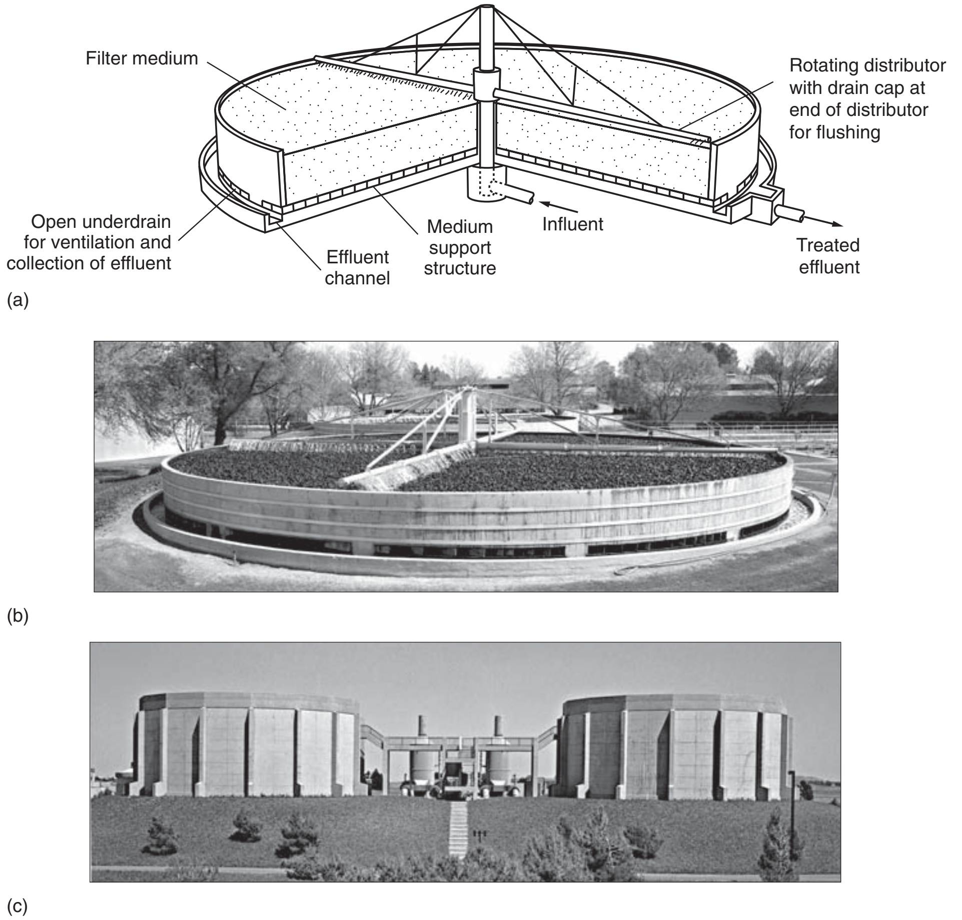
General Process Description

As noted previously, the trickling filter is a nonsubmerged fixed film biological reactor using rock or plastic packing over which wastewater is distributed continuously. The physical features and functional description of the trickling filter process are introduced in this section, and considered further subsequently.

Physical Features. The three principal features of trickling filters, as illustrated on Fig. 9–2(a), are (1) the filter media, (2) the wastewater distribution system, and (3) the underdrain system.

Figure 9–2

Typical trickling filters: (a) cutaway view of a rock trickling filter (b) view of conventional rock-filled trickling filter, and (c) view of tower trickling filters with plastic filter media.



Filter Media. The ideal filter packing is a material that has a high specific surface area (m^2 exposed surface/m^3 bulk volume), is low in cost, has a high durability, has a high enough porosity to avoid clogging, and allows good air ventilation by natural draft or low pressure blowers. The media serves as a structure for the attachment and growth of a biofilm. Treatment occurs as the liquid flows over the attached biofilm. Many conventional trickling filters using rock as the packing material [see Fig. 9–2(b)] have been converted to plastic packing to increase treatment capacity. Virtually all new trickling filters in the United States are now constructed with plastic packing in a tower arrangement [see Fig. 9–2(c)]. Primary clarification is used before application of wastewater to the trickling filter media to prevent clogging, although in some cases primary clarification has been omitted where fine screens are used.

Distribution System. The distributor is used to apply the wastewater to be treated uniformly over the surface of the filter media. The flow-driven rotary distributor for trickling filtration has become a standard for the process because it is reliable and easy to maintain. A distributor consists of two or more arms that are mounted on a pivot in the center of the filter and revolve in a horizontal plane [see Fig. 9–2(b)]. The arms are hollow and contain nozzles through which the wastewater is discharged over the filter bed. The distributor assembly may be driven either by the dynamic reaction of the wastewater discharging from the nozzles or by an electric motor.

Underdrain System. The functions of the underdrain system are twofold: (1) to collect wastewater that has passed through the filter, along with any biofilm that has sloughed off from the media, and (2) to provide an open area for the movement of air, either by natural draft or by forced air, needed for treatment of the applied wastewater.

Functional Description. Functionally, trickling filter media is covered with a very visible *biofilm* as a result of the BOD removal. In the literature, the biofilm is often identified as a *slime layer (*also a *zoogleal film)*, especially in those cases where an especially thick slimy biofilm layer develops. Soluble substrates and dissolved oxygen diffuse into the biofilm to support biomass growth. During the time when the wastewater flow is not applied or the *resting period*, oxygen continues to diffuse into the biofilm. Colloidal and particulate substances are removed by attachment and trapping in the biofilm. Effluent solids that become detached from the biofilm support are removed in a secondary clarifier, but clarifier underflow solids are not recycled to the trickling filter as they are in the activated sludge process. Recirculation of the trickling filter effluent to the trickling filter influent is done to maintain wetting of the media at low flows, and because it also enhances the supply of oxygen and treatment performance.

Biofilm Development. Depending on operating conditions, the thickness of the biofilm can reach 10 mm. In the outer portions of the biofilm (0.1 to 0.2 mm), the organic material is degraded by aerobic microorganisms. As the microorganisms grow and the biofilm thickness increases, oxygen is consumed before it can penetrate the full depth, and an anaerobic environment is established near the surface of the packing. As the biofilm thickness increases, the substrate in the wastewater is used before it can penetrate the inner depths of the biofilm, where the biofilm then enters an endogenous respiration state and lose their ability to cling to the packing surface.

Sloughing. When the microorganisms can no longer cling to the media, the applied liquid washes the biofilm off the packing, and a new biofilm layer starts to grow. The phenomenon of losing a portion of a biofilm layer is called *sloughing* and is primarily a function of the organic and hydraulic loading on the filter. The hydraulic loading affects the shear velocities, and the organic loading affects the rate of metabolism and rate of biofilm growth. Methods to control the hydraulic loading and sloughing are discussed subsequently.

Process Microbiology. The biological community in the biofilm is very diverse and includes aerobic and facultative bacteria, protozoan, fungi, and algae. Higher life forms, such as worms, insect larvae, and snails, are also present. Facultative bacteria are the predominating organisms in trickling filters, and decompose the organic material in the wastewater along with aerobic and anaerobic bacteria. *Achromobacter, Flavobacterium, Pseudomonas,* and *Alcaligenes* are among the bacterial species commonly associated with the trickling filter. The fungi present are also responsible for waste stabilization, but their role is usually important only under low-pH conditions or with certain industrial wastes. At times, their growth can be so rapid that the filter clogs and ventilation becomes restricted. Among the fungi species that have been identified are *Fusazium, Mucor, Pencillium, Geotrichum, Sporatichum*, and various yeasts (Hawkes, 1963; Higgins and Burns, 1975). For low-loaded trickling filters where little organic substrate remains nitrifying bacteria will be established on the media. At higher organic substrate concentrations the heterotrophic bacteria outcompete the nitrifying bacteria for growth on the media surface area due to their faster growth rates and higher biomass yields.

Filamentous Bacteria. When adverse dissolved oxygen conditions prevail within the biofilm due to high organic loading, the filamentous bacteria forms *Sphaerotilus natans* and *Beggiatoa* will be found. The *Beggiatoa* are favored due to their ability to oxidize hydrogen sulfide and other reduced organic substances that can be produced within anaerobic layers in the depth of the biofilm.

Presence of Higher Forms. Higher life forms, such as protozoa, worms, snails, and insects feed on the biological film. The protozoa in the filter are predominantly of the ciliate group, including *Vorticella, Opercularia*, and *Epistylis* (Hawkes, 1963; Higgins and Burns, 1975). Their function is to feed on the biological films and free bacteria to decrease effluent turbidity and help keep the film in a higher growth state. Snails are especially troublesome in trickling filters used mainly for nitrification, where they have been known to consume enough of the nitrifying bacteria to significantly reduce treatment efficiency (Timpany and Harrison, 1989).

Presence of Algae. Algae can grow only in the upper reaches of the filter where sunlight is available. *Phormidiun, Chlorella*. and *Ulothrix* are among the algae species commonly found in trickling filters (Hawkes, 1963; Higgins and Burns, 1975). Generally, algae do not take a direct part in waste degradation, but during the daylight hours they add oxygen to the percolating wastewater. From an operational standpoint, the algae may be troublesome because they can cause clogging of the media surface, which produces odors.

Trickling Filter Classification and Applications

Trickling filters are often classified according to the applied organic loading rates expressed as kg BOD applied/m^3 bulk media volume·d (in US Customary units as lb BOD/1000 ft^3·d). Trickling filter applications and loadings, based on historical terminology developed originally for rock filter designs, are summarized in Table 9–1. As shown in Table 9–1, trickling

Table 9–1

Trickling filter process applications and representative design parameter when treating primary effluent[a]

Design parameter	Unit	Low rate BOD removal	High rate BOD removal	High rate BOD removal	BOD removal and nitrification	Partial BOD removal
BOD removal efficiency	%	80–90	80–90	70–90	85–90	40–70
Type of packing		Rock	Rock	Plastic	Plastic (P)/rock (R)	Plastic
Ventilation	Type	Natural	Forced air	Forced air	Forced air	Forced air
Organic loading	kg BOD/m^3·d (lb BOD/10^3 ft^3·d)	0.08–0.3 (5–20)	0.6–1.6 (40–100)	0.6–2.4 (50–150)	0.08–0.4 (5–25)	1.6–3.5 (100–220)
Hydraulic loading	m^3/m^2·d (gal/ft^2·d)	1–4 (25–100)	4–40 (100–1000)	15–75 (350–1850)	5–16 (125–400)	40–100 (1000–2500)
Recirculation ratio	Q_R/Q	0–1	1–2	1–2	1–2	0–2
Depth	m (ft)	1–2.5[b] (3–8)	1–2.5[b] (3–8)	3–12 (8–40)	P, 3–12 (8–40) R, 1–2.5 (3–8)	0.9–6 (3–20)
Effluent quality	BOD, mg/L	<30	<30	<30	<20	>30
	NH_4-N, mg/L	<5	>5	>5	<3	

[a] Adapted in part from Tchobanoglous et al. (2003) and Daigger and Boltz (2011).

[b] Depth is limited due the weight of the rock.

Note: kg/m^3·d × 62.4280 = lb/10^3 ft^3·d

m^3/m^2·d × 24.5424 = gal/ft^2·d

filter applications and the corresponding loading rates will vary depending on whether the treatment objective is (a) BOD removal with natural draft aeration, (b) BOD removal with forced draft aeration, (c) BOD removal and nitrification with forced draft aeration, and (d) partial BOD removal, or *roughing* treatment with forced draft aeration. Each type of loading is considered briefly in the following discussion.

BOD Removal with Natural Draft Aeration. Historically, rock trickling filters have been designed and operated with natural draft aeration. Because of low air movement during parts of the day when the temperature of the wastewater and air are within 1.7°C (3°F) of each other, the organic loading rates were low. These filters are often identified in the literature as *low rate* filters. Although relatively simple, a low rate filter is a highly dependable device that produces an effluent of consistent quality with an influent of varying strength. Low-rate filters with rock as the filter medium are usually operated without recirculation, although some recirculation has been used to keep the filters moist during low flow periods. Circular filters are the most common shape but rectangular and polygonal shapes have been used. In some cases, filters have been constructed without a surrounding enclosure [see Fig. 9–3(a), see also Fig. 7–15 in Chap. 7]. Often rock is replaced with plastic packing material [see Fig. 9–3(b)].

At lower organic loading rates, in the range of 0.07–0.25 kg $\text{BOD/m}^3\cdot\text{d}$, secondary treatment is achieved with 85 to 90 percent BOD removal. In most low-rate filters, only the top 0.6 to 1.2 m (2 to 4 ft) of the filter packing will have appreciable biofilm. As a result, the lower portions of the filter may be populated by autotrophic nitrifying bacteria, which oxidize ammonia nitrogen to nitrite and nitrate forms. Heterotrophic bacteria, with higher yield coefficients and faster growth rates, are more competitive than nitrifying bacteria for space on the fixed film packing. Thus, significant nitrification only occurs after the BOD concentration is reduced appreciably. Based on soluble BOD, Harremöes (1982) concluded that a concentration of less than 20 mg/L is needed to initiate nitrification. Low rate filters are seldom used these days, primarily due to odor issues and space constraints.

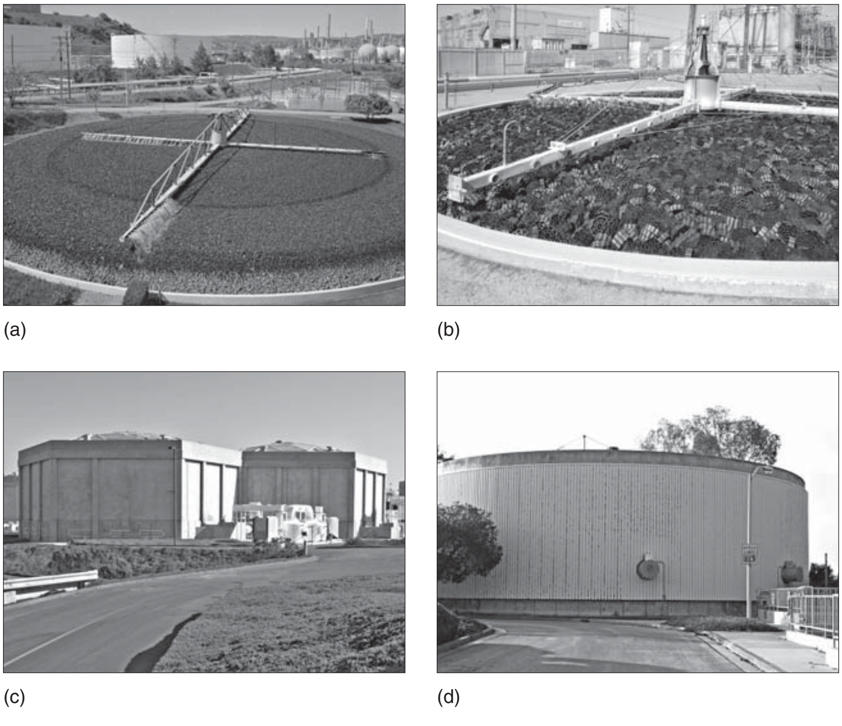

(a) (b) (c) (d)

Figure 9–3

Typical examples of trickling filters: (a) conventional shallow-depth, rock trickling filter without containment structure (see also Fig. 7–15 in Chap. 7), (b) conventional shallow-depth filter in which the original rock media (see Table 9–2) has been replaced with random pack plastic media (see Table 9–2), (c) multisided tower trickling filter employing plastic media (air pollution control equipment shown in foreground), and (d) tower trickling filter with fans for forced air aeration located around periphery of the filter.

BOD Removal with Forced Draft Aeration. Significantly higher organic loading rates (see Table 9–1) can be achieved with forced draft aeration. Although either rock or plastic packing can be used, the current trend is to use a plastic packing material in tall circular towers [see Figs. 9–3(c) and (d)]. The use of plastic packing allows for the construction of deeper filters with less land area required. Recirculation of the trickling filter effluent permits higher organic loadings, provides higher dosing rates on the filter to improve the liquid distribution and better control of the biofilm thickness, provides more oxygen in the influent wastewater flow, and returns viable organisms [see Figs. 9–4(a) and (b)]. It should be noted that other recirculation patterns have been used (Tchobanoglous et al., 2003).

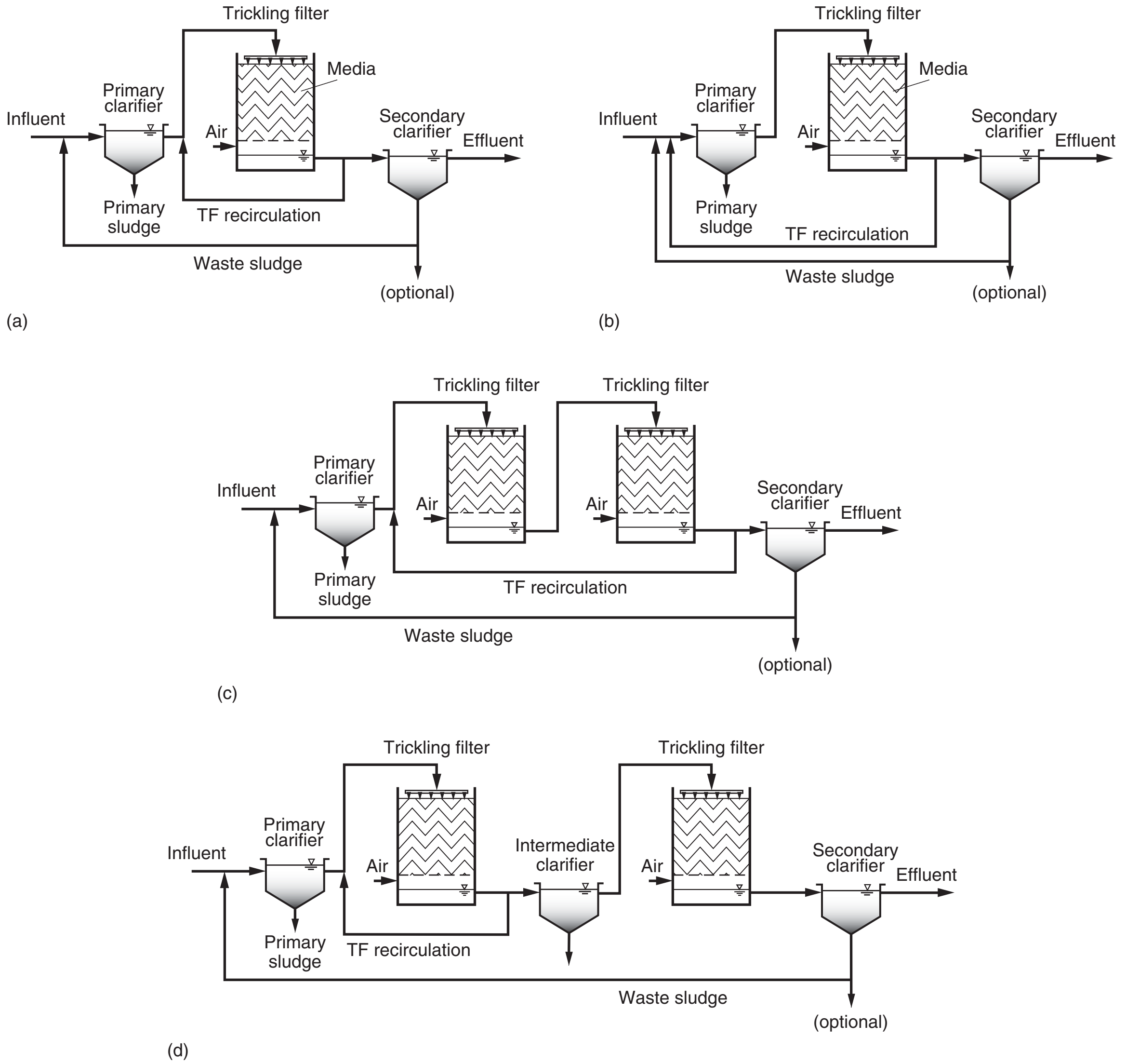


Figure 9–4

Typical trickling filter process schemes: (a) single trickling filter with waste sludge to the primary clarifier, (b) single trickling filter with recirculation and waste sludge to the primary clarifier, (c) two-stage trickling filter with waste sludge to the primary clarifier, and (d) two-stage trickling filter with intermediate clarifier. Note that many other flow schemes have been used (Tchobanoglous et al., 2003).

Recirculation also helps prevent ponding in the filter and to reduce the nuisance from odors and flies.

BOD Removal and Nitrification with Forced Draft Aeration. Both BOD removal and nitrification can be accomplished in rock or plastic packing trickling filters operated at reduced organic loading rates (Stenquist et al., 1974, Parker and Richards, 1986). As the BOD loading rate is decreased to what has been historically referred to as a *low-rate* trickling filter, as discussed above, nitrification also occurs in the same trickling filter or in the second stage of a system with two trickling filters in series [see Figs. 9–4(c) and (d)]. Plastic or rock media may be used in these trickling filters.

The hydraulic loading decreases with lower organic loadings. At very low hydraulic loading rates problems can occur due to improper media wetting and filter fly growth, unless the design adequately addresses filter effluent recirculation flows. Recirculation ratios (Q_R/Q) used range from 0.5 to 4.0. Vertical-flow media require an average total hydraulic loading exceeding 1.8 $m^3/m^2 \cdot h$ to maximize wetting and the BOD removal efficiency, and total hydraulic loadings in the range 0.4 to 1.1 $m^3/m^2 \cdot h$ have been used in shallow towers with cross-flow media.

Partial BOD Removal with Forced Draft Aeration. At organic loading rates typically greater than 1.6 kg $BOD/m^3 \cdot d$, trickling filters have been used to achieve partial BOD removal in the range of 50 to 70 percent. These highly loaded filters have been identified in the literature as *roughing* (or *high-rate*) filters. Most roughing filters are designed using plastic media (WEF, 2011). A roughing filter can be an attractive alternative for the pretreatment of modest to high strength industrial wastewaters. One of the advantages of roughing filters is the low energy requirement for BOD removal of higher strength wastewaters as compared to activated sludge aeration. Because the energy required is only for pumping the influent wastewater and recirculation flows, the amount of BOD removal per unit of energy input can increase as the wastewater strength increases until more recirculation is needed. The energy requirement for a roughing application may range from 2 to 5 kg BOD applied/kWh versus 1.2 to 2.4 kg BOD/kWh for activated sludge treatment.

Advantages and Disadvantages of Trickling Filters

Trickling filters, as noted above, have been used as aerobic attached growth processes for BOD removal only, combined BOD removal and nitrification, and for tertiary nitrification after secondary treatment by suspended growth or attached growth processes. The advantages and disadvantages of the use of trickling filters are as follows.

Advantages. The principal advantages claimed for trickling filters over the activated sludge process are as follows:

1. Less energy required
2. Simpler operation with no issues of mixed liquor inventory control and sludge wasting
3. No problems of bulking sludge in secondary clarifiers
4. Better sludge thickening properties
5. Less equipment maintenance needs
6. Better recovery from shock toxic loads

Disadvantages. Many disadvantages often cited for trickling filters, such as a poorer effluent quality in terms of BOD and TSS concentrations, greater sensitivity to lower temperatures, odor production, and uncontrolled solids sloughing events, are related more to the specific process and final clarifier designs than to the actual process capabilities

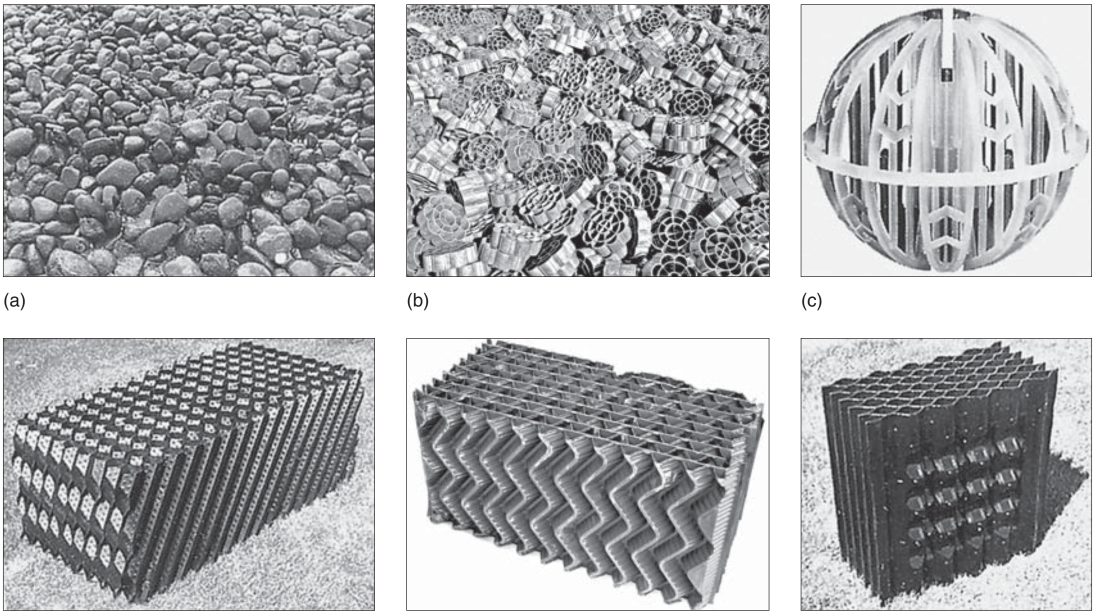
(WEF, 2011). In general, the actual limitations of the processes are the difficulty in accomplishing biological nitrogen and phosphorus removal compared to single-sludge biological nutrient removal suspended growth designs, and the production of an effluent with a higher suspended solids concentration than activated sludge treatment.

Physical Facilities for Trickling Filters

Factors that must be considered in the design of trickling filters include (1) type and physical characteristics of filter packing to be used, (2) type and dosing characteristics of the distribution system, and (3) configuration of the underdrain system. The dosing rate, the provision of adequate airflow (i.e., ventilation) by either natural or forced air and settling tank design are considered under design considerations.

Filter Packing Material. Typical trickling filter packing materials, including rock, random pack plastic media, and cross- and vertical-flow plastic bundles, are shown on Fig. 9–5. Redwood slates have also been used in the past, but are seldom used today. The physical characteristics of commonly used trickling filter packings, including those shown on Fig. 9–5, are reported in Table 9–2. Since the 1960s, plastic packing material, either cross-flow or vertical-flow, has become the packing of choice in the United States.

Rock. Where locally available and cost effective, rock media may be used. The most suitable material is rounded river rock or crushed stone, graded to a uniform size so that 95 percent is within the range of 75 to 100 mm (3 to 4 in.). The specification of size uniformity is a way of ensuring adequate pore space for wastewater flow and air circulation.


(a) (b) (c)

(d) (e) (f)

Figure 9–5

Typical packing material for trickling filters: (a) rock, (b) and (c) random pack plastic media, (d) and (e) plastic cross-flow bundle, and (f) plastic vertical-flow bundle. Media characteristics are summarized in Table 9–2.

Table 9–2

Physical properties of trickling-filter packing materials[a]

Packing material	Nominal size, cm (in.)	Approx bulk weight, kg/m^3 (lb/ft^3)	Approx specific surface area, m^2/m^3 (ft^2/ft^3)	Void space, %	Application[b]
River rock [see Fig. 9–5(a)]	7.5–10 (3–4)	1000–1300 (62–90)	50 (15)	55	C, CN, N
Plastic random packing [see Fig. 9–5(b)]	18.5 dia × 5.1 (7.3 dia × 2)	27 (1.7)	98 (30)	95	C, CN, N
Plastic random packing [see Fig. 9–5(c)]	9 (3.5 dia)	53 (3.0)	125 (38)	95	N
Plastic cross-flow [see Figs. 9–5(d) and (e)]	61 × 61 × 122 (24.0 × 24.0 × 48)	25–45 (1.6–2.8)	100, 138, 223 (30, 42, 68)	>95	C, CN, N
Plastic vertical flow [see Fig. 9–5(f)]	61 × 61 × 122 (24.0 × 24.0 × 48)	25–45 (1.6–2.8)	102, 131 (31, 40)	>94	C, CN, N

[a] Adapted from Tchobanoglous et al. (2003) and WEF (2011).

[b] C = BOD removal; N = tertiary nitrification; CN = combined BOD and nitrification.

Note: kg/m^3 × 0.06246 = lb/ft^3
m^2/m^3 × 0.305 = ft^2/ft^3

Plastic Packing. Plastic packing materials come in two forms: random pack and molded plastic bundles. A number of different random pack materials are available. Two different examples are illustrated on Figs. 9–5(b) and (c). The molded plastic packing bundles have the appearance of a honeycomb. Flat and corrugated sheets of polyvinyl-chloride are bonded together in rectangular modules. The sheets usually have a corrugated surface for enhancing slime growth and retention time. Each layer of modules is turned at right angles to the previous layer to further improve wastewater distribution. The two basic types of corrugated plastic sheet packing are cross- and vertical-flow [see Figs. 9–5(d), (e), and (f)].

Both types of plastic sheet packings are reported to be effective in BOD and TSS removal over a wide range of loadings (Harrison and Daigger, 1987; Aryan and Johnson, 1987). Biotowers as deep as 12 m (40 ft) have been constructed using plastic packing, with depths in the range of 6 m (20 ft) being more common. In biotowers with vertical plastic packing, cross-flow packing can be used for the upper-most layers to enhance the distribution across the top of the filter. The high hydraulic capacity, high void ratio, and resistance to plugging offered by these types of packing can best be used in a high-rate type filter.

Plastic packing has the advantage of requiring less land area for filter structure than rock due to the ability to use taller trickling filters. Grady et al. (1999) noted that when loaded at similar low organic loadings rates (less than 1.0 kg BOD/m^3·d), the performance of rock filters and plastic packing filters was similar. At higher organic loading rates, however, the performance of filters with plastic packing is superior. The higher porosity, which provides for better air circulation and biofilm sloughing, is a likely explanation for the improved performance.

Strength and Durability of Packing Materials. Other important characteristics of filter packing materials are strength and durability. Durability may be determined by the sodium sulfate test, which is used to test the soundness of concrete aggregates (U.S. EPA, 1974).

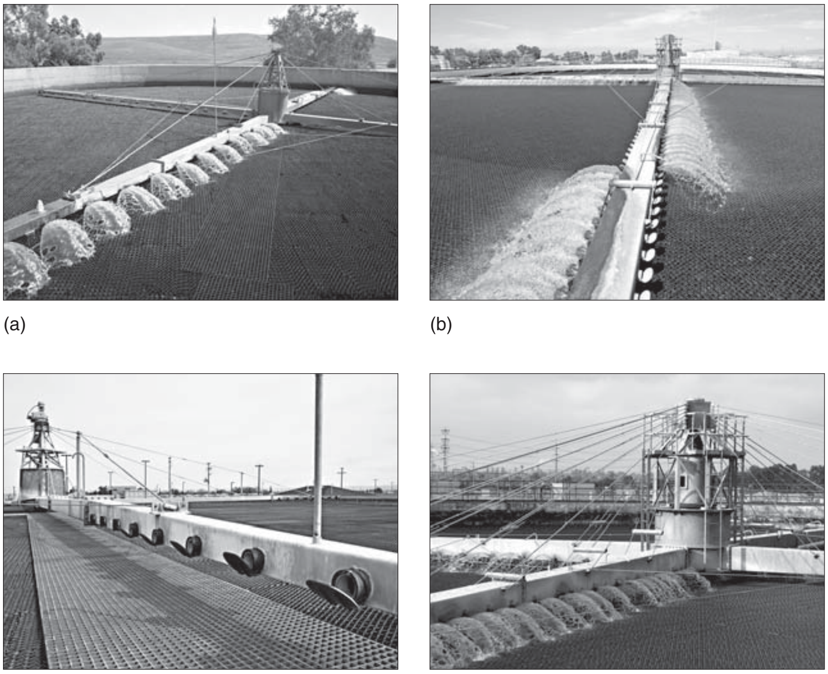
Because of the weight of the packing, the depth of rock filters is usually on the order of 2 m (6 ft). The low void volume of rock limits the space available for airflow and increases the potential for plugging and flow short-circuiting. Because of plugging, the organic loadings to rock filters are more commonly in the range of 0.3 to 1.0 kg BOD/m^3·d.

Distribution Systems. Distributors are manufactured for trickling filters with diameters up to 60 m (200 ft). Distributor arms may be of constant cross section for small units, or they may be tapered to maintain minimum transport velocity.

Flow Distribution Nozzles. Nozzles are spaced unevenly so that greater flow per unit of length is achieved with increasing length (and area) from the center. For uniform distribution over the area of the filter, the flowrate per unit of length should be proportional to the radius from the center [see Fig. 9–6(a)]. Headloss through the distributor is in the range of 0.6 to 1.5 m (2 to 5 ft). Forward-facing flow nozzles are also used to control the rotational speed of the rotary distributor [see Fig. 9–6(b) and (c)].

Fixed Nozzle Distribution Systems. Fixed-nozzle distribution systems consist of a series of spray nozzles located at the points of equilateral triangles covering the filter bed. A system of pipes placed in the filter is used to distribute the wastewater uniformly to the nozzles. Special nozzles having a flat spray pattern are used, and the pressure is varied systematically so that the spray falls first at a maximum distance from the nozzle and then at a decreasing distance as the head slowly drops. In this way, a uniform dose is applied over the whole area of the bed. Half-spray nozzles are used along the sides of the filter.

Distributor Drive Systems. Drive systems for trickling filter rotary distributors are of two types: (1) hydrostatic pressure and (2) a combination of hydrostatic pressure and electric



(a) (b) (c) (d)

Figure 9–6

Typical distributors used to supply wastewater to trickling filter packing: (a) view of two-arm tapered rectangular shaped rotary distributor for tower trickling filter, (b) view of top of tower trickling filter with four-arm rotary distributor (water jets on the left hand side are used to slow down the rotational speed of the distributor), (c) view of splash plates used to distribute water over media, and (d) view of variable speed electric motor drive for distributor arms.

motor drives. Historically, trickling filter distributor arms were driven with water pressure. One of the problems with the early pressure driven units was that a mercury seal was used around the distributor center column. Because mercury was released from seals that were occasionally blown out by excessive hydrostatic pressure, mechanical seals are now used on hydrostatic pressure driven distributor arms. Electric motor drives, both fixed and variable speed provide operational flexibility, and are almost always used with tower tricking filters with plastic packing [see Fig. 9–6(d)].

Other Important Features. Important features that should be considered in selecting a distributor are the ruggedness of construction, ease of cleaning, ability to handle large variations in flowrate while maintaining adequate rotational speed, and corrosion resistance of the material and its coating system. Drain caps should be provided at the end of the distribution arms for flushing, as needed. Clearance of 150 to 225 mm (6 to 9 in.) should be allowed between the bottom of the distributor arm and the top of the bed. This clearance permits the wastewater streams from the nozzles to spread out and cover the bed uniformly, and it prevents ice accumulations from interfering with the distributor motion during freezing weather.

Underdrains. The wastewater collection system in a trickling filter consists of underdrains that catch the treated wastewater and solids discharged from the filter packing for conveyance to the final sedimentation tank. The floor and underdrain slope to a central or peripheral drainage channel at a 1 to 5 percent grade. The effluent channels are sized to produce a minimum velocity of 0.6 m/s (2 ft/s) at the average daily flowrate (WPCF, 1988).

Underdrains for Rock Trickling Filter. Underdrains may be open at both ends, so that they may be inspected easily and flushed out if they become plugged. All underdrain channels should be designed to flow half full at the maximum anticipated hydraulic application rate. The underdrains also ventilate the filter, providing the air for the microorganisms that live in the filter slime, and they should at least be open to a circumferential channel for ventilation at the wall as well as to the central collection channel. All underdrain systems should be designed so that forced air ventilation can be added at a later date if filter operating conditions should change. The floor and underdrains must have sufficient strength to support the packing, slime growth, and the wastewater. The underdrain system for a rock filter usually has precast blocks of vitrified clay or fiberglass grating laid on a reinforced concrete subfloor [see Figs. 9–7(a), (b), and (c)]. The specific type of underdrain used will vary, depending on the location in the United States and from country to country.

Underdrains for Plastic Packing. The underdrain and support system for plastic packing consists of either a beam and column or grating. A typical underdrain system for a biotower filter is shown on Fig. 9–7(d). The beam and column system typically has precast concrete beams supported by columns or posts [see Fig. 9–7(e)]. The plastic packing is placed over the beams, which have channels in their tops to ensure free flow of wastewater and air. Air ventilation ports on a tower tricking filter are located around the periphery of the filter.

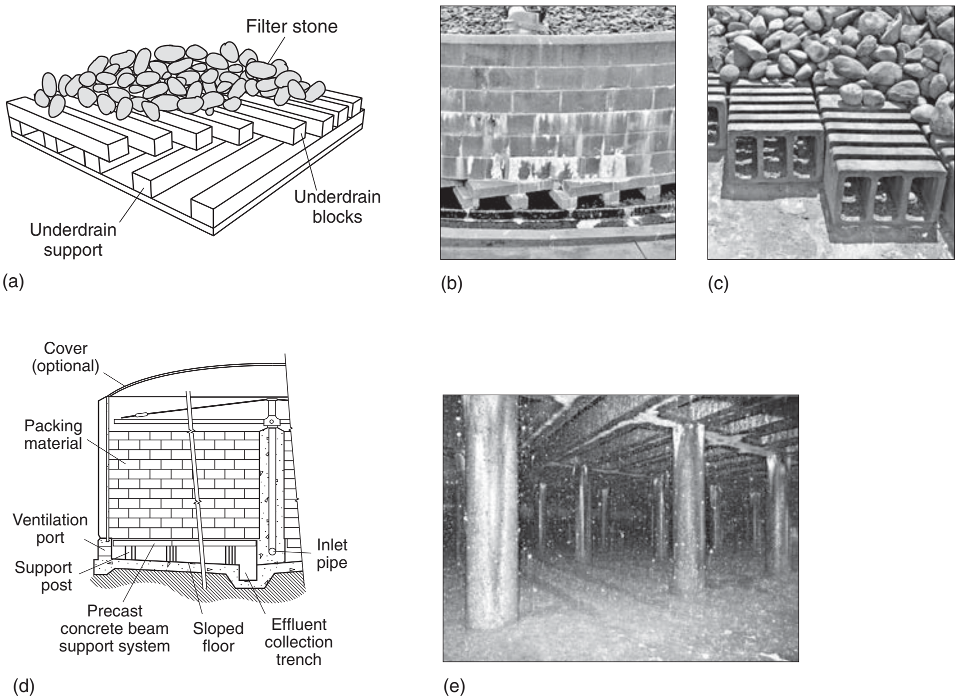
Design Considerations for Physical Facilities

Major design issues with trickling filters involve (1) providing an adequate dosing strategy and hydraulic load, (2) providing adequate airflow to meet oxygen requirements, (3) oxygen transfer, (4) pressure drop, (5) ponding, (6) odor control, and (7) control of predators. Each of these issues is considered below.

Hydraulic Application Rate. An adequate and uniform hydraulic application rate must be maintained to provide full wetting of the packing media, to promote efficient

Figure 9–7

Typical trickling filter underdrains: (a) schematic of underdrain for rock tricking filter, (b) view of concrete beam underdrain system for rock filter, (c) view of vitrified clay block underdrain system for rock trickling filter, (d) section through tower trickling filter (note ventilation port, media support sloped bottom, and effluent collection trench), and (e) view of underdrain plenum showing media support columns. [The crawl space between the media support and the floor is 1.25 m (4 ft) to allow for underdrain maintenance.]



treatment, to control the biofilm thickness, to minimize odors, and to minimize the nuisance *Psychoda* and *Anisopus* fly population. Sustained low hydraulic application rates and low dosing rates can result in excessive biofilm thickness that can lead to over weighting and collapse of plastic media packing and to excessive worm growth. The dosing rate is the depth of liquid discharged on top of the packing for each pass of the distributor. For higher distributor rotational speeds, the dosing rate is lower.

In the past, typical rotational speeds for distributors were about 0.5 to 2 min per revolution (WEF, 2000). With two to four arms, the trickling filter is dosed every 10 to 60 s. Based on the results from various investigators, it has been shown that reducing the distributor speed results in better filter performance. Hawkes (1963) found that rock trickling filters dosed every 30 to 55 min/rev outperformed a more conventional operation of 1 to 5 min/rev. Besides improved BOD removal, there were dramatic reductions in the *Psychoda* and *Anisopus* fly population, biofilm thickness, and odors. Similar advantages were found by Albertson and Davies (1984) from an investigation of reduced distributor speed. At a higher dosing rate, the larger water volume applied per revolution: (1) provides greater wetting efficiency, (2) results in greater agitation, which causes more solids to flush out of the packing, (3) results in a thinner biofilm, and (4) helps to wash away fly eggs. The thinner biofilm creates more surface area and results in a more aerobic biofilm.

Control of Biofilm Thickness. If the high dosing rate preferred to control the biofilm thickness is sustained the treatment efficiency may be decreased because the liquid contact time in the filter would be less. For low-loaded trickling filters, a daily intermittent high dose, referred to as a *flushing dose,* is used to control the biofilm thickness and solids inventory. A combination of a once per day high flushing rate and a lower daily sustained dosing rate is recommended as a function of the BOD loading as shown in Table 9–3 (WEF, 2011). The data in Table 9–3 can be used as guidance in establishing an appropriate

Table 9–3

A guideline for the trickling filter dosing rate as a function of BOD loading[a]

BOD loading, $kg/m^3 \cdot d$	Operating dose, mm/pass[b]	Flushing dose, mm/pass[b]
<0.4	25–75	100
0.8	50–150	150
1.2	75–225	220
1.6	100–300	300
2.4	150–450	450
3.2	200–600	600

[a] Adapted from WEF (2011).

[b] mm/pass represents the amount of liquid applied for each pass of each distributor arm.

dosing range. Optimization of the dosing rate and flushing rate and frequency is best determined from field operation. Flexibility in the distributor design is needed to provide a range of dosing rates to optimize the trickling filter performance.

Dosing Rate as a Function of the Hydraulic Loading Rate. The dosing rate is a function of the total hydraulic loading (influent flow plus recycle flow), the number of arms on the flow distributor, and the rotational speed of the distributor (WEF, 2011).

$$DR = \frac{(1 + R)(q)(10^3 \text{ mm/1 m})}{(N_A)(n)(60 \text{ min/h})} \tag{9–1}$$

where DR = dosing rate, mm/pass of distributor arm
n = rotational speed, rev/min
q = influent applied hydraulic loading rate, $m^3/m^2 \cdot h$
R = recirculation ratio
N_A = number of arms in rotary distributor assembly

The dosing rate has also been referred to as the SK value, which stands for *Spulkraft,* a term used in the German regulations to define dosing in the early 1980s. To achieve the suggested dosing rates, the speed of the rotary distributor can be controlled by: (1) reversing the location of some of the existing orifices to the front of the distributor arm [see Figs. 9–6(b) and (c)], (2) adding reversed deflectors to the existing orifice discharges, or (3) converting the rotary distributor to a variable-speed electric drive [see Fig. 9–6(d)] (Albertson, 1995). An advantage of the variable-speed electric drive is the wide range of operational flexibility it provides and the ease of controlling the dosing rate without having to change the distributor arm orifice discharge design.

Airflow. An adequate flow of air is of fundamental importance to maintaining aerobic conditions within the trickling filter biofilm to provide efficient treatment and to prevent odors. Natural draft has historically been the primary means of providing air flow for rock media filters, but it is not always adequate. Forced ventilation using low-head fans provides more reliable and controlled airflow.

Natural Draft. In the case of natural draft, the driving force for airflow is the temperature difference between the ambient air and the air inside the pores. If the wastewater is colder than the ambient air, the pore air will be cold and the direction of flow will be downward. If the ambient air is colder than the wastewater, the flow will be upward. The latter is less desirable from a mass transfer point of view because the partial pressure of oxygen (and thus the oxygen transfer rate) is lowest in the region of highest oxygen demand. In many areas in the United States, there are periods, especially during the summer, when essentially no airflow occurs through the trickling filter because temperature differentials are negligible.

Draft, which is the pressure head resulting from the temperature and moisture differences, may be determined from Eq. (9–2) (Schroeder and Tchobanoglous, 1976):

$$D_{\text{air}} = 353\left(\frac{1}{T_c} - \frac{1}{T_h}\right)Z \tag{9–2}$$

where D_{air} = natural air draft, mm of water
T_c = cold temperature, K
T_h = hot temperature, K
Z = height of the filter, m

A more conservative estimate of the average pore air temperature is obtained by using the log mean temperature, T_m for T_h in Eq. (9–2).

$$T_m = \frac{T_2 - T_1}{\ln (T_2/T_1)} \tag{9–3}$$

where T_2 = colder temperature, K
T_1 = warmer temperature, K

The volumetric air flowrate may be estimated by setting the draft equal to the sum of the headlosses that result from the passage of air through the filter and underdrain system (Albertson and Okey, 1988).

Where natural draft is used the following needs to be included in the design:

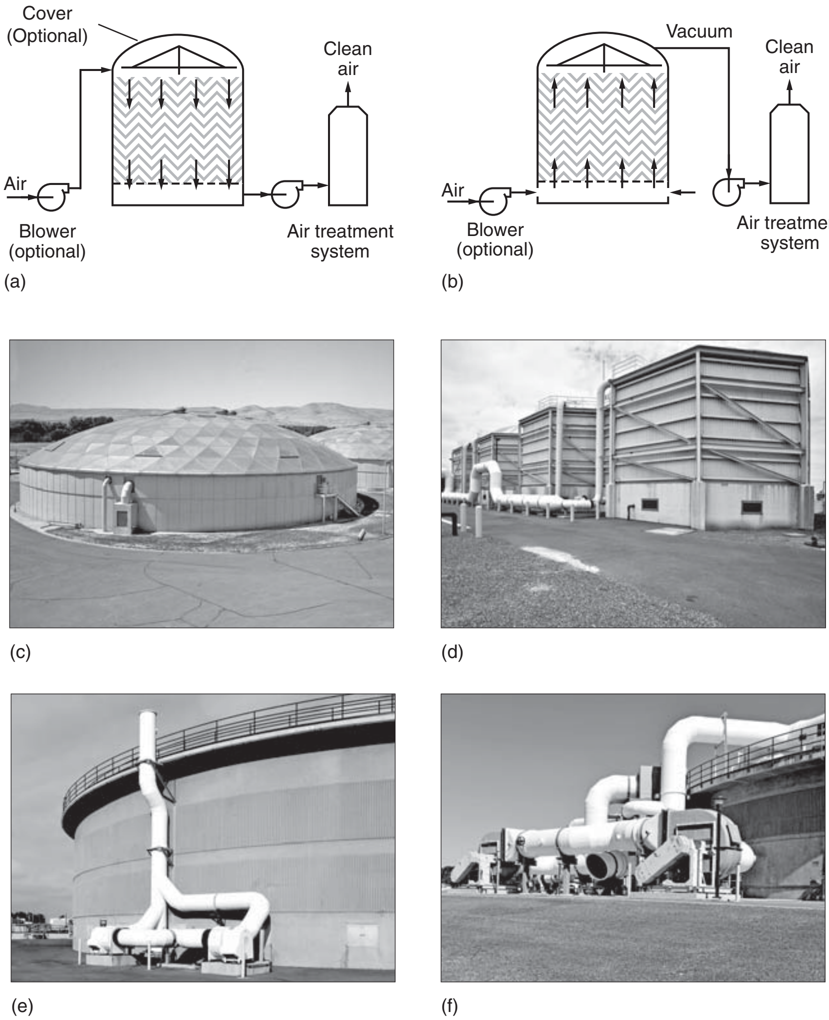
1. Underdrains and collecting channels should be designed to flow no more than half full to provide a passageway for the air.
2. Ventilating access ports with open grating types of covers should be installed at both ends of the central collection channel.
3. Large diameter filters should have branch collecting channels with ventilating manholes or vent stacks installed at the filter periphery.
4. The open area of the slots in the top of the underdrain blocks should be at least 15 percent of the area of the filter.
5. One square meter gross area of open grating in ventilating access ports and vent stacks should be provided for each 23 m^2 (250 ft^2) of filter area.

Forced Air. The use of forced or induced draft fans are recommended for trickling filter designs to provide a reliable supply of oxygen. As an approximation, an air flow of 0.3 $m^3/m^2 \cdot min$ (1 $ft^3/ft^2 \cdot min$) of filter area in either direction is recommended. The costs for a forced draft air supply are minimal compared to the benefits. For a 3800 m^3/d (1.0 Mgal/d) wastewater treatment flow the estimated power requirement is only about 0.15 kW (0.2 hp) (WEF, 2000). Both upflow and downflow forced-air aeration systems are used (see Fig. 9–8). A downflow direction is preferred as it provides contact time for treating odorous compounds released at the top of the filter and a richer air supply at the top where the oxygen demand is highest [see Figs. 9–8(a) and (c)]. If an upflow air direction is used, the trickling filter must be covered to collect odorous off gas [see Figs. 9–8(b) and (d)]. Covers also help to maintain water temperatures in the trickling filter. For applications with extremely low air temperature, it may be necessary to restrict the flow of air through the filter to keep it from freezing.

Forced air designs should provide multiple air distribution points by the use of fans around the periphery of the tower [see Figs. 9–3(d) and 9–8(e)] or the use of air openings

Figure 9–8

Examples of forced air aeration systems: (a) schematic of downflow forced-air aeration system, (b) schematic of upflow forced-air aeration system, (c) view of fan used to blow air into top of covered rock trickling filter, (d) view of covered tower trickling filter with air pulled up through filter by a vacuum withdrawal system leading to foul air treatment system, (e) view of large fans used to pull air through the tower trickling filter underdrain plenum and discharge it to the atmosphere, and (f) view of covered tower trickling filter with large fans used to blow air into the top of the filter and to withdraw air from the underdrain plenum for treatment.



below the packing material with suction from the top of the filter [see Fig. 9–8(d)]. The opening in the airflow distribution piping should be selected to provide equal airflow distribution. A pipe air flow velocity in the range of 1100 to 2200 m/h is typically used to further promote uniform airflow distribution (WEF, 2011).

Oxygen Transfer in Trickling Filters. Little has been done to quantify the amount of oxygen used in trickling filters and the actual oxygen transfer efficiency. The following formulations are based on earlier work by Dow Chemical during the development of plastic packing material for trickling filter applications. In developing these formulations for BOD removal applications, it was assumed that the oxygen transfer efficiency was about 5 percent. A more conservative transfer efficiency of 2.5 percent may be assumed as a higher dissolved oxygen concentration is desired in the gas-liquid interface to maximize nitrification efficiency (WEF, 2011). The required oxygen supply is given as follows:

BOD removal only:

$$R_o = (20\ \text{kg/kg})\,[0.80\,e^{-9L_B} + 1.2\,e^{-0.17L_B}](\text{PF}) \tag{9–4}$$

BOD removal and nitrification:

$$R_o = (40\,\text{kg/kg})[0.80e^{-9L_B} + 1.2e^{-0.17L_B} + 4.6N_{\text{ox}}/\text{BOD}](\text{PF}) \tag{9–5}$$

where R_o = oxygen supply, kg O_2/kg BOD applied
L_B = BOD loading to filter, kg BOD/m^3·d
N_{ox}/BOD = ratio of influent nitrogen oxidized to influent BOD, mg/mg
PF = Peaking factor, maximum to average load

The air application rate at 20°C and 1.0 atm is computed as follows. From Appendix B, the density of air at 20°C and 1.0 atm is 1.204 kg/m^3, and the percent of oxygen by weight in air is 23.18 percent. Thus, the volume of oxygen per kg of air is 3.58 m^3/kg {1/[(1.204 kg/m^3)(0.2318)]} and the required air flow is given by:

$$AR_{20} = \frac{(R_o)(Q)(S_o)(3.58\ \text{m}^3/\text{kg O}_2)}{(10^3\ \text{g}/1\ \text{kg})(1440\ \text{min/d})} \tag{9–6}$$

where AR$_{20}$ = air flowrate at 20°C and 1.0 atm, m^3/min
Q = wastewater flowrate, m^3/d
S_o = primary effluent BOD, g/m^3

The air flowrate is corrected for temperature and pressure according to the ideal gas law:

$$AR_T = AR_{20}\left(\frac{273.15 + T_A}{273.15}\right)\left(\frac{P_a}{P_b}\right) \tag{9–7}$$

where AR_T = air flowrate at ambient air temperature, °C
T_A = ambient air temperature, °C
P_a = atmospheric pressure, 1.0 atm (101.325 kPa)
P_b = atmospheric pressure at treatment plant site, atm (kPa)

A further correction to the calculated air flow is recommended for temperatures above 20°C to account for the lower oxygen saturation concentration at higher temperatures and the higher biological uptake rates in the filter. For each degree centigrade above 20°C the air flowrate is increase by 1 percent.

$$AR_{T>20°\text{C}} = AR_T\left(1 + \frac{T_A - 20}{100}\right) \tag{9–8}$$

Pressure Drop in Trickling Filters. The pressure drop through the packing is related to the superficial air velocity as follows:

$$\Delta P = N_p\left(\frac{v^2}{2g}\right) \tag{9–9}$$

where ΔP = total headlosses, kPa
g = acceleration of gravity, 9.81 m/s^2
v = superficial velocity, Q/A, m/s
N_P = tower resistance; number of velocity heads

The tower resistance term, N_p, is the sum of all the individual headlosses related to the air flow. Headloss occurs as air moves through the inlet, underdrain, and packing material. The packing loss in terms of number of velocity headlosses was developed by Dow Chemical for the original vertical packing:

$$N_p = 10.33(D)e^{(1.36 \times 10^{-5})(L/A)} \tag{9–10}$$

where N_p = packing headloss in terms of velocity heads
D = packing depth, m
L = liquid loading rate, kg/h
A = tower cross-section area, m^2

Table 9–4

Correction factors for computing headloss in nonvertical trickling filter packings based on Eq. (9–10)[a]

Packing	Specific surface, area, m^2/m^3	Correction factor
Rock	45	2.0
Plastic cross-flow	100	1.3
Plastic cross-flow	140	1.6
Plastic random	100	1.6

[a] Adapted from WEF (2010).

Although similar correlations have not been developed for other packing materials, recommended correction factors that can be used to obtain N_p values for other packing materials based on the value determined using Eq (9–10) are given in Table 9–4. To estimate the total headloss for the trickling filter, the value of N_p, computed using Eq. (9–10), is often multiplied by a factor of 1.3 to 1.5 to include inlet, underdrain, and other minor losses. The determination of airflow and pressure drop in trickling filters with forced aeration ventilation is illustrated in Example 9–1.

EXAMPLE 9–1 **Determine Trickling Filter Airflow Rate Requirements and Pressure Drop for Forced Ventilation** Determine the forced-ventilation air requirement and pressure drop in a trickling filter with cross-flow plastic packing designed for carbonaceous BOD removal, given the following design and operating information.

Wastewater characteristics:

1. Wastewater flowrate = 15,000 m^3/d (174 L/s)
2. Primary effluent BOD = 140 g/m^3
3. Warm month wastewater temperature = 20°C

Design assumptions:

1. BOD loading = 0.60 kg BOD/m^3·d (from Table 9–1)
2. Organic loading peaking factor = 1.4
3. Tower diameter = 20 m
4. Number of towers = 2
5. Depth of packing = 6.1 m
6. Headloss correction factor for inlet and other minor losses = 1.5
7. Headloss correction factor for cross-flow packing = 1.3 (see Table 9–4)
8. Air temperature = −7 to 28°C
9. Atmospheric pressure at treatment plant site = 1 atm

Solution

1. Determine the required oxygen supply rate using Eq. (9–4).

$$R_o = (20\text{ kg/kg})[0.80e^{-9L_B} + 1.2e^{-0.17L_B}](\text{PF})$$

$$L_B = 0.60\text{ kg BOD/m}^3\cdot\text{d}$$

$$\text{PF} = 1.4$$

$$R_o = 20[0.80e^{-9(0.60)} + 1.2e^{-0.17(0.60)}](1.4) = 41.6\text{ kg O}_2\text{/kg BOD applied}$$

2. Determine airflow rate for the given conditions.
   a. Using Eq. (9–6), determine the airflow rate at standard conditions

$$AR_{\mathrm{STD}} = \frac{R_o(Q)(S_o)(3.58\ \mathrm{m^3/kg\ O_2})}{(10^3\ \mathrm{g/1\ kg})(1440\ \mathrm{min/d})}$$

$$= \frac{(41.6\ \mathrm{kg/kg})(15{,}000\ \mathrm{m^3/d})(140\ \mathrm{g/m^3})(3.58\ \mathrm{m^3/kg\ O_2})}{(10^3\ \mathrm{g/1\ kg})(1440\ \mathrm{min/d})}$$

$$= 217\ \mathrm{m^3/min}$$

   b. Correct the air flowrate for temperature and pressure using Eq. (9–7).

$$AR_{TA} > AR_{\mathrm{STD}}\left(\frac{273.15 + T_A}{273.15}\right)\left(\frac{1\ \mathrm{atm}}{1\ \mathrm{atm}}\right)$$

$$T_A = 28°\mathrm{C}$$

$$AR_{28} = (217\ \mathrm{m^3/min})\left(\frac{273.15 + 28}{273.15}\right) = 239.2\ \mathrm{m^3/min}$$

   c. Correct the air flowrate for lower oxygen saturation using Eq. (9–8).

$$AR = AR_{TA}\left(1 + \frac{T_A - 20}{100}\right)$$

$$T_A = 28°\mathrm{C}$$

$$AR = (239.2)\left(1 + \frac{28 - 20}{100}\right) = 258.3\ \mathrm{m^3/min}$$

3. Determine pressure drop in the cross-flow packing.
   a. Determine the value of N_p using Eq. (9–10). The value of N_p will then be used to determine the pressure drop using Eq. (9–9).

$$N_p = 10.33(D)e^{(1.36\times 10^{-5})(L/A)}$$

Solve for L/A, kg/m²·h

Hydraulic loading $= q = Q/A$

$$Q = (15{,}000\ \mathrm{m^3/d})(1\ \mathrm{d/24\ h}) = 625\ \mathrm{m^3/h}$$

$$\text{Area of single tower } A = \frac{\pi D^2}{4} = \frac{3.14(20.0\ \mathrm{m})^2}{4} = 314\ \mathrm{m^2}$$

$$q = \frac{625}{314} = 1.99\ \mathrm{m^3/m^2{\cdot}h}$$

$$\frac{L}{A} = (1.99\ \mathrm{m^3/m^2{\cdot}h})(10^3\ \mathrm{L/1\ m^3})(1\ \mathrm{kg/L})$$

$$= 1990\ \mathrm{kg/m^2{\cdot}h}$$

Packing depth = 6.1 m

$$N_p = 10.33(6.1)e^{(1.36\times 10^{-5})(1990)}$$

$$= 64.7$$

Headloss correction factor for cross-flow packing = 1.3 (Table 9–4)

Headloss correction factor for inlet and other losses = 1.5

$N_P = (1.5)(1.3)(64.7) = 126.2$

b. Solve for the pressure drop using Eq. (9–9).

$$\Delta P = N_p\left(\frac{v^2}{2g}\right)$$

$$\text{Superficial velocity } v = \frac{\text{(air flowrate/tower)}}{\text{area of tower}}$$

$$v = \frac{(258.3/2)}{314.0} = 0.41\text{ m/min} = 0.0069\text{ m/s}$$

$N_P = 126.2$

$$\Delta P = 126.2\left(\frac{v^2}{2g}\right)$$

$$= \frac{(126.2)(0.0069\text{ m/s})^2}{2(9.8\text{ m/s}^2)} = 0.0003\text{ m}$$

Air density at 28°C = 1.175 kg/m^3 (see Appendix B)

$$\Delta P = 0.0003\text{ m }(1.175\text{ kg/m}^3)(9.8\text{ m/s}^2)$$

$$= 0.00346\text{ N/m}^2 = 0.00346\text{ Pa} = 3.46 \times 10^{-6}\text{ kPa}$$

4. Compare to natural draft pressure per Eqs. (9–2) and (9–3).

a. Determine log mean temperature using Eq. (9–3).

Wastewater temperature = 20°C

Air temperature = 28°C

$$T_m = \frac{T_2 - T_1}{\ln(T_2/T_1)} = \frac{28 - 20}{\ln(28/20)} = 23.8^\circ\text{C}$$

b. Determine draft using Eq. (9–2).

$$D_{\text{air}} = 353\left(\frac{1}{T_C} - \frac{1}{T_m}\right)Z$$

$T_C = 273.15 + 20 = 293.15\text{ K}$

$T_m = 273.15 + 23.8 = 296.95$

$$D_{\text{air}} = 353\left(\frac{1}{293.15} - \frac{1}{296.95}\right)6.1 = 0.094\text{ mm}$$

c. Compare draft to estimated headloss.
Convert mm of water to pressure expressed in Pa

$$\text{Draft} = (0.0094\text{ mm H}_2\text{O})\left(\frac{9.797\text{ Pa}}{\text{mm H}_2\text{O}}\right) = 0.0921\text{ Pa}$$

Thus, draft (0.0921 Pa) > headloss (0.00346 Pa)

Comment More draft is available for these temperature differences than is needed, but for periods where the wastewater and air temperatures are very close sufficient air flow and oxygen will not be available. Note that the pressure drop for the necessary airflow rate is very low and multiple air feed points are needed to assure uniform air distribution.

Ponding. On occasion, excessive growth and/or the presence of large suspended material in the influent wastewater can cause ponding to occur. Ponding occurs when a bridge is formed between the individual media particles or when the pore space between media particles is filled. Ponding is a more serious problem with rock filters, especially those with smaller rock sizes. Ponding can lead to poor air and water distribution within the filter which, in turn, can lead to reduced treatment performance, the breeding of mosquitoes, and the formation of odors.

Methods used to control ponding include (1) improved pretreatment (e.g. screening following primary sedimentation), (2) reducing the organic loading, (3) increasing the hydraulic loading to induce sloughing, (4) use of high pressure steam to breakup pond sites, (5) maintaining a chlorine residual of 1 to 2 mg/L to breakup the ponded sites, (6) shutting off the filter to allow the ponded sites to dry out so that they can be washed off, and (7) manually removing the material causing the ponding.

Odor Control. An extremely important consideration in the implementation of trickling filters is the control of odors, which can develop from time to time for a variety of reasons, in both natural and force draft filters, including the presence of odorous compounds and gases in the influent wastewater, excessive organic loading, unequal hydraulic distribution, inadequate air flow (draft through the filter), and ponding as described above.

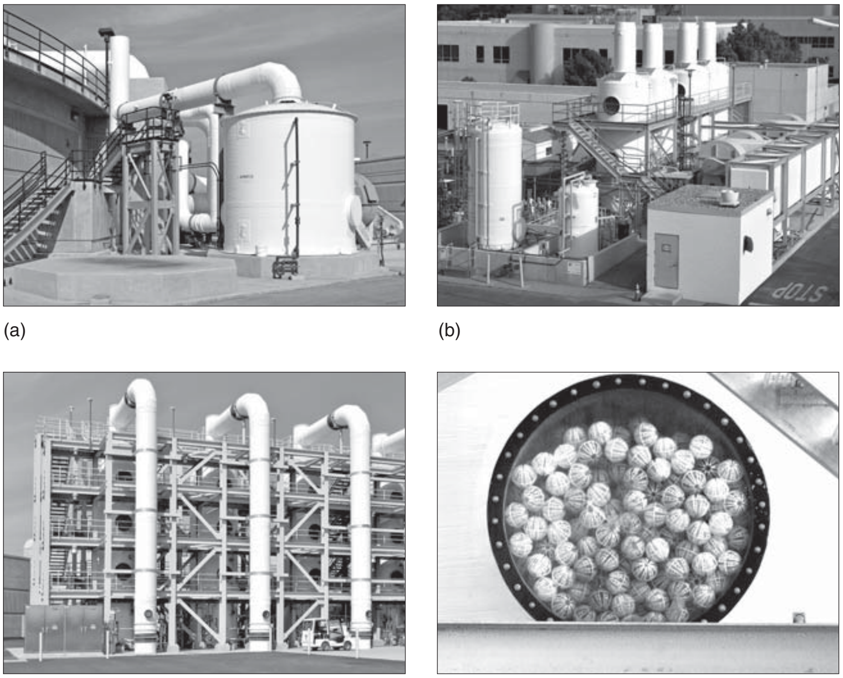
Corrective measures that have been used include (1) temporarily increasing the hydraulic loading to flush the filter, (2) reducing the organic loading while increasing the hydraulic loading rate by increasing the rate of recirculation (use of temporary pumps may be necessary), (3) increasing the draft though the filter (if that option is available), (4) adding chemical oxidants, (5) switching from rock to plastic media, and (6) covering the filter and treating the odors through an odor scrubber. Although listed last, the covering of trickling filters is now common, especially where they are located near residential areas [see Figs. 9–3(c) and (d) and 9–8(c) and (d)]. Off gases from tower trickling filters are usually treated in either carbon absorbers, chemical scrubbers or biotrickling filters (see Fig. 9–9). Additional details on off-gas treatment are presented in Chap. 16.

Predator Problems. A significant problem for nitrifying filters is the development of a snail population, which may graze on the biofilm to reduce the nitrifying bacteria population and nitrification performance [see Figs. 9–10(a) and (b)]. In addition, snails can cause problems with plugging channels and pumps, accumulate in digesters, and cause wear and tear on equipment. A sump can be provided in an effluent collection chamber upstream of the secondary clarifiers to facilitate removal of snails from the effluent. Snail separation has also been done by passing the trickling filter clarifier underflow waste sludge through screens or a vortex classifier (Daigger and Boltz, 2011).

Methods to control snail growth involve the use of toxic chemicals or adverse environmental conditions. These include alkaline treatment, high ammonia concentration and pH, saline water dosing, chlorination, and copper sulfate at 0.4 g/L. The use of molluscicides such as metaldehyde, niclosamide and trifenmorph can also suppress snail growth (WEF, 2011). The use of toxic inhibitors must be short term and consider the effect on the

Figure 9–9

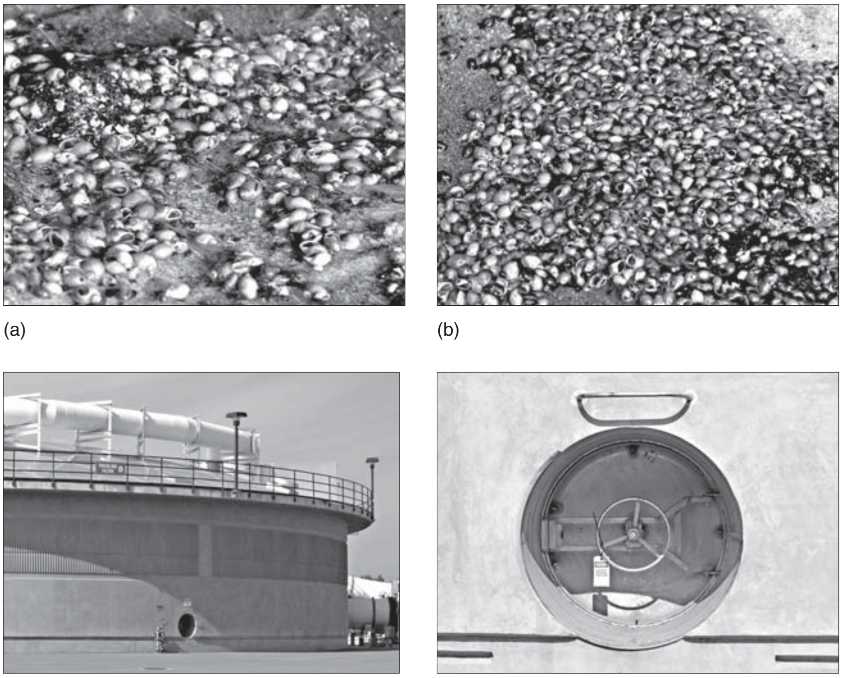
Trickling filter odor management units: (a) view of carbon treatment system, (b) view of acid scrubber, (c) view of biotrickling filter, and (d)view of media used in chemical scrubber.



(a) (b) (c) (d)

Figure 9–10

Trickling filter pests and their control: (a) and (b) view of snails on ledge inside tower trickling filter shown on Fig.9–8(e). The length of the snails varies from 3 to 5 mm, (c) view of tower trickling filter with water-tight access ports designed to be flooded, and (d) view of water-tight access port.



(a) (b) (c) (d)

nitrifying bacteria and WWTP effluent quality. It has been shown that at an aqueous free ammonia (NH_3) concentration in excess of 150 mg/L resulted in 100 percent snail mortality (Lacan et al., 2000). Successful snail control using this method at the Reno, Nevada Truckee Meadows Water Reclamation Facility was first reported by Gray et al. (2000). A tertiary nitrification trickling filter was contacted for 2 h with an ammonia-rich anaerobic digester centrate with sodium hydroxide addition to a pH of 9.2 to shift a high fraction of the total ammonia nitrogen concentration to free ammonia. Annual treatment of an isolated nitrifying trickling filter with a high ammonia concentration sludge dewatering liquor and sodium hydroxide additions was reported by Pearce and Jarvis (2011) to maintain successful snail growth control. Snail control by flooding at high pH (pH 9 to 10) was demonstrated by Parker et al. (1997) at the Littleton-Englewood WWTP in Colorado. Both of these methods require that the trickling filter tower tanks can be flooded fully and withstand hydrostatic pressures [see Fig. 9–10(c) and (d)].

Process Design Considerations for BOD Removal

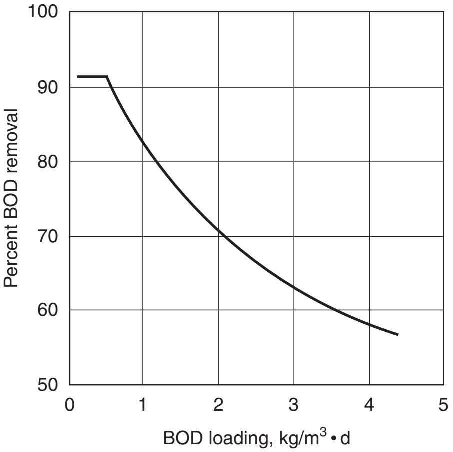
The trickling filter process appears simple, consisting of a bed of packing material through which wastewater flows and a clarifier. However, a trickling filter is a very complex system in terms of the physical characteristics of the attached growth and internal hydrodynamics. The biofilm thickness, surface contours, and degree of media coverage are difficult to predict. The attached growth is not uniformly distributed in the trickling filter (Hinton and Stensel, 1994), the biofilm thickness can vary, and the biofilm solids concentration may range from 40 to 100 g/L. The liquid flowrate through the media depth varies with time after dosing and does not uniformly flow over the entire packing surface area, which is referred to as the wetting efficiency. With the inability to quantify the biological and hydrodynamic properties of field trickling filter systems, broader parameters such as volumetric organic loading, unit area loading, and hydraulic loading have been used as design and operating parameters to relate to treatment efficiency. In view of these complexities, trickling filter designs and performance are based mainly on empirical relationships derived from pilot plant and full-scale plant experience. In this section, trickling filter performance for BOD removal and nitrification, factors that affect performance, and commonly used process design approaches are reviewed.

Effluent Characteristics. Historically, trickling filters have been considered to have major advantages of using less energy than activated sludge treatment and being easier to operate, but have disadvantages of more potential for odors and lower quality effluent. Some of these shortcomings, however, have been due more to inadequate ventilation, poor clarifier design, inadequate protection from cold temperatures, and the dosing operation. With proper design and within the volumetric loadings given in Table 9–1, trickling filters have successfully demonstrated the ability to produce effluent BOD concentrations of ≤ 30 mg/L for BOD removal designs and effluent BOD and ammonia-N concentrations of ≤ 20 mg/L and ≤ 3 mg/L, respectively, for combined BOD and ammonia removal designs. Tertiary nitrification systems have been designed to achieve effluent ammonia-N concentrations less than 1.0 mg/L.

Volumetric Loading Criteria. Volumetric BOD loading has been correlated well with treatment performance for both BOD removal and nitrification in combined BOD and nitrification trickling filter designs. Bruce and Merkens (1970 and 1973) found that the organic loading rate controlled trickling filter performance and not the hydraulic loading rate. The original design model for rock trickling filters was developed by the National Research Council (1946) from field data that correlated BOD removal efficiency with the

Figure 9–11

Example of trickling filter performance at 20°C. Effect of BOD loading on BOD removal efficiency for plastic media trickling filter.



volumetric organic loading rate. For combined BOD removal and nitrification systems, nitrification efficiency has also been related to the volumetric BOD loading (Stenquist and Kelly, 1980; U.S. EPA, 1975; and Daigger et al., 1993). An example of the effect of BOD loading on BOD removal efficiency is illustrated on Fig. 9–11. At low BOD loadings, BOD removal efficiency reaches a maximum plateau level of about 90 percent. In actual plant operations considerable scatter will exist around the curve due to variations in solids sloughing, wastewater characteristics (sBOD fraction), and clarification efficiency.

Trickling Filter Effluent Recirculation. Recirculation is an important aspect of trickling filter design that affects the wetting efficiency, controls the biofilm thickness, and helps maintain aerobic conditions at the upper, more highly-loaded section of the biotower. Wetting efficiency, defined as the ratio of the wetted area to the specific surface area, ranged from 0.2 to 0.6 with the lowest value for high density random pack media in the study by Crine et al. (1990). An increase in the total hydraulic loading to the trickling filter by recirculation flow reduces the liquid residence time, but increases the wetting efficiency. Recirculation flow also assures that a sufficient dosing rate, as shown in Table 9–3, is provided to control biofilm thickness. Insufficient dosing rates and a greater biofilm thickness can decrease the effective treatment area of trickling filter plastic packing (Daigger and Boltz, 2011).

Dilution of the influent BOD concentration of higher strength wastewaters by recirculation decreases the biological oxidation rates in the upper initial sections of a trickling filter to help maintain aerobic operating conditions. In addition, more oxygen to the tower is provided by the greater amount of flow leaving the distributor. A total hydraulic loading rate (influent flow plus recirculation flow) of greater than 0.5 $L/m^2 \cdot s$ is recommended for vertical-flow plastic media. Shallow cross-flow plastic media is able to provide improved spreading and wetting efficiency, and have been operated at about 0.25 $L/m^2 \cdot s$ in shallow trickling filter applications.

A trickling filter pumping station is designed to lift primary effluent and recirculated trickling filter effluent to the distributor system above the trickling filter packed media. Relatively low-head submersible or vertical turbine pumps are used. Variable frequency drives may be used on the pumps for operational flexibility.

Solids Production. Solids production from trickling filter processes will depend on the wastewater characteristics and the trickling filter loading. At lower organic loading rates, a greater amount of the particulate BOD is degraded, and because the biomass has a longer SRT, less biomass is produced. A procedure that can be used to evaluate the solids production for trickling filters is presented later in this chapter under in Sec. 9–4, which deals with combined trickling filter activated sludge processes.

Secondary Clarification. The function of clarifiers that follow trickling filters is to produce a clarified effluent. They differ from activated sludge settling tanks in that the clarifier has a much lower suspended solids content in the feed, and sludge thickening and recirculation is not necessary. The clarifier underflow is for the purpose of transporting waste sludge to the primary clarifier or sludge processing facilities.

Problems with Shallow Clarifiers. Trickling filter performance has historically suffered from poor clarifier designs. The use of shallow clarifiers (about 2.1 m depth) for trickling filter applications with relatively high surface overflow rates (about 1.7 m/h) was recommended in previous versions of the "Ten State Standards" (GLUMRB, 1997). Unfortunately, the use of shallow clarifiers typically resulted in poor clarification efficiency. Clarifier overflow rates recommended currently in the Ten State Standards are more inline with those used for the activated sludge process.

Use of Deep Clarifiers. Deeper clarifiers and lower surface overflow rates are now generally recommended with an interest in obtaining effluent TSS and BOD concentrations well below 20 mg/L. Surface overflow rates as a function of side water depth, as recommended in the WEF Biofilm Reactors MOP 35 (2011) are summarized in Table 9–5 and illustrated on Fig. 9–12. Clarifier designs for trickling filters should be similar to designs used for activated sludge process clarifiers (see Sec. 8–11 in Chap. 8), with appropriate feedwell size and depth and increased sidewater depth.

Mass-Transfer Limitations. One of the concerns in the process design for trickling filters is at what organic loading the filter performance becomes limited by oxygen transfer. When this condition occurs, treatment efficiency at the higher organic load is limited and odors may be produced due to anaerobic activity in the biofilm. Based on an evaluation of the data in the literature, for influent BOD concentrations in the range of 400 to 500 mg/L oxygen transfer may become limiting (Schroeder and Tchobanoglous, 1976). Hinton and Stensel (1994) reported that oxygen availability controlled organic substrate removal rates at soluble biodegradable COD loadings above 3.3 kg/m^3·d, which is on the upper end of BOD loadings for partial BOD removal applications as shown in Table 9–1. Note that recirculation flows can help reduce mass transfer limitations at the top of the trickling filter biotower.

Table 9–5

Recommended surface overflow rate (SOR) as a function of side water depth (SWD) for trickling filter and RBC secondary clarifiers with floor slope ≥1:1.2[a]

SWD, m	Average SOR, m/h	Maximum SOR, m/h
1.83 to 3.05	$\leq 0.092(\text{SWD})^2$	$\leq 0.182(\text{SWD})^2$
3.05 to 4.57	$\leq 0.278(\text{SWD})$	$\leq 0.556(\text{SWD})$

[a] Adapted from WEF (2011).

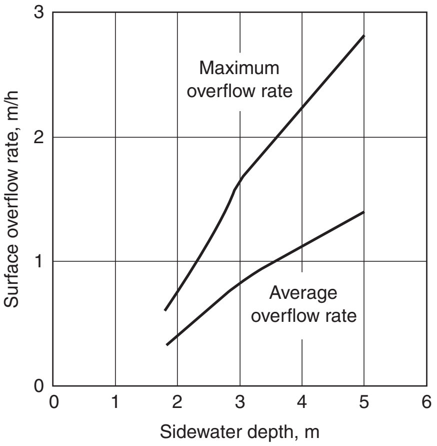


Figure 9–12

Typical secondary clarifier surface overflow rates as a function of clarifier sidewater depth.

EXAMPLE 9–2 **Trickling Filter Loadings** A 10-m diameter single stage trickling filter contains cross-flow plastic packing at a depth of 6.1 m. Primary effluent with the characteristics given below is applied to the filter. What is the volumetric BOD and TKN loading? What is the approximate BOD removal efficiency at 20°C? Can nitrification be expected?

Primary effluent wastewater characteristics:

Item	Unit	Value
Flowrate	m^3/d	4000
Temperature	°C	15
BOD	g/m^3	120
TSS	g/m^3	80
TKN	g/m^3	25

Solution

1. Determine the volume of the trickling filter packing material.

   Volume, $V = (A)(D)$

   $$A = \frac{\pi(10\text{ m}^2)}{4} = 78.5\text{ m}^2$$

   $$V = (78.5\text{ m}^2)(6.1\text{ m}) = 479\text{ m}^3$$

2. Determine the BOD Loading.

   BOD loading rate $= QS_O/V$

   $$= \frac{(4000\text{ m}^3/\text{d})(120\text{ g/m}^3)(1\text{ kg}/10^3\text{ g})}{479\text{ m}^3} = 1.0\text{ kg/m}^3\cdot\text{d}$$

3. Estimate the approximate BOD removal efficiency.

   From Fig. 9–11, at a loading of 1.0 kg BOD/m^2·d, the BOD removal efficiency is about 82 percent.

4. Can nitrification be expected?

   No. Based on the data given in Table 9–1, the BOD loading for combined BOD removal and nitrification is too high (1.0 $kg/m^3 \cdot d$, which is much greater than 0.25 $kg/m^3 \cdot d$). At the higher BOD loading, the heterotrophic bacteria outcompete the nitrifying bacteria for sites on the packing surface and, thus, control the slime biomass population.

Process Analysis for BOD Removal

The first empirical design equations were developed from field analyses of rock trickling filters. Later models were developed and applied for plastic packing trickling filters based on pilot plant and full scale plant operation and performance. The first empirical design equations for a rock trickling filter were developed by the National Research Council (1946) based on a data analysis on the performance of 34 systems in military installations treating domestic wastewater. These formulations can be found in a number of literature sources including WEF (2011) and Tchobanoglous et al. (2003). Because plastic media is used most often today the design equations for plastic media towers are the focus of this section.

Formulations for Plastic Packing. In the general equations developed for trickling filters with plastic packing, BOD removal was related to the hydraulic loading rate. The formulations are based on the early work of Velz (1948), who observed that the BOD remaining with depth in a trickling filter could be modeled as a first-order relationship, and Schulze (1960), who described the hydraulic detention time.

Schulze Equation. Schulze (1960) proposed that the liquid contact time with the biofilm is proportional to the filter depth and inversely proportional to the hydraulic loading rate as follows:

$$t = \frac{CD}{\text{THL}^n} \tag{9–11}$$

$$\text{THL} = \frac{Q(1 + R)}{A} = (1 + R)q \tag{9–12}$$

where THL = total hydraulic loading rate, $m^3/m^2 \cdot d$
t = liquid contact time, d
C = constant for packing used
D = depth of packing, m
n = hydraulic constant for the packing material used
R = recirculation ratio
q = hydraulic loading rate based on primary clarifier effluent flow, $m^3/m^2 \cdot d$

According to Eq. 9–11, as the influent flowrate to the trickling filter increases the detention time does not decrease in direct proportion to the flow, because the liquid film thickness increases.

The change in BOD concentration in the filter with time is described by a first-order reaction as follows:

$$\frac{dS}{dt} = -kS \tag{9–13}$$

where k = an experimentally determined rate constant
S = BOD concentration at time t

and using Eq. (9–11) to determine the time in the filter, Schulze derived the following equation:

$$\frac{S_e}{S_i} = \exp\frac{-kD}{(\text{THL})^n} \tag{9–14}$$

where S_e = BOD concentration of settled filter effluent, g/m^3
S_i = influent BOD concentration in total flow to trickling filter, g/m^3
k = an experimentally determined rate constant
D = packing depth, m

The values of k and n determined by Schulze at 20°C were 0.69/d and 0.67, respectively.

Germain Equation. Germain applied the Schulze equation in 1966 to trickling filters with plastic packing (WEF, 2000) as follows:

$$\frac{S_e}{S_i} = \exp\left\{\frac{-kD}{[(1 + R)q]^n}\right\} \tag{9–15}$$

where S_e = BOD concentration of settled filter effluent, g/m^3
S_i = influent BOD concentration in total flow to trickling filter, g/m^3
k = wastewater treatability and packing coefficient, (L/s)$^{0.5}$/m^2 (based on $n = 0.5$)
D = packing depth, m
n = constant characteristic of packing used

The value for n is normally assumed to be 0.50 and pilot plant or full-scale plant influent and effluent BOD concentration data are used to solve for k. Values for k were developed from more than 140 pilot plant studies by Dow Chemical Company with vertical plastic packing with a specific surface area of about 90 m^2/m^3. Similar tests have been done by other suppliers for a variety of packings. Most of these tests have been done with packing depths of 6.1 to 6.7 m (20 to 22 ft).

It should be noted that the clarifier design, solids loading, and dosing cycle and method can affect pilot plant results used to calculate a value for k (Harrison and Daigger, 1987). In summary, the value for k is affected by many factors including the wastewater characteristics, filter and clarifier design, and operating conditions.

The commonly accepted temperature correction for k is as follows:

$$k_T = k_{20}(1.035)^{T-20} \tag{9–16}$$

Other Formulations. Other formulations have been proposed (WEF, 2011) to describe the performance of plastic packing filters, including models by Eckenfelder (1961) and Eckenfelder and Barnhart (1963). One of the modified equations termed the modified Velz equation as given below in Eq. (9–19) relates the trickling filtered clarifier effluent BOD concentration as a function of the primary effluent BOD, the recirculation ratio, a factor for the specific area of the packing, and temperature. The equation is derived from Eq. (9–15) in which $(k_{20}A_s)$ equals k and S_e/S_o is used instead of S_e/S_i by developing a relationship between S_e/S_i and S_e/S_o from a BOD mass balance of the primary effluent and recirculation flows entering the trickling filter.

$$QS_o + Q_R S_e = (Q + Q_R)S_i \tag{9–17}$$

$$\frac{S_i}{S_e} = \frac{RS_e + S_o}{(1 + R)S_e} \tag{9–18}$$

Inverting Eq. (9–15) and substituting Eq. (9–18) for S_i/S_e relates the trickling filter clarifier effluent BOD concentration to the primary effluent BOD concentration and the recirculation ratio.

$$S_e = \frac{S_o}{(R + 1)\exp\left\{\dfrac{k_{20}A_s D\theta^{T-20}}{[q(R + 1)]^n}\right\} - R} \tag{9–19}$$

where S_o = influent BOD based on primary effluent flow, g/m^3
S_e = effluent BOD, g/m^3
k_{20} = filter treatability constant at 20°C, (L/s)$^{0.5}$/m
A_s = clean packing specific surface area, m^2/m^3
D = depth of packing, m
θ = temperature correction coefficient, 1.035
q = hydraulic loading rate based on primary effluent flow, L/m^2·s
R = ratio of recirculation flowrate to primary effluent flowrate
n = constant characteristic of packing used

Because the BOD removal is determined as a function of hydraulic loading rate, application of Eqs. (9–15) and (9–19) without regard to the fundamental effect of organic loading can lead to erroneous designs. For example, to achieve the same BOD removal efficiency, Eq. (9–15) would predict a smaller packing volume requirement by increasing the packing depth to greater than 6.1 m. However, as the volume is reduced, the organic loading increases and thus the treatment efficiency should decline. By assuming that the BOD removal efficiency is equal at the same organic loading, the value for k had to be adjusted for depth and influent BOD concentration. The k value is normalized to a specified depth and influent BOD concentration as follows (WEF, 2011):

$$k_2 = k_1\left(\frac{D_1}{D_2}\right)^{0.5}\left(\frac{S_1}{S_2}\right)^{0.5} \tag{9–20}$$

where k_2 = normalized value of k for the site-specific packing depth and influent BOD concentration
k_1 = k value at depth of 6.1 m (20 ft) and influent BOD of 150 g/m^3
S_1 = 150 g BOD/m^3
S_2 = site-specific influent BOD concentration, g BOD/m^3
D_1 = 6.1 m (20 ft) packing depth, m
D_2 = site-specific packing depth, m

Normalized values of k at 20°C determined from Dow Chemical Company pilot plant studies are summarized in Table 9–6. These are examples of pilot plant results and provide a rough approximation of treatability differences for different wastewaters. Wastewaters that would have the lowest degradation rates are from refineries, Kraft pulp mills, and textile mills.

Table 9–6

Normalized Germain equation $k_{20}A_S$ values [Eq. (9–19)] for plastic packing media (100 m^2/m^3) for different wastewaters

Type of wastewater	value, $(L/s)^{0.5}/m^2$
Domestic	0.210
Front canning	0.181
Kraft mill	0.108
Meat packing	0.216
Pharmaceutical	0.221
Potato processing	0.351
Refinery	0.059
Sugar processing	0.165
Synthetic dairy	0.170
Textile mill	0.107

Note: $[(L/s)^{0.5}/m^2] \times 0.3704 = (gal/min)^{0.5}/ft^2$.

EXAMPLE 9–3 **Design of Trickling Filter with Plastic Packing** Given the following design flowrates and primary effluent wastewater characteristics, determine the following design parameters for a trickling filter design assuming 2 towers at 6.1 m depth, cross-flow plastic packing with a specific surface area of 90 m^2/m^3, a packing coefficient n value of 0.5, a recirculation ratio of 1.0, and a 2-arm distributor system. The required minimum wetting rate = 0.5 $L/m^2 \cdot s$. Assume a secondary clarifier depth of 4.0 m.

Design Criteria:

Parameter	Unit	Primary effluent	Target effluent
Flow	m^3/d	15,140	
BOD	g/m^3	125	20
TSS	g/m^3	65	20
Minimum. temp.	°C	14	

Using the above information determine:

1. Diameter of tower trickling filter, m
2. Volume of packing required, m^3
3. Total pumping rate, m^3/h
4. Flushing and normal dose rate, mm/pass
5. Flushing and normal distributor speeds, min/rev
6. Clarifier diameter, m (Assume the ratio of the peak to average flowrate is 1.5)

Solution

1. Determine k_{20} for the design conditions using Eq. (9–20).

$$k_2 = k_1\left(\frac{D_1}{D_2}\right)^{0.5}\left(\frac{S_1}{S_2}\right)^{0.5}$$

   a. Solve for k_2.

   From Table 9–6, $k = 0.210\ (L/s)^{0.5}/m^2$ [Note: $k = k_{20}A_s$ in Eq. (9–19)]

Trickling filter depth = 6.1 m

$$= 0.210\left(\frac{6.1}{6.1}\right)^{0.5}\left(\frac{150}{125}\right)^{0.5} = 0.230\ (\text{L/s})^{0.5}/\text{m}^2$$

b. Correct k_2 for temperature effect using Eq. (9–16).

i. $k_T = k_{20}\ (1.035)^{T-20}$

ii. $k_{14} = 0.230(1.035)^{14-20} = 0.187\ (\text{L/s})^{0.5}/\text{m}^2$

2. Determine the hydraulic loading rate and the filter area, volume, and diameter.

a. Using Eq. (9–19) with $k_T = (k_{20}A_s)\theta^{T-20} = 0.187\ (\text{L/s})^{0.5}/\text{m}^2$ determine the hydraulic loading rate.

$$S_e = \frac{S_o}{(R+1)\exp\left\{\dfrac{k_T(D)}{[q(R+1)]^n}\right\} - R}$$

Rearrange to get following:

$$[q(1+R)] = \left\{\frac{k_T D}{\ln\left[\dfrac{S_o + RS_e}{S_e(1+R)}\right]}\right\}^{1/n}$$

$$[q(1+1)] = \left\{\frac{(0.187\ \text{L/m}^2\cdot\text{s})(6.1\ \text{m})}{\ln\left[\dfrac{(125\ \text{g/m}^3) + (1)(20\ \text{g/m}^3)}{(20\ \text{g/m}^3)(1+1)}\right]}\right\}^2$$

$$q = 0.443\ \text{L/m}^2?\text{s}$$

b. Determine the tower area.

$Q = 15{,}140\ \text{m}^3/\text{d} = 175.2\ \text{L/s}$

Filter area $= Q/q = 175/0.443 = 395.5\ \text{m}^2$

c. Determine the packing volume.

Packing volume $= (395.5\ \text{m}^2)\ (6.1\ \text{m}) = 2412\ \text{m}^3$

d. Determine the tower diameter.

Area/tower $= 395.5\ \text{m}^2/2 = 197.75\ \text{m}^2$

Diameter = 15.9 m each

Two towers each with a diameter of 16 m

3. Determine the pumping rate.

$q + q_r = (1+R)q = (1+1)0.443\ \text{L/m}^2\cdot\text{s} = 0.886\ \text{L/m}^2\cdot\text{s}$

Total pumping rate $= (0.886\ \text{L/m}^2\cdot\text{s})(\ 395.5\ \text{m}^2)$

$= 350.4\ \text{L/s} = 1261\ \text{m}^3/\text{h}$

4. Determine flushing and normal dose rate using the data given in Table 9–3.
   a. Determine BOD loading.

$$\text{BOD loading} = Q\,S_o/V$$

$$= \frac{(15{,}140\ \text{m}^3/\text{d})(125\ \text{mg/L})(1\ \text{kg}/10^3\ \text{g})}{2412\ \text{m}^3}$$

$$= 0.79\ \text{kg/m}^3\cdot\text{d}$$

   b. Determine the dosing rates.

   From Table 9–3, the estimated flushing and operation dose rates are:

   i. Flushing dose = 150 mm/pass

   ii. Operating dose = 75 mm/pass

5. Determine the distributor speed using Eq. (9–1).
   a. For normal operation:

$$n = \frac{(1 + R)\,q(1000\ \text{mm/min})}{(N_A)(DR)(60\ \text{min/h})},\ \text{where } q = \text{m}^3/\text{m}^2\cdot\text{h}$$

$$q = (0.443\ \text{L/m}^2\cdot\text{s})\left(\frac{3600\ \text{s}}{\text{h}}\right)\left(\frac{1\ \text{m}^3}{10^3\ \text{L}}\right) = 1.6\ \text{m}^3/\text{m}^2\cdot\text{h}$$

$$R = 1.0$$

$$n = \frac{(1 + 1)(1.6)(1000)}{(2)(75)(60)} = 0.36\ \text{rev/min (i.e. 2.8 min/rev)}$$

   b. For flushing operation:

$$n = \frac{(1 + 1)(1.6)(1000)}{(2)(150)(60)} = 0.18\ \text{rev/min (i.e. 5.6 min/rev)}$$

   Note because of the different speed requirements for normal and flushing operation, a distributor drive with variable speed capability should be used.

6. Determine clarifier diameter using the average curve from Fig. 9–12.

   Clarifier depth = 4.0 m

   From Fig. 9–12, the average SOR = 1.1 m/h

   The recommended ratio of the peak to average overflow rate from Table 9–5 is 2.0 (0.556/0.278). Because the ratio of the peak to average flowrate is 1.5, the average overflow rate controls the design.

   Flowrate = $(15{,}140\ \text{m}^3/\text{d})/(24\ \text{h/d}) = 630.8\ \text{m}^3/\text{h}$

   Clarifier area = $630.8/1.1 = 573.5\ \text{m}^2$

   Use 2 clarifiers

   Area for each = $573.5\ \text{m}^2/2 = 286.7\ \text{m}^2$

   Diameter of each = 14.1 m

7. Design Summary.

Parameter	Unit	Value
Number of filters	number	2
Diameter	m	16
Depth	m	6.1
Total packing volume	m^3	2412
BOD loading	$kg/m^3·d$	0.79
Hydraulic loading rate	$L/m^2·s$	0.886
Total pumping rate	m^3/h	1261
Recirculation ratio	unitless	1.0
Distributor arms	number	2
Normal distributor speed	min/rev	2.8
Flushing distributor speed	min/rev	5.6
Clarifiers	number	2
Clarifier depth	m	4.0
Clarifier diameter	m	14.1

Process Analysis for Nitrification

Two types of process design approaches have been used to accomplish biological nitrification in trickling filters, either in a combined system along with BOD removal or in a tertiary application following secondary treatment and clarification for BOD removal. Tertiary nitrification can be a low energy, cost-effective process for ammonia removal after secondary treatment by suspended growth or fixed-film processes. Empirical design approaches based on pilot plant and full-scale plant results are again used to guide nitrification designs in view of the difficulty in predicting the actual biofilm coverage area, wetting efficiency, and biofilm thickness and density. The nitrification efficiency in combined systems has been correlated with the BOD volumetric loading (kg BOD/m^3·d) and BOD surface area loading (kg BOD/1000 m^2·d) from full-scale and pilot plant results. For tertiary nitrification applications the ammonia surface loading is the common design parameter (g N/m^2·d).

Combined BOD Removal and Nitrification. In combined BOD removal and nitrification trickling filters, the heterotrophic bacteria will outcompete the nitrifying bacteria for media surface area due to their faster growth rate and higher biomass yield. Most of the nitrifying biomass growth occurs only after the soluble BOD is removed to below 5 to 10 mg/L (Harremoës, 1982; Figueroa and Silverstein, 1991; and Parker and Richards, 1986). Much lower BOD volumetric loading rates than that used for BOD removal only are used in combined BOD removal and nitrification trickling filters to provide additional media area for nitrifying bacteria.

For 90 percent nitrification efficiency, a BOD loading of less than 0.08 kg BOD/ m^3·d (5 lb BOD/1000 ft^3·d) is recommended (WEF, 2011) for rock media. At a loading of about 6.3 kg BOD/ m^3·d (14 lb BOD/1000 ft^3·d), about 50 percent nitrification efficiency could be expected. An evaluation of nitrification in a combined system using low-density, crossflow packing by Daigger et al. (1994) suggested a loading of less than 0.20 kg BOD/ m^3·d (12.5 lb BOD/1000 ft^3·d) for 90% nitrification efficiency.

Nitrification efficiency has also been related to the BOD loading based on the packing surface area. In comparing nitrification performance for both rock and cross-flow plastic packing, Parker and Richards (1986) found that the nitrification efficiency was similar at similar BOD surface loading rates (g BOD/m^2·d) for both packings. A surface loading rate as low as 2.0 kg BOD/ m^2·d (0.5 lb BOD/1000 ft^2·d) is necessary for ≥ 90 percent ammonia-N removal (Daigger and Boltz, 2011). Recirculation and a dosing rate strategy to control biofilm thickness improves nitrification performance (Parker et al., 1997).

Daigger et al. (1994) found that the oxidation of BOD and ammonia-N in trickling filters with plastic packing could be characterized by a volumetric oxidation rate defined as follows:

$$\text{VOR} = \frac{[S_o + 4.6(\text{NO}_\text{x})]Q}{(V)(10^3\ \text{g/1 kg})} \tag{9–21}$$

where VOR = volumetric oxidation rate, kg/m^3·d
S_O = influent BOD concentration, g/m^3
NO_x = amount of influent ammonia-N oxidized, g/m^3
Q = influent flowrate, m^3·d
V = packing volume, m^3

Using Eq. (9–21), the volumetric oxidation rate for three plants was determined and ranged from 0.4 to 1.3 kg/m^3·d. The amount of nitrification can then be estimated from an assumed VOR value and the influent BOD concentration.

Okey and Albertson (WEF, 2000) found a relationship between the specific nitrification rate (g/m^2·d) and the influent BOD/TKN ratio for combined systems based on data from four different studies.

$$R_n = 0.82\left(\frac{\text{BOD}}{\text{TKN}}\right)^{-0.44} \tag{9–22}$$

where R_n = specific nitrification rate, g/m^2·d
$\frac{\text{BOD}}{\text{TKN}}$ = influent BOD to TKN ratio, g/g

The data included in this correlation were for operation at temperatures ranging from 9 to 20°C. The authors concluded that the DO concentration had a greater effect on the nitrification rates than temperature. The effect of DO concentration is supported by fundamental mass transfer considerations in which it can be shown that a bulk liquid DO concentration of 2.8 mg/L is required for nitrification without oxygen diffusion limitations, at a liquid ammonia-N concentration of 1.0 mg/L.

Pearce and Edwards (2011) proposed a model for predicting the effluent ammonia-N concentration as a function of ammonia-N, BOD, and hydraulic loading rates relative to the packing surface area and temperatures.

$$\text{NH}_4\text{-N}_\text{e} = 20.81(\text{BOD}_\text{L})^{1.03}(\text{NH}_4\text{-N}_\text{L})^{1.52}(\text{Iv})^{-0.36}(\text{T})^{-0.12} \tag{9–23}$$

Where, $\text{NH}_4\text{-N}_\text{e}$ = average effluent ammonia-N concentration, mg/L
BOD_L = BOD specific surface loading rate, g/m^2·d
$\text{NH}_4\text{-N}_\text{L}$ = ammonia-N specific surface loading rate, g/m^2·d
Iv = specific hydraulic surface loading rate L/m^2·d
T = filter effluent temperature, °C

The model fitted experimental data for various hydraulic and organic loading conditions ($R^2 = 0.78$). The temperature effect is not as pronounced as for suspended growth systems due the combined effects of the high biomass inventory, microbial kinetics, and mass transfer in the trickling filter.

EXAMPLE 9–4 **Combined BOD Removal and Nitrification in a Trickling Filter with Plastic Packing** Determine the volume and area of plastic packing required for 90 percent TKN removal in trickling filter with a depth of 6.1 m for the wastewater characteristics given. Assume low-density cross-flow packing with a BOD volumetric loading of 0.20 kg BOD/m^3·d. How does the computed volume compare to the volume predicted based on a volumetric oxidation rate of 0.40 kg/m^3·d? How does the specific nitrification rate compare to the relationship shown in Eq. 9–22? What percent nitrification is estimated from the nitrification model in Eq. 9–23? Assume the specific surface area of the plastic packing material is 90 m^2/m^3 from Table 9–2 and the trickling filter effluent temperature is 20°C.

Also, determine the hydraulic loading rate.

Wastewater characteristics:

Parameter	Unit	Value
Flowrate	m^3/d	8000
BOD	g/m^3	160
TKN	g/m^3	25
TSS	g/m^3	70

Solution

1. Determine the packing volume based on a loading of 0.20 kg BOD/m^3·d.

$$\text{Volumetric BOD loading} = 0.20\ \text{kg/m}^3\cdot\text{d} = \frac{(8000\ \text{m}^3/\text{d})(160\ \text{g/m}^3)(1\ \text{kg}/10^3\ \text{g})}{\text{Packing volume, m}^3}$$

$$\text{Packing Volume} = \frac{(8000\ \text{m}^3/\text{d})(160\ \text{g/m}^3)(1\ \text{kg}/10^3\ \text{g})}{(0.20\ \text{kg/m}^3\cdot\text{d})} = 6400\ \text{m}^3$$

2. Determine the volume based on the volumetric oxidation rate using Eq. (9–21).

$$\text{VOR} = \frac{[S_o + 4.6(\text{NO}_x)]Q}{(V)(10^3\ \text{g/1 kg})} = 0.40\ \text{kg/m}^3\cdot\text{d}$$

$$V = \frac{[S_o + 4.6(\text{NO}_x)]Q}{(0.40\ \text{kg/m}^3\cdot\text{d})(10^3\ \text{g/1 kg})}$$

$$V = \frac{[160\ \text{g/m}^3 + 4.6(25\ \text{g/m}^3)](8000\ \text{m}^3/\text{d})}{(0.40\ \text{kg/m}^3\cdot\text{d})(10^3\ \text{g/1 kg})} = 5500\ \text{m}^3$$

3. Determine the specific nitrification rate at 90% removal.

Packing surface area = 6400 m^3 (90 m^2/m^3) = 576,000 m^2

$$R_n = \frac{(Q)(\text{NO}_x)}{\text{Surface Area}} = \frac{(0.90)(8000\ \text{m}^3/\text{d})(25\ \text{g/m}^3)}{576{,}000\ \text{m}^2} = 0.32\ \text{g/m}^2\cdot\text{d}$$

4. Determine the specific nitrification rate from Eq. (9–22).

BOD/TKN = 160/25 = 6.4

$$R_n = 0.82\left(\frac{\text{BOD}}{\text{TKN}}\right)^{-0.44}$$
$$= 0.82\,(6.4)^{-0.44}$$
$$= 0.36\ \text{g/m}^2\cdot\text{d}$$

5. Determine the hydraulic loading rate.

$$\text{Filter area} = \frac{\text{volume}}{\text{depth}} = \frac{6400\ \text{m}^3}{6.1\ \text{m}} = 1049\ \text{m}^2$$

Hydraulic loading rate, q

$$q = \frac{Q}{A} = \frac{(8000\ \text{m}^3/\text{d})(10^3\ \text{L}/1\ \text{m}^3)(\text{d}/1440\ \text{min})(\text{min}/60\ \text{s})}{1049\ \text{m}^2} = 0.09\ \text{L/m}^2\cdot\text{s}$$

To meet the minimum hydraulic loading rate given previously as 0.5 L/m^2·s, recirculation will be required.

6. Estimate the effluent NH_4-N concentration using Eq. (9–23).

$$\text{NH}_4\text{-N}_e = 20.81(\text{BOD}_L)^{1.03}(\text{NH}_4\text{-N}_L)^{1.52}(\text{Iv})^{-0.36}(T)^{-0.12}$$

$$\text{BOD}_L = \frac{(8000\ \text{m}^3/\text{d})(160\ \text{g/m}^3)}{576{,}000\ \text{m}^2} = 2.22\ \text{g/m}^2\cdot\text{d}$$

$$\text{NH}_4\text{-N}_L = (0.32\ \text{g/m}^2\cdot\text{d})/0.9 = 0.36\ \text{g/m}^2\cdot\text{d (from step 3)}$$

$$\text{Iv} = \frac{(8000\ \text{m}^3/\text{d})(10^3\ \text{L}/1\ \text{m}^3)}{576{,}000\ \text{m}^2} = 13.9\ \text{L/m}^2\cdot\text{d}$$

$$\text{NH}_4\text{-N}_e = 20.81(2.22)^{1.03}(0.36)^{1.52}(13.9)^{-0.36}(20)^{-0.12} = 1.2\ \text{mg/L}$$

$$\text{Percent nitrification} = \frac{(100)[(25 - 1.2)\ \text{mg/L}]}{(25\ \text{mg/L})} = 95.2\%$$

Comments The computed value for the volume required based on the BOD volumetric loading rate is higher than that predicted by the volumetric oxidation rate (Daigger et al. (1994). For the wastewater BOD/N ratio, the specific nitrification rate is close to that predicted based on the BOD/TKN ratio using Eq. (9–22) and the percent nitrification is close to that predicted by Eq. (9–23).

Tertiary Nitrification. A number of facilities exist where trickling filters with plastic packing are used for nitrification after secondary treatment. Advantages for tertiary nitrification include (1) low energy consumption, (2) simplicity of operation, and (3) stable performance.

For tertiary nitrification applications, very little BOD is applied to the trickling filter and a thin biofilm develops on the packing that consists of a high proportion of nitrifying bacteria. Effluent NH_4-N concentrations will vary with summer and winter operation and can range from <1.0 mg/L at warm temperatures and from <1 to 4 mg/L at cold temperatures. Hydraulic loading rates may range from 0.40 to 1.0 L/m^2·s and recycle is commonly used to maintain packing surface wetting. Some tertiary nitrification systems have been constructed without downstream liquid-solids separation because of the low net biomass yield by the nitrifying bacteria. This depends on site specific wastewater characteristics and treated effluent goals.

Design and Operation. In the design and operation of tertiary nitrifying trickling filters it is important to consider (1) the media type and surface area density, (2) mechanical ventilation, (3) hydraulic loading rates and recycle, (4) minimizing slug ammonia loads, and (5) control of predatory micro fauna. It is generally well accepted that in the upper portion of the trickling filter the nitrification rate is limited by oxygen availability and diffusion into the biofilm. To mitigate the oxygen limitation, forced draft air is generally used to assure maximum oxygen availability. Higher hydraulic rates including recirculation that provide better wetting efficiency and agitation of the biofilm surface generally produce better performance. Distributor speed control to provide a dosing rate in the range 25 to 75 mm/pass and flushing intensity ≥300 mm/pass is recommended (WEF, 2011). Because plugging is less of an issue, a medium density packing material is preferred (i.e., specific surface area of 138 m^2/m^3) to provide more area as a function of the percent of the reactor volume. Equalization of high ammonia concentration flows from solids processing is recommended to minimize slug loads and diurnal load fluctuations to obtain a low and consistent effluent ammonia concentration. Snail growth [see Figs. 9–10(a) and (b)] in nitrifying trickling filters has occurred and can result in a serious loss of the nitrifying bacteria population and treatment efficiency. Predation control methods are discussed after the process design methodology.

Nitrification Rate. The rate of nitrification in a trickling filter varies with packing media depth and is related to bulk liquid oxygen and ammonia-N concentrations and hydraulic application rate. In the upper portion of a nitrification tower, the ammonia-N concentration may be high enough so that the nitrification rate is oxygen limited, and thus zero order with respect to the ammonia-N concentration. Further down in the packing, as the ammonia-N concentration decreases, the nitrification rate is limited by the ammonia-N concentration and thus decreases. The decline in nitrification rate is further affected by less growth of nitrifying bacteria due to the low amount of ammonia-N available. The use of nitrification trickling filters in series with operational modifications has been shown to compensate for this limitation (Boller and Gujer, 1986). The order of operation of the towers is reversed every few days so that a higher nitrifying bacteria population can be developed and be available where the ammonia-N concentration is low. Anderson et al. (1994) showed a 20 percent improvement in nitrification efficiency with this method.

The nitrification removal efficiency has been related to the packing surface area and correlated with the nitrogen removal rate per unit of surface area (g $N/m^2 \cdot d$) (Okey and Albertson, 1989; Parker et al., 1990; WEF, 2011). Boller and Gujer (1986) developed an empirical equation that relates the ammonia-N removal flux from the bulk liquid as a function of the bulk liquid ammonia-N concentration. The ammonia-N removal flux is equal to the nitrification rate per unit area in the biofilm.

$$J_N(z) = J_{N,\max}\left(\frac{N}{K_N + N}\right) \tag{9–24}$$

where $J_N(z)$ = NH_4-N removal flux, $g/m^2 \cdot d$

$J_{N,\max}$ = maximum ammonia-N removal flux at temperature T, $g/m^2 \cdot d$

N = bulk liquid ammonia-N concentration, g/m^3

K_N = half-velocity ammonia-N coefficient, g/m^3

The value of $J_{N,\max}$ has been determined by observations on zero-order nitrification rates in tertiary trickling filter operations where the ammonia-N concentration is significantly greater than K_N. As noted above, approximate zero-order nitrification rates may occur in the upper portion of the trickling filter tower until the ammonia-N concentration decreases to below 6.0 mg/L. After zero order removal, the packing media surface biofilm growth declines due to the lower ammonia-N concentration and lower ammonia-N removal rate. The value of

Table 9–7

Reported maximum ammonia-N removal flux for tertiary nitrification trickling filters

Location	Packing[a]	Range of $J_{N,\max}$ values, g N/m²·d	Reference
Central Valley, UT	XF 138	2.1–2.9	Parker (1990)
Malmö, Sweden	XF 138	1.6–2.8	Parker (1990)
Littleton/Englewood, CO	XF 136	1.2–2.3	Parker (1997)
Midland, MI	VF 89	1.1–1.8	WEF (2011)
Lima, OH	VF 89	1.2–1.8	WEF (2011)
Zurich, Switzerland	VF 92	1.6	WEF (2011)
Zurich, Switzerland	XF 223	1.2	WEF (2011)

[a] XF = cross-flow, VF = vertical flow. Numbers correspond to specific surface area in m²/m³.

the maximum ammonia-N removal flux is decreased with depth to account for the decline in the surface biofilm growth. A reduction of 0.1 g/m²·d per meter of increasing depth has been proposed for a correction to the maximum ammonia-N removal flux (Daigger and Boltz, 2011).

Maximum surface nitrification fluxes ($J_{N,\max}$) vary widely as reported in Table 9–7. Cross-flow plastic packing appears to produce higher nitrification flux. The higher nitrification flux may be due to better flow distribution and greater wetting of the packing to develop a higher biofilm surface area. However, the flux was worst with the XF 223, higher density, cross-flow packing.

No temperature correction is used with Eq. (9–24) between 10 and 25°C. Other investigators have also observed minimal temperature effects for tertiary nitrification, and have attributed the minimal observed rate change more to the effect of dissolved oxygen concentration and hydraulics (Okey and Albertson, 1989). Below 10°C the following temperature correction coefficient is recommended (WEF, 2011).

$$J_{N,\max(T)} = J_{N,\max(10)}(1.045)^{(T-10)} \tag{9–25}$$

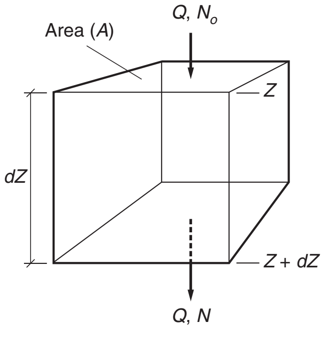


Figure 9–13

Schematic of trickling filter incremental depth section for derivation of Eq. (9–29).

Equation (9–24) is applied in a mass balance across an incremental depth of packing media, as shown on Fig. 9–13, to develop a general equation for the bulk liquid ammonia-N concentration with media depth. The ammonia-N removal rate across the incremental depth is equal to the flux of ammonia-N into the biofilm.

$$V\frac{dN}{dt} = -J_{N,\max}\left(\frac{N}{K_N + N}\right)(aV) \tag{9–26}$$

where V = volume of incremental section, m³
a = specific surface area of media, m²/m³

The volume of the incremental section is equal to Adz and dt equals dz(A)/Q. Substitution for dt in Eq. (9–26) with q equal to the hydraulic load (m³/m²) as defined previously yields:

$$dN = -J_{N,\max}\left(\frac{N}{K_N + N}\right)(a)\left(\frac{dZ}{q}\right) \tag{9–27}$$

Rearranging yields:

$$dN\left(\frac{K_N + N}{N}\right) = -\frac{J_{N,\max}}{q}(a)dZ \tag{9–28}$$

where Z = packing media depth, m
q = hydraulic loading rate, m³/m²·d

Integration of Eq. (9–28) with limits of N equal the influent concentration (N_o) at $Z = 0$ and N equal N at $Z = Z$ results in the following general expression for the ammonia-N concentration with depth.

$$(N_o - N) + K_N Ln\left(\frac{N_o}{N}\right) = \frac{ZaJ_{N,\max}}{q} \tag{9–29}$$

When recirculation is used an iterative solution approach is necessary, in which an assumed effluent ammonia-N concentration is first assumed to calculate N_o.

$$N_o = \frac{N_{\text{sec}} + RN}{1 + R} \tag{9–30}$$

where N_{sec} = NH_4-N concentration in secondary effluent feed to the nitrification tower, g/m^3
R = Recirculation ratio

The total hydraulic load is also used in place of q in Eq. (9–29) and equals $(1 + R)q$.

EXAMPLE 9–5 **Tertiary Nitrification Trickling Filter Design** Determine the total plastic packing media depth for a nitrification filter is to treat a flow of 6000 m^3/d after secondary treatment at a temperature of 15°C. The influent NH_4-N concentration is 25 g/m^3 and an effluent NH_4-N concentration of 1.0 g/m^3 is desired. Assume a $J_{N,\max}$ value of 1.8 $g/m^2{\cdot}d$ from information in Table 9–7 and a K_N value of 1.5 g/m^3. A medium density cross-flow plastic packing material with a specific surface area of 138 m^2/m^3 will be used, as little BOD or TSS exists in the influent wastewater to pose plugging problems. Assume that the $J_{N,\max}$ value is decreased by 0.10/m after the NH_4-N concentration is decreased to 6.0 g/m^3. Compare the plastic tower height and total packing volume needed for two design assumptions in which both provide a total hydraulic loading rate of 1.0 $L/m^2{\cdot}s$ to provide a sufficient media packing wetting efficiency: (a) the recirculation ratio = 0 and (b) the recirculation ratio = 1.0.

Solution with No Recycle

(a) $R = 0$ and $q = 1.0\ L/m^2{\cdot}s$

1. Convert q to units of $m^3/m^2{\cdot}d$ for use in Eq. (9–29).

$$q = (1.0\ L/m^2{\cdot}s)\left(\frac{1\ m^3}{10^3\ L}\right)\left(\frac{60\ s}{min}\right)\left(\frac{1440\ min}{d}\right) = 86.4\ m^3/m^2{\cdot}d$$

2. Solve for depth Z in Eq. (9–29), where N_o is 25.0 g/m^3 and N is 6.0 g/m^3. Because no recirculation is used, the NH_4-N concentration at the top of the tower is 25.0 g/m^3. $J_{N,\max} = 1.8\ g/m^2{\cdot}d$

$$(N_o - N) + K_N \ln\left(\frac{N_o}{N}\right) = \frac{ZaJ_{N,\max}}{q}$$

$$[(25.0 - 6.0)\ g/m^3] + (1.5\ g/m^3)\ln\left(\frac{25.0}{6.0}\right) = \frac{Z(138\ m^2/m^3)(1.8\ g/m^2{\cdot}d)}{(86.4\ m^3/m^2{\cdot}d)}$$

$Z = 7.4$ m

3. For the next 1 m of depth $J_{N,\max} = (1.8 - 0.1) g/m^2{\cdot}d = 1.7\ g/m^2{\cdot}d$

Solve for N using Eq. (9–29).

$$[(6.0 - N)\ g/m^3] + (1.5\ g/m^3)\ln\left(\frac{6.0}{N}\right) = \frac{1.0\ m(138\ m^2/m^3)(1.7\ g/m^2{\cdot}d)}{(86.4\ m^3/m^2{\cdot}d)}$$

$N = 3.9\ g/m^3$

4. Continue this calculation procedure until N is $\leq 1.0\ \mathrm{g/m^3}$.
5. The volume of packing media for each depth increment is the depth times the tower cross-section area.

$$\text{Area} = \frac{\text{flow}}{\text{hydraulic application rate}}$$

$$\text{Area} = \frac{Q}{q}$$

$$\text{Area} = \frac{(6000\ \mathrm{m^3/d})}{(86.4\ \mathrm{m^3/m^2{\cdot}d})} = 69.4\ \mathrm{m^2}$$

Plastic packing volume at 7.4 m depth = (A)(Z) = $(69.4\ \mathrm{m^2})(7.4\ \mathrm{m}) = 513.6\ \mathrm{m^3}$

Plastic packing volume for 1.0 m depth increment = $(69.4\ \mathrm{m^2})(1.0\ \mathrm{m}) = 69.4\ \mathrm{m^3}$

The results are summarized in the following table:

$J_{N,max}$, g/m²·d	NH_4-N, mg/L	Incremental depth, m	Volume, m³
1.8	6.0	7.4	513.6
1.7	3.9	1.0	69.4
1.6	2.2	1.0	69.4
1.5	1.0	1.0	69.4
	Total	10.4	721.8

The plastic tower depth = 10.4 m (or use two in series at 5.2 m each)

6. Determine the tower diameter.

Assume two towers in parallel.

Cross-section area per tower = $69.4\ \mathrm{m^2}/2 = 34.7\ \mathrm{m^2}$

$$\text{Area} = A = \frac{\pi D^2}{4},\ \text{Diameter} = \sqrt{\frac{4A}{\pi}}$$

$$\text{Diameter} = \sqrt{\frac{4(34.7\ \mathrm{m^2})}{3.14}} = 6.64\ \mathrm{m}$$

Solution with Recycle

(b) $R = 1.0,\ q = 0.50\ \mathrm{L/m^2{\cdot}s} = \dfrac{(0.50\ \mathrm{L/m^2{\cdot}s})(86.4\ \mathrm{m^3/m^2{\cdot}d})}{(1.0\ \mathrm{L/m^2{\cdot}s})} = 43.2\ \mathrm{m^3/m^2{\cdot}d}$

The total hydraulic load with the recycle = $86.4\ \mathrm{m^3/m^2{\cdot}d}$ for use in Eq. (9–29)

1. Determine the NH_4-N concentration at the top of the tower due to influent dilution by the recycle using Eq. (9–30).

Assume effluent NH_4-N = $1.0\ \mathrm{g/m^3}$.

$$N_o = \frac{N_{sec} + RN}{1 + R} = \frac{(25.0\ \mathrm{g/m^3}) + (1.0)(1.0\ \mathrm{g/m^3})}{1.0 + 1.0} = 13.0\ \mathrm{g/m^3}$$

2. Solve for depth Z in Eq. (9–29), where N_o is 13.0 g/m^3 and N is 6.0 g/m^3.

$$(N_o - N) + K_N \ln\left(\frac{N_o}{N}\right) = \frac{ZaJ_{N,\max}}{q}$$

$$[(13.0 - 6.0)\ \text{g/m}^3] + (1.5\ \text{g/m}^3)\ln\left(\frac{13}{6}\right) = \frac{Z(138\ \text{m}^2/\text{m}^3)(1.8\ \text{g/m}^2\cdot\text{d})}{(86.4\ \text{m}^3/\text{m}^2\cdot\text{d})}$$

$Z = 2.8$ m

3. For the next 1 m of depth $J_{N,\max} = (1.8 - 0.1)\ \text{g/m}^2\cdot\text{d} = 1.7\ \text{g/m}^2\cdot\text{d}$

Solve for N using Eq. (9–29).

$$[(6.0 - N)\ \text{g/m}^3] + (1.5\ \text{g/m}^3)\ln\left(\frac{6}{N}\right) = \frac{1.0\ \text{m}(138\ \text{m}^2/\text{m}^3)(1.7\ \text{g/m}^2\cdot\text{d})}{(86.4\ \text{m}^3/\text{m}^2\cdot\text{d})}$$

$N = 3.9$, Note: this now yields the same solution at the greater depths as in (a) as the NH_4-N concentrations and total hydraulic load are the same.

4. The volume of packing media for each depth increment is the depth times the tower cross-section area.

$$\text{Area} = \frac{Q}{q}$$

$$\text{Area} = \frac{(6000\ \text{m}^3/\text{d})}{(43.2\ \text{m}^3/\text{m}^2\cdot\text{d})} = 138.8\ \text{m}^2$$

Plastic packing volume at 2.8 m depth = (A)(Z) = (138.8 m^2)(2.8 m) = 388.6 m^3

Plastic packing volume for 1.0 m depth increment = (138.8 m^2)(1.0 m) = 138.8 m^3

The results are summarized in the following table:

$J_{N,max}$, g/m^2·d	NH_4-N, g/m^3	Incremental depth, m	Volume, m^3
1.8	6.0	2.8	388.6
1.7	3.9	1.0	138.8
1.6	2.2	1.0	138.8
1.5	1.0	1.0	138.8
	Total	5.8	805.0

The plastic tower depth = 5.8 m (This is in the range of common tower heights used)

5. Determine the tower diameter.

Assume two towers in parallel.

Cross-section area per tower = 138.8 m^2/2 = 69.4 m^2

$$\text{Area} = A = \frac{\pi D^2}{4},\ \text{Diameter} = \sqrt{\frac{4A}{\pi}}$$

$$\text{Diameter} = \sqrt{\frac{4(69.4\ \text{m}^2)}{3.14}} = 9.4\ \text{m}$$

6. Summary and comparison of the plastic tower designs.

Design parameter	(a)	(b)
Recirculation ratio	0.0	1.0
Hydraulic load, $m^3/m^2 \cdot d$	86.4	43.2
Total hydraulic load, $m^3/m^2 \cdot d$	86.4	86.4
Number of towers	2	2
Tower diameter, m	6.64	9.40
Total tower depth, m	10.4	5.8
Plastic packing volume per tower, m^3	360.9	402.5

Comment The effect of using the 100 percent recirculation results in a shorter tower and larger tower diameter and an 11.5 percent increase in the total packing media volume required. Note that the system design with no recirculation would still have an effluent recirculation wet well and pump so that the tower design total hydraulic load can be maintained during periods of low influent flowrates.

9–3 SEQUENTIAL COMBINED TRICKLING FILTER AND SUSPENDED SOLIDS PROCESSES

Several treatment process combinations have been developed that couple trickling filters with suspended solids or activated sludge process. The three principal types of combined processes are the (1) trickling filter/solids contact (TF/SC) process, (2) trickling filter/activated sludge (TF/AS) process, and (3) series trickling filter/activated sludge process (Series TF/AS). The combined biological processes are commonly referred to as dual processes or coupled trickling filter/activated sludge systems. Topics considered in this section include (1) process applications, (2) the trickling filter/solids contact process, (3) the trickling filter/activated sludge process, and (4) the series trickling-filter activated sludge process.

Process Development

In addition to the TF/AS process described in the Sec. 9–1, another sequential combined trickling filter and activated sludge process is the *trickling filter/solids contact process*. The TF/SC process was developed in the late 1970s from research in Corvallis, OR aimed at a higher quality effluent after trickling filter treatment of domestic wastewater (Norris et al., 1982). In this case the trickling filter was designed to remove most of the soluble BOD, and was followed by an aerated solids contact channel also receiving return activated sludge from secondary clarifiers. The primary purpose of the aerated solids contact channel was to flocculate trickling filter effluent suspended solids in the activated sludge.

Process Applications

Sequential combined processes have often been the result of plant upgrading where either a trickling filter or suspended solids process is added, but they have also been provided in new facility designs (Parker et al., 1994). Combined processes are mainly applicable for BOD removal applications as the consumption of influent BOD in the trickling filter is a detriment to biological nutrient removal processes. The advantages of combined fixed film and suspended growth processes include (1) the stability and resistance to shock loads of the attached growth process, (2) the volumetric efficiency and low energy requirement of attached growth process for partial BOD removal, (3) the role of attached growth pretreatment as a biological selector to improve activated sludge settling characteristics, and (4) the high quality effluent possible with the downstream suspended growth or activated sludge treatment.

Trickling Filter/Solids Contact Process

The first full-scale field demonstration of a trickling filter/solids contact (TF/SC) process was at Corvallis, OR., where it was demonstrated that the combined process could provide advanced secondary treatment performance with effluent TSS and BOD concentrations below 10.0 mg/L (Norris et al., 1982).

Process Description. The process consists of a low-loaded trickling filter followed by a short-duration, aerated suspended growth solids contact tank prior to secondary clarification as shown on Figs. 9–14(a) and (b). The majority of the soluble BOD is removed in the trickling filter and the aerated solids contact tank is operated with a sufficient aerobic SRT for the production of biomass exocellular polymer to promote bioflocculation of dispersed colloids and particulates in the trickling filter effluent. Residual

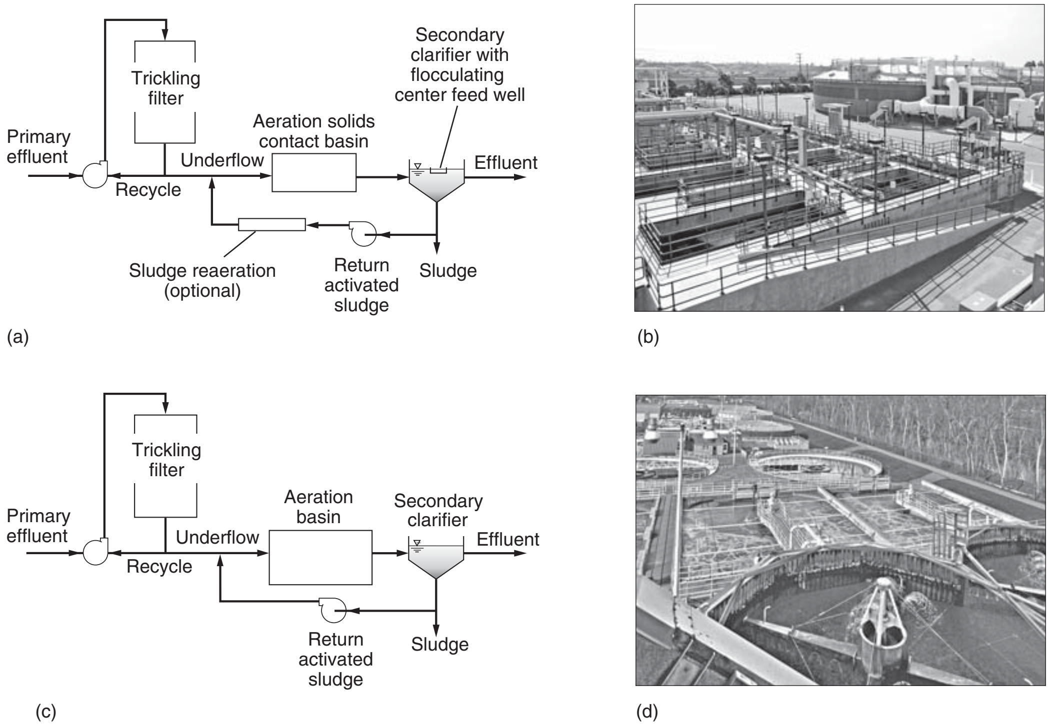


Figure 9–14

Combined trickling filter/activated sludge processes: (a) schematic flow diagram of trickling filter/solids contact (TF/SC) process with option to include return sludge reaeration, (b) view of solids contact channel with tower trickling filters in the background, (c) schematic flow diagram of trickling filter/activated sludge (TF/AS) process, and (d) view of trickling filter/solids contact installation taken from above the tower trickling filter.

sBOD in the trickling filter effluent is also removed in the solids contact tank. Newbry et al. (1988) found that with an SRT of 1.0 d or more for the aeration basin, the effluent sBOD concentration could be minimized.

Secondary Clarifier. The solids contact unit is generally followed by a secondary clarifier with a flocculating feed well to promote additional solids contact under gentle mixing conditions and bioflocculation of dispersed solids. The process is referred to as solids contact and bioflocculation treatment instead of an activated sludge step, because most of the BOD removal and biomass growth occurs in the trickling filter (Parker and Bratby, 2001).

Use of Sludge Recycle. About half of the TF/SC processes installed have used recycle sludge reaeration, as shown as an optional operating mode on Fig. 9–14(a). Sludge recycle is used to increase the sludge inventory and SRT without an increase in the solids contact detention time. A higher solids inventory is preferred for applications with higher trickling filter BOD loadings and to help attenuate the effects of occasions with increased amounts of trickling filter solids sloughing. For low-loaded trickling filter conditions (0.30 to 0.70 kg BOD/m^3·d), nitrifying bacteria may grow in the trickling filter and seed the solids contact unit to result in significant nitrification at a lower than expected aerobic SRT.

Design Considerations. Key design considerations for the TF/SC process are summarized in Table 9–8. Solids contact hydraulic retention times used range from 30 to 120 min, with 60 min close to many designs. The aerobic SRT is based on the solids inventory in the solids contact and sludge aeration tanks. Parker and Bratby (2001) noted that an SRT above 1.0 to 1.2 d is needed to obtain effluent TSS concentrations below 10 mg/L. They also advocated higher trickling filter BOD loadings of up to at least 2.0 kg BOD/m^3·d, while still achieving low effluent TSS concentrations. The secondary clarifier area requirement was found to be limited by solids loading instead of surface overflow rates (SOR), with allowable peak values as high as 12.2 kg/m^2·h due to the good settling sludge resulting from the trickling filter biomass growth. The 90th percentile SVI values for results at four TF/SC WWTPs ranged from 110 to 130 mL/g (Parker and Bratby, 2001).

Process Advantages. The major advantages of the TF/SC process are: (1) it provides an optimal design for the trickling filter and suspended growth process with minimal energy requirements due to most of the BOD oxidation occurring in the trickling filter, and (2) it achieves low effluent BOD and suspended solids with a small suspended growth aeration tank (5 to 20 percent of that needed for activated sludge treatment) (Daigger and Boltz, 2011).

Table 9–8

Process design criteria for trickling filter-solids contact (TF/SC) process[a]

Parameter	Units	Range	Common
Trickling filter BOD loading	kg BOD/m^3·d	0.4–1.8	0.80
Solids contact average detention time	min	45–120	60
Peak flow solids contact detention time	min	15–30	30
Total solids contact/reaeration tank SRT	d	1.0–2.0	1.2
Solids contact DO concentration	mg/L	1.0–2.0	2
Solids contract MLSS concentration	mg/L	1500–3000	2000
Secondary clarifier average SOR[b]	m/h	0.9–1.7	1.3

[a] Adapted from Parker and Bratby (2001) and Daigger and Boltz (2011).

[b] SOR = surface overflow rate, m^3/m^2·h.

Trickling Filter/Activated Sludge Process

The trickling filter/activated sludge process (TF/AS), as shown on Fig. 9–14(c), is similar to the TF/SC described above with the exception that instead of a contact process, a conventional activated sludge process is used.

Process Description. In contrast to the TF/SC process, the trickling filter in a TF/AS process is designed at a much higher BOD loading to serve as a roughing filter (40 to 70 percent BOD removal) and an appreciable amount of soluble BOD removal also occurs in the activated sludge process. The activated sludge SRT is designed to achieve effluent goals for BOD removal and possibly nitrification. Little reduction in the needed aeration basin size occurs as compared to that without the trickling filter because the overall solids production is only slightly less than what would have occurred without the upstream trickling filter.

Design Considerations. Common design conditions used for TF/AS processes are summarized in Table 9–9. The SRT needed is a function of the mixed liquor temperature and effluent BOD or ammonia-N concentration needed. The trickling filter acts as a bioselector for good settling floc in the activated sludge process and sustained SVI values of 100 to 120 mL/g are possible to allow for a higher MLSS concentration in the activated sludge system compared to that without the trickling filters (Biesinger et al., 1980).

Oxygen Required. The amount of oxygen required in the activated sludge aeration tank depends on how much influent BOD remains after the trickling filter treatment, and the amount and endogenous oxygen demand of trickling filter biomass fed to the activated sludge tank. The net amount of biomass produced in the trickling filter is a function of the amount of BOD removed and the trickling filter biomass SRT.

Removal of BOD. The amount of BOD removed and metabolized in the trickling filter is difficult to predict. Both particulate and soluble BOD are removed by biomass in the trickling filter and current empirical design models (Sec. 9–2) are generally based on influent and final settled BOD, and thus do not distinguish between particulate (pBOD) and soluble BOD (sBOD) removal rates. These models may be used to estimate the sBOD removal by subtracting the estimated BOD of the effluent suspended solids. However, pBOD not degraded in the trickling filter will most likely be degraded in the activated sludge process, therefore affecting the oxygen demand. Thus, to determine the oxygen required for the activated sludge process, the amount of pBOD degraded in the trickling filter is critical.

Table 9–9

Process design criteria for trickling filter-activated sludge (TF/AS) process

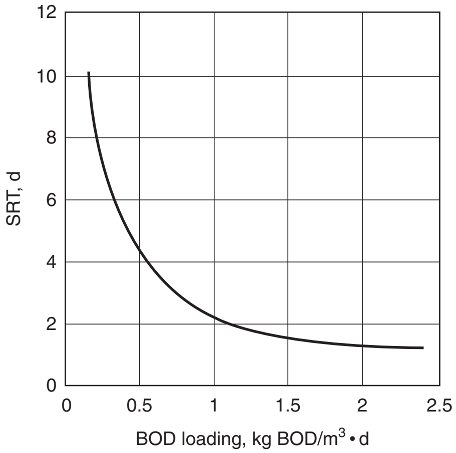
Parameter	Units	Range	Common
Trickling filter BOD loading	kg BOD/m^3·d	1.6–4.0	2.5
Activated sludge SRT	d	3.0–10.0	Varies[a]
Aeration tank DO concentration	mg/L	1.0–2.0	2
Aeration tank MLSS concentration	mg/L	2500–5000	3500
Secondary clarifier average SOR[b]	m/h	0.9–1.7	1.3

[a] Function of temperature, SRT selected to stay below nitrification.

[b] SOR = surface overflow rate, m^3/m^2·h.

Figure 9–15

Equivalent SRT for biomass in a trickling filter as a function of the BOD loading. (Adapted from WEF, 2000.)



Removal of pBOD. The removal of pBOD was studied in a combined trickling filter-activated sludge pilot plant over a wide range of trickling filter BOD loadings. Intensive sampling with COD and BOD solids balances on the trickling filter were used to determine the amount of pBOD degradation (Bogus, 1989). The amount of pBOD degraded increased as the BOD loading to the trickling filter was decreased. An estimate of the percent of influent pBOD degraded as a function of the BOD loading, is provided on Fig. 9–16.

Trickling Filter/Activated Sludge Design Approach. The design procedure to determine the oxygen requirements, sludge production and aeration volume of the activated sludge basin for a TF/AS process is summarized in Table 9–10. With an estimate of the trickling filter SRT and amount of pBOD and sBOD removal in the trickling filter, the amount of biomass produced can be calculated. With that information the amount of oxygen demand satisfied in the trickling filter can be estimated. The trickling filter effluent biomass and non-degraded pBOD and sBOD concentrations can then be used to estimate the activated sludge aeration basin oxygen demand. A solids balance is also done to determine the basin volume as a function of the design SRT and MLSS concentration. The biomass yield and endogenous decay coefficients and basic equations are the same as those

Figure 9–16

Approximate amount of particulate BOD degraded in a trickling filter as a function of organic loading. (From Bogus, 1989).

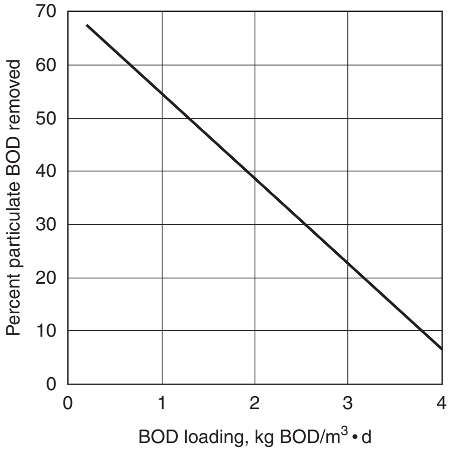


Table 9–10

Computation procedure for the design of a trickling filter/activated sludge process

Item	Description
1.	Select a BOD loading for the trickling filter that is compatible with the combined process selection.
2.	Select the SRT of the activated sludge process. For highly loaded trickling filters longer SRTs will be used in the activated sludge process.
3.	Determine the trickling filter tower size and hydraulic loading rate.
4.	Estimate the soluble BOD removal in the trickling filter.
5.	Based on the BOD loading to the trickling filter use Fig. 9–16 to determine the fraction of particulate BOD removed in the trickling filter.
6.	Use Fig. 9–15 to estimate the biomass SRT in the trickling filter. Use this value to estimate the biomass production.
7.	From the biomass production in the trickling filter and the amount of BOD removed, perform a mass balance on the ultimate BOD, including biomass production, to determine the amount of oxygen demand satisfied in the trickling filter.
8.	Assume that the portion of influent BOD not degraded in the trickling filter is degraded in the activated sludge basin if the SRT is ≥ 4 d. From the BOD removal calculate the biomass production in the activated sludge tank. Adjust the biomass produced from the trickling filter by the loss due to endogenous decay in the activated sludge tank.
9.	Based on the total biomass produced, perform an ultimate BOD mass balance to determine the total oxygen demand for the entire system. Subtract the amount of oxygen demand satisfied in the trickling filter to obtain the oxygen requirements for the activated sludge tank. Add a sufficient peaking factor to account for variable loadings.
10.	Sum the sludge production from biomass, nonbiodegradable VSS (nbVSS) in the influent wastewater and inorganic TSS (TSS–VSS) in the influent wastewater.
11.	Using the net sludge production, SRT, and assumed MLSS concentration calculate the volume of the activated sludge aeration tank.
12.	Evaluate solids and hydraulic loadings for the secondary clarifier design.

presented in Table 8–14 in Chap. 8. A TF/AS process design using the procedure outlined in Table 9–10 is illustrated in Example 9–6.

EXAMPLE 9–6 **Trickling Filter/Activated Sludge (TF/AS) Process Design** Determine the following design elements for a TF/AS process treating a high strength combined domestic and industrial food processing wastewater to meet an effluent BOD and TSS concentration of ≤ 20 mg/L. No nitrification is required:

a. Trickling filter diameter and hydraulic loading rate
b. The amount of oxygen demand satisfied in the trickling filter, kg/d
c. The amount of oxygen required in the activated sludge aeration tank, kg/d
d. The amount of solids wasted per day, kg/d
e. The volume and hydraulic retention time of the aeration tank
f. Compare the energy required for the trickling filter and activated sludge treatment steps in terms of kW/kg O_2 demand satisfied

Wastewater characteristics after primary treatment are given below for the domestic and industrial wastewaters and for their combined flow to the trickling filter.

Item	Unit	Domestic	Industrial	Combined
Flow	m^3/d	6000	1000	7000
BOD	g/m^3	130	600	197
sBOD	g/m^3	90	480	146
TSS	g/m^3	60	120	69
VSS	g/m^3	52	110	60
nbVSS	g/m^3	20	5	18

For TF/AS design assume the following conditions apply:

1. Number of trickling filters = 2
2. Domestic wastewater plastic packing coefficient $(k_{20}A_S) = 0.21\ (\text{L/s})^{0.5}/\text{m}^2$ (see Table 9–6)
3. Industrial wastewater plastic packing coefficient $(k_{20}A_S) = 0.181\ (\text{L/s})^{0.5}/\text{m}^2$ (see Table 9–6)
4. Depth of packing = 6.1 m
5. Biomass yield $Y_H = 0.6$ g VSS/g BOD
6. Endogenous decay $b_H = 0.08$ g/g·d
7. UBOD/BOD = 1.6
8. MLSS = 3500 g/m^3
9. Biomass VSS/TSS ratio = 0.85
10. Temperature = 15°C

Solution

1. Determine the weighted average trickling filter BOD removal coefficient $(k_{20}A_S)$ for the combined wastewater.

$$\text{Net } k_{20}A_S = \frac{(6000\ \text{m}^3/\text{d})(130\ \text{g/m}^3)(0.21(\text{L/s})^{0.5}/\text{m}^2) + (1000\ \text{m}^3/\text{d})(600\ \text{g/m}^3)[0.181(\text{L/s})^{0.5}/\text{m}^2]}{(7000\ \text{m}^3/\text{d})(197.1\ \text{g/m}^3)}$$

$$\text{Net } k_{20}A_S = 0.197(\text{L/s})^{0.5}/\text{m}^2$$

2. Select the organic loading rate for a partial BOD removal filter application from Table 9–1.

$$L_{\text{org}} = 2.5\ \text{kg BOD/m}^3\cdot\text{d}$$

3. Select the SRT of the activated sludge based on avoiding nitrification.

$$\text{SRT} = 5.0\ \text{d}$$

4. Determine the trickling filter tower size and hydraulic loading rate.
   a. Determine the trickling filter volume

   The organic volumetric loading rate, as given by Eq. (7–69), is

$$L_{\text{org}} = \frac{QS_o}{(V)(10^3\ \text{g/1 kg})} = 2.5\ \text{kg BOD/m}^3\cdot\text{d}$$

$$V = \frac{QS_o}{(2.5\ \text{kg BOD/m}^3\cdot\text{d})(10^3\ \text{g/1 kg})}$$

$$= \frac{(7000\ \text{m}^3/\text{d})(197.1\ \text{g/m}^3)}{(2.5\ \text{kg BOD/m}^3\cdot\text{d})(10^3\ \text{g/1 kg})} = 551.9\ \text{m}^3$$

b. Determine the hydraulic loading rate.

$$V = AD$$

$$A = \frac{V}{D} = \frac{551.9\ \text{m}^3}{6.1\ \text{m}} = 90.48\ \text{m}^2$$

$$q = \frac{Q}{A} = \left[\frac{(7000\ \text{m}^3/\text{d})}{(90.48\ \text{m}^2)}\right]\left(\frac{10^3\ \text{L}}{1\ \text{m}^3}\right)\left(\frac{\text{d}}{1440\ \text{min}}\right)\left(\frac{\text{min}}{60\ \text{s}}\right) = 0.90\ \text{L/m}^2\cdot\text{s}$$

c. Determine the trickling filter diameter.

$$\text{Area per tower} = \frac{(90.48\ \text{m}^2)}{2\ \text{towers}} = 45.24\ \text{m}^2/\text{tower}$$

Diameter/tower = 7.6 m

5. Determine the amount of soluble and particulate BOD removed in the trickling filter using Eq. (9–19). Note Eq. (9–19) is based on an effluent settled BOD. Assume a settled effluent TSS concentration of 30 mg/L.

a. Determine the effluent BOD concentration from Eq. (9–19) where $R = 0$, the effluent sBOD from Eq. (8–26) and the amount of sBOD removed in the trickling filter.

$$S_e = \frac{S_o}{(R+1)\exp\left\{\dfrac{k_{20}A_sD\theta^{T-20}}{[q(R+1)]^n}\right\} - R}$$

where S_o = influent BOD based on primary effluent flow, g/m^3
S_e = effluent BOD, g/m^3
k_{20} = filter treatability constant at 20°C, (L/s)$^{0.5}$/m
A_s = clean packing specific surface area, m^2/m^3
D = depth of packing, m
θ = temperature correction coefficient, 1.035
q = hydraulic loading rate based on primary effluent flow, L/m^2·s
R = ratio of recirculation flowrate to primary effluent flowrate, 0
n = constant characteristic of packing used = 0.50

$$S_e = \frac{(197.1\ \text{g BOD/m}^3)}{\exp\left\{\dfrac{(0.197(\text{L/s})^{0.5}/\text{m}^2)(6.1\ \text{m})1.035^{15-20}}{[(0.90\ \text{L/m}^2)\cdot\text{s}]^{0.5}}\right\}} = 67.8\ \text{g BOD/m}^3$$

Determine the soluble BOD from Eq. (8–26)

$$\text{BODe} = \text{sBODe} + \left(\frac{0.60\ \text{g BOD}}{\text{g UBOD}}\right)\left(\frac{1.42\ \text{g UBOD}}{\text{g VSS}}\right)\left(\frac{0.85\ \text{g VSS}}{\text{g TSS}}\right)\left(\frac{30\ \text{g TSS}}{\text{m}^3}\right)$$

$$67.8\ \text{g/m}^3 = \text{sBODe} + 21.7\ \text{g/m}^3,\ \text{sBODe} = 46.1\ \text{g/m}^3$$

$$\text{sBOD removal} = (145.7 - 46.1) = 99.6\ \text{g/m}^3$$

b. Determine the particulate BOD removed.

Particulate BOD in influent = BOD – sBOD

$$\text{pBOD} = (197.1 - 145.7) = 51.4\ \text{g/m}^3$$

From Fig. 9–16, amount of pBOD removed at 2.5 kg BOD/m^3·d = 30%

pBOD in trickling filter effluent = (1 – 0.30) (51.4 g/m^3) = 36.0 g/m^3

The pBOD removed in the trickling filter effluent is:

pBOD removed = 51.4 g/m^3 – 36.0 g/m^3= 14.4 g/m^3

c. Determine the undegraded influent BOD in the trickling filter effluent.

Trickling filter effluent BOD = sBOD + pBOD = (46.1 + 36.0) g/m^3

= 82.1 g/m^3

6. Determine oxygen demand satisfied in the trickling filter.

From Fig. 9–15, the trickling filter SRT = 1.2 d at an organic rate of 2.5 kg BOD/m^3·d

Determine biomass produced using Eq. (7–42) in Table 8–10. (Note: SRT = τ),:

$$X = \frac{Y_H(S_o - S)}{1 + b_H(\text{SRT})}$$

Substrate removal (S_o – S) in trickling filter

= (197.1 – 82.1) = 115.0 g BOD/m^3

$$X_{TF} = \frac{(0.6 \text{ g VSS/g BODr})(115 \text{ g BOD/m}^3)}{[1 + (0.08 \text{ g/g·d})(1.2 \text{ d})]} = 63.0 \text{ g VSS/m}^3$$

For short SRT values, cell debris is very small and is not included here.

Determine the oxygen satisfied in trickling filter with a COD balance.

$$\text{O}_2 \text{ used} = \text{bCOD}_{\text{IN}} - \text{bCOD}_{\text{OUT}} - 1.42X_{TF}$$

$$= 1.6(197.1 \text{ g/m}^3 - 82.1 \text{ g/m}^3) - 1.42(63.0 \text{ g/m}^3)$$

Oxygen used in trickling filter = 94.5 g/m^3

7. Determine biomass produced in the activated sludge aeration tank.

Approximate BOD removed = 82.1 g/m^3, SRT = 5.0 d

Biomass due to oxidation of organic matter

$$X_{AS} = \frac{(0.6 \text{ g VSS/g BODr})(82.1 \text{g BOD/m}^3)}{[1 + (0.08 \text{ g/g·d})(5.0 \text{ d})]} = 35.2 \text{ g VSS/m}^3$$

Trickling filter biomass remaining in the activated sludge aeration tank after endogenous decay:

$$X_{TF,AS} = \frac{(63.0 \text{ g VSS/m}^3)}{[1 + (0.08 \text{ g/g·d})(5.0 \text{ d})]} = 45.0 \text{ g VSS/m}^3$$

Net biomass produced or fed to the aeration tank = 35.2 g/m^3 + 45.0 g/m^3 = 80.2 g VSS/m^3

8. Determine oxygen demand in the aeration tank in mg/L and kg O_2/d.

Total oxygen consumed = 1.6(197.1 g/m^3) – 1.42(80.2 g/m^3)

= 201.5 g O_2/m^3

Activated-sludge oxygen demand = total demand − *TF* demand

$$= 201.5 \text{ g O}_2/\text{m}^3 - 94.5 \text{ g O}_2/\text{m}^3 = 107.0 \text{ g O}_2/\text{m}^3$$

$$\text{kg O}_2/\text{d} = (107.0 \text{ g O}_2/\text{m}^3)(7000 \text{ m}^3/\text{d})(1 \text{ kg}/10^3 \text{ g}) = 749 \text{ kg O}_2/\text{d}$$

9. Determine the sludge production rate in the activated sludge.

Amount of solids wasted per day (TSS) using Eq. (8–21) in Table 8–10.
(Cell debris term ignored)

$$P_{X,\text{TSS}} = \frac{P_{X,\text{Bio}}}{0.85} + Q(\text{nbVSS}) + Q(\text{TSS}_o - \text{VSS}_o)$$

$$P_{X,\text{TSS}} = \frac{(7000 \text{ m}^3/\text{d})(80.2 \text{ g VSS/m}^3)}{(0.85 \text{ g VSS/g TSS})} + (7000 \text{ m}^3/\text{d})(18.0 \text{ g VSS/m}^3)$$
$$+ (7000 \text{ m}^3/\text{d})[(69.0 - 60.0) \text{ g TSS/m}^3] = 849{,}470 \text{ g TSS/d}$$

The total solids wasted per day expressed in kg/d = 849.5 kg TSS/d

10. Determine the aeration tank volume and the corresponding hydraulic detention time
    a. Determine the aeration tank volume using Eq. (7–57) in Table 8–10.

$$(X_{\text{TSS}})(V) = P_{X,\text{TSS}}(\text{SRT})$$

$$V = \frac{P_{X,\text{TSS}}(\text{SRT})}{X_{\text{TSS}}} = \frac{(849.5 \text{ kg/d})(5.0 \text{ d})(10^3 \text{ g/1 kg})}{(3500 \text{ g/m}^3)} = 1213.6 \text{ m}^3$$

    b. Determine the hydraulic detention time in the solids contact aeration tank

$$\tau = \frac{(1213.6 \text{ m}^3)(24 \text{ h/d})}{(7000 \text{ m}^3/\text{d})} = 4.2 \text{ h}$$

11. Compare the energy needed per kg O_2 satisfied in the trickling filter versus in the activated sludge step.
    a. The energy required for the trickling filter is to pump the feed flow to above the 6.1 m tower height, which is approximately 1.38 kW/1000 m^3.

The energy needed equals (1.38 kW/km^3)(7000 m^3/d) = 9.66 kW

$$\text{Oxygen demand satisfied} = (94.5 \text{ g/m}^3)\left[\frac{(7000 \text{ m}^3/\text{d})}{(10^3 \text{ g/1 kg})}\right] = 661.5 \text{ kg/d} = 27.6 \text{ kg/h}$$

$$\text{O}_2 \text{ supply efficiency} = \frac{(27.6 \text{ kg/h})}{9.66 \text{ kW}} = 2.85 \text{ kg O}_2/\text{h}\cdot\text{kW}$$

O_2 supply efficiency in activated sludge aeration equals 0.7–1.5 kg O_2/h·kW (Table 5–31, Chap. 5)

Determine the net oxygen supply efficiency for the TF/AS process assuming the high-end efficiency of 1.5 kg O_2/h·kW for the AS process.

$$(0.47)(2.85 \text{ kg O}_2/\text{h}\cdot\text{kW}) + (0.53)(1.5 \text{ kg O}_2/\text{h}\cdot\text{kW}) = 2.13 \text{ kg O}_2/\text{h}\cdot\text{kW}$$

The net energy saving is

$$\left(1 - \frac{1.5}{2.13}\right)100 = 30\%$$

Comments Almost half of the total oxygen demand is satisfied in the trickling filter (94.5 g O_2/m^3 versus a total of 201.5 g O_2/m^3, or about 47 percent. The oxygen demand is provided with a higher oxygen transfer efficiency in the trickling filter than in the activated sludge step. Use of the TF/AS process for this higher strength wastewater reduces the aeration energy required by at least 30 percent compared to treatment with only activated sludge. In addition the activated sludge settling properties are better for the combined TF/AS process so that a higher MLSS concentration and smaller aeration tank volume can be used in the activated sludge process.

Process Advantages. The TF/AS process is attractive for treating higher strength industrial wastewater because of the relatively low energy use per quantity of BOD removed on the trickling filter. The use of the trickling filter also results in good SVI values for the activated sludge mixed liquor, as it acts as a biological selector in removing soluble BOD.

Series Trickling Filter and Activated Sludge Process

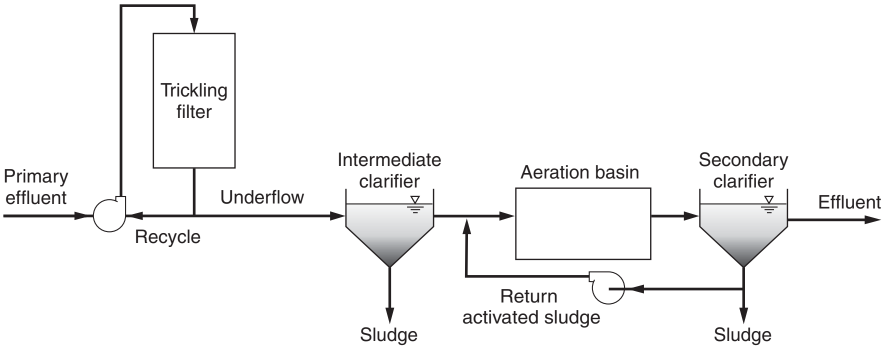
In the third approach employing combined processes, a trickling filter and activated sludge process are operated in series, with an intermediate clarifier between the trickling filter and activated sludge process (see Fig. 9–17). The combination of a trickling filter process followed by an activated sludge process is often used (1) to upgrade an existing activated-sludge system, (2) to reduce the strength of wastewater where industrial and domestic wastewater is treated in common treatment facilities, and (3) to protect a nitrification activated sludge process from toxic and inhibitory substances. In systems treating high-strength wastes, intermediate clarifiers are used between the trickling filters and the activated-sludge units to reduce the solids load to the activated sludge system and to minimize the aeration volume required.

9–4 INTEGRATED FIXED FILM ACTIVATED SLUDGE PROCESS

An *integrated fixed film activated sludge* (IFAS) process, or *hybrid* process, consists of an activated sludge system in which a material to support attached biomass growth has been added in addition to the suspended biomass growth in an activated sludge reactor. A variety of synthetic media for attached growth have been developed for applications in activated sludge processes. These materials may be suspended in the activated sludge or fixed in the aeration tank as shown on Fig. 9–18. The addition of attached growth to the activated

Figure 9–17

Schematic flow diagram of a series trickling filter and activated sludge process with intermediate clarifier.



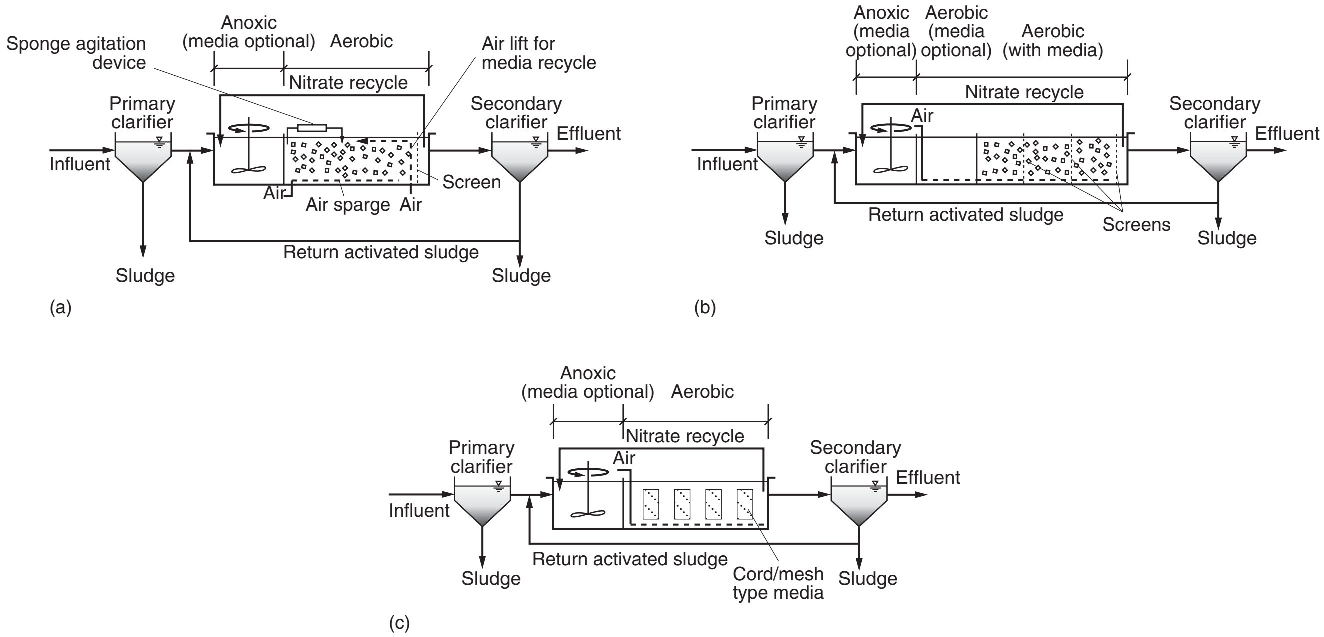


Figure 9–18

Schematic flow diagrams of examples of integrated fixed-film activated sludge (IFAS) systems: (a) suspended sponge biofilm carriers, (b) suspended plastic biofilm carriers, and (c) fixed media. (Adapted from WEF, 2011; and Phillips et al., 2010.)

sludge reactor results in a total equivalent MLSS concentration that may be 1.5 to 2.0 times the activated sludge MLSS concentration alone. The higher biomass inventory in the aeration tank can result in an increase in the effective SRT to provide nitrification or an increase in the system's treatment capacity. Because the additional solids are from attached growth, the higher MLSS inventory occurs without an increase in solids loading on the secondary clarifier. The following topics are considered in this section: (1) process development, (2) process applications, (3) process advantages and disadvantages, (4) design of physical facilities, (5) process analysis, and (6) BOD removal and nitrification design.

Process Development

The attached growth media in an IFAS process may be stationary or moving due to air sparging or mixing. Though most any type of media can provide an effective surface area for biofilm growth in an activated sludge process, the main applications of the IFAS process involve designs using polymeric, petroleum-based proprietary media. The suspended or *floating* media are called biofilm carriers and are typically either a sponge-type polyurethane or polyethylene foam or specially designed polypropylene plastic. The properties of representative types of biofilm carriers used in integrated fixed film activated sludge (IFAS) and moving bed bioreactor (MBBR) processes are summarized in Table 9–11.

Sponge-Type Biofilm Carriers. Linpor™ and Captor™ are IFAS processes using proprietary sponge-type media. The sponge-type biofilm carriers are foam cuboids with a specific gravity of about 0.95 g/cm^3 (see Fig. 9–19). The sponge biofilm carriers are placed in the activated sludge aeration tank with a filling fraction of 15 to 30 percent of the reactor volume. At a 20 percent fill volume with sponges having a bulk specific surface area of about 800 m^2/m^3, the biofilm surface area per unit reactor volume is about 160 m^2/m^3.

Table 9–11

Typical biofilm carriers used for IFAS and MBBR processes

Type of biofilm carrier[a]	Specific gravity	Example of nominal dimensions, mm	Example bulk specific surface area, m^2/m^3
Sponge[a]	0.95	15 × 15 × 12 depth	850
Plastic wheel (K1)[b]	0.96–0.98	7 × 10 dia.	500
Plastic wheel (K3)[b]	0.96–0.98	4 × 25 dia.	800
Plastic wheel (K5)[b]	0.96–0.98	9 × 25 dia.	500
BioChip (P)[b]	0.96–1.02	3 × 45 dia.	900
BioChip (M)[b]	0.96–1.02	2 × 48 dia.	1200
Plastic square	0.96	15 × 15 × 10 depth	680
Rope[c]		45 dia. rope	2.85 m^2/m

[a] From Linpor. See Fig. 9–19 for views of biofilm carriers.

[b] From Kaldness. See Fig. 9–20 for views of biofilm carriers.

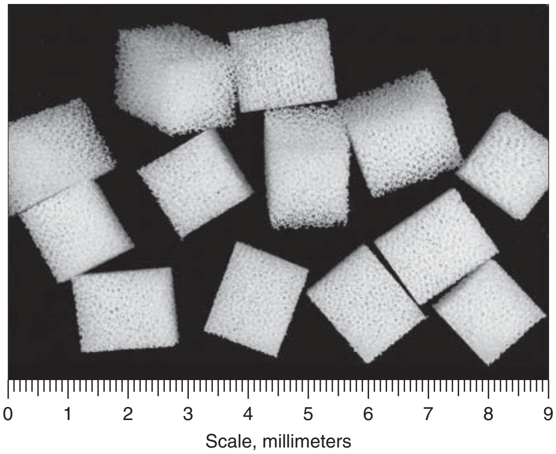
[c] From Ecologix.

As shown on Figs. 9–18(a) and (b), screens are required upstream and downstream of the IFAS reactor to retain the carriers. Coarse or fine bubble air sparging diffusers are used to provide oxygen and promote mixing of the sponges. The diffusers are arranged to promote a rolling mixing pattern. The effluent flow tends to pack the sponges up against the screens at the effluent, where they tend to float on the surface if not removed (WEF, 2011). An air-lift pump has been used to return sponges upstream and an air knife has been installed to clean the effluent screens continuously. A submersible pump is often installed to periodically pump the sponges to agitate and remove the accumulated excess biomass. An impingement plate at the end of the air lift media recirculation line has also been used to remove excess growth from the sponge surface (Warakomski, 2005).

Plastic-Type Biofilm Carriers. Plastic biofilm carriers with a specific gravity of 0.96 to 0.98 g/cm^3 were originally developed by a Norwegian company, Kaldnes Miljøteknolog, for use in the moving bed bioreactor process, but have since been used in IFAS systems with the first U.S. installation at Broomfield, CO. (Phillips et al., 2008). The original designs were a wagon-wheel configuration of about 10-mm diameter and

Figure 9–19

Sponge biofilm carriers used for integrated fixed film activated sludge and moving bed bioreactor processes. (Media samples provided courtesy of Mixing & Mass Transfer Technologies).



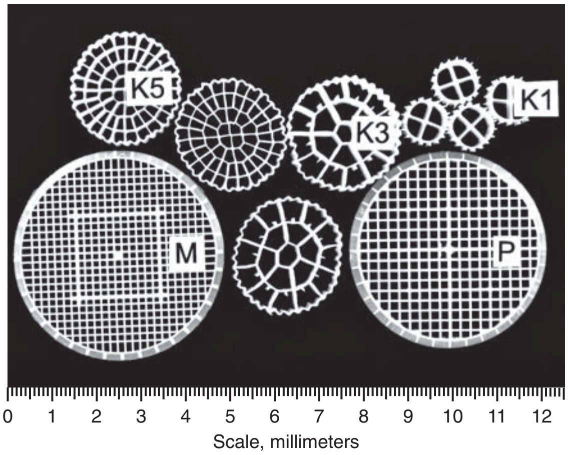


Figure 9–20

Examples of various plastic suspended biofilm carriers used in integrated fixed film activated sludge and moving bed bioreactor processes. The characteristics of the biofilm carriers are presented in Table 9–11. (AnoxKaldness™ Biofilm carriers provided courtesy of Veolia, Inc.)

7 mm height, but the 25 mm diameter design indicated in Table 9–11 is now more common (see Fig. 9–20). A number of companies provide plastic biofilm carrier media, and other shapes including chips and square configurations may be used. Plastic biofilm carriers used in aerobic zones have a bulk specific surface area of 500 to 700 m^2/m^3. With the plastic type carriers, the media fill volume can be higher than that used for sponge media, as they have less of a tendency to bind the downstream screens, with fill fractions ranging from 30 to 60 percent of the tank volume. Coarse or intermediate bubble size diffusers provide mixing in aeration zones and the plastic biofilm carriers are maintained in the reactor by the use of a perforated plate (5 x 25 mm slots) at the tank outlet.

Fixed Material. Some of the early IFAS processes used plastic sheets or asbestos sheets in activated sludge tanks. A Submerged RBC (SRBC) IFAS system has been created by placing standard RBC units in activated sludge with approximately 85 percent submergence instead of the typical operation at 40 percent submergence. The rotation is driven by aeration and may be mechanically assisted. The submerged operation reduces the load on the packing shaft. The more common IFAS process design with fixed material use a rope- or mesh-type media.

Examples of proprietary rope-like material used in IFAS processes are the Ringlace® and BioMatrix® and the mesh-type include the AccuWeb® and BioWeb® designs. The Ringlace and Biomatrix products consist of approximately 5 mm diameter strands of polyvinyl chloride material that are looped around aluminum or stainless-steel frames. The web media, which resembles netting woven in a hexagonal pattern, is also installed as sheets of media with the top and bottom of the media supported by the bars in the frames. An example of a web-mesh fixed material installation is shown on Fig. 9–21. The use of dedicated coarse-bubble sparging underneath the media is often used to ensure good mixing and oxygen transfer and media agitation to control the biofilm thickness and minimize the potential for red worm predators.

Process Applications

A common IFAS process application has been in the conversion of conventional aerobic activated sludge treatment processes to biological nitrification-denitrification processes for nitrogen removal, and especially where space is limited for adding additional tank volume. As an example, the addition of biofilm carriers to the activated sludge aeration tank at the Bloomfield, CO installation allowed a reduction of aerobic volume needed for nitrification,

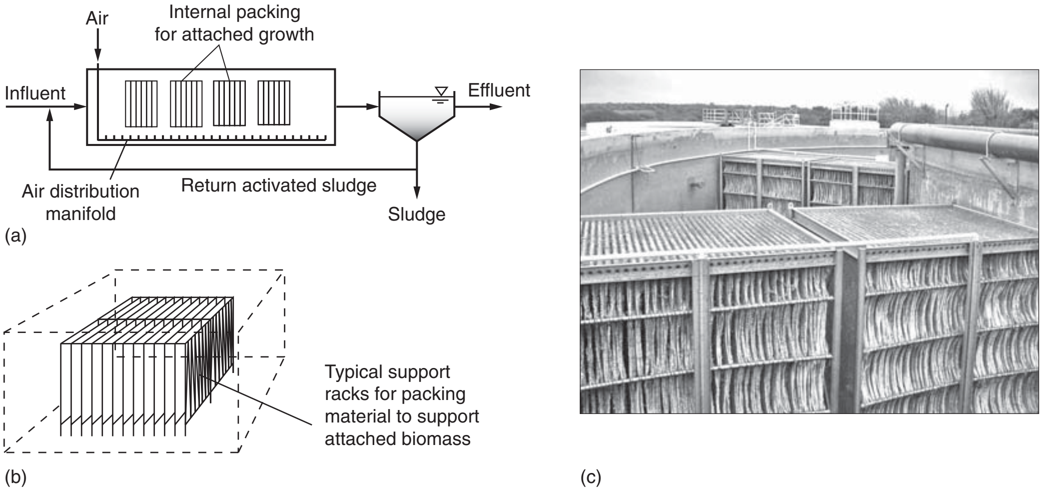


Figure 9–21

Example of a fixed attached growth media: (a) schematic showing the placement in activated sludge reactor (b) isometric view, and (c) view of installation in a circular configuration activated sludge tank (courtesy of J. Barnard).

so that a portion of the existing aeration tank could then be used for a preanoxic tank to convert the system to an MLE nitrogen removal process (Phillips et al., 2008).

Fixed or floating media can be applied in a number of activated sludge configurations to increase the effective volumetric biomass concentration and to reduce the suspended growth SRT needed for nitrification. When conventional activated sludge processes must be upgraded to a higher level of treatment for nitrification, nitrogen removal or phosphorus removal, an IFAS process may be considered as an alternative to provide the higher level of treatment without additional tank construction. Treatment capacity is added without the need to add new tankage or increase the secondary clarifier area. IFAS also offers flexibility in design in terms of the location and amount of biofilm media to be added to an existing or new system.

Nitrification and Biological Nitrogen Removal. The use of IFAS processes for nitrification and biological nitrogen removal is of primary interest due to the potential for upgrading BOD removal only facilities without construction of additional treatment tanks to accommodate longer SRTs and hydraulic retention times. IFAS configurations may include single or multiple-staged aerobic zones, and require higher DO concentrations in the aerobic zones, in the range of 4 to 6 mg/L, to overcome the biofilm substrate diffusion limitations so that the biomass on the media surface area can be utilized more fully for nitrification.

Flow schematics of possible IFAS configurations for nitrogen removal with single or multiple-stage aerobic zones are shown on Fig. 9–22. If a sufficient DO concentration is available, a two-stage aerobic zone provides higher nitrification efficiency due to a higher ammonia-N concentration and removal in the first stage. There are two advantages for using the attached growth process only in the intermediate stage of a three-stage aerobic zone as shown on Fig. 9–22(c). The first is more efficient use of the biofilm carrier surface area for nitrifying bacteria growth by having most of the heterotrophic growth for BOD removal in the first stage by the suspended biomass. The second advantage is that a lower DO concentration can be used in the smaller third aeration zone so that less dissolved oxygen is added to the preanoxic zone in the nitrate recycle flow.

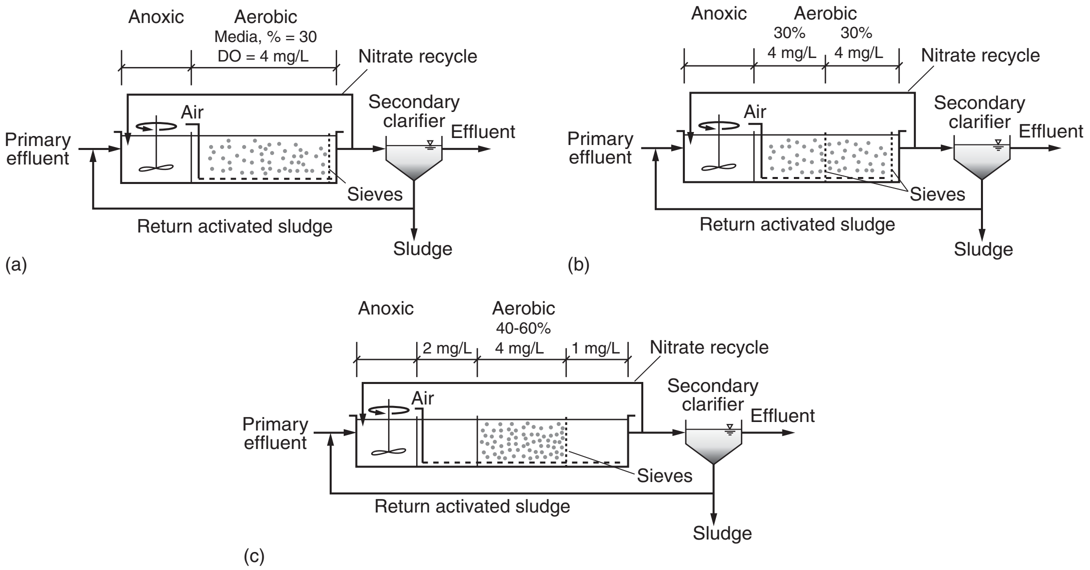


Figure 9–22

Examples of biofilm carrier media locations in the MLE biological nitrogen removal process: (a) single aeration reactor with carrier media, (b) two aeration reactors in series with carrier media, and (c) three aeration reactors in series with carrier media in larger, middle reactor. Typical media percentages and DO values are also shown each of the figures. (Adapted from Phillips et al., 2010.)

Enhanced Biological Phosphorus Removal. Enhanced biological phosphorus removal (EBPR) is possible in IFAS systems using EBPR process schemes (as described in Sec. 8–8) that have anaerobic contact zones with no media and only suspended growth (Christensson and Welander, 2004; Rogalla et al., 2006; and Pastorelli et al., 1999). In a pilot plant comparison of EBPR performance between an activated sludge system, an IFAS system with fixed BioWeb® media in the aerobic zone, and an IFAS system with fixed BioWeb® media in the anoxic and aerobic zones, similar performance was found for the activated sludge and IFAS system with media in the aerobic zone. The IFAS system with media in the anoxic and aerobic zone had slightly less EBPR efficiency, presumably due to anaerobic conditions within the biofilm in the anoxic zone causing phosphorus release (Sriwiriyarat and Randall, 2005). From an investigation of an IFAS system with suspended and attached growth autotrophic nitrifiers, heterotrophic PAOs, and denitrifier populations with plastic biofilm carriers in the aerobic zone, it was found that the slower-growing nitrifiers were mainly attached to the carrier media and the faster-growing heterotrophs were mainly in the suspended mixed liquor (Onnis-Hayden et al., 2011). The need for the PAOs to be exposed to anaerobic and aerobic conditions likely favored their growth in the shorter SRT suspended growth mixed liquor. Because of this decoupling ability, the advantage of operating at shorter SRTs based on suspended growth mixed liquor in an IFAS system, but having a longer SRT for nitrification is of importance.

IFAS Process Advantages and Disadvantages

Several advantages can be gained from an IFAS process compared to an activated sludge process including (1) an increased treatment capacity in a smaller space, (2) an increase in the effective mixed liquor concentration to as much as 4000 to 8000 mg/L without a significant increase in solids loading to the secondary clarifier, (3) the ability to phase in treatment capacity or improve performance by adding media, (4) a more stable nitrification community due to a high attached growth inventory, and (5) the potential for simultaneous nitrification and denitrification by controlling loading and DO conditions.

Process disadvantages compared to an activated sludge treatment process include (1) a higher energy demand due to the need to operate at elevated DO concentrations, (2) the need to use proprietary media, (3) issues of media removal for diffuser maintenance, (4) the need for improved influent wastewater screening, and (5) additional hydraulic profile headlosses due to flow through the media screening devices.

Design of Physical Facilities

Physical facility design considerations for an IFAS process include (1) pretreatment, (2) media retention, (3) aeration and mixing, (4) foam control, (5) biofilm control, and (6) liquid-solids separation.

Pretreatment. Appropriate pretreatment, including screening, grit removal, and primary sedimentation, is necessary to prevent the accumulation of inert material, such as rags, plastics, and sand on the media and in the tank. These materials would be difficult to remove once in the tank because of the presence of the fixed or suspended media. The maximum preferred bar screen spacing or perforation diameter is 6 mm with primary sedimentation and 3 mm without primary treatment (WEF, 2011).

Media Retention. Critical design elements that must be considered for the sponge or plastic biofilm carriers are effluent screens or sieves and the tank forward flow velocity (Phillips et al., 2010; McQuarrie and Boltz, 2011; and WEF, 2011) [see Figs. 9–18(a) and (b)]. Stainless steel effluent sieves are located at the end of the tank to collect the tank effluent flow and retain the media. The effluent sieve can be either a flat plate with orifices or horizontal cylinders, with the top of the cylinder at 35 to 65 percent of the liquid depth. Typical cylinder diameters are 0.30 to 0.40 m with lengths of 1.5 m, 3.0 m, and 3.65 m. The hydraulic application rate across the screen area is limited to 50 to 60 m/h at peak flow conditions including the recycle and secondary treatment influent flows. The sieve orifice size is selected to keep the headloss below 50–150 mm. An orifice flow velocity of less than 0.50 m/s is typically used to control headloss. Coarse bubble air diffusers located on the floor along the length of the flat plate screen provide air agitation to prevent the accumulation of debris and media and has been referred to as an *air knife*.

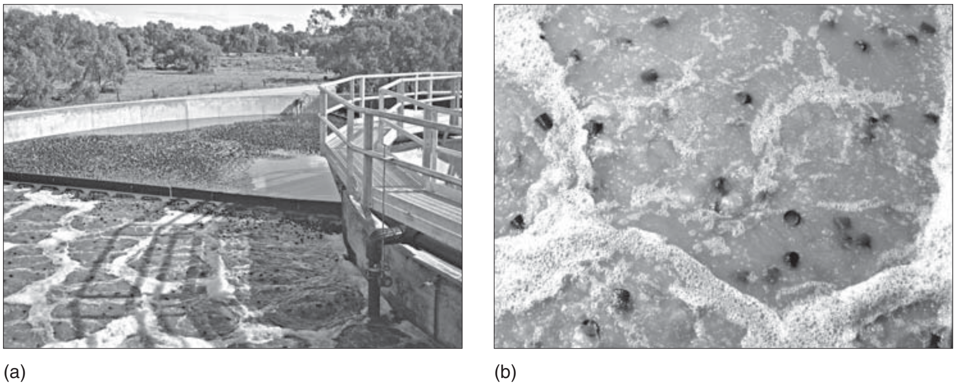
The tank forward flow velocity is the total flowrate out of the tank, including recycle flows, divided by the tank cross-section area. High forward flow velocities push more media to the effluent end of the tank, where it can bunch up and build up on the liquid surface. A forward flow velocity of less than 30 to 35 m/h is recommended (McQuarrie and Boltz, 2011; and WEF, 2011).

The fixed media is installed on frames that are placed at fixed locations in the IFAS aeration tank. The frames may be supported off the floor or supported by attaching structural members to the aeration tank walls. The frames must be sufficiently anchored to withstand the forward liquid flow velocity and the fluid agitation caused by aeration.

Aeration and Mixing. For floating media applications, sparged aeration is used to provide oxygen and to also mix the media. The sparging design is also used to provide agitation and fouling control of the effluent sieves (see Fig. 9–23). Coarse bubble aeration is most common for floating media applications but fine bubble aeration has also been used. The use of coarse bubble aeration avoids the need to move the media out of the basin

Figure 9–23

View of bioreactor with suspended biofilm carriers: (a) view of a bioreactor with a screen to retain the biofilm carriers and (b) suspend biofilm carriers in the air sparging zone.



for diffuser cleaning and maintenance as would be the case for fine bubble aeration, while fine bubble aeration is more energy efficient. In either case the diffuser aeration layout should be done to encourage a rolling action of the media across the basin width. Full floor coverage with fine bubble diffusers is discouraged. The IFAS media does not appear to have a significant impact on the sparged aeration oxygen transfer efficiency. In side-by-side testing of activated sludge and IFAS for a coarse bubble application (Rosso et al., 2011) and for a fine bubble application (Phillips et al., 2010), similar oxygen transfer efficiency was found at standard conditions.

In some fixed media designs, coarse bubble aeration facilities are placed below the fixed racks of media to provide mixing and agitation of the biofilm. In other designs, the objective is to create a flow pattern so that mixed liquor moves up and through the fixed media.

Mixing of floating media in anoxic zones is provided by slow speed mechanical mixers, which are generally placed near the top (but submerged) of the tank with a slight inclination to create a directional flow across the tank and rolling action. Typical mixing energy may range from 15 to 25 W/m^3 tank volume (McQuarrie and Boltz, 2011).

Scum and Foam Control. The use of effluent sieves for IFAS systems with suspended media results in trapping foam in the aeration basin. Trapping of foam in activated sludge basins normally encourages the growth of foam-causing bacteria to enhance the foaming problem. Serious foaming problems have not been reported in the early IFAS installations except for during start up operations. Provisions for chlorinated sprays or defoaming chemicals may be considered (Phillips et al., 2008).

Biofilm Control. Agitation and mixing caused by the sparged aeration system have been shown to be capable of controlling excessive biofilm growth in fixed media and plastic biofilm carriers. The lighter sponge media often needs additional agitation to dislodge excessive biofilm growth, which is done when media is circulated to the front of the aeration tank by an airlift or submersible pump. An impingement plate at the recirculation pipe discharge helps to dislodge the attached solids.

Predators. Blooms of a type of red-worm population have been observed with fixed media in IFAS systems. The occurrence of predators is less of a problem with the floating media, presumably due to the higher level of agitation. The worm growth results in a lower net sludge production, but can reduce the effective biomass and treatment performance. The worms are obligate aerobes and operating changes, such as a decrease in the influent load, that result in an increase in DO concentration may favor their growth. Turning the air off for several hours and chlorinating the return activated sludge is recommended for worm control. This method of control must be repeated to coincide with the worm egg cycle of approximately two weeks (WEF, 2011).

Liquid Solids Separation. In spite of the higher equivalent biomass concentration in the aeration tank, the solids loading to the secondary clarifier can remain within normal values used for activated sludge treatment. Based on limited studies it appears that the mixed liquor from an IFAS process may have a slightly higher SVI than from activated sludge treatment alone (Kim et al., 2009; Parker et al., 2011). The difference in SVI was attributed to more floc disruption from the flow through the effluent sieves and the high air agitation rates in the aeration basin (Parker et al., 2011).

IFAS Process Design Analysis

An IFAS process is typically used where fixed or biofilm carrier media is added to an activated sludge system to enable nitrification without the construction of additional aeration tanks. In such systems the suspended growth SRT is too short to support an adequate ammonia oxidizing bacteria (AOB) population for nitrification. Nitrifying bacteria can grow on the IFAS media to provide nitrification in spite of the limited SRT in the suspended growth process.

Examples of situations in which the benefits of an IFAS retrofit for nitrification have been realized are (1) the effluent treatment requirements have been changed from only BOD removal to include a low effluent ammonia-N concentration, (2) the system must be converted to nitrogen removal such that a portion of the existing aerobic tanks must be used for a preanoxic zone to thus lower the SRT to below that needed for nitrification, and (3) the flowrate and loading to the activated sludge process are increased and the increased sludge production and aeration tank MLSS concentration limits result in an SRT that is too low for nitrification.

Examples of IFAS retrofit applications to accomplish nitrification in activated sludge systems with limited SRT have been described for sponge or plastic biofilm carriers used at Westerly, RI (Masterson et al., 2004), Broomfield, CO (Phillips et al., 2008), the Region of Peel in Ontario, Canada (Stricker et al., 2009), and Conselve, Italy (Falletti and Conte, 2007), and for fixed media at Annapolis, MD (Randall and Sen, 1996) and Blacklick, OH (Sen et al., 2006). In most cases the media is added in the last two-thirds or half of the aeration basin after most of the soluble BOD has been removed by the suspended growth bacteria, so that the media surface area is available for growth by nitrifying bacteria and not overwhelmed by growth of heterotrophic bacteria.

Substrate Removal. The removal of substrate in an IFAS system is a complex process involving both substrate uptake by the suspended biomass and diffusion and consumption of substrate in the biofilm. The sloughing of nitrifying bacteria from the attached growth biofilm results in nitrification in the suspended mixed liquor at low SRTs for which nitrification would not normally be maintained. As was shown in Sec. 9–2 for attached growth nitrification in trickling filters, substrate removal across the biofilm surface area is defined in terms of the removal flux as g substrate/m^2·d. The actual effective surface area is less than the clean surface area of the media due to the biofilm thickness. Excess biofilm thickness growth is controlled in an IFAS system by the aeration designs, which causes hydraulic shearing and turbulent motion of the biofilm floating carriers or agitation of the biofilm in the fixed media designs. Reported biofilm surface concentration for attached growth on plastic carriers ranges from 13 to 39 g TSS/m^2 media surface for BOD removal zones and 7 to 13 g TSS/m^2 for nitrification only zones.

Table 9–12

Representative design conditions for IFAS systems with nitrification[a]

Type of media	Media fill fraction, %	Applied specific surface area, m^2/m^3 tank volume	MLSS concentration, mg/L	Aerobic hydraulic retention time, h
Fixed rope or web	70–80	50–100	3000	5.0
Sponge carrier	20–30	100–150	2500	4.0
Plastic carrier	30–60	150–300	2500	4.0

[a] Adapted from WEF (2011).

Substrate Removal Flux. The substrate removal flux is affected by the diffusion rate of substrate into the biofilm, the biofilm thickness, the external fluid mixing intensity and thickness of the biofilm stagnant layer, biokinetics and biomass density within the biofilm, and the availability of the electron acceptor. At a bulk DO concentration of 2.0 mg/L, the BOD removal flux is not normally impeded by the diffusion rate of the electron acceptor. However, for nitrification a DO concentration above 2 mg/L is normally needed to realize the benefit of an IFAS system for nitrification. Data on substrate removal flux has been gained from pilot- and full-scale plant evaluations.

IFAS Process Design Parameters. Typical process design conditions for different types of IFAS media are summarized in Table 9–12. Key process design parameters are (1) the substrate removal flux, (2) the media specific surface area (m^2/m^3), (3) the amount of media added to the activated sludge tank, which is also expressed in terms of the tank media bulk volume fill fraction or percent fill volume, (4) the aerobic tank DO concentration, and (5) the suspended growth MLSS or biomass concentration. The activated sludge tank is normally divided into stages to prevent influent short circuiting and to take advantage of the biokinetic benefits of staging by having higher soluble substrate concentrations in the upstream stages versus that needed in the final effluent.

For a given removal rate required and possible removal flux by the attached growth in an IFAS treatment zone, the required media surface area can be determined as shown by the following for ammonia-N nitrification.

$$A_{BF} = \frac{Q(N_o - N_e)}{J_N} \tag{9–31}$$

where A_{BF} = biofilm surface area, m^2
N_o = influent NH_4-N concentration to treatment zone, g/m^3
N_e = effluent NH_4-N concentration from treatment zone, g/m^3
J_N = NH_4-N removal flux, $g/m^2 \cdot d$

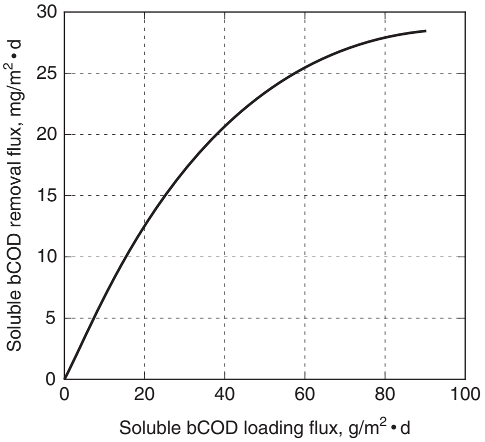
The media fill fraction can then be determined based on the specific surface area of the media and treatment zone reactor volume.

$$\text{Fill fraction (\%)} = \frac{(100)A_{BF}}{V(SS_A)} \tag{9–32}$$

where SS_A = media specific available surface area, m^2/m^3
V = reactor liquid volume, m^3

Figure 9–24

Soluble bCOD removal flux versus loading flux for plastic biofilm carriers. (Adapted from Ødgaard, 2006.)

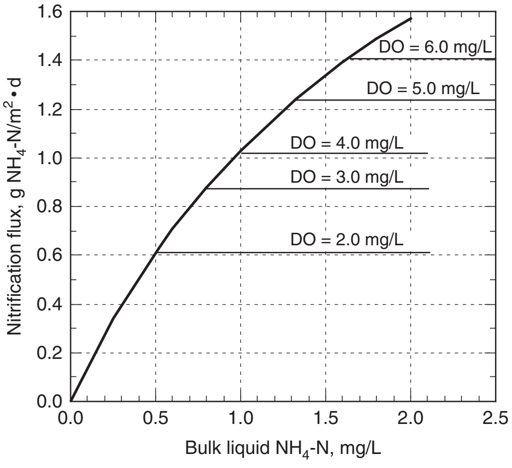


Biofilm Removal Flux. The soluble bCOD loading flux should not exceed 20 to 25 g/m^2·d with biofilm carrier media to achieve a secondary effluent BOD concentration (Ødegaard, 2006). A relationship between soluble bCOD removal flux and loading flux reported by Ødegaard (2006) is shown on Fig. 9–24. At a soluble bCOD loading flux of 10 g/m^2·d, the total BOD loading flux would be approximately 10.4 g/m^2·d, assuming 60 percent of the BOD after primary treatment is soluble. Assuming a media fill fraction of 50 percent and specific surface area of 500 m^2/m^3·d, the volumetric BOD loading would be 2.6 kg BOD/m^3·d. These loading values can be compared to a loading value of 0.30 to 0.70 kg BOD/m^3·d for secondary treatment with conventional activated sludge (see Table 8–19 in Chap. 8).

The nitrification ammonia-N removal flux is much slower than that for soluble bCOD and depends on the bulk liquid DO, ammonia-N, and soluble BOD concentrations. The nitrification ammonia-N removal flux is not affected by the soluble BOD concentration if it is less than or equal to 10 mg/L. A soluble BOD concentration of less than 10 mg/L can normally be achieved in an IFAS process with upstream separate stage BOD removal or with operation in a single tank at a low enough BOD loading. Observed ammonia-N removal flux as a function of bulk liquid DO and ammonia-N concentrations can be approximated by the process relationships presented by Ødegaard (2006) and summarized on Fig. 9–25. For a given DO concentration, the removal flux increases with increased ammonia-N concentration

Figure 9–25

Nitrification flux in biofilm as function of bulk liquid ammonia-N and DO concentration at low soluble BOD. (Adapted from Ødegaard, 2006.)



at an approximate first order relationship, until at some ammonia-N concentration the reaction is DO limited and the removal flux is then fixed by that DO concentration. Based on diffusion limitations and stoichiometric oxygen requirements for nitrification in Sec. 7–7 the theoretical transition point occurs at a ratio of 2.8 g O_2/g NH_4-N. However, the value for the relationship of flux versus NH_4-N on Fig. 9–25 is approximately 3.7. A value of 3.2 has been reported by Rusten et al. (2006). The higher value is related to biofilm substrate diffusivity and oxygen demand by carbonaceous substrates.

BOD and Nitrification Design Approach

The design of IFAS systems has generally been based on empirical approaches that incorporate performance data from pilot plant or full-scale facilities. The empirical models have used an equivalent MLSS concentration for the IFAS process based on media area and removal flux. Semi-empirical process kinetic-based models have also been used to analyze IFAS designs and to evaluate process performance under dynamic conditions. A third approach is the use of simulation software that models substrate diffusion and removal within the biofilm as well as the activated sludge suspended growth reactions.

Equivalent MLSS Approach. An example of the equivalent mixed liquor approach is that advanced by M^2T Technologies for sponge media (WEF, 2011). In this approach an equivalent MLSS is determined as follows:

$$\text{Equivalent MLSS} = (V_M)X_M + (V_{\text{AS}})(AS_{\text{MLSS}}) \tag{9–33}$$

$$V_T = V_M + V_{\text{AS}} \tag{9–34}$$

$$\text{Media fill fraction} = (V_M)/(V_T) \tag{9–35}$$

where V_M = media volume, m^3
X_M = solids concentration within the media volume, g TSS/m^3
V_{AS} = activated sludge volume, m^3
AS_{MLSS} = activated sludge MLSS concentration, g TSS/m^3
V_T = total tank volume, m^3

The activated sludge MLSS concentration can be determined from the design SRT, influent wastewater characteristics, and aeration tank volume using the procedures presented in Sec. 8–6. Assuming the IFAS supplier suggested value of 18,000 g/m^3 for the media solids, a suspended MLSS concentration of 3000 g/m^3, and a 25 percent fill volume, the aeration tank equivalent MLSS concentration is (18,000 g/m^3)(0.25) + (3000 g/m^3)(0.75) = 6750 g/m^3. Empirical relationships for specific removal rates of BOD and ammonia (i.e. g NH_4-N/g MLSS·d) are then used to calculate the amount of BOD and ammonia removal for the given aeration volume and media fill volume.

Empirical Design Approach Based on Media Flux and SRT. A range of removal flux values based on pilot plant or full-scale testing results are used with an empirical approach to estimate the media volume needed as a function of the activated sludge SRT (WEF, 2011). Flux removal values of 0.50 to 5.0 g COD/m^2·d and 0.05 to 0.50 g NH_4-N/m^2·d may be used at a temperature of 15°C and bulk liquid DO concentration of 3.0 mg/L. The fraction of COD and ammonia-N removed on the installed media is a function of the temperature and activated sludge SRT. The percent of total ammonia-N nitrified in the media biofilm is 80, 50, and 20 at suspended growth SRTs of 2, 4, and 8 d. For every °C increase the SRT is decreased by 3.0 percent. The corresponding percent COD removals on the media are 50 and 25 at the 2- and 4-d SRTs and no significant removal on the media at an 8.0-d SRT.

Semi-Empirical Process Kinetic-Based Approach. A semi-empirical process design approach for IFAS and submerged attached growth processes have been presented by Sen and Randall (2008a, 2008b, and 2008c) in a series of papers that describe the model developments, and application and verification. A similar simplified version is presented here for nitrification in the IFAS process. This kinetic approach includes the effect of nitrifying bacteria sloughing from the attached growth to seed or enhance nitrification in the suspended growth. The model provides a means to analyze the tradeoff between the amount of media installed in the activated sludge aeration tank and the operating SRT.

The process modeling approach involves mass balances on (1) ammonia oxidizing bacteria (AOB) on the biofilm media, (2) AOB in the bulk liquid, and (3) the ammonia-N concentration in the bulk liquid. The rate of growth from the nitrification flux, the rate of cell loss due to endogenous decay, and sloughing of the AOB from the biofilm to the bulk liquid are considered in the mass balance for AOB in the biofilm. An average biofilm SRT is used to account for the AOB residence time in the biofilm and sloughing rate. The mass balance is as follows:

$$(SS_A)V_m\frac{dX_{BF}}{dt} = Y_nJ_N(SS_A)V_m - b_nX_{BF}(SS_A)V_m - \frac{(SS_A)V_mX_{BF}}{\mathrm{SRT}_{BF}} \tag{9–36}$$

where, SS_A = bulk specific surface area of the media, $\mathrm{m^2/m^3}$
X_{BF} = AOB concentration in the biofilm, $\mathrm{g\ VSS/m^2}$
Y_n = AOB growth yield, g AOB VSS/g NH_4-N
J_N = nitrification flux to biofilm, g NH_4-N/$\mathrm{m^2 \cdot d}$
V_m = bulk volume of media added to the aeration tank, $\mathrm{m^3}$
b_n = AOB specific endogenous decay rate, g/g·d
SRT_{BF} = solids retention time of the AOB in the biofilm, d

At steady state conditions, X_{BF} is given by

$$X_{BF} = \frac{Y_nJ_N}{b_n + \left(\dfrac{1}{\mathrm{SRT}_{BF}}\right)} \tag{9–37}$$

As shown on Fig. 9–25, the nitrification flux is a function of the bulk liquid ammonia-N and DO concentrations and is described as follows:

$$J_N = \left(\frac{N}{k_{n,BF} + N}\right)J_{N,\max} \tag{9–38}$$

where, N = bulk liquid NH_4-N concentration, $\mathrm{g/m^3}$
$K_{n,BF}$ = biofilm half-velocity constant for nitrification, $\mathrm{g/m^3}$

Thus X_{BF} is determined as follows:

$$X_{BF} = \frac{Y_n\left(\dfrac{N}{k_{n,BF} + N}\right)J_{N,\max}}{b_n + \left(\dfrac{1}{\mathrm{SRT}_{BF}}\right)} \ or \ \frac{Y_nJ_{N,\mathrm{DOmax}}}{b_n + \left(\dfrac{1}{\mathrm{SRT}_{BF}}\right)} \tag{9–39}$$

For a given bulk liquid DO concentration, there is a bulk liquid ammonia-N concentration where the nitrification flux becomes DO limited and at that ammonia-N concentration, J_N equals $J_{N,\mathrm{DOmax}}$. The value for $J_{N,\mathrm{DOmax}}$ is indicated by the horizontal line on Fig. 9–25 for each bulk liquid DO concentration. If the bulk liquid ammonia-N concentration is below the inflection point for the respective DO concentrations shown on Fig. 9–25, the nitrification flux is ammonia limited and the J_N value is determined from Eq. (9–38). The $K_{n,\mathrm{BF}}$ value and $J_{N,\,\max}$ values for Fig. 9–25 are 2.2 mg/L and 3.3 g/m^2·d.

The suspended growth AOB concentration is the net result of the AOB sloughing rate from the biofilm, the growth rate from ammonia-N uptake from the bulk liquid, the rate of loss due to endogenous decay, and rate of solids wasting for SRT control. The mass balance assumes that suspended growth AOB activity does not occur in the bulk volume occupied by the IFAS media.

$$(V - V_m)\frac{dX_n}{dt} = \frac{(SS_A)V_m X_{BF}}{\mathrm{SRT}_{BF}} + Y_n r_N(V - V_m) - b_n X_n(V - V_m) - \frac{X_n(V - V_m)}{\mathrm{SRT}_{AS}} \tag{9–40}$$

$$(V - V_m) = \left(1 - \frac{V_m}{V}\right)V \tag{9–41}$$

where, V = total tank volume, m^3
X_n = suspended growth AOB concentration, g VSS/m^3
V_m/V = media fill fraction, m^3 media volume/m^3 tank volume
r_N = bulk liquid nitrification rate, g/m^3·d
SRT_{AS} = activated sludge SRT, d

Assuming steady state, substituting the relationship for r_N as given by Eq. (7–101) in Table 8–10, dividing by V, and rearranging yields

$$X_n = \frac{SS_A\left(\frac{V_m}{V}\right)\left(\frac{X_{BF}}{\mathrm{SRT}_{BF}}\right)}{b_n\left(1 - \frac{V_m}{V}\right) + \frac{1}{\mathrm{SRT}_{AS}}\left(1 - \frac{V_m}{V}\right) - \left(\frac{\mu_{\max}N}{K_n + N}\right)\left(\frac{\mathrm{DO}}{K_o + \mathrm{DO}}\right)\left(1 - \frac{V_m}{V}\right)} \tag{9–42}$$

The mass balance for the bulk liquid ammonia-N concentration accounts for the rate of ammonia-N added in the tank influent, rate of ammonia-N leaving in the tank effluent, the rate of ammonia-N consumption by the suspended growth AOB, the rate of ammonia-N removed from the bulk liquid by the biofilm nitrification flux, and the rate of ammonia-N released due to endogenous decay of the heterotrophic bacteria (due to the relatively small concentration of AOB, the ammonia-N released from AOB is ignored to simplify the solution).

$$(V - V_m)\frac{dN}{dt} = Q(N_o) - Q(N) - r_N(V - V_m) - J_N(SS_A)V_m + 0.12b_{\mathrm{H}}X_{\mathrm{H}}(V - V_m) \tag{9–43}$$

where N_o = influent available NH$_4$-N concentration, g/m^3
Q = influent flowrate, m^3/d
b_{H} = heterotrophic bacteria specific endogenous decay rate, g/g·d
X_{H} = heterotrophic bacteria concentration, g VSS/m^3
0.12 = g N released/g bacteria VSS loss by endogenous decay

Assuming steady state, substituting for J_N and r_N, and rearranging results in the following:

$$\begin{aligned} N_o = N &+ \left[\frac{(\mu_{\max}/Y_n)N}{K_n + N}\right]\left(\frac{\mathrm{DO}}{K_o + \mathrm{DO}}\right)X_n(\tau) + \left(\frac{N}{K_{n,BF} + N}\right)J_{N,\max}\left(\frac{V_m}{V}\right)(SS_A)(\tau) \\ &- 0.12b_{\mathrm{H}}X_{\mathrm{H}}\left(1 - \frac{V_m}{V}\right)(\tau) \end{aligned} \tag{9–44}$$

Again, if the value for N is above the ammonia-N concentration at the inflection point for a given bulk liquid DO concentration on Fig. (9–25) the value for $J_{N,\text{DOmax}}$ is substituted for the J_N relationship in Eqs. (9–43) and (9–44) reduces to:

$$N_o = N + \left[\frac{(\mu_{\max}/Y_n)N}{K_n + N}\right]\left(\frac{\text{DO}}{K_o + \text{DO}}\right)X_n(\tau) + J_{N,\,\text{DOmax}}\left(\frac{V_m}{V}\right)(SS_A)(\tau) - 0.12b_{\text{H}}X_{\text{H}}\left(1 - \frac{V_m}{V}\right)(\tau) \tag{9–45}$$

The heterotrophic biomass concentration is determined from Eq. (7–42) in Table 8–10, where $S_o - S \simeq S_o$.

$$X_{\text{H}} = \frac{Y_{\text{H}}(S_o)(\text{SRT}_{AS})}{[1 + b_{\text{H}}(\text{SRT}_{AS})]\left(1 - \frac{V_m}{V}\right)\tau} \tag{9–46}$$

The ammonia-N concentration available in the influent to the reactor is equal to the rate of bioavailable TKN added to the reactor minus the amount of nitrogen removed via the waste sludge production.

$$N_o = \text{TKN} - 0.12\left[\frac{X_{\text{H}}(\tau)\left(1 - \frac{V_m}{V}\right)}{\text{SRT}_{AS}}\right] \tag{9–47}$$

The solution for an IFAS design using the above relationships is illustrated in Example 9–7.

EXAMPLE 9–7 **Effect of Media Fill Volume Fraction in IFAS Process** Determine the aeration tank media fill volume fraction needed to provide nitrification to an effluent NH_4-N concentration below 0.70 g/m^3 for an existing activated sludge process following an increase in the average primary effluent flowrate from 20,000 to 30,000 m^3/d. The MLSS must be maintained at 3000 g/m^3 due to clarifier constraints. Provide a graph of effluent NH_4-N concentration and fraction of the IFAS nitrification occurring in the attached growth biofilm versus the media fill volume percent.

The aeration tank volume of the activated sludge system is 6940 m^3 and the primary effluent characteristics are given in the table below. The SRT at the 20,000 m^3/d flowrate with a MLSS concentration of 3000 g/m^3 is determined to be 9.5 d using the activated sludge design procedure shown in Example Problem 8–3 in Sec. 8–6. At the higher influent flowrate and assuming a MLSS concentration of 3000 g/m^3, the SRT is 4.1 d, which is not sufficient to meet the nitrification goals.

Design parameter	Unit	Design condition for activated sludge nitrification	Design condition for IFAS process analysis
Average flowrate	m^3/d	20,000	30,000
BOD concentration	g/m^3	140	140
TKN concentration	g/m^3	35	35
Nonbiodegradable VSS	g/m^3	25	25

(*continued*)

(*Continued*)

Design parameter	Unit	Design condition for activated sludge nitrification	Design condition for IFAS process analysis
TSS	g/m^3	70	70
VSS	g/m^3	60	60
Minimum design temperature	°C	12.0	12.0
Suspended growth aerobic SRT	d	9.5	6.0^a
Aeration tank volume	m^3	6940	6940
Hydraulic detention time, τ	d	0.35	0.233
MLSS	g/m^3	3000	3000
Heterotrophic biomass VSS	g/m^3	1740	1860
X_H, VSS/MLSS	g/g	0.58	0.62
BOD loading	kg BOD/m^3·d	0.40	0.60
Effluent NH$_4$-N	g/m^3	≤ 0.7	

[a] Suspended growth SRT is for without any media addition.

Other assumptions are as follows:

1. Plastic biofilm carrier media is used with a specific surface are of 500 m^2/m^3
2. Aeration tank DO concentration = 4.0 mg/L
3. Critical NH$_4$-N concentration for zero order flux = 1.07 mg/L (Fig. 9–25)
4. Maximum biofilm nitrification flux at DO = 4.0 mg/L is 1.08 g/m^2·d (Fig. 9–25)
5. bCOD/BOD = 1.6
6. Heterotrophic bacteria and AOB kinetic coefficients are from Table 8–14

   $Y_H = 0.45$ g VSS/g bCOD, $b_{H,20} = 0.12$ g/g·d, $\theta = 1.04$ for temperature correction for b_H.

7. AOB kinetic coefficients are from Table 8–14

   $\mu_{max,20} = 0.90$ g/g·d, $b_{n,20} = 0.17$ g/g·d, $\theta = 1.72$ and 1.029 for temperature correction for μ_{max} and b_n, respectively.

   $K_n = 0.50$ mg/L, $K_o = 0.50$ mg/L, $Y_n = 0.15$ g VSS/g NH$_4$-N oxidized

8. $K_{n,BF} = 2.2$ mg/L and $J_{N,\,max} = 3.3$ g/m^2·d
9. Biofilm AOB SRT$_{BF}$ = 6.0 d
10. The first 1/3rd of the aeration tank has no media addition and the soluble BOD leaving that zone is less than 10.0 mg/L, and thus does not interfere with AOB growth on the biofilm carriers.
11. The maximum media fill volume percent allowed for the plastic biofilm carriers in an IFAS process is 60 percent.

Solution

1. Assume a media fill volume percent in 2/3 of tank = maximum at 60 percent. Average fill volume percent = 2/3(60) = 40 percent. $V_M/V = 0.40$
2. Determine the kinetic coefficients at 12°C.

   $b_{H,12} = (0.12)(1.04^{(12-20)}) = 0.088$ g/g·d

   $\mu_{max,12} = (0.90)(1.072^{(12-20)}) = 0.516$ g/g·d

   $b_{n,12} = (0.17)(1.029^{(12-20)}) = 0.135$ g/g·d

3. Determine the influent bCOD concentration.

$$\text{bCOD} = 1.6(140 \text{ g BOD/m}^3) = 224.0 \text{ g/m}^3$$

4. Determine the suspended growth SRT using Eq. (9–46) and $X_H = 1860 \text{ g/m}^3$ from the above table.

$$X_{\text{H}} = \frac{Y_{\text{H}}(S_o)(\text{SRT}_{AS})}{[1 + b_{\text{H}}(\text{SRT}_{AS})]\left(1 - \dfrac{V_m}{V}\right)\tau}$$

$$1860 \text{ gVSS/m}^3 = \frac{(0.45 \text{ g VSS/g bCOD})(224 \text{ g bCOD/m}^3)(\text{SRT}_{AS})}{[1 + (0.088 \text{ g/g·d})(\text{SRT}_{AS})](1 - 0.4)(0.233 \text{ d})}$$

$$\text{SRT}_{AS} = 3.34 \text{ d}$$

5. Determine N_o from Eq. (9–47).

$$N_o = \text{TKN} - 0.12\left[\frac{X_{\text{H}}(\tau)\left(1 - \dfrac{V_m}{V}\right)}{\text{SRT}_{AS}}\right]$$

$$= 35.0 \text{ g/m}^3 - 0.12\left[\frac{(1860 \text{ g/m}^3)(0.233 \text{ d})(1 - 0.40)}{3.34 \text{ d}}\right]$$

$$= 25.7 \text{ g/m}^3$$

6. Assume the bulk liquid N concentration is below the critical value and use the left side of Eq. (9–39) to calculate X_{BF}.

$$X_{BF} = \frac{Y_n\left(\dfrac{N}{k_{n,BF} + N}\right)J_{N,\max}}{b_n + \left(\dfrac{1}{\text{SRT}_{BF}}\right)}$$

$$= \frac{(0.15 \text{ g VSS/g } N_{\text{oxidized}})\left[\dfrac{N}{(2.2 \text{ g/m}^3) + N}\right](3.3 \text{ g N/m}^2\text{·d})}{(0.135 \text{ g/g·d}) + \left(\dfrac{1}{6.0 \text{ d}}\right)}$$

7. Calculate the bulk liquid AOB concentration from Eq. (9–42).

$$X_n = \frac{SS_A\left(\dfrac{V_m}{V}\right)\left(\dfrac{X_{BF}}{\text{SRT}_{BF}}\right)}{b_n\left(1 - \dfrac{V_m}{V}\right) + \dfrac{1}{\text{SRT}_{AS}}\left(1 - \dfrac{V_m}{V}\right) - \left(\dfrac{\mu_{\max} N}{K_n + N}\right)\left(\dfrac{\text{DO}}{K_o + \text{DO}}\right)\left(1 - \dfrac{V_m}{V}\right)}$$

$$= \frac{(500 \text{ m}^2/\text{m}^3)(0.40)(X_{BF}/6 \text{ d})}{(0.135 \text{ g/g·d})(1 - 0.40) + \left(\dfrac{1}{3.34 \text{ d}}\right)(1 - 0.40) - \left[\dfrac{(0.516 \text{ g/g·d})N(4.0 \text{ g/m}^3)(1 - 0.40)}{(0.50 \text{ g/m}^3 + N)(0.50 \text{ g/m}^3 + 4.0 \text{ g/m}^3)}\right]}$$

8. Put in known values for Eq. (9–44).

$$N_o = N + \left[\frac{(\mu_{\max}/Y_n)N}{K_n + N}\right]\left(\frac{\mathrm{DO}}{K_o + \mathrm{DO}}\right)X_n(\tau) + \left(\frac{N}{K_{n,BF} + N}\right)J_{N,\max}\left(\frac{V_m}{V}\right)(SS_A)(\tau)$$
$$- 0.12 b_{\mathrm{H}} X_{\mathrm{H}}\left(1 - \frac{V_m}{V}\right)(\tau)$$

$$25.7\ \mathrm{g/m^3} = N + \frac{[(0.516\ \mathrm{g/g\cdot d})/(0.15\ \mathrm{g\ VSS/g}\ N)](N)(4.0\ \mathrm{g/m^3})X_n(0.233\ \mathrm{d})}{(0.50\ \mathrm{g/m^3} + N)(0.50\ \mathrm{g/m^3} + 4.0\ \mathrm{g/m^3})}$$
$$+ \frac{(N)(3.3\ \mathrm{g\ N/m^2\cdot d})(0.40)(500\ \mathrm{m^2/m^3})(0.233\ \mathrm{d})}{(2.2\ \mathrm{g/m^3} + N)}$$
$$- 0.12(0.088\ \mathrm{g/g\cdot d})(1860\ \mathrm{g/m^3})(0.60)(0.233\ \mathrm{d})$$

9. For steps 6, 7, and 8, there are three unknowns (X_{BF}, X_n, and N) and three equations. These equations can be solved on an excel spreadsheet using a solver function. The result is

$$X_n = 18.0\ \mathrm{g/m^3}$$
$$X_{BF} = 0.20\ \mathrm{g/m^2}$$
$$N = 0.30\ \mathrm{g/m^3}$$

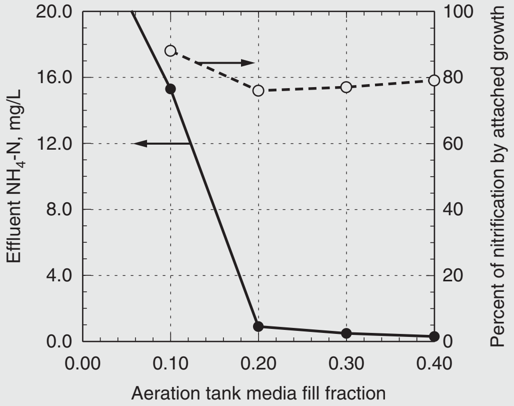
The fraction of the amount of NH_4-N nitrified that occurred in the biofilm = 0.79.

The bulk liquid N concentration is below the critical value of 1.07 g/m^3 so the appropriate equations were selected.

10. Similar calculations were done for V_M/V = 0.30, 0.20, and 0.10 and the results are summarized in the following table, followed by the graph required in the problem statement.

		V_m/V			
Parameter	Unit	0.10	0.20	0.30	0.40
SRT_{AS}	d	5.9	4.9	4.1	3.3
X_{BF}	g/m^2	0.54	0.47	0.29	0.20
X_n	g/m^3	6.6	15.5	17.4	18.0
N	g/m^3	15.3	0.89	0.48	0.30
Fraction of N oxidized in biofilm	—	0.88	0.76	0.77	0.79

Note: for V_m/V = 0.10, the critical NH_4-N value was exceeded and the biofilm nitrification flux was equal to $J_{N,\ DOmax}$.



Comment A fill volume fraction for the plastic biofilm carriers needs to be greater than 0.20 to meet the effluent NH_4-N concentration goal of $\leq 0.70\ g/m^3$. At a fill volume fraction of 0.30, the effluent NH_4-N is within the treatment goal. Because the media is added to only the last 2/3 of the aeration tank the fill volume fraction in the aeration tank with media is 0.45, which is acceptable as it is within the maximum value of 0.60 for plastic biofilm carriers in IFAS. The sloughing of nitrifying bacteria from the biofilm provided bioaugmentation of the suspended growth nitrification with 12 to 21 percent of the nitrification occurring in the suspended growth.

Mechanistic Modeling and Simulation Software. Simulation software is available for dynamic process modeling and analysis of attached growth systems, such as trickling filters, moving bed bioreactors, biological aerated filters and also for the combined attached growth/activated sludge IFAS systems. These are provided by most of the companies that provide activated sludge simulation software and also by others. A description of biofilm model development and types and complexity of current models is provided by Wanner et al. (2006).

In these models the biofilm is defined by several layers in which calculations are carried out for mass transport by diffusion and substrate utilization using the same biokinetic relationships as those used for suspended growth. The models provide calculations for substrate removal flux values for bCOD, ammonia-N, and oxidized-N for anoxic biofilms. The biofilm model is set up with a selected number of reactors in series. Key factors which must be selected by the model user and affect the prediction substrate removal rates within the biofilm are the biofilm thickness and biomass concentration. Assumptions of a thicker film can lead to predicting simultaneous nitrification-denitrification within the film due to the prediction of anoxic conditions in the deeper biofilm layers. Pilot plant or full-scale plant data, including bulk liquid DO concentrations and biofilm growth, can be used to calibrate the models to then analyze the effect of process configuration, media fill volumes, staging, DO concentration, temperature, and SRT for IFAS processes on system performance.

9–5 MOVING BED BIOFILM REACTOR (MBBR)

The MBBR process is similar to the IFAS process with mixed, suspended media contained within the reactor by effluent sieves, as described in Sec. 9–4, with the exception that there is no return activated sludge. The media fill volume fraction is generally higher (up to 70 percent), and the suspended solids concentration in the flow to the secondary clarifier may be in the range of from 100 to 250 mg/L versus 2500 to 3500 mg/L. Process designs for MBBR also include the suspended media in anoxic zones for fixed film biological denitrification. With MBBR reactor effluent, filtration processes, including granular media and membrane filtration, and dissolved air flotation can be used in lieu of gravity settling. Much of the information presented on the IFAS process with suspended media in Sec. 9–4 is applicable to the MBBR process, including the characteristics of the biofilm carrier media, media retention, aeration and mixing, and substrate flux into the biofilm. The topics considered in this section are (1) background, (2) process applications, (3) process advantages, (4) physical facilities, (5) process design analysis, and (5) nitrogen removal for MBBR systems.

Background

A variety of biofilm carrier media are possible for use in MBBR processes but much of the original research and most of the present installations involved the use of plastic biofilm carriers as described in Sec. 9–4 for IFAS applications and also indicated in Table 9–11. Much of the development work on MBBR processes with plastic biofilm carriers was done at the Norwegian University of Science and Technology (NTNU) in Trondheim, Norway with the aim to provide a compact biological nitrogen removal process to reduce point-source discharges of nitrogen to the North Sea. The first patents and commercialization of the MBBR technology was in 1989, through the Norwegian company, Kaldnes Miljøteknolog. Since then, a number of companies provide MBBR processes with proprietary floating plastic biofilm carrier designs, but most out of the hundreds of world-wide MBBR installations have used the Kaldnes media (WEF, 2011).

MBBR Process Applications

Flow schematics of examples of MMBR process configurations that can be used for BOD removal, biological nitrification and biological nitrogen removal are given in Table 9–13. All of the treatment schemes, with the exception of tertiary nitrification [see Table 9–13(d)] and tertiary biological denitrification after secondary treatment [see Table 9–13(h)] require raw wastewater pretreatment with influent screening, grit removal and primary treatment or fine mesh influent screening and grit removal. Chemical addition to the primary clarification step is an option for employing chemically enhanced primary treatment to reduce the BOD load to the MBBR process or as a first step for phosphorus removal. Metal salts and possibly polymer may be added before the MBBR process secondary clarifier for phosphorus removal and/or to improve the effluent suspended solids removal. A highly loaded, high rate BOD removal MBBR process [see Table 9–13(b)] can produce a more dispersed turbid effluent and would likely require chemicals to improve flocculation and effluent suspended solids removal.

Removal of BOD and Nitrification. For BOD removal and nitrification process designs, staged reactors may be used with the first stage primarily for removing most of the soluble BOD to minimize the heterotrophic bacteria growth competing with nitrifying bacteria for surface area on the biofilm carriers in the downstream nitrification reactors. The use of staged reactors [see Table 9–13(c)] for the nitrification zones improves the volumetric efficiency of the MBBR process due to having a higher ammonia-N concentration which can results in a greater nitrification flux in the stages before the final nitrification stage, provided that the nitrification reaction is not DO limited due to an insufficient bulk liquid DO concentration. Another potential MBBR process application for nitrification is tertiary nitrification after an activated sludge process for BOD removal.

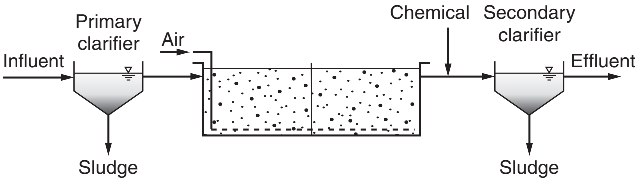
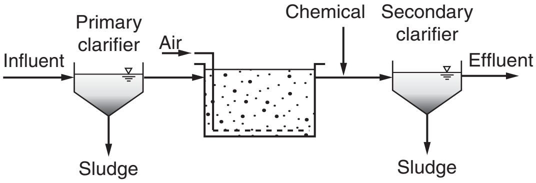
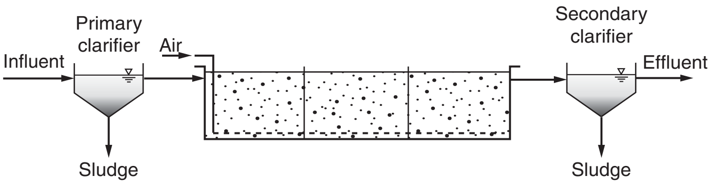
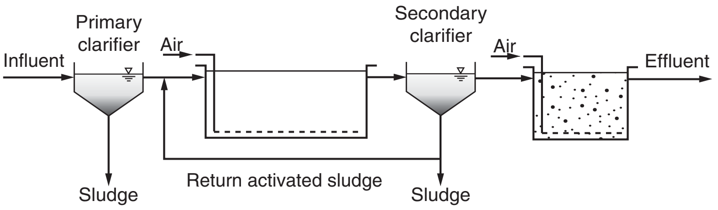
Removal of Nitrogen with Activated Sludge. The MBBR process can be designed with similar process configurations used with activated sludge for biological nitrogen removal. An MLE process configuration is shown in Table 9–13(e); it has a preanoxic zone at the front of the system for biological denitrification by biomass growth from influent BOD and consumption of nitrates in the nitrate recycle flow from the last aerobic reactor. A preanoxic and postanoxic configuration shown in Table 9–13(g) is similar to the Bardenpho process, which is used over the MLE process to minimize the effluent total nitrogen concentration by minimizing the effluent nitrate- and nitrite-N concentrations. An external carbon must be added to the postanoxic zones to promote bacteria growth and biological denitrification. An MBBR process can also be used for tertiary denitrification after BOD removal and nitrification in an activated sludge process as shown in Table 9–13(h).

MBBR Process Advantages and Disadvantages

MBBR processes are able to provide similar BOD removal and nitrogen removal treatment performance as activated sludge processes.

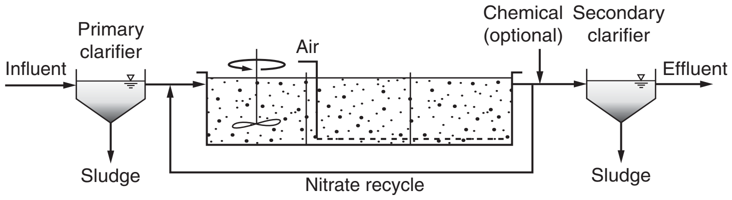
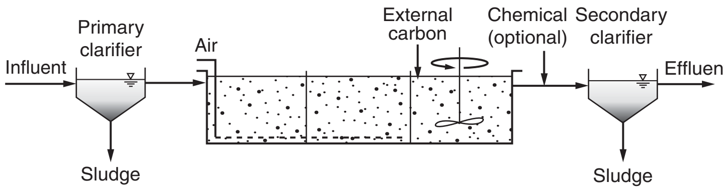
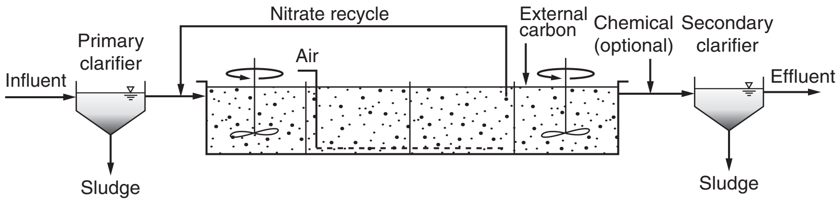
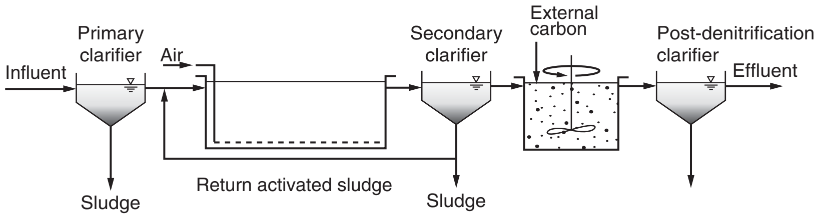
Table 9–13

MBBR systems configurations for different biological treatment goals

Process	Description
BOD removal	
(a) 	MBBR system for BOD removal and phosphorus removal by chemical precipitation. The aerobic tank may be a single or two-stage design.
(b) 	MBBR system for high rate BOD removal. Chemicals are added to provide sufficient suspended solids removal and also for phosphorus removal if needed.
BOD removal and nitrification	
(c) 	MBBR system for BOD removal and nitrification. Most of the BOD removal occurs in the first stage before subsequent single or multiple stage reactors for nitrification.
(d) 	MBBR system used as a tertiary nitrification process after secondary treatment for BOD removal.

(*continued*)

Table 9–13 (*Continued*)

Process	Description
Nitrogen removal	
(e) 	MBBR system used for MLE process with preanoxic zone and nitrate recirculation. Chemical addition before the secondary clarifier may be used for phosphorus removal (total nitrogen <10 mg/L).
(f) 	MBBR system used for postanoxic biological denitrification following BOD removal and nitrification. External carbon must be added to the postanoxic reactor (total nitrogen <3 mg/L).
(g) 	MBBR Bardenpho system for biological nitrogen removal with preanoxic and postanoxic reactors. External carbon must be added to the postanoxic reactor. Chemical addition may be used for effluent phosphorus removal and improved suspended solids removal (total nitrogen <3 mg/L).
(h) 	MBBR system for biological denitrification after activated sludge for BOD removal and nitrification and possibly EBPR. External carbon must be added (total nitrogen <3 mg/L).

Advantages. The major advantages of an MBBR process compared to activated sludge are (1) the small space needed, (2) the simplicity of operation with no need for manual sludge wasting and SRT control and sludge recycle, (3) the elimination of concerns of sludge bulking in the secondary clarifiers and its effects on operation and effluent quality, and (4) its ability to withstand peak wet weather flow variations. An MBBR process is well

Table 9–14

Example of hydraulic retention times used for full-scale MBBR process installations

Facility in Norway	Design flow, m^3/d	Process application	MBBR hydraulic retention time, h	Effluent total N, mg/L	Reference
Lillehammer	28,800	Bardenpho	3.2	4.5	Ødegaard (2006)
Nordre Folio	18,000	Bardenpho	4.9	8.0	Ødegaard (2006)
Gardermoen	22,100	Bardenpho	6.3	10.0	Ødegaard (2006)
Sjolunda	126,000	Post Anoxic	1.2	6.8	McQuarrie and Boltz (2011)
Klagshamm	23,800	Post Anoxic	1.1	5.8	McQuarrie and Boltz (2011)

suited for retrofit applications for activated sludge processes with reduced time and little if any tank construction. Because the MBBR process is basically an attached growth treatment process it deserves comparison to other attached growth processes such as trickling filters, rotating biological contactors, and biological aerated filters that are discussed in the following section. Compared to these processes, the MBBR process is much more versatile and adaptable for biological nitrogen removal and is a continuous operation that does not require special operational attention or interruption of treatment for biofilm thickness control or flushing out excess solids. Examples of MBBR full-scale plants in Table 9–14 show the relatively low hydraulic retention times used in MBBR processes.

Disadvantages. Process disadvantages compared to activated sludge treatment include (1) a higher energy demand due to the need to operate at elevated DO concentrations, (2) the need to use proprietary media, (3) issues of media removal for diffuser maintenance, (4) the need for improved influent wastewater screening, (5) additional hydraulic profile headlosses due to flow through the media screening devices, and (6) the limitation of phosphorus removal only by chemical addition.

Design of Physical Facilities

The physical facility design considerations discussed in Sec. 9–4 for an IFAS process with suspended biofilm carriers with regard to (1) pretreatment, (2) media retention and effluent sieves, (3) aeration and mixing, and (4) foam control, apply to the MBBR process. Coarse bubble or intermediate bubble aeration is generally used in MBBR aerobic tanks. Based on off gas testing of a system with plastic biofilm carriers, the SOTE values ranged from 11 to 15 percent, where SOTE is the oxygen transfer efficiency at standard conditions.

The amount of plastic biofilm carriers added is defined in terms of its bulk volume filling fraction or percent of the tank volume. The plastic media carrier is normally added to an empty reactor for the specified volumetric filling percentage. The coarse aeration pipes are designed to withstand the weight of carrier media after the tank is drained for maintenance. During maintenance, plastic carrier can be transferred to a carrier storage with a recessed impeller pump. Media replacement has not been a significant issue in MBBR processes, though the design of mixers must be done to prevent media abrasion. After 10 years of operation of the first MBBR plant installed in Norway in 1996, no observed biofilm carrier wear has been reported (Rusten et al., 2006).

Because of the relatively low suspended solids concentration in the effluent from the MBBR process (typically in the range of 100 to 250 mg/L), solids recycle is not required as it is for activated sludge and IFAS processes, and liquid-solids separation methods other than secondary clarification may be used. Flocculation and flotation, dual media sand filters, cloth disc filters, and ballasted flocculation have been used at full-scale facilities (McQuarrie and Boltz, 2011).

MBBR Process Design Analysis

The mechanisms and important factors for substrate removal by suspended biofilm carriers are discussed in Sec. 9–4 for the IFAS process. The major differences between the IFAS and MBBR processes are that the MBBR normally has a higher biofilm carrier density, with up to a 70 percent media fill volume, and it does not have return activated sludge and a significant suspended mixed liquor or biomass concentration. In the MBBR process, the substrate removal is primarily by the attached growth and does not have the benefit of activated sludge to enhance the flocculation and settling of solids that slough from the attached growth. As described in Sec. 9–4, the amount of substrate removal by biofilm carriers is a function of the available biofilm surface area and substrate flux. The biofilm surface area is the product of the reactor volume, media fill volume fraction, and the media bulk specific surface area. The substrate flux is a key process design parameter and is a function of the bulk liquid substrate and DO concentrations, the reactor mixing conditions, and the biofilm characteristics. Substrate flux relationships determined for BOD removal, nitrification, and nitrate reduction from the analysis of pilot plant and full-scale plant MBBR processes can be used to develop MBBR process designs. As described by Eq. (9–31) the required total surface area of biofilm carriers in an MBBR reactor is the rate of substrate removal (g/d) needed divided by the removal flux ($g/m^2{\cdot}d$).

Typical substrate removal flux values for BOD removal, nitrification, and denitrification are shown in Table 9–15. Note that the surface area loading rate (SALR) is higher than the substrate removal flux and can be estimated by dividing the removal flux by the estimated treatment efficiency. The equivalent volumetric removal rates are also shown for a 60 percent media fill volume. The secondary treatment volumetric BOD load for the MBBR secondary treatment SALR is about 1.7 to 5.0 times greater than a typical conventional activated sludge loading of 1.0 kg $BOD/m^3{\cdot}d$.

The ability of the MBBR process to handle much higher volumetric BOD loadings than a conventional activated sludge process can only be partially explained by differences in reactor biomass concentrations. The biofilm areal solids concentration may range from 12 g TSS/m^2 for a combined carbon and nitrification operation to as high as 28 g TSS/m^2 for a high rate BOD removal operation (McQuarrie and Boltz, 2011). Assuming a media specific

Table 9–15

Typical removal flux for BOD, nitrification and denitrification in MBBR processes[a]

Application	Substrate	Removal flux, $g/m^2{\cdot}d$	Volumetric removal rate,[b] $kg/m^3{\cdot}d$
BOD removal, partial	BOD	15–20	4.5–6.0
Secondary treatment	BOD	5–15	1.7–5.0
Prenitrification	BOD	4–5	1.2–1.5
Nitrification	NH_4-N	0.4–1.4	0.1–0.4
Predenitrification	NO_3-N	0.20–1.0	0.1–0.3
Postdenitification	NO_3-N	1.0–2.0	0.3–0.6

[a] Adapted from McQuarrie and Boltz (2011) and WEF (2011).

[b] Based on 60 percent media fill fraction.

surface area of 500 m^2/m^3 and a 60 percent fill volume, the corresponding volumetric TSS concentrations can vary from 3870 to 8400 mg/L. As the higher volumetric BOD removal rates in the MBBR cannot be related entirely to differences in reactor suspended solids, it has also been suggested that it is due to the use of highly compartmentalized designs and the presence of a higher viable biomass on the media (Ødegaard, 2006).

BOD Removal and Nitrification Design

Designs for BOD removal can be based on appropriate SALRs based on observed substrate removal flux information. The design for nitrification is more complex and requires special consideration to the amount of soluble BOD entering the MBBR nitrification reactor and the DO concentration in the MBBR reactor. To assure effective nitrification and to minimize the volume of biofilm carrier needed, operation at elevated DO concentrations, in the range of 4.0 to 6.0 mg/L, must be considered. In addition, the use of MBBR reactors in series for removal of soluble BOD in upstream reactors prior to nitrification is normally done. Substrate removal flux information for BOD removal and nitrification, operating conditions, and design considerations for MBBR systems for BOD removal only, BOD removal and nitrification, and tertiary nitrification are discussed below. Denitrification in MBBR reactors is discussed in Sec. 9–7. As discussed in Sec. 9–4, computer simulation software is also available for analysis of MBBR designs and can be calibrated to existing plant performance data.

BOD Removal. Three levels of BOD removal design loadings are considered: (1) partial or high rate SALR, (2) secondary treatment or normal SALR, and (3) prenitrification or low SALR. An MBBR process may be added before activated sludge treatment for partial BOD removal to increase the treatment capacity of an industrial or combined domestic/industrial wastewater. It may also be used in initial reactors for BOD removal prior to downstream nitrification reactors. Assuming 70 percent BOD removal, the BOD roughing SALR may be up to 28 g $BOD/m^2 \cdot d$. At this loading, depending on the wastewater BOD concentration, the hydraulic retention time, τ, could be less than 30 min. However, a τ of at least 45 to 60 min is used, as oxygen transfer issues can limit the maximum loading and minimal τ needed for partial BOD removal. A DO concentration of 2 to 3 mg/L has been considered adequate for BOD removal in MBBR designs (Ødegaard, 2006). At the high BOD SALRs the sloughed solids from the MBBR media are more dispersed and poor setting occurs. Thus, if the MBBR is designed as only a roughing BOD treatment, it is necessary to provide chemicals and flocculation before the secondary clarifier.

The BOD SALR for secondary treatment to produce an effluent BOD and TSS concentration of less than 25 mg/L is in the range of 6 to 16 $g/m^2 \cdot d$, based on the assumption of 90 percent BOD removal. Prenitrification designs use much lower BOD SALRs to minimize the MBBR reactor effluent soluble bCOD to downstream nitrification reactors, so that there is less competition by heterotrophic bacteria over nitrifying bacteria for growth on the media surface area. A higher nitrification ammonia-N flux and greater nitrification efficiency is then possible. At a soluble BOD concentration of less than 10 mg/L maximum use of the biofilm carrier surface area by ammonia-oxidizing bacteria is possible. If the soluble COD concentration of the wastewater is known, the relationship shown on Fig. 9–24 can be used to determine the amount of media and the reactor volume needed to reduce the soluble COD to a low level before nitrification (Ødegaard, 2006). The use of designs with multiple MBBR reactors in series is recommended to provide high rate BOD uptake, a low soluble BOD prior to nitrification, and more efficient ammonia removal to thus provide more process stability and to minimize reactor volume and media requirements. Another advantage of using reactors in series is to minimize the possibility of short circuiting of the influent flow.

Nitrification. The MBBR process has been used for nitrification in various process schemes (Table 9–13), including combined with BOD removal, in the later stages of an MBBR process with reactors in series after depletion of most of the soluble BOD, and for tertiary nitrification. Nitrification substrate flux design values that can be used to estimate the reactor volume and the volume of biofilm carriers needed for nitrification in different process modes are considered below. The factors that govern the nitrification substrate flux include the bulk liquid ammonia-N and DO concentrations, temperature, soluble BOD concentration which is related to the reactor BOD loads, and pH. As described in Sec. 8–6, an alkalinity balance should be done for each application to assure that a minimum alkalinity of 70 mg/L as $CaCO_3$ is present in the nitrification reactor so that a pH of at least 6.8 is present to allow uninhibited nitrification due to pH limitations. Higher nitrification flux occurs for nitrification reactor designs with a (1) lower organic loading rate, (2) minimal soluble BOD concentration, (3) higher bulk liquid DO concentration, (4) higher temperature, and (5) higher bulk liquid ammonia-N concentration. The bulk liquid ammonia-N concentration may be the rate-limiting factor at low concentrations (<1–3 mg/L), and thus the use of nitrification reactors in series is recommended to provide a more optimal reactor volume and media addition to achieve a low effluent ammonia concentration of less than 1.0 mg/L.

Tertiary Nitrification. Tertiary nitrification in an MBBR process involves growth of primarily AOB on the plastic carrier biofilm. The growth of AOB can occur when the MBBR nitrification reactor receives clarifier effluent following secondary treatment [see Table 9–13(d)] or after extensive BOD removal and a low organic loading in an upstream reactor before nitrification in a MBBR process with a number of reactors in series. The relationship shown previously on Fig. 9–25 can be used to estimate the nitrification flux as a function of the bulk liquid DO and ammonia-N concentration. The nitrification flux on Fig. 9–25 at ammonia-N concentrations for which the rate is not DO limited can be determined from the Eq. (9–48).

$$J_N = \left[\frac{N}{(2.2 \text{ g N/L}) + N}\right](3.3 \text{ g N/m}^2\cdot\text{d}) \tag{9–48}$$

The data on Fig. 9–25 was developed at 15°C. A temperature adjustment factor, according to Eq. 2–25 of 1.098 was found by Salvetti et al. (2006) for ammonia concentration limited nitrification flux and 1.058 for oxygen limited nitrification flux.

In MBBR reactors with combined BOD removal and nitrification, the heterotrophic bacteria compete with the AOB for dissolved oxygen and space on the biofilm which can hinder the nitrification flux. The effect of BOD SALR and DO concentration is shown on Fig. 9–26. As the BOD SALR increases to above 2.0 g/m²·d, a bulk liquid DO concentration of 4.0 to 6.0 mg/L is needed to provide an appreciable nitrification flux. An empirical equation, developed by Rusten et al. (1995), can be used to estimate the nitrification flux under ammonia rate limiting conditions in a combined BOD removal and nitrification process at 10°C.

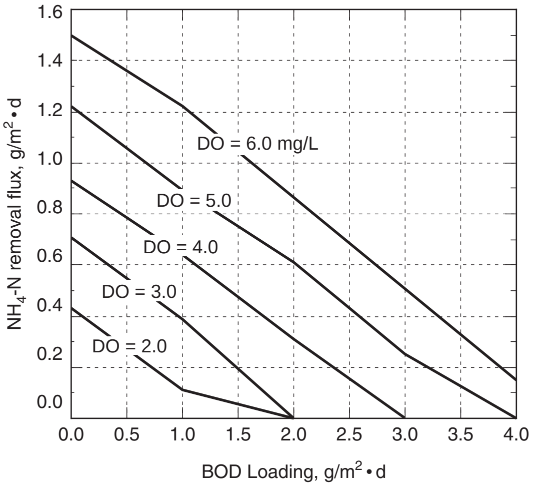
$$J_N = k_{nf}(N)^{n'} \tag{9–49}$$

where k_{nf} = rate coefficient dependent on pretreatment
n' = reaction rate coefficient, 0.70

Values for k_{nf} were given as 0.40, 0.47, 0.50, and 0.53, for (1) screening but no primary clarifier, (2) primary clarification or predenitrification, (3) primary clarification and predenitrification and (4) chemically enhanced primary treatment, respectively. The temperature correction value is 1.09. Determination of the MBBR media and reactor volumes for BOD removal and nitrification is illustrated in Example 9–8.

Figure 9–26

Effect of DO concentration and BOD surface area loading rate on nitrification flux. (Adapted from Ødegaard, 2006.)



EXAMPLE 9–8 **MBBR Process Media and Reactor Volumes for BOD Removal and Nitrification** Using the same primary effluent wastewater characteristics as for Example Problem 8–3, determine the plastic carrier media volume required and the aeration tank volume for an MBBR process to produce an effluent NH_4-N concentration of 0.70 mg/L. Use four reactors in series with the first two of equal size for BOD removal and the last two of equal size for nitrification. Assume a 50 percent media fill volume for the BOD removal reactors and 60 percent for the nitrification reactors. Assume a DO concentration of 4.0 mg/L for the nitrification reactors.

Design parameter	Unit	Design condition
Average flow	m^3/d	30,000
BOD concentration	g/m^3	140
TKN concentration	g/m^3	35
Nonbiodegradable VSS	g/m^3	25
TSS	g/m^3	70
VSS	g/m^3	60
Minimum design temperature	°C	12.0
Effluent NH_4-N	g/m^3	≤0.7

Note: g/m^3 = mg/L

Other assumptions are as follows:

1. Plastic biofilm carrier media specific surface area = 500 m^2/m^3.
2. First reactor roughing BOD removal flux (Table 9–15) = 12 $g/m^2 \cdot d$ at 12°C for 75% BOD removal.
3. Second reactor for prenitrification BOD removal flux (Table 9–15) = 4.0 $g/m^2 \cdot d$ at 12°C for 90 percent BOD removal.
4. Use Fig. 9–25 for nitrification flux as function of bulk liquid DO concentration and NH_4-N concentration. Assume ammonia removal in first nitrification reactor is DO limited. Thus the biofilm nitrification flux at 15°C = 1.07 $gN/m^2 \cdot d$.

5. Nitrification flux temperature (θ) correction for DO limited condition = 1.058 (Salvetti et al., 2006).
6. Nitrification flux temperature (θ) correction for ammonia limited condition = 1.098 (Salvetti et al., 2006).
7. bCOD/BOD = 1.6.
8. Heterotrophic bacteria synthesis yield and specific endogenous decay coefficients from Table 8–14.

   $Y_H = 0.45$ g VSS/g bCOD, $b_{H,20} = 0.12$ g/g·d, $\theta = 1.04$ for temperature correction for b_H.
9. BOD removal biofilm SRT = 6.0 d.

Solution

1. Determine plastic media volume needed and tank volume for 1st reactor.

   a. Applied BOD flux $= \dfrac{\text{BOD removal flux}}{(\%\text{ BOD removal}/100)} = \dfrac{(12\ \text{g/m}^2\cdot\text{d})}{0.75} = 16.0\ \text{g BOD/m}^2\cdot\text{d}$

   b. Media area:

   $$= \frac{\text{BOD application rate}}{\text{Applied BOD flux}} = \frac{(30{,}000\ \text{m}^3/\text{d})(140.0\ \text{g BOD/m}^3)}{(16.0\ \text{g BOD/m}^2\cdot\text{d})} = 262{,}500\ \text{m}^2$$

   c. Media volume $= \dfrac{262{,}500\ \text{m}^2}{(500\ \text{m}^2/\text{m}^3)} = 525\ \text{m}^3$

   d. First reactor tank volume $= \dfrac{525\ \text{m}^3}{(0.50\ \text{m}^3/\text{m}^3)} = 1050\ \text{m}^3$

   e. Hydraulic retention time, $\tau = V/Q = \dfrac{1050\ \text{m}^3}{30{,}000\ \text{m}^3}\left(\dfrac{24\ \text{h}}{\text{d}}\right) = 0.84\ \text{h}$

2. Determine plastic media volume needed and tank volume for 2nd reactor.

   a. Applied BOD flux $= \dfrac{(4.0\ \text{g/m}^2\cdot\text{d})}{0.90} = 4.44\ \text{g BOD/m}^2\cdot\text{d}$

   b. Media area: BOD remaining to 2nd reactor

   $$= 0.25(30{,}000\ \text{m}^3/\text{d})(140\ \text{g/m}^3) = 1{,}050{,}000\ \text{g BOD/d}$$

   $$= \frac{(1{,}050{,}000\ \text{g BOD/d})}{(4.44\ \text{g BOD/m}^2\cdot\text{d})} = 236{,}486\ \text{m}^2$$

   c. Media volume $= \dfrac{236{,}486\ \text{m}^2}{(500\ \text{m}^2/\text{m}^3)} = 473\ \text{m}^3$

   d. Second reactor tank volume $= \dfrac{473\ \text{m}^3}{(0.50\ \text{m}^3/\text{m}^3)} = 946\ \text{m}^3$

   use same volume as first reactor = 1050 m^3

   e. Hydraulic retention time, $\tau = V/Q = \dfrac{1050\ \text{m}^3}{30{,}000\ \text{m}^3}\left(\dfrac{24\ \text{h}}{\text{d}}\right) = 0.84\ \text{h}$

3. Determine available NH_4-N for nitrification after nitrogen consumption for heterotrophic bacteria synthesis from BOD removal using Eq. (8–20), Table 8–10.

   $$P_{x,\text{bio}} = \frac{Q(Y_H)(\text{BOD})}{[1 + b_H(\text{SRT})]} + \frac{f_d(b_H)Q(Y_H)(\text{BOD})\text{SRT}}{[1 + b_H(\text{SRT})]}$$

   $$\frac{P_{x,\text{bio}}}{Q} = \frac{(Y_H)(\text{BOD})[1 + f_d(b_H)(\text{SRT})]}{[1 + b_H(\text{SRT})]}$$

$$b_{H,12} = b_{H,20}(1.04)^{(12-20)} = (0.12\text{ g/g·d})(1.04)^{(12-20)} = 0.087\text{ g/g·d}$$

$$Y_H = 0.45\frac{\text{g VSS}}{\text{g bCOD}}\left(\frac{1.6\text{ g bCOD}}{\text{g BOD}}\right) = 0.72\text{ g VSS/g BOD}$$

$$\frac{P_{x,\text{bio}}}{Q} = \frac{(0.72\text{ g VSS/g BOD})(140\text{ g/m}^3)[1 + 0.15(0.087\text{ g/g·d})(6\text{ d})]}{[1 + (0.087\text{ g/g·d})(6\text{ d})]} = 71.4\text{ g VSS/m}^3$$

Ammonia available for nitrification (Eq. 8–24, Table 8–10):

$$\text{NH}_o = \text{TKN} - (0.12\text{ g N/g biomass})\frac{P_{x,\text{bio}}}{Q}$$

$$\text{NH}_o = 35.0\text{ gN/m}^3 - 0.12(71.4\text{ g VSS/m}^3) = 26.4\text{ g N/m}^3$$

4. Determine media volume and tank volume for nitrification reactors.
   a. Adjust DO limited flux for temperature

$$J_{N,12} = 1.07\text{ g N/m}^2\text{·d}(1.058)^{(12-15)} = 0.90\text{ g N/m}^2\text{·d}$$

   b. Determine nitrification flux in 2nd nitrification reactor under ammonia limited conditions from Eq. (9–48).

$$J_{N,15} = \left[\frac{N}{(2.2\text{ g/m}^3) + \text{N}}\right]3.3\text{ g N/m}^2\text{·d}$$

$$= \left[\frac{(0.70\text{ g/m}^3)}{(2.2\text{ g/m}^3 + 0.70\text{ g/m}^3)}\right](3.3\text{ g N/m}^2\text{·d}) = 0.797\text{ g N/m}^2\text{·d}$$

$$J_{N,12} = 0.797\text{ g N/m}^2\text{·d}(1.098)^{(12-15)} = 0.60\text{ g N/m}^2\text{·d}$$

   c. Design the two tanks with equal media area; they will have same volume as they have the same fill volume fraction.

$$\text{Media area} = \frac{(\text{g N removed/d})}{(\text{flux, g N/m}^2\text{·d})}$$

Reactor 3

$$\text{Media area} = A_3 = \frac{[(26.4 - X)\text{g N/m}^3](30{,}000\text{ m}^3\text{/d})}{(0.90\text{ g N/m}^2\text{·d})},$$

where $X = \text{NH}_4\text{-N}$ concentration in reactor

Reactor 4

$$\text{Media area} = A_4 = \frac{[(X - 0.70)\text{ g N/m}^3](30{,}000\text{ m}^3\text{/d})}{(0.60\text{ g N/m}^2\text{·d})},$$

$$A_3 = A_4$$

$$\frac{[(26.4 - X)\text{ g N/m}^3](30{,}000\text{ m}^3\text{/d})}{(0.90\text{ g N/m}^2\text{·d})} = \frac{[(X - 0.70)\text{g N/m}^3](30{,}000\text{ m}^3\text{/d})}{(0.60\text{ g N/m}^2\text{·d})}$$

solve for X; $X = 10.98\text{ g N/m}^3$

$$\text{Media area for each tank} = \frac{[(10.98 - 0.70)\text{g N/m}^3](30{,}000\text{ m}^3\text{/d})}{(0.60\text{ g N/m}^2\text{·d})} = 514{,}000\text{ m}^2$$

   d. Media volume $= \dfrac{514{,}000\text{ m}^2}{(500\text{ m}^2\text{/m}^3)} = 1028\text{ m}^3$

e. Nitrification reactors tank volume $= \dfrac{1028\ \text{m}^3}{(0.60\ \text{m}^3/\text{m}^3)} = 1713\ \text{m}^3$

f. Hydraulic retention time, $\tau = V/Q = \dfrac{1713\ \text{m}^3}{30{,}000\ \text{m}^3}\left(\dfrac{24\ \text{h}}{\text{d}}\right) = 1.37\ \text{h}$

5. Summary of media and tank volumes.

Reactor	Function	DO, mg/L	Media volume, m^3	Tank volume, m^3	τ, h
1	BOD removal	2.0	525	1050	0.8
2	BOD removal	3.0	473	1050	0.8
3	Nitrification	4.0	1030	1720	1.4
4	Nitrification	4.0	1030	1720	1.4
Total			3058	5540	4.4

9–6 SUBMERGED AEROBIC ATTACHED GROWTH PROCESSES

A number of other aerobic attached growth process designs have been used in addition to the moving bed bioreactor process presented in Sec. 9–5. These are compact biological treatment processes with a major difference from the MBBR process being that they do not have secondary clarification. Suspended solids are removed by filtration within the process media or by side stream solids harvesting in the case of the fluidized bed reactor. The purpose of this section is to (1) describe the more common designs used, (2) identify advantages and disadvantages, (3) address the design physical facilities, and (4) provide a process analysis of removal mechanisms and basic design considerations. The more commonly used aerobic attached growth processes are presented in this section and include the downflow *biological aerated filter* (BAF), the upflow BAF, and a fluidized bed biological reactor (FBBR).

Process Development

The term biological aerated filter refers to the fact that the attached growth process is aerated to provide oxygen for BOD removal and nitrification. The biological aerated filter falls within a broader term for these types of systems, which is a *biological active filter*. The term biological active filter includes the biological aerated filter and similar designs that have been employed under anoxic conditions for biological denitrification for nitrogen removal. Attached growth processes for biological denitrification are discussed in Sec. 9–7.

The development of submerged aerobic attached growth processes began in the late 1970s. Examples of such systems reported in the literature are the FBBR with preoxygenation of the influent (Jeris et al., 1977), an upflow granular media reactor with air sparging from the bottom of the media (Young and Steward (1979), a downflow granular media reactor with air sparging near the bottom of the bed above the effluent collection nozzles (Leglise et al., 1980), and an upflow plastic media reactor with air sparging from the bottom (Rusten, 1984). The downflow granular media design by Leglise et al. (1980) was termed the Biocarbone® process because of the early use of granular activated carbon for the media. Later, the media was changed to a fired clay material and the process name was changed to *biological aerated filter* (BAF). The predecessor to wide commercial applications of submerged attached growth processes was the start up of the first commercial

Biocarbone® process near Paris, France, in 1982 (Stephenson et al., 2004). The term *sunken media* has been applied to this application in reference to the immobile high specific density of the fired clay material used.

A similar design using an expanded shale media was developed with its first installations in England (Smith and Edwards, 1994). The trapping of rising air in the downflow operation mode could result in undesirable increased headloss in the downflow BAF, and led to upflow BAF designs, in France and in the United Kingdom, with both sunken and a lighter plastic media (Rogalla and Bourbigot, 1990; Meaney and Strickland, 1994; and WEF 2011). Since the mid-1980s, hundreds of BAFs have been installed, most of which are in Europe.

Process Applications

The aerobic attached growth BAF processes presented in this section have been used primarily for municipal wastewater treatment for the following applications: (1) secondary treatment for BOD removal, (2) secondary treatment for BOD removal and nitrification, and (3) tertiary treatment for nitrification. In addition they have found some use for BOD removal from dilute industrial wastewaters. Phosphorus removal can be accomplished by chemical addition to the influent of the aerobic attached growth processes. While FBBR systems have been used for biological denitrification in tertiary treatment applications, their more common use has been for nitrification and to treat groundwater contaminated with hazardous substances. In some of these applications activated carbon media has been used to provide both carbon adsorption and biological degradation (Sutton and Mishra, 1994).

Most of the aerobic attached growth processes are proprietary designs from various suppliers. Factors that distinguish the different designs and products are (1) the flow direction; upflow or downflow, (2) media density, (3) media size, (4) media material, (5) media depth, (6) fluid velocity, and (7) the method used for excess solids removal. No clarification is used with aerobic submerged attached growth processes and excess solids from biomass growth and influent suspended solids trapped in the system must be removed periodically. Most designs require a backwashing system much like that used in a water filtration plant to flush out accumulated solids, usually on a daily basis.

Process design features for typical aerobic attached growth processes are summarized in Table 9–16. The BAF processes are capable of producing effluent BOD and TSS concentrations of less than 10 mg/L for secondary treatment. Lower loaded single BAF units can provide combined BOD and nitrification or tertiary BAF and FBBR units can provide nitrification to produce effluent ammonia-N concentrations less than 1.0 mg/L or between 1.0 to 4.0 mg/L, depending on the ammonia-N loading and dissolved oxygen concentration.

The high hydraulic application rates possible in BAF and FBBR processes result in empty bed contact times from 0.50 to 1.5 h or even less for FBBR applications. The upflow BAFs are more popular due to their lower and more reliable headloss but they may require a higher amount of product water for backwashing. During demonstration testing in San Diego, California, of upflow BAFs for BOD removal, the floating media BAF backwash water needs were between 10.3 and 13.9 percent of the product water compared to 7.4 to 7.9 percent for an upflow sunken media BAF (Newman et al., 2005).

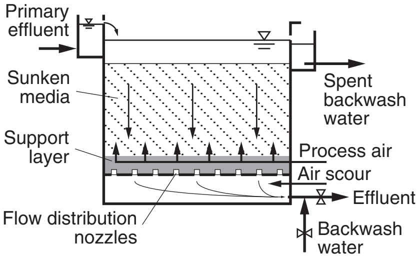
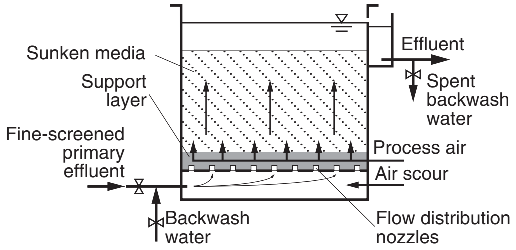
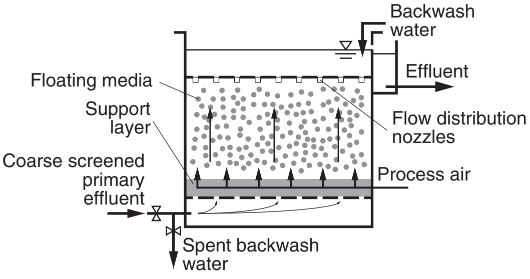
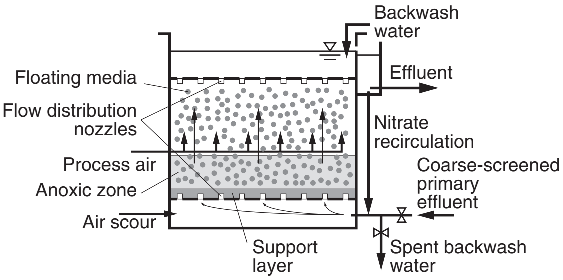
Process Advantages and Disadvantages

The advantages and disadvantages of BAF and FBBR processes compared to activated sludge treatment are as follows:

Advantages for BAF Processes. The major advantages of BAF processes include (1) their relatively small space requirement, (2) the ability to effectively treat dilute

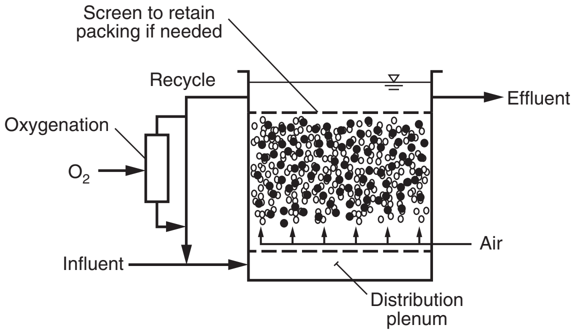
Table 9–16

Summary of design and operating features of commonly used aerobic submerged attached growth processes[a]

Process	Description
Downflow BAF	
(a)Downflow, sunken media  	Primary effluent wastewater is applied above a bed of heavy media at a depth of 1.6 to 2.0 m. Typical media is a 3.0 to 5.0 mm expanded clay or shale with a specific gravity of about 1.6. Air distribution piping is located about 30 cm above underdrain and uniformly distributed across bottom area. Downflow velocity ranges 2.4 to 4.8 m/h. Backwashing with air scour at about 90 $m^3/m^2 \cdot h$ followed by water at about 15 $m^3/m^2 \cdot h$. Backwash once per d or headloss at about 1.8 m. Commercial products include the Biocarbone®, and Biodrof® systems.
Upflow BAF	
(b) Upflow, sunken media  	Primary effluent wastewater is applied through nozzles embedded in a plenum floor below the biofilter bed. Commonly used media are an expanded clay or shale material of 3.5 or 4.5 mm spherical grains, or 2.7 mm angular grains with a specific gravity of 1.6. Typical bed depth is 3 m but range from 2 to 4 m. Upflow velocity ranges from 4 to 6 m/h. Backwashing is typically done once per d with a water flush rate of 10 to 30 $m^3/m^2 \cdot h$ to expand the bed. The backwash usually consists of a drain down, then air scour to breakup the media, followed by concurrent scour air and backwash water, and a final water rinse. Commercial products include the Biofor® system.
(c) Upflow, floating media  	This type uses a floating bed of spherical media that are lighter than water. Two of the major suppliers with similar process designs are: (1) Biostyr® uses a 3.0 to 6.0 mm polystyrene bead with a specific gravity of about 0.5 and (2) Biobead® uses a 2.3 to 2.7 mm polyethylene bead with a specific gravity of about 0.95. The beads form a floating bed in the upper portion of the reactor; typical height of 3 to 4 m. The top of the bed is restrained by a plenum fitted with filtration nozzles to collect the treated wastewater. The media is compressed with the upflow of water during treatment and provides filtration. Plate covered holes in a bottom support plate distribute influent flow and also collect the downflow spent backwash water. Process air is distributed through diffusers located along the bottom of the reactor. Upflow velocity ranges from 4 to 6 m/h. Other commercial products include the Biopure®, and Biolest® systems.
(d) Upflow, floating media with anoxic zone  	A variation of the upflow floating media design for Biostyr systems has been done in full-scale facilities to accomplish nitrification and denitrification in the same attached growth reactor. The process air header is located at about 2 m below the top of the media and a bottom portion of media is a nonaerated, preanoxic zone for biological denitrification. Nitrified effluent is recycled to the influent feed and BOD in the primary effluent provides the electron donor for nitrate reduction.

(*continued*)

Table 9–16 (*Continued*)

Process	Description
Fluidized bed bioreactor	
(e) Upflow fluidized bed bioreactor 	The wastewater is fed upward to a fluidized bed of 0.3 to 0.7 mm sand or 0.6 to 1.4 mm activated carbon. Bed depths range from 3.0 to 4.0 m. The upflow velocity used for 0.5 mm silica sand is from 30 to 40 m/h. Recycle flowrate is 2 to 5 times the influent flowrate to maintain media fluidization. Influent hydraulic retention times range from 10 to 20 min. The influent is introduced through a perforated plate or downward facing nozzle.

[a] Borregaard (1997), Freihammer et al. (2007), Holbrook et al. (1998), Lazarova et al. (2000), Mendoza and Stephenson (1999), Pujol et al. (1994), Sutton et al. (1981), and U.S. EPA (1993).

wastewaters, (3) no issues with regard to sludge settling characteristics, (4) simplicity of operation, and (5) ability to house it due to the small space needs. The ability to house it has the advantage of aesthetics and less effect of weather on operating personnel. Also for many processes solids filtration occurs to produce a high quality effluent.

Disadvantages for BAF Processes. Their disadvantages include (1) a more complex system in terms of operation and maintenance of instrumentation and controls, (2) limitations of economies of scale for application to larger facilities, (3) generally a higher capital cost unless land is at a premium or not available, and (4) vulnerable to high headloss from high solids loadings. The design and cost of BAF and FBBR processes are impacted directly by the hydraulic flowrate. Flow equalization should be considered for high hydraulic peak flows from wet weather events.

Advantages for FBBR Process. The main advantages for the FBBR technology for treatment of dilute organic substrates and for biodegradation of xenobiotics are (1) it provides an extraordinary long SRT for microorganisms necessary to degrade the xenobiotic and toxic compounds, (2) shock loads or non-biodegradable toxic compounds can be absorbed onto the activated carbon, (3) a high quality effluent with low TSS and COD concentration is possible, (4) the oxygenation method prevents stripping and emission of toxic organic compounds to the atmosphere, and (5) the system operation is simple and reliable.

Disadvantages for FBBR Process. The main disadvantages recognized for FBBR are (1) high energy requirement for high recycle flow pumping, (2) higher oxygen supply rate required, and (3) operational challenges in biomass and process control.

Design of Physical Facilities

The design components of a BAF system are very similar to those for water treatment filtration facilities, including influent flow distribution, media selection, and backwash water storage, water pumps, air scour compressors, and automated controls and valves.

It is common for a supplier to provide package or modular designs for standard BAF treatment units with necessary piping, valving, pumps, blowers and controllers. For wastewater treatment applications the physical facility designs depend on the characteristics of the influent wastewater, methods to remove accumulated solids, and process air supply. Important physical facility design issues addressed in this section are (1) pretreatment requirements, (2) methods to remove excess solids, and (3) oxygen supply.

Pretreatment Requirements. Pretreatment is required for both the upflow and downflow BAF processes to avoid clogging of the influent or effluent flow distribution nozzles from the collection of fibers or large influent particle size. In addition, primary treatment is normally used to control the amount of solids that must be removed by filtration. Higher influent suspended solids concentrations would result in more frequent backwashing and the use and recycle of a greater portion of the effluent product water. The net effect would be the need for more BAF treatment area and possibly more treatment units.

The need for and type of additional pretreatment required following primary treatment depends on the type of BAF process selected. A downflow BAF with sunken media does not require additional pretreatment as the primary effluent solids are removed in the filter bed before the effluent flow reaches the underdrain collection nozzles. An upflow sunken media design introduces the primary effluent flow through underdrain nozzles and thus fine screening with a screen size of 2.5 mm is typically recommended. The upflow floating media BAF design does not have influent flow distribution nozzles, but prescreening with a screen size of about 10 mm is recommended to prevent the collection of fibers on the effluent nozzles or on the media.

Excess Solids Removal. Periodic backwashing of a BAF is needed to remove accumulated solids from the influent wastewater filtration and produced by biological growth on influent wastewater soluble and particulate substrates. The backwash procedure varies depending on whether sunken or floating media is used, but in both cases air scour and water flushing are employed. The backwash water source is treatment effluent water from a backwash water storage tank or from the effluent flow from a number of BAF modules in larger plants. The backwash frequency may be set for 24 h increments or governed by a preset headloss value.

Backwashing of the sunken media BAF involves taking the unit off line and subjecting it to an air scour and water backwash flushing cycle that is automatically controlled by timers and actuated valves. During backwashing of BAFs with floating media, treated water flows down through the media at a very high rate to cause a downward expansion of the originally compressed material. The solids retained in the lower portion of the reactor and excess biomass produced on the packing is flushed out through the influent underdrain plate. The normal backwash procedure consists of repeated rinse (water flushing) and air agitation steps. The air flowrate in the process air header is increased during backwashing. The downward water flow expands the bed to release solids held in the media pore space and air scouring agitates the bed to dislodge solids from the media. Typically, 4 water phases and 3 air phases are used (WEF, 2011). Similar backwash air and water application rates are used for both types of BAF processes. Air and water flush rates vary from 0.8 to 1.0 m/min and water flush rates vary from 0.6 to 1.0 m/min (Stensel et al., 1988; Mendoza-Espinosa and Stephenson, 1999).

The biofilm thickness on FBBR media increases with time as substrate is removed, which then lowers the particle specific density and the lighter particles migrate to the top of the bed. Wasting of excess biogrowth is then done by transporting the biofilm carrier media from the top of the bed to an external device, such as vibrating screen, hydrocyclone or airlift system, for agitation and stripping off the excess biomass. The media carrier is then recycled back to the fluidized bed reactor.

Processing Backwash Water. Spent backwash water is returned to head of the treatment plant prior to primary settling tanks or is treated in a separate unit. Recycle of BAF backwash water to the primary treatment step is thought to improve BOD removal in primary treatment due to adsorption of BOD by the BAF waste biomass. Treatment in a separate unit is done if the existing primary treatment has limited hydraulic and/or solids loading capacity or may be preferred for larger facilities. The types of technologies used include ballasted flocculation and settling, a solids contact/sludge recirculation system, dissolved air flotation thickening, and gravity settling (WEF, 2011). The treatment of return flows is considered in Chap. 15.

Aeration Equipment. Process air is typically supplied by coarse bubble aeration through a piping grid system located at the bottom of the sunken media or below the floating media. The sparge air pipes may be placed 250 to 300 mm apart with holes for coarse bubble air drilled at equal intervals to provide a uniform air application rate across the bed area. The use of fine bubble diffusers does not provide any improvement in oxygen transfer efficiency because the bubbles coalesce into large bubble size in the media (Harris et al., 1996). Blower design for a downflow sunken media BAF must consider the effect of headloss on the blower air supply rate. As solids accumulate in the bed, the headloss and liquid level above the bed increases, which can decrease the blower air supply rate. The upflow BAF designs with sunken and floating media do not have this problem as the liquid level on top of the bed is set with effluent overflow.

BAF Process Design Analysis

A BAF process is a high rate treatment system that employs dual functions of biological oxidation of BOD and/or ammonia and physical removal of particulate and colloidal solids by absorption and filtration. Important issues that affect the design and performance of BAF processes are discussed in this section and include (1) media characteristics, (2) process loadings, (3) aeration design, and (4) sludge production.

Media Size. The average media size affects (1) the surface area available per unit of reactor volume for biofilm growth and substrate removal rates and (2) the pore size and filtration efficiency. The specific surface area of a sphere is given by Eq. 9–50.

$$SSA_{sp} = \frac{6000}{D_{sp}} \tag{9–50}$$

where: SSA_{sp} = specific surface area of a spherical particle, m^2/m^3
D_{sp} = diameter of particle, mm

As an example of the magnitude of surface area available for biomass growth in submerged attached growth processes, a BAF system with 3-mm sunken media and bed porosity of 40 percent would have a bulk reactor volume specific media surface area of 1800 m^2/m^3. For a FBBR process with a 0.50 mm sand and 100 percent bed expansion, the bulk reactor specific surface area is estimated at 6000 m^2/m^3. The actual area would be less as the media is normally given in terms of the effective size and the biofilm growth increases the effective particle diameter.

The media size selection in a BAF design must be based on a consideration of the effluent TSS concentration needed and filtration efficiency, the organic removal efficiency needed and organic loading, and backwash requirements. A smaller media size allows a greater organic loading due the higher bulk specific surface area and produces a lower effluent TSS concentration. However, the BAF run time before backwashing would be less, a greater fraction of the effluent would be needed for backwashing, and the operating costs would be higher. A media size of about 6 mm for partial BOD removal, an intermediate size of 4 to 5 mm for secondary treatment, and less than or equal to 3 mm for effluent polishing or tertiary nitrification is recommended (Mendoza-Espinosa and Stephenson, 1999).

Process Design Loadings. Substrate removal mechanisms in a BAF process include removal of soluble substrates by diffusion and biodegradation kinetics in the media biofilm area, and adsorption and filtration of colloidal and particulate material. Soluble substrate removal is a diffusion-limited biological treatment process with removal rates affected by the biofilm area, bulk liquid substrate and DO concentrations, upflow or downflow fluid velocity, and temperature. The particulate BOD removal is affected by media size, fluid velocity, operating headloss and backwash frequency. Because of the process complexity with regard to predicting the biofilm growth, biofilm thickness and biomass density, and hydraulic flow patterns process, designs are based on volumetric process loading rates and performance observed in full-scale and pilot-scale facilities. BAF performance for secondary treatment, combined BOD removal and nitrification, and tertiary nitrification is usually measured as a function of kg BOD/m^3·d or kg N/m^3·d. For more dilute wastewaters, hydraulic loading rates may govern the design. Typical hydraulic application rates for each type of BAF process are included with the process descriptions in Table 9–16.

A range of volumetric BOD loading rates reported for BAF processes is given in Table 9–17 for treatment for BOD removal only and for treatment at lower BOD loadings to accomplish BOD removal and nitrification in the same BAF unit. Though the loadings are typically based on total BOD, the soluble BOD loading is of importance in determining the effluent soluble BOD and final total BOD concentration. It has been found that for a downflow sunken media BAF process the soluble BOD volumetric loadings below 1.0, 1.2, and 1.4 kg/m^3·d were needed to achieve a secondary effluent BOD at 10–12, 12–16, and >16°C, respectively (Stensel et al., 1988). Thus, the volumetric BOD loading selected should be at the lower range of the values shown in Table 9–18 for wastewater with a higher soluble BOD fraction.

BAF processes have been operated at low BOD loadings to achieve BOD removal and nitrification. As was discussed in Sec. 9–4 and 9–5 for trickling filters and MBBR processes, nitrification in a BAF process does not occur until most of the soluble BOD is first removed so that surface area is available for nitrifying bacteria. The net effect on the process design is that a much lower overall volumetric BOD loading is used for combined BOD removal and nitrification BAF processes as shown in Table 9–17. The DO concentration for nitrification applications should be at least 3 to 4 mg/L to obtain reasonable nitrification rates as the process is often DO limited. Maintaining elevated DO concentration is difficult

Table 9–17

Process volumetric loadings for biological aerated filters[a]

Process application	Loading units	Range	Removal efficiency, %
BOD removal	kg BOD/m^3·d	3.5–5.5	≥85
BOD removal and nitrification	kg BOD/m^3·d	1.8–2.5	≥85
Tertiary nitrification	kg NH_4-N/m^3·d	1.0–1.5	≥90

[a] Mendoza and Stephenson (1999), WEF (1998), Tchobanoglous et al. (2003), WEF (2011).

Table 9–18

Oxygen transfer efficiencies expressed in percent/m of depth observed for biological aerated filters

BAF Design	Test system depth, m	O_2 transfer efficiency, %/m	Test system	Reference
DF, sunken	1.6	3.4–5.5	Full-scale	Stensel et al. (1988)
DF, sunken	2.0	5.0–8.5	Full-scale	WER (2011)
UF, sunken	3.6	1.6–5.8	Lab-scale	Stenstrom et al. (2008)
UF, sunken	4.0	5.0	Full-scale	Laurence et al. (2003)
UF, floating	3.6	3.6–8.0	Lab-scale	Stenstrom et al. (2008)
UF, floating	3.0	6.7	Full-scale	Laurence et al. (2003)

in combined BOD removal and nitrification processes due to the high volumetric oxygen demand for BOD removal.

An alternative to a combined BOD removal and nitrification process is to use two BAF units in series with the first unit designed at a BOD loading to meet a secondary effluent BOD concentration of less than 15 mg/L and the second stage unit designed for nitrification. With two units, design and operating conditions for nitrification are more optimal, with each unit operating at different DO concentrations, hydraulic application rates, and backwash frequency. Based on nitrification studies, 85 to 90 percent nitrogen oxidation was found at loadings of 1.5 to 1.8 kg N/m^3·d for tertiary nitrification applications (Payraudeau et al., 2000). A range of ammonia-N loadings used for tertiary nitrification is shown in Table 9–17. The volumetric nitrification rate was found to increase by about 3 percent for each degree centigrade temperature increase from 10°C for all types of BAFs (Tschui et al., 1994).

Aeration Design. The volumetric oxygen uptake rates for BAFs are very high compared to activated sludge processes due to their relatively short hydraulic retention times and high organic loading rates. Volumetric oxygen uptake rates as high as 250 mg/L·h have been measured in a 1.7 m deep downflow BAF unit (Stensel et al., 1984). Even though coarse bubble aeration is used, the actual oxygen transfer efficiency is greater than what would be expected for fine bubble diffused aeration at the same depth because of bubble hold up in the BAF media (Stensel et al., 1984; Lee and Stensel, 1986; and Stenstrom et al., 2008). Actual process oxygen transfer efficiencies are higher at lower air application rates per unit area due to greater bubble hold up time. A range of reported oxygen transfer efficiencies are summarized in Table 9–18. When higher air application rates are used to maintain DO concentrations at higher loadings, the lower range of values should be used for design.

The oxygen demand in a BAF process is a function of the wastewater characteristics, BOD loading, and backwash frequency. It has been found that for a downflow sunken media system loaded at 3.5 kg BOD/m^3·d and once per day backwash, only about 20 percent of the influent volatile suspended solids were degraded due to the short solids retention time (Stensel et al., 1984). The following equation can be used to estimate the oxygen demand associated with BOD and volatile suspended solids removal.

$$\mathrm{OR} = 0.82\frac{\mathrm{sBOD}_o}{\mathrm{TBOD}_o} + \frac{1.6(\mathrm{BF}_{\mathrm{VSS}})X_o}{\mathrm{TBOD}_o} \qquad (9\text{–}51)$$

where OR = oxygen required, g O_2/g BOD applied
sBOD_o = influent soluble BOD concentration, g/m^3
TBOD_o = influent total BOD concentration, g/m^3
$\mathrm{BF}_{\mathrm{VSS}}$ = fraction of influent volatile solids degraded, g/g
X_o = influent volatile suspended solids concentration, g/m^3

Sludge Production. As shown above for the oxygen requirements, the sludge production is also related to the influent characteristics and fraction of influent VSS degraded. The following equation can be used to estimate the sludge production (Stensel et al., 1984). The VSS/TSS ratio may be in the range of 0.80 to 0.85 following primary treatment.

$$P_{X,\text{VSS}} = [0.60(\text{sBOD}_o) + (1 - \text{BF}_{\text{VSS}})(X_o)]Q \tag{9–52}$$

where $P_{X,\text{VSS}}$ = volatile solids production rate, g/d
Q = influent flowrate, m^3/d

The solids accumulation capacity before backwashing is needed may be in the range of 2.4 to 3.0 kg TSS/m^3 (Stensel et al., 1984; WEF, 2011). The backwash suspended solids concentration may range from 500 to 1500 mg/L, depending on the influent wastewater characteristics, BOD loading, and backwash frequency.

FBBR Process Design Analysis

Aerobic FBBR applications involve an external oxygen transfer device, bed fluidization, and solids removal for biofilm control. The media biofilm thickness affects the reaction rate, and it can be controlled and optimized by the selection of the bioparticle composite concentration based on support media volume and degree of bed expansion. The bed expansion typically ranges from 50 to 100 percent. A lower degree of expansion allows a lower upflow velocity and requires less energy, but has a higher volumetric oxygen demand due to the increases in biomass concentration.

The application of aerobic treatment with FBBR systems for biodegradation of contaminated groundwaters or dilute industrial wastewater is very site specific and requires an understanding of the degradation characteristics of the compounds of interest. FBBR systems have been used for ammonia removal.

A bed depth of 5 to 6 m has been used for nitrification applications at an ammonia loading in the range of 1.0 kg NH_4-N/m^3·d. The recommended hydraulic loading rate is less than 40 m^3/m^3 bed·d and 25 m^3/m^3 bed·d to achieve effluent ammonia-N concentration of ≤5 mg/L and 0.5 mg/L, respectively (Dempsey et al., 2006).

9–7 ATTACHED GROWTH DENITRIFICATION PROCESSES

Over the years, a number of attached growth denitrification processes have been developed for both postanoxic and preanoxic applications. Topics included in this section include (1) process development, (2) common types of attached growth processes used for biological denitrification, (3) process design analysis, and (4) process operation considerations. The physical facility design considerations are covered elsewhere, either in the previous sections in this chapter or in Chap. 11.

Process Development

The development of attached growth biological denitrification processes began in the early 1970s following an increase in regulations to restrict effluent nitrogen discharges from municipal wastewater treatment plants for control of eutrophication. The aim at that time was to remove nitrate from secondary effluents, and the first two process concepts developed were a downflow denitrification filter patented by Dravo Corp in 1973 (WERF, 2010) and an anoxic fluidized bed reactor (Jeris et al., 1974). The first full-scale anoxic fluidized bed reactor was installed at the Truckee Meadows Water Reclamation Facility near Reno, NV, in the early 1980s following a nitrifying trickling filter (Sedlak, 1991). Since then a number of anoxic attached growth denitrification process designs have been developed and marketed by various suppliers. These include upflow and downflow filters, an anoxic moving bed bioreactor, and an anoxic submerged rotating biological contactor.

Description and Application of Attached Growth Denitrification Processes

Attached growth denitrification processes have been applied most often as a postanoxic treatment step after nitrification. However, in some cases attached growth denitrification has been used in a preanoxic mode prior to nitrification. Both types are discussed in this section.

Attached Growth for Postanoxic Denitrification. The most commonly used attached growth process designs for biological denitrification, shown on Fig. 9–27. can be classified into three types: (1) denitrification filter, (2) suspended media denitrification, and (3) fluidized bed denitrification. All of these systems are provided by different manufacturers, with each specifying their selection of media type and size, media depth, hydraulic application rates, and backwashing methods for denitrification applications.

Denitrification filters have been included under a general category of biological active filters in the Water Environment Federation MOP 35 on biofilm reactors (WEF, 2011). The terminology for biological active filters in the document also includes the biological aerated filters (BAFs) that were discussed in Sec. 9–6. All of the types of BAFs presented in Table 9–16 in Sec. 9–6 are changed to denitrification filters by eliminating the addition of process air, and having nitrate and an exogenous carbon feed in their influent.

As done for the BAFs, these denitrification filters are distinguished by their treatment flow direction and media characteristics and are shown as (1) downflow sunken media [see Fig. 9–27(a)], (2) upflow sunken media [see Fig. 9–27(b)], and (3) upflow floating media [see Fig. 9–27(c)]. In addition, a fourth type of denitrification filter is the continuous

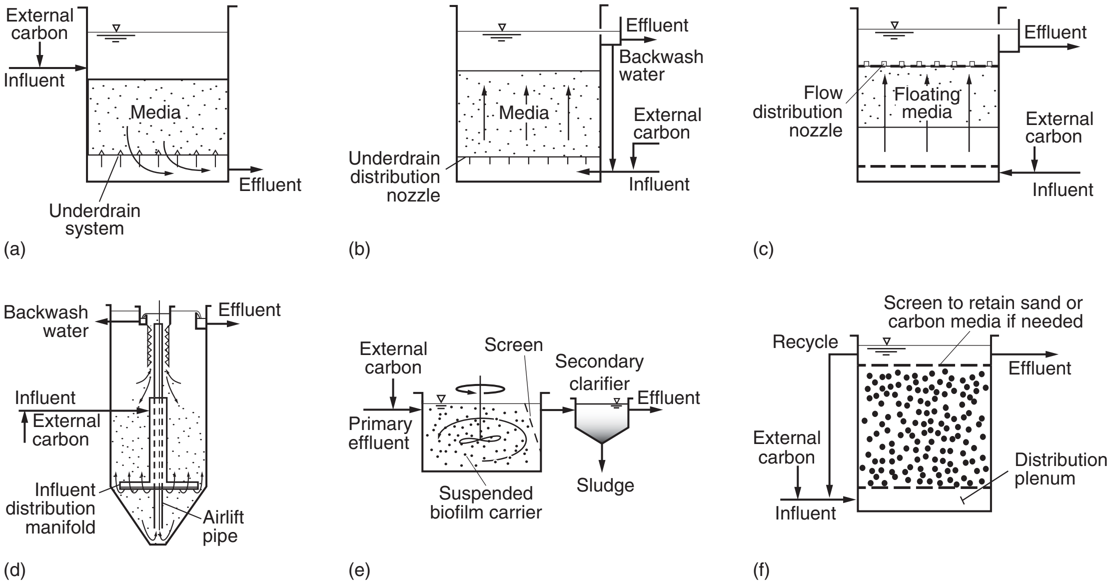


Figure 9–27

Commonly used attached growth processes for biological denitrification. Denitrification filters (DNFs): (a) downflow sunken media, (b) upflow sunken media, (c) upflow floating media, and (d) continuous backwash sand filter. Suspended media denitrification processes: (e) anoxic moving bed bioreactor (AnoxMBBR) and (f) anoxic fluidized bed reactor (AnoxFBBR).

backwash sand filter shown on Fig. 9–27(d). Downflow and continuous backwash sand filters for suspended solids removal are described in Sec. 11–3 in Chap. 11. In the anoxic continuous backwash filter, the wastewater enters the filter at the top and is conveyed to the bottom of the bed where it is distributed through feed radials and then travels upward for treatment through the filter bed. An airlift in a center column in the filter provides a continuous backwash of the media so that the filtration operation is not interrupted and taken off line for backwashing, which is in contrast to the other denitrification filters that must be removed from service for backwashing.

The AnoxMBBR reactor [see Fig. 9–27(e)] has been described in Sec. 9–5 and has been used in both preanoxic and postanoxic zones in an MBBR biological nitrogen removal treatment system and as a postanoxic denitrification process after secondary or tertiary nitrification (see Table 9–13). The AnoxFBBR [see Fig. 9–27(f)] uses a much smaller media size and thus has the highest reactor specific surface area of the attached growth biological denitrification systems.

Need for Carbon Source. A carbon source must be added to the influent of postanoxic attached growth denitrification reactors to support biological growth and create a demand for nitrite/nitrate as an electron acceptor. The use of nitrate and nitrite as an electron acceptor by heterotrophic bacteria for organic substrate consumption with the evolution of nitrogen gas has been discussed in Sec. 7–10 and in Sec. 8–7 for suspended growth anoxic processes. Methanol is the most commonly used exogenous carbon source for postanoxic denitrification because of its low cost per unit of nitrate-N removed and lower solids yield. However other carbon sources, such as ethanol and glycerol, have been considered for postanoxic attached growth denitrification.

Influent Feed Conditions. Typical upstream treatment and influent feed conditions for postanoxic attached growth processes fall into one of two scenarios: Scenario I is treatment of effluent from an activated sludge biological nitrogen removal process (i.e. MLE process) that has a NOx-N (NO_3-N plus NO_2-N) concentration below 10 mg/L, and Scenario II is treatment of an attached growth nitrified process effluent with a NOx-N concentration in the range of 20 to 35 mg/L. Many of the applications with Scenario I also require a low effluent TSS concentration and thus a denitrification filter is often used to remove both NOx-N and TSS to meet low total nitrogen and low TSS effluent concentrations. More stringent effluent TN requirements, such as less than or equal to 3.0 mg/L, cannot be met without effluent filtration. For Scenario II, a greater amount of carbon must be added to the postanoxic denitrification reactor, which results in a higher volumetric denitrification rate and greater amount of biomass production. The AnoxMBBR and AnoxFBBR processes can accommodate the higher nitrate and carbon loadings but normally need an effluent polishing step for suspended solids removal.

Typical Performance. Survey of full-scale facilities with postanoxic attached growth denitrification and with over 3 years of performance was done with data from 4 facilities to provide a statistical treatment performance evaluation. The 50th percentile values for the effluent TN concentrations were less than or equal to 1 mg/L. The 90th percentile effluent TN concentrations ranged from 2.7 to 4.2 mg/L for the different facilities (Bott and Parker, 2010).

Attached Growth for Preanoxic Denitrification. Attached growth denitrification processes that have been used for preanoxic denitrification include applications with (1) trickling filters (Nasr et al., 2000; and Dorias and Baumenn, 1994), (2) denitrification filters (Ninassi et al., 1998), and (3) applications with MBBRs (Lazarova et al., 1998). As shown in Table 9–13(e) and (g) in Sec. 9–5, an AnoxMBBR can be used for preanoxic nitrate removal in an MLE- or Bardenpho-type nitrogen removal process.

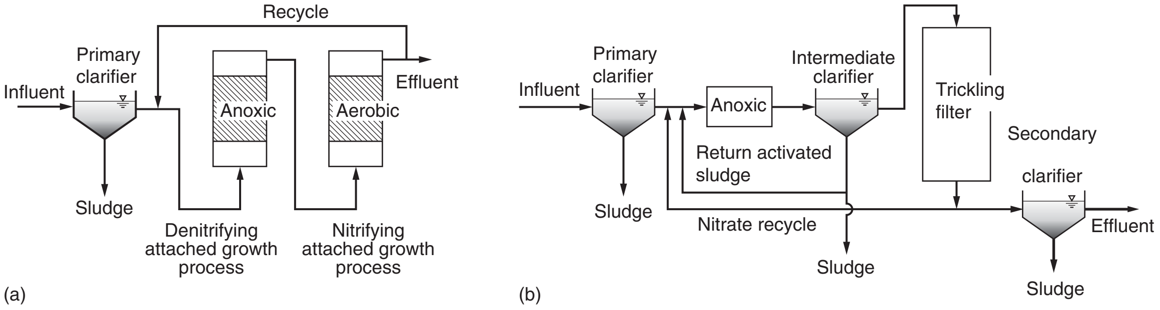


Figure 9–28

Example of preanoxic attached growth processes: (a) sequential upflow denitrification-nitrification biological active filters and (b) recirculation of nitrifying trickling filter effluent to a preanoxic suspended growth process.

Other examples of preanoxic denitrification with attached growth processes are illustrated on Fig. 9–28. In the first case [see Fig. 9–28(a)], an attached growth process is used for denitrification using organic material in the primary effluent wastewater for electron donors to reduce NOx-N that is contained in the recycle from the downstream nitrifying attached growth process. The recirculation rate is about 3 to 4 times the influent flowrate and the nitrogen removal performance is similar to that for the MLE process used in activated sludge nitrogen removal (Sec. 8–7). A nitrate-N loading of 1.0 to 1.5 kg/m^3·d and hydraulic application rate of 20 to 30 m/h has been reported for a preanoxic upflow sunken media process (Ninassi et al., 1998). The high recirculation rate increases the energy needs for pumping flow to the attached growth processes and the hydraulic load.

In the second case [see Fig. 9–28(b)] nitrified trickling filter effluent is recycled to the upstream suspended growth preanoxic system (Melhart, 1994). An intermediate clarifier is used to separate the denitrifying mixed liquor and to provide return activated sludge to the anoxic tank. The recycle flow needed to provide nitrate to the preanoxic zone has a significant effect on the clarifier size and pumping requirements to the trickling filter, and thus the overall system economics.

Examples of converting full-scale trickling filter systems to nitrogen removal have been given by Nasr et al. (2000) and Dorias and Baumenn (1994). Recycle from a second stage nitrifying trickling filter was returned to an upstream first-stage preanoxic submerged or covered gas-tight trickling filter. The recirculation flows greatly increased the trickling filter energy requirements and high DO concentrations in the trickling filter effluent affected the preanoxic nitrate removal efficiency.

An important advantage of using a preanoxic attached growth denitrification process is that influent BOD is used for nitrate reduction, which eliminates the cost of an exogenous carbon source. Major disadvantages are the effects of the nitrate recycle flow on the design and operating cost and the impact of the higher influent solids concentration on attached growth preanoxic designs.

Process Design Analysis of Postanoxic Attached Growth Denitrification

Important design factors for postanoxic attached growth processes discussed here are (1) hydraulic application rate, (2) nitrate loading rate, (3) exogenous carbon supply, (4) excess solids accumulation and backwash needs, and (5) phosphorus limitations.

Table 9–19

Range of reported process loadings for postanoxic attached growth processes

Type of system	Hydraulic application rate, m/h	NO_3-N loading rate, kg/m^3·d	Minimal EBCT[b], min	References
Downflow sunken media DNF	2.4–4.8	0.3–3.2	25	Falk et al. (2011)
Upflow sunken media DNF	4.0–6.0	0.8–5.0	20	WEF (2011)
Upflow floating media DNF	4.0–6.0	1.5–2.0	20	WEF (2011)
Continuous backwash DNF	2.4–8.0	0.3–2.0	15	deBarbadillo et al. (2005)
AnoxFBBR	15.0–25.0	3.0–5.0	6	U.S. EPA (1993)
AnoxMBBR		0.25–0.5	30[a]	Stinson et al. (2009)

[a] Assumes media specific surface area of 500 m^2/m^3 and 50 percent fill volume.

[b] EBCT = empty bed contact time.

Hydraulic Application and Nitrate Loading Rates. Two key design considerations for attached growth postanoxic denitrification processes are the hydraulic application rate and volumetric nitrate loading rate. The nitrate loading rate to achieve a target effluent NO_3-N concentration is the key factor for sizing the AnoxMBBR and AnoxFBBR processes, whereas both the hydraulic loading rate and volumetric nitrate loading rate must be considered for the denitrification filters when the filters must also produce a low effluent TSS concentration. In the case of Scenario I described above, the hydraulic loading rate required to meet a low effluent TSS concentration is the controlling design parameter because after the upstream nitrogen removal system the feed nitrate-N concentration is low enough so that the nitrate loading at the filtration hydraulic applications rates does not stress the nitrate removal performance.

Hydraulic Loading Rates. A range of hydraulic loading rates and volumetric nitrate loading rates reported for attached growth are summarized in Table 9–19. The minimal empty bed contact time (EBCT), which is the tank volume divided by the influent flowrate, is based on assuming a media depth of 2.0 m for the DNFs and using the higher respective hydraulic application rate. The AnoxMBBR EBCT is based on assuming the higher volumetric nitrate-N loading rate shown, a suspended media specific surface area of 500 m^2/m^3, and a 50 percent fill volume.

Nitrate Loading Rates. A wide range of volumetric nitrate loading rates are reported for the denitrification filters; from 0.3 to 5.0 kg NO_3-N/m^3·d. These values are based on pilot plant and full-scale facility observations and represent varying conditions of influent nitrate-N concentration, carbon dose, temperature, hydraulic application rates, and effluent goals. Performance results of a number of postanoxic denitrification filters were evaluated as a function of the EBCT by deBarbadillo et al. (2005). Based on the data from 14 test sites, it was found that greater than 90 percent nitrate-N removal efficiency could be obtained at EBCTs of 10 min or greater. In addition pilot plant performance data for a continuous flow backwash upflow denitrification filter was evaluated with a plug flow, substrate diffusion half-order reaction kinetics model by Harremöes (1976) and applied by Hultman et al. (1994) as follows:

$$NO_e^{(1/2)} = NO_o^{(1/2)} - (1/2)k_{DN}(EBCT) \qquad (9\text{–}53)$$

Where NO_e = effluent NO_3-N concentration, g/m^3
NO_o = influent NO_3-N concentration, g/m^3
k_{DN} = half-order kinetic coefficient, mg/L·min
EBCT = empty bed contact time, min

Note the nitrate-N load is defined as follows:

$$\mathrm{NL} = \frac{Q(\mathrm{NO}_o)}{V} = \frac{1.44(\mathrm{NO}_o)}{\mathrm{EBCT}} \tag{9–54}$$

where NL = volumetric nitrate-N load, kg NO_3-N/m^3·d

By combining Eq. (9–53) and (9–54), the volumetric nitrate-N load can be related to the influent and effluent NO_3-N concentrations and the kinetic coefficient:

$$\mathrm{NL} = \frac{-0.5k_{\mathrm{DN}}\,(1.44\mathrm{NO}_o)}{\mathrm{NO}_e^{(1/2)} - \mathrm{NO}_o^{(1/2)}} \tag{9–55}$$

A value of 0.36 mg/L·min was obtained by deBarbadillo et al. (2005) from the pilot test results when operating at a higher loading of 1.6 kg NO_3-N/m^3·d to stress the system performance.

The required nitrate-N volumetric loading rate as a function of effluent nitrate-N concentration for influent nitrate-N concentrations from 10 to 30 mg/L was calculated using Eq. (9–55) and k_{DN} = 0.36 mg/L·min. The results are shown on Fig. 9–29. An effluent concentration of 1.0 mg/L can be met at volumetric nitrate-N loadings of 1.2 to 1.7 kg NO_3-N/m^3·d, which is within the range of loadings given in Table 9–19. The EBCT calculated for the same conditions are shown on Fig. 9–30. The calculated EBCT of 12 min for an effluent nitrate-N concentration of 1.0 mg/L generally agrees with the observations noted above by deBarbadillo et al. (2005) that 90 percent nitrate removal can be achieved at EBCTs of 10 min or more. For a 2-m denitrification filter bed depth and nominal hydraulic application rate of 4 m/h, the EBCT is 30 min. Based on these results, it was found that the denitrification filter hydraulic loading design needed to meet a low effluent TSS filtration goal will result in sufficient contact time for postanoxic nitrate removal needs. In the above analysis it was assumed that the carbon dose was sufficient to drive the nitrate-N reduction reaction.

Exogenous Carbon Supply. A carbon source must be added to the postanoxic denitrification process in proportion to the rate of nitrate, nitrite, and oxygen addition in the influent feed. Methanol has been the most common carbon source used in postanoxic

Figure 9–29

Predicted denitrification filter nitrate volumetric loading (kg NO_3-N/m^3·d) versus effluent nitrate-N concentration for influent NO_3-N concentrations from 10 to 30 mg/L (based on Harremöes (1976) substrate diffusion half-order kinetics model).

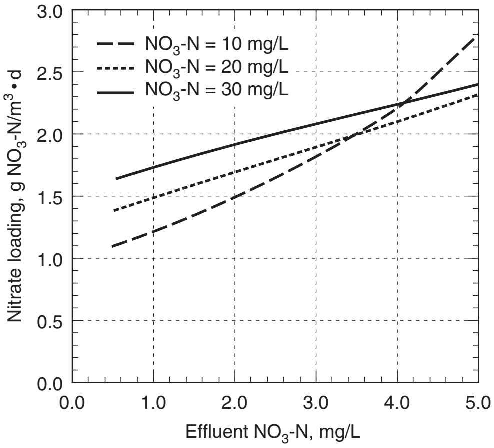
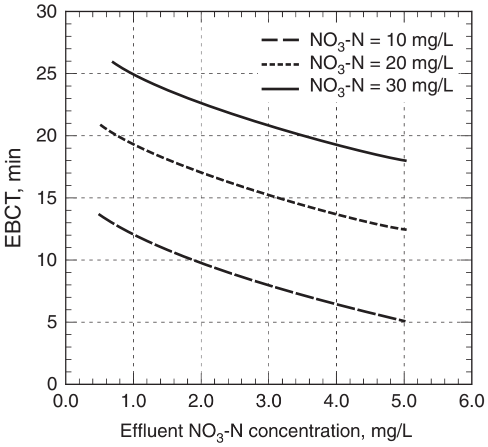


Figure 9–30

Calculated necessary denitrification filter empty bed contact time (EBCT) versus effluent nitrate-N concentration for influent NO_3-N concentrations from 10 to 30 mg/L (based on Harremöes (1976) substrate diffusion half-order kinetics model).



attached growth denitrification processes, in view of its lower cost per unit of nitrate removed and lower solids production. A typical methanol to NO_3-N dose ratio that accounts for some influent DO is in the range of 3.0 to 3.5 kg methanol/kg NO_3-N removed. However, issues with methanol handling and the possible need for intermittent nitrate removal has spurred interest in other carbon sources, such as ethanol and glycerol or commercial products with glycerol. The impact of other carbon sources on dose requirements and biomass solids production has been addressed in Sec. 8–7. The use of ethanol or glycerol increases the carbon dose ratio and sludge production by approximately 22 and 30 percent, respectively, compared to that for methanol. A higher sludge production rate can increase backwashing needs and backwash energy for the removal of excess solids.

Efforts should be made to optimize the upstream nitrogen removal performance to minimize the carbon dose and cost for a postanoxic denitrification process. Issues to address for an upstream MLE process include the internal recycle ratio, management of recycle flows from digester biosolids dewatering, and DO addition to the preanoxic zone from the internal nitrate recycle flow and influent flow.

Excess Solids Control. During operation of a denitrification filter, headloss gradually increases because of solids accumulation (filtration), biomass growth, and accumulation of nitrogen gas due to denitrification. The nitrogen gas accumulation is more pronounced in downflow filters. The downflow filter is "bumped" periodically by an upflow hydraulic surge. A water-only flush is used at a rate of 12 m/h for about 3 to 5 min for a "bump" to release the accumulated nitrogen gas. The bump frequency may vary from once every 2 to 4 h. No special backwash "bumping" is provided for nitrogen gas release for the upflow and continuous backwash filters.

Upflow and downflow denitrification filters must be taken off line for an air and water backwash about every 24 to 48 h, depending on the solids accumulation and headloss. An advantage of the continuous backwash filter is that it does not have to be taken off line. The solids storage capacity is estimated to be about 4.0 kg TSS/m^3 before an unacceptable headloss occurs. Most of the solids accumulated are from the biomass growth from the external carbon source. For example, at a nitrate-N loading of 1.5 kg NO_3-N/m^3·d, the biomass solids accumulation in 24 h would be about 2.8 kg TSS/m^3; assuming a methanol dose of 3.5 kg/kg NO_3-N, and biomass yield of 0.30 g VSS/g CH_3OH COD. Assuming the influent TSS concentration is the same as the influent NO_3-N concentration at 10 mg/L and the effluent TSS is 5 mg/L, the additional solids accumulation from removal of influent solids is an additional 0.75 kg TSS/m^3 for a total of 3.5 kg TSS/m^3. The backwash typically

Table 9–20

Backwash requirements for postanoxic attached growth denitrification filters[a]

Type of system	Backwash rate, m/h	Air scour rate, m/h	Approx. backwash water duration, min	Approx. backwash water, % of feed flow[b]
Downflow sunken media	18	90	15	4.7
Upflow sunken media	20	97	10	3.5
Upflow floating media	55	12	12	11.4
Continuous backwash	0.4	air lift	continuous	10

[a] Adapted from WERF (2010).

[b] Based on average hydraulic application rate of 4 m/h.

consists of an air scour followed by an air and water backwash. Typical backwash rates and fraction of product water use for backwashing are summarized in Table 9–20.

Phosphorus Limitations. Combined biological denitrification and chemical phosphorus removal has been used by several municipal WWTPs to meet effluent TN and TP concentrations of less than 3.0 and 0.10 mg/L, respectively. In such cases, upstream processes may be used before a postanoxic attached growth denitrification process to meet low effluent TN and TSS concentrations in addition to a low effluent TP concentration. The upstream process may be enhanced biological phosphorus removal, chemical treatment, or a combination of processes. However, it should be noted that the postanoxic denitrification process can become phosphorus-limited if the influent phosphorus concentration is too low to support the biogrowth needed for biological denitrification. During pilot testing of continuous backwash denitrification filters at the Hagerstown, MD, WWTP, the effluent NOx-N concentration remaining in the effluent stream increased when the influent PO_4-P/NOx-N ratio dropped below 0.02 g P/g N (deBarbadillo et al., 2006). Other investigators reported acceptable influent P/NOx-N ratios: 0.005 by (Scherrenberg et al., 2008); 0.01 to 0.02 (Husband and Becker, 2007); and 0.023 and 0.026 (Peric et al., 2009). Using a mechanistic modeling approach and data from various studies, Boltz et al. (2012) determined that biological denitrification performance declines greatly due to phosphorus limitation at an influent ratio of 0.009 g P/g NOx-N. One operating alternative used to overcome phosphorus limitations on postanoxic denitrification has been to dose phosphoric acid to the postanoxic process influent (WERF, 2010).

Operational Considerations for Postanoxic Attached Growth Denitrification

Key operational considerations for optimal performance of a postanoxic attached growth denitrification process are (1) backwash frequency and biofilm control, (2) control of the carbon dose, and (3) preventing phosphorus limitation.

Biofilm Control. Backwashing postanoxic denitrification filters, solids harvesting in an AnoxFBBR and scouring the media by mixing in an AnoxMBBR is a balancing act between minimizing costs associated with backwash/scour and maintaining the appropriate amount of biomass. Too frequent backwashing uses more energy and reduces the available biomass, which could have an effect on treatment performance. On the other hand, infrequent backwash/scour may result in thick biofilms that are diffusion-limited with poor removal performance and/or cause low hydraulic efficiency in the denitrification filters.

Carbon Dose Control. Overdosing of the carbon can lead to odor production from biological sulfate reduction and increase the effluent BOD concentration. Too low of a dose will result in higher effluent NOx-N concentrations and possibly an increase in effluent NO_2-N concentration. The rate of carbon addition can vary diurnally due to changes in the influent TKN concentration with time and to seasonal or wet weather loading variations. Manual operation or flow-paced chemical addition relies on an operator's experience with changes in the facility and finding the appropriate dose ratios. Feed forward control with online or manual measurements of nitrate-N or NOx-N in the upstream tank or feed stream to the postanoxic process provides a more effective approach to optimize performance and to reduce chemical wastage.

Preventing Phosphorus Limitation. As discussed above an insufficient influent phosphorus ratio to the postanoxic attached growth process can result in lowered performance. Modifications in the upstream treatment process and controls or adding phosphorus to the postanoxic process are options for assuring sufficient phosphorus for nitrogen removal. Online monitoring of upstream phosphorus concentrations is also used in such cases. The design of a denitrification filter is illustrated in Example 9–9.

EXAMPLE 9–9 **Denitrification Filter Design** Given the following flow and secondary effluent characteristics for feed to a 2 m deep downflow postanoxic denitrification filter, determine the following design parameters to achieve effluent TSS and NO_3-N concentrations of less than 5.0 and 1.0 mg/L, respectively. Assume a filtration rate of 4 m/h at average flow and a half-order nitrate removal kinetic coefficient of 0.27 mg/L·min at 15°C.

Design parameters to be determined:

1. Filter dimensions
2. Backwash water rate and percent of product water
3. Backwash air rate
4. Backwash water bump rate
5. Methanol requirements, kg/d
6. Daily solids production, kg/d
7. Backwash water TSS concentration, g/m^3

Wastewater characteristics:

Item	Unit	Value
Flowrate	m^3/d	8000
TSS	g/m^3	20
NO_3-N	g/m^3	25
Temperature	°C	15

Note: $g/m^3 = mg/L$.

Operational parameters and assumptions

1. Filter backwash frequency = 1/d
2. Water backwash flowrate/time = 18 $m^3/m^2 \cdot h$ for 15 min (see Table 9–20)
3. Air backwash rate = 90 $m^3/m^2 \cdot h$ (see Table 9–20)
4. Bump water flush rate/frequency = 12 m/h at once/3h
5. One filter required for standby

6. Net yield with methanol = 0.25 g VSS/g CODr
7. VSS/TSS = 0.85

Solution

1. Determine the dimensions of the denitrification filter.
   a. Determine filter size based on nitrogen loading (NL). Apply Eq. (9–55):

$$NL = \frac{-0.5k_{DN}(1.44\ NO_o)}{NO_e^{1/2} - NO_o^{1/2}}$$

$$NL = \frac{-0.5(0.27\ g/m^3{\cdot}min)(1.44)(25.0\ g/m^3)}{(1.0\ g/m^3)^{(1/2)} - (25.0\ g/m^3)^{(1/2)}} = 1.21\ kg/m^3{\cdot}d$$

$$NO_3\text{-N applied} = [(25.0)\ g/m^3](8000\ m^3/d)(1\ kg/10^3\ g)$$
$$= 200\ kg/d$$

$$\text{Volume} = \frac{(200\ kg/d)}{(1.21\ kg/m^3{\cdot}d)} = 165.3\ m^3$$

$$\text{Area} = V/D = 165.3\ m^3/2.0\ m$$
$$= 82.7\ m^2$$

   b. Determine filter size based on filtration hydraulic loading bases.

Hydraulic application rate = 4 m/h

Filtration rate = 8000 m^3/d = 333.33 m^3/h

Filter area = (333.33 m^3/h)/(4 m/h)

= 83.3 m^2

Thus, the filter size is controlled by the hydraulic application rate.

Assume 5 filters installed with 1 filter used as standby. Use of the filters should be rotated so that active biofilm is maintained in all of the filters.

$$\text{Area/filter} = \frac{83.3\ m^2}{4} = 20.8\ m^2$$

Use a square configuration.

Filter bed dimensions = 4.6 × 4.6 × 2.0 m

(check supplier standard modules)

2. Determine air and water backwash flowrates and percent of product water used as backwash water.
   a. Determine backwash air flowrate at an air backwash application rate of 90 $m^3/m^2{\cdot}h$.

Air flowrate = (90 $m^3/m^2{\cdot}h$)(20.8 m^2/filter) = 1870 m^3/h

   b. Determine backwash water flowrate at a water backwash application rate of 18 $m^3/m^2{\cdot}h$.

Water flowrate = (18 $m^3/m^2{\cdot}h$)(20.8 m^2/filter) = 374.4 m^3/h

= 6.24 m^3/min

   c. Determine volume of backwash water at 1 backwash/24 h for each filter with a duration of 15 min

Backwash water volume = (6.24 m^3/min)(15 min/filter)(4 filters)

= 374.4 m^3d

d. Determine percent of product water used as backwash water.

$(374.4\ \text{m}^3/\text{d})(8000\ \text{m}^3/\text{d}) = 0.0468$

$= 4.68\%$

3. Determine volume of water used to flush (bump) filters at 1 bump/3 h with a duration of 4 min/bump and a flowrate of 12 m/h.

Total bump water volume

$= (83.2\ \text{m}^2)(12\ \text{m/h})(3\ \text{min/bump})(1\ \text{bump}/3\ \text{h}\cdot\text{filter})(24\ \text{h/d})(\text{h}/60\ \text{min})$

$= 399.4\ \text{m}^3/\text{d}$

Percent of product water used for nitrogen release bumping:

$= (399.4\ \text{m}^3/\text{d})/(8000\ \text{m}^3/\text{d}) = 0.0499$

$= 5.0\%$

4. Determine the total product water used as percent of influent treated.

Percent backwash plus bump water $= 4.68 + 5.0 = 9.7\%$

Actual filtration rate required including product water used for backwashing and bumping:

$(4\ \text{m/h})(1.097) = 4.39\ \text{m/h} = 105\ \text{m}^3/\text{m}^2\cdot\text{d}$

Note: An iterative calculation can be performed to develop an actual filtration rate approaching 4 m/h by using a lower rate in step 1b.

5. Methanol (CH_3OH) requirements.

Nitrate removal $= (25.0\ \text{g/m}^3 - 1.0\ \text{g/m}^3)(8000\ \text{m}^3/\text{d}) = 192\ \text{kg/d}$

Calculate methanol dose from Eq. (8–69)

$$C_{R,\text{NO}_3} = \frac{2.86}{1 - 1.42(Y_\text{H})} = \frac{2.86}{1 - 1.42(0.25\ \text{g VSS/g COD})}$$

$$= 4.43\ \text{g methanol COD/g NO}_3\text{-N}$$

add 10% for DO and NO_2-N in influent

Methanol dose $= (1.1)(4.43) = 4.88$ g methanol COD/g NO_3-N

$$\text{As methanol} = \frac{(4.88\ \text{g methanol COD/g NO}_3\text{-N})}{(1.5\ \text{g COD/g methanol})} = 3.25\ \text{g methanol/g NO}_3\text{-N}$$

Methanol $= (3.25\ \text{kg/kg})(192\ \text{kg/d}) = 624.0\ \text{kg/d}$

6. Determine the solids production.

Solids = filtered solids + biomass production

Use effluent TSS = 5 mg/L (g/m^3)(given value)

Filter solids $= [(20 - 5)\ \text{g/m}^3](8000\ \text{m}^3/\text{d})(1\ \text{kg}/10^3\ \text{g})$

$= 120\ \text{kg/d}$

Biomass production: based on 0.25 g VSS/g methanol COD (given)

Biomass produced =

$$P_{X,\text{bio}} = \frac{\left(\dfrac{0.25\text{ g VSS}}{\text{g COD}}\right)\left(\dfrac{1.5\text{ g COD}}{\text{g CH}_3\text{OH}}\right)(624\text{ kg CH}_3\text{OH/d})}{(0.85\text{ g VSS/g TSS})} = 275.3\text{ kg TSS/d}$$

Total solids = 120 kg TSS/d + 275.3 kg TSS/d = 395.3 kg TSS/d

Volume per filter = $(20.8\text{ m}^2)(2\text{ m}) = 41.6\text{ m}^3$

$$\text{Solids Storage/24 h} = \left(\frac{395.3\text{ kg}}{\text{d}}\right)\left(\frac{1}{4\text{ filters}}\right)\left[\frac{\text{filter}}{(41.6\text{ m}^3\text{/d})}\right] = 2.4\text{ kg TSS/m}^3$$

Comment Because the solids storage value is well below 4.0 kg TSS/m^3, backwashing may be required only once every 1.5 d. However, to maintain the health of the filter, backwashing once per day is recommended (see Sec. 11–4 in Chap. 11).

9–8 EMERGING BIOFILM PROCESSES

There are several emerging biofilm reactors that are briefly described below, including (1) membrane biofilm reactors, (2) biofilm airlift reactors, and (3) aerobic granule process.

Membrane Biofilm Reactors

Membrane biofilm reactors (MBfR) refers to processes that use membrane to deliver gaseous substrate such as oxygen and hydrogen and as biofilm growth support (Timberlake et al., 1988, Brindle and Stephenson, 1996; Lee and Rittmann, 2000; Syron and Casey, 2008). Microporous, hydrophobic materials and hollow-fiber membranes are typically used in MBfRs because of the high specific surface areas (up to 5000 m^2/m^3). The hollow-fibers are collected into bundles with one end as gas-supplying manifold and the other end sealed. Pressurized gas (i.e. hydrogen or oxygen) diffuses through the dry membrane pores and into the biofilm on the other side of the membrane. This mode of operation leads to counter-diffusion, where one substrate (electron donor or acceptor) diffuses into the biofilm from the dry membrane and the other one diffuses through the bulk liquid into the biofilm.

The advantage of an MBfR is the greater fluxes possible by supplying the gaseous electron donor or electron acceptor directly to the biofilm without traversing a liquid layer. However, this advantage can be lost due to biofilm diffusion from the back to the liquid side where substrate is most consumed if the biofilm is too thick. Therefore, control of biofilm accumulation is important in MBfRs. Hydrogen-based MBfRs have been investigated for water treatment such as denitrification and, for remediation processes that use hydrogen as electron donor for reduction such as for perchlorate, trichloroethane, selenite, arsenate etc. (Ergas and Reuss, 2001; Chung et al., 2006a; Chung et al., 2006b; Nerenberg and Rittmann 2004; Chung and Rittmann, 2007).

Oxygen-based MBfR has been demonstrated for concurrent carbon removal, nitrification and denitrification (Timberlake et al., 1988; Suzuki et al., 1993; Brindle et al., 1998; Schramm et al., 2000). The oxygen concentration gradient, which is higher near the membrane and lower closer to the bulk liquid, leads to nitrification in the inner portion of the biofilm and BOD removal and denitrification activities in the outer portion of the biofilm. A study by Downing and Nerenberg (2007) showed the possibility of using MBfR in a hybrid system that incorporates hollow-fibers into the activated sludge process. MBfR applications have been evaluated in lab and pilot testing and challenges remain for full-scale configurations and process control.

Biofilm Airlift Reactors

Biofilm airlift reactors were developed in the Netherlands in the late 1980s and contain a center column air lift zone and small-sized sand media which supports a very high effective volumetric biomass concentration. The concept can be used for both aerobic and anaerobic processes. For aerobic processes, the airlift in the center column provides oxygen and recirculation of the biofilm and media. It has been used for aerobic wastewater treatment for BOD removal, nitrification and denitrification (Heijnen et al., 1993; Frijters et al., 2000; Nicolella et al., 2000). The commercial CIRCOX process is based on this type of reactor. The CIRCOX process has a high loading capacity (4 to 10 kg COD/m^3·d), short HRTs (0.5 to 4 h), high biomass settling velocities (50 m/h), and high biomass concentrations (15 to 30 g/L) (Frijters et al., 2000; Nicolella et al., 2000).

Aerobic Granules Reactor

Aerobic reactors with dense granular biomass particles can be considered with attached growth processes because the granular biomass behaves as a biofilm with substrate removal being diffusion limited. A number of aerobic granular biomass reactor operations have been reported (Liu and Tay 2002; Morgenroth et al., 1997; Beun et al., 2002; de Kreuk and van Loosdrecht, 2006, Adav et al., 2008). The advantage of aerobic granular biomass reactors is a high treatment capacity with a much smaller footprint than an activated sludge system. Post treatment is normally required to obtain low effluent suspended solids. A combination of an aerobic granular biomass process with membrane separation has been proposed to achieve high effluent quality (Wang et al., 2008). Aerobic granular processes have been investigated for simultaneous removal of organics, nitrogen and phosphorus in wastewater treatment as shown in Table 8–27, and for industrial applications (de Kreuk et al., 2005; Yilmaz et al., 2008; Schwarzenbeck et al., 2005).

PROBLEMS AND DISCUSSION TOPICS

9–1 A 20-m diameter plastic packing trickling filter, containing cross-flow plastic packing at a 6.1-m depth with a specific surface area of 100 m^2/m^3, receives domestic wastewater after primary treatment. The average flowrate is 390, 440, or 490 m^3/h (value to be selected by instructor) and the BOD concentration is 150 mg/L. Determine and compare the effluent BOD concentration and percent BOD removal at 20°C and 15°C. Assume an n value of 0.5 and recirculation ratio equal to zero.

9–2 Two 15-m diameter trickling filters containing conventional cross-flow plastic packing at a 6-m depth are used to treat a pharmaceutical wastewater at an average flowrate of 2120 m^3/d. The influent BOD concentration is 600, 900, or 1200 mg/L (value to be selected by instructor) and the temperature is 20°C. Each tower has a 2-arm distributor. Determine: (a) the operating dose and flushing dose from Table 9–3 and distributor speeds in revolutions/min for each case; and (b) the recirculation ratio and total pumping rate in m^3/h to each filter.

9–3 For Example 9–3, using the same design criteria given for the trickling towers, influent BOD concentration and temperature, and effluent BOD concentration, determine the following using a plastic packing depth of 4.0, 5.0, or 7.0 m (value to be selected by the instructor):

a. Volume of plastic packing, m^3

b. Hydraulic loading rate in L/m^2·s

c. Volumetric BOD loading in kg/m^3·d

Compare the results to the values determined for the same design parameters in the example.

9–4 The following data were obtained for a pilot plant study involving the treatment of a combined domestic-industrial wastewater with a tower trickling filter filled with plastic packing at a depth of 6.1 m. The diameter of the pilot plant tower was 1 m and the specific surface area of the plastic packing was 90 m^2/m^3. The wastewater temperature at the time of the test was 12, 18, or 24°C (value to be determined by the instructor). During the testing the influent BOD concentration was 350 mg/L. The following table summarizes the average BOD removal efficiency at different flowrates. Using these data, determine the value of the wastewater treatability coefficient, *k* (assume the value of *n* is 0.50), at the test temperature and at 20°C.

Pilot-plant test results:

Flowrate, m^3/d	BOD removal efficiency, %
6	88
12	82
18	67
24	63
48	54

9–5 The following design information is given for a domestic wastewater (wastewater 1, 2, or 3 to be selected by instructor):

		Wastewater		
Design information	Unit	1	2	3
Flowrate	m^3/d	10,000	10,000	10,000
BOD	mg/L	270	300	220
TSS	mg/L	240	280	210
Minimum temp.	°C	15	12	15

Using the above information, and assuming 30 percent BOD removal in primary clarification, a packing depth of 6.1 m, a packing specific surface area of 100 m^2/m^3, and a value for *n* of 0.50, determine the following design parameters for a trickling filter treatment system.

a. Primary and secondary clarifier diameters, m

b. Trickling filter tower diameter, m

c. Packing volume, m^3

d. Recirculation ratio, if required

e. Total pumping rate, m^3/h

f. Flushing and normal dose rates, mm/pass

9–6 Determine the diameters and packing volume of two 4.0-m deep plastic-media filters, operated in series, for a domestic wastewater treatment application with the following characteristics and requirements. Assume a plastic media vertical-flow packing from Table 9–2 with a specific surface area of 102 m^2/m^3.

Parameter	Unit	Value
Flowrate	m^3/d	5000
Influent BOD	mg/L	
Wastewater 1		220
Wastewater 2		200
Wastewater 3		180

(*continued*)

(*Continued*)

Parameter	Unit	Value
Temperature	°C	14
Primary clarifier BOD removal efficiency	%	35
Trickling filter average effluent BOD	mg/L	20

Note: Wastewater 1, 2 or 3 to be selected by instructor.

9–7 An 18-m diameter trickling filter tower containing conventional plastic packing at a 6.1-m depth and a packing specific surface area of 100 m^2/m^3 treats a primary effluent flowrate of 7600 m^3/d. The primary effluent BOD is 100, 120, or 150 mg/L (value to be selected by instructor) and the wastewater temperature is 18°C. The air temperature varies from 2 to 23°C. The BOD loading peaking factor is 1.5. Assume a factor of 1.5 with Eq. (9–10) to account for the inlet and outlet pressure losses for the tower. Determine the following:

a. The required oxygen supply rate, kg/h

b. The required air flowrate at the warmest temperature, m^3/min

c. The airflow pressure drop across the packing, Pa

9–8 Two 20-m diameter plastic tower trickling filters containing 6.1 m of conventional plastic media (100 m^2 area/m^3 volume) receive a primary clarifier effluent at an average flowrate of 11,200 m^3/d. The TKN concentration is 24 mg/L and the BOD concentration is 150, 130, or 120 mg/L (value to be selected by the instructor). The temperature is 18°C. Evaluate the trickling filter BOD loading and determine the nitrogen removal efficiency due to nitrification.

9–9 Compare the design of a trickling filter alternative to the activated sludge designs provided in Example 8–3 for (a) BOD removal, (b) combined BOD removal and nitrification, and/or (c) BOD removal followed by tertiary nitrification (to be selected by the instructor). Use the same wastewater characteristics and temperature. Prepare a summary table comparing (1) the total volume of the biological process units, (2) the area used for the biological process units, (3) the secondary clarifier area, and (4) the monthly energy requirement in kW for the secondary treatment processes. Assume the following:

a. Trickling filter packing depth equals 6.1 m

b. Trickling filter packing specific surface area equals 100 m^2/m^3 for BOD removal and combined BOD removal/nitrification and equals 138 m^2/m^3 for tertiary nitrification

c. Trickling filter effluent BOD equals 25 mg/L

d. Trickling filter effluent NH_4-N equals 1.0 mg/L

e. The trickling filter recirculation ratio equals 0.50

f. The amount of the influent TKN used for heterotrophic bacteria growth in the trickling filter is 8.0 mg/L

g. The aeration tank depth is 5 m

h. The blower energy requirement for activated sludge aeration, considering the piping headlosses and diffuser submergence depth is 1.80 kW per m^3/min air flowrate.

i. The tricking filter feed flow pumping energy requirement is 1.58 kW per 1000 m^3/d.

9–10 Prepare a table to compare the advantages and limitations of a tower trickling filter with plastic packing versus activated sludge treatment in terms of space, ease of process operation, sludge settling characteristics, energy requirements, maintenance, treatment flexibility, nitrification reliability, potential odors, and potential for future nitrogen or phosphorus removal.

9–11 A 20-m diameter trickling filter tower containing a 5-m depth of high-density plastic packing (138 m^2 area/m^3 packing volume) is used in a tertiary nitrification application. The influent flowrate is 37,000, 39,000, or 41,000 m^3/d (value to be selected by the instructor) and the NH_4-N concentration is 20 mg/L. Assuming that the nitrification $J_{N,\max}$ value is 1.8 $g/m^2 \cdot d$ from information in Table 9–7 and the K_N value is 1.5 g/m^3, compare the trickling filter effluent NH_4-N concentration (mg/L) with no recycle and with a 100 percent recirculation ratio. Why is no gravity settling tank used for the trickling filter effluent?

9–12 A TF/AS process is used to treat a combined domestic and industrial wastewater after primary clarification, and the primary effluent wastewater characteristics are given below. For the following design parameters, compare the effect of designing the plastic tower trickling filter step for 40 percent versus 80 percent BOD removal:

a. Trickling filter diameter, m, and hydraulic loading rate, $L/m^2 \cdot s$

b. Oxygen required in the activated sludge aeration tank, kg/d

c. The amount of solids wasted per day, kg/d

d. The volume, m^3, and hydraulic retention time, h, of the aeration tank

Use the following assumptions for the trickling filter and activated-sludge designs.

Trickling filter:
Plastic packing treatability coefficient, $k_{20} = 0.18(L/s)^{0.5}/m^2$
Packing depth = 6.1 m
Number of towers = 2

50 percent of theoretical effluent BOD is soluble.
Activated sludge:
SRT = 5.0 d (no nitrification)
MLSS = 3000 mg/L
Biomass yield, Y, = 0.6 g VSS/g BOD removed
Endogenous decay, b = 0.12 g VSS/g VSS·d
UBOD/BOD = 1.6

Wastewater characteristics:

Item	Unit	Value
Flowrate	m^3/d	8000
BOD	mg/L	
Wastewater 1		400
Wastewater 2		500
Wastewater 3		600
sBOD	mg/L	60% of BOD
TSS	mg/L	65
VSS	mg/L	55
nbVSS	mg/L	22
Temperature	°C	12

Note: Wastewater 1, 2 or 3 to be selected by instructor.

Which design is preferred? State your reasons.

9–13 An existing activated-sludge facility with primary treatment is operated at an SRT of 18 d and minimal temperature of 12°C so that complete nitrification is maintained. The system is operated with an MLSS concentration of 2200 mg/L, and the SVI is generally in the range of 180 to 200 mL/g. At these conditions, the treatment flowrate is 8000 m^3/d. The city engineer requests that your firm design a plastic tower pretreatment system for about 60 percent BOD removal before the activated sludge unit. You are to list for the city engineer the potential impacts of converting the existing activated sludge process to a TF/AS process, including flow capacity, solids production, oxygen requirements, energy demand, sludge settling characteristics, and effluent NH_4-N and BOD concentrations. Provide an explanation of the basis of your opinion on the potential impacts.

9–14 An industrial waste is to be treated with a tower trickling filter followed by an activated-sludge process (TF/AS process). The wastewater flowrate is 20,000 m^3/d and is equalized. Primary settling is not used because the wastewater contains mainly soluble organic substances. The tower trickling filter contains conventional plastic packing with a specific surface area of 100 m^2/m^3 and the operational SRT for the activated sludge process is to be 5 d during the critical summer period and a maximum of 15 d during the winter. The lowest average sustained winter temperature (at least two weeks) is 5°C and the highest average sustained summer temperature is 26°C. The characteristics of the industrial waste, data derived from pilot plant studies, and related design data are presented below. Using these data, size the units and determine the following:

a. Concentration of mixed liquor suspended solids to be maintained during summer and winter operation, mg/L

b. Recycle flowrates around the filter and activated-sludge process, m^3/d

c. Quantity of sludge to be disposed daily, kg/d

d. Effluent BOD concentration from the trickling and activated-sludge processes, mg/L

e. Quantity of nutrients that must be added daily, kg/d

Wastewater characteristics:
BOD = 1200, 1500, or 1800 mg/L (value to be selected by instructor)
TSS = 100 mg/L
VSS = 0 mg/L
Total nitrogen as N = 10 mg/L
Total phosphorus as P = 4 mg/L

Trickling filter pilot plant results:
$k_{20^\circ C} = 0.075\ (L/s)0.5/m^2$
Net solids yield = 0.5 g VSS/g BOD removal
Temperature correction value, $\theta = 1.06$

Activated sludge pilot plant results:
Solids synthesis yield, $Y_H = 0.6$ g VSS/g BOD removal
Endogenous decay, $b_H = 0.12$ g VSS/g VSS·d
$k = 6.0$ g BOD/g VSS·d
$K_s = 90$ mg BOD/L
$\theta = 1.035$

Design parameters:
Trickling filter hydraulic application rate = 0.10 m^3/m^2·min

9–15 For the IFAS process analysis in Example 9–7, determine the effect of having a DO concentration of 3.0, 5.0 or 6.0 mg/L (to be selected by the instructor) on the amount of plastic carrier media needed, the V_M/V ratio needed, and the media fill volume fraction needed in the aeration tank with the plastic carrier addition.

9–16 The following biological treatment processes are being considered for a decentralized wastewater treatment plant to treat a domestic wastewater with an average daily flowrate of 1000 m^3/d.

The present requirement is to discharge the treated effluent to a nearby surface water and the effluent BOD, TSS, and NH_4-N concentrations must be less than 25, 25, and 1.0 mg/L, respectively. The plant site available has limited space and it is close to a residential area. Prepare a table that compares the advantages and disadvantages of the following three processes of interest: (a) activated sludge with conventional secondary clarification, (b) membrane bioreactor, and (c) moving bed bioreactor. Include among the criteria considerations for space, operations, and energy requirements. Which process would you recommend and why?

9–17 Use the same wastewater characteristics and effluent goals as used for Example 9–8 and design an MBBR system with only a single stage for BOD removal followed by only a single stage for nitrification. Use a DO concentration of 2.0 mg/L in the BOD removal zone and 4.0 mg/L in the nitrification zone. Compare the design to that for Example 9–8 in terms of media volume and tank volume requirements and determine the sludge production rate and oxygen required. Assume a 50 percent carrier media fill fraction. Assume a biofilm biomass effective SRT of 4.0 d for BOD removal and 8.0 d for nitrifying bacteria.

9–18 Evaluate the effect of the media fill fraction and nitrification zone DO concentration on the amount of carrier media required and tank volumes needed for Example 9–8. Using the given data (a) compare the effect of using a DO of 3.0, 5.0, or 6.0 mg/L (to be determined by instructor) for the nitrification zones with the same carrier media fill fraction used in the example problem and (b) use the same DO concentrations as used in the example problem and compare the effect of using a carrier media fill fraction of 65 percent.

9–19 Perform an MBBR design at 18°C using the same influent wastewater conditions and assumptions as given Example 9–8. Provide a table to compare the designs at 12 and 18°C.

9–20 Design an upflow attached growth BAF process for BOD removal only (assume a floating media process) to achieve an effluent BOD concentration of 20 mg/L or less. Use the wastewater characteristics and average flowrate given for Example 9–8 and a soluble BOD concentration of 80 mg/L. Assume an unexpanded media depth of 2 m, and hydraulic application rate of 6.0 m/h. Design the system with 4 operating units and one standby. Provide a design summary table with (a) the reactor volume in m^3 and equivalent hydraulic retention time in h, (b) the dimensions of each BAF reactor assuming square configuration, (c) the oxygen required in kg/d, (d) the air application rate per treatment unit in m^3/min, (d) the sludge production rate in kg TSS/d, (f) the amount of backwash water used daily per unit in m^3/d, and (g) the backwash water suspended solids concentration in mg/L. Use the following assumptions:

BOD loading equals 3.5 kg BOD/m^3·d from Table 9–17

Actual oxygen transfer efficiency equals 6.0 percent from Table 9–18

Fraction of influent VSS destroyed equals 0.25

A backwash water time of 15 min/d and average flush rate of 40 m/h

9–21 A secondary effluent has the following characteristics:

Parameter	Unit	Value
Flowrate	m^3/d	5000
TSS	mg/L	15
NO_3-N	mg/L	30
Temperature	°C	18

Design a downflow denitrification filter with a sand media depth of 1.6 m to reduce the NO_3-N to an effluent concentration of 3.0, 2.0, or 1.0 mg/L (value to be selected by instructor). Assume a hydraulic application rate of 4.0 m/h and determine the following:

a. Denitrification filter media volume, m^3

b. Number of filters and filter dimensions, assuming square tanks with a maximum dimension of 10 m by 10 m

c. Methanol dose in mg/L and kg/d

d. Amount of solids produced, kg/d

The following assumptions apply:
Methanol biomass synthesis yield equals 0.25 g VSS/g CODr.
The half-order nitrate removal kinetic coefficient equals 0.30 mg/L·min.
Effluent TSS concentration equals 5.0 mg/L

9–22 For Example 9–9 assume that a glycerol type feed is used instead of methanol for the external carbon source for the same denitrification filter design produced in the example. Determine the effect of the different carbon source on (a) the effluent NO_3-N concentration, (b) the carbon dose requirement in kg/d as glycerol and as COD, (c) the solids production rate in kg/d, and (d) the solids storage in kg/m^3 between a once per day backwash event. Is once per day backwashing acceptable?

Use the following assumptions:
The glycerol feed results in higher denitrification rates than methanol and thus the half-order nitrate removal kinetic coefficient is estimate at 0.40 mg/L·min.
The synthesis yield is higher at 0.36 g VSS/g CODr.
The COD to glycerol feed mixture ratio is 1.4 g COD/g glycerol feed.

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10 Anaerobic Suspended and Attached Growth Biological Treatment Processes

WORKING TERMINOLOGY

Term	Definition
Ammonia toxicity	Free ammonia (NH_3), at high enough concentrations, is considered toxic to acetoclastic methanogenic organisms.
Anaerobic expanded bed process	An anaerobic upflow process in which the medium, usually silica sand, is expanded but not fluidized by the upward liquid velocity.
Anaerobic fluidized bed process	An anaerobic upflow process which operates in a mode similar to the expanded bed, but the medium is fully fluidized by the upward velocity of the fluid.
Anaerobic granular sludge	Dense 0.50 to 4.0 mm particles in anaerobic upflow reactors containing fermentation, hydrogenotrophic, and methanogenic organisms in close proximity.
Anaerobic sequencing batch reactor	An anaerobic suspended growth process with reaction and solids-liquid separation in the same vessel, much like that for aerobic sequencing batch reactors (SBR).
Anaerobic sludge blanket processes	Influent wastewater is distributed at the bottom of an anaerobic reactor and travels in an upflow mode through a sludge blanket zone that typically contains dense granular biomass particles.
Anaerobic suspended growth processes	A mixed anaerobic reactor containing a suspension of fermentation and methanogenic organisms and feed particulate matter.
Anaerobic treatment processes	Any of a number of biological treatment processes carried out in the absence of oxygen.
Attached growth anaerobic process	Anaerobic treatment processes in which the biomass responsible for treatment is attached to some type of medium. The medium may be fixed, expanded or fluidized. Where fixed medium is used, the flow can either be downflow or upflow.
Covered anaerobic lagoon process	Covered liner earthen lagoon typically used for high-strength industrial wastewaters, such as meat processing wastewaters.
Expanded granular sludge blanket process (EGSB)	An anaerobic sludge blanket process with higher upflow velocities than the UASB process.
Hydrogen sulfide (H_2S)	A malodorous gas toxic gas formed under anaerobic conditions. Sulfur containing compounds in wastewater serve as electron acceptors for sulfate-reducing bacteria, which consume organic compounds and produce hydrogen sulfide (H_2S).
Membrane separation anaerobic treatment process	A suspended growth anaerobic process in which a synthetic membrane is used to separate the treated wastewater from the solids to achieve complete effluent suspended solids removal.
Methane (CH_4)	Methane and carbon dioxide along with cell biomass are the principal carbon end products of anaerobic conversion processes.
Organic loading rate (OLR)	The mass rate of organic substrate (COD) addition per unit volume of an anaerobic reactor.
Solids retention time (SRT)	The average period of time in which solids remain in a biological reactor.
Upflow anaerobic sludge blanket (UASB) reactor	The name of the first and most common of the anaerobic sludge blanket processes (see above).

Anaerobic biological reactions involve specialized bacteria and archaea that use a variety of electron acceptors in the absence of molecular oxygen for energy production. They are used in a number of different anaerobic processes in wastewater treatment. These include processes for nitrate/nitrite reduction to nitrogen gases, fermentation processes to produce volatile fatty acids for use in enhanced biological phosphorus removal, anaerobic contacting for acetate and propionate uptake in enhanced biological phosphorus removal, anaerobic oxidation of organic compounds in municipal and industrial wastewaters, anaerobic digestion of waste sludge, and anaerobic digestion of other organic wastes.

Anaerobic wastewater treatment processes include suspended growth, upflow and downflow attached growth, fluidized bed attached growth, upflow sludge blanket, lagoon, suspended growth with membrane separation, and many other proprietary processes. The purpose of this chapter is to describe and present typical design loadings and treatment process capabilities for the principal anaerobic processes, excluding conventional anaerobic digestion processes which are considered in Chap. 13. Before considering the individual anaerobic treatment processes, it will be helpful to consider the rational for the use of anaerobic treatment processes, the evolution of anaerobic treatment technologies, a brief review of the principal public and commercial processes, and general considerations for the application of anaerobic treatment processes.

10–1 THE RATIONALE FOR ANAEROBIC TREATMENT

The rationale for and interest in the use of anaerobic treatment processes can be explained by considering the advantages and disadvantages of these processes. The principal advantages and disadvantages of anaerobic treatment are listed in Table 10–1 and are discussed below.

Advantages of Anaerobic Treatment Processes

Anaerobic treatment processes have been used as an alternative to aerobic treatment for applications varying from low to extremely high strength wastes. Of the advantages cited in Table 10–1, energy considerations, lower biomass yield, less nutrients required, higher volumetric loadings, and effective pretreatment are examined briefly in the following discussion. The subjects are also considered in Chap. 13, 14, and 17.

Energy Considerations. Anaerobic processes may be net energy producers instead of energy users as is the case for aerobic processes. The potential net energy production that can be achieved with anaerobic treatment depends on the strength of the wastewater, the operating temperature, and whether energy recovery is practiced. An energy balance comparison between anaerobic and aerobic treatment for various wastewater strengths is presented in Sec. 10–4.

Table 10–1

Advantages and disadvantages of anaerobic processes compared to aerobic processes

Advantages	Disadvantages
1. Less energy required	1. Longer startup time to develop necessary biomass inventory
2. Less biological sludge production	2. May require alkalinity addition
3. Less nutrients required	3. May require further treatment with an aerobic treatment process to meet discharge requirements
4. Methane production, a potential energy source	4. Biological nitrogen and phosphorus removal is not possible
5. Smaller reactor volume required	5. Much more sensitive to the negative effect of lower temperatures on reaction rates
6. Elimination of off-gas air pollution	6. May be more susceptible to upsets due to toxic substances or wide feeding changes
7. Able to respond quickly to substrate addition after long periods without feeding	7. Potential for odor production and corrosiveness of gas
8. Effective pretreatment process	
9. Potential for lower carbon footprint	

Lower Biomass Yield. Because the energetics of anaerobic processes result in lower biomass production by a factor of about 6 to 8 times, sludge processing and disposal costs are greatly reduced. Given the major environmental and monetary issues associated with the use or disposal of biomass produced from aerobic processes as discussed in Chap. 14, the fact that less sludge is produced in anaerobic treatment is a significant advantage for anaerobic treatment.

Less Nutrients Required. Because many industrial wastewaters lack sufficient nutrients, the lower cost for nutrient addition due to less biomass production from anaerobic treatment is a clear benefit.

Higher Volumetric Organic Loadings. Anaerobic processes generally have higher volumetric organic loading rates than aerobic processes so that smaller reactor volumes and less space may be required for treatment. Organic loading rates of 3.2 to 32 kg COD/m^3·d may be used for anaerobic processes, compared to 0.5 to 3.2 kg COD/ m^3·d for aerobic processes (Speece, 1996).

Effective Pretreatment Process. Often, anaerobic treatment processes are used in combination with aerobic treatment processes to achieve specific treatment goals. A common application in the wastewater field is the treatment of high strength waste before discharge to a municipal wastewater treatment facility. A typical example is shown on Fig. 10–1. Gas, recovered from the covered anaerobic section of the large anaerobic lagoon is used to produce electricity.

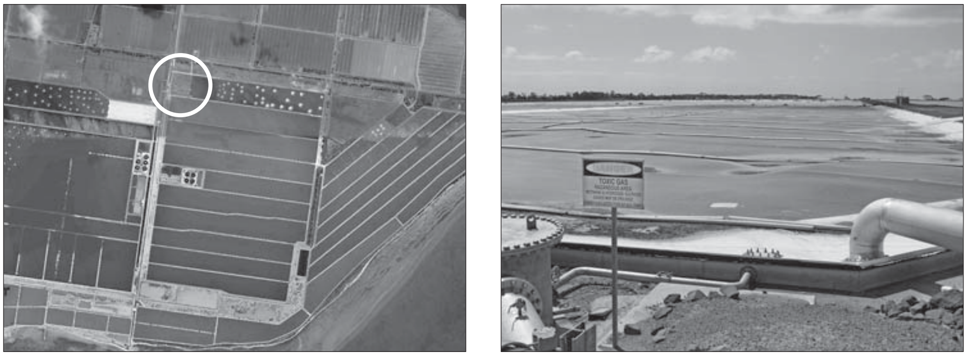
Disadvantages of Anaerobic Treatment Processes

Potential disadvantages also exist for anaerobic processes as reported in Table 10–1. Operational considerations, the need for alkalinity addition, and the need for further treatment are highlighted further in the following discussion.

Operational Considerations. The major concerns with anaerobic processes is their longer start-up time (months for a anaerobic versus days for aerobic processes), their sensitivity to possible toxic compounds, operational stability, the potential for odor production, and corrosiveness of the digester gas. However with proper wastewater characterization and process design these problems can be avoided and/or managed. Operational process knowledge and skill is also needed to maintain anaerobic process stability by

Figure 10–1

Combined anaerobic lagoon/aerobic pond wastewater treatment system: (a) aerial view of large treatment ponds with covered anaerobic pretreatment lagoon, shown in circled area. The white dots are the plumes from large turbine type floating aerators (coordinates 37.9788 S, 144.6417 E). (b) View from inlet with gas recovery facilities in the foreground, floating membrane cover, and large surface floating aerators in the background.



(a) (b)

proper control of the feed, temperature, and pH to maintain a balance between volatile fatty acid production by the acidogens and the capacity of the methanogenic organisms.

Need For Alkalinity Addition. The most significant negative factor that can affect the economics of anaerobic versus aerobic treatment is the possible need to add alkalinity. Alkalinity concentrations of at least 2000 to 3000 mg/L as $CaCO_3$ may be needed in anaerobic processes to maintain an acceptable pH with the characteristic high gas phase CO_2 concentration. If this amount of alkalinity is not in the influent wastewater or cannot be produced by the degradation of proteins and amino acid, a significant cost may be incurred to purchase alkalinity, which can affect the overall economics of the process.

Need For Further Treatment. Anaerobic wastewater treatment processes may have effluent BOD concentrations ranging from 50 to 150 mg/L due to the presence of residual volatile fatty acids and dispersed solids. The process can be followed by an aerobic process for effluent polishing to utilize the benefits of both processes. Series reactors of anaerobic-aerobic processes have been shown feasible for treating municipal wastewaters in both temperate and warmer climates resulting in less energy needs and less sludge production (Lew et al., 2003; Chong et al., 2012). More recently, as discussed subsequently, a number of integrated combined single tank anaerobic-aerobic reactors have also been developed.

Summary Assessment

In general, for municipal wastewaters with lower concentrations of degradable COD, lower temperatures, higher effluent quality needs, and nutrient removal requirements, aerobic processes are favored at present. For industrial wastewaters with much higher degradable COD concentrations and elevated temperatures, anaerobic processes may be more economical. In the future, with further developments in anaerobic treatment processes, it is anticipated that their use will become more widespread in a variety of applications, because of the overwhelming advantages from energy savings and less sludge production.

10–2 DEVELOPMENT OF ANAEROBIC TECHNOLOGIES

The earliest engineered anaerobic technologies were designed for and applied to the treatment of wastewater. At the time of their development in the late 1800s and early 1900s, a community's wastewater was an unhealthy combination of untreated sanitary wastes, animal manure, and various local industrial discharges. The historical development of the early anaerobic technologies, the application of anaerobic treatment for sludges, use of anaerobic treatment for high strength wastes, and some thoughts on the future are considered in this section. The types of anaerobic technologies currently available are considered in the following section.

Historical Developments in Liquefaction

Some of the noteworthy developments in the early stages of anaerobic treatment technology are summarized in Table 10–2. Schematic diagrams of some of the early treatment process developments are illustrated on Fig. 10–2. The early developments in anaerobic treatment were focused on liquefying the solids in wastewater to reduce or eliminate the need for sludge management and make the effluent suitable for subsequent treatment and or reuse for irrigation. The automatic scavenger developed by Mouras in France and patented in the 1880s [see Fig. 10–2(a)] was perhaps the first purposeful attempt to liquefy

Table 10–2

Important milestones in the development of anaerobic technologies[a]

Period	Event
Early developments	
1852	Henry Austin in England designed and built a tank that allows accumulation of solid matter layer on the bottom and scum layer on the top, with liquid draw off between layers. Closely resembles modern septic tanks (Kinnicutt et al., 1913).
1881	Jean-Louis Mouras obtained a French patent for an "automatic scavenger—an automatic and odorless cesspit" to treat wastewater anaerobically, a development that became one of the earliest known domestic wastewater treatment systems [see Fig. 10–1(a)]. Perhaps the first purposeful attempt to liquefy sludge. (Moigno, 1881,1882; Kinnicutt et al., 1913).
1887	The first Dortmund tank was designed and built by Kniebuhler in Germany. The advantage of the Dortmund tank was that sludge could be removed without stopping the flow. Dortmund tanks are still being built today [see Fig. 10–1(b)] (Kinnicutt et al., 1913).
1887	Lawrence Experimental Station established on the bank of the Merrimac River at Lawrence, MA. In 1890, the first report of the work at the Lawrence Station was published, which Winslow considered the "the most important single document in the history of sewage treatment" (Winslow, 1938).
1887	A sand bed, upflow anaerobic filter was constructed by the Lawrence Experimental Station to treat domestic wastewater and was operated for about 14 y (McCarty, 2001).
1891	Scott-Moncrieff of England constructed a two-tiered tank (empty lower level sludge compartment and upper level upflow anaerobic rock filter) to treat domestic wastewater anaerobically followed by an aerobic coke tray trickling filter [see Fig. 10–1(c)] (Kinnicutt et al., 1913). The complete treatment system comprised of the anaerobic rock filter and aerobic treatment unit is perhaps the first hybrid system ever built (McCarty, 2001).
Developments in the treatment of wastewater sludges	
1895	In Exeter, England, Donald Cameron installed a water-tight, covered basin to treat wastewater anaerobically, naming the device a "septic tank" [see Fig. 10–1(d)] (Kinnicutt et al., 1913).
1899	H. W. Clark at the Lawrence Experimental Station, noted that sludge should be fermented in a separate tank. Sludge lagoons had been used up to that time and later (Imhoff, 1938; Winslow, 1938).
1904	Travis developed a two story tank for the liquefaction of sludge [see Fig. 10–1(e)]. In the original design about one sixth of the flow passed through the lower chamber) (Kinnicutt et al., 1913). In one configuration effluent was directed to anaerobic filters [see Fig. 10–1(f)].
1906	Dr. Karl Imhoff of Germany patented a wastewater treatment device (the Imhoff tank) that anaerobically treated sewage while separating solids from the liquid phase prior to discharge [see Fig. 10–1(g)]. The Imhoff tank was based on earlier work by Travis, but avoided the flow through the liquefaction chamber (Imhoff, 1938).
1909–1913	Production of combustible gas by methane fermentation of strawboard was demonstrated at working scale in Netherlands in 1909. The gas was utilized for power generation. Gas utilization occurred at treatment plants for a diary and slaughterhouse in 1912, and at a wastewater treatment plant in 1914. Both fixed and floating gasholders were used (Kessener, 1938).
1914	Early experiments were conducted on the collection and heating of gas at Emschergenossenschaft in Germany (Imhoff, 1938).
1915	Experiments in heating digesters were done to increase gas production in the Netherlands (Kessener, 1938).
1927	The Ruhrverband, a German water management association, constructed a separate heated digestion tank at Essen—Rellinghausen to anaerobically digest treatment plant sludge and utilize the generated biogas for power and heating (Imhoff, 1938).
1929	A long rectangular digester with paddles on a horizontal shaft for seeding, stirring, and scum destruction was built in the Netherlands (Kessener, 1938).

(*continued*)

Table 10–2 (*Continued*)

Period	Event
1930 and 1932	Buswell of the Illinois State Water Survey started a series of reports on the fundamentals of anaerobic digestion of solids (Buswell and Neave, 1930; Buswell and Boruff, 1932).
1950	Stander of South Africa demonstrated the full-scale process benefits obtained by separating anaerobic solids externally and returning them to the digester tank (Stander, 1950; Standar and Snyders, 1950).
Early 1950s	Morgan and Torpey both demonstrated through their research the increased process performance achievable by adding mixing to an anaerobic biosolids digester (Morgan, 1954; Torpey, 1954).
Developments in the treatment of high strength wastes	
1955	Schroepfer and others were involved in the first full-scale application of the anaerobic contact process at a meat packing plant in Minnesota (Schroepfer et al., 1955).
1969	Young and McCarty developed the anaerobic filter process to make attached growth (fixed-film) anaerobic biomass available for high strength wastewater treatment (Young and McCarty, 1969; Young, 1991).
1978	Grethlein experimented with the anaerobic treatment of wastewater using a septic tank and an external cross-flow membrane (Grethlein, 1978).
Late 1970s, early 1980	Lettinga developed the upflow anaerobic sludge (UASB) process [see Fig. 10–1(h)] using waste from the sugar beet industry (Lettinga et al., 1980). His work on anaerobic processes has been instrumental in the development of many commercial anaerobic technologies, especially for the treatment of high strength wastes (see Table 10–3).
1980	Switzenbaum and Jewell developed the anaerobic fluidized bed process, particularly useful in the treatment of high strength wastewater (Switzenbaum and Jewell, 1980).

[a] Adapted in part from Totzke (2012), McCarty (2001), Metcalf & Eddy (1915).

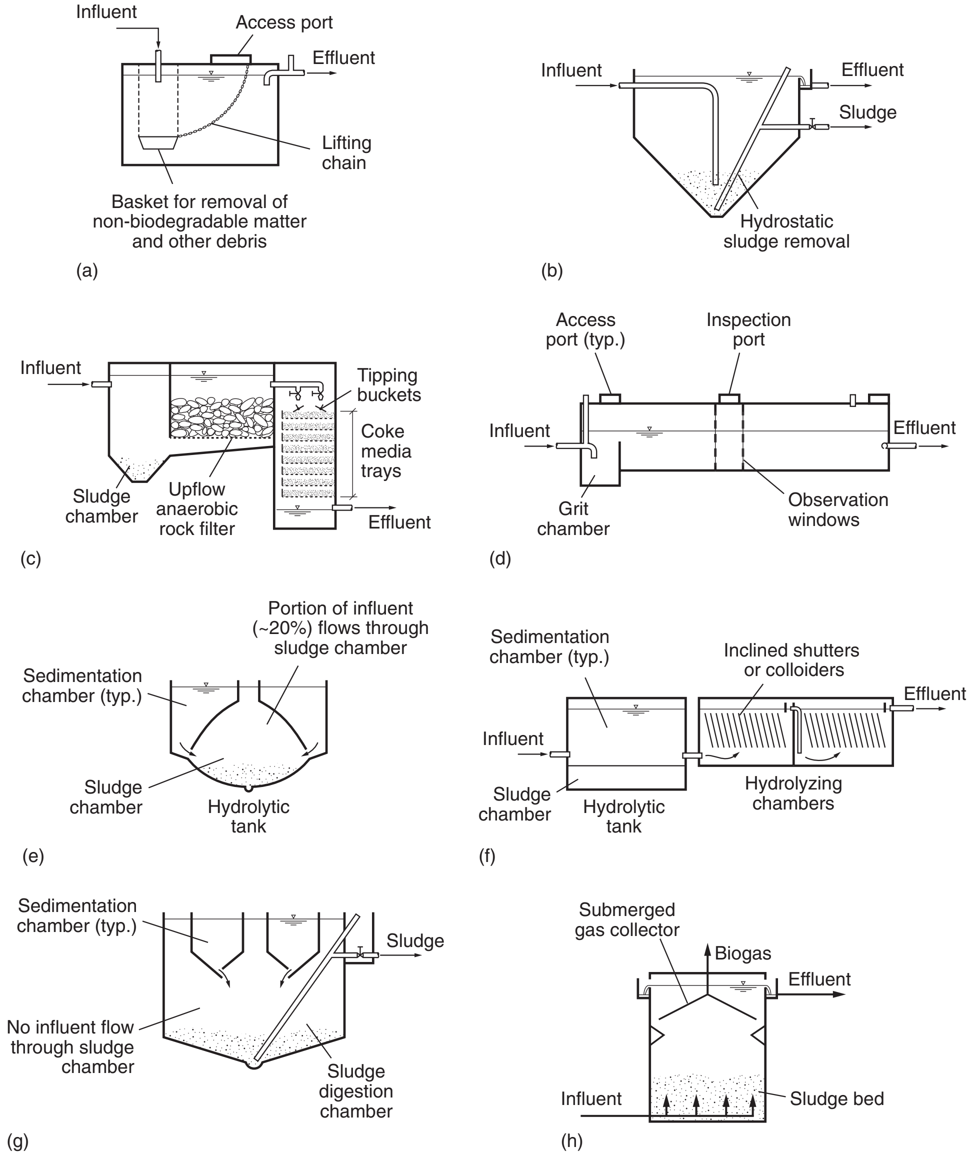
sludge. In the Dortmund tank [see Fig. 10–2(b)], sludge could liquefy and residual matter could be removed without stopping the process. In the Scott Moncrieff tank and filter [see Fig. 10–2(c)], the first tank was known as the liquefying tank. Referring to Fig. 10–2(c) wastewater entered the first compartment, which served as both a grease trap and sedimentation tank, and from there flowed into the space below the anaerobic rock filter. The wastewater then flowed up through the anaerobic rock filter. Bacteria attached to the rocks brought about the liquefaction of the colloidal material in the wastewater passing through the bed. Wastewater from above the stone bed then flowed and was distributed with tilting buckets over the uppermost series of nine perforated trays contain coke. The reactor was defined as anaerobic and the coke tray filter as aerobic. The time for the wastewater to pass through the nine trays was about 10 min.

Treatment of Wastewater Sludges

The first tank to be identified as a *septic tank* was built by Donald Cameron in England in 1895 [see Fig. 10–2(d)]. The next significant development in anaerobic technology was the Travis hydrolytic tank [see Fig. 10–2(e)] developed by W. O. Travis. As early as 1899, W. H. Clark, a chemist at the Lawrence laboratory, had suggested that sludge should be fermented in a separate tank. Travis partially implemented Clark's idea in the hydrolytic tank with the exception that in the original design a portion of the influent wastewater flowed through the liquefaction chamber. The Travis hydrolytic tank was followed by a series of hydrolyzing chambers containing an early form of inclined settlers to bring about additional treatment [see Fig. 10–2(f)]. It is interesting to note that the Imhoff tank, shown on Fig. 10–2(g), was basically an improvement of the Travis hydrolytic tank. K. Imhoff, who knew of Clark's idea, also recognized that the effluent quality was deteriorated by allowing a portion of the flow to pass through the liquefying chamber of the hydrolytic tank.

Figure 10–2

Evolution of anaerobic technology for wastewater treatment in process schematics: (a) Mouras automatic scavenger (1881), (b) Dortmund tank (1887), (c) Scott-Moncrieff two-tiered tank (1891), (d) Cameron septic tank (1895, first use of the name septic tank), (e) Travis two story hydrolytic tank (1904), (f) Travis two story hydrolytic tank with hydrolyzing chambers (1904), (g) Imhoff tank (1906), and (h) upflow anaerobic sludge blanket reactor (1980).



Imhoff tanks are still in common use at the present time. To increase gas production, experimental work on the collection, combustion, and utilization of digester gas was conducted in both Germany and the Netherlands in the early 1900s. The first heated digester constructed in Germany in 1927 evolved from early experimental work.

Treatment of High Strength Wastes

Following the stock market crash of the 1920s and the depression of the 1930s the rate of development in the anaerobic treatment of wastes slowed. Interest in wastewater treatment increased following the Second World War and considerable work was carried out in the 1950s by Stander (1950), Morgan (1954), and Torpey (1954). Concomitant with the work of Stander, Morgan, Torpey, and others in the 1950s, the principles of anaerobic degradation began to be applied to the private marketplace, a unique repository of high-strength, readily biodegradable wastes. Important technological developments in the anaerobic treatment of high strength wastes include the work of Schroepfer in the treatment of meat packing waste with the anaerobic contact process (Schroepher et al., 1955); the development of the anaerobic filter by Young and McCarty in the late 1960s (Young and McCarty, 1969; Young, 1991); and the development of the upflow anaerobic sludge blanket treatment process by Lettinga and his associates at Wageningen University in the Netherlands in the 1970s (Lettinga et al., 1980). Lettinga's work was motivated by the publications of Young and McCarty and McCarty.

The upflow anaerobic sludge blanket (UASB) reactor developed by Lettinga in 1980s [see Fig. 10–2(h)], is the most significant modern development in anaerobic process technology. Lettinga's work on the UASB and other anaerobic processes has been instrumental in the development of the many commercial anaerobic technologies discussed in the following section. Since these seminal advances, numerous researchers, consulting engineers, and manufacturers have pushed the development of anaerobic technologies to the point where there are many available suppliers and thousands of full-scale applications (Totzke, 2012).

Future Developments

In the evolution of anaerobic technologies it is interesting to note the significant amount of physical experimentation that took place in the late 1800s and early 1900s. One of the most interesting examples is the glass model [essentially the same as Fig. 10–2(f)] of the hydrolytic tank constructed by Travis to observe what was occurring in the process. The inclined glass plates, called *colloiders,* in the second and third chamber were used to remove colloidal material by interception and sedimentation, a precursor of the modern day lamella or tube settler described in Chap. 5. It should also be noted that the use of rock, coke, lath and slate filters are cited often in the literature from the late 1800s through the early 1900s. From this brief review, it is clear that many of the early process developments or variants thereof are in use today. Also, it is important to recognize that just as much or more research is going on today, but the primary focus is on understanding the fundamental process biochemistry, molecular microbiology, and physiology.

10–3 AVAILABLE ANAEROBIC TECHNOLOGIES

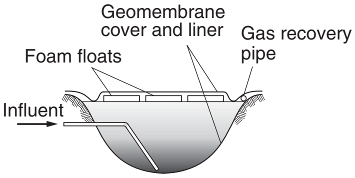
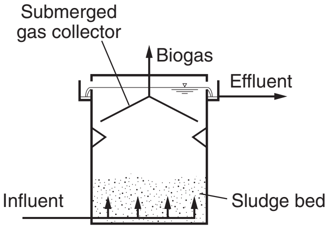
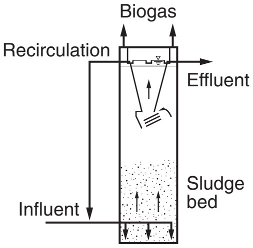
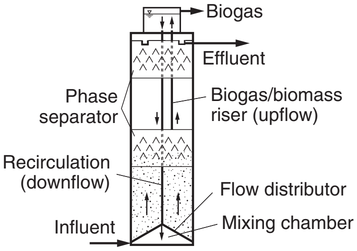
Currently, anaerobic treatment technology can be classified as public or proprietary and commercial. Public technologies are those technologies available to any experienced designer. Proprietary and commercial technologies are those technologies supplied as a complete package with little or no input from the designer. The available types of anaerobic technologies and their application are discussed below.

Types of Anaerobic Technologies

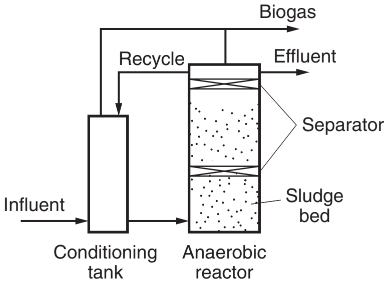
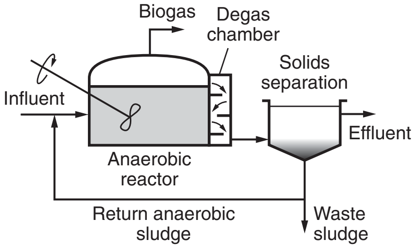
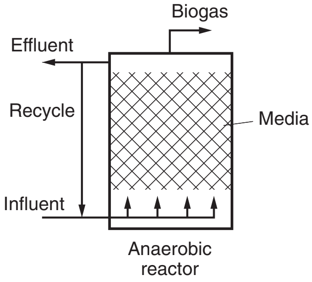
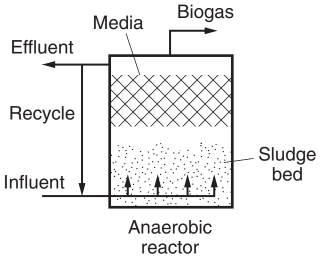
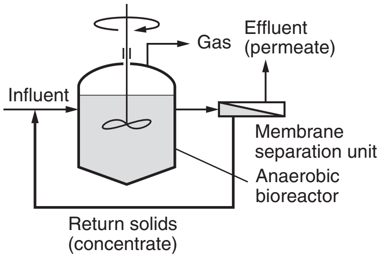
The principal types of anaerobic technologies used for the treatment of wastes, with the exception of the conventional complete-mix digester used for the treatment of municipal sludge's, are illustrated and described in Table 10–3. The digestion of municipal sludge is considered in detail in Chap. 13. The technologies identified in Table 10–3 are listed in their approximate order of application prevalence. In reviewing the technologies in Table 10–3, the lineage to Lettinga's UASB [see Table 10–3(b)] is apparent. The expanded granular sludge blanket [EGSB, see Table 10–3(c)] process is a modified UASB process with effluent recycle and higher upflow velocity to fluidize the granules within the reactor. The EGSB process was introduced by Lettinga and coworkers for the treatment of low strength brewery wastewater but it can handle both lower and higher

Table 10–3

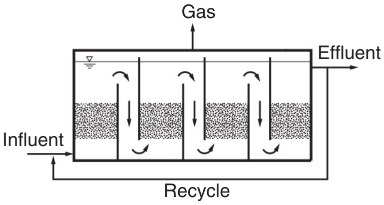
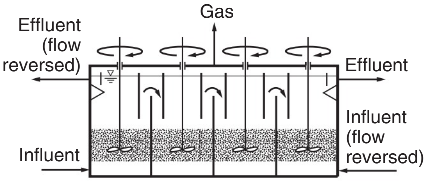
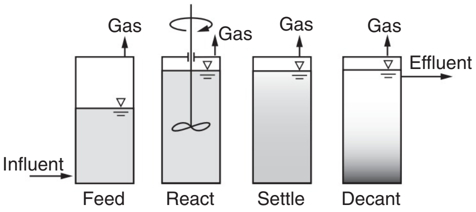
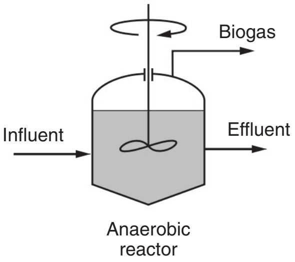
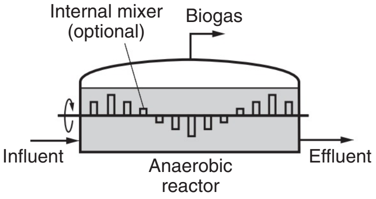
Description of the principal types of commercially available anaerobic treatment technologies[a]

Process	Description
(a) Low loaded anaerobic lagoon system (ANL)  	Generally unmixed reactor system employing suspended/flocculating anaerobic biomass and settled anaerobic solids with hydraulic retention times of 20 to 50 d and average SRTs of 50 to 100 d. Can handle a wider range of wastes including solids and soluble wastewaters. Designed for a total chemical oxygen demand (COD) loading of less than 2 kg/m^3·d. Systems can be covered with synthetic membranes for gas collection.
(b) Upflow anaerobic sludge blanket (UASB)  	An upflow reactor with a bottom sludge bed and dense, granular anaerobic biomass with good mixing provided by the upflow velocity and biogas generation. The reactor effective anaerobic sludge concentration may be in the range of 35 to 40 kg/m^3. A gas-liquid-solid separator at the top separates granular solids from the effluent and collects biogas. The sludge blanket has SRTs in excess of 30 d with hydraulic retention times in the range of 4 to 8 h. Designed for a COD loading of 5 to 20 kg/m^3·d. Upflow velocities can vary from 1 to 6 m/h and reactor heights of 5 to 20 m have been used.
(c) Expanded granular sludge blanket (EGSB)  	The EGSB is a commonly used modification of the UASB by employing a higher upflow velocity, a greater height to diameter ratio, and recirculation of effluent. Upflow velocities may be the range of 4 to 10 m/h and reactor heights up to 25 m have been used. The higher velocity provides a more efficient reactor for treatment of soluble substrates by improving mixing, reducing dead volume, and increasing diffusion rates from the bulk liquid to the granular biofilm. Was originally developed to treat low strength wastes but has been used for high strength as well and at low temperatures of 10°C. Organic loading rates as high as 35 kg/m^3·d have been used. Not as effective as the UASB for colloidal and particulate solids capture.
(d) Internal circulation UASB (IC)  	An IC reactor consists of two stacked UASB reactors in series, each with a gas separator at the top. The system uses a down comer pipe from a top chamber to the bottom inlet and a riser pipe from the first gas separator to induce recirculation and high upflow velocities in the lower granular sludge blanket reactor. Gas produced from the lower reactor is captured in the first separator and creates a gas lift for water and biosolids in the riser pipe. The gas is separated (released) from the biosolids in the chamber above the second reactor gas separator. From there the biosolids/water mixture enters the down comer pipe to provide internal recirculation to the bottom compartment. The high recirculation ratio results in high upflow velocities in the bottom chamber, 8 to 20 times higher than in conventional UASB unit, to provide good mixing and a very efficient reactor operation. The upper reactor provides a second stage anaerobic treatment for more efficient overall COD removal and the lower upflow velocity and lower biogas production rate improves effluent solids and biomass capture. Reactor heights of up to 25 m have been used.

(*continued*)

Process	**Description**
(e) Fluidized bed (FB) 	These systems are based on the development of a dense anaerobic biomass on small size (0.10 to 0.30 mm) inert particles of fine sand, basalt, pumice, or plastic. The particles are kept in suspension and mixed by a high upward velocity. The higher velocities leads to what is called a fluidized bed with 25 to 300 percent bed expansion and the expanded fluidized bed refers to operation at lower velocity with 15 to 25 percent bed expansion. These reactors are applicable for soluble wastes or easily degraded small particulates, such as whey. Upflow velocities may be in the range of 10 to 20 m/h and COD loadings of 20 to 40 kg/m^3·d have been used.
(f) Anaerobic contact process (ANCP) 	A completely mixed reactor system employing suspended anaerobic biomass, a mixing/flocculator degassing chamber, liquid-solids separation, and solids recycle so that the SRT is longer than the hydraulic retention time. Designed for a COD loading of in the range of 2 to 5 kg/m^3·d.
(g) Anaerobic filter (ANF) 	An anaerobic filter (ANF) system is an unmixed reactor system employing fixed film anaerobic biomass attached to supporting media, so that a large anaerobic biomass and long SRT can be maintained to allow treatment at hydraulic retention times in the range of 1 to 3 d and designed for a COD loading of 5 to 20 kg/m^3·d. It is available in upflow (ANFU) and downflow (ANFD) configurations.
(h) Anaerobic hybrid process (ANHYB) 	A combination of stand-alone anaerobic technologies employing a combination of an upflow anaerobic sludge blanket reactor and anaerobic filter to provide a high biomass concentration and high volumetric organic removal rates in the lower portion and further removal of volatile fatty acids and capture of suspended solids in the upper anaerobic filter portion.
(i) Anaerobic membrane process (ANMBR) 	A mixed reactor system employing suspended/flocculating anaerobic biomass and a synthetic membrane solids-liquid separation with solids recycle to provide a long SRT with the short hydraulic retention time. Designed for a COD loading of 5 to 15 kg/m^3·d.

(*continued*)

Process	Description
(j) Anaerobic baffled reactor (ABR)  	Baffles are used to direct the flow of wastewater in an upflow mode through a series of upflow anaerobic sludge blanket reactors. The sludge in the reactor rises and falls with gas production and flow, but moves through the reactor at a slow rate. Reactor volatile solids concentrations vary from 2 to 10 percent. Systems have been operated with τ values in the range of 6 to 24 h and SRTs in excess of 30 d. Designed for a COD loading of 5 to 10 kg/m^3·d. The main limitations with the ABR process are that many studies have been limited do laboratory- and pilot-scale treatment units.
(k) Anaerobic migrating blanket reactor (AMBR)  	Process is similar to the ABR with the added features of mechanical mixing in each stage and an operating approach to maintain the sludge in the system without resorting to packing or settlers for additional solids capture. When a significant quantity of solids accumulates in the last stage, the influent feed point is changed to the effluent side, which helps to maintain a more uniform sludge blanket. Organic loading rates from 1.0 to 3.0 kg COD/m^3·d with hydraulic retention times ranging from 4 to 12 h are possible.
(l) Anaerobic sequencing batch reactor (ANSBR)  	A mixed suspended growth anaerobic process with reaction and solids-liquid separation in the same vessel, much like that for aerobic sequencing batch reactors (SBRs) (see Chap. 8). The operation of SBRs consists of four steps: (1) feed, (2) react, (3) settle, and (4) decant/effluent withdrawal. The settling velocity of the sludge during the settle period before decanting the effluent is critical. Settling times used are about 30 min. After sufficient operating time, a dense granulated sludge develops that improves the liquid-solids separation. At τ values from 6 to 24 h, the SRT may range from 50 to 200 d, respectively. At 25°C, 92 to 98 percent COD removal was achieved at volumetric organic loadings of 1.2 to 2.4 kg COD/m^3·d. At 5°C, COD removal ranged from 85 to 75 percent for COD loadings from 0.9 to 2.4 kg/m^3·d, respectively.
(m) Continuously stirred tank anaerobic reactor (ANCSTR)  	A completely mixed reactor system treating semi-solids wastes with suspended anaerobic biomass. The reactor detention time equals the SRT, which may range from 15 to 30 d, with resulted COD loadings typically less than 4 kg/m^3·d.
(n) Plug flow anaerobic system (ANPF)  	Generally an unmixed rectangular reactor system treating semi-solids waste with high (10 to 18 percent) total solids concentration. In some cases the rectangular reactor is slightly inclined. Recycle of effluent solids may be done to seed the influent feed. The feed retention time equals the SRT, which may range from 20 to 30 d with COD loadings generally less than 4 kg/m^3·d.

Adapted from Nicolella et al. (2000), Totzke (2012), Tauseef et al. (2013).

strength wastewaters (Kato et al., 1999). Similarly, the internal recycle (IC) reactor [see Table 10–3(d)] is essentially two UASB reactors in series with internal recycle and it has also been successful for the treatment of low and very high strength wastewaters. The principal advantages in the development of EGSB and IC processes has been to increase the volumetric organic loading and treatment efficiency. Other anaerobic processes have been developed to treat wastes with specific characteristic (e.g., colloidal and particulate wastes).

Combined Processes. Process additions to the upflow granular sludge blanket have been made to improve treatment performance to approach or meet secondary treatment levels. These include hybrid anaerobic processes and a combined anaerobic-aerobic process. Hybrid processes typically involve two stages of anaerobic treatment, such as the one shown in Table 10–3(h) which involves an anaerobic sludge blanket process in the lower portion of the reactor followed by anaerobic attached growth process for polishing in the upper portion. A sequential anaerobic-aerobic process was reported as early as 1992 (Garuti et al., 1992) to provide a secondary effluent quality for anaerobic treatment in warm climates. More recently the aerobic process has been incorporated as an integral part of a combined anaerobic-aerobic process in a single tank (Tauseef, et al., 2013). Although there are a number of combined aerobic-anaerobic process the focus of this chapter is on the anaerobic processes.

Commercial Technologies. Compiling a list of commercially available anaerobic technologies is beyond the scope of this book task, and is complicated by the wide range of technical descriptions, trade names, and research descriptions that are available in the literature and in the marketplace. Most of the technologies listed in Table 10–3 can be defined by (1) the organic loading criteria, (2) the methods used to condition the influent waste (e.g., dilution, pH adjustment, nutrient addition), (3) the method used to introduce and distribute the influent into the reactor; (4) the method used to contact the waste to be treated with the biomass, (5) the method used to retain and separate the anaerobic biomass, a key factor in a successful operation of an anaerobic treatment process, (6) the characteristics of the biological reactor, (7) the gas management system, and (8) the ultimate management of the residual solids. Because so many new processes are coming on the market it is important to review the current literature. Pilot-plant testing is recommended for new designs and for wastewater with no full-scale treatment experience. Fundamental considerations for anaerobic treatment are presented and discussed in the following section.

Application of Anaerobic Technologies

The technologies described in Table 10–3 are used for (1) the treatment of high-strength wastes from a variety of industries, (2) the pretreatment of high-strength wastes, (3) the treatment of domestic wastewater in combination with other aerobic processes, and (4) treatment of domestic wastewater. Of these applications, most of the activity has been for the treatment of high-strength and specialized industrial wastes, but there are also a number of installations for anaerobic treatment of domestic wastewater.

Applications for High Strength Wastes. In the last 25 to 30 y, the number of industrial anaerobic installations worldwide has increased by nearly an order-of-magnitude and is as of 2013 close to 4750 (see Table 10–4). A representative listing of the industries in which anaerobic technologies are now used is presented in Table 10–5. Different types of anaerobic processes may be employed for the same type of industrial wastewater as indicated in Table 10–6. Because of their relatively small footprint, the EGSB and

Table 10–4

Estimated number of anaerobic process installations by type as of mid 2013[a] (exclusive of conventional anaerobic digesters used for wastewater sludges)

Process category	Installation
Anaerobic lagoon	>50,000
UASB	2000
EGSB	1500
Anaerobic contact process	500
Anaerobic filters (Upflow and downflow)	250
Anaerobic hybrid process	200
Anaerobic membrane process	50
Other	250

[a] Adapted from Totzke (2012).

IC processes have become quite popular for the treatment of industrial wastewaters. Typical views of EGSB and IC reactors are shown on Fig. 10–3. The use of UASB process for the treatment of high strength waste and domestic wastewater is considered below.

To meet current water quality requirements for most effluent discharges, an anaerobic process used for the treatment of high strength wastes may need to be followed by an aerobic process. A variety of aerobic processes have been used for post treatment of effluent from high rate anaerobic industrial pretreatment processes. These include conventional or biological nutrient removal activated sludge, sequencing batch reactors. trickling filters, biological aerated filters, rotating biological contactors, and wetlands (Chong et al., 2012). An example is shown on Fig. 10–4 in which a UASB reactor, located at a municipal wastewater treatment, is used to treat high strength food processing wastewater from a dedicated pipeline from the industry. The UASB effluent is directed to a biological nutrient removal activated sludge process receiving the municipal wastewater. The residual volatile fatty acids in the UASB effluent are expected to help the performance of the enhanced biological phosphorus removal process in the municipal plant.

Applications for Domestic Wastewater. In many parts of the world, especially in less developed countries with warm climates, anaerobic treatment may be the

Table 10–5

Representative examples of wastewater sources treated by anaerobic processes

Food and brewage industry	
Alcohol distillation	Slaughterhouse and meatpacking
Breweries	Soft drink beverages
Dairy and cheese processing	Starch production
Food processing	Sugar processing
Fish and seafood processing	Vegetable processing
Fruit processing	
Other applications	
Chemical manufacturing	Landfill leachate
Contaminated groundwater	Pharmaceuticals
Domestic wastewater	Pulp and paper

Table 10–6

Relative use of UASB, EGSB, and anaerobic contact process technologies in various applications[a]

Industry	UASB	EGSB	Anaerobic contact
Food and beverage industry			
Breweries	305	210	1
Candy	22	13	
Dairy and cheese processing	36	16	14
Food processing	61	29	8
Fruit	18	29	3
Meat/poultry/fish processing	8	1	11
Soft drink beverage	253	97	49
Starch production	59	30	13
Sugar processing	55	18	78
Vegetable processing	108	63	12
Yeast production	26	37	9
Other applications			
Chemical manufacturing	39	87	9
Pulp and paper	101	225	37
Miscellaneous	95	29	15
Total	1186	884	259

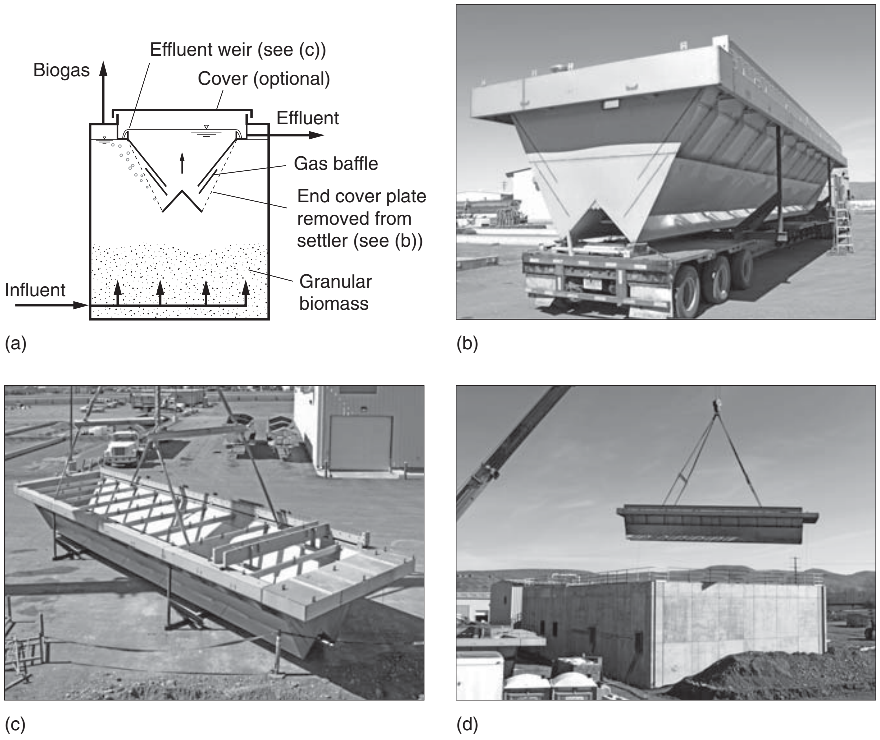
[a] Adapted from Totzke (2012).

most attractive option for domestic wastewater treatment. The USAB process is the most common anaerobic process used for treatment of domestic wastewater. Following its development for the treatment of high-strength wastewaters, it was successfully demonstrated for treating domestic wastewater at 25°C in a tropical climate during the early 1980s in Columbia (Gomec, 2010). The first full-scale facility was installed in


(a) (b) (c)

Figure 10–3

Various anaerobic reactors used for the treatment of high strength industrial wastewaters: (a) view of expanded granular sludge blanket (EGSB) treatment system. From right to left, the three tanks are (i) conditioning tank, (ii) EGSB reactor [see Table 10-3(c)], and (iii) sulfide oxidation tank (courtesy of Robert Pharmer of Pharmer Engineering); (b) view of EGSB reactor at cheese factory; and (c) view of internal circulation UASB [see Table 10-3(d)] (courtesy of Paques, BV).



(a) (b) (c) (d)

Figure 10–4

Installation of upflow anaerobic sludge blanket (UASB) reactor for the pretreatment of industrial waste: (a) schematic of settling unit in UASB reactor (adapted from Biothane, BV), (b) view of settler as delivered to the construction site, (c) view of top of settler (Note the effluent weir on the right hand side of the settler, and (d) settler in position to be placed in UASB reactor. (Photographs courtesy of City of Yakima, WA.)

1989 in Kanpur, India, and is still in operation with a treatment capacity of 5000 m^3/d. As of 2006, Aiyuk et al. (2006) reported that over 200 facilities were in operation worldwide. Many of these are in warm climates, but the UASB process had been applied for the treatment of domestic wastewaters at temperatures as low as 10°C. In a summary of UASB installations by Gomec (2010), 9 out of 35 facilities were operating at temperatures below 15°C. The hydraulic retention times in 75 percent of these facilities ranged from 2 to 10 h. A low energy wastewater treatment system employing a UASB reactor with a trickling filter treating the UASB effluent is shown on Fig. 10–5. The trickling filter effluent is used for crop irrigation following sedimentation. In the future, it is anticipated that many more combined anaerobic/aerobic treatment systems will be used and new ones developed, based on energy concerns and evolving treatment objectives.

The high rate anaerobic treatment process can be an attractive alternative for domestic wastewater treatment because of the relatively small size and low construction costs, low excess sludge production, low energy demand, and potential for biogas recovery. For countries with low water use and more concentrated wastewater, the advantages for the process increase. However, where a secondary effluent treatment quality is required, the anaerobic treatment process must be followed by an aerobic system for post treatment to further remove colloids, suspended solids and soluble BOD (Lew et al., 2002). Another emerging alternative is the use of synthetic membranes with a suspended growth anaerobic treatment process operated with a long SRT [see Table 10–3(i)]. The membrane separation can result in greater than 99 percent suspended solids removal to meet secondary treatment levels (Visvanathan and Abeynayaka, 2012). A number of processes that may be used after anaerobic treatment for nutrient removal, such as the Anammox process, are currently under investigation.

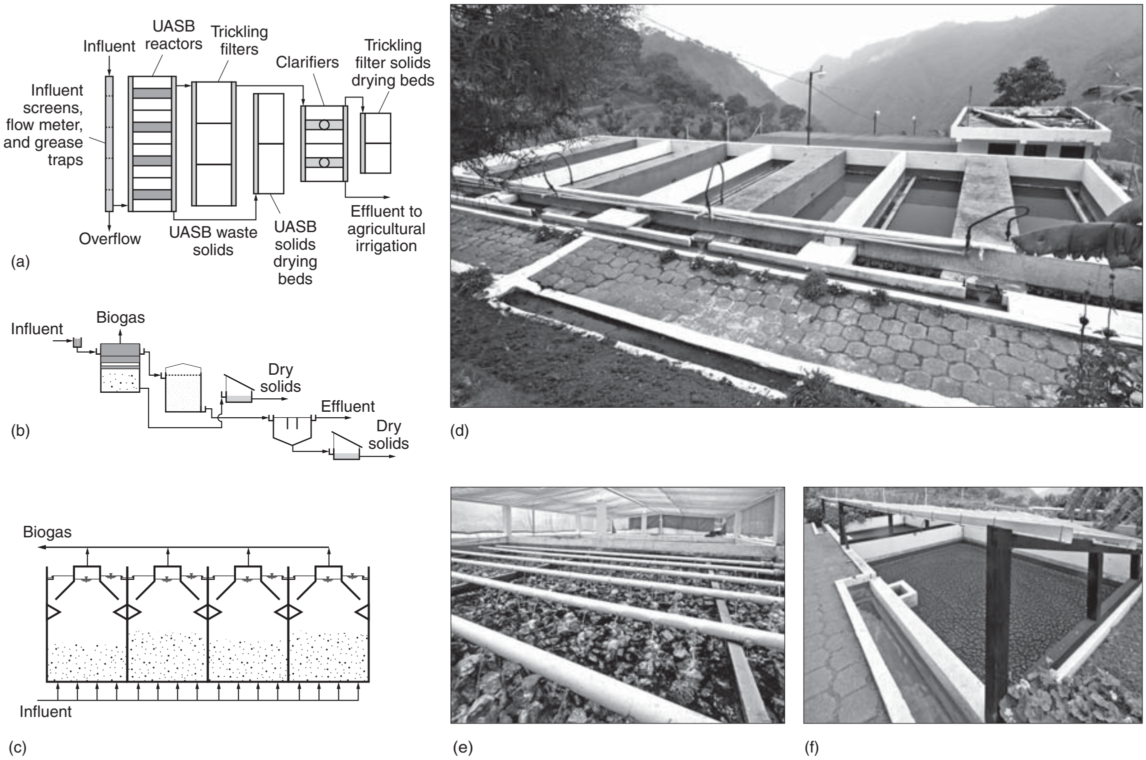


Figure 10–5

Combined UASB/trickling filter treatment process (a) schematic of the treatment system, (b) schematic of flow diagram, (c) cross-section through UASB reactor, (d) view of UASB reactor with four compartments, (e) view of gravity fed trickling filter, and (f) view of sludge drying beds. Settled trickling filter effluent is used to irrigated agricultural fields. (Photographs courtesy of S. Oakley and H. Leverenz, Coordinates: 14.7722 N, 91.1917 W; another similar plant can be seen at coordinates: W 14.7646 N, 91.1797 W.)

10–4 FUNDAMENTAL CONSIDERATIONS IN THE APPLICATION OF ANAEROBIC TREATMENT PROCESSES

The type of wastewater and its characteristics are important in the evaluation, design, and implementation of anaerobic processes. The wastewater characteristics affect the economics of selecting an anaerobic process over an aerobic treatment process, the type of anaerobic treatment process preferred, and the operational costs and concerns for using an anaerobic treatment process. The focus of the discussion in this section is on the effect of the wastewater characteristics on (1) important anaerobic process design issues, (2) the need for pretreatment and alkalinity and/or nutrient addition, and (3) the gas production and amount of energy that can be gained by treating the wastewater with an anaerobic treatment process. The topics discussed apply to the granular sludge blanket, suspended growth, attached growth, and membrane separation anaerobic processes presented in Table 10–3.

Characteristics of the Wastewater

As shown in Table 10–5, a wide variety of wastewaters have been treated by anaerobic processes. Due to the energy savings by eliminating aeration and minimal sludge production, anaerobic processes become more attractive with increasing organic concentration and temperature. Food processing and distillery wastewaters can have COD concentrations ranging from 3000 to 30,000 g/m^3. Other considerations within the type of wastewater are potential toxic streams, daily flowrate and loading variations, inorganic concentrations, and seasonal load variations.

Flowrate and Loading Variations. Wide variations in influent flowrate and organic loads can upset the balance between acid fermentation and methanogenesis in anaerobic processes. The acidogenic reactions can be much faster to possibly depress the pH and increase VFA and hydrogen concentration to levels that inhibit methanogenesis. A more conservative design or flow equalization must be considered for such situations. Flow equalization tanks can store wastewaters from batch wasting in industrial process operations and then allow a more uniform feed flowrate and feed strength to the anaerobic treatment process. More uniform feeding provides a more stable anaerobic process operation and also allows higher average organic loading rates. Anaerobic processes have been shown to be able to respond quickly to incremental increases in wastewater feed after long periods without substrate addition.

Organic Concentration and Temperature. As discussed in Sec. 10–1, the wastewater strength and temperature greatly affect the economics and choice of anaerobic treatment over aerobic treatment. Reactor temperatures of 25 to 35°C are generally preferred to support more optimal biological reaction rates and to provide more stable treatment. Generally, biodegradable COD concentrations greater than 1500 to 2000 mg/L are needed to provide sufficient methane production to heat the wastewater assuming it is at ambient temperature.

Anaerobic treatment can be applied at lower temperatures and has been sustained at 10 to 20°C in granular sludge blanket, suspended growth, and attached growth reactors. At the lower temperatures slower reaction rates occur and longer SRTs, larger reactor volumes, and lower organic COD loadings are needed (Banik et al., 1996; Collins et al., 1998; and Alvarez et al., 2008).

When higher SRTs are needed, the solids loss in an anaerobic reactor can become a critical limiting factor. Anaerobic reactors generally produce more dispersed, less flocculent solids than aerobic systems with effluent TSS concentrations for suspended growth processes in the 100 mg/L range.

For dilute wastewaters, the effluent TSS concentration will limit the possible SRT of the process and treatment potential. Either a lower treatment performance occurs or it is necessary to operate the reactor at a higher temperature. Thus, the method to retain solids in the anaerobic reactor is important in the overall process design and performance. The maximum SRT possible is when biomass lost via the effluent is equal to the solids produced, which can be calculated from Eq. 8–20.

$$Q(\text{VSSe}) = \frac{QY_{\text{H}}(b\text{CODr})}{1 + b_{\text{H}}(\text{SRT})} + \frac{(f_d)(b_{\text{H}})QY_{\text{H}}(b\text{CODr})\text{SRT}}{1 + b_{\text{H}}(\text{SRT})} \tag{10–1}$$

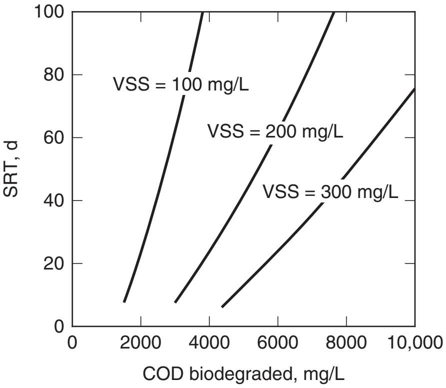
Where VSSe = effluent volatile suspended solids, g/m^3
bCODr = COD degraded in anaerobic reactor, g/m^3

Solving for SRT as a function of effluent volatile suspended solids concentration, biodegradable COD removed, and growth and decay coefficients yields

$$\text{SRT} = \frac{Y_{\text{H}}(b\text{CODr}) - \text{VSSe}}{(b_{\text{H}})(\text{VSSe}) - (f_d)(b_{\text{H}})(Y_{\text{H}})(b\text{CODr})} \tag{10–2}$$

Figure 10–6

Estimate of biomass SRT in an anaerobic reactor as a function of the amount of COD degraded and effluent biomass VSS concentration. (Based on assumptions of a synthesis yield coefficient of 0.08 g VSS/g CODr and specific endogenous decay rate coefficient of 0.03 g VSS/g VSS·d from Table 10–13 and an f_d value of 0.10.)



The maximum SRT as a function of bCOD removed and effluent VSS concentration is shown on Fig. 10–6. To achieve an SRT of 40 d, which would be desirable at a temperature of about 30°C, the amount of COD that must be degraded increases from about 2400 mg/L for an effluent VSS concentration of 100 mg/L to about 7400 mg/L if the effluent VSS concentration is 300 mg/L. For operation at lower temperatures much higher SRT values are needed, which requires a very low effluent VSS concentrations for weaker wastewaters or a higher biodegradable COD in the influent.

Fraction of Non-Dissolved Organic Matter. The composition of the wastewater in terms of its particulate and soluble fractions affects the type of anaerobic reactor selected and its design. Wastewaters with high solids concentrations are treated more appropriately in suspended growth and UASB reactors than by upflow or downflow attached growth processes. Where more particulate conversion is required, longer SRT values may be needed if solids hydrolysis is the rate limiting step as compared to acid fermentation or methanogenesis in anaerobic treatment. In such cases, it may be appropriate to use a two phase anaerobic treatment process with hydrolysis and acid fermentation in a sludge bed or stirred reactor, followed by UASB reactor or other type of reactor for methanogenesis (Shuizhou and Zhou, 2005; Alvarez et al., 2008).

Wastewater Alkalinity. During anaerobic treatment gas bubbles containing a high CO_2 content (25 to 35 percent) are produced in the liquid from fermentation reactions, which results in an elevated dissolved CO_2 concentration in the liquid. A high alkalinity concentration, in the range from 2000 to 4000 mg/L as $CaCO_3$, is typically needed to offset the dissolved carbonic acid and maintain the pH at or near neutral. For wastewaters low in alkalinity, proteins or amino acids can be metabolized and deaminated during anaerobic treatment to produce alkalinity as $NH_4(HCO_3)$. Where lacking, the purchase of alkalinity for pH control is necessary, which can have a significant impact on the economics of anaerobic treatment.

The relationship between pH and alkalinity as outlined in Appendix F is controlled by the bicarbonate chemistry as follows:

$$\frac{[HCO_3^-][H^+]}{[H_2CO_3]} = K_{a1} \tag{10–3}$$

where K_{a1} = first acid dissociation constant, which is a function of ionic strength and temperature

The carbonic acid (H_2CO_3) concentration is determined using Henry's law [Eq. (2–46)] and the partial pressure of the CO_2 in the atmosphere above the water.

$$x_g = \frac{P_T}{H} p_g \tag{2–46}$$

where x_g = mole fraction of gas in water, mole gas/mol water

$$= \frac{\text{mole gas } (n_g)}{\text{mole gas } (n_g) + \text{mole water } (n_w)}$$

P_T = total pressure, usually 1.0 atm

H = Henry's las constant, $\dfrac{\text{atm (mole gas/mole air)}}{\text{(mole gas/mole water)}}$

P_g = mole fraction of gas in air, mole gas / mole of air (Note: The mole fraction of a gas is proportional to the volume fraction)

Once the carbonic acid concentration is known, the bicarbonate (HCO_3^-) alkalinity needed to maintain the required pH can be estimated. The use of the above equations is illustrated in Example 10–1.

EXAMPLE 10–1 **Alkalinity and pH in Anaerobic Process** Determine the alkalinity required in kg $CaCO_3$/d to maintain a pH value of 7.0 in an anaerobic suspended growth process at 35°C, with a 30 percent CO_2 content in the gas phase. The influent wastewater flowrate is 2000 m^3/d and the alkalinity is 400 mg/L as $CaCO_3$. At 35°C, the Henry's constant for CO_2, computed using Eq. (2–28) and the data given in Table 2–8, is 2092 atm and the value of K_{a1} is 4.85×10^{-7} (see Appendix F).

Solution

1. Determine the concentration of HCO_3^- required to maintain the pH at or near a value of 7.0.
   a. Determine the concentration of H_2CO_3 using Eq. (2–46)

$$x_{H_2CO_3} = \frac{P_T}{H} P_g = \frac{(1 \text{ atm})(0.30)}{2092 \text{ atm}} = 1.434 \times 10^{-4}$$

Because one liter of water contains 55.6 moles [1000 g/(18 g/mole)], the mole fraction of H_2CO_3 is equal to

$$x_{H_2CO_3} = \frac{\text{mole gas } (n_g)}{\text{mole gas } (n_g) + \text{mole water } (n_w)}$$

$$1.434 \times 10^{-4} = \frac{[H_2CO_3]}{[H_2CO_3] + (55.6 \text{ mole/L})}$$

Because the number of moles of dissolved gas in a liter of water is much less than the number of moles of water,

$$[H_2CO_3] \approx (1.434 \times 10^{-4})(55.6 \text{ mole/L}) \approx 7.97 \times 10^{-3} \text{ mole/L}$$

   b. Determine the concentration of HCO_3^- required to maintain the pH at or near a value of 7.0 using Eq. (10–1)

$$[HCO_3^-] = \frac{(4.85 \times 10^{-7})(7.97 \times 10^{-3} \text{ mole/L})}{(10^{-7} \text{ mole/L})}$$

$$= 0.03863 \text{ mole/L}$$

$$HCO_3^- = 0.03863 \text{ mole/L } (61 \text{ g/mole})(10^3 \text{ mg/1 g}) = 2356 \text{ mg/L}$$

2. Determine the amount of alkalinity required per day

$$\text{Equivalents of } HCO_3^- = \frac{(2356 \text{ g/L})}{(61 \text{ g/eq})} = 0.03863$$

$$1 \text{ eq. } CaCO_3 = \frac{\text{m.w.}}{2} = \frac{(100 \text{ g/mole})}{2} = 50 \text{ g } CaCO_3/\text{eq}$$

$$\text{Alkalinity as } CaCO_3 = (0.03863 \text{ eq/L})\,(50 \text{ g/eq})(10^3 \text{ mg/1 g})$$

$$= 1931 \text{ mg/L as } CaCO_3$$

$$\text{Alkalinity needed} = (1931 - 400) \text{ mg/L}$$

$$= 1531 \text{ mg/L as } CaCO_3$$

Daily alkalinity addition:

$$= (1531 \text{ g/m}^3)\,(2000 \text{ m}^3\text{/d})\,(1 \text{ kg}/10^3 \text{ g})$$

$$= 3062 \text{ kg/d}$$

Comment Based on the results of the above analysis, it is clear that the quantity of alkalinity required can be significant, and, as a consequence, a significant cost can be incurred.

The results of similar calculations to those presented in Example 10–1 for different temperatures and gas phase CO_2 concentrations are reported in Table 10–7. The data presented in Table 10–7 were derived using the bicarbonate dissociation constants given in Table F–2 in Appendix F and Henry's constants derived from the data given in Table 2–8 in Chap. 2. The values presented in Table 10–7 can be used to estimate the alkalinity requirements. For wastewaters with a higher total dissolved solids concentration and ionic strength, the alkalinity requirements will generally be much greater.

Nutrients. Though anaerobic processes produce less sludge and thus require less nitrogen and phosphorus for biomass growth, many industrial wastewaters may lack sufficient nutrients. Thus, the addition of nitrogen and/or phosphorus may be needed.

Macronutrients. The presence of the iron, nickel, cobalt, and molybdenum at trace concentrations are needed for the growth of methanogenic bacteria in anaerobic processes (Demirel and Scherer, 2011). The addition of trace metals has been shown in a number of cases to increase COD removal efficiency in anaerobic processes including granular

Table 10–7

Estimated minimum alkalinity as $CaCO_3$ required to maintain a pH of 7.0 as a function of temperature and percent carbon dioxide during anaerobic digestion

	Gas phase CO_2, %			
Temperature, °C	25	30	35	40
20	2040	2449	2857	3265
25	1913	2295	2678	3061
30	1761	2113	2465	2817
35	1609	1931	2253	2575
40	1476	1771	2066	2362

Table 10–8

Toxic and inhibitory inorganic compounds and concentrations harmful to methanogenesis in anaerobic processes[a]

Substance	Moderately inhibitory concentration, mg/L	Strongly inhibitory concentration, mg/L
Na^+	3500–5500	8000
K^+	2500–4500	12,000
Ca^{2+}	2500–4500	8000
Mg^{2+}	1000–1500	3000
Ammonia-nitrogen	1500–3000	3000
Sulfide, S^{2-}	200	200
Copper, Cu		0.5 (soluble)
		50–70 (total)
Chromium, Cr(VI)		3.0 (soluble)
		200–250 (total)
Chromium, Cr(III)		180–420 (total)
		2.0 (soluble)
Nickel, Ni		30.0 (total)
Zinc, Zn		1.0 (soluble)

[a] From Parkin and Owen (1986).

sludge reactors (Osuna et al., 2003; Fermoso et al., 2008) and a suspended growth process for the digestion of food waste (Evans et al., 2012). Digester studies by Takashima et al. (2011) led to a recommendation of having the following ratios available for iron, nickel, cobalt, and zinc in mg/g COD removed, respectively for efficient anaerobic degradation: 0.20, 0.0063, 0.017, and 0.049 in a mesophilic system and 0.45, 0.049, 0.054, and 0.24 in a thermophilic system.

The exact amounts of trace metals needed can vary for different wastewaters and, thus, successive trials are used to assess their benefit for high rate anaerobic processes.

Inorganic and Organic Toxic Compounds. Proper waste analysis and treatability studies are needed to assure that a chronic or serious transient toxicity does not exist for wastewater treated by anaerobic processes. At the same time, the presence of a toxic substance does not mean the process cannot function. Some toxic compounds inhibit anaerobic methanogenic reaction rates, but with a high biomass inventory and low enough loading, the process can be sustained. Toxic and inhibitory inorganic and organic compounds of concern for anaerobic processes are presented in Tables 10–8 and 10–9, respectively. Acclimation to toxic concentrations has also been shown (Speece, 1996) but it may be necessary to apply pretreatment steps to prevent toxicity problems in the anaerobic degradation process.

Pretreatment of Wastewater

Wastewater pretreatment requirements depend on the waste source, the type of anaerobic treatment process employed, and the need to prevent an anaerobic treatment process failure or unstable operation. Pretreatment considerations include screening, solids conditioning or reduction, pH and temperature adjustment, nutrient addition, and fats, oil, and grease (FOG) control, and toxicity reduction.

Table 10–9

Toxic and inhibitory organic compounds and concentrations harmful to methanogenesis in anaerobic processes[a]

Compound	Concentration resulting in 50% activity, mM
1-Chioropropene	0.1
Nitrobenrene	0.1
Acrolein	0.2
1-Chioropropane	1.9
Formaldehyde	2.4
Lauric acid	2.6
Ethyl benzene	3.2
Acrylonitrile	4
3-Chlorol- 1, 2-propandiol	6
Crotonaldehyde	6.5
2-Chioropropionic acid	8
Vinyl acetate	8
Acetaldehyde	10
Ethyl acetate	11
Acrylic acid	12
Catechol	24
Phenol	26
Aniline	26
Resorcinol	29
Propanol	90

[a] From Parkin and Owen (1986).

Screening. Some form of screening is normally used to removed objectionable material that could cause interference with the flow distribution in a granular sludge reactor, mixing problems in suspended growth reactors, or plugging of attached growth reactors. Fine screening, in the range of 2 to 3 mm, should be considered for anaerobic membrane reactors to prevent membrane fouling problems.

Solids Conditioning or Reduction. A solids conditioning pretreatment step may be considered in the processing of wastes that are high in solids content and/or lignin material, such as agriculture waste or certain pulp and paper mill streams, to enhance methane production and anaerobic degradation reaction rates. These may include mechanical, chemical, thermal, and biological processes or a combination of them (Sambusiti et al., 2013). Solids conditioning technology is addressed in Chap. 13 for anaerobic digestion of wastewater treatment plant sludges.

Solids reduction by a two-step process with solids removal or solids removal and hydrolysis in the first step before downstream granular sludge and attached growth anaerobic processes can be beneficial to COD removal performance and operational stability. High suspended solids concentration in the influent to a granular sludge anaerobic process may cause clogging and channeling in the sludge blanket to reduce treatment effectiveness. In addition the adsorption and incorporation of suspended and colloidal solids may impair the sludge granulation process and the density of the granules. Minimal influent solids concentrations are desired for the operation of packed bed anaerobic treatment processes to prevent plugging and channeling of the wastewater flow. One option is to provide gravity settling for solids removal and thickening with the settled solids treated in a separate sludge digester and another option is to provide solids reduction by anaerobic hydrolysis in the first step.

A two-step processes with solids contact and hydrolysis in the first step was proposed by van Haandel and Lettinga (1994) and has been used for both industrial and domestic wastewater treatment with downstream UASB, EGSB, and packed bed anaerobic processes (Seghezzo et al., 1998). A hydrolysis upflow sludge blanket (HUSB) reactor or alternatively referred to as an upflow anaerobic solids removal (UASR) reactor provides solids capture and hydrolysis prior to a downstream anaerobic process. It has been demonstrated in domestic wastewater treatment applications at temperatures from 14 to 26°C and with hydraulic retention times from 3 to 10 h (Alvarez et al., 2008; Zeeman et al., 1997). A greater solids accumulation occurs under cold temperature operation due to the slower hydrolysis rates and may require solids wasting and further treatment elsewhere.

pH Adjustment. The ability to operate the anaerobic process with little variation in the reactor pH and temperature leads to a more stable operation and better process efficiency. Based on knowledge of the wastewater characteristics and the anaerobic reactor operating conditions the amount of alkalinity, if needed, can be determined and provided in the operational protocols. For more dilute wastewaters with lower gas production rates relative to the wastewater flowrates, the percent CO_2 can be minimal compared to the amount when treating a wastewaters with high COD concentrations, such that the alkalinity demands are none or modest. The influent alkalinity should be controlled to maintain reactor operating pH values between 6.8 and 7.8 for which stable methanogenic activity occurs (Leitão et al., 2006).

Temperature Adjustment. Operation at constant temperature provides a more stable and more efficient process performance. A decrease in the reactor temperature is more detrimental than an increase. A sudden temperature drop of 1 to 2°C results in a slower acetate uptake rate by the methanogenic bacteria and accumulation of VFAs. Depending on the system alkalinity and buffer capacity, this could result in a pH decrease that could further slow down the methanogenic activity leading to a path of digester instability and potential failure if not adjusted soon enough. Sudden temperature drops can also affect the integrity of granular sludge in UASB, EGSB, and IC reactors.

Nutrient Addition. Many high strength wastewaters, such as those from food processing, brewery, beverage, and distillery operations, will require the addition of nitrogen and phosphorus to support growth of the anaerobic bacteria. The amount needed is higher during start up due to more rapid growth and less supply of nutrients within the reactor by endogenous decay. An influent COD:N:P ratio of 600:5:1 is recommended during start up and 300:5:1 during long term operation (Annachhatre, 1996).

Fats, Oil, and Grease (FOG) Control. The presence of FOG in wastewaters fed to anaerobic processes can be of concern for two reasons: (1) inhibition of methanogenesis due to inhibition by long chain fatty acids (LCFAs) and (2) sludge flotation due to the hydrophobicity and lower density of FOG components. The inhibition is thought to be due to sorption of LCFAs on the methanogenic organism's cell wall and membranes to interfere with cell substrate transport functions (Hanaki et al., 1981). Further, the presence of FOG can also be detrimental to the integrity of granular sludge particles in UASB and EGSB reactors and can cause fouling of the synthetic membranes used in anaerobic membrane reactors.

Anaerobic biodegradation of FOG has been demonstrated in the treatment of rendering waste with sustained FOG loadings of less than 1.0 kg FOG/m^3·d, but unacceptable solids loss occurred due to flotation at higher loadings (Jeganathan et al., 2006). Upon initial exposure, FOG inhibition of methanogenic activity occurs at much lower loadings until FOG degradation occurs following an acclimation period (Evans et al., 2012). Thus, intermittent or variable FOG loadings can cause methanogenic inhibition and unstable operation of the anaerobic process. Source control and dissolved air flotation pretreatment must be considered for wastewater with high FOG concentration.

Toxicity Reduction. As shown in Tables 10–8 and 10–9 a wide range of inorganic and organic substances can be toxic to anaerobic processes, including certain heavy metals, high dissolved solids, chlorinate organic compounds, high nitrogen concentration from ammonia, amino acids, and or proteins, and industrial chemical products. Control of toxicity for anaerobic processes requires a careful evaluation of the wastewater and its sources. Source control is a first step and where necessary pretreatment steps for toxicity reduction may include for example dilution, air stripping, and chemical precipitation processes. Air stripping was found for example to be effective for the reduction of high ammonia concentration toxicity in piggery wastes (Zhang and Jahng, 2010).

A two step process in the anaerobic treatment system may be used to remove or biodegrade toxic substances before a final methanogenic treatment step. A first step solids removal/hydrolysis or acid phase anaerobic treatment may remove the toxin to a sufficient level before exposure of the more sensitive methanogenic bacteria to the toxic compound (Lettinga and Hulshoff Pol, 1991).

Expected Gas Production

Higher strength wastewaters will produce a greater amount of methane per volume of liquid treated to provide a relatively higher amount of energy to raise the liquid temperature if needed. Gas composition and volume relationships for methane are discussed below.

Gas Composition. Anaerobic degradation of organic substances results in the production of a gaseous product with the conversion of carbon to methane (CH_4) and carbon dioxide (CO_2), nitrogen to ammonia (NH_3), and sulfur to hydrogen sulfide (H_2S). The energy value of the gas produced is proportional to its methane content. Buswell and Boruff (1932) were the first to recognize that the composition of the gaseous product is a function of the organic compound type and composition. Buswell and Mueller (1952) later developed a molar stoichiometric relationship between the carbon, hydrogen, and oxygen in an organic compound to the volume of methane, and carbon dioxide produced in anaerobic degradation. Their relationship was further modified to include the organic compound nitrogen and sulfur content and the volume of ammonia and hydrogen sulfide produced (Parkin and Owen, 1986; Tchobanoglous et al., 2003):

$$\begin{aligned} &C_vH_wO_xN_yS_z + \left(v - \frac{w}{4} - \frac{x}{2} + \frac{3y}{4} + \frac{z}{2}\right)H_2O \rightarrow \\ &\quad \left(\frac{v}{2} + \frac{w}{8} - \frac{x}{4} - \frac{3y}{8} - \frac{z}{4}\right)CH_4 + \left(\frac{v}{2} - \frac{w}{8} + \frac{x}{4} + \frac{3y}{8} + \frac{z}{4}\right)CO_2 + yNH_3 + zH_2S \end{aligned} \tag{10–4}$$

The gaseous ammonia (NH_3) that is formed will react with the carbon dioxide to form the ammonium ion and bicarbonate according to the following relationship.

$$NH_3 + H_2O + CO_2 \rightarrow NH_4^+ + HCO_3^- \tag{10–5}$$

The reaction given by Eq. (10–5) is representative of the formation of alkalinity under anaerobic conditions, due to the conversion of organic compounds containing proteins, peptides or amino acids (i.e., nitrogen). The expected mole fractions of methane and carbon dioxide, and hydrogen sulfide are given by the following three expressions, respectively. Ammonia gas production from Eq. (10–4) is not included as most of it will remain in solution as ammonium bicarbonate. In general, the mole fraction of hydrogen sulfide will be somewhat less because of metal complexation/precipitation.

$$f_{CO_2} = \frac{4v - w + 2x + 2z}{8(v + z)} \tag{10–6}$$

$$f_{CH_4} = \frac{4v + w - 2x - 2z}{8(v + z)} \tag{10–7}$$

$$f_{H_2S} = \frac{z}{(v + z)} \tag{10–8}$$

Using approximate formulas for lipid, carbohydrate, and protein compounds of $C_{18}H_{33}O_2$, $C_6H_{10}O_5$, and $C_{11}H_{24}O_5N_4$, the percent methane in the anaerobic process gas calculated from Eq. (10–7) are 70, 50, and 66, respectively. These values compare well to respective values of 70, 50, and 68 percent in studies by Li et al. (2002). For carbohydrate, starch, and FOG wastes, alkalinity will be a problem due to the lack of ammonia production.

Volume of Methane Gas. As derived by Eq. (7–142) in Sec. 7–14 in Chap. 7, the amount of methane (CH_4) produced from anaerobic oxidation of COD is equal to 0.35 L CH_4/g COD at standard conditions (0°C and 1 atm). The quantity of methane at other than standard conditions is determined by using the universal gas law [Eq. (2–44)] to determine the volume of gas occupied by one mole of CH_4 at the temperature in question.

$$V = \frac{nRT}{P} \tag{2–44}$$

Where V = volume occupied by the gas, L, m^3
n = moles of gas, mole
R = universal gas law constant, 0.08205 atm·L/g mole·K
= 0.08205 atm·L/g mole·K
T = temperature, °K (273.15 + °C)
P = absolute pressure, atm

Thus, for 35°C the volume occupied by one mole of CH_4 is:

$$V = \frac{(1\text{ mole})(0.082056\text{ atm·L/g mole·K})(273.15 + 35)}{1.0\text{ atm}} = 25.29\text{ L/mole}$$

Because the COD of one mole of CH_4 is equal to 64 g, the amount of CH_4 produced per unit of COD converted under anaerobic conditions at 35°C is equal to 0.40 L as determined below.

$$(25.29\text{ L/mole})/(64\text{ g COD/mole }CH_4) = 0.40\text{ L }CH_4\text{/g COD}$$

The total gas production rate is commonly estimated by calculating the methane production rate and dividing it by the fraction of methane in the gas phase, typically assuming a value in the range of 60 to 65 percent methane.

Table 10–10

Comparison of energy balance for aerobic and anaerobic processes. Wastewater at 20°C, 100 m^3/d and 10,000 g/m^3 COD concentration

Energy	Value, kJ/d	
	Anaerobic	Aerobic
Aeration		-1.9×10^6
Methane produced	12.5×10^6	
Increase wastewater temperature to 30°C	-5.3×10^6	
Net energy, kJ/d	7.2×10^6	-1.9×10^6

Energy Production Potential

In contrast to aerobic biological treatment very little electrical energy is used for anaerobic biological treatment, being limited to energy for pumping and tank mixing, depending on the anaerobic process. The methane produced in the anaerobic process is often used to heat the wastewater or sludge where required. In some cases for larger installations, the methane may be used in reciprocating engine or microturbine equipment to produce electricity. When the heat generated during electrical generation is captured and used for heating, the energy producing process is called cogeneration.

An energy balance comparison for aerobic and anaerobic treatment of a high strength wastewater at 20°C is presented in Table 10–10. For the conditions given, the aerobic process requires 1.9×10^6 kJ/d of energy. On the other hand, the anaerobic process produces a total of 12.5×10^6 kJ/d. Of the total energy produced anaerobically, about 5.3×10^6 kJ/d is required to raise the wastewater temperature from 20 to 30°C, assuming an 80 percent energy utilization efficiency after accounting for losses in the boiler and heat exchanger and reactor heat losses. Thus, the potential net energy production that can be achieved with anaerobic treatment is on the order of 7.2×10^6 kJ/d, or about 3.8 times the energy required for aerobic treatment. General considerations used for the energy balance are shown as follows.

A large portion of the energy demand for an aerobic activated sludge process is for aeration to supply oxygen and can be calculated as follows:

$$E_{\mathrm{AER}}(\mathrm{kJ/d}) = Q(\mathrm{CODr})(A_n)(3600\ \mathrm{kJ/kWh})/\mathrm{AOTE} \tag{10–9}$$

Where E_{AER} = daily energy demand for oxygen transfer, kJ/d
Q = wastewater flowrate, m^3/d
CODr = COD removed by biodegradation, kg/m^3
A_n = net oxygen required, kg O_2/kg CODr
AOTE = actual oxygen transfer efficiency, kg O_2/kWh

The value for A_n can be calculated for a given activated sludge design SRT using Eq. (8–67). The anaerobic process produces energy in the form of methane but may use some portion of the methane produced to heat the wastewater to a more optimal operating temperature, typically in the range of 30 to 35°C. The net energy production, accounting for the energy from methane production and energy used for heating is as follows:

$$\begin{aligned} E_{\mathrm{ANAER}},\ \mathrm{kJ/d} &= (Q)(\mathrm{CODr})\left(\frac{0.35\ \mathrm{m^3\ CH_4}}{\mathrm{kg\ CODr}}\right)\left(\frac{35{,}846\ \mathrm{kJ}}{\mathrm{m^3\ CH_4}}\right) \\ &\quad -(Q)(\Delta T)(C_p)\left(\frac{10^3\ \mathrm{kg}}{\mathrm{m^3\ H_2O}}\right)\left(\frac{1}{\mathrm{Eff_{heat}}}\right) \end{aligned} \tag{10–10}$$

Where: E_{ANAER} = net energy available, kJ/d
ΔT = temperature increase for influent wastewater, °C
C_p = specific heat of water, 4.2 kJ/°C·kg
Eff_{heat} = fraction of heat available after losses from vessel and heat exchanger

The effect of the wastewater strength on the net energy consumption or production between aerobic and anaerobic treatment is illustrated in Example 10–2. The same assumptions used to generate the comparison in Table 10–10 are used in the example.

EXAMPLE 10–2 **Comparison of Energy Consumption and Production in Aerobic and Anaerobic Treatment** Compare the energy needed for aeration for treatment of a high strength wastewater by an activated sludge aerobic treatment process to the net energy production for an anaerobic treatment process. The net energy production for anaerobic treatment accounts for the energy produced by methane generation minus the energy used to heat the wastewater from 20 to 30°C.

a. Express the net energy as a positive or negative production in kJ/m^3 of wastewater treated and tabulate the results for aerobic and anaerobic treatment to achieve a COD biodegradation of 3800, 4200, 6000, 8000, and 10,000 mg/L. Include the volume of methane produced in m^3/d for a wastewater flowrate of 400 m^3/d.
b. Using similar calculations as for the above, prepare a graph that can be used to illustrate the net energy for anaerobic treatment in terms of kJ/m^3 for three wastewater conditions that require heating from 25 to 30°C, 20 to 30°C and 10 to 30°C. Use an influent wastewater COD biodegradation range of 1500 to 10,000 mg/L. Where the net energy production is negative limit the *Y* axis to –5000 kJ/m^3.

The following assumptions apply:

1. A_n = 0.80 g O_2/g COD removal
2. Aeration actual oxygen transfer efficiency (OTE) = 1.52 kg O_2/kWh
3. Net heat loss for methane utilization for heating = 20 percent
4. Heat capacity of water = 4.2 kJ/°C·kg
5. Energy content of methane at standard conditions = 38,846 kJ/m^3
6. Methane production rate at standard conditions = 0.35 m^3 CH_4/kg CODr (ignore COD to biomass, which is about 3 to 4 percent of total COD degraded)

Solution

1. Determine the energy needed for aeration per unit flow in the aerobic treatment process using Eq. 10–9 for a CODr of 3800 mg/L (3.8 kg/m^3).

$$\frac{E_{\text{AER}}}{Q} = (\text{CODr})(A_n)(3600 \text{ kJ/kWh})/\text{AOTE}$$

$$\frac{E_{\text{AER}}}{Q} = (3.8 \text{ kg COD/m}^3)(0.8 \text{ kg O}_2\text{/kg COD})\left[\frac{1}{(1.52 \text{ kg O}_2\text{/kWh})}\right](3600 \text{ kJ/kWh})$$

$$= 7200 \text{ kJ/m}^3$$

The aeration energy needed for the other CODr removal levels is directly proportional to the COD concentration and is tabulated in the summary table in Step 4a.

2. Determine the net energy produced per unit flow from methane production and wastewater heating using Eq. (10–10) for a CODr = 3800 mg/L for heating the wastewater from 20 to 30°C (ΔT = 10°C). With a 20 percent heat loss, the methane utilization efficiency for heating is 80 percent.

a. Net energy for a CODr = 3800 mg/L

$$\frac{E_{\text{ANAER}}}{Q} = (3.8\text{ kg COD/m}^3)(0.35\text{ m}^3\text{ CH}_4\text{/kg COD})(35{,}846\text{ kJ/m}^3\text{ CH}_4)$$

$$-(10^\circ\text{C})(4.2\text{ kJ/}^\circ\text{C}\cdot\text{kg})(10^3\text{ kg/m}^3\text{ H}_2\text{O})\left(\frac{1}{0.80}\right)$$

$$= -4825\text{ kJ/m}^3$$

b. Net energy for a CODr = 4200 mg/L

$$\frac{E_{\text{ANAER}}}{Q} = (4.2\text{ kg COD/m}^3)(0.35\text{ m}^3\text{ CH}_4\text{/kg COD})(35{,}846\text{ kJ/m}^3\text{ CH}_4)$$

$$-(10^\circ\text{C})(4.2\text{ kJ/}^\circ\text{C}\cdot\text{kg})(10^3\text{ kg/m}^3\text{ H}_2\text{O})\left(\frac{1}{0.80}\right)$$

$$= 194\text{ kJ/m}^3$$

Similarly for a CODr of 6000, 8000 and 10,000 mg/L, the specific net energy production is 22,777, 47,869, and 72,961 kJ/m^3, respectively.

3. Determine the volume of methane produced for a flowrate of 400 m^3/d as a function of temperature using the ideal gas law.

a. Methane volume for a CODr = 3800 mg/L at a temperature of 30°C
From the ideal gas law, $V_2 = (V_1/T_1)\,T_2$

$$\text{CH}_4\text{ production at }30^\circ\text{C} = (\text{CODr})Q(0.35\text{ m}^3\text{ CH}_4\text{/kg COD})\left(\frac{273.15^\circ\text{C} + 30^\circ\text{C}}{273.15^\circ\text{C}}\right)$$

$$= (3.8\text{ kg/m}^3)(400\text{ m}^3\text{/d})(0.35\text{ m}^3\text{ CH}_4\text{/kg COD})(1.1098)$$

$$= 590.4\text{ m}^3\text{ CH}_4\text{/d}$$

b. Methane volume for a CODr = 4200 mg/L at a temperature of 30°C

$$\text{CH}_4\text{ production at }30^\circ\text{C} = (\text{CODr})Q(0.35\text{ m}^3\text{ CH}_4\text{/kg COD})\left(\frac{273.15^\circ\text{C} + 30^\circ\text{C}}{273.15^\circ\text{C}}\right)$$

$$= (4.2\text{ kg/m}^3)(400\text{ m}^3\text{/d})(0.35\text{ m}^3\text{ CH}_4\text{/kg COD})(1.1098)$$

$$= 652.6\text{ m}^3\text{ CH}_4\text{/d}$$

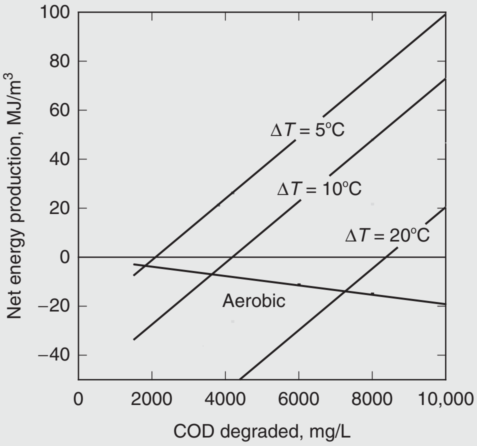
Similarly for a CODr of 6000, 8000 and 10,000 mg/L, methane production is 930, 1240, and 1550 m^3/d, respectively.

4. Prepare a summary table and a graphic plot to illustrate the results.

a. Prepare summary table

CODr, mg/L	Aeration energy, kJ/m^3	Net anaerobic treatment energy, kJ/m^3	Methane production, m^3/d
3800	−7200	−4830	590
4200	−7960	190	650
6000	−11,370	22,780	930
8000	−15,160	47,870	1240
10,000	−18,950	72,960	1550

b. Similar calculations used in Step 2a were prepared for $\Delta T = 5^\circ$C and 10°C. The net energy produced from an anaerobic process as a function of the wastewater temperature increase needed to bring the temperature up to 30°C and the amount of COD degraded is shown on the following plot.



Comment The wastewater strength and temperature is important for comparing energy balances for aerobic and anaerobic processes, where the wastewater temperature must be increased. Based on the results presented in the summary table, both the aerobic and anaerobic processes require energy input for the degradation of a wastewater with a COD of 3800 mg/L COD. It can be shown that if the amount of COD degraded is 3640 mg/L. the energy input is equal for both systems. At lower COD concentrations, the aerobic process requires less energy and the anaerobic process has less methane production.

The effect of the influent wastewater temperature and the amount of COD that can be degraded on the net energy produced by anaerobic treatment is shown in the above figure. A net positive energy production for the anaerobic process requires biodegrading 2100, 4200, and 8400 mg/L COD if it is necessary to raise the influent wastewater temperature 5, 10, and 20°C, respectively. However, heat recovery from the anaerobic effluent stream can modify these values. Even at a break even net energy production, the lower biomass yield discussed below is still a major advantage offered by anaerobic treatment. Consideration should also be given to whether alkalinity addition is necessary for the anaerobic process.

Sulfide Production

Oxidized sulfur compounds, such as sulfate, sulfite, and thiosulfate, may be present in significant concentrations in various industrial wastewaters and to some degree in municipal wastewaters. These compounds can serve as electron acceptors for sulfate-reducing bacteria, which consume organic compounds in the anaerobic reactor and produce hydrogen sulfide (H_2S). For example, using methanol as the electron donor and an f_s value of 0.05 (see Sec. 7–4 in Chap. 7), the overall reaction for the reduction of sulfate to H_2S can be illustrated by the following expression:

$$0.119\ SO_4^{2-} + 0.167CH_3OH + 0.010CO_2 + 0.003NH_4^+ + 0.003\ HCO_3^- +$$
$$0.178H^+ = 0.003C_5H_7NO_2 + 0.060H_2S + 0.060HS^- + 0.331H_2O \qquad (10–11)$$

From Eq. (10–11), the amount of COD used for sulfate reduction is 0.89 g COD/g sulfate which is higher than the reported value of 0.67 g COD/g sulfate reduced (Arceivala, 1998) and is due to the lower biomass yield coefficient associated with methanol oxidation. Based on the following stoichiometry for H_2S oxidation, 2 moles of oxygen are required per mole of H_2S, as was the case for methane oxidation,

$$H_2S + 2O_2 \rightarrow H_2SO_4 \qquad (10–12)$$

Thus, the amount of H_2S produced per unit COD is the same as that for methane (0.40 L H_2S/g COD used at 35°C).

Hydrogen sulfide is malodorous and corrosive to metals. In contrast to methane, H_2S is highly soluble in water with a solubility of 2650 mg/L at 35°C, for example.

The concentration of oxidized sulfur compounds in the influent wastewater to an anaerobic treatment process is important as high concentrations can have a negative effect on anaerobic treatment. Sulfate reducing bacteria compete with the methanogenic bacteria for COD, and, thus, can decrease the amount of methane gas production. While low concentrations of sulfide (less than 20 mg/L) are needed for optimal methanogenic activity, higher concentrations can be toxic Speece (1996, 2008). Methanogenic activity has been decreased by 50 percent or more at H_2S concentrations ranging from 50 to 250 mg/L (Arceivala, 1998). A comprehensive evaluation of the dynamics of competition between sulfate-reducing and methanogenic bacteria and toxicity effects is given in Maillacheruvu et al. (1993).

Because unionized H_2S is considered more toxic than ionized sulfide, pH is important in determining H_2S toxicity. The degree of H_2S toxicity is also complicated by the type of anaerobic biomass present (granular versus dispersed), the particular methanogenic population, and the feed COD/SO_4 ratio. With higher COD concentrations, more methane gas is produced to dilute the H_2S and transfer more of it to the gas phase. Hydrogen sulfide exists in aqueous solution as either the hydrogen sulfide gas (H_2S), the ion (HS^-) or as the sulfide ion (S^{2-}), depending on the pH of the solution, in accordance with the following equilibrium reactions:

$$H_2S \rightleftarrows HS^- + H^+ \tag{10–13}$$

$$HS^- \rightleftarrows S^{2-} + H^+ \tag{10–14}$$

Applying the law of mass action [Eqs. (10–13)] to Eq. (10–14) yields

$$\frac{[HS^-][H^+]}{[H_2S]} = K_{a1} \text{ and } \frac{[S^{2-}][H^+]}{[HS^-]} = K_{a2} \tag{10–15}$$

where K_{a1} = first acid dissociation constant, 1×10^{-7}
K_{a2} = second acid dissociation constant, $\sim 10^{-19}$ (value uncertain)

The percentage H_2S as a function of the pH can be determined using the following relationship:

$$H_2S, \% = \frac{[H_2S](100)}{[H_2S] + [HS^-]} = \frac{100}{1 + [HS^-]/H_2S} = \frac{100}{1 + [H^+]/K_{a1}} \tag{10–16}$$

Dissociation constants for hydrogen sulfide as a function of temperature are presented in Table 10–11, along with values for ammonia. As illustrated on Fig. 10–7, at a pH value of 7, at 30°C, about 60 percent of the total H_2S is present as gaseous H_2S.

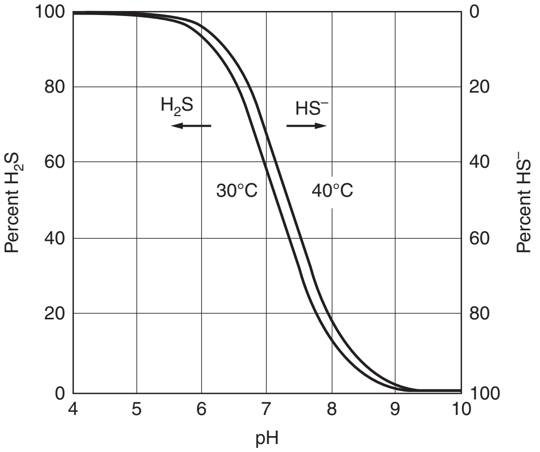
Table 10–11

Acid equilibrium constants for ammonia (NH_3) and hydrogen sulfide (H_2S)

Temperature, °C	$K_{NH_3} \times 10^{10}$, mole/L	$K_{H_2S} \times 10^7$, mole/L
0	7.28	0.262
10	6.37	0.485
20	5.84	0.862
25	5.62	1.000
30	5.49	1.480
40	5.37	2.440

Figure 10–7

Fraction of hydrogen sulfide in H_2S form as function of pH.



Ammonia Toxicity

Ammonia toxicity may be of concern for anaerobic treatment of wastewaters containing high concentrations of ammonium or proteins and/or amino acids, which can be degraded to produce ammonium. Free ammonia (NH_3), at high enough concentrations, has been found to be toxic to aceticlastic methanogens (Angelidaki and Arhing, 1994; Steinhaus et al., 2007; Lu et al., 2008), but the hydrogenotrophic organisms appear to be less sensitive (Sprott and Patel, 1986). As described in Chap. 2, ammonia is a weak acid and dissociates in water to form ammonium (NH_4^+) and hydroxyl ions. The amount of free ammonia is a function of temperature and pH. Dissociation constants for NH_3 as a function of temperature are given in Table 10–11. Based on the constants in Table 10–11, the free ammonia content decreases from 1.8 to 1.7 percent at a pH of 7.5 as the temperature increases from 20 to 35°C. At a pH of 7.8, the range is from 3.5 to 3.3 percent. Thus, the total ammonium- plus ammonia-N concentration (TAN) that can be tolerated is a function of the free ammonia toxicity concentration. Reported values for the threshold of NH_3-N toxicity range from 100 mg/L (McCarty, 1964) to 250 mg/L (Garcia and Angenent, 2009; Wilson et al., 2012). The higher values have been observed with more acclimation time, which may be due to acclimation by the aceticlasts or changes in the anaerobic process population with a shift to organisms more tolerant of free ammonia. These observations suggest that TAN concentrations of 3000 to 7000 mg/L may be tolerated in anaerobic treatment processes. With long-term acclimation non-inhibitory TAN concentrations of about 4000 mg/L have been reported by Garcia and Angement (2009) and 5000 to 8000 mg/L by van Velsen (1977) and Parkin and Miller (1982).

10–5 DESIGN CONSIDERATIONS FOR IMPLEMENTATION OF ANAEROBIC TREATMENT PROCESSES

Design of anaerobic treatment processes involves identification of process elements, determination of organic loading rate and other design parameters, and considerations for the conditions specific to anaerobic processes. When proprietary processes are considered, engineers must identify the quality and quantity of the waste stream to be treated, as well as the treatment goals. The purpose of this section is to provide a brief review of design parameters and considerations necessary to implement anaerobic treatment processes. The topics covered are (1) treatment efficiency needed, (2) general process design parameters and anaerobic degradation kinetics, and (3) other implementation issues.

Treatment Efficiency Needed

Anaerobic treatment processes are capable of very high COD conversion efficiency to methane production with minimal biomass production and relatively short hydraulic retention time reactors compared to aerobic treatment. Operation at high SRT values, greater than 20 to 50 d is possible with granular sludge, packed bed, and anaerobic membrane reactors, to provide maximum conversion of solids at temperatures above 25°C. However, the ability to meet secondary effluent discharge standards by anaerobic treatment alone is limited by high effluent suspended solids (50 to 150 mg/L) and a residual volatile fatty acid concentration that are common for anaerobic processes. These issues are magnified for anaerobic treatment at temperatures below 20°C. Some form of aerobic treatment is necessary to provide effluent polishing, either attached growth or suspended growth processes (Chong et al., 2012). For high strength wastewaters the combination of anaerobic and aerobic treatment can be very economical in terms of both capital and operating costs (Obayashi et al., 1981).

General Process Design Parameters

In the introduction to this chapter it was noted that the anaerobic digestion of wastewater treatment plant sludges is considered in Chap. 13. The material presented below is provided to serve as general guidance for comparing commercially available anaerobic technologies. Important design parameters used to size anaerobic processes include organic loading rate, hydraulic loading rate, and SRT. Biological kinetics and stoichiometric relationships between substrate removal and biomass growth are also useful for determining excess solids production and soluble substrate concentrations.

Organic Loading Rate. The volumetric organic loading rate (OLR) is the key design parameter used to determine the reactor volume for granular sludge and attached growth anaerobic treatment processes. Generally, organic loading rates for anaerobic treatment processes are significantly higher than those used for aerobic processes. Organic loading rates used in anaerobic processes can vary from 1 to 50 kg COD/m^3·d, whereas typical volumetric organic loading rates used for aerobic processes will vary from 0.5 and 3.2 kg COD/m^3·d.

Organic loading rates are affected by the type of anaerobic process used, type of wastewater, and temperature. A range of organic loading rates are included with the process descriptions for commonly used anaerobic processes in Table 10–3. The wide range of organic loading rates reported for a given anaerobic treatment technology is illustrated in a comparison of organic loading rates for the UASB, EGSB, and anaerobic contact processes on Fig. 10–8. The values for the UASB and EGSB processes are of a similar

Figure 10–8

Range of organic loading rates reported for anaerobic treatment of various wastewaters with UASB, EGSB, and anaerobic contactor processes. (Data from Tauseef et al., 2013.)

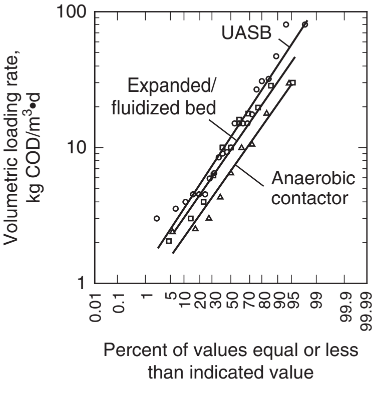


Table 10–12

Recommended UASB reactor SRTs for stable operation for treatment of domestic wastewater (van Lier et al., in Henze et al., 2011)

Temperature, °C	SRT, d
35	25
30	30
25	60
20	100
15	140

magnitude and generally higher than that used for anaerobic contactor reactors. The use of higher loadings for the granular sludge systems are in part due to the ability to carry a very high biomass concentration in the reactor as a result of the high biomass density developed in the granular sludge. The mixing regime for the EGSB and IC granular sludge systems also promote higher mass transfer of soluble substrate into the biofilm. The lower loading rates for the anaerobic contactor process may also be related to the more frequent application of this process for wastewaters with high solids concentration. The solids can limit the sludge concentration possible in the anaerobic reactor to thus require a larger reactor volume for a sufficient SRT for the necessary solids destruction, which then results in a lower organic loading rate.

Solids Retention Time. The solids retention time is a fundamental parameter for all anaerobic processes. The performance of an anaerobic treatment process is related to SRT in three ways: (1) from the basic biological kinetics discussed in Sec. 7–6, soluble substrate concentration is lower with increased SRT; (2) a higher SRT provides a greater mass of methanogenic organisms in the reactor, which is then better able to handle variations in process operating conditions to maintain a stable balance between volatile fatty acid production and utilization; and (3) the hydrolytic destruction efficiency of solids fed to an anaerobic reactor is increased at higher SRTs. At 30°C, SRT values greater than 20 d are needed for anaerobic processes for effective treatment performance. At lower temperatures much longer SRT values are needed. Recommended SRT values as a function of temperature for stable treatment of domestic wastewater by a UASB process are given in Table 10–12. Lower SRTs can be used for wastewaters that contain mainly soluble substrates.

Because of the nature of the biomass accumulation in granular sludge and attached growth anaerobic systems, it is impossible to predict the biomass concentration and SRT accurately. In addition, the substrate removal rate is not just a function of SRT as in a completely-mixed suspended growth reactor, but is also a function of substrate diffusion, mass transfer, and biofilm characteristics. Thus, the organic loading rate is the primary design parameter used for these systems, with the recognition that longer SRTs are commensurate with lower OLR values.

The use of SRT as design parameter is feasible for completely-mixed reactors such as the anaerobic contact process [see Table 10–3(f)] and the anaerobic membrane reactor [see Table 10–3(i)]. In such cases, the effect of the SRT on the effluent soluble substrate concentrations can be determined using Eq. (7–56) and appropriate biokinetic coefficients. The final step in anaerobic processing of organic material is the conversion of acetate and hydrogen to methane. The reactor acetate concentration is affected by the SRT and biokinetics of the acetoclastic methanogenic organisms. Biokinetic information for methanogenesis is summarized in Table 10–13 for different temperatures. Values for synthesis yield and specific endogenous decay coefficient coefficients for fermentation and methanogenesis

Table 10–13

Summary of biokinetic coefficients and methane stoichiometry for anaerobic treatment reactions[a]

Parameter	Unit	Value Range	Value Typical
Synthesis yield, Y_H			
Fermentation	g VSS/g COD	0.06–0.12	0.10
Methanogenisis	g VSS/g COD	0.02–0.06	0.04
Overall combined	g VSS/g COD	0.05–0.10	0.08
Decay coefficient, b_H			
Fermentation	g/g·d	0.02–0.06	0.04
Methanogenisis	g/g·d	0.01–0.04	0.02
Overall combined	g/g·d	0.02–0.04	0.03
Maximum specific growth rate, μ_m			
35°C	g/g·d	0.03–0.38	0.35
30°C	g/g·d	0.22–0.28	0.25
25°C	g/g·d	0.18–0.24	0.20
Half-velocity constant, K_S	mg COD/L	60–500	120
Methane			
Production at std. conditions	m^3/kg CODr	—	0.35
Content of gas	%	60–70	65
Energy content at std.conditions	kJ/m^3	—	38,846

[a] Tchobanoglous et al. (2003); Batstone et al. (2002).

are also included in Table 10–13. The methanogenic biokinetic values shown are based on *Methanosaeta* as the dominant organisms for acetoclastic methanogenesis, which are very common. However, under certain conditions, such as highly loaded anaerobic reactors, *Methanosarcina* may dominate. It has been shown that the maximum growth rate of *Methanosarcina* is about 2.5 times that for *Methanosaeta* at 35°C and its half-velocity coefficient for acetate utilization is about 3.5 times greater (Conklin et al., 2006).

The information given in Table 10–13 can also be used to approximate the amount of biomass and excess sludge production for completely-mixed anaerobic reactors, and for other types of anaerobic processes, if the SRT values can be estimated The use of SRT for the design of an anaerobic contact process is illustrated in Example 10–4 in Sec. 10–6.

Hydraulic Retention Time. The hydraulic retention time is directly related to the anaerobic process organic loading rates and wastewater strength:

$$\tau = S_o/\text{OLR} \tag{10–17}$$

Wastewaters with higher substrate concentration will require a longer hydraulic retention time for a given design OLR.

Process Implementation Issues

In addition to the process design consideration discussed above a number of other issues must be addressed when considering the implementation of anaerobic treatment technologies. Important issues include: solids separation, temperature management, corrosion control, and odor management. Additional issues are identified in Table 10–14.

Table 10–14

Design issues for anaerobic treatment of industrial wastewaters[a]

Design issue	Remarks
Flow equalization	
Pretreatment	TSS removal, FOG removal
Anaerobic process	
Corrosion control	Material selection
Temperature control	Preheating of waste stream, reactor heating
Chemical addition	Alkalinity control, nutrient, struvite control
Odor control	Odor removal processes (Chap. 16)
Sludge wasting	
Gas collection	Gas storage, cleaning, utilization (Chap. 17)

[a] Adapted, in part, from Totzke (2012).

Liquid-Solids Separation. Solids separation is a critical element in the application of anaerobic processes. With respect to energy intensity, liquid sludge separation processes can vary from simple evaporation and/or decanting in sludge lagoons to centrifugation and electro-dewatering, depending on the application. A range of dewatering options for digested wastewater sludge, as well as other types of sludge, is considered in Sec. 14–2 in Chap. 14. Other solids separation processes that are an integral part of one or more of the processes identified in Table 10–3 are conventional sedimentation and membranes. Where sedimentation is used some form of degasification will be needed.

Temperature Management. Anaerobic microorganisms responsible for the anaerobic treatment processes are in general temperature sensitive and anaerobic processes are often maintained at a significantly higher temperature than the ambient temperature, in the range of 30 to 36°C, or higher. Heating of the influent and the reactor may be achieved by using a heater directly in the reactor vessel, or using heat exchangers to preheat the influent. Heat in the effluent flow is used commonly to preheat the influent through a heat exchanger. Digester heating is considered in detail in Sec. 13–9 in Chap. 13.

Corrosion Control. Anaerobic treatment usually generates hydrogen sulfide and H_2S, both of which can causes corrosion of various materials. Hydrogen sulfide in the anaerobic conditions can be converted biologically to sulfuric acid in the presence of moisture, also resulting in corrosion from sulfuric acid (U.S. EPA, 1991). Appropriate materials must be used for the equipment exposed to the anaerobic condition and the pipes to convey biogas containing hydrogen sulfide. Hydrogen sulfide removal in the gas phase is practiced commonly for odor control, and it is considered in Chap. 16. The use of materials such corrosion resistant concrete, fiber-reinforced plastic (FRP), polyvinyl chloride (PVC), high density polyethylene (HDPE) and a variety of similar materials and coatings in parts of the process exposed to corrosive gases must be considered to minimize the potential risk of corrosion and minimize the use of chemicals for corrosion control.

Odor Management. Much of the early impetus for contained anaerobic process applications was related to complying with waste discharge regulations and minimizing odor complaints. With development of techniques to contain and remove odorous gases, odor has become less of the impetus but the odor control unit is a critical process element for the anaerobic treatment processes. As with the issue of corrosion, hydrogen sulfide is the most common odorous compound found in anaerobic processes, although a variety of mercaptan based odors have also been reported. The threshold values for a variety of odors are given in Sec. 16–3 in Chap. 16.

The principal methods used for the control of odors from anaerobic processes include the use of compost filters (biofilters), direct oxidation, carbon adsorption, acid scrubbing, and biotrickling filters. Relatively inexpensive, compost filters are generally used for smaller plants. In some wastewater treatment plants, odorous gases are oxidized by introducing them into sparged aeration diffusers in the activated sludge process. Views of carbon adsorption, acid scrubbing, and biotrickling filter facilities are shown on Fig. 9–9 in Chap. 9. All of these options for odor control are considered in greater detail in Sec. 16–3 in Chap. 16.

10–6 PROCESS DESIGN EXAMPLES

With the exception of the anaerobic lagoon process and anaerobic digesters used for wastewater treatment sludge, the three most common types of commercial anaerobic processes are: (1) the upflow anaerobic sludge blanket process, (2) the expanded granular sludge blanket process, and (3) the anaerobic contact process (see Table 10–4). The complete mix process which corresponds to anaerobic digestion of wastewater treatment plant sludges is considered separately in Chap. 13, along with sludge processing in Chap. 14. The purpose of this section is to illustrate the design of the upflow anaerobic sludge blanket process and the anaerobic contact process. It should be noted that because most commercial anaerobic processes illustrated in Table 10–3 are proprietary, there is little opportunity for individual process design as the treatment units are provided as a package.

Upflow Anaerobic Sludge Blanket Process

The UASB process and its evolution to the EGSB and IC processes have been described in Sec. 10–3. Some commercial and domestic installations have been illustrated previously on Figs. 10-3, 10–4, and 10–5. Some additional UASB and UASB type installations are illustrated on Fig. 10–9. The purpose of this section is to discuss key process elements

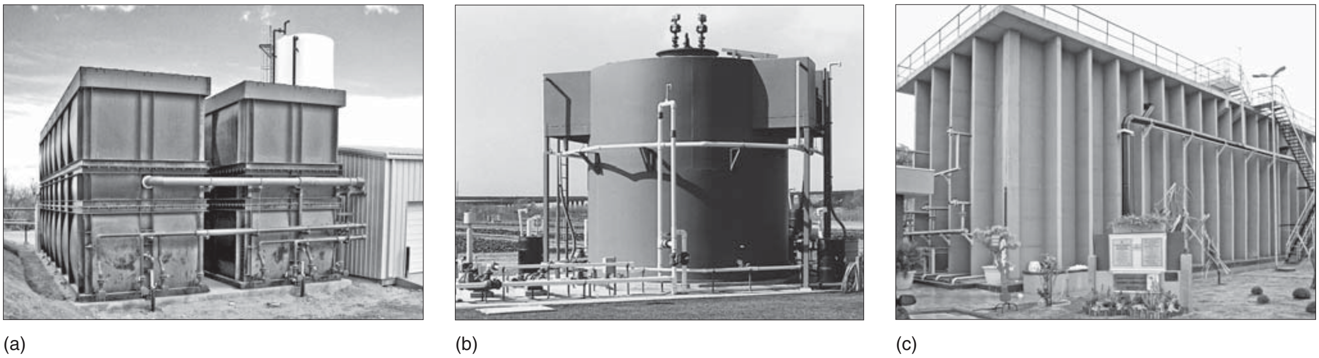

(a) (b) (c)

Figure 10–9

Granular sludge blanket anaerobic treatment systems: (a) UASB process in modular steel tanks, before insulation and hookup (courtesy of Robert Pharmer of Pharmer Engineering); (b) view of hybrid reactor equipped with internal packing above the granular sludge blanket [see Table 10-3(h)]. The exterior physical appearance of a UASB reactor without and with internal packing is similar; and (c) combined anaerobic-aerobic single-tank reactor treatment system (courtesy of Paques BV).

and process design considerations for granular sludge treatment systems. Important topics addressed are (1) the development and maintenance of the granular sludge, (2) physical design components, and (3) process design considerations.

Process Description. The heart of the UASB, EGSB, and IC, and similar processes is the development and maintenance of the dense granular sludge particles, which results in a high reactor biomass concentration. Because of the high biomass concentration these anaerobic processes can be operated at high organic loading rates. The granular sludge particle size is generally in the range of 1 to 2 mm but may range from 0.10 to 8 mm depending on the waste treated and hydraulic and gas shear forces. Particle densities are in the range of 1.0 to 1.05 g/L with settling velocities of 15 to 50 m/h (Henze et al., 2011). Because of the granulated floc formation, the solids concentration at the bottom of the reactor may range from 50 to 100 kg/m^3. Above the sludge blanket, a more diffused layer forms containing particles with lower settling velocities. The solids concentration in this layer may range from 10 to 30 kg/m^3 (Aiyuk et al., 2006).

From studies on the microbial composition of the granules it appears that the surface is made up of coccid bacteria, while rod-shaped *Methanosaeta* are dominant in the interior and provide a filamentous structural backbone for the granulation (O'Flaherty et al., 2006). The physical characteristics of the granules provide a complex microbial ecology with methane-producing organisms in close proximity to hydrogen- and acetic acid-producers. A specific methane production activity of 0.10 g COD/VSS·d in the granular particles has been reported by Seghezzo et al. (2001).

Development of Granular Sludge. Schmidt and Ahring (1996) describe a four-step process for the development of granular sludge: (1) attachment of cells to an uncolonized inert material or other cells, (2) initial adsorption of other colloidal or bacteria particles by reversible physiochemical forces, (3) irreversible attachment of microbial organisms due to microbial extracellular polymers, and (4) multiplication of cells from substrate diffusion into the granular structure. The substrate removal kinetics and mass transfer characteristics of the granular particles are similar to that described for biofilms in Sec. 7–7.

The development of a granular sludge bed can take many months, but this is normally avoided today by seeding with granular sludge wasted from other UASB reactors. The sludge blanket development is more rapid at higher temperatures (above 20°C) and with the presence of readily degradable soluble COD in the feed. The high upflow velocities washout unattached organisms and favor the growth of the dense granular particles.

Impact of Wastewater Characteristics. The development and maintenance of granular sludge is affected by the wastewater characteristics. Granulation is very successful with high carbohydrate or sugar wastewaters, but less so with wastewaters high in protein resulting in a more fluffy floc instead (Thaveesri et al., 1994). Other factors affecting the development of granulated solids are pH, divalent cations and nutrient addition (Annachhatre, 1996). The pH should be maintained near 7.0 and a recommended COD/N/P ratio during startup is 300:5:1, while a lower ratio, 600:5:1, can be used during steady-state operation. Studies have shown that within certain concentrations, ferrous iron and calcium can enhance granular sludge formation; about 300 mg/L for Fe^{2+} and about 250 mg/L for Ca^{2+} (Yu et al., 2000; Yu et al., 2001).

The presence of suspended solids in the wastewater can adversely affect granular sludge formation and density, and the affects are more pronounced at lower temperatures

Table 10–15

Recommended design considerations for the gas-solids separators for UASB reactors[a]

1. The slope of the settler bottom, i.e., the inclined wall of the gas collector, should be between 45 to 60°.
2. The surface area of the apertures between the gas collectors should not be smaller than 15 to 20 percent of the total reactor surface area.
3. The height of the gas collector should be between 1.5 to 2 m at reactor heights of 5 to 7 m.
4. A liquid gas interface should be maintained in the gas collector in order to facilitate the release and collection of gas bubbles and to combat scum layer formation.
5. The overlap of the baffles installed beneath the apertures should be 100 to 200 mm to avoid upward flowing gas bubbles entering the settler compartment.
6. Generally scum layer baffles should be installed in front of the effluent weirs.
7. The diameter of the gas exhaust pipes should be sufficient to guarantee the easy removal of the biogas from the gas collection cap, particularly also in the case where foaming occurs.
8. Anti-foam spray nozzles should be installed in the upper part of the gas cap where the treatment of the wastewater is accompanied by heavy foaming.

[a] Adapted from Malina and Pohland (1992).

due to slower solids hydrolysis rates and solids accumulation (Letting and Hulshoff-Pol, 1991; Elmitwalli et al., 2002). For wastewaters with higher influent suspended solids and with lower temperatures, a two-step process, with a UASB reactor at a lower upflow velocity for solids capture and hydrolysis prior to a final UASB reactor, may provide more stable performance and operation flexibility. At certain feed solids concentrations (above 6 g TSS/L), an anaerobic contract process may be more appropriate.

Physical Design Considerations. The main physical design considerations are the feed inlet, gas separation, gas collection, and effluent withdrawal. The gas separation and effluent withdrawal functions are done with specifically designed gas-solids separators. Specific design features of these components are provided by the suppliers of proprietary upflow granular sludge anaerobic treatment processes. Gas-solids separator designs are located at the top of the liquid layer overlying the sludge blanket zone. Designs are employed to direct and trap the rising gas bubbles into hoods or collection zones [see Table 10–3(b) and (c)], and to capture the effluent from quiescent zones that allow solids settling and return to the reactor process volume below. Methods applied to improve effluent solids capture include a UASB-hybrid design with plastic packing at the top of the reactor (Tauseef et al., 2013) and the use of lamella-type plates in the effluent settling zone (Gomec, 2010). Design considerations for gas-solids separators are listed in Table 10–15.

The feed inlet designs must provide uniform flow distribution across the bottom of the granular sludge treatment process column to avoid channeling or formation of dead zones. The distribution is even more critical for designs with lower organic loading rates, as there would be less gas production to help mix the reactor contents. As shown in Table 10–16, the recommended bottom area served by individual feed pipes is a function of the bottom sludge layer density and the organic loading rate.

Process Design Considerations. Specific process designs will be influenced by the type of upflow granular sludge process used and the experience with the particular wastewater by the anaerobic treatment system supplier. Selection of the upflow granular sludge process reactor volume will be controlled by either (1) the allowable upflow velocity

Table 10–16

Guidelines for area served by feed inlet pipes for UASB reactors as a function of sludge density and organic loading rate[a]

Sludge type	COD loading, kg/m³·d	Area per feed inlet, m²
Dense flocculent sludge, (> 40 kg TSS/m³)	< 1.0	0.5–1
	1–2	1–2
	> 2	2–3
Medium flocculent sludge, (20 to 40 kg TSS/m³)	< 1–2	1–2
	> 2	2–5
Granular sludge	1–2	0.5–1.0
	2–4	0.5–2.0
	> 4	> 2.0

[a] Adapted from Lettinga and Hulshoff Pol (1991).

or (2) the organic loading rate (OLR). Limiting values for these parameters are affected by temperature and the type of wastewater being treated.

Upflow Velocity. The upflow velocity, based on the influent flowrate, is a critical design parameter. Design upflow velocities recommended for UASB reactors are shown in Table 10–17. Upflow velocities would be much higher in EGSB and IC reactors, which are more likely applied for higher strength industrial wastewaters. When the UASB reactor is applied for domestic wastewater treatment or for wastewater with higher influent solids concentrations, a lower velocity is needed to better retain the solids to provide sufficient time for solids capture and reduction by hydrolysis. The maximum allowable upflow velocity determines the cross-sectional area of the reactor, which is the feed rate divided by the upflow velocity:

$$A = \frac{Q}{v} \tag{10–18}$$

where v = maximum design upflow superficial velocity, m/h
A = reactor cross-section area, m^2
Q = influent flowrate, m^3/h

The reactor process volume is equal to the cross-sectional area times the process reactor height (H) where V_v = reactor volume controlled by the maximum upflow velocity, m^3.

$$V_v = H(A) \tag{10–19}$$

The total reactor height is greater than this, as additional depth is added at the top for the gas-solids separator zone. The process volume must be large enough so that the allowable

Table 10–17

Design upflow velocities and reactor heights recommended for UASB reactors[a]

	Upflow velocity, m/h		Reactor height, m	
Wastewater type	Range	Typical	Range	Typical
COD near 100% soluble	1.0–3.0	1.5	6–10	8
COD partially soluble	1.0–1.25	1.0	3–7	6
Domestic wastewater	0.8–1.0	0.7	3–5	5

[a] Adapted from Lettinga and Hulshoff Pol (1991).

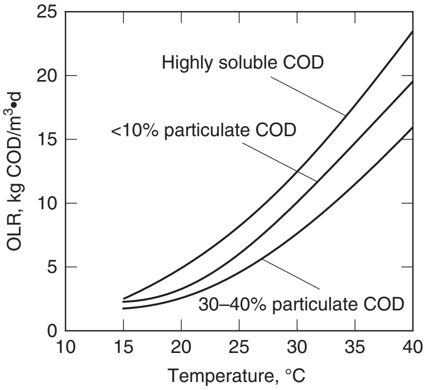


Figure 10–10

Effect of temperature on organic loading rate for UASB process (Henze et al., 2011).

reactor organic loading rate is not exceeded. The reactor design is ultimately controlled by either the maximum allowable upflow velocity or by the required organic loading rate. For weaker wastewaters the upflow velocity will be the controlling parameter.

Organic Loading Rate. As discussed in Sec. 10–5, a wide range of organic loading rates have been applied for upflow granular sludge processes, depending on the wastewater characteristics, reactor type, and temperature. In addition, Lettinga and Hulshoff-Pol (1991) point out that higher organic loadings can be used if higher effluent TSS concentrations are acceptable and also that higher loadings can be used for wastewaters with a lower fraction of particulate COD. As shown in Table 10–3, organic loading rates for UASB systems may range from 5 to 15 kg COD/ m^3·d and from 10 to 40 kg COD/ m^3·d for EGSB systems. An example of the effect of temperature on organic loading rate is shown on Fig. 10–10. For a highly soluble wastewater the organic loading rate is reduced by a factor of 5.0 for operation at 15 versus 30°C and by a factor of 4.3 for a wastewater with 30–40% particulate COD.

The reactor process volume is related to the organic loading rate as follows:

$$V_{\mathrm{OLR}} = \frac{Q(S_o)}{\mathrm{OLR}} \tag{10–20}$$

where V_{OLR} = reactor process volume controlled by the organic loading rate, m^3. The process design for a UASB treatment process is illustrated in Example 10–3.

EXAMPLE 10–3 **UASB Treatment Process Design** Determine the following for a UASB treatment process used to treat a sugar beet wastewater, with the characteristic given below, to achieve 90 percent COD removal:

a. Reactor process volume
b. Process hydraulic retention time
c. Reactor dimensions
d. Reactor SRT
e. Daily sludge production rate in kg VSS/d
f. Excess sludge waste volume in m^3/d
g. Methane gas production rate in m^3/d
h. Total gas production rate in m^3/d
i. Energy available from methane production in kJ/d
j. Alkalinity requirements

Wastewater characteristics:

Item	Unit	Value
Flowrate	m^3/d	500
COD	g/m^3	12,000
TSS	g/m^3	600
nbVSS	g/m^3	500
Alkalinity	g/m^3 as $CaCO_3$	500
Temperature	°C	25

Use the design parameters given below and typical values from Table 10–13:

1. From Table 10–14
   $Y_H = 0.08$ g VSS/g COD
   $b_H = 0.03$ g VSS/g VSS·d,
2. $f_d = 0.10$ g VSS cell debris/g VSS biomass decay
3. Methane production at 0°C = 0.35 L CH_4/g COD
4. Energy content of methane at 0°C = 38,846 kJ/m^3
5. Percent methane in gas phase = 65%
6. Height of reactor process volume = 8 m
4. Height of clear zone above the sludge blanket = 0.50 m
8. Height of gas-solids separator = 2.5 m
9. Reactor length:width ratio = 2.0
10. Maximum reactor upflow velocity = 1.0 m/h
11. Average solids concentration in process volume = 30 kg VSS/m^3

Based on data from the treatment of sugar beet wastewater in other UASB facilities, 90 percent COD removal at 25°C, can be achieved with a design organic loading rate of 8.0 kg COD/m^3·d. The wastewater is mainly soluble containing carbohydrate compounds, and a granular sludge is expected. Assume an effluent VSS concentration of 120 g/m^3.

Solution

1. Determine the reactor process volume.
   a. Determine the reactor volume based on the maximum upflow velocity Eq. (10–18) and Eq. (10–19)

$$A = \frac{Q}{v} = \frac{(500\ \text{m}^3/\text{d})}{(1.0\ \text{m/h})(24\ \text{h/d})} = 20.8\ \text{m}^2$$

$$V_v = A(H) = 20.8\ \text{m}^2(8\ \text{m}) = 166.7\ \text{m}^3$$

   b. Determine the reactor volume based on the organic loading rate.
   From Eq. (10–20)

$$V_{\text{OLR}} = \frac{Q\,S_o}{\text{OLR}} = \frac{(500\ \text{m}^3/\text{d})(12\ \text{kg COD/m}^3)}{(8.0\ \text{kg COD/m}^3\cdot\text{d})} = 750\ \text{m}^3$$

   The organic loading rate controls the reactor volume design.
2. Determine the process hydraulic retention time.

$$\frac{V}{Q} = \frac{750\ \text{m}^3}{(500\ \text{m}^3/\text{d})} = 1.5\ \text{d}$$

3. Determine the reactor dimensions.
   a. Reactor area = $(L)(W) = 2W(W) = 2W^2$

$$\text{Area} = \frac{V}{H} = \frac{750\ \text{m}^3}{8\ \text{m}} = 93.75\ \text{m}^2$$

$2W^2 = 93.75\ \text{m}^2$, $W = 6.85$ m, $L = 13.7$ m

b. Total reactor height

H_T = processing hgt + clear zone hgt + separator hgt

$H_T = 8\ \text{m} + 0.5\ \text{m} + 2.5\ \text{m} = 11\ \text{m}$

Reactor dimensions = 13.7 m × 6.85 m × 11 m

4. Determine the reactor SRT.

a. From Eq. (7–56), $X(V) = P_x\ \text{SRT}$

b. From Eq. (8–20),

$$P_x = \frac{Q(Y_\text{H})(S_o - S)}{1 + b_\text{H}(\text{SRT})} + \frac{f_d b_\text{H}(Q)(Y_\text{H})(S_o - S)(\text{SRT})}{1 + b_\text{H}(\text{SRT})} + (\text{nbVSS})Q$$

c. Substituting Eq. (8–20) into Eq. (7–56),

$$X_\text{VSS}(V) = \frac{Q(Y_\text{H})(S_o - S)(\text{SRT})[1 + f_d b_\text{H}(\text{SRT})]}{1 + b_\text{H}(\text{SRT})} + (\text{nbVSS})\text{Q}(\text{SRT})$$

$$S_o - S = 0.90\ S_o = 0.90(12{,}000\ \text{g COD/m}^3)$$

$$S_o - S = 10{,}800\ \text{g COD/m}^3$$

$$(30{,}000\ \text{g VSS/m}^3)(750\ \text{m}^3) =$$

$$\frac{(500\ \text{m}^3/\text{d})(0.08\ \text{g VSS/g COD})(10{,}800\ \text{g COD/m}^3)(\text{SRT})[1 + 0.10(0.03\ \text{g/g}\cdot\text{d})(\text{SRT})]}{1 + (0.03\ \text{g/g}\cdot\text{d})(\text{SRT})}$$

$$+\ 500\ \text{g VSS/m}^3\ (500\ \text{m}^3/\text{d})\text{SRT}$$

Solving: SRT = 50.2 d

5. Determine the daily sludge production rate from Eq. (7–56).

$$P_{X,\text{VSS}} = \frac{X_\text{VSS}(V)}{\text{SRT}}$$

$$= \frac{(30{,}000\ \text{g VSS/m}^3)(750\ \text{m}^3)(1\ \text{kg}/10^3\ \text{g})}{50.2\ \text{d}}$$

$$P_{X,\text{VSS}} = 448.2\ \text{kg VSS/d}$$

6. Determine the excess sludge daily waste volume.

$$P_{X,\text{VSS}} = Q(X_e) + (X)Q_W$$

$$Q_W = \frac{P_{X,\text{VSS}} - Q(X_e)}{X}$$

$$= \frac{(448{,}200\ \text{g VSS/d}) - (500\ \text{m}^3/\text{d})(0.120\ \text{g VSS/m}^3)}{(30{,}000\ \text{g VSS/m}^3)}$$

$$Q_W = 14.9\ \text{m}^3/\text{d}$$

7. Determine the methane gas production rate by COD balance.

COD removal = methane COD + biomass COD

$$P_{X,\text{bio}} = P_{X,\text{VSS}} - \text{nbVSS(Q)}$$

$$P_{X,\,\text{bio}} = 448{,}200 \text{ g VSS/d} - 500 \text{ g VSS/m}^3 \text{ (500 m}^3\text{/d)}$$

$$P_{X,\,\text{bio}} = 448{,}200 \text{ g VSS/d} - 250{,}000 \text{ g VSS/d}$$

$$= 198{,}200 \text{ g VSS/d}$$

Methane COD = COD removed − biomass COD

CH_4 COD/d

$$= 500 \text{ m}^3\text{/d}(10{,}800 \text{ g COD/m}^3) - 1.42 \text{ g COD/g VSS (198,200 g VSS/d)}$$

$$CH_4 \text{ COD} = 5{,}118{,}556 \text{ g } CH_4 \text{ COD/d}$$

At standard conditions, methane production rate =

$$(5{,}118{,}556 \text{ g } CH_4 \text{ COD/d})(0.35 \text{ L } CH_4\text{/g COD})(1 \text{ m}^3/10^3 \text{ L})$$

$$= 1719.5 \text{ m}^3 \text{ } CH_4\text{/d at } 0°\text{C}$$

methane production rate at 25°C =

$$(1719.5 \text{ m}^3 \text{ } CH_4\text{/d})\frac{(273.15 + 25)°\text{C}}{273.15°\text{C}} = 1955 \text{ m}^3 \text{ } CH_4\text{/d}$$

8. Determine the total gas production rate; Percent methane = 65%

$$\text{Total gas production rate} = \frac{(1955 \text{ m}^3 \text{ } CH_4\text{d})}{(0.65 \text{ m}^3 \text{ } CH_4\text{/m}^3 \text{ gas})} = 3008 \text{ m}^3 \text{ gas/d}$$

9. Energy content of methane production

$$\text{Energy} = (38{,}846 \text{ kJ/m}^3)(1719.5 \text{ m}^3 \text{ } CH_4\text{/d}) = 66.8 \times 10^6 \text{ kJ/d}$$

10. Determine alkalinity requirements.

Assume pH = 7.0

From Table 10–7 at pH = 7.0, T = 25°C,

percent CO_2 = 35%, alkalinity = 2678 g/m^3 as $CaCO_3$

Influent alkalinity = 500 g/m^3 as $CaCO_3$

$$\text{Alkalinity needed} = (2678 - 500)\text{g/m}^3 \text{ as } CaCO_3$$

$$= 2178 \text{ g/m}^3 \text{ as } CaCO_3$$

$$\text{Alkalinity in kg/d} = (2178 \text{ g/m}^3)(500 \text{ m}^3\text{/d})(1 \text{ kg}/10^3 \text{ g})$$

$$= 1098 \text{ kg/d}$$

11. Summary of results.

Parameter	Unit	Value
Reactor process volume	m^3	750.0
Reactor total height	m	11.0
Reactor LxW	m	13.7 × 6.85
Hydraulic retention time	d	1.5

(continued)

Parameter	Unit	Value
SRT	d	50.2
Excess waste sludge	m^3/d	14.9
Total gas production rate	m^3/d	3008
Methane production rate	m^3/d	1955
Energy production rate	kJ/d	66.8×10^6
Alkalinity needed as $CaCO_3$	kg/d	1089

Comment A significant amount of energy is produced daily in the form of methane gas. If the gas can be used as an energy source by the industrial facility, it could help to offset the cost of adding a considerable amount of alkalinity to maintain the anaerobic reactor pH near 7.

Anaerobic Contact Process

The anaerobic contact process is a completely-mixed reactor system with suspended/flocculating anaerobic biomass and a liquid-solids separation step for biomass capture and recycle.

Process Description. The process flow diagram for the ANCP is illustrated in Table 10–3(f). As shown, the anaerobic contact process overcomes the disadvantages of equal values of hydraulic retention time and SRT in the completely mixed process. Biomass is separated and returned to the completely mixed or contact reactor so that the process SRT is longer than the hydraulic retention time. By separating the hydraulic retention time and SRT, the anaerobic reactor volume can be reduced. Gravity separation is the most common approach for solids separation and thickening prior to sludge recycle, however, the success is dependent on the settling properties of the anaerobic reactor solids.

Because the reactor sludge contains gas produced in the anaerobic process and gas production can continue in the separation process, solids-liquid separation can be inefficient and unpredictable. Various methods have been used to minimize the effect of trapped gas bubbles in the sludge settling step. These include gas stripping by agitation or vacuum degasification, inclined plate separators, and the use of coagulant chemicals. Clarifier hydraulic application rates range from 0.5 to 1.0 m/h. Practical reactor MLVSS concentrations are 4000 to 8000 mg/L (Malina and Pohland, 1992). Organic loading rates are typically $\leq$ 4.0 kg COD/m^3·d and SRTs range from 15 to 30 d as indicted in Table 10–3(f).

Design Considerations for Anaerobic Contact Process. The anaerobic contact process may be designed in a manner similar to completely mixed aerobic activated sludge processes, because the hydraulic regime and biomass concentration can be reasonably defined. The design procedure is as follows:

1. Select an SRT to achieve a given effluent concentration and percent COD removal.
2. Determine the daily solids production and mass of solids in the system to maintain the desired SRT.
3. Select the expected solids concentration in the reactor and determine the reactor volume.
4. Determine the gas production rate.
5. Determine the amount of excess sludge wasted and the nutrient needs.
6. Check the volumetric organic loading rate.
7. Determine alkalinity needs.

These same design considerations apply to other types of anaerobic treatment processes, with the main difference being the need to rely on using organic loading rates instead of SRT to size the reactor volume.

Information in Table 10–13 provides a summary of kinetic coefficients and other design values that may be used to design an anaerobic contactor process treating wastewater that is comprised of mostly soluble biodegradable COD. For wastewater with high solids concentrations, the design methods presented in Chap. 13 for anaerobic digestion may be applicable. Laboratory treatability studies or pilot plant testing is normally recommended for wastewater containing soluble and particulate constituents that are significantly different from past experience with anaerobic degradation of other wastewaters. The design of an anaerobic contact process is presented in Example 10–4.

EXAMPLE 10–4 **Suspended Growth Anaerobic Contact Reactor Process** Determine the reactor volume and hydraulic retention time, the gas production rate and energy available, the solids production rate, and the amount of alkalinity and nutrient addition needed for an anaerobic contact process [see Table 10–3(f)] treating the following wastewaters to achieve 90 percent COD removal.

Wastewater characteristics:

Item	Unit	Value
Flowrate	m^3/d	500
COD	g/m^3	6000
Soluble COD	g/m^3	4000
COD/VSS ratio	g/g	1.8
Degradable fraction of VSS	%	80
Nitrogen	g/m^3	10
Phosphorus	g/m^3	20
Alkalinity	g $CaCO_3/m^3$	500
Temperature	°C	25

Design assumptions:

1. Effluent VSS concentration = 150 g/m^3.
2. Factor of safety for design SRT = 3.0
3. f_d = 0.15 g VSS cell debris /g VSS biomass decay
4. At SRT $\geq$ 30, >99% of degradable VSS is transformed
5. MLVSS = 6000 g/m^3
6. Settling rate = 24 m/d
7. Gas composition = 65% CH_4 and 35% CO_2
8. Use kinetic coefficients and methane production assumptions from Table 10–13.
9. Biomass nutrient content = 12% N and 2.4% P

Solution

1. Determine design SRT at 25°C.

   At 90 percent COD removal the effluent COD is:

   = (1.0 – 0.90) (6000 mg/L) = 600 g/m^3

   The assumed effluent VSS concentration equals 150 g/m^3.

Effluent COD from VSS = (150 g/m^{3}L) 1.8 g COD/g VSS

= 270 g/m^3

Allowable effluent soluble COD = (600 – 270) g/m^3

= 330 g/m^3

Solving for SRT in Eq. (7–46) and substituting $\mu_{\max} = Y_H k$;

$$S = \frac{K_s[1 + b_H(\text{SRT})]}{\text{SRT}(\mu_{\max} - b_H) - 1}$$

$$\text{SRT} = \left[\frac{\mu_{\max}(S)}{K_s + S} - b_H\right]^{-1}$$

Use kinetic coefficients from Table 10–13,

$\mu_{\max}$ = 0.20 g/g·d

K_s = 120 g/m^3

b_H = 0.03 g/g·d

$$\text{SRT} = \left[\frac{(0.20\text{ g/g·d})(330\text{ g COD/m}^3)}{(120 + 330)\text{g COD m}^3} - (0.03\text{ g/g·d})\right]^{-1} = 8.6\text{ d}$$

With a safety of 3.0

Minimal design SRT = 3.0 (8.6) = 25.7 d

Use SRT = 30 d to complete degradable VSS transformation

2. Determine sludge production rate.

Calculate nondegraded VSS concentration

Nonsoluble COD = (6000 − 4000) g/m^3

= 2000 g/m^3

Nonsoluble COD as VSS = (2000 g/m^3 COD) / (1.8 g COD/g VSS)

= 1110 g/m^3 VSS

Degradable fraction of VSS = 0.8 (given)

Nondegraded VSS = 0.20 (1110) = 222 g VSS/m^3

Use Eq. (8–20) to determine solids production:

$$P_{X,\text{VSS}} = \frac{Q(Y_H)(S_o - S)}{1 + b_H(\text{SRT})} + \frac{f_d b_H(Q)(Y_H)(S_o - S)(\text{SRT})}{1 + b_H(\text{SRT})} + (\text{nbVSS})Q$$

$S_o - S$ = COD degraded

= Influent COD − nondegraded VSS COD − effluent soluble COD

= 6000 g COD/m^3 − 222 g VSS/m^3 −330 g COD/m^3 = 5270 g COD/m^3

Use following coefficients from Table 10–14 and assume $f_d = 0.15$

Y_H = 0.08 g VSS/g COD

b_H = 0.03 g/g·d

$$P_{X,\text{VSS}} = \frac{Q(Y_H)(S_o - S)[1 + f_d b_H(\text{SRT})]}{1 + b_H(\text{SRT})} + (\text{nbVSS})Q$$

$$P_{X,\text{VSS}} = \frac{(500\ \text{m}^3/\text{d})(0.08\ \text{g VSS/g COD})(5270\ \text{g COD/m}^3)[1 + 0.15(0.03\ \text{g/g}\cdot\text{d})(30.0\ \text{d})]}{[1 + (0.03\ \text{g/g}\cdot\text{d})(30.0\ \text{d})]}$$

$$+ (222\ \text{g VSS/m}^3)(500\ \text{m}^3/\text{d})$$

$$P_{X,\text{VSS}} = 125{,}925\ \text{g VSS/d} + 111{,}000\ \text{g VSS/d} = 236{,}925\ \text{g VSS/d}$$

3. Determine reactor volume and τ.
   a. Determine the volume using Eq (7–56).

$$V = \frac{(P_{X,\text{VSS}})\text{SRT}}{\text{MLVSS}} = \frac{(236{,}925\ \text{g VSS/d})(30\ \text{d})}{6000\ \text{g/m}^3} = 1184.6\ \text{m}^3$$

   b. Determine the hydraulic detention time, τ.

$$\tau = \frac{V}{Q} = \frac{1184.6\ \text{m}^3}{(500\ \text{m}^3/\text{d})} = 2.4\ \text{d}$$

4. Determine the methane and total gas production rate and energy production rates.
   a. Determine the methane gas production rate.

From Table 10–13, 0.35 $\text{m}^3\ \text{CH}_4$/kg COD at 0°C

COD removal = methane COD + biomass COD

biomass COD = (1.42 g COD/g VSS)($P_{x,\text{bio}}$)

$P_{x,\text{bio}}$ = first term in $P_{X,\text{VSS}}$ calculation = 125,925 g VSS/d

Methane COD = COD removed − biomass COD

CH_4 COD/d

$= 500\ \text{m}^3/\text{d}\ (5270\ \text{g COD/m}^3) - 1.42\ \text{g COD/g VSS}\ (125{,}925\ \text{g VSS/d})$

$= 2{,}456{,}186\ \text{g CH}_4\ \text{COD/d}$

At standard conditions, methane production rate =

$(2{,}456{,}186\ \text{g CH}_4\ \text{COD/d})(0.35\ \text{L CH}_4/\text{g COD})(1\ \text{m}^3/10^3\ \text{L})$

$= 859.7\ \text{m}^3\ \text{CH}_4/\text{d}$ at 0°C

Methane production rate at 25°C =

$$(859.7\ \text{m}^3\ \text{CH}_4/\text{d})\frac{(273.15 + 25)^\circ\text{C}}{273.15^\circ\text{C}} = 938.3\ \text{m}^3\ \text{CH}_4/\text{d}$$

   b. Determine the total gas production rate.

Gas composition = 65% methane (given)

$$\text{Total gas production rate at 25°C} = \frac{(938.3\ \text{m}^3\ \text{CH}_4/\text{d})}{(0.65\ \text{m}^3\ \text{CH}_4/\text{m}^3\ \text{gas})} = 1443.6\ \text{m}^3\ \text{gas/d}$$

(Note gas rate = 1443.6/500 = 2.9 times liquid flowrate.)

c. Determine the energy production rate.

From Table 10–13, energy content of methane = 38,846 kJ/m^3 at 0°C.

Energy production rate = (859.7 m^3 CH_4/d)(38,846 kJ/m^3) = 33.4 × 10^6 kJ/d

5. Determine nutrient requirements.

Biomass production rate = 125,925 gVSS/d

Given: biomass N = 12% and P = 2.4% of VSS

N required = (125,925)(0.12) = 15,111 g/d

P required = (125,925)(0.024) = 3022 g/d

Influent nutrients:

N = (10 g/m^3)(500 m^3/d) = 5000 g/d

P = (20 g/m^3)(500 m^3/d) = 10,000 g/d

There is sufficient phosphorus in the influent, but nitrogen must be added.

N addition = (15,111 – 5000) g N/d

= 10,111 g N/d

= 10.1 kg N/d

6. Determine alkalinity requirement.
From Table 10–7 at pH = 7.0, T = 25°C, percent CO_2 = 35, alkalinity = 2678 g/m^3 as $CaCO_3$

Influent alkalinity = 500 g/m^3 as $CaCO_3$

Alkalinity needed = (2678 − 500) g/m^3 as $CaCO_3$

= 2178 g/m^3 as $CaCO_3$

$$\text{As NaHCO}_3 = \left[\frac{(2178 \text{ g as CaCO}_3/\text{m}^3)}{(50 \text{ mg/meq CaCO}_3)}\right](84 \text{ mg NaHCO}_3/\text{meq}) = 3659 \text{ g NaHCO}_3/\text{m}^3$$

$NaHCO_3$/d = (3659 g/m^3)(500 m^3/d)(1 kg/10^3 g) = 1830 kg/d

7. Determine clarifier diameter.

(Assume degasifier used before clarifier)

$$\text{Area} = \frac{(Q, \text{m}^3/\text{d})}{(\text{settling rate, m/d})} = \frac{(500 \text{ m}^3/\text{d})}{(24 \text{ m/d})} = 20.83 \text{ m}^2$$

Diameter = 5.2 m

Comments A considerable amount of energy (i.e., kJ) is generated by the production of methane (CH_4). The methane could be used to heat the anaerobic process, which would provide more rapid degradation and, thus, reduce the anaerobic bioreactor size.

Use of Simulation Models

The relatively straightforward design procedures described above can be used to obtain a reasonable estimate of reactor volume requirements, effluent soluble bCOD concentration, and gas production. However, as discussed in Sec. 8–5, in Chap. 8, for aerobic activated sludge processes, computer aided mechanistic dynamic simulation models have been developed and applied for anaerobic reactors. The most common is the model developed for anaerobic sludge digestion, termed ADM1, by an International Water Association task group (Batstone et al., 2002a). ADM1 has also been applied to other wastes with high solids contents such as pig slurry wastes (Girault et al., 2011) and waste mixtures of biodegradable soluble and particulate COD (Batstone et al., 2002b; Fezzani and Cheikh, 2008).

The ADM1 model follows the fate of feed COD according to the schematic given on Fig. 7–26 and also accounts for inert nondegradable soluble COD and volatile solids. Changes in concentration of biodegradable solid and dissolved COD components of the waste and intermediate degradation products are determined by the application of biokinetic equations along the various degradation pathways that include the type of microorganisms involved and effects of pH and temperature. The particulate COD is considered to be a homogeneous mixture of carbohydrates, proteins, and lipids. Kinetic relationships described a disintegration rate of particles to a production rate of carbohydrate, protein and lipid components. Another set of equations is used to describe the rate of hydrolysis of carbohydrates, proteins and lipids to sugars, amino acids, and long chain fatty acids (LCFA), respectively. Monod-based biokinetic equations are then applied for acidogenesis of sugars and amino acids to volatile fatty acids (VFA) and hydrogen and acetogenesis of LCFA and VFAs to acetate. Separate Monod-based kinetics are used to describe methanogenesis by acetate and hydrogen utilizing organisms.

The model also includes physicochemical processes to (1) calculate the reactor pH as a function of alkalinity and VFA concentrations and gas phase carbon dioxide concentration and (2) gas-liquid transfer of carbon dioxide, methane and hydrogen sulfide generated in the process. The model is applied as a series of differential equations encompassing 32 dynamic concentration state variables. Application of the model is most useful for dynamic simulations to analyze changes in VFA and hydrogen concentrations following transient feed and loading variations to evaluate conditions that can result in an imbalance between VFA production rates and methanogenesis acetate utilization rates that could lead to digester instability (Straub et al., 2006).

10–7 CODIGESTION OF ORGANIC WASTES WITH MUNICIPAL SLUDGE

Codigestion refers to an anaerobic digestion process in which different types of wastes from at least two different sources are combined and treated in a common anaerobic reactor. The main practice of codigestion occurs in municipal anaerobic sludge digesters. If there is excess capacity in municipal facility anaerobic digesters, codigestion of other wastes in the community can be an attractive means for increasing the methane production and energy available for the facility or for other community uses such as gas powered vehicles. Typical applications of codigestion have been for processing FOG wastes and food wastes. Codigestion applications with municipal sludge digesters is discussed in Chap. 13.

Benefits of Codigestion. A major benefit of codigestion is the ability to turn a waste product into a source of energy, while at the same time reducing the region carbon footprint by the replacement of other fuels with methane from the anaerobic conversion of the waste material and curtailing the carbon dioxide release from the waste decomposition in other ways (Rosso and Stenstrom, 2008). A variety of wastes are available in local communities for processing in codigestion. Examples are given in the list of wastes shown in Table 10–18 from a modest sized community that were found to be highly amenable for codigestion (Muller et al., 2009).

Table 10–18

Example of types of highly biodegradable wastes evaluated for a municipal codigestion application[a]

Description of wastes	Comment
Flower and vegetable wastes	Requires nutrients and alkalinity
Blood product from animal processing	High in nitrogen
Dissolved air flotation sludge from rendering plant	High in nitrogen
Brown grease from grease traps	Difficult to degraded alone, requires nutrients and alkalinity
Chili, soup and salad dressing production wastes	Requires nutrients and alkalinity
Confectionary sugar wastes	Requires nutrients and alkalinity
Beer, wine, soda and juice production wastes	Requires nutrients and alkalinity

[a] Muller et al. (2009).

Codigestion can also be an attractive alternative for various food processing operations in lieu of constructing their own onsite anaerobic treatment process. The advantages include economies of scale for the facility, the elimination of onsite operational requirements, and the elimination of operating cost for the addition of alkalinity and nutrients for high carbohydrate wastes. Without a sufficient amount of protein or amino acids in the waste to produce ammonium bicarbonate at a level needed to maintain a proper pH in anaerobic treatment, the cost for alkalinity addition can be prohibitive.

Operation of Digestion Process. Alkalinity production in municipal anaerobic digesters is normally sufficient due to the degradation of organic nitrogen in primary sludge and waste activated sludge. In some cases, such as oily wastewaters (Jeganathan et al., 2006), the waste is difficult to treat by itself, but can be handled within proper proportions in municipal anaerobic digesters. The economical impact to the municipal facility must also be considered and includes costs for feed stock management, storage, and pretreatment such as screening or heating, and increased operational requirements for the codigestion operation.

PROBLEMS AND DISCUSSION TOPICS

10–1 The alkalinity concentration in an anaerobic suspended growth reactor operated at 30°C, is 2200, 2600, or 2800 mg as $CaCO_3$/L (value to be selected by instructor). Assuming equilibrium between the liquid and gas phase with a CO_2 content in the gas phase of 35 percent, determine the reactor pH.

10–2 An industrial wastewater with a flowrate of 4000 m^3/d has a soluble degradable COD concentration of 10,000, 5000, and 2500 mg/L (value to be selected by instructor), 20°C temperature, and 200 mg/L alkalinity concentration as $CaCO_3$. Determine and compare the net operating costs or revenue for anaerobic versus aerobic treatment based on the following key parameters and assumptions (labor and maintenance costs are omitted here) for each:

Anaerobic process:
Anaerobic operating cost items are related to raising the liquid temperature, and adding alkalinity, versus the revenue from methane production. The following assumptions apply:

1. Reactor temperature = 35°C
2. Heat exchanger recovery efficiency for raising liquid temperature = 80 percent
3. COD removal efficiency = 95 percent
4. CO_2 of gas phase = 35 percent and pH = 7.0

5. Value of methane $= \$5/10^6$ kJ
6. Alkalinity is provided as $NaHCO_3$ at \$0.90/kg

Aerobic

Major aerobic treatment operating cost items are energy for aeration and sludge processing and disposal. The following assumptions apply:

1. COD removal efficiency = 99 percent
2. gO_2/g COD removal = 1.2
3. Actual aeration efficiency = 1.2 kgO_2/kWh
4. Electricity costs = \$0.08/kWh
5. Net sludge production = 0.3 g TSS/g COD removed
6. Sludge processing/disposal cost = \$0.10/kg dry solids

10–3 A wastewater has a daily average flowrate of 1000, 2000, or 3000 m^3/d (value to be selected by instructor) and 4000 mg/L of an organic substance with the following approximate composition: $C_{50}H_{75}O_{20}N_5S$. For anaerobic treatment at 95 percent degradation determine (a) the alkalinity production in mg/L as $CaCO_3$; and (b) the approximate mole fraction of CO_2, CH_4, and H_2S in the gas phase.

10–4 An industrial wastewater has an average flowrate of 2000 m^3/d, an influent COD concentration of 4000, 6000, or 8000 mg/L (value to be selected by instructor), and influent sulfate concentration of 500 mg/L. The percent of COD degraded in an anaerobic treatment process at 35°C is 95 percent, and 98 percent of the sulfate is reduced. Determine (a) the amount of methane produced in m^3/d; (b) the amount of methane produced in m^3/d, if the sulfate reduction is not accounted for; and (c) the amount of H_2S in the gas phase at a reactor pH value of 7.0.

10–5 A suspended growth anaerobic reactor is operated at an SRT of 30 d at a temperature of 30°C. On a given day, the methane gas production rate (m^3/d) decreases by 30 percent. List at least four possible causes that should be investigated and briefly explain the mechanism behind each one.

10–6 A 100 percent soluble industrial wastewater is to be treated by an anaerobic contact process consisting of a mixed covered reactor, a degasifier, and gravity settling. The effluent TSS concentration from the clarifier is 120 mg/L. For the following wastewater characteristics and design assumptions, determine and compare the following design parameters for treatment at 25 and 35°C to meet an effluent soluble COD concentration of ≤ 50 mg/L.

a. The design SRT, d
b. The reactor volume, m^3
c. The reactor hydraulic detention time τ, d
d. The methane gas production rate, m^3/d
e. The total gas production rate, m^3/d
f. The amount of solids and to be manually wasted daily, kg/d
g. The nitrogen and phosphorus requirements, kg/d.

Wastewater characteristics:

Parameter	Unit	Value
Flowrate	m^3/d	2000
Degradable COD	mg/L	

(*continued*)

(*Continued*)

Parameter	Unit	Value
Wastewater 1		4000
Wastewater 2		6000
Wastewater 3		8000
Percent sCOD	%	100
Alkalinity	mg/L as $CaCO_3$	500

Note: Wastewater 1, 2 or 3 to be selected by instructor.

Other design assumptions:

1. Reactor MLSS concentration = 5000 mg/L
2. Factor of safety for SRT = 1.5
3. VSS/TSS ratio = 0.85
4. f_d = 0.15 g VSS cell debris/g VSS biomass decay
5. Gas phase methane = 65 percent
6. Nitrogen content of biomass = 0.12 g N/g VSS5
7. Phosphorus content of biomass = 0.02 g P/g VSS
8. Use the appropriate kinetic coefficients and design information provided in Table 10–13

10–7 An anaerobic process is being considered for the treatment of a soluble industrial wastewater at 30°C. A design SRT of 30 d is required to provide the desired level of 95 percent soluble COD degradation. An effluent VSS concentration of 100, 150 or 200 mg/L (value to be selected by instructor) from biomass growth is assumed. Using the appropriate kinetic coefficient values from Table 10–13, determine the influent COD concentration that must be present to allow operation at a 30-d SRT, if all the biomass wasted is via the effluent solids losses.

10–8 Design a single UASB reactor to treat an industrial wastewater at 30°C with the following wastewater characteristics and using the assumptions given below. Assume 97 percent degradation of the soluble COD, 60 percent particulate COD degradation, and an effluent VSS concentration of 200 mg/L. Using the given information, determine:

1. The reactor liquid volume, m^3
2. The reactor area (assume a circular reactor will be used), m^2
3. The reactor area diameter and total height, m
4. The hydraulic retention time, d
5. The average SRT, d
6. The amount of solids to be manually wasted daily, kg VSS/d
7. The methane gas production rate, m^3/d
8. The energy value of the gas, kJ/d
9. The alkalinity requirement, kg as $CaCO_3/d$

Design Assumptions:

1. Kinetic coefficients from Table 10–13.
2. f_d = 0.15 g VSS/g VSS biomass decayed
3. Maximum organic loading rate = 6.0 kg $COD/m^3 \cdot d$
4. Maximum upflow velocity = 0.50 m/h
5. pH = 7.0
6. CO_2 in gas phase = 35 percent
7. Process reactor liquid height = 8 m
8. Average solids concentration in process reactor = 50 g VSS/L

Wastewater characteristics:

Parameter	Unit	Value
Flowrate	m^3/d	500
Total bCOD	mg/L	
Wastewater 1		6000
Wastewater 2		7000
Wastewater 3		8000
Particulate COD	Percent	40
Particulate COD/VSS ratio	g/g	1.8
Particulate VSS/TSS ratio	g/g	0.85
Alkalinity	mg/L as $CaCO_3$	300

Note: Wastewater 1, 2, or 3 to be selected by instructor.

10–9 A domestic wastewater is to be treated using the UASB process at 25°C. The wastewater characteristics are given in the following table. Determine: (a) the reactor hydraulic retention time (hours); (b) the COD loading rate (kg COD/m^3•d); and (c) the process reactor liquid height (m) and diameter (m). What effluent BOD and TSS concentration may be expected from the UASB reactor? Describe an aerobic secondary treatment process you would select to add after the UASB process to meet an effluent BOD concentration of 20 mg/L or less. Would alkalinity have to be added to the UASB reactor to maintain the pH near 7.0? Explain the basis for your answer.

Wastewater characteristics:

Parameter	Unit	Value
Flowrate	m^3/d	
Wastewater 1		3000
Wastewater 2		4000
Wastewater 3		5000
COD	mg/L	450
BOD	mg/L	180
TSS	mg/L	180
Alkalinity	mg/L as $CaCO_3$	150

Note: Wastewater 1, 2, or 3 to be selected by instructor.

10–10 A brewery wastewater with a flowrate of 1000 m^3/d and COD (mainly soluble) of 4000 mg/L is to be treated at 35°C in a 4 m upflow attached growth anaerobic reactor, which contains cross-flow plastic packing, with the aim of 90 percent COD removal. Assume that the attached growth SRT is 30 d. Determine (a) the reactor volume (m^3) and dimensions; (b) the methane gas production rate (m^3/d); and (c) the effluent TSS concentration (mg/L).

10–11 An industrial wastewater has a degradable COD concentration of 8000 mg/L and 4000 mg/L VSS concentration with 50 percent of the VSS degradable. Briefly critique the compatibility of the following processes for treatment of this wastewater and describe the potential impact of the influent solids on the reactor operation and performance.

Processes:
- UASB
- Anaerobic fluidized bed reactor
- Anaerobic baffled reactor
- Upflow packed bed reactor
- Downflow attached growth reactor
- Anaerobic covered lagoon

10–12 From the literature within the past three years, identify and summarize an application of an anaerobic membrane process. Include a description of the wastewater treated, the reactor design, the organic loading rate, the temperature, the membrane fouling control strategy, the membrane flux rate over time, the reactor solids concentration, the membrane cleaning method or restoration method, and any significant operating and performance issues.

10–13 From a review of the literature summarize the wastewater type and characteristics, system design and operating conditions and the treatment performance of a UASB, EGSB, or anaerobic contact process (to be selected by the instructor).

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11

Separation Processes for Removal of Residual Constituents

WORKING TERMINOLOGY

Term	Definition
Absorption	The process by which atoms, ions, molecules, and other constituents are transferred from one phase and are distributed uniformly in another phase (see also adsorption).
Activated carbon	A substance used commonly in adsorption processes for the removal of trace constituents from water and odor compounds from air. Activated carbon is derived from an organic base material, prepared using a high temperature pyrolysis process and activated at high temperature in the presence of steam resulting in properties conducive to mass transfer.

Term	Definition
Adsorption	The process by which atoms, ions, molecules, and other constituents are transferred from one phase and accumulate on the surface of another phase (see also absorption).
Backwash	The process of removing solids accumulated on or in a filtration medium by applying air and/or clean water in the opposing flow direction.
Brine	Concentrated liquid waste stream containing elevated concentrations of total dissolved solids
Depth filtration	The removal of particulate matter suspended from a liquid by passing the liquid through a granular medium such as sand or anthracite coal.
Electrodialysis (ED)	A process that moves ions (charged molecular species) from one solution to another by employing an electrical potential as the driving force and using a semipermeable membrane as a separator.
Flux	The mass or volume rate of transfer through the membrane surface, usually expressed as $m^3/m^2 \cdot h$ or $L/m^2 \cdot h$ ($gal/ft^2 \cdot d$).
Fouling	The accumulation of solid matter on the surface of or within the pores of a membrane that impedes the flow of permeate through the membrane.
Gas stripping	A process to remove a volatile constituent from a liquid phase, such as in the removal of ammonia from water in a packed column using air as the gas phase.
Ion exchange	A process used for the removal of dissolved ionic constituents where ions of a given species are displaced from a solid phase material by ions of a different species from solution.
Isotherm	A function used to relate the amount of a given constituent adsorbed from water per concentration of adsorbent at a given temperature.
Membrane	A device, usually made of an organic polymer, that allows the passage of water and certain constituents, but rejects others above a certain physical size or molecular weight.
Microfiltration (MF)	A membrane separation process used typically to remove particulate material from the feed water; microfiltration pore sizes range approximately from 0.05 to 2 μm.
Nanofiltration (NF)	A pressure-driven membrane separation process used to remove colloidal and dissolved material as small as approximately 0.001 μm.
Residuals	Waste streams produced by wastewater treatment processes. For depth and surface filtration, the residual waste stream is filter waste washwater. For membrane systems, residual waste streams include waste washwater, concentrate, and chemical cleaning wastes.
Reverse osmosis (RO)	The rejection of dissolved constituents by preferential diffusion using a pressure-driven, semipermeable membrane.
Semipermeable membrane	A membrane that is permeable to some components in a feed solution and impermeable to other components.
Separation processes	Physical and chemical processes used in water reclamation that bring about treatment by the isolation of particular constituents. The isolated constituents are concentrated into a waste stream that must be managed.
Sodium adsorption ratio (SAR)	A measure of the sodicity of the soil; the SAR is the ratio of the sodium cation to the calcium and magnesium cations.
Surface filtration	The removal of particulate matter suspended in a liquid by passing the liquid through a thin septum, usually a cloth or metal medium.
Synthetic organic compounds (SOCs)	Compounds of synthetic origin used extensively in industrial processes and contained in numerous manufactured consumer products. The presence of SOCs in drinking water as well as reclaimed water is of concern due to toxicity and unknown effects.
Ultrafiltration (UF)	A membrane separation process similar to MF except the membrane pore sizes can range from approximately 0.005 to 0.1 μm. Generally, UF membranes are able to achieve higher levels of separation than MF, particularly for bacteria and viruses.

The effluent from conventional secondary treatment contains varying amounts of residual suspended, colloidal, and dissolved constituents. Suspended and colloidal matter can reduce the effectiveness of downstream disinfection processes or make the effluent unsuitable for discharge or reuse. Dissolved constituents may range from relatively simple inorganic ions, such as calcium, potassium, sulfate, nitrate, and phosphate to an ever-increasing number of highly complex synthetic organic compounds. Research is ongoing to determine (1) the environmental effects of potentially toxic and biologically active substances found in wastewater and (2) how these substances can be removed by both conventional and advanced wastewater treatment processes. In recent years, the effects of many of these substances on the environment have become understood more clearly. As a result, wastewater treatment requirements are becoming more stringent in terms of limiting effluent concentrations of many of these substances.

To meet new treatment requirements, many of the existing secondary treatment facilities will have to be retrofit and new advanced wastewater treatment facilities will have to be constructed. The purpose of this chapter is to present an introduction to the unit processes used for the removal and/or treatment of residual particulate, colloidal, and dissolved constituents in treated wastewater. However, before discussing the individual unit processes, it will be helpful to review the need for additional wastewater treatment and the reasons that specific constituents are of concern.

11–1 NEED FOR ADDITIONAL WASTEWATER TREATMENT

Residual constituents found in secondary effluent can be grouped into four broad categories: (1) organic and inorganic suspended and colloidal particulate matter, (2) dissolved organic constituents, (3) dissolved inorganic constituents, and (4) biological constituents. Constituents within each category are reported in Table 11–1, along with the reasons for their removal. The potential impacts of the residual constituents identified in Table 11–1 will vary considerably depending on local conditions. The list of constituents presented in Table 11–1 is not meant to be exhaustive; rather, it is meant to highlight that a wide variety of substances must be considered in establishing and meeting discharge requirements. Also, based on the accumulation of scientific knowledge concerning the impacts of the residual constituents found in secondary effluent, derived from laboratory studies and environmental monitoring, it is anticipated that many of the treatment methods now classified as tertiary or advanced will be considered conventional within the next 10 to 20 years. For example, effluent filtration has become more commonplace within the past 20 years.

11–2 OVERVIEW OF TECHNOLOGIES USED FOR REMOVAL OF RESIDUAL PARTICULATE AND DISSOLVED CONSTITUENTS

Over the past 20 years, a wide variety of treatment technologies have been studied, developed, and applied for the removal of the residual constituents found in secondary and tertiary effluent. The unit processes used for the removal of residual constituents from water may be classified as (1) mass transfer separation processes and (2) chemical and biological transformation processes.

Separation Processes Based on Mass Transfer

The removal of constituents by the transfer of mass from one phase to another or by the concentration of mass within a phase is accomplished with various unit processes.

Table 11–1

Typical residual constituents found in treated wastewater effluents and reasons that additional treatment may be required

Residual constituent	Effect and/or need for additional treatment
Inorganic and organic suspended and colloidal particulate matter	
Suspended solids	• Can impact disinfection by shielding organisms • May cause sludge deposits or interfere with receiving water clarity • May affect effluent turbidity
Colloidal solids	• May affect effluent turbidity
Organic matter (particulate)	• May shield bacteria during disinfection, may deplete oxygen resources
Dissolved organic matter	
Total organic carbon	• May deplete oxygen resources
Refractory organics	• Toxic to humans; carcinogenic
Volatile organic compounds	• Toxic to humans; carcinogenic; form photochemical oxidants
Pharmaceutical compounds	• Impacts to aquatic species (e.g., endocrine disruption)
Surfactants	• Cause foaming and may interfere with coagulation
Dissolved inorganic matter	
Ammonia	• Increases chlorine demand • Can be converted to nitrates and, in the process, can deplete oxygen resources • With phosphorus, can lead to the development of undesirable aquatic growth • Toxic to fish
Nitrate	• Can stimulate algal and other aquatic growth • Can cause methemoglobinemia in infants (blue babies)
Phosphorus	• Can stimulate algal and other aquatic growth • Increases chemical requirements • Interferes with lime-soda softening
Calcium and magnesium	• Increase hardness and total dissolved solids • Can affect sodium adsorption ratio
Chloride and sulfate	• Can impart salty taste
Total dissolved solids	• Interfere with agricultural and industrial processes • Can interferes with coagulation
Biological	
Bacteria	• Can cause disease
Protozoan cysts and oocysts	• Can cause disease
Viruses	• Can cause disease

The principal mass transfer processes used for the separation (removal) of residual constituents are summarized in Table 11–2. It is important to note that a key characteristic of most separation processes is the generation of a waste stream that will require subsequent management (e.g., processing, disposal, reuse). The particular waste stream generated will depend on the type and effectiveness of the separation process used. For example,

Table 11–2

Unit processes based on mass transfer used for the removal of particulate and dissolved constituents in wastewater treatment and water reclamation[a]

Unit process	Phase	Application
Absorption	Gas → liquid	Aeration, O_2 transfer, SO_2 scrubbing, chlorination, chlorine dioxide and ammonia addition, ozonation
Adsorption	Gas → solid Liquid → solid	Removal of inorganic and organic compounds using activated carbon, activated alumina, granular ferric hydroxide, or other adsorbent material
Distillation	Liquid → gas	Demineralization of water, concentrating of waste brines
Electrodialysis	Liquid → liquid	Removal of dissolved species, removal of salts
Filtration, depth	Liquid → solid	Removal of particulate material
Filtration, surface	Liquid → solid	Removal of particulate material
Flotation	Liquid → solid	Removal of particulate constituents
Gas stripping	Liquid → gas	Removal of NH_3 and other volatile inorganic and organic chemicals
Ion exchange	Liquid → solid	Demineralization of water, removal of specific constituents, softening
Microfiltration ultrafiltration	Liquid → liquid	Removal of particulate and colloidal species
Nanofiltration	Liquid → liquid	Removal of dissolved and colloidal species; softening
Precipitation, chemical	Liquid → solid	Removal of particulate and dissolved species; softening
Reverse osmosis	Liquid → solid	Removal of dissolved constituents
Sedimentation	Liquid → solid	Removal of particulate constituents

[a] Adapted in part from Crittenden et al. (2012).

adsorption results in a medium saturated with removed constituents, chemical precipitation produces a sludge containing both the precipitated constituents as well as the chemical(s) added to cause the precipitation, and reverse osmosis produces a liquid waste (brine) containing concentrated rejected constituents. In many cases, the management of waste streams resulting from separation processes, as discussed in Chap. 15, can present a significant technological challenge and cost.

Transformation Based on Chemical and Biological Processes

The second group of processes used for the removal of residual constituents make use of chemical and biological reactions to transform or destroy trace constituents in water, typically through oxidation and reduction reactions. Conventional chemical oxidants that have been used for constituent transformation include hydrogen peroxide, ozone, chlorine, chlorine dioxide, and potassium permanganate. Chemical oxidation processes that utilize hydroxyl radical species, referred to as advanced oxidation processes (AOPs), or photons generated through UV photolysis, are particularly effective for the transformation and destruction of trace constituents, often resulting in the complete mineralization of trace constituents to carbon dioxide and mineral acids. Chemical treatment processes including advanced oxidation and photolysis are considered in Chap. 6. Chemical disinfection of wastewater is considered separately in Chap. 12. Biological treatment and conversion processes are considered in Chaps. 7 through 10.

Application of Unit Processes for Removal of Residual Constituents

Information on the application of the unit processes identified in Table 11–2 is presented in Table 11–3. Selection of a given unit process or combination thereof depends on (1) the use to be made of the treated effluent; (2) the constituent(s) of concern; (3) the compatibility of the various operations and processes; (4) the available means for management of any process residuals; and (5) the environmental and economic feasibility of the various systems. Specific factors that should be considered in the selection of treatment processes were identified and discussed previously in Table 4–2 in Chap. 4. It should be noted that in some situations, economic feasibility may not be a controlling factor in the design of advanced wastewater treatment systems, especially where specific constituents must be removed to protect the environment and/or to meet discharge requirements. Because of the variations in performance observed in the field, bench-scale and pilot-plant testing is recommended for the development of local treatment performance data and design criteria. Representative performance data for the processes identified in Table 11–3 are presented in the discussion of the individual technologies that follows and in the indicated sections in other chapters.

11–3 UNIT PROCESSES FOR THE REMOVAL OF RESIDUAL PARTICULATE AND DISSOLVED CONSTITUENTS

The principal unit processes used for the removal of residual particulate matter discussed in this chapter include (1) depth filtration (passing the liquid through a filter bed comprised of a granular or compressible filter medium); (2) surface filtration (the removal of particulate material suspended in a liquid by mechanical sieving by passing the liquid through a thin septum); and (3) membrane filtration (passing the liquid through porous material to exclude particles ranging in size from 0.005 to 2.0 μm). Each of these processes is illustrated and described in Table 11–4. Flotation (attaching air bubbles to particulate matter to provide buoyancy so the particles can be removed by skimming) is included in Table 11–4 for completeness, but has been considered previously in Chap. 5.

The principal unit processes used for the removal of dissolved constituents, as discussed in this chapter, include (1) reverse osmosis (passing the liquid through semipermeable membranes to exclude particles ranging in size from 0.0001 to 0.001 μm), (2) electrodialysis (transport of ionic species through an ion-selective membranes), (3) adsorption (the accumulation of constituents on a solid phase), (4) gas stripping (transfer of a constituent from a liquid to a gas phase), (5) ion exchange (the exchange of ionic species), and (6) distillation (constituents are separated by evaporation).

Table 11–3

Application of the unit processes for the removal of residual particulate and dissolved constituents found in treated wastewater effluents[a]

Residual constituent	Unit process (Section discussed)					
	Depth filtration (11–4)	Surface filtration (11–5)	Micro and ultra-filtration (11–6)	Reverse osmosis (11–6)	Electro-dialysis (11–7)	Adsorption (11–8)
Inorganic and organic suspended and colloidal particulate matter						
Suspended solids	✔	✔	✔	✔		✔
Colloidal solids	✔	✔	✔	✔		✔
Dissolved organic matter						
Total organic carbon				✔	✔	✔
Refractory organics				✔	✔	✔
Volatile organic compounds				✔	✔	✔
Dissolved inorganic matter						
Ammonia[a]				✔	✔	
Nitrate[a]				✔	✔	
Phosphorus[a]	✔[b]			✔	✔	
Totals dissolved solids				✔	✔	
Biological						
Bacteria			✔	✔	✔	
Protozoan cysts and oocysts	✔		✔	✔	✔	✔
Viruses				✔	✔	

[a] The biological removal of nitrogen and phosphorus is considered in Chaps. 7 through 10.

[b] Phosphorous removal is accomplished in a two-stage filtration process.

[c] Some carryover can occur.

Typical Process Flow Diagrams

Typical treatment process flow diagrams that incorporate the unit processes discussed above are illustrated on Fig. 11–1. The combination of unit processes will depend on the treatment objective. For example, in the flow diagram shown on Fig. 11–1(b), electrodialysis is used to remove salts to reduce the total dissolved solids in the treated effluent. On Fig. 11(d), a number of unit processes have been combined to produce a potable water. Where reverse osmosis is used, some form of membrane filter will be used upstream to mitigate the effects of particulate matter that tends to clog the membrane. On Fig. 11–1(f), two stages of reverse osmosis have been combined to produce water suitable for use in

Table 11–3 (*Continued*)

Unit process (Section discussed)						
Gas stripping (11–9)	Ion exchange (11–10)	Distillation (11–11)	Chemical precipitation (6-3, 4, 5)	Chemical oxidation (6–7)	Advanced oxidation processes (6–8)	Photolysis (6–9)
	✔	✔	✔			
	✔	✔	✔			
	✔	✔	✔	✔	✔	✔
		✔		✔	✔	✔
✔		✔[c]		✔	✔	✔
✔	✔	✔				
	✔	✔				
		✔	✔			
	✔	✔				
		✔				
	✔	✔	✔			
		✔		✔	✔	✔

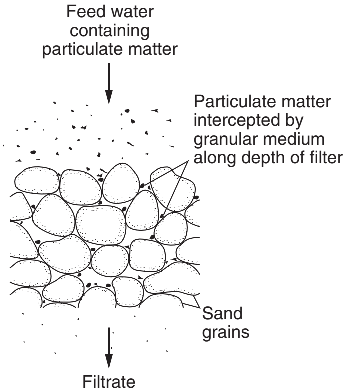
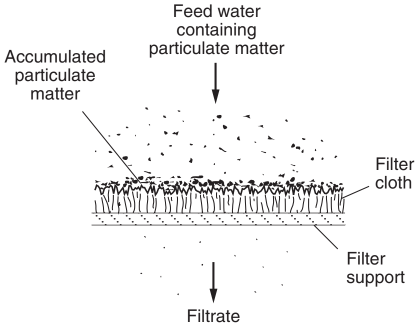
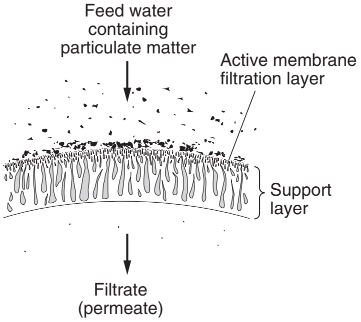
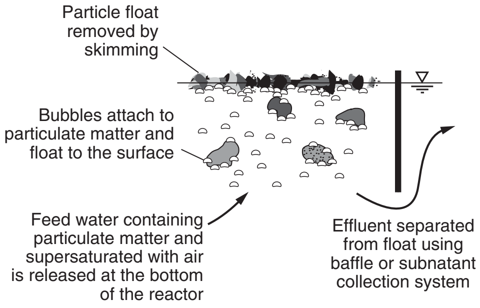
high-pressure boilers. Clearly, a wide variety of treatment process flow diagrams can be developed, depending on the specific requirements. Other examples of process flow diagrams are presented and discussed throughout this chapter.

Process Performance Expectations

It is important to know what typical mean effluent constituent values can be expected and the variability in those values for a given unit process. Information on the constituent values and variability is of importance in meeting effluent requirements and in the selection of technologies that might be used to further process the treated effluent. Typical mean

Table 11–4

Description of commonly used processes for the removal of residual suspended and colloidal solids

Unit process	Description
(a) Depth filtration  	Depth filtration was developed originally for the treatment of surface water for potable uses and later adapted for wastewater treatment applications. Depth filtration is used in to achieve supplemental removal of suspended solids (including particulate BOD) from wastewater effluents for the following purposes: (1) to allow more effective disinfection; (2) as a pretreatment step for subsequent treatment steps such as carbon adsorption, membrane filtration, or advanced oxidation; and (3) to remove chemically precipitated phosphorus.
(b) Surface filtration  	Surface filtration is used to remove the residual suspended solids from secondary effluents and stabilization pond effluents, and as an alternative to depth filtration. Surface filtration, a relatively new technology, involves a sieving action similar to a kitchen colander.
(c) Membrane filtration  	Membrane filtration with microfiltration (MF) and ultrafiltration (UF) membranes is being used increasingly for water and wastewater applications. Microfiltration and UF membrane filters are also surface filtration devices but are differentiated on the basis of the sizes of the pores in the filter medium; the pore size can vary from 0.005 to 2.0 μm. In water reuse applications, MF and UF usually follow biological treatment and are used to remove particulates, including pathogens; organic matter; and some nutrients, not removed by secondary clarification. Product water from MF and UF may be used directly for a variety of reuse applications (after disinfection) or used as pretreated feed water for further treatment by nanofiltration (NF) or reverse osmosis (RO).
(d) Dissolved air flotation  	Dissolved air flotation is a gravity separation process in which gas bubbles attach to solid particles to cause the density of the bubble-solid agglomerates to be lighter than water. For water reuse applications, DAF has been used principally for treating pond effluents containing algae and for low density particles that are difficult to remove by gravity sedimentation, as a replacement for conventional primary sedimentation, and as a pretreatment step for depth or surface filtration. Dissolved air flotation is considered in Sec. 5–7 in Chap. 5.

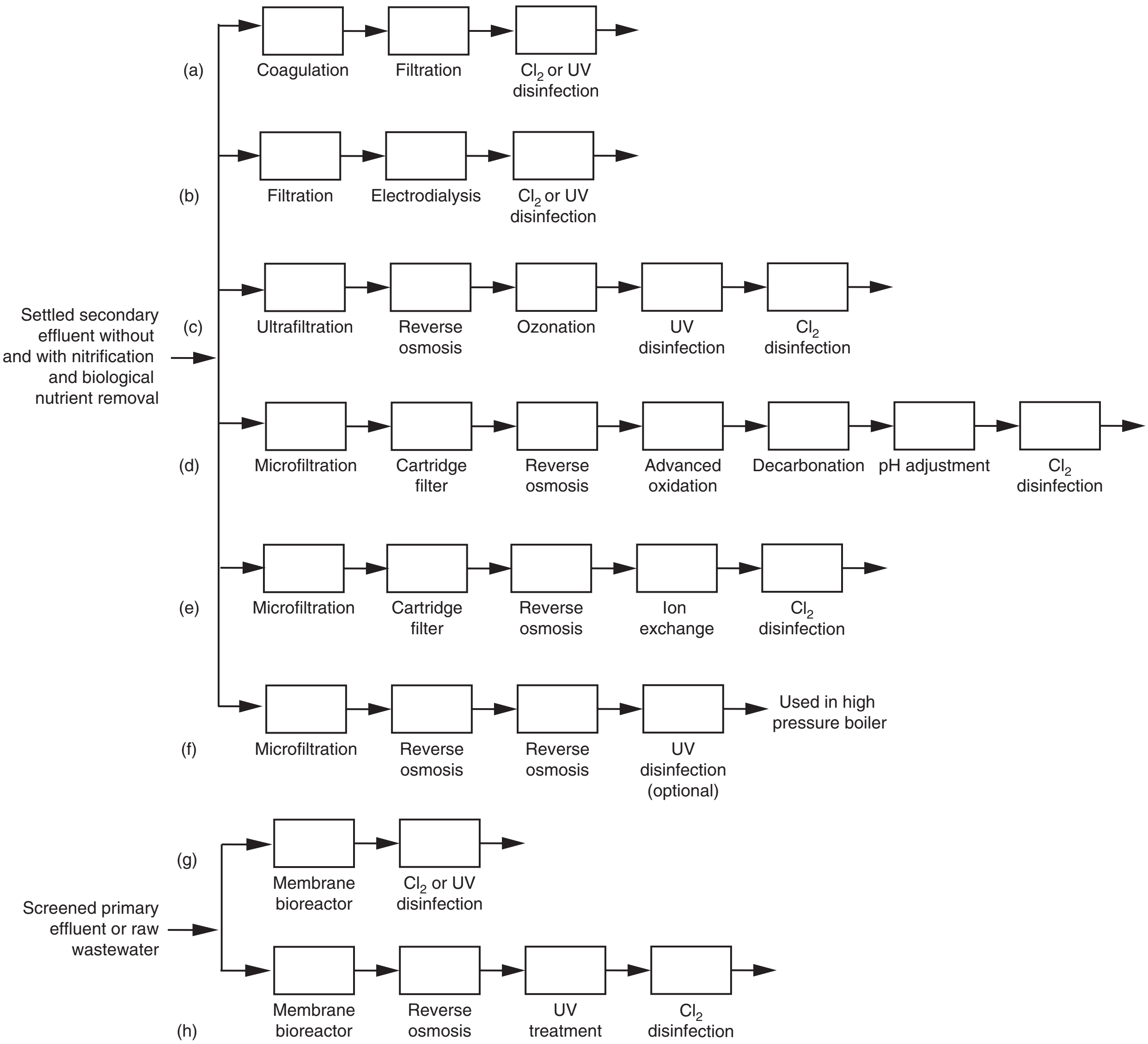


Figure 11–1

Typical process flow diagrams for wastewater treatment employing advanced treatment processes. All of the flow diagrams have been used at one time or another. For example, in flow diagram number (d) advanced oxidation is used for the destruction of NDMA. In flow diagram number (e) ion exchange is used for the removal of nitrate.

effluent constituent values that can be achieved with depth, surface, and membrane filtration following various forms of biological treatment are reported in Table 11–5. The variability observed in the performance of various particulate removal processes with respect to TSS and turbidity in the treated effluent is discussed in the sections which deal with depth, surface, and membrane filtration.

Table 11–5

Typical range of effluent quality after various levels of treatment

			Range of effluent quality after indicated treatment						
Constituent	**Unit**	**Untreated wastewater[a]**	**Conventional activated sludge[b]**	**Conventional activated sludge with filtration[b]**	**Activated sludge with BNR[c]**	**Activated sludge with BNR and filtration[c]**	**Mem-brane bioreactor**	**Activated sludge with microfiltration and reverse osmosis**
Total suspended solids (TSS)	mg/L	130–389	5–25	2–8	5–20	1–4	<1–5	≤1
Colloidal solids	mg/L		5–25	5–20	5–10	1–5	0.5–4	≤1
Biochemical oxygen demand (BOD)	mg/L	133–400	5–25	<5–20	5–15	1–5	<1–5	≤1
Chemical oxygen demand (COD)	mg/L	339–1016	40–80	30–70	20–40	20–30	<10–30	≤2–10
Total organic carbon (TOC)	mg/L	109–328	20–40	15–30	10–20	1–5	<0.5–5	0.1–1
Ammonia nitrogen	mg N/L	14–41	1–10	1–6	1–3	1–2	<1–5	≤0.1
Nitrate nitrogen	mg N/L	0–trace	5–30	5–30	<2–8	1–8	<8[d]	≤1
Nitrite nitrogen	mg N/L	0–trace	0–trace	0–trace	0–trace	0.001–0.1	0–trace	≤0.001
Total nitrogen	mg N/L	23–69	15–35	15–35	3–8	2–5	<10[d]	≤1
Total phosphorus	mg P/L	3.7–11	3–10	3–8	1–2	≤2	<0.3[e]–5	≤0.5
Turbidity	NTU		2–15	0.5–4	2–8	0.3–2	0.1–1	0.01–1
Volatile organic compounds (VOCs)	μg/L	<100–>400	10–40	10–40	10–20	10–20	10–20	≤1
Metals	mg/L	1–2.5	1–1.5	1–1.4	1–1.5	1–1.5	trace	trace
Surfactants	mg/L	4–10	0.5–2	0.5–1.5	0.1–1	0.1–1	0.1–0.5	≤1
Totals dissolved solids (TDS)	mg/L	374–1121	500–700	500–700	500–700	500–700	500–700	≤5–40
Trace constituents[f]	μg/L	10–50	5 to 40	5–30	5–30	5–30	0.5–20	~0.1
Total coliform	No./100 mL	10^6–10^{10}	10^4–10^5	10^3–10^5	10^4–10^5	10^4–10^5	<100	~0
Protozoan cysts and oocysts	No./100 mL	10^1–10^5	10^1–10^2	0–10	0–10	0–1	0–1	~0
Viruses	PFU/100 mL[g]	10^1–10^4	10^1–10^3	10^1–10^3	10^1–10^3	10^1–10^3	10^0–10^3	~0

[a] From Table 3–18 in Chap. 3.

[b] Conventional activated sludge treatment includes nitrification.

[c] BNR is defined as biological nutrient removal for the removal of nitrogen and phosphorus.

[d] With anoxic stage.

[e] With coagulant addition.

[f] For example, fire retardants, personal care products, and prescription and non-prescription drugs (see also Table 2–16 in Chap. 2).

[g] PFU = plaque forming units.

11–4 INTRODUCTION TO DEPTH FILTRATION

Depth filtration with a non-compressible filter medium or media is one of the oldest unit process used in the treatment of potable water and is commonly used for the filtration of effluents from wastewater treatment processes, especially in low-level nutrient removal and water reuse applications. Depth filtration is used most commonly to (1) achieve supplemental removals of residual suspended solids (including particulate BOD and phosphorus), (2) reduce the mass discharge of solids, and (3) perhaps more importantly, as a conditioning step that will allow for the effective disinfection of the filtered effluent, especially with UV disinfection (see Chap. 12). Single- and two-stage filtration is used to remove chemically precipitated phosphorus. The relationship between depth filtration and other forms of filtration is illustrated on Fig. 11–2. In the past, depth filtration was used almost exclusively for effluent filtration. However, with the development of modern surface filtration technologies, as discussed in Sec. 11–5, depth filtration is no longer the dominant filtration technology.

To introduce the subject of depth filtration, the purpose of this section is to present (1) a general introduction to the depth filtration process, (2) an introduction to filter clean-water hydraulics, and (3) an analysis of the filtration process. The types of filters that are available and issues associated with their selection and design, including a discussion of the need for pilot-plant studies, are considered in the following section.

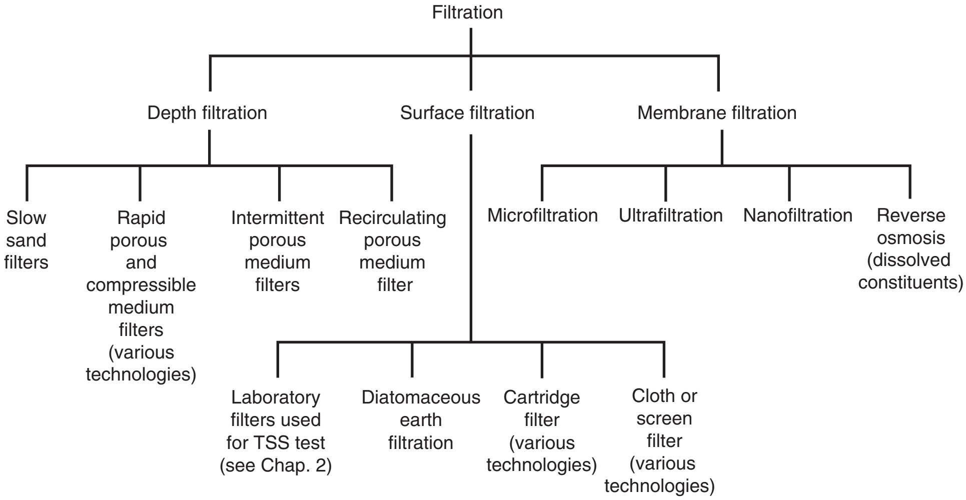
Description of the Filtration Process

The basics of the depth filtration can be understood by considering the (1) physical features of a conventional granular medium depth filter, (2) characteristics of the filter-medium, (3) the process by which suspended material is removed from the liquid, and (4) the backwash process in which material retained within the filter is removed.

Physical Features of a Depth Filter. The general features of a conventional granular medium depth filter are illustrated on Fig. 11–3. As shown, the filtering medium (sand in this case) is supported on a gravel layer, which, in turn, rests on the filter underdrain system. The water to be filtered enters the filter from an inlet channel. Filtered water

Figure 11–2

Classification of filtration processes used in wastewater management. Note: Intermittent and recirculating porous medium filters are used for small systems and are not considered in this text.



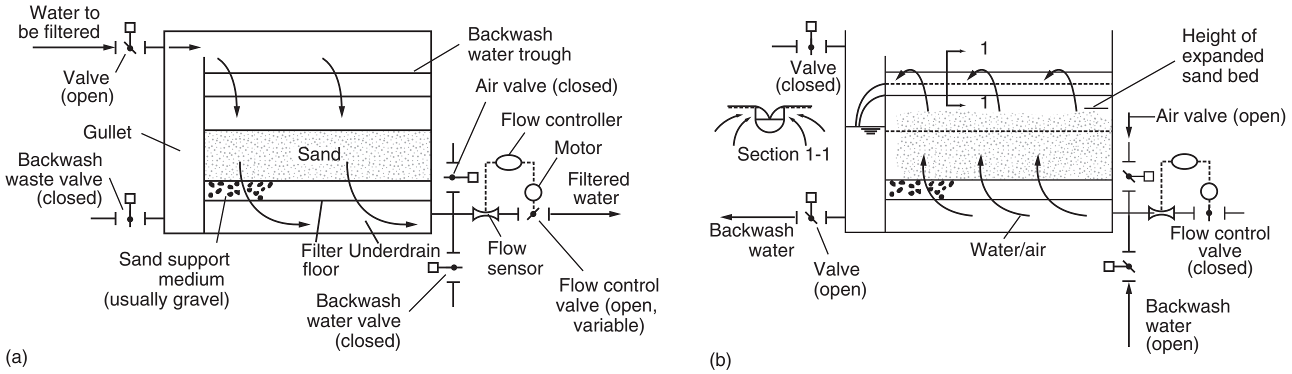


Figure 11–3

General features and operation of a conventional rapid granular medium depth filter: (a) flow during filtration cycle, and (b) flow during backwash cycle. (From Tchobanoglous and Schroeder, 1985.)

is collected in the underdrain system which is also used to reverse the flow to backwash the filter. Filtered water typically is disinfected before being discharged to the environment. If the filtered water is to be reused, it can be discharged to a storage reservoir or to the reclaimed water distribution system.

Characteristics of the Filter Medium. Grain size is the principal filter-medium characteristic that affects the filtration operation including the removal of suspended and colloidal material, the clear-water headloss, and the buildup of headloss during the filter run. If the size of the filtering medium is too small, much of the driving force will be wasted in overcoming the frictional resistance of the filter bed. If the size of the medium is too large, many of the small particles in the influent will pass directly through the filter bed. Thus, the selection of media size must balance the need for a target filtered water quality with an acceptable rate of filter headloss development. The size distribution of the filter material is usually determined by sieve analysis using a series of decreasing sieve sizes. The designation and size of opening for U.S. sieve sizes are given in Table 11–6. The results of a sieve analysis are usually analyzed by plotting the cumulative percent passing a given sieve size on arithmetic-log or probability-log paper (see Example 11–1).

The effective size of a filtering medium, $d_{10,}$ is defined as the 10 percent size based on weight. For sand, it has been found that the 10 percent size by weight corresponds approximately to the 50 percent size by count. The uniformity coefficient (UC) is defined as the ratio of the 60 percent size to the 10 percent size ($\mathrm{UC} = d_{60}/d_{10}$). Sometimes it is advantageous to specify the 99 percent passing size and the 1 percent passing size to define the gradation curve for each filter medium more accurately. Additional information on filter medium characteristics is presented in the following section dealing with the design of depth filters.

The Filtration Process. During filtration in a conventional downflow depth filter, wastewater containing suspended and colloidal material is applied to the top of the filter bed [see Fig. 11–3(a)]. As the water passes through the filter bed, the suspended matter (measured as turbidity) in the wastewater is removed by a variety of removal mechanisms,

Table 11–6

Designation and size of opening of US sieve sizes[a]

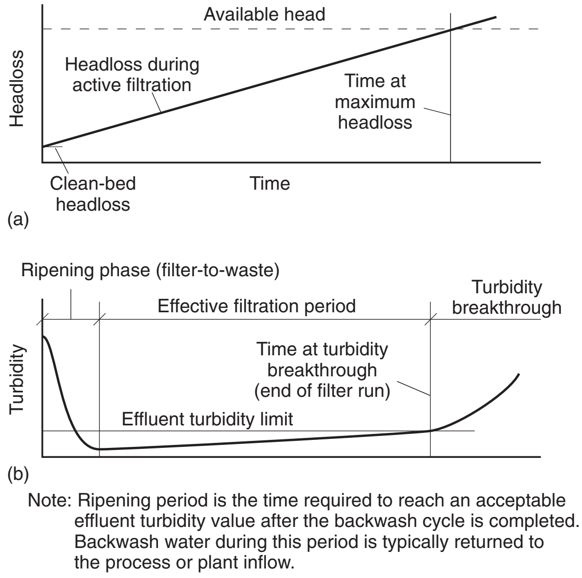
Sieve size or number	Size of opening in.	Size of opening mm
3/8 in.	0.375[a]	9.51[b]
1/4 in.	0.250[a]	6.35[b]
4	0.187	4.76
6	0.132	3.36
8	0.0937	2.38
10	0.0787[a]	2.00[b]
12	0.0661	1.68
14	0.0555[a]	1.41[b]
16	0.0469	1.19
18	0.0394[a]	1.00[b]
20	0.0331	0.841
25	0.0280[a]	0.710[b]
30	0.0234	0.595
35	0.0197[a]	0.500[b]
40	0.0165	0.420
45	0.0138[a]	0.350[b]
50	0.0117	0.297
60	0.0098[a]	0.250[b]
70	0.0083	0.210
80	0.0070[a]	0.177[b]
100	0.0059	0.149
140	0.0041	0.105
200	0.0029	0.074

[a] Adapted from ASTM (2001b).

[b] Size does not follow the ratio $(2)^{0.5}$.

as described below. With the passage of time, as material accumulates within the interstices of the granular medium, the headloss through the filter starts to build up beyond the initial value, as shown on Fig. 11–4.

Headloss and Turbidity Considerations. After some period of time, the operating headloss or effluent turbidity reaches some predetermined headloss or turbidity value, and the filter must be cleaned. Under ideal conditions, the time required for the headloss buildup to reach the preselected terminal value should correspond to the time when the turbidity or suspended solids in the effluent reach the preselected terminal value for acceptable quality. Turbidity breakthrough occurs when the interstitial spaces within the filter bed fill to a point where the shearing force of the liquid passing through the filter exceeds the strength of the bond formed between the material being filtered and the accumulated material. At breakthrough, accumulated material will dislodge only to be replaced by new material so



Note: Ripening period is the time required to reach an acceptable effluent turbidity value after the backwash cycle is completed. Backwash water during this period is typically returned to the process or plant inflow.

Figure 11–4

Definition sketch for length of filter run based on: (a) headloss buildup and (b) effluent turbidity breakthrough.

that an equilibrium condition is maintained. In actual practice, one or the other event will govern the backwash cycle.

Particle Removal Mechanisms. The principal particle-removal mechanisms, believed to contribute to the removal of material within a granular medium filter, are identified and described in Table 11–7. The major removal mechanisms (the first five listed in Table 11–7) are illustrated pictorially on Fig. 11–5. Straining has been identified as the principal mechanism that is operative in the removal of suspended solids during the filtration of settled secondary effluent from biological treatment processes (Tchobanoglous and Eliassen, 1970; Tchobanoglous, 1988).

Other mechanisms including interception, impaction, and adhesion are also operative even though their effects are small and, for the most part, masked by the straining action. The removal of the smaller particles found in wastewater (see Fig. 11–5) must be accomplished in two steps involving (1) the transport of the particles to or near the surface where they will be removed and (2) the removal of particles by one or more of the operative removal mechanisms. This two-step process has been identified as transport and attachment (O'Melia and Stumm, 1967).

Conventional down-flow filters, dual- and multi-media and deep-bed mono-medium depth filters (see Fig. 11–6) were developed to allow the suspended solids in the liquid to be filtered to penetrate further into the filter bed, and thus use more of the solids-storage capacity available within the filter bed. The deeper penetration of the solids into the filter bed also permits longer filter runs because the buildup of headloss is reduced. By comparison, in shallow mono-medium beds, most of the removal occurs in the upper few millimeters of the bed.

Backwash Process. The end of the filter run (filtration phase) is reached when the suspended solids in the effluent start to increase (breakthrough) beyond an acceptable level, or when a limiting headloss occurs across the filter bed (see Fig. 11–4). Once either of these conditions is reached, the filtration phase is terminated, and the filter must be cleaned (backwashed) to remove the material (suspended solids) that has accumulated within the granular medium filter bed. Backwashing is accomplished by reversing the flow

Table 11–7

Principal mechanisms and phenomena contributing to removal of material within a granular medium depth filter

Mechanism/phenomenon	Description
1. Straining	
a. Mechanical	Particles larger than the pore space of the filtering medium are strained out mechanically.
b. Chance contact	Particles smaller than the pore space are trapped within the filter by chance contact.
2. Sedimentation or impaction	Heavy particles that do not follow the flow streamlines settle on the filtering medium within the filter.
3. Interception	Many particles that move along in the streamline are removed when they come in contact with the surface of the filtering medium.
4. Adhesion	Particles become attached to the surface of the filtering medium as they pass by. Because of the force of the flowing water, some material is sheared away before it becomes firmly attached and is pushed deeper into the filter bed. As the bed becomes clogged, the surface shear force increases to a point at which no additional material can be removed. Some material may break through the bottom of the filter, causing the sudden appearance of turbidity in the effluent.
5. Flocculation	Flocculation can occur within the interstices of the filter medium. The larger particles formed by the velocity gradients within the filter are then removed by one or more of the above removal mechanisms.
6. Chemical adsorption	Once a particle has been brought in contact with the surface of the filtering medium or with other particles, either one of these mechanisms, or both, may be responsible for holding it there.
a. Bonding	
b. Chemical interaction	
7. Physical adsorption	
a. Electrostatic forces	
b. Electrokinetic forces	
c. van der Waals forces	
8. Biological growth	Biological growth within the filter will reduce the pore volume and may enhance the removal of particles with removal mechanisms 1 through 5.

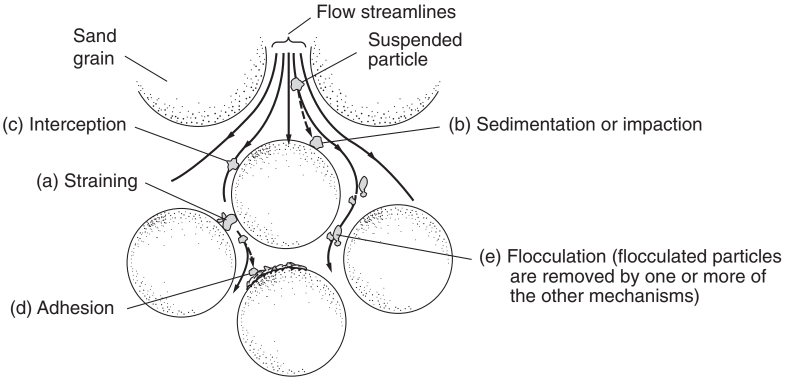
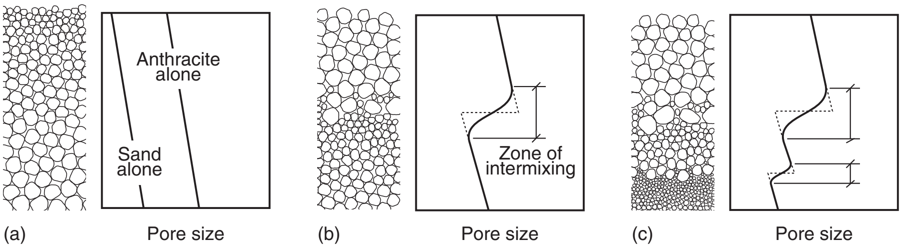


Figure 11–5

Removal of suspended particulate matter within a granular filter by: (a) straining, (b) sedimentation or inertial impaction, (c) interception, (d) adhesion, and (e) flocculation with subsequent removal by one or more of the previous mechanisms. (Adapted from Tchobanoglous and Schroeder, 1985.)

Figure 11–6

Schematic diagram of filter beds illustrating potential increase in storage capacity: (a) single medium, (b) dual media, and (c) multi-media.



through the filter [see Fig. 11–3(b)]. A sufficient flow of washwater is applied until the granular filtering medium is fluidized (expanded), causing the particles of the filtering medium to abrade against each other. The backwash water flow requirements are related to water temperature and the desired bed expansion during high-rate washing. Unless the filter bed is cleaned properly, fine material, grease, and bacterial slimes can accumulate within the bed, ultimately leading to the formation of mudballs.

The suspended matter retained within the filter is removed by the shear forces created by backwash water as it moves up through the expanded bed and by abrasion as the grains of the filter medium interact with each other. The material that has accumulated within the bed is then washed away. Surface washing with water and air scouring are often used in conjunction with the water backwash to enhance the cleaning of the filter bed. Air scour in particular will often reduce the required backwash duration and, therefore, will reduce the amount of washwater required. Following backwashing, the practice of filter-to-waste is often used to prevent any residual backwash particles from entering the filtered water. In most wastewater treatment plant flow diagrams, the washwater containing the suspended solids that are removed from the filter is returned either to the primary settling facilities or to the biological treatment process. Increasingly, especially at larger treatment plants, separate treatment facilities are provided to remove the solids from the backwash water.

Filter Hydraulics

During the past 60 years considerable effort has been devoted to the modeling of the filtration process. The models fall into two general categories: those models used to predict the clean-water headloss through a granular medium filter bed (*clean water* referring to water absent of suspended particles that would otherwise create headloss) and the filter backwash expansion, and those models used to predict the performance of filters for the removal of suspended solids. Headloss and backwash hydraulics are considered in the following discussion.

Clean-Water Headloss. Over the years a number of equations have been proposed to describe the flow of clean-water through a porous medium (Darcy, 1856; Hazen, 1905; Fair and Hatch, 1933; Kozeny-Carman, 1937; Rose, 1945; Ergun, 1952). The equations developed by these early researchers are summarized in Table 11–8. In most cases, the equations for the flow of clean water through a porous medium are derived from a consideration of the Darcy-Weisbach equation [Eq. (5–78)] for flow in a closed conduit and from dimensional analysis.

Application of Headloss Equations. The equations given in Table 11–8 apply to different flow regimes. All of the equations presented in Table 11–8 can be used for laminar flow conditions, with Reynolds numbers typically below 6. However, only the Rose and Ergun equations apply to the laminar, transitional, and turbulent flow regimes. The equations for

Table 11–8

Formulas used to compute the clear-water headloss through a granular porous medium

Equation	No.
Hazen (Hazen, 1905)	
$h = \frac{1}{C}\left(\frac{60}{T + 10}\right)\frac{L}{d_{10}^2}\upsilon_h$	(11–1)
Fair-Hatch (Fair and Hatch, 1933)	
$h = k\nu S^2\frac{(1 - \alpha)^2}{\alpha^3}\frac{L}{d^2}\frac{\upsilon_s}{g}$	(11–2)
$h = k\nu\frac{(1 - \alpha)^2}{\alpha^3}\frac{L\upsilon_s}{g}\left(\frac{6}{\psi}\right)^2\Sigma\frac{p}{d_g^2}$	(11–3)
Kozeny-Carman[a] (Carman, 1937)	
$h = \frac{k}{g}\frac{\mu}{\rho}\frac{(1 - \alpha)^2}{\alpha^3}(S_v)^2 L\,\upsilon_s\ (k = 5)$	(11–4)
Rose (Rose, 1945, 1949)	
$h = \frac{1.067}{\psi}C_d\frac{1}{\alpha^4}\frac{L}{d}\frac{\upsilon_s^2}{g}$	(11–5)
$h = \frac{1.067}{\psi}\frac{L\upsilon_s^2}{\alpha^4 g}\Sigma C_d\frac{p}{d_g}$	(11–6)
$C_d = \frac{24}{N_R} + \frac{3}{\sqrt{N_R}} + 0.34$	(11–7)
$N_R = \frac{\psi d\upsilon_s\rho}{\mu}$	(11–8)
Ergun (Ergun, 1952)	
$h = \frac{f}{\psi}\frac{(1 - \alpha)}{\alpha^3}\frac{L}{d}\frac{\upsilon_s^2}{g}$	(11–9)
$f = 150\frac{(1 - \alpha)}{N_R} + 1.75$	(11–10)
N_R = See Eq.(11–8)	

C = coefficient of compactness (varies from 600 for very closely packed sands that are not quite clean to 1200 for very uniform clean sand)
C_d = coefficient of drag
d = grain size diameter, m (ft)
d_g = geometric mean diameter between sieve sizes d_1 and d_2, $\sqrt{d_1 d_2}$, mm (in.)
d_{10} = effective medium size diameter, mm (in.)
f = friction factor
g = acceleration due to gravity, 9.81 m/s^2 (32.2 ft/s^2)
h = headloss, m (ft)
k = filtration constant, 5 based on sieve openings, 6 based on size of separation
L = depth of filter bed or layer, m (ft)
N_R = Reynolds number
p = fraction of particles (based on mass) within adjacent sieve sizes
S = shape factor (varies between 6.0 for spherical particles and 6/ψ for nonshperical particles)
S_v = specific surface area (A_p/V_p) is equal to 6/d for spheres and 6/d ψ for nonshperical particles
T = temperature, °C [°F in Eq. (11–10)]
υ_h = superficial (approach) filtration velocity, m/d (ft/d)
υ_s = superficial (approach) filtration velocity, m/s (ft/s)
α = porosity
μ = viscosity, N·s/m^2 (lb·s/ft^2)
ν = kinematic viscosity, m^2/s (ft^2/s)
ρ = density, = kg/m^3 (slug/ft^3, lb·s^2/ft^4)
ψ = sphericity, often identified as ϕ in the literature (1.0 for spheres, 0.94 for worn sand, 0.81 for sharp sand, 0.78 for angular sand, 0.70 for crushed coal and sand)

[a] Although known as the Kozeny-Carman equation, Blake (1922) should also be credited with its development.

the transitional or turbulent flow regimes are important because many of the newer filters are deeper with larger filter media and operate at higher filtration rates. In the Rose equation [Eq. (11–5)], use of the coefficient of drag, C_d [Eq. (11–7)], makes it possible to capture the affect of varying flow regimes from viscous to inertial. Similarly, in the Ergun friction equation [Eq. (11–10)], the first term accounts for viscous energy losses and the second term accounts for inertial energy losses. Based on a review of the literature, Trussell and Chang (1999) proposed some different coefficients for sand and anthracite for use in the Ergun equation [Eq. (11–10)].

The summation term in Eqs. (11–3) and (11–6) is included to account for the stratification that occurs in filters. To account for stratification, the mean size of the material retained between successive sieve sizes is assumed to correspond to the mean size of the successive sieves (see Table 11–6), assuming that the particles retained between sieve sizes are substantially uniform (Fair and Hatch, 1933). The Ergun equation [Eq. (11–9)] can also be applied to successive layers in a stratified filter.

Sphericity, Specific Surface Area, and Shape Factor. In applying the equations given in Table 11–8, some confusion exists over the definition of sphericity, ψ, specific surface area, S_v, and shape factor, S. The sphericity factor is defined as the ratio of the surface area of a sphere with the same volume as a given particle to the surface area of the particle and is given by the following formula (Wadell, 1935).

$$\psi = \frac{\pi^{1/3}(6V_p)^{2/3}}{A_p} \tag{11–11}$$

where ψ = sphericity, dimensionless
V_p = equivalent volume sphere, L^3 (m^3)
A_p = actual surface area of particle, L^2 (m^2)

Thus, for a spherical particle, the sphericity factor is equal to 1.0. Typically, sphericity factors can be applied to discrete particles and can vary from 1.0 for spheres to 0.70 for crushed sand. Further, because sphericity is difficult to measure, typical values are derived from experimental observations (Carman, 1937).

The specific surface area, S_v, defined as the area to volume ratio, is given by the following expressions for spherical and nonspherical particles.
For spherical particles

$$S_v = \frac{A_p}{V_p} = \frac{\pi d^2}{(\pi d^3/6)} = \frac{6}{d}, \text{ and} \tag{11–12a}$$

For nonspherical (irregular) particles

$$S_v = \frac{A_p}{V_p} = \frac{6}{\psi d} \tag{11–12b}$$

where S_v = specific surface area, m, mm
A_p = surface area of filter medium particle, m^2, mm^2
V_p = volume of filter medium particle, m^3, mm^3
d = diameter of filter medium particle, m, mm
ψ = sphericity, dimensionless

In the literature, the number 6 that appears in the above equations has been identified as a shape factor S for spherical particles and $6/\psi$ for nonspherical particles [see Eq. (11–2) in Table 11–8] (Fair et al., 1968). Computation of the clean-water headloss through a filter is illustrated in Example 11–1.

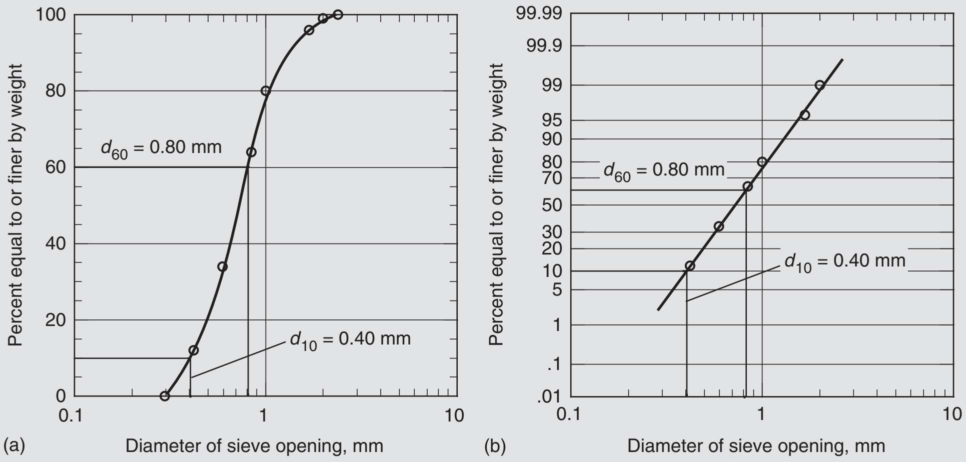
EXAMPLE 11–1 **Determination of Clean-water Headloss in a Granular Medium Filter** Determine the effective size, the uniformity coefficient, and the clean-water headloss in a filter bed composed of 0.75 m of uniform sand with the size distribution given below for a filtration rate of 160 L/m^2·min. Assume that the operating temperature is 20°C. Use the Rose equation [Eq. (11–6)] given in Table 11–8 for computing the headloss. Assume the porosity of the sand in the various layers is 0.40 and use a value of 0.85 for the sphericity factor for sand.

Sieve size or number	Percent of sand retained	Cumulative percent passing	Geometric mean size^a, mm
6–8	0	100	
8–10	1	99	2.18
10–12	3	96	1.83
12–18	16	80	1.30
18–20	16	64	0.92
20–30	30	34	0.71
30–40	22	12	0.50
40–50	12		0.35

^{a} Using sieve size data from Table 11–6, the geometric mean size $= \sqrt{d_1 d_2}$

Solution

1. Determine the effective size and the uniformity coefficient of the sand. Plot the cumulative percent passing versus the corresponding sieve size. Two different methods of plotting the data are presented below.



a. The effective size, d_{10}, read from the graphs is 0.40 mm
b. The uniformity coefficient is

$$\text{UC} = \frac{d_{60}}{d_{10}} = \frac{0.80\text{ mm}}{0.40\text{ mm}} = 2.0$$

2. Determine the clean-water headloss using Eq. (11–6).

$$h = \frac{1.067}{\psi}\frac{Lv_s^2}{\alpha^4 g}\sum C_d \frac{p}{d_g}$$

a. Set up computation table to determine the summation term in Eq. (11–6)

Sieve size or number	Fraction of sand retained	Geometric mean size, mm	Reynolds number	C_d	$C_d\left(\frac{p}{d}\right)$, m^{-1}
8–10	0.01	2.18	4.93	6.56	30
10–12	0.03	1.83	4.15	7.60	124
12–18	0.16	1.30	2.93	10.28	1268
18–20	0.16	0.92	2.08	13.99	2441
20–30	0.30	0.71	1.60	17.71	7509
30–40	0.22	0.50	1.13	24.38	10,729
40–50	0.12	0.35	0.80	33.73	11,459
Sum					33,560

b. Determine the Reynolds number for each geometric mean as illustrated below.

$$N_R = \frac{\psi d v_s \rho}{\mu} = \frac{\psi d v_s}{\nu}$$

$d = 2.18$ mm

$$v_s = \left(\frac{160\,\text{L}}{\text{m}^2\cdot\text{min}}\right)\left(\frac{1\ \text{m}^3}{1000\,\text{L}}\right)\left(\frac{1\ \text{min}}{60\,\text{s}}\right) = 0.00267\ \text{m/s}$$

$\nu = 1.003 \times 10^{-6}\ \text{m}^2/\text{s}$ (see Appendix C)

$$N_R = \frac{(0.85)(0.00218\,\text{m})(0.00267\,\text{m/s})}{(1.003 \times 10^{-6}\,\text{m}^2/\text{s})}$$

$N_R = 4.93$

c. Determine C_d using Eq. (11–7)

$$C_d = \frac{24}{N_R} + \frac{3}{\sqrt{N_R}} + 0.34$$

$$C_d = \frac{24}{4.93} + \frac{3}{\sqrt{4.93}} + 0.34 = 6.56$$

d. Determine the headloss through the stratified filter bed using Eq. (11–6).

$L = 0.75$ m
$v_s = 0.00267$ m/s
$\psi = 0.85$
$\alpha = 0.40$
$g = 9.81\ \text{m/s}^2$

$$h = \frac{1.067(0.75\,\text{m})(0.00267\,\text{m/s})^2}{(0.85)(0.4)^4(9.81\,\text{m/s}^2)}(33{,}560/\text{m})$$

$h = 0.90$ m

Comment Given that the Reynolds numbers are less than 6 (i.e., 4.93 or less), most of the headloss in this example is due to viscous forces operative in the laminar flow region, which is reflected by the first term in the C_d relationship [Eq. (11–7)]. As larger media are used and the flowrate increases, the second and third terms of the C_d relationship will have a greater impact. Although many equations have been proposed over the years, the Rose equation has proven to be quite satisfactory for estimating the clear-water headloss in granular medium filter beds for a variety of flow regimes.

Backwash Hydraulics. To understand what happens during the backwash operation it will be helpful to refer to Fig. 11–7 in which the pressure drop across a packed bed is illustrated as the upward backwash velocity through it increases. Between points *A* and *B*, the bed is stable, and the pressure drop and Reynolds number N_R are related linearly. At point *B*, the pressure drop essentially balances the weight of the filter. Between points *B* and *C* the bed is unstable, and the particles adjust their position to present as little resistance to flow as possible. At point *C*, the loosest possible arrangement is obtained in which the particles are still in contact. Beyond point *C*, the particles begin to move freely but collide frequently so that the motion is similar to that of particles in hindered settling. Point *C* is referred to as the "point of fluidization." By the time point *D* is reached, the particles are all in motion, and, beyond this point, increases in N_R result in very small increases in ΔP as the bed continues to expand and the particles move in more rapid and more independent motion. Ultimately, the particles will stream with the fluid, and the bed will cease to exist at point *E*.

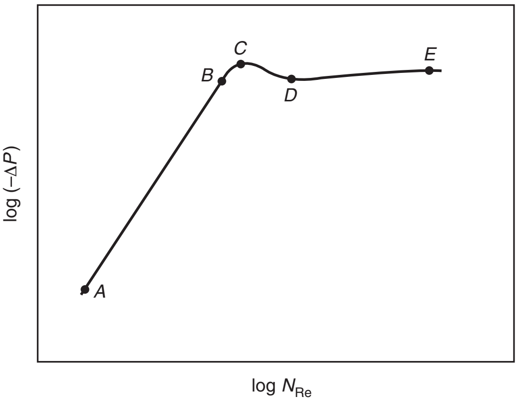
To expand a filter bed comprised of a uniform filter medium hydraulically, the headloss must equal the buoyant mass of the granular medium in the fluid. Mathematically this relationship can be expressed as

$$h = L_e(1 - \alpha_e)\left(\frac{\rho_m - \rho_w}{\rho_w}\right) \tag{11–13}$$

where h = headloss required to expand the bed
L_e = the depth of the expanded bed
α_e = the expanded porosity
ρ_m = density of the medium
ρ_w = density of water

Figure 11–7

Schematic diagram illustrating the fluidization of a filter bed. (Adapted from Foust et al., 1960.)



Because the individual particles are kept in suspension by the drag force exerted by the rising fluid it can be shown from settling theory (see Sec. 5–5 in Chap. 5) that

$$C_D A_p \rho_w \frac{v^2}{2} \phi(\alpha_e) = (\rho_m - \rho_w) g v_p \tag{11–14}$$

where v = face velocity of backwash water, m/s
$\phi(\alpha_e)$ = correction factor to account for the fact that v is the velocity of the backwash water and not the particle-settling velocity v_p

other terms are as defined previously.

From experimental studies (Fair, 1951; Richardson and Zaki, 1954) it has been found that the expanded bed porosity can be approximated using the following relationships, assuming the Reynolds number is approximately one.

$$\phi(\alpha_e) = \left(\frac{v_s}{v}\right)^2 = \left(\frac{1}{\alpha_e}\right)^9 \tag{11–15}$$

Thus

$$\alpha_e = \left(\frac{v}{v_s}\right)^{0.22} \tag{11–16}$$

or

$$v = v_s \alpha_e^{4.5} \tag{11–17}$$

where v_s = settling velocity of particle

However, because the volume of the filtering medium per unit area remains constant, $(1 - \alpha)L$ must be equal to $(1 - \alpha e)L_e$ so that

$$\frac{L_e}{L} = \frac{1 - \alpha}{1 - \alpha_e} = \frac{1 - \alpha}{1 - (v/v_s)^{0.22}} \tag{11–18}$$

Where the filter medium is stratified, the smaller and lighter particles in the upper layers expand first. To expand the entire bed, the backwash velocity must be sufficient to lift the largest and heaviest particle. To account for filter bed stratification, Eq. (11–18) is modified assuming that particles retained between sieve sizes are substantially uniform (Fair and Hatch, 1933).

$$\frac{L_e}{L} = (1 - \alpha) \sum \frac{p}{(1 - \alpha_e)} \tag{11–19}$$

Where p = fraction of filter medium retained between sieve sizes

Thus, the required backwash velocity and expanded depth can be estimated using Eqs. 11–18 and 11–19, respectively, as illustrated in Example 11–2. Additional details on filter bed expansion may be found in Amirtharajah (1978), Cleasby and Fan (1982), Dharmarajah and Cleasby (1986), Kawamura (2000), Leva (1959), and Richardson and Zaki (1954).

EXAMPLE 11–2 **Determination of Required Backwash Velocities for Filter Cleaning** A stratified sand bed with the size distribution given below is to be backwashed at a rate of 0.75 $m^3/m^2 \cdot min$. Determine the degree of expansion and whether the proposed backwash rate will expand all of the bed. Assume the following data are applicable:

Sieve size or number	Percent of sand retained	Geometric mean size[a], mm
8–10	1	2.18[b]
10–12	3	1.83
12–18	16	1.30
18–20	16	0.92
20–30	30	0.71
30–40	22	0.50
40–50	12	0.35

[a] Based on sieve sizes given in Table 11–6.

[b] $2.18 = \sqrt{2.38 \times 2.0}$.

1. Granular medium = sand
2. Specific gravity of sand = 2.65
3. Depth of filter bed = 0.90 m
4. Temperature = 20°C

Solution

1. Set up computation table to determine the summation term in Eq. (11–19).

$$\frac{L_e}{L} = (1 - \alpha)\sum \frac{p}{(1 - \alpha_e)}$$

Sieve size or number	Percent of sand retained[a]	Geometric mean size, mm	v_s, m/s	v/v_s	α_e	$p/(1 - \alpha_e)$
8–10	1	2.18	0.304	0.041	0.496	1.98
10–12	3	1.83	0.270	0.046	0.509	6.11
12–18	16	1.30	0.210	0.060	0.538	34.62
18–20	16	0.92	0.157	0.080	0.573	37.51
20–30	30	0.71	0.123	0.102	0.605	75.97
30–40	22	0.50	0.085	0.146	0.655	63.81
40–50	12	0.35	0.055	0.227	0.722	43.15
Summation						263.15

[a] For ease of computation, the percentage value is used instead of the decimal fractional value.

a. Determine the particle settling velocity using Fig. 5–20 in Chap. 5. Alternatively the particle settling velocity can be computed as illustrated in Example 5–5. The settling velocity values from Fig. 5–20 are entered in the computation table.

b. Determine the values of v/v_s and enter the computed values in the computation table.

The backwash velocity is

$v = 0.75 \text{ m/min} = 0.0125 \text{ m/s}$

c. Determine the values of α_e and enter the computed values in the computation table.

$$\alpha_e = \left(\frac{v}{v_s}\right)^{0.22} = \left(\frac{0.0125}{0.304}\right)^{0.22} = 0.496$$

d. Determine the values for column 7 and enter the computed values in the computation table.

$$\frac{p}{1 - \alpha_e} = \frac{0.01}{1 - 0.496} = 0.02$$

2. Determine the expanded bed depth using Eq. (11–19).

$$\frac{L_e}{L} = (1 - \alpha)\sum\frac{p}{(1 - \alpha_e)}\left(\frac{1}{100}\right)$$

$$L_e = (0.9 \text{ m})\ (1 - 0.4)(263.15)(1/100) = 1.42 \text{ m}$$

3. Because the expanded porosity of the largest size fraction (0.496) is greater than the normal porosity of the filter material, the entire filter bed will be expanded.

Comment The expanded depth needs to be known to establish the minimum height of the washwater troughs above the surface of the filter bed. In practice, the bottom of the backwash water troughs is set from 50 to 150 mm (2 to 6 in) above the expanded filter bed and filter expansion of 30 to 50 percent is used commonly. The width and depth of the troughs should be sufficient to handle the volume of backwash water used to clean the bed, with a minimum freeboard of 600 mm (24 in) above the top of the trough.

Modeling the Filtration Process

The modeling of the filtration process involves the development of equations to describe the (1) removal of suspended solids with time and distance within the filter bed and (2) buildup of headloss as suspended solids are removed from the liquid passing through the filter.

Removal of Suspended Solids. In general, the mathematical modeling of the time-space removal of particulate matter within the filter is based on a consideration of the equation of continuity, together with an auxiliary rate equation. The equation of continuity for the filtration operation may be developed by considering a suspended solids mass balance for a section of filter of cross-sectional area A, and of thickness Δz, measured in the direction of flow as illustrated on Fig. 11–8. Following the approach outlined in Chap. 1, the resulting equation is:

$$-v\frac{\partial C}{\partial z} = \frac{\partial q}{\partial t} + \alpha(t)\frac{\partial \overline{C}}{\partial t} \qquad (11\text{–}20)$$

where v = filtration velocity, L/m²·min

$\partial C/\partial z$ = change in concentration of suspended solids in fluid stream with distance, g/m³·m

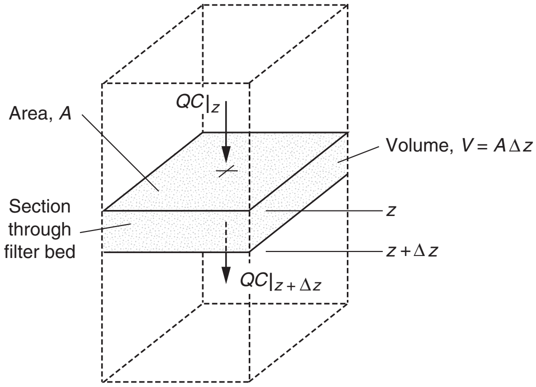
$\partial q/\partial t$ = change in quantity of solids deposited within the filter with time, g/m³·min

$\alpha(t)$ = average porosity as a function of time

$\partial \overline{C}/\partial t$ = change in average concentration of solids in pore space with time, g/m³·min

Figure 11–8

Definition sketch for the analysis of the filtration process.



In Eq. (11–20), the first term represents the difference between the mass of suspended solids entering and leaving the section, the second term represents the time rate of change in the mass of suspended solids accumulated within the interstices of the filter medium, and the third term represents the time rate of change in the suspended solids concentration in the pore space within the filter volume.

Because the quantity of fluid contained within the bed in a flowing process is usually small compared with the volume of liquid passing through the bed, the materials balance equation can be written as:

$$-v\frac{\partial C}{\partial z} = \frac{\partial q}{\partial t} \tag{11–21}$$

This equation is the one most commonly found in the literature dealing with filtration theory.

To solve Eq. (11–21), an additional independent equation is required. The most direct approach is to derive a relationship that can be used to describe the change in concentration of suspended matter with distance, such as

$$\frac{\partial C}{\partial z} = \phi(V_1, V_2, V_3 \ldots) \tag{11–22}$$

in which V_1, V_2, and V_3 are the variables governing the removal of suspended matter from solution. An alternative approach is to develop a complementary equation in which the pertinent process variables are related to the amount of material retained (accumulated) within the filter at various depths. In equation form, this may be written as

$$\frac{\partial q}{\partial t} = \phi(V_1, V_2, V_3 \ldots) \tag{11–23}$$

Using one or the other of the above expressions, [Eq. (11–21) or (11–22)], a number of solutions have been proposed for the continuity equation [Eq. (11–21)] (Caliskaner and Tchobanoglous, 2000).

Headloss Development. In the past, the most commonly used approach to determine headloss in a clogged filter was to compute it with a modified form of the equations used to evaluate the clear-water headloss (see Table 11–8). In all cases, the difficulty encountered in using these equations is that the porosity must be estimated for various degrees of clogging. Unfortunately, the complexity of this approach renders most of these formulations useless or, at best, extremely difficult to use. An alternative approach is to relate the development of headloss to the amount of material removed by the filter. The headloss would then be computed using the expression

$$H_t = H_o + \sum_{i=i}^{n} (h_i)_t \tag{11–24}$$

where H_t = total headloss at time t, m (ft)
H_o = total initial clean-water headloss, m (ft)
$(h_i)_t$ = headloss in the ith layer of the filter at time t, m (ft)

From an evaluation of the incremental headloss curves for uniform sand and anthracite, the buildup of headloss in an individual layer of the filter was found to be related to the amount of material contained within the layer. The form of the resulting equation for headloss in the ith layer is

$$(h_i)_t = a(q_i)_t^b \tag{11–25}$$

where $(q_i)_t$ = amount of material deposited in the ith layer at time t, mg/cm^3
a, b = constants

In this equation, it is assumed that the buildup of headloss is only a function of the amount of material removed. The application of these modeling equations may be found in the 3rd and 4th editions of this textbook.

11–5 DEPTH FILTRATION: SELECTION AND DESIGN CONSIDERATIONS

The ability to select and design filter technologies must be based on (1) knowledge of the types of filters that are available, (2) a general understanding of their performance characteristics, and (3) an appreciation of the process variables controlling depth filtration. Important design considerations for effluent filtration systems include (1) filter influent wastewater characteristics, (2) design and operation of the biological treatment process, (3) type of filtration technology to be used, (4) available flow control options, (5) type of filter backwashing system to be employed, (6) necessary filter appurtenances, and (7) filter control systems and instrumentation (not considered in this textbook). An understanding of issues related to effluent filtration with chemical addition, the type of filter problems encountered in the field, and the importance of pilot plant studies is also necessary. These subjects are presented and discussed in this section.

Available Filtration Technologies

The principal types of depth filters that have been used for the filtration of wastewater are described in Table 11–9. As shown in Table 11–9, the filters can be classified in terms of their operation as semi-continuous or continuous. Filters that must be taken offline periodically to be backwashed are classified operationally as semi-continuous. Filters in which the filtration and backwash operation occurs simultaneously are classified as continuous. Within each of these two classifications there are a number of different types of filters depending on bed depth (e.g., shallow, conventional, and deep bed), the type of filtering medium used (mono-medium, dual-, and multi-media), whether the

Table 11–9

Comparison of principal types of granular and synthetic medium filters

Type of filter	Type of filter operation	Filter bed details[a]		Typical direction of flow	Backwash operation	Flowrate through filter	Solids storage location	Type of design	Remarks
		Type	Filtering medium						
Conventional	Semi-continuous	Mono-medium (stratified or unstratified)	Sand or anthracite	Downward	Batch	Constant/ variable	Surface and upper bed	Individual	Rapid headloss buildup
Conventional	Semi-continuous	Dual-media (stratified	Sand and anthracite	Downward	Batch	Constant/ variable	Internal	Individual	Dual-media design used to extend length of filter run
Conventional	Semi-continuous	Multi-media (stratified)	Sand, anthracite, and garnet	Downward	Batch	Constant/ variable	Internal	Individual	Multi-media design used for particle depth penetration
Deep bed	Semi-continuous	Mono-medium (stratified or unstratified)	Sand or anthracite	Downward	Batch	Constant/ variable	Internal	Individual	Deep bed used to store solids and extend length of filter run
Deep bed	Semi-continuous	Mono-medium (stratified)	Sand	Upward	Batch	Constant	Internal	Proprietary	Deep bed used to store solids and extend length of filter run
Deep bed	Semi-continuous	Mono-medium (unstratified)	Sand	Upward	Continuous	Constant	Internal	Proprietary	Sand bed moves in countercurrent direction to fluid flow
Pulsed bed	Semi-continuous	Mono-medium (stratified)	Sand	Downward	Batch	Constant	Surface and upper bed	Proprietary	Air pulses used to break up surface mat and increase run length
Fuzzy filter	Semi-continuous	Mono-medium (unstratified)	Synthetic fiber	Upward	Batch	Constant	Internal	Proprietary	Perferorated plate is used to retain the filter medium during backwash
Traveling bridge	Continuous	Mono-medium (stratified)	Sand	Downward	Semi-continuous	Constant	Surface and upper bed	Proprietary	Individual filter cells backwashed sequentially
Traveling bridge	Continuous	Dual-media (stratified)	Sand and anthracite	Downward	Semi-continuous	Constant	Surface and upper bed	Proprietary	Individual filter cells backwashed sequentially
Pressure filters	Semi-continuous	Mono medium or dual media	Sand and/or anthracite	Downward	Batch	Constant/ variable	Surface and upper bed	Individual and proprietary	Used for small plants

[a] For filter bed depths, see Tables 11–15 and 11–16.

filtering medium is stratified or unstratified, the type of operation (downflow or upflow), and the method used for the management of solids (i.e., surface or internal storage). For the mono-medium and dual-media semi-continuous filters, a further classification can be made based on the driving force (e.g., gravity or pressure) although most of the filters used commonly in wastewater applications are gravity flow. Another important distinction that must be noted for the filters identified in Table 11–9 is whether they are proprietary or individually designed.

The five types of depth filters used most commonly for wastewater filtration at larger treatment plants [greater than 1000 m^3/d (0.25 Mgal/d)] are (1) conventional downflow filters (mono-medium, dual-, and multi-media), (2) deep-bed downflow filters, (3) deep-bed upflow continuous-backwash filters, (4) synthetic medium filters, (5) the pulsed bed filter, and (6) traveling bridge filters. A two-stage deep-bed filtration system which incorporates phosphorus removal is also used. Pressure filters, which operate in the same manner as gravity filters, are used at smaller plants. Many of the filters are proprietary and are supplied by the manufacturer as a complete unit. Each of these eight filter types is described in greater detail in Table 11–10. Views of several different types of filter installations are shown on Fig. 11–9.

Performance of Different Types of Depth Filters

The critical question associated with the selection of any depth filter is whether it will perform as anticipated. Performance of depth filters can be assessed from a review of the (1) hydraulic loading rate, (2) removal of turbidity and total suspended solids, (3) variability of turbidity and TSS removal, (4) removal of different particle sizes, (5) removal of microorganisms, and (6) backwash water requirements.

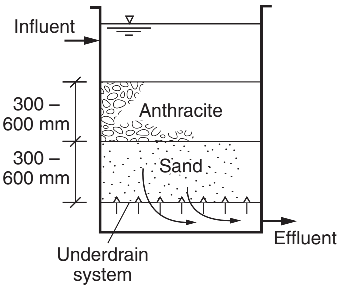
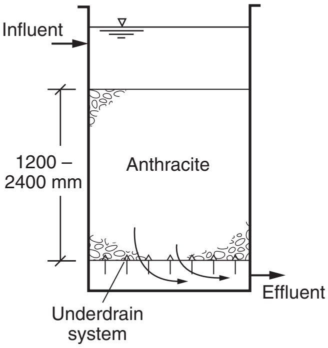
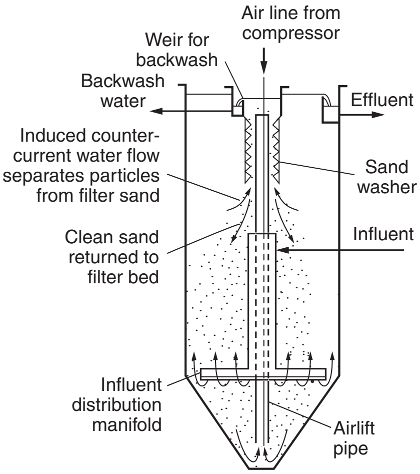
Hydraulic Loading Rate. The principal operational considerations for a depth filter are the volume of water produced in a given time period at a specified quality and the volume of washwater used to clean the filter. The volume of water produced is related to the development of headloss and filter performance, typically measured in terms of turbidity (see Fig. 11–4). The objective of a balanced filter design is to have the limiting headloss and turbidity breakthrough occur at or near the same time. In small plants, the water filtered during the ripening period is wasted during the filter-to-waste step (usually returned to the plant inflow). In large plants with many filters, the filter to waste cycle is often omitted. Chemical addition has also been used to extend the time to turbidity breakthrough and to achieve a variety of other treatment objectives including the removal of specific contaminants such as phosphorus, metal ions, and humic substances. Chemicals used commonly in effluent filtration include a variety of organic polymers, alum, and ferric chloride. It should be noted that the use of filter aid chemicals will normally result in more rapid headloss development and overuse of these chemicals can result in mudball formation.

Both the volume of water filtered and the rate at which headloss increases are related to the hydraulic loading rate (HLR). Typical operating characteristics, including hydraulic loading rates, for depth filters are reported in Table 11–11. Also reported in Table 11–11 are the filtration rates allowed by the California Department of Public Health for various filters in reuse applications. Because of the wide variation in the allowable rates, pilot plant studies are recommended.

Removal of Turbidity and Total Suspended Solids. The results of long-term testing of seven different types of pilot-scale filters on the effluent from the same activated sludge process ($SRT > 8$ d), without chemical addition, are shown on Fig. 11–10.

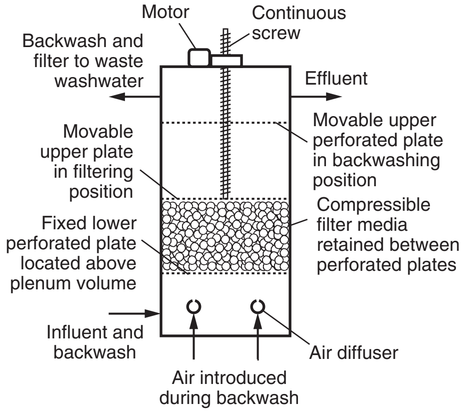
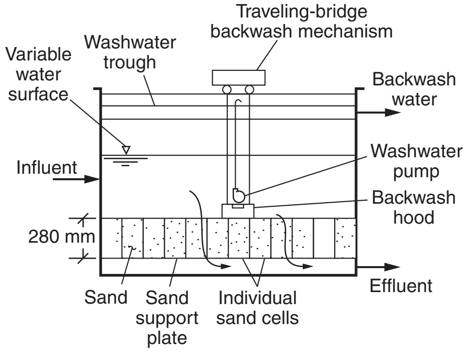
Table 11–10

Description of commonly used depth filters for reclaimed water applications[a]

Filter type	Description
(a) Conventional downflow  	Wastewater containing suspended matter is applied to the top of the filter bed. Mono-medium, dual-, or multi-media filter materials are used. Typically sand or anthracite is used as the filtering material in single-medium filters. Dual-media filters usually consist of a layer of anthracite over a layer of sand. Other combinations include (1) activated carbon and sand, (2) resin beads and sand, and (3) resin beads and anthracite. Multi-media filters typically consist of a layer of anthracite over a layer of sand over a layer of garnet or ilmenite. Other combinations include (1) activated carbon, anthracite, and sand, (2) weighted, spherical resin beads, anthracite, and sand, and (3) activated carbon, sand, and garnet.
(b) Deep-bed downflow  	The deep-bed downflow filter is similar to the conventional downflow filter with the exception that the depth of the filter bed and the size of the filtering medium (usually anthracite) are greater than the corresponding values in a conventional filter. Because of the greater depth and larger medium size (i.e., sand or anthracite), more solids can be stored within the filter bed and the run length can be extended. The maximum size of the filter medium used in these filters depends on the ability to backwash the filter. In general, deep-bed filters are not fluidized completely during backwashing. To achieve effective cleaning, air scour plus water is used in the backwash operation.
(c) Deep-bed upflow continuous backwash  	Wastewater to be filtered is introduced into the bottom of the filter where it flows upwards through a series of riser tubes and is distributed evenly into the sand bed through the open bottom of an inlet distribution hood. The water then flows upward through the downward moving sand. Clean filtrate exits from the sand bed, overflows a weir, and is discharged from the filter. At the same time sand particles, along with trapped solids, are drawn downward into the suction of an airlift pipe which is positioned in the center of the filter. A small volume of compressed air, introduced into the bottom of the airlift, draws sand, solids, and water upward through the pipe by creating a fluid with a density less than one. Impurities are scoured (abraded) from the sand particles during the turbulent upward flow. Upon reaching the top of the airlift, the dirty slurry spills over into the central reject compartment. A steady stream of clean filtrate flows upward, countercurrent to the movement of sand, through the washer section. The upflow liquid carries away the solids and reject water. Because the sand has a higher settling velocity than the removed solids, the sand is not carried out of the filter. The sand is cleaned further as it moves down through the washer. The cleaned sand is redistributed onto the top of the sand bed, allowing for a continuous uninterrupted flow of filtrate and reject water.

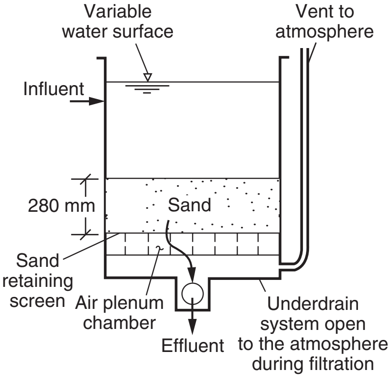
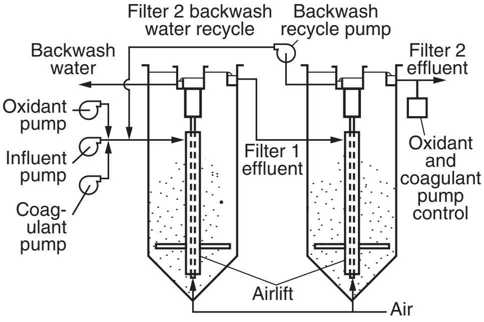
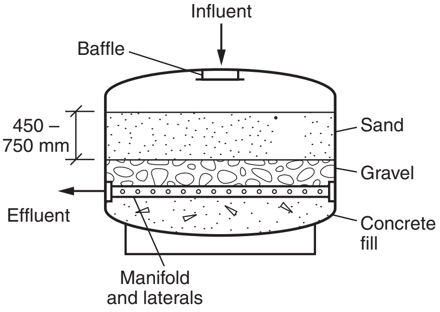
(*continued*)

Table 11–10 (*Continued*)

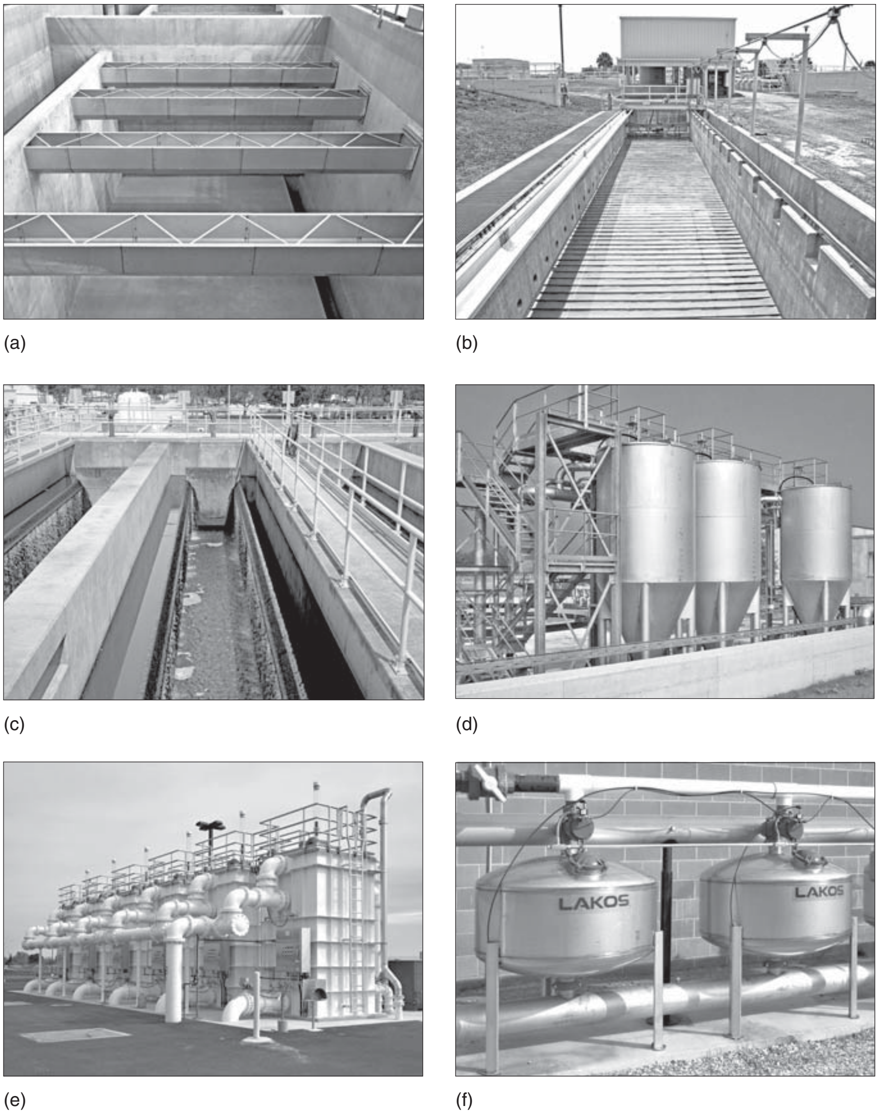
Filter type	Description
(d) Synthetic medium (Fuzzy filter)  	A synthetic medium filter, developed originally in Japan, is used for reclaimed water filtration. Unusual features of the filter are: (1) the porosity of the filter bed can be modified by compressing the filter medium and (2) the size of the filter bed is increased mechanically to backwash the filter. The filter medium, a highly porous synthetic material made of polyvaniladene, allows the influent to flow through the medium as opposed to flowing around the filtering medium, as in sand and anthracite filters. The porosity of the uncompacted quasi-spherical filter medium itself is estimated to be about 88 to 90 percent, and the porosity of the filter bed is approximately 94 percent. In the filtering mode, secondary effluent is introduced in the bottom of the filter. The influent wastewater flows upward through the filter medium, retained by two porous plates, and is discharged from the top of the filter. To backwash the filter, the upper porous plate is raised mechanically. While flow to the filter continues, air is introduced sequentially from the left and right sides of the filter below the lower porous plate, causing the filter medium to move in a rolling motion. The filter medium is cleaned by the shearing forces as the backwash water moves past the filter and by abrasion as the filter medium rubs against itself. Backwash water containing the solids removed from the filter is diverted for subsequent processing. To put the filter back into operation after the backwash cycle has been completed, the raised porous plate is returned to its original position. After a short flushing cycle, the filtered effluent valve is opened, and filtered effluent is discharged.
(e) Pulsed-bed (PBF)  	The pulsed bed filter is a proprietary downflow gravity filter with an unstratified shallow layer of fine sand as the filtering medium. The shallow bed is used for solids storage, as opposed to other shallow-bed filters where solids are principally stored on the sand surface. An unusual feature of this filter is the use of an air pulse to disrupt the sand surface and thus allow penetration of suspended solids into the bed. The air pulse process involves forcing a volume of air, trapped in the underdrain system, up through the shallow filter bed to break up the surface mat of solids and renew the sand surface. When the solids mat is disturbed, some of the trapped material is suspended but the most of solids are entrapped within the filter bed. The intermittent air pulse causes a folding over the sand surface, burying solids within the medium and regenerating the filter bed surface. The filter continues to operate with intermittent pulsing until a terminal headloss limit is reached. The filter then operates in a conventional backwash cycle to remove solids from the sand. During normal operation the filter underdrain is not flooded as it is in a conventional filter.

(*continued*)

Table 11–10 (*Continued*)

Filter type	Description
(f) Traveling bridge  	The traveling bridge filter is a proprietary continuous downflow, automatic backwash, low-head, granular medium depth filter. The bed of the filter is divided horizontally into long independent filter cells. Each filter cell contains approximately 280 mm (11 in.) of medium. Treated wastewater flows through the medium by gravity and exits to the clearwell plenum via a porous plate, polyethylene underdrain. Each cell is backwashed individually by an overhead, traveling bridge assembly, while all other cells remain in service. Water used for backwashing is pumped directly from the clearwell plenum up through the medium and deposited in a backwash trough. During the backwash cycle, wastewater is filtered continuously through the cells that are not being backwashed. The backwash mechanism includes a surface wash pump to assist in breaking up of the surface matting and "mudballing" in the medium. Because the backwashing operation is performed on an "as needed" basis, the backwash cycle is termed semi-continuous.
(g) Two-stage  	A proprietary two-stage filtration process is used for the removal of turbidity, total suspended solids, and phosphorus. Two deep-bed upflow continuous backwash filters are used in series to produce a high quality effluent. A large size sand diameter is used in the first filter to increase the contact time and to minimize clogging. A smaller sand size is used in the second filter to remove residual particles from the first stage filter. The waste washwater from the second filter which contains small particles and residual coagulant is recycled to the first filter to improve floc formation within the first stage filter and the influent to waste ratio. Based on full scale installations the reject rate has been found to be less than 5 percent. Phosphorus levels equal to or less than 0.02 mg/L have been achieved in the final filter effluent.
(h) Pressure filters  	Pressure filters operate in the same manner as gravity filters and are used at smaller plants. The only difference is that, in pressure filters, the filtration operation is carried out in a closed vessel under pressurized conditions achieved by pumping. Pressure filters normally are operated at higher terminal headlosses, resulting in longer filter runs and reduced backwash requirements. If however, they are not backwashed on a regular basis, problems have been experienced with the formation of mudballs.

Long-term data from other water reclamation plants are also shown. The principal conclusions to be reached from an analysis of the data presented on Fig. 11–10 are that (1) given a high quality filter influent (turbidity less than 5 to 7 NTU) all of the filters tested are capable of producing an effluent with an average turbidity of 2 NTU or less without chemical addition; (2) when the influent turbidity is greater than about 7 to 10 NTU,



(a) (b) (c) (d) (e) (f)

Figure 11–9

Views of typical filtration installations: (a) view of empty conventional gravity filter without underdrain system (see Fig. 11–20), but with washwater troughs in place; (b) typical traveling bridge filter (empty) with individual cells exposed; (c) deep-bed denitrifying filter; (d) continuous backwash upflow filters (courtesy Austep, Italy); (e) Fuzzy filter installation comprised of six filters; and (f) bank of small pressure filters used at small wastewater treatment plants. Additional information on these filters is presented in Tables 11–9 and 11–10.

chemical addition is required with all of the filters to achieve an effluent turbidity of 2 NTU or less; and (3) effluent quality is directly related to influent quality if chemical addition is not used. Typical values of effluent quality for turbidity and total suspended solids for depth filtration using granular media are presented in Table 11–12. For comparison, comparable data for other advanced filtration processes (e.g. membranes) used for the removal of particulate matter are presented in Table 11–31 in Sec 11–7.

Keeping in mind the limitations associated with turbidity measurements, the following two relationships can be used to approximate TSS values from measured turbidity values.

Settled secondary effluent

$$\text{TSS, mg/L} = (2.0 \text{ to } 2.4) \times (\text{turbidity, NTU}) \qquad (11\text{–}26)$$

Table 11–11

Comparison of operational characteristics for selected depth filters when filtering settled activated sludge effluent

	Filter bed details[a]		Typical operational filtration rate		Maximum filtration rate approved by CDPH[c,d]		
Type of filter	Type	Filtering medium[b]	gal/ft^2·min	m^3/m^2·min	gal/ft^2·min	m^3/m^2·min	Backwash percentage
Conventional shallow	Mono-medium	S	2–6	0.08–0.24	5	0.20	4–8
Conventional	Dual-media	S and A	2–6	0.08–0.24	5	0.20	4–8
Conventional	Multi-media	S, A, and G	2–6	0.08–0.24	5	0.20	4–8
Deep bed	Mono-medium	S	5–8	0.20–0.33	5	0.20	4–8
Deep bed	Mono-medium	A	5–8	0.20–0.33	5	0.20	4–8
Deep bed, upflow	Mono-medium	S	4–6	0.16–0.15	5	0.20	8–15
Fuzzy filter	Mono-medium	SM	15–40	0.60–1.60	40	1.60	2–5
Pulsed bed	Mono-medium	S	2–6	0.08–0.24	5	0.08	4–8
Traveling bridge	Mono-medium	S	2–5	0.08–0.2	2	0.08	4–8
Traveling bridge	Dual-media	S and A	2–5	0.08–0.2	2	0.08	4–8
Pressure filters	Mono medium or dual media	S and A, A	2–6	0.08–0.24	5	0.20	4–8

[a] For filter bed depths, see Tables 11–15 and 11–16.

[b] S =sand, A = anthracite, G = garnet, SM = synthetic medium.

[c] California Department of Public Health.

[d] For Title 22 wastewater reuse applications.

Figure 11–10

Performance data for seven different types of depth filters used for wastewater applications tested using the effluent from the same activated sludge plant at filtration rate 160 L/m^2·min (4 gal/ft^2·min) with the exception of the Fuzzy Filter which was operated at 800 L/m^2·min (20 gal/ft^2·min).

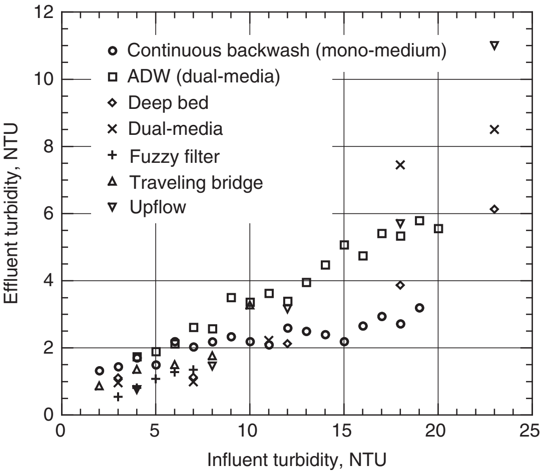


Table 11–12

Typical range of effluent quality variability observed from particulate removal processes

Particulate removal process	Unit	Typical range of effluent values	Geometric standard deviation, s_g[a] Range	Typical
Depth filtration following activated sludge process				
Turbidity	NTU	0.5–4	1.2–1.4	1.25
TSS	mg/L	2–8	1.3–1.5	1.4
Depth filtration following activated sludge with BNR				
Turbidity	NTU	0.3–2	1.2–1.4	1.25
TSS	mg/L	1–4	1.3–1.5	1.35
Surface filtration following activated sludge process				
Turbidity	NTU	0.5–2	1.2–1.4	1.25
TSS	mg/L	1–4	1.3–1.5	1.25

[a] s_g = geometric standard deviation; $s_g = P_{84.1}/P_{50}$.

Filter effluent

$$\text{TSS, mg/L} = (1.3 \text{ to } 1.6) \times (\text{turbidity, NTU}) \tag{11–27}$$

Using the above approximations, turbidity values of 5 to 7 NTU in the settled secondary effluent, which is the influent to the filter, correspond to TSS concentrations varying from about 10 to 17 mg/L, and an effluent turbidity of 2 NTU corresponds to TSS concentrations varying from 2.8 to 3.2 mg/L.

Variability in Turbidity and Total Suspended Solids Removal. In water reuse applications the variability of filter performance is of critical importance because there are specific effluent turbidity limits that must be met consistently. For example, the turbidity standard for reclaimed water for unrestricted use in California is equal to or less than 2 NTU. Because the required turbidity value is written without a decimal point, a turbidity value of 2.49 NTU is reported as 2 NTU. The variability observed in the operating data from a large water reclamation facility is illustrated on Fig. 11–11, for the years 2010 and 2011.

Comparing the mean turbidity and TSS values for the two different years, the TSS/turbidity ratios are 1.51 and 1.32, respectively, which is consistent with the range given in Eq. (11–27). The corresponding geometric standard deviations, s_g, for turbidity for the two years are 1.26 and 1.23, respectively. Similarly, the geometric standard deviations for TSS are 1.37 and 1.42, respectively. Both sets of values are consistent with the range of s_g values reported in the literature, as given in Table 11–12. Characterization of the variability in effluent constituents using the geometric standard deviation, s_g, is discussed in Appendix D. The greater the numerical value of s_g, the greater the observed range in the measured values. Use of the data in Table 11–12 is illustrated in Example 11–3.

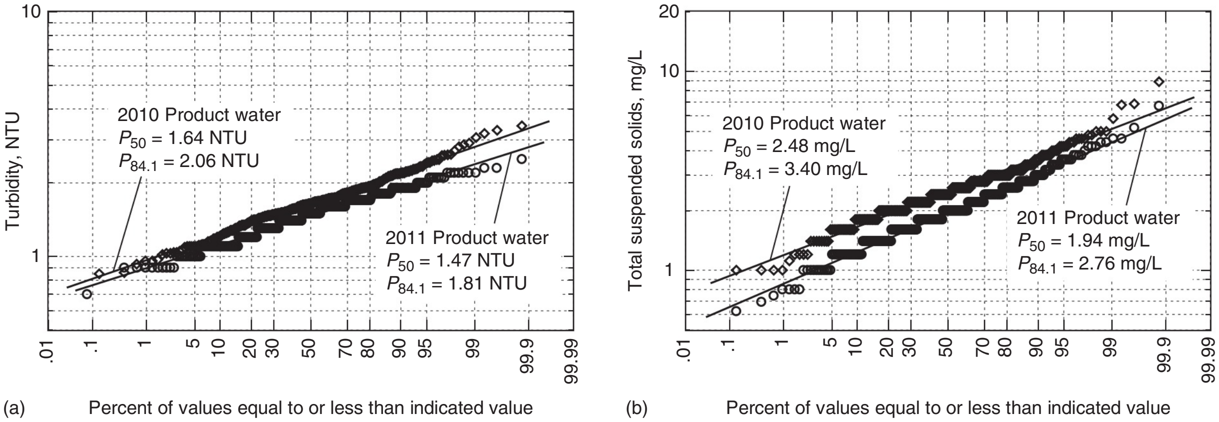


Figure 11–11

Probability distributions for filter performance for the filtration of settled activated sludge effluent from a large water reclamation facility: (a) turbidity and (b) total suspended solids.

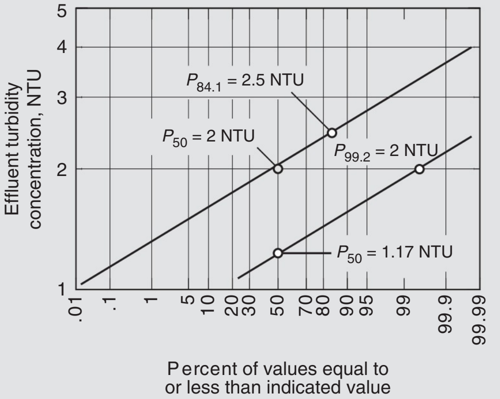
EXAMPLE 11–3 **Evaluation of the Effluent Variability of an Activated Sludge Process with Mono-medium Filtration** An activated sludge process with mono-medium filtration has been designed to have a mean effluent turbidity value of 2 NTU. Determine the maximum turbidity value that is expected to occur with a frequency of (a) once per year and (b) once every three years. If the effluent turbidity standard is 2.49 NTU, estimate how often the process will exceed the turbidity limit.

Solution

1. Select an s_g value from Table 11–12 that corresponds to the effluent turbidity for an activated sludge with filtration process. From Table 11–12, use the typical s_g value of 1.25.
2. Determine the probability distribution of the effluent turbidity values.
   a. Using the s_g value, compute the turbidity value corresponding to the plotting position on $P_{84.1}$ (see Appendix D).

   $$P_{84.1} = s_g \times P_{50} = 1.25 \times 2 \text{ NTU} = 2.5 \text{ NTU}$$

   b. Estimate the distribution of effluent turbidity values by plotting the $P_{84.1}$ and P_{50} values. As the effluent turbidity values are expected to follow a log normal distribution, a straight line can be drawn through the $P_{84.1}$ and P_{50} values, as shown on the following plot.



3. Compute the effluent turbidity value expected to occur with the frequency of interest.
   a. The probability of occurrence of a given event with a frequency of once per year is (1 / 365) x 100 = 0.3 percent. Using the plot developed in step 2, an effluent turbidity value equal to or greater than 3.5 NTU will occur 0.3 percent of the time.
   b. Similarly, turbidity values equal to greater than 3.7 NTU will occur with a frequency of once in three years (i.e., 99.9 percent).
4. Estimate how often the combined treatment process will exceed the turbidity standard of 2.49 NTU. From the plot presented in step 2, the effluent turbidity will exceed 2.49 NTU approximately 16 (100–84) percent of the time.

Comment Recognition of the variability in performance is of importance in the design of filtration systems, especially where more stringent mean effluent turbidity values must be met. For example, it the turbidity standard had been 2.0 NTU at a reliability of at least 99.2 percent (three exceedances per year), the mean design value, as illustrated in the above figure, would have to be about 1.17 NTU, assuming that the geometric standard deviation remained constant and was equal to 1.25. To reach a mean turbidity value of 1.17 NTU would, in most cases, require the addition of chemicals, although many plants with deep secondary clarifiers reach these values consistently without chemical addition.

Removal of Different Particle Sizes. Although all of the filters shown on Fig. 11–10 can produce an effluent with an average turbidity of two or less with a suitable secondary effluent, the filtered effluent particle size distribution is different for each of the filters. Typical data on the removal of particle sizes from activated sludge effluent using depth filtration are shown on Fig. 11–12. As shown, the particle removal rate is essentially independent of the filtration rate in the range from 100 to about 260 L/m^2·min. It is significant to note that most depth filters will pass some particles with diameters greater than 15 to 20 μm.

Depending on the quality of the settled secondary effluent, chemical addition has been used to improve the performance of effluent filters, with respect to turbidity. An example of the change in the distribution of particle sizes in the effluent from an activated sludge process following depth filtration without and with chemical coagulation is illustrated on

Figure 11–12

Particle size removal efficiency for a depth filter for effluent from an activated sludge plant at two different filtration rates.

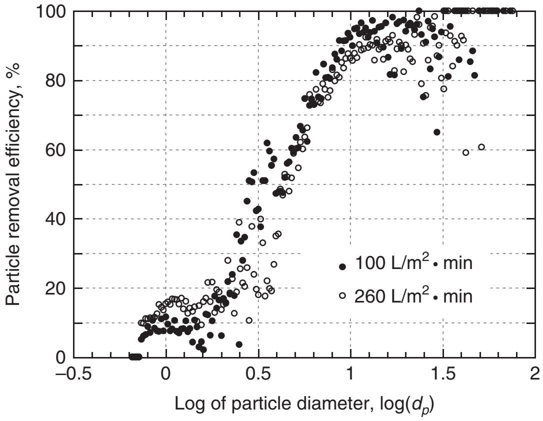


Figure 11–13

Effect of the use of chemicals on filter particle size removal performance (courtesy of K. Bourgeous, 2005): (a) original data as collected and, (b) the original data, plotted functionally according to the power law (see Example 2–4 in Chap. 2).

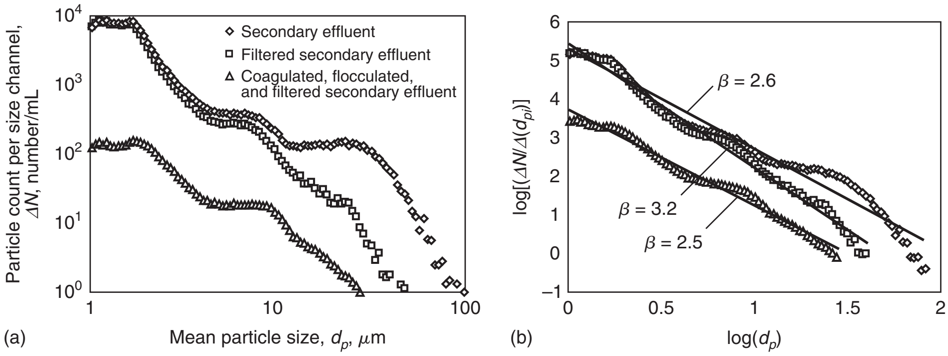


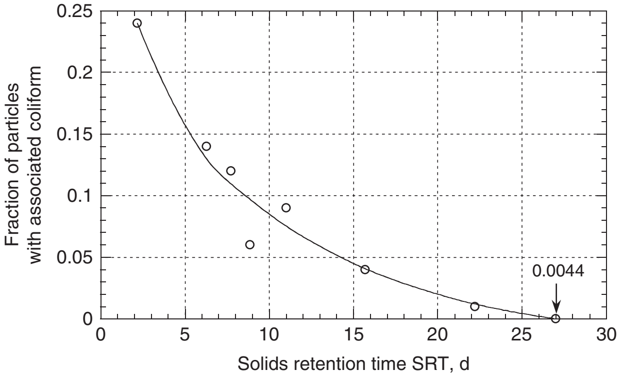
Fig. 11–13. The original data, as collected, are shown on Fig. 11–13(a). The original data, plotted functionally according to the power law (see Example 2–4 in Chap. 2), are presented on Fig. 11–13(b). As shown on Fig. 11–13(a), filtration alone only affected the larger particles, whereas with chemical coagulation all of the particles were affected more-or-less uniformly. As shown on Fig. 11–13(b), even though the number of particles in each size range was reduced by an order of magnitude, a significant number of particles remains in each size range.

Removal of Microorganisms. Where chemicals are not used, the removal of coliform bacteria and viruses from biologically treated secondary effluent is on the order of 0 to 1.0 and 0 to 0.5 logs, respectively. The degree of removal depends on the solids retention time (SRT) at which the biological process is operated. For example, as shown on Fig. 11–14, as the SRT is increased, fewer of the particles have one or more associated coliform bacteria. Typical data on the removal of the bacteriophage MS2 are illustrated on Fig. 11–15. As shown, the mean removal of MS2 across the effluent filters is about 0.3 logs. However, what is of more interest is the distribution of the removal data. Based on the distribution shown on Fig. 11–15, which is also typical for the removal of coliform organisms, allowing a disinfection credit of one log of removal for filtration in water reuse applications may not be protective of public health. Where chemicals are used, the data on the removal for microorganisms is confounded statistically. In general, it is not possible to separate the effect of chemical addition from the performance of the filter.

Backwash Water Requirements. The amount of backwash water needed to clean the filter bed, expressed as a percentage of plant flow, will depend on the characteristics of filter influent and the design of the filter bed. Typical backwash water percentages for

Figure 11–14

Number of particles with one or more associated coliform organisms as a function of the solids retention time for the activated sludge process. (Adapted from Darby, et al., 1999.)



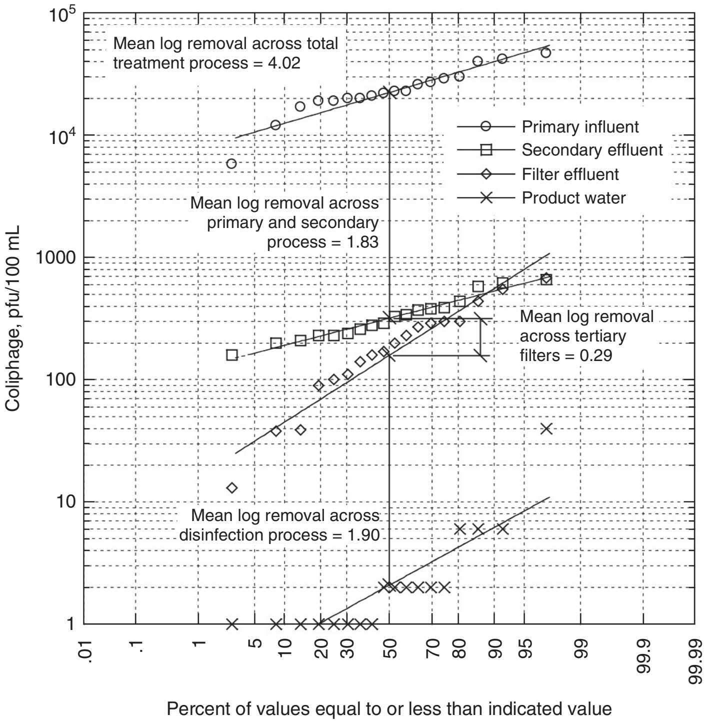


Figure 11–15

Removal of MS2 coliphage through a treatment process comprised of an activated sludge process, depth filtration, and chlorine disinfection.

depth filters shown in Table 11–11 will vary from 4 to 15 percent. By comparison, the backwash percentages for surface filters, discussed in the following section, will typically vary from 1 to 4 percent. When adding effluent filtration to an existing plant, the impact of the return backwash water on plant hydraulics must be evaluated carefully. In many locations, the percentage of backwash water will control filter selection. The volume of backwash water required to wash a filter is related to filter area and desired degree of backwash bed expansion, the size of the media, and water temperature.

Considerations Related to Design and Operation of Treatment Facilities

The design and operation of the treatment facilities can have a significant impact on the performance of depth filters. Issues related to the design and operation of wastewater treatment facilities will depend on whether filtration is being considered for an existing or new facility. Both existing and new facilities are considered in the following discussion.

Existing Treatment Plants. Where effluent filtration must be added to or retrofitted to an existing treatment plant it will be important to consider the factors that can impact the effluent quality including (1) the design and operation of the secondary sedimentation facilities, (2) the type and operation of biological treatment process, (3) the operation of the treatment plant with respect to return flows, and (4) the potential use of flow equalization.

Design and Operation of Secondary Sedimentation Facilities. The concentration and variability of the suspended solids and colloidal material in the effluent from sedimentation facilities will vary depending on the characteristics of the biological solids to be settled; the depth of the sedimentation basins; the use of energy dissipation inlets (see Fig. 8–54); the presence of sludge density currents, as described in Chap. 5; short circuiting caused by a variety of factors and the placement of the effluent weirs. Typically, the TSS concentration in the effluent from activated sludge and trickling filter plants varies between 6 and 30 mg/L. Corresponding turbidity values, as discussed previously, can vary from 3 to 15 NTU. It should be noted that properly designed deep secondary sedimentation tanks 6 to 7 m (19.5 to 23 ft) can produce an average settled effluent with turbidity of 2 NTU or less consistently, and in some cases less than 1 NTU. Depending on the concentration and variability of the suspended solids in the effluent from the secondary sedimentation it may be appropriate to consider a filter that can continue to function even when heavily loaded. Both downflow and upflow deep-bed coarse medium filters, the fuzzy filter, and the pulsed-bed filter have been used in such applications.

Type and Operation of Biological Treatment Process. The three principal types of biological treatment processes used for secondary treatment are suspended growth, as exemplified by the activated sludge process; attached growth processes, as exemplified by the trickling filter process; and pond processes. The floc strength and filterability of the effluents from each of these processes will vary with the mode of operation. For example, the strength of the biological floc from the activated sludge process will vary with the mean cell-residence time, increasing with longer mean cell residence time. The increased strength derives in part from the production of extracellular polymers as the mean cell residence time is lengthened. At extremely long mean cell residence times (15 d and longer), it has been observed that the floc strength will decrease due to floc breakup. It has also been observed that the residual suspended solids remaining after secondary clarification from treatment processes with extremely long mean cell residence times are far more difficult to filter and as a result can impact the performance of disinfection systems (Emerick, 2012). Further, the residual floc from the chemical precipitation of biologically processed wastewater may be considerably weaker than the residual biological floc before precipitation. If there is uncertainty about the characteristics of the effluent from the secondary sedimentation facilities, pilot plant studies should be conducted, as discussed subsequently.

Management of Return Flows. Currently, in most treatment plants, return flows from sludge thickeners, sludge dewatering (e.g., centrifuges, and belt presses), sludge stabilization (e.g., digester supernatant), and sludge drying facilities are returned to the wastewater treatment plant headworks for reprocessing. In many instances, these return flows contain constituents that deteriorate overall plant performance (e.g., nitrogenous compounds, colloidal material and total dissolved solids). Unfortunately, these return flows can impact the overall performance of the biological treatment process especially with respect to the removal of nitrogen and colloidal material. Flow equalization facilities and or separate systems for the treatment of return flows, as discussed in Chap. 15, are now being installed at a number of treatment plants that need to meet more stringent discharge requirements. (Tchobanoglous et al., 2011).

Flow Equalization. Flow equalization is a method used to improve the performance and variability of the downstream treatment processes and to reduce the size and cost of treatment facilities (see extended discussion in Sec. 3–7 in Chap. 3). In advanced wastewater treatment, the principal benefits include (1) reduced variability of incoming water quality; (2) enhanced performance at constant flow operation, especially for membrane processes; and (3) reduced wear and tear on membranes due to fluctuating flows and loads (Tchobanoglous et al., 2003). If full plant flow equalization is not feasible, consideration should be given to flow equalization of the return flows.

New Treatment Plants. For new wastewater treatment plants all of the factors discussed above should be considered. However, extra care should be devoted to the design of the secondary settling facilities. With properly designed settling facilities resulting in an effluent with low TSS (typically 3 to 4 mg/L or less) and turbidity (less than 1 to 2 NTU), the decision on what type of filtration system is to be used is often based on plant-related variables, such as the space available, duration of filtration period (seasonal versus year-round), the time available for construction, and costs. The most important influent characteristics in the filtration of treated secondary effluents are the suspended-solids concentration, particle size and distribution, and floc strength.

Selection of Filtration Technology

In selecting a filter technology, important factors that must be considered include (1) the required effluent quality, (2) the influent wastewater characteristics, (3) type of filter to be used: proprietary or individually designed, (4) the filtration rate, (5) filtration driving force, (6) number and size of filter units, (7) the backwash water requirements, (8) the need for pilot-plant studies, and (9) system redundancy. Each of these issues is described in Table 11–13. The importance of the influent characteristics has been discussed previously. The backwash requirements, the need for chemical addition, and the need to conduct pilot plant studies are considered in the following discussion because of the impact these factors can have on the selection of a depth filtration process.

Table 11–13

Important factors in selecting filter technology for effluent filtration applications[a]

Factor	Remarks
Required effluent quality	Usually fixed regulatory requirement, depending on the final use of the effluent.
Influent wastewater characteristics	The required effluent quality will impact the selection process, as some filters are more able to withstand periodic shock loadings. For example, wider variations in effluent quality would be expected where shallow clarifiers are used. More predictable effluent quality can be expected from deep clarifiers. In recent designs employing deep clarifiers (5 to 6 m side water depths), effluent turbidity values of less than 2 NTU are achieved consistently.
Type of filter: proprietary vs. individually designed	Currently available filter technologies are either proprietary or individually designed. With proprietary filters, the manufacturer is responsible for providing the complete filter unit and its controls, based on basic design criteria and performance specifications. In individually designed filters, the design engineer is responsible for working with several suppliers in developing the design of the system components. Contractors and suppliers then furnish the materials and equipment in accordance with the engineer's design.
Filtration rate	The filtration rate affects the areal size of the filters that will be required. For a given filter application, the rate of filtration depends primarily on floc strength and the size of the filtering medium. For example, if the strength of the floc is weak, high filtration rates tend to shear the floc particles and carry much of the material through the filter. Filtration rates generally in the range of 80 to 330 $L/m^2 \cdot min$ will not affect the effluent quality when filtering settled activated sludge effluent (see Table 11–11).
Filtration driving force	Either the force of gravity or an applied pressure force can be used to overcome the frictional resistance to flow offered by the filter bed. Gravity filters of the type discussed in Table 8-5 are used most commonly for the filtration of treated effluent at large plants. Pressure filters operate in the same manner as gravity filters and are used at smaller plants. In pressure filters, the filtration operation is carried out in a closed vessel under pressurized conditions achieved by pumping.

(*continued*)

Table 11–13 (*Continued*)

Factor	Remarks
Number and size of filtration units	The number of filter units generally should be kept to a minimum to reduce the cost of piping and construction, but it should be sufficient to assure that (1) backwash flowrates do not become excessively large and (2) that when one filter unit is taken out of service for backwashing, the transient loading on the remaining units is not excessive. Transient loadings due to backwashing are not an issue with filters that backwash continuously. To meet redundancy requirements, a minimum of two filters should be used. The sizes of the individual filter units should be consistent with the sizes of equipment available for use as underdrains, washwater troughs, and surface washers. Typically, width-to-length ratios for individually designed gravity filters vary from 1:1 to 1:4. A practical limit for the surface area on an individual depth filter (or filter cell) is about 100 m^2 (1075 ft^2), although larger filters units have been built. For proprietary filters, use standard sizes that are available from manufacturers. The surface area of a depth filter is based on the peak filtration and peak plant flowrates. The allowable peak filtration rate is usually established on the basis of regulatory requirements. Operating ranges for a given filter type are based on past experience, the results of pilot-plant studies, manufacturers recommendations, and regulatory constraints.
Backwash water requirements	As noted in Table 8–4, depth filters operate in either a semi-continuous or continuous mode. In semi-continuous operation, the filter is operated until the effluent quality starts to deteriorate or the headloss becomes excessive at which point the filter is taken out of service and backwashed to remove the accumulated solids. With filters operated in the semi-continuous mode, provision must be made for the backwash water needed to clean the filters. Typically, the backwash water is pumped from a filtered water clearwell or obtained by gravity from an elevated storage tank. The backwash storage volume should be sufficient to backwash each filter every 12 h. For filters that operate continuously such as the upflow filter and the traveling bridge filter, the filtering and backwashing phases take place simultaneously. In the traveling bridge filter, the backwash operation can either be continuous or semi- continuous as required. For filters that operate continuously, there is no turbidity breakthrough or terminal headloss.
Chemical addition	Need for chemical addition is site specific. Depending on the final use of the effluent, provision of chemical dosing facilities may be mandated by local and/or state regulations.
Pilot-plant studies	Because of the many variables involved, pilot-plant studies are often conducted when filtration facilities are to be added to an existing facility. For new plants, pilot-plant studies can be conducted at treatment plants of similar design.
System redundancy	System redundancy is related to uninterruptible power and the need to provide standby capacity for routine maintenance. Most water reclamation plants in continuous service have emergency storage and onsite power generation to operate process equipment. In general, one standby filter, as a minimum, is recommended for standby service. Where the provision of standby facilities is not possible due to space or other limitations, the filters and related piping should be sized to handle periodic overloads during maintenance periods.

[a] Adapted, in part, from Tchobanoglous et al. (2003).

Filter Backwash Water Requirements. Methods commonly used for backwashing granular medium filter beds are considered subsequently. In general, depth filters require more backwash water as compared to surface filters. For existing plants with limited hydraulic capacity, the percentage of backwash water that can be processed may be the limiting factor in the selection of a filter technology. Because it is impossible to predict a priori what percentage backwash water will be needed, pilot-plant studies must be conducted to resolve the issue.

Effluent Filtration with Chemical Addition. Depending on the quality of the settled secondary effluent, chemical addition has been used to improve the performance of effluent filters. Chemical addition has also been used to achieve specific treatment objectives including the removal of specific contaminants such as phosphorus, metal ions, and humic substances. The removal of phosphorus by chemical addition is considered in Chap. 6. To control eutrophication, the contact filtration process is used in many parts of the country to remove phosphorus from wastewater treatment plant effluents which are discharged to sensitive water bodies. The two-stage filtration process, described in Table 11–10, has proven to be very effective, achieving phosphorus levels of 0.2 mg/L or less in the filtered effluent. Chemicals commonly used in effluent filtration include a variety of organic polymers, alum, and ferric chloride. Use of organic polymers and the effects of the chemical characteristics of the wastewater on alum addition are considered in the following discussion.

Use of Organic Polymers. Organic polymers are typically classified as long chain organic molecules with molecular weights varying from 10^4 to 10^6. With respect to charge, organic polymers can be cationic (positively charged), anionic (negatively charged), or nonionic (no charge). Polymers are added to settled effluent to bring about the formation of larger particles by bridging as described in Chap. 6. Because the chemistry of the wastewater has a significant effect on the performance of a polymer, the selection of a given type of polymer for use as a filter aid generally requires experimental testing (e.g., jar testing).

Common test procedures for polymers involve adding an initial dosage (usually 1.0 mg/L) of a given polymer and observing the effects. Depending upon the effects observed, the dosage should be increased by 0.5 mg/L increments or decreased by 0.25 mg/L increments (with accompanying observation of effects) to obtain an operating range. After the operating range is established, additional testing can be done to establish the optimum dosage. Great care must be taken to insure that the polymer is well dispersed before reaching the filter to avoid the formation of mudballs.

A recent development is the use of lower molecular weight polymers that are intended to serve as alum substitutes. When these polymers are used, the dosage is considerably higher ($\geq$10 mg/L) than with higher molecular weight polymers (0.25 to 1.25 mg/L). As with the mixing of alum, the initial mixing step is critical in achieving maximum effectiveness of a given polymer. In general, mixing times of less than 1 second with G values of $>2500\ \text{s}^{-1}$ are recommended (see Table 5–9 in Chap. 5). It should be noted that, as a practical matter, as treatment plants get larger it is difficult to achieve mixing times less than one second unless multiple mixing devices are used.

Effects of Chemical Characteristics of Wastewater on Alum Addition. As with polymers, the chemical characteristics of the treated wastewater effluent can have a significant impact on the effectiveness of aluminum sulfate (alum) when it is used as an aid to filtration. For example, the effectiveness of alum is dependent on pH (see Fig. 6–9 in Chap. 6). Although Fig. 6–9 was developed for water treatment applications, it has been found to apply to most wastewater effluent filtration uses with minor variations. As shown on Fig. 6–9, the approximate regions in which the different phenomena associated with particle removal in conventional sedimentation and filtration processes are operative are plotted as a function of the alum dose and the pH of the treated effluent after alum has been added. For example, optimum particle removal by sweep floc occurs in the pH range of 7 to 8 with an alum dose of 20 to 60 mg/L. Generally, for many wastewater effluents that have high pH values (e.g. 7.3 to 8.5), low alum dosages in the range of 5 to 10 mg/L will not be effective. To operate with low alum dosages, pH control will generally be required.

Need for Bench-Scale and Pilot-Plant Studies. Although the information presented earlier in this section and previously in Sec. 11–3 will help the reader understand the nature of the filtration operation as it is applied to the filtration of treated wastewater, it must be stressed that there is no generalized approach to the design of full-scale filters. The principal reason is the inherent variability in the characteristics of the influent suspended solids to be filtered. For example, changes in the degree of flocculation of the suspended solids in the secondary settling facilities will significantly affect the particle sizes and their distribution in the effluent, which in turn will affect the performance of the filter. Further, because the characteristics of the effluent suspended solids will also vary with the organic loading on the process as well as with the time of day, filters must be designed to function under a rather wide range of operating conditions. The best way to ensure that the filter configuration selected for a given application will function properly is to conduct pilot-plant studies (see Fig. 11–16).

Because of the many variables that can be analyzed, care must be taken not to change more than one variable at a time so as to confound the results in a statistical sense. Bench-scale and pilot-plant testing should be carried out at several intervals, ideally throughout a full year, to assess seasonal variations in the characteristics of the effluent to be filtered. All test results should be summarized and evaluated in different ways to ensure their proper analysis. Because the specific details of each test program will be different, no generalization on the best method of analysis can be given.

Design Considerations for Granular Medium Filters

As noted in Table 11–9, the currently available filter technologies are either proprietary or individually designed. With proprietary filters, the manufacturer is responsible for providing the complete filter unit and its controls, based on basic design criteria and performance specifications. In individually designed filters, the designer is responsible for working with several suppliers in developing the design of the system components. Contractors and suppliers then furnish the materials and equipment in accordance with the engineer's design.

Because granular medium filters are still designed individually, important design considerations for depth filters are summarized in Table 11–14. Although some of the factors listed in Table 11–14 have been discussed previously and other design details are beyond the scope of this textbook, it is nevertheless important to consider the selection of the type of filter bed and filter medium(s) used in depth filters, the backwashing operation, and filter appurtenances.

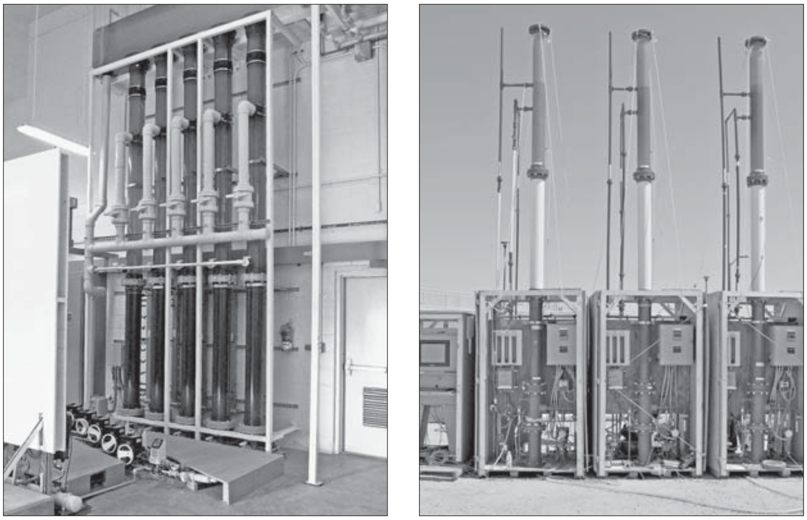


Figure 11–16

Views of filtration pilot plant: (a) filter columns fed from the source and (b) instrumentation used to monitor filter performance including turbidity and particle size counting.

Table 11–14

Design considerations for granular medium filters for effluent for filtration

Variable	Significance
1. Required effluent quality	Usually fixed regulatory requirement, depending on the final use of the effluent.
2. Influent wastewater characteristics	Considered in previous section.
3. Filter medium characteristics a. Effective size, d_{10} b. Uniformity coefficient, UC c. Type, grain shape, density, and composition	Affects clean-water headloss, particulate matter removal efficiency, and headloss buildup.
4. Filter-bed characteristics a. Bed depth b. Porosity c. Stratification d. Degree of medium intermixing	Porosity affects the amount of solids that can be stored within the filter. Bed depth affects initial headloss, length of run. Degree of intermixing will affect performance of filter bed.
5. Filtration rate	Used in conjunction with variables 2, 3, and 4 to compute clean-water headloss. Maximum rate typically specified by regulatory agency (see Table 11–11).
6. Flowrate control	The principal methods now used to control the rate of flow through downflow gravity filters may be classified as (1) constant-rate filtration with fixed head, (2) constant-rate filtration with variable head, and (3) variable-declining-rate filtration. Other control methods are also in use.
7. Allowable headloss	Design variable, depends on whether driving force will be gravity or applied pressure.
8. Backwashing system	Methods commonly used for backwashing granular medium filter beds operated in the semi-continuous mode include (1) water backwash with auxiliary surface washwater agitation, (2) water backwash with auxiliary air scour, and (3) combined air-water backwashing. With the first two methods, fluidization of the granular medium is necessary to achieve effective cleaning of the filter bed at the end of the run. With the third method, fluidization is not necessary. Typical backwash flowrates required to fluidize various filter beds are reported in Table 11–11.
9. Backwash requirements	Affects size of filter piping and pipe gallery.
10. Filter appurtenances	Filter appurtenances include: (1) the underdrain system used to support the filtering materials, collect the filtered effluent, and distribute the backwash water and air (where used); (2) the washwater troughs used to remove the spent backwash water from the filter; and (3) the surface washing systems used to help remove attached material from the filter medium.

Filter Bed Configuration and Filter Medium. Important considerations in individually designed depth filters are the selection of the type of filter bed and the corresponding media characteristics.

Selection of Filter Bed Configuration. The principal types of non-proprietary filter bed configurations now used for wastewater filtration may be classified according to the number of filtering media that are used as mono-medium, dual-media, or multi-media beds (see Fig. 11–6). In conventional downflow filters, the distribution of grain sizes for each medium after backwashing is from small to large. Typical design data for mono-medium, and dual- and multi-media filters are presented in Tables 11–15 and 11–16, respectively.

Table 11–15

Typical design data for depth filters with mono-medium[a]

Characteristic	Value		
	Unit	Range	Typical
Shallow bed (stratified)			
Anthracite			
Depth	mm	300–500	400
Effective size	mm	0.8–1.5	1.3
Uniformity coefficient	unitless	1.3–1.8	≤1.5
Filtration rate	$m^3/m^2 \cdot min$	0.08–0.24	
Sand			
Depth	mm	300–360	330
Effective size	mm	0.45–0.65	0.45
Uniformity coefficient	unitless	1.2–1.6	≤1.5
Filtration rate	$m^3/m^2 \cdot min$	0.08–0.24	
Conventional (stratified)			
Anthracite			
Depth	mm	600–900	750
Effective size	mm	0.8–2.0	1.3
Uniformity coefficient	unitless	1.3–1.8	≤1.5
Filtration rate	$m^3/m^2 \cdot min$	0.08–0.40	
Sand			
Depth	mm	500–750	600
Effective size	mm	0.4–0.8	0.65
Uniformity coefficient	unitless	1.2–1.6	≤1.5
Filtration rate	$m^3/m^2 \cdot min$	0.08–0.24	
Deep-bed (unstratified)			
Anthracite			
Depth	mm	900–2100	1500
Effective size	mm	2–4	2.7
Uniformity coefficient	unitless	1.3–1.8	≤1.5
Filtration rate	$m^3/m^2 \cdot min$	0.08–0.40	
Sand			
Depth	mm	900–1800	1200
Effective size	mm	2–3	2.5
Uniformity coefficient	unitless	1.2–1.6	≤1.5
Filtration rate	$m^3/m^2 \cdot min$	0.08–0.40	
Fuzzy filter			
Depth	mm	600–1080	800
Effective size	mm	25–30	28
Uniformity coefficient	unitless	1.1–1.2	1.1
Filtration rate	$m^3/m^2 \cdot min$	0.60–1.60	

[a] Adapted in part from Tchobanoglous (1988) and Tchobanoglous et al. (2003).

Note: $m^3/m^2 \cdot min \times 24.5424 = gal/ft^2 \cdot min$.

Table 11–16

Typical design data for dual- and multi-media depth filters[a]

Characteristic	Value[b]		
	Unit	Range	Typical
Dual-media			
Anthracite ($\rho = 1.60$)			
Depth	mm	360–900	720
Effective size	mm	0.8–2.0	1.5
Uniformity coefficient	unitless	1.3–1.6	≤1.5
Sand ($\rho = 2.65$)			
Depth	mm	180–360	360
Effective size	mm	0.4–0.8	0.65
Uniformity coefficient	unitless	1.2–1.6	≤1.5
Filtration rate	$m^3/m^2 \cdot min$	0.08–0.40	0.20
Multi-media			
Anthracite (top layer of quad-media filter, $\rho = 1.60$)			
Depth	mm	240–600	480
Effective size	mm	1.3–2.0	1.6
Uniformity coefficient	unitless	1.3–1.6	≤1.5
Anthracite (second layer of quad-media filter, $\rho = 1.60$)			
Depth	mm	120–480	240
Effective size	mm	1.0–1.6	1.1
Uniformity coefficient	unitless	1.5–1.8	1.5
Anthracite (top layer of tri-media filter, $\rho = 1.60$)			
Depth	mm	240–600	480
Effective size	mm	1.0–2.0	1.4
Uniformity coefficient	unitless	1.4–1.8	≤1.5
Sand ($\rho = 2.65$)			
Depth	mm	240–480	300
Effective size	mm	0.4–0.8	0.5
Uniformity coefficient	unitless	1.3–1.8	≤1.5
Garnet ($\rho = 4.2$)			
Depth	mm	50–150	100
Effective size	mm	0.2–0.6	0.35
Uniformity coefficient	unitless	1.5–1.8	≤1.5
Filtration rate	$m^3/m^2 \cdot min$	0.08–0.40	0.20

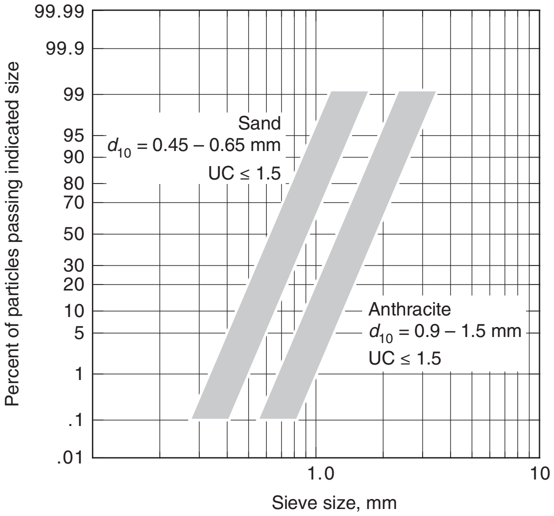
[a] Adapted from Tchobanoglous (1988) and Tchobanoglous et al. (2003).

[b] Anthracite, sand, and garnet sizes selected to limit the degree of intermixing. Use Eq. (11–28) for other values of density, ρ.

Note: $m^3/m^2 \cdot min \times 24.5424 = gal/ft^2 \cdot min$.

Figure 11–17

Typical particle size distribution ranges for sand and anthracite used in dual medium depth filters. Note that for sand the 10 percent size by weight corresponds approximately to the 50 percent size by count.



Selection of Filter Medium. Once the type of filter to be used has been selected, the next step is to specify the characteristics of the filter medium, or media, if more than one is used. Typically, this process involves the selection of the grain size as specified by the effective size, d_{10}, uniformity coefficient, UC, the 90 percent size, the specific gravity, solubility, hardness, and depth of the various materials used in the filter bed. Typical particle size distribution ranges for sand and anthracite filtering material are shown on Fig. 11–17. The 90 percent size designated, d_{90}, as read from a grain size analysis is used commonly to determine the required backwash rate for depth filters. The physical properties of filter materials used in depth filters are summarized in Table 11–17.

To avoid extensive intermixing of the individual mediums in multi-media filter beds, the settling rate of the filter media comprising the dual- and multi-media filters must have essentially the same settling velocity. Some intermixing is unavoidable, and the degree of intermixing in the dual- and multi-media beds depends on the density and size differences of the various media. The following relationship can be used to establish the appropriate sizes (Kawamura, 2000).

$$\frac{d_1}{d_2} = \left(\frac{\rho_2 - \rho_w}{\rho_1 - \rho_w}\right)^{0.667} \tag{11–28}$$

where d_1, d_2 = effective size of filter medium
ρ_1, ρ_2 = density of filter medium
ρ_w = density of water

The application of Eq. (11–28) is illustrated in Example 11–4.

Table 11–17

Typical properties of filter materials used in depth filtration[a]

Filter material	Specific gravity	Porosity, α	Sphericity
Anthracite	1.4–1.75	0.56–0.60	
Sand	2.55–2.65	0.40–0.46	0.75–0.85
Garnet	3.8–4.3	0.42–0.55	0.75–0.85
Ilmenite	4.5	0.40–0.5	
Fuzzy filter medium		0.87–0.89	

[a] Adapted in part from Cleasby and Logsdon (1999).

EXAMPLE 11–4 **Determination of Filter Medium Sizes** A dual media filter bed comprised of sand and anthracite is to be used for the filtration of settled secondary effluent. If the effective size of the sand in the dual medium filter is to be 0.55 mm, determine the effective size of the anthracite to avoid significant intermixing.

Solution

1. Summarize the properties of the filter media
    a. For sand
        i. Effective size = 0.55 mm
        ii. Specific gravity = 2.65 (see Table 11–17)
    b. For anthracite
        i. Effective size = to be determined, mm
        ii. Specific gravity = 1.7 (see Table 11–17)
2. Compute the effective size of the anthracite using Eq. (11–28)

$$d_1 = d_2\left(\frac{\rho_2 - \rho_w}{\rho_1 - \rho_w}\right)^{0.667}$$

$$d_1 = 0.55\text{ mm}\left(\frac{2.65 - 1}{1.7 - 1}\right)^{0.667}$$

$$d_1 = 0.97\text{ mm}$$

Comment Another approach that can be used to assess whether intermixing will occur is to compare the fluidized bulk densities of the two adjacent layers (e.g., upper 450 mm sand and lower 100 mm of anthracite).

Filter Flowrate Control. The principal methods now used to control the rate of flow through downflow gravity filters may be classified as (1) constant-rate filtration with fixed head, (2) constant rate filtration with variable head, and (3) variable-declining-rate filtration. A variety of other control methods are also in use (Cleasby and Logsdon, 1999; Kawumura, 2000).

Constant Rate Filtration with Fixed Head. In constant-rate filtration with fixed head [see Fig. 11–18(a)], the flow through the filter is maintained at a constant rate. Constant-rate filtration systems are either influent controlled or effluent controlled. Pumps or weirs are used for influent control whereas an effluent modulating valve that can be operated manually or mechanically is used for effluent control. In effluent control systems, at the beginning of the run, a large portion of the available driving force is dissipated at the valve, which is almost closed. The valve is opened as the headloss builds up within the filter during the run. Because the required control valves are expensive and because they have malfunctioned on a number of occasions, alternative methods of flowrate control involving pumps and weirs have been developed and are coming into wider use.

Constant Rate Filtration with Variable Head. In constant-rate variable head filtration head [see Fig. 11–18(b)], the flow through the filter is maintained at a constant rate. Pumps or weirs are used for influent control. When the head or effluent turbidity reaches a preset value, the filter is backwashed.

Variable Rate Filtration with Fixed or Variable Head. In variable-declining-rate filtration [see Fig. 11–18(c)], the rate of flow through the filter is allowed to decline as the

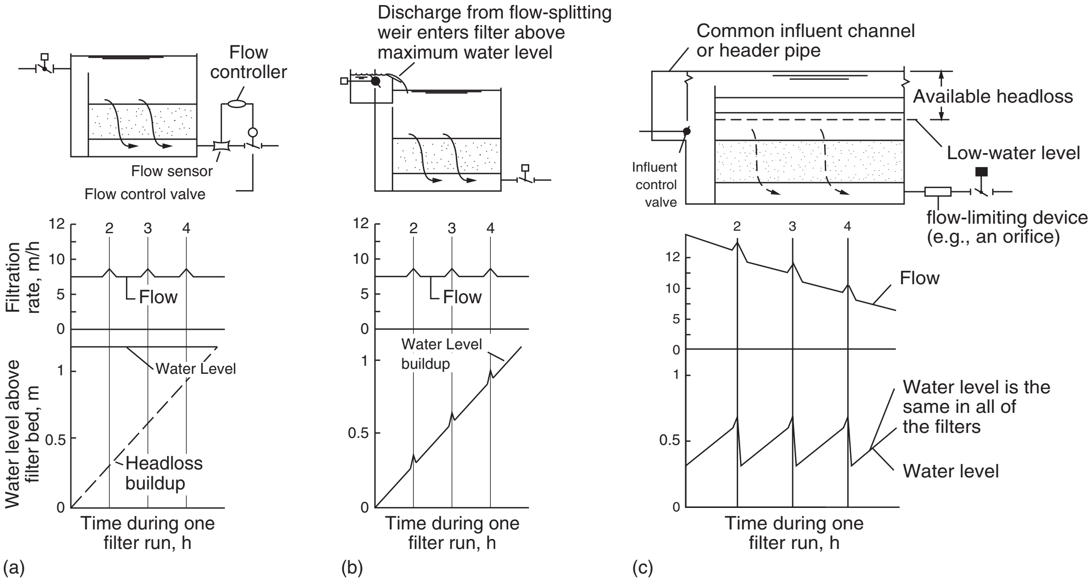


Figure 11–18

Definition sketch for filter operation: (a) fixed head, (b) variable head, and (c) variable-flow variable-head. Curves for filters in (a), (b), and (c) are for the operation of one filter in a bank of four filters. The numbers represent the filter that is backwashing during the filter run. In practice, the time before backwashing will not be the same for all of the filters. (Adapted from Tchobanoglous and Schroeder, 1985.)

rate of headloss builds up with time. Declining-rate filtration systems are either influent controlled or effluent controlled. When the rate of flow is reduced to the minimum design rate, the filter is removed from service and backwashed.

Filter Backwashing Systems. Methods commonly used for backwashing granular-medium filter beds operated in the semi-continuous mode include (1) water backwash only, (2) water backwash with auxiliary surface water-wash agitation, (3) water backwash with auxiliary air scour, and (4) combined air-water backwashing. With the first three methods, fluidization of the granular medium is necessary to achieve effective cleaning of the filter bed at the end of the run. With the fourth method, fluidization is not necessary.

Water Backwash Only. In the past, the most common method used to clean a filter of accumulated material was to backwash it with filtered water. Based on experimental studies, it has been found that the optimum cleaning of a conventional filter bed occurs when the expanded porosity of the bed is in the range of 0.65 to 0.70 (Amirtharajah, 1978). At this degree of expansion, it has been found that the shearing action of the rising backwash water and particle abrasion is most effective in removing the accumulated material from the filtering medium. Approximate backwash water flowrates required to fluidize various filter beds are reported in Table 11–18. To reduce the potential for the formation of mud balls and to enhance the removal of accumulated material either surface washers or air scour, as described below, are now used in conjunction with the water backwash.

Water Backwash with Auxiliary Surface Wash. Surface washers (see Fig. 11–19) are often used to provide the shearing force required to clean the grains of the filtering medium

Table 11–18

Approximate backwash water flowrates required to fluidize various filter beds at 20°C

Filter media	Effective size, mm	Minimum backwash velocity needed to fluidize bed[a] gal/ft²·min	Minimum backwash velocity needed to fluidize bed[a] m³/m²·min
Sand	1	24–27	1.0–1.2
Sand	2	44–50	1.8–2.0
Anthracite	1.7	22–24	0.9–1.0
Anthracite and sand	1.5 (A) and 0.65 (S)	15–30	0.8–1.2
Anthracite, sand, and garnet	1.4 (A), 0.5 (S), and 0.35 (G)	15–30	0.6–1.2
Fuzzy filter	28–30	10–15	0.4–0.6

[a] Varies with size, shape, and specific gravity of the medium and the temperature of the backwash water.

Note: $m^3/m^2 \cdot min \times 24.5424 = gal/ft^2 \cdot min$.

used for wastewater filtration. Surface washers for filters can be fixed or mounted on rotary sweeps. According to data on a number of systems, rotary sweep washers appear to be the most effective. Operationally, the surface washing cycle is started about 1 or 2 min before the water backwashing cycle is started. Both cycles are continued for about 2 min, at which time the surface wash is terminated. Water usage for a single-sweep surface backwashing system varies from 0.02 to 0.04 $m^3/m^2 \cdot min$ (0.5 to 1.0 $gal/ft^2 \cdot min$) and from 0.06 to 0.08 $m^3/m^2 \cdot min$ (1.5 to 2.0 $gal/ft^2 \cdot min$) for a dual-sweep surface backwashing system.

Water Backwash with Auxiliary Air Scour. The use of air to scour the filter provides a more vigorous washing action than water alone. Operationally, the water level above the filter bed is lowered to within 150 mm (6 in.) of the top of the media and air is usually applied for 3 to 4 min before the low-rate water backwashing cycle begins. In some

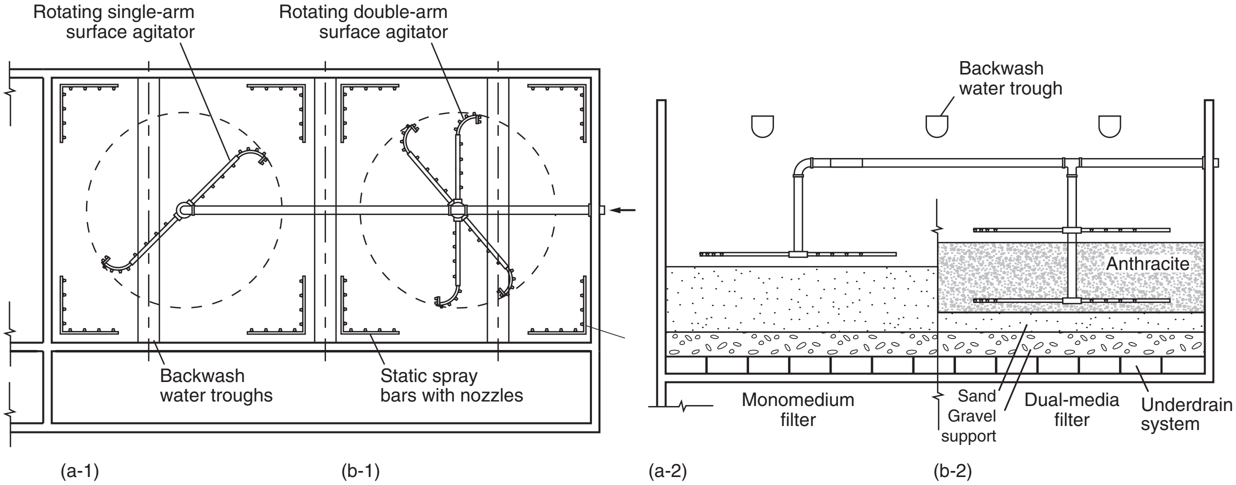


Figure 11–19

Surface washing facilities used to clean conventional granular medium filters: (a-1) and (a-2) single arm washer for a mono-medium filter and (b-1) and (b-2) a dual arm washer for a dual-medium filter.

Table 11–19

Approximate air and water flowrates used to backwash various filter beds at 20°C

	Medium characteristics			Backwash rates, $m^3/m^2·min$	
Filter media	Effective size, mm	Uniformity coefficient	Backwash sequence	Air	Water
Sand	1	1.4	1st–air + water	0.8–1.3	0.25–0.3
			2nd–water		0.5–0.6
Sand	2	1.4	1st–air + water	1.8–2.4	0.4–0.6
			2nd–water		0.8–1.2
Anthracite	1.7	1.4	1st–air + water	1.0–1.5	0.35–0.5
			2nd–water		0.6–0.8
Sand and	0.65(S)	1.4	1st–air	0.8–1.6	
anthracite[b]	1.5(A)	1.4	2nd–air + water	0.8–1.6	0.3–0.5
			3rd–water		0.6–0.9

[a] Adapted in part from Dehab and Young (1977) and Cleasby and Logsdon (2000).

[b] Dual medium filter bed is fluidized.

Note: $m^3/m^2·min \times 24.5424 = gal/ft^2·min$

$m^3/m^2·min \times 3.2808 = ft^3/ft^2·min$.

systems, air is also injected during the first part of the low-rate water-washing cycle, referred to as combined or concurrent air scour (see below). Typical air flowrates range from 0.9 to 1.6 $m^3/ m^2·min$ (3 to 5 $ft^3/ft^2·min$). A typical operating sequence for a dual media filter is given in Table 11–19. Also, as noted in Table 11–19, a water wash at the end of the air/water backwash cycle is used at the end to purge the filter of any residual air which could cause air binding. The reduced washwater requirements for the air-water backwash system can be appreciated by comparing the values given in Table 11–18 with those given in Table 11–19.

Combined Air-Water Backwash. The combined air-water backwash system is used in conjunction with the single-medium unstratified filter bed. Operationally, air and water are applied simultaneously for several minutes. The specific duration of the combined backwash varies with the design of the filter bed. Ideally, during the backwash operation, the filter bed should be agitated sufficiently so that the grains of the filter medium move in a circular pattern from the top to the bottom of the filter as the air and water rise up through the bed. Some typical data on the quantity of water and air required are reported in Table 11–19. At the end of the combined air-water backwash, a 2- to 3-min water a low-rate backwash at sub-fluidization velocities [typically 0.2 $m^3/ m^2·min$ (5 $gal/ft^2·min$)] is used to remove any air bubbles that may remain in the filter bed. This step is required to eliminate the possibility of air binding within the filter. Normally, a high-rate washing step follows the low-rate washing. High-rate washing is typically conducted at [$0.6 - 0.8$ $m^3/m^2·min$ ($15-20$ $gal/ft^2·min$)]. Air scour combined with high-rate washing is not conducted because of the possibility of excessive expansion of media resulting in media loss into the washwater troughs.

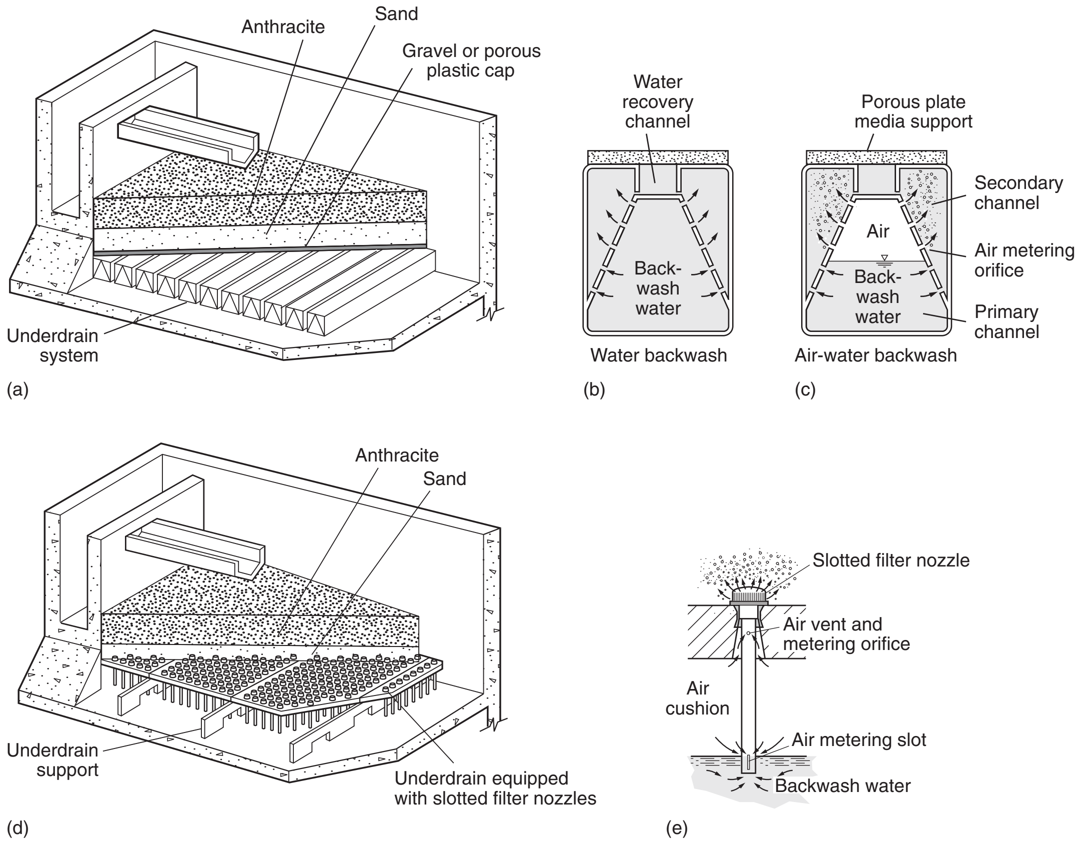


Figure 11–20

Typical underdrain systems used for granular media filters: (a) underdrain system used with gravel or porous plastic cap support for filter media, (b) underdrain system shown in (a) during water backwash, (c) underdrain system shown in (a) during air-water backwash, (d) underdrain system equipped with slotted air-water nozzles used without gravel support layer, and (e) air-water nozzle used in underdrain system shown in (d) without gravel support layer [(b) and (c) adapted from Leopold; (e) adapted from Infilco-Degremont, Inc.].

Filter Appurtenances. The principal filter appurtenances are as follows: (1) the underdrain system used to support the filtering materials, collect the filtered effluent, and distribute the backwash water and air (where used); (2) the washwater troughs used to remove the spent backwash water from the filter; and (3) the surface washing systems or air scour blower used to help remove attached material from the filter medium.

Underdrain Systems. The type of underdrain system to be used depends on the type of backwash system. In conventional water backwashed filters without air scour, it is common practice to place the filtering medium on a support consisting of several layers of graded gravel. The design of a gravel support for a granular medium is delineated in the AWWA Standard for Filtering Material B100–96 (AWWA, 1996). Typical underdrain systems are shown on Fig. 11–20. The gravel is not intended to aid in filtration. Rather, the purpose of the gravel is to prevent media from entering the underdrain. Air scour for filters with a gravel support system can be challenging, and would normally be comprised of the air scour piping grid laid on top of the gravel itself. Disruption of gravel support will

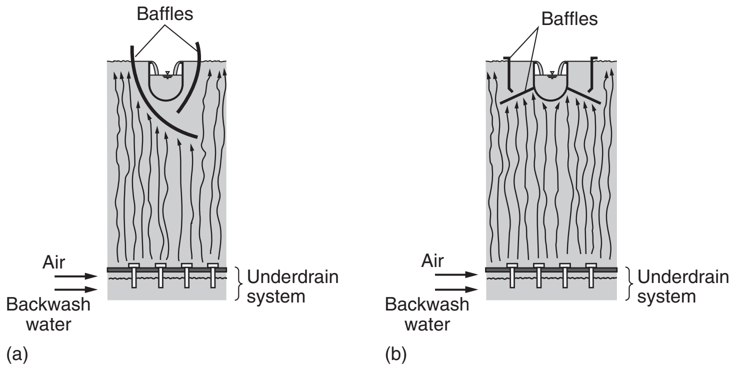


Figure 11–21

Details of baffle systems developed to minimize loss of filtering medium during backwash operation: (a) section through dual-baffle system and . (b) more elaborate baffle with two wings and side baffles.

lead to potential media passage to the underdrain or media plugging the underdrain apertures. As an alternative to gravel, newer filter underdrains are often equipped with a porous, HDPE direct media retention plate, typically 25mm (1 in.) thick and fastened to the top of each underdrain block. This arrangement precludes the need for gravel and therefore can allow for a deeper filter bed.

Washwater Troughs. Washwater troughs are constructed of fiberglass, plastic, or sheet metal or of concrete with adjustable weir plates. The particular design of the trough will depend to some extent on the other equipment to be used in the design and construction of the filter. Loss of filter material during backwashing is a common operating problem. To reduce this problem, baffles can be placed on the under side of the washwater troughs as shown on Fig. 11–21.

Operational Problems with Depth Filters. The principal operational problems encountered in wastewater filtration with depth filters are (1) turbidity breakthrough; (2) mudball formation; (3) buildup of emulsified grease; (4) development of cracks and contraction of the filter bed; (5) loss of filter medium or media, by mechanical and operational means; and (6) gravel mounding. Because these problems can affect both the performance and operation of a filter system, care should be taken in the design phase to provide the necessary facilities to minimize their impact. These issues are considered further in Table 11–20. Because of the inherent variability in the wastewater characteristics and their potential impact upon filter design and operation, the best way to ensure the filter configuration selected for a given application will function properly is to conduct pilot plant studies representative of the range of operating conditions.

11–6 SURFACE FILTRATION

Surface filtration, as shown in Table 11–4, involves the removal of particulate material suspended in a liquid by mechanical sieving by passing the liquid through a thin septum (i.e., filter material). The mechanical sieving action is similar to a kitchen colander. Membrane filters, microfiltration, and ultrafiltration, discussed in Sec. 11–6, are also surface filtration devices but are differentiated on the basis of the sizes of the pores in the filter medium. Surface filter mediums typically have openings in the size range from 5 to 30 μm or larger; in microfiltration and ultrafiltration, the pore size can vary from 0.05 to 2.0 μm for MF and 0.005 to 0.1 μm for UF.

Surface filtration has been used in several applications including (1) as a replacement for depth filtration to remove residual suspended solids from secondary effluents, (2) for the

Table 11–20

Summary of commonly encountered problems in depth filtration of wastewater and control measures for those problems

Problem	Description/control
Turbidity breakthrough[a]	Unacceptable levels of turbidity are recorded in the effluent from the filter, even though the terminal headloss has not been reached. To control the buildup of effluent turbidity levels, chemicals and polymers have been added to the filter. The point of chemical or polymer addition must be determined by testing.
Mudball formation	Mudballs are an agglomeration of biological floc, dirt, and the filtering medium or media. If the mudballs are not removed, they will grow into large masses that often sink into the filter bed and ultimately reduce the effectiveness of the filtering and backwashing operations. The formation of mudballs can be controlled by auxiliary washing processes such as air scour or water surface wash concurrent with, or followed by, water wash.
Buildup of emulsified grease	The buildup of emulsified grease within the filter bed increases the headloss and thus reduces the length of filter run. Both air scour and water surface wash systems help control the buildup of grease. In extreme cases, it may be necessary to steam clean the bed or to install a special washing system.
Development of cracks and contraction of filter bed	If the filter bed is not cleaned properly, the grains of the filter bed filtering medium become coated. As the filter compresses cracks develop, especially at the sidewalls of the filter. Ultimately, mudballs may develop. This problem can be controlled by adequately backwashing and scouring.
Loss of filter medium or media (mechanical)	In time, some of the filter material may be lost during backwashing and through the underdrain system (where the gravel support has been upset and the underdrain system has been installed improperly). The loss of the filter material can be minimized through the proper placement of washwater troughs and underdrain system. Special baffles have also proved effective.
Loss of filter medium or media (operational)	Depending on the characteristics of the biological floc, grains of the filter material can become attached to it, forming aggregates light enough to be floated away during the backwashing operations. The problem can be minimized by the addition of an auxiliary air and/or water scouring system.
Gravel mounding	Gravel mounding occurs when the various layers of the support gravel are disrupted by the application of excessive rates of flow during the backwashing operation. A gravel support with an additional 60 to 75 mm (2 to 3 in.) layer of high density material, such as ilmenite or garnet, can be used to overcome this problem.

[a] Turbidity breakthrough does not occur with filters that operate continuously.

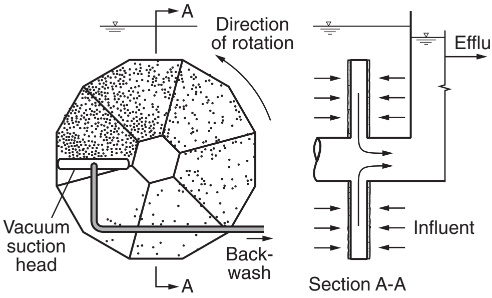
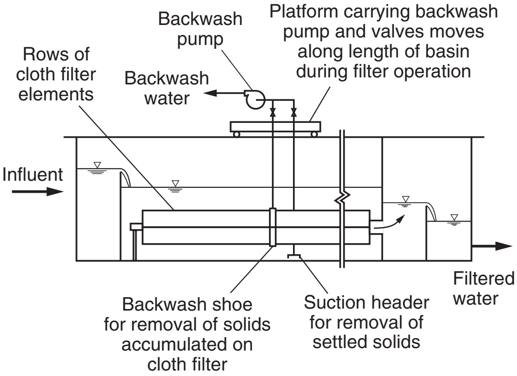
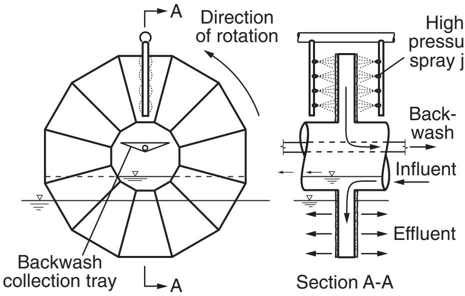
removal of suspended solids and algae from stabilization pond effluents, and (3) as a pretreatment operation before microfiltration or UV disinfection. Surface filtration is gaining in popularity because of the high quality effluent produced, smaller footprint, low backwash rates, and reduced maintenance requirements. Information on surface filtration technologies, their performance, and design considerations is presented and discussed in this section.

Available Filtration Technologies

The principal types of surface filtration devices are identified and described in Table 11–21. With the exception of the inclined surface and cartridge filters, all of the other surface filters have been used for the filtration of secondary effluent. Some of the surface filters have also been used for the filtration of algae for lagoon effluents. The inclined surface

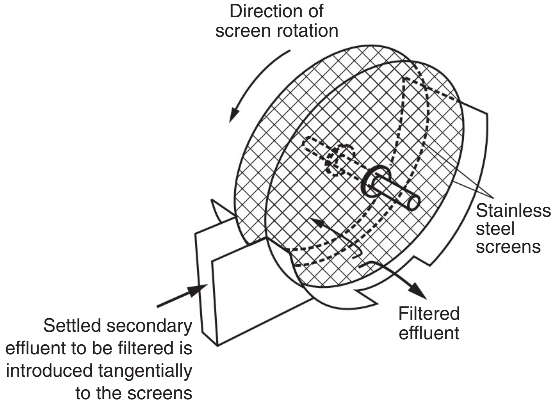
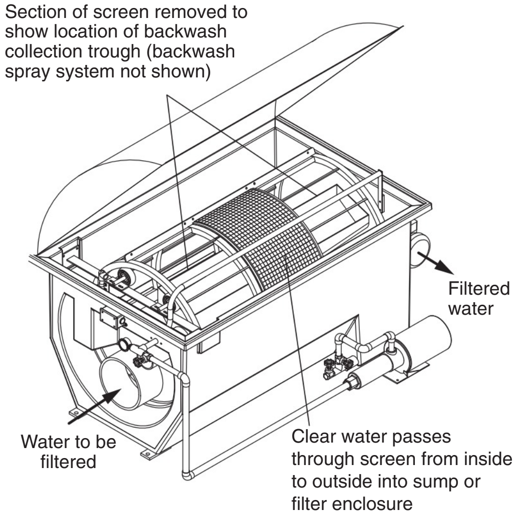
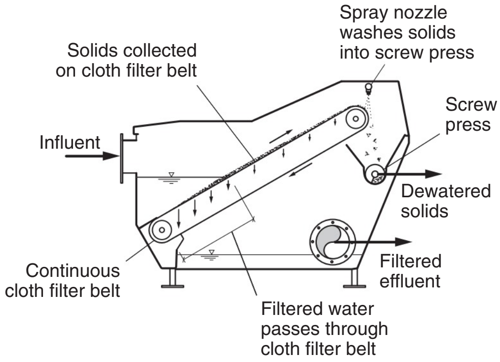
Table 11–21

Description of some surface filters used in effluent filtration applications

Type	Description
(a) Cloth Media Filter (CMF)  	The CMF, marketed under the trademark AquaDisk® by Aqua-Aerobic Systems, also consists of several disks mounted vertically in a tank. Each disk is comprised of six equal segments. Operationally, water flows by gravity from the exterior of the disks through the filter medium to an internal collection system. Typically, two types of filter cloth are used: (1) a needle felt cloth made of polyester or (2) synthetic pile fabric cloth. A vacuum system is. Vacuum suction heads, located on either side of the disk, are used to remove the accumulated solids by drawing filtrate water from the filtrate header back through the cloth media while the disk is rotating. Solids are removed when a predetermined increase in headloss has occurred.
(b) Diamond Cloth Media Filter (DCMF)  	The DCMF, marketed under the trademark AquaDisk® by Aqua-Aerobic Systems, consists of cloth filter elements, which have a diamond shaped cross section. The filter elements are cleaned by a vacuum sweep which moves back and forth along the length of the filter, when a predetermined increase in headloss has occurred. Solids that settle to the bottom of the reactor below the filter element are removed periodically by a vacuum header. Using a diamond shape for the filter, it is possible to increase the cloth filter surface area per unit of aerial surface area. Because higher volumes for filtered water can be produced per unit area, the DCFM is used in new installations and as a replacement for existing sand filters.
(c) Discfilter® (DF)  	The DF, developed by Hydrotech and marketed in the U.S. by Veolia Water Systems, consists of a series of disks comprised of two vertically mounted parallel disks that are used to support the filter cloth Each disk is connected to a central feed tube. The cloth screen material used can be of either polyester or Type 304 or 316 stainless steel. Accumulated solids are removed from the screen by high-pressure water jets. The filter mechanism can be furnished with a self-contained tank or for installation in a concrete tank. In cold climates or where odor control is a consideration, an enclosure can be provided for the disks.

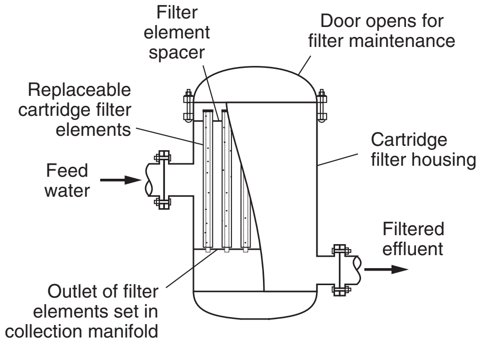
(*continued*)

Table 11–21 (*Continued*)

Type	Description
(d) Ultrascreen®  	The Ultrascreen® developed by Nova Water technologies that consists of two continuously rotating circular screens of woven stainless steel mesh. The liquid to be filtered is introduced between the two screens at right angles to the screens. The filtered effluent that flows through the screen flows out and is directed to a collection chamber below the screen. Unlike other disk type screens, the discharge from the screen is by gravity as there is no water on the discharge side of the screen. High- pressure water jets are used to clean the accumulated solids from the screens.
(e) Drum Filter (DF)[a]  	As the name implies the drum filter is in the shape of a drum. The liquid to be filtered is introduced on the inside of the drum and flows out thought the periphery of the drum, through a filter cloth of polyester or polypropylene or stainless steel, as the drum rotates slowly. When the water level within the drum rises to specified level, a backwash cycle is initiated to remove the accumulated solids. A high pressure water spray is used to dislodge and remove the accumulated solids as the drum rotates. The solids removed from the drum are collected in a collection trough on the inside of the drum. Drum filter can be installed in concrete, stainless steel, or fiberglass tankage. The range of pore openings for the filter cloth range from 10 μm to 1 mm, depending on the application.
(f) Inclined Cloth Media Screen  	Developed by M2 Renewables, the inclined screen is used for the filtration of untreated wastewater. As the moving screen rotates, solids are accumulated on the screen. When the screen exits the water pool, the accumulated solids are partially dewatered by the force of gravity. The accumulated solids are removed from the screen as it passes over the upper roller. High-pressure water jets can also be used. As noted in Chap. 5, the fact that the screen alters the particle size distribution of the solids to be treated and has a relatively small footprint is significant relative to conventional primary clarification.

(*continued*)

Table 11–21 (*Continued*)

Type	Description
(g) Cartridge Filter  	Most cartridge filters are usually polypropylene wound cartridges from 800 to 1000 mm in length housed inside a vertical or horizontal stainless steel or fiberglass vessel. They are employed in a number of different applications, typically to protect downstream applications. In advanced water treatment, they are employed to remove contaminates found in the chemicals added to control scaling in reverse osmosis membranes. Cartridge filters are not considered further in the discussion of surface filtration, but are included because they are used as a pretreatment step for reverse osmosis. Pleated cartridge filters are used almost exclusively to concentrate virus from treated wastewater for analysis.

[a] Courtesy of Xylem.

filter [see Table 11–21(f)], discussed in Sec. 5–9 in Chap. 5, is used for the filtration of untreated wastewater following coarse and intermediate screening. Cartridge filters [see Table 11–21(g)], are used for pretreatment prior to membrane filtration, particularly where reverse osmosis (RO) is used.

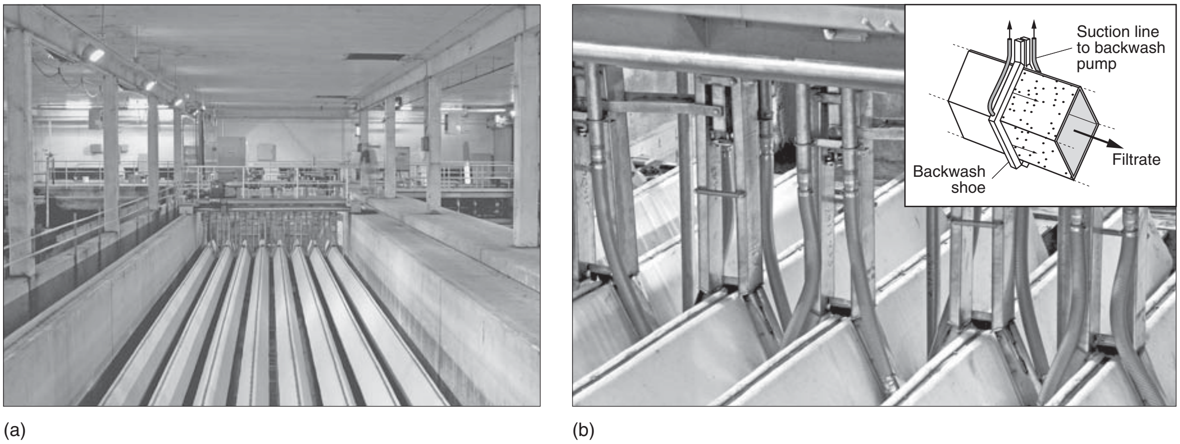
Description of the Surface Filtration Process

The key features of surface filters are (1) the filter configuration, (2) the filter medium, (3) the method used to introduce the liquid to be filtered, (4) the method used to clean the filtering medium, and (5) the impact of the accumulation of solids has on process performance.

Filter Configurations. Surface filters are available in a variety of configurations. The most common type of surface filter is comprised of a series of disks attached to a central shaft. The individual disks are made up of two filtering surfaces attached to a metal support frame as shown in Table 11–21 (a) and (c). The diamond cloth-media filter (DCMF), a relatively recent development, is shown schematically in Table 11–21(b) and pictorially on Fig. 11–22. The drum filter for effluent filtration [see Table 11–21(e)] is a relatively recent development, although drum filters are used in a number of filtering applications. Operation of the different surface filters depends on how the liquid to be filtered is applied to the filtering medium and how the filtering material is cleaned of accumulated material.

Filter Materials. The filter material used in surface filters can be categorized as two dimensional and three dimensional. Two dimensional mediums are typically made of synthetic fabrics of different weaves and woven metal fabrics (most commonly stainless steel). The most common weave for synthetic materials is known as a plain weave, which is similar to broadcloth. Stainless steel weaves can include plain weave, twilled weave, and Dutch weave wire meshes. Three dimensional filtering mediums include polyester needle felt cloth and synthetic pile fabric cloth.

Flow Path for Liquid to be Filtered. The flow path can also be used to classify surface filters. Basically, two methods are used to apply the liquid to be filtered to the



(a) (b)

Figure 11–22

Views of Diamond Cloth Media Filter: (a) view of filters installed in an existing sand filter basin and (b) view of backwash shoe.

filtering medium. In the first, the liquid enters the feed tank and flows from the outside through the filter cloth into a receiving area (out-in) [see Table 11–21(a), (b) and (f)]. In the second method, the liquid to be filtered is introduced into the annular volume between the two filtering surfaces and flows outward through the filtering medium into the collection vessel (in-out) [see Table 11–21 (c), (d), and (e)]. In either case, solids accumulate on the surface in the direction of flow. The direction of flow affects the method to be used for the removal of the accumulated material, the submergence (i.e. the active filter area), and the overall depth of the unit.

Cleaning the Filter Medium. Two types of methods are used to remove the accumulated material removed from the fluid: (1) vacuum removal and (2) intermittent and/or continuous high-pressure spray washing. The vacuum removal system is used for surface filters where the flow is from the outside in whereas the high-pressure water spray nozzles are used where the flow is from the inside to the outside.

Vacuum Removal. When the headloss through the CMF reaches a predetermined set point, the disks are cleaned. As the disks rotate, solids are removed from both sides of the disk by liquid vacuum suction heads, located on either side of each disk, which draws filtered water from the filtrate header back through the cloth media while the disk is rotating. This reversal of flow removes particles that have become entrapped on the surface and within the cloth medium. The diamond cloth filter is also cleaned by a vacuum sweep which moves back and forth along the length of the filter. Solids that settle to the bottom of the reactor below the filter element are removed periodically by a vacuum header.

Over time, particles will accumulate in the cloth medium that can not be removed by a typical backwash. This accumulation of particles leads to increased headloss across the filter, an increase in the backwash suction pressure, and shorter run times between backwashes. When the backwash suction pressure or operating time reaches predetermined setpoints, a high pressure spray wash is initiated automatically. The high pressure spray wash flushes the particles that have become lodged inside the cloth filter media in 2 rev of the disk. The time interval between high pressure spray washes is a function of the feed water quality.

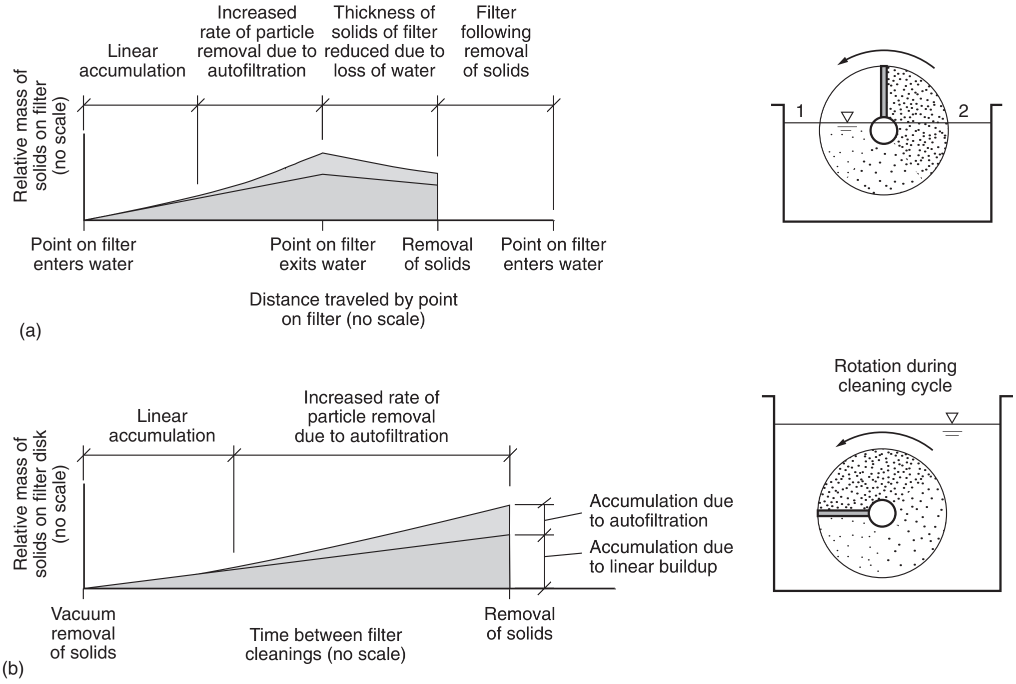


Figure 11–23

Generalized schematic of the operation of disc type surface filters based on following a point with time on the filter cloth between cleaning events, including the removal of particles by size exclusion and finer particles due to autofiltration: (a) for surface filters that are partially submerged and (b) for submerged filter with vacuum removal of accumulated solids.

High-Pressure Water Sprays. In surface filters where the flow is from inside to outside, high-pressure water sprays are used to remove the material that accumulates on the inside of the disk. Most high-pressure water spray wash systems can operate in either an intermittent or continuous backwash mode. In the intermittent mode, high-pressure backwash spray jets are activated only when headloss through the filter reaches a preset level or time. Once activated, washwater is sprayed through the filter material from the outside as the disk rotates. The accumulated solids that are dislodged fall into a collection trough. When operating in a continuous backwash mode, the production of filtered water and backwashed occur simultaneously. The location and configuration of the high-pressure spray nozzles and solids collection trough are manufacturer specific.

Impact of the Accumulation of Solids on Process Performance. The removal of particulate matter with surface filters is illustrated schematically on Fig. 11–23 for partially and completely submerged surface filters. For partially submerged filters [see Fig. 11–23(a)], the accumulation of solids on the filter medium occurs between point 1, where the clean filter comes in contact with the fluid to be filtered, and point 2, where the filter leaves the fluid. For the completely submerged filter [see Fig. 11–23(b)], the accumulation of solids occurs with time until the backwash headloss is reached, at which time the filter is cleaned. In both cases, the accumulated material on the surface of the filter begins to act as a filter. The filtering action of the accumulated material is known as *autofiltration*. Autofiltration can be used to explain why material of a smaller size than the pore size of the filter medium can be removed by surface filtration. For both types of surface filters, the onset of autofiltration, and the degree of additional removal that can be achieved, will depend on the mesh size of the filtering medium, the characteristics of the wastewater, and the filtration rate.

Performance of Surface Filters

In investigations of surface filters in comparison to granular medium filters in filtering secondary effluent (Riess et al., 2001 and Olivier et al., 2003), it has been observed that surface filters performed as well or better than granular medium filters in removing turbidity and the number and size of particles, with increased surface loading and reduced backwash water requirements. As with depth filters, insight into the operation of surface filters can be gained from a review of performance data with respect to the (1) hydraulic loading rate, (2) removal of turbidity and total suspended solids, (3) variability of turbidity and TSS removal, (4) removal of different particle sizes, (5) removal of microorganism, and (6) backwash water requirements.

Hydraulic Loading Rate. As with depth filters, there is a considerable difference in the HLRs over which surface filters operate. Typical HLRs for surface filters are reported in Table 11–22. For example, while two surface filters can produce a

Table 11–22

Comparison of operational characteristics for selected surface filters

Parameter	Unit	Cloth media filter®	Diamond cloth media filter®	Diskfilter®	Ultrascreen®	Drumfilter®
Typical hydraulic loading rate	$m^3/m^2 \cdot min$	0.08–0.20	0.08–0.20	0.08–0.20	0.20–0.65	0.08–0.26
	$gal/ft^2 \cdot min$	2–5	2–5	2–5	5–16	2–6.5
Peak HLR	$m^3/m^2 \cdot min$	0.26	0.26	0.24	0.65	0.26
	$gal/ft^2 \cdot min$	6.5	6.5	6	16	6.5
CDPH[a] allowable average HLR	$m^3/m^2 \cdot min$	—	—	—	0.32	—
	$gal/ft^2 \cdot min$	—	—	—	8	—
CDPH allowable peak HLR	$m^3/m^2 \cdot min$	0.24	0.24	0.24	0.65	—
	$gal/ft^2 \cdot min$	6	6	6	16	—
Influent TSS concentration	mg/L	5–20	5–20	5–20	5–20	5–20
Filter material	Type	Nylon and/or Polyester	Nylon and/or Polyester	Polyester or stainless steel	Stainless steel	Polyester or stainless steel
Nominal pore size of screen	μm	5–10	5–10	10–40	10–20	10–40
Direction of flow		out–in	out–in	in–out	in–out	in–out
Submergence	%	100	100	60–70	45	60–70
Headloss	mm	50–300	50–300	75–300	650	300
Disk diameter	m	0.90 or 1.80	na	1.75–3.0	1.6	
Backwash requirement	% of throughput	2–5	2–5	2–4	2–4	2–4

[a] CDPH = California Department of Public Health.

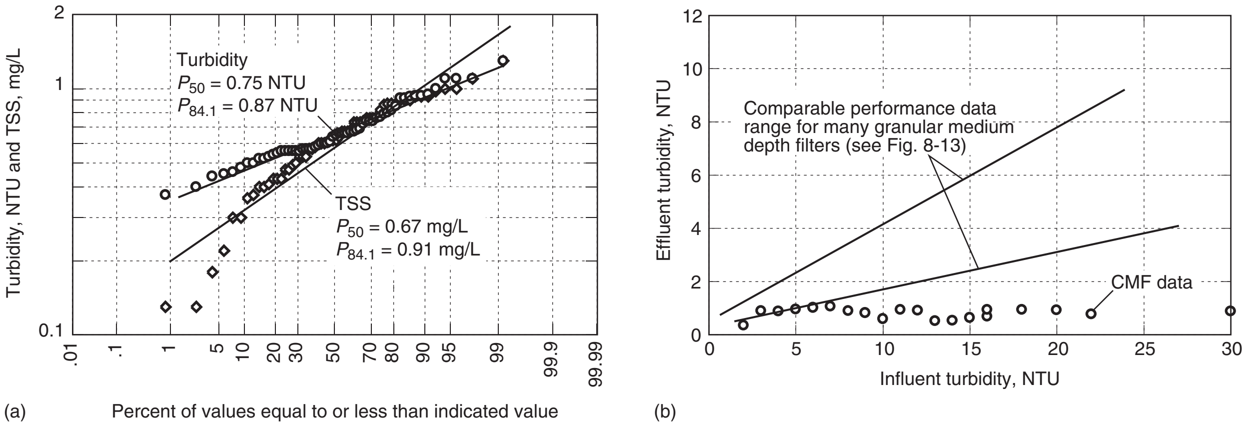


Figure 11–24

Performance data for cloth-media filter for secondary effluent: (a) effluent turbidity as a function of influent turbidity at a filtration rated of 176 L/min·m^2 and (b) effluent probability distributions for turbidity and TSS.

filtered effluent with turbidity of 2 NTU or less filtering the same effluent, the HLRs can vary by more that a factor of four or five. As with depth filters, the HLR and the backwash water requirements for surface filters will impact cost and the carbon footprint significantly.

Removal of Turbidity and Total Suspended Solids. To evaluate performance capabilities of surface filtration, a CMF pilot plant was tested using secondary effluent from an extended aeration activated sludge process with a solids retention time greater than 15 d. Effluent TSS and turbidity values from the activated sludge process ranged from 3.9 to 30 mg/L and 2 to 30 NTU, respectively. Based on a long-term study, it was found, as shown on Fig. 11–24(a), that both TSS and turbidity values of the filtered effluent were less than 1, 92 percent of the time (Riess et al., 2001). The performance of the CMF as compared to depth filters, all tested with the same activated sludge effluent, is shown on Fig. 11–24(b). As shown, the effluent turbidity from the CMF remained constant over a range of influent turbidity values that tested up to 30 NTU. The degree of removal of TSS from settled activated sludge effluent with surface filters, as with depth filters, will depend on the SRT at which the activated sludge process is operated. Similar results have been reported for other surface filtration technologies.

Variability in Turbidity and Total Suspended Solids Removal. The performance variability of surface filters is of critical importance where specific effluent turbidity limits must be met consistently. The variability observed in the operating data for surface filters as reported in Table 11–12, presented previously, is similar to that observed for depth filters. However it should be noted that average turbidity and TSS values tend to be lower.

Removal of Different Particle Sizes. In comparative testing with a granular medium filter, the surface filter consistently out-performed the granular medium filter in respect to particle removal (see Fig. 11–25). The particle size reduction also had a

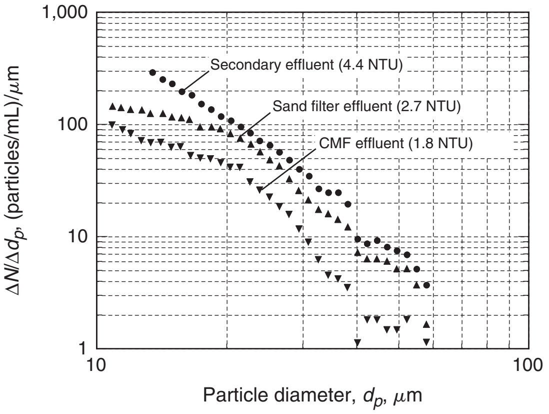


Figure 11–25

Comparison of particle sizes in effluent from secondary treatment, granular medium filter, and cloth media filter (Olivier et al., 2003).

significant impact on the inactivation of total coliform bacteria when used with UV disinfection (Olivier et al., 2002, 2003).

Removal of Microorganisms. Where chemicals are not used, the removal of coliform bacteria and viruses from biologically treated secondary effluent is on the order of 0 to 1.0 and 0 to 0.5 logs, respectively, similar to the values observed for depth filters.

Backwash Water Requirements. The amount of backwash water needed to clean surface filters, expressed as a percentage of plant flow, will depend on the characteristics of filter influent and the design of the surface filter. Typical backwash water percentages for surface filters as shown in Table 11–22 will vary, from 1 to 4 percent.

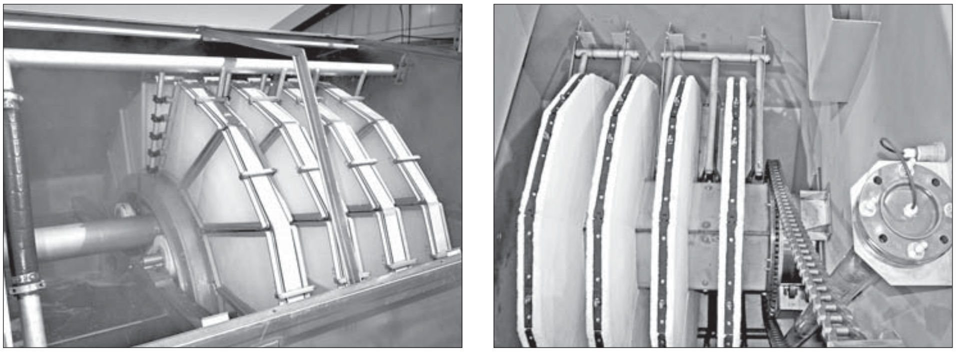
Design Considerations

Pilot studies are recommended in developing design and operating parameters for new installations. Useful data for design includes (1) the variability of the characteristics of the feed water to be treated and (2) the amount of backwash water required for normal operation. The backwash water requirements are a function of the TSS in the feed water and the solids loading on the filters. If the secondary treatment system is effective in TSS removal, the volume of backwash water can be reduced substantially.

Because cloth-media surface filtration is a relatively new technology, little long-term data are available on the life of the filter cloth. Where surface filtration is being considered, performance should be evaluated from operating installations using a similar type of cloth medium. One operating advantage cited for cloth-media filters is that the filter cloth can be removed and washed in a heavy-duty washing machine.

Pilot Plant Studies

As with granular media filtration, discussed above, there is no generalized approach to the design of full-scale filters for the treatment of wastewater. The discussion presented in the previous section on pilot plant testing also applies to the cloth filter. Typical cloth filter test facilities are illustrated on Fig. 11–26. It should be noted that the single disk shown on Fig. 11–26(b) is full sized. In a larger installation, a number of disks would be arranged on the center shaft.



(a) (b)

Figure 11–26

View of cloth filter pilot test filters. It should be noted that the cloth filters shown are full size.

11–7 MEMBRANE FILTRATION PROCESSES

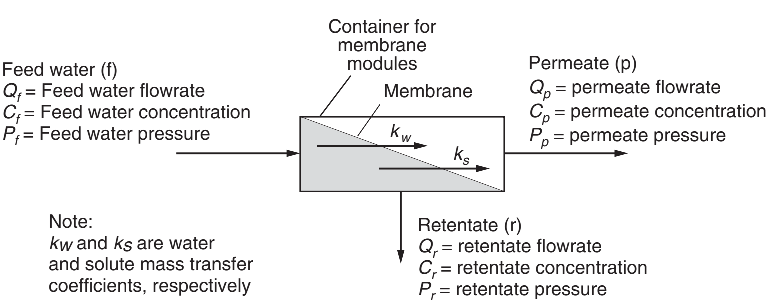
Filtration, as defined in Sections 11–3 and 11–5, involves the separation (removal) of particulate and colloidal matter from a liquid. In membrane filtration the range of particle sizes is extended to include dissolved constituents (typically from 0.0001 to 1.0 μm). The role of the membrane, as shown in Table 11–4, is to serve as a selective barrier that will allow the passage of certain constituents and will retain other constituents found in the liquid. To introduce membrane technologies and their application, the following subjects are considered in this section: (1) membrane process terminology, (2) membrane classification, (3) membrane configurations, (4) application of membrane technologies, and (5) the need for pilot-plant studies. The disposal of concentrated waste streams is considered at the end of this section. Electrodialysis, also a membrane process, used typically for the removal of dissolved constituents, is considered separately in Sec 11–7 following the discussion of the application of pressure driven membranes.

Membrane Process Terminology

Terms used commonly in the membrane technology field include: *feed water*, *permeate*, and *retentate*. These terms are illustrated on Fig. 11–27. The influent water to supplied to the membrane system for treatment is known as the *feed water*. The liquid that has passed through the membrane is known as the *permeate*. The portion of the feed water that does not pass through the membrane is known as the *retentate* (also referred to as concentrate, reject, or waste stream). Flux, the rate at which permeate flows through the membrane expressed as L/m^2·h or L/m^2·d, is the principal measure of membrane performance. Flux is synonymous with the concept of filter hydraulic loading rate presented in the discussion of depth and surface filtration.

Figure 11–27

Definition sketch for operation of a membrane process.



Membrane Process Classification

Membrane processes include microfiltration (MF), ultrafiltration (UF), nanofiltration (NF), reverse osmosis (RO), and electrodialysis (ED). Membrane processes can be classified in a number of different ways including (1) membrane configuration, (2) the type of material from which the membrane is made, (3) the nature of the driving force, (4) the separation mechanism, and (5) the nominal size of the separation achieved. Each of these methods of classifying membrane processes is considered in the following discussion. The general characteristics of membrane processes including typical operating ranges are reported in Table 11–23. The focus of the following discussion is on pressure driven membrane processes used for the removal of residual TSS, colloidal matter and dissolved solids. Pressure driven membranes are further defined as "low pressure," which includes MF and UF, and "high pressure," which includes NF and RO.

Membrane Configurations. In the membrane field, the term *module* is used to describe a complete unit comprised of the membrane elements (or modules), the pressure support structure for the membranes, the feed inlet and outlet permeate and retentate ports, and an overall support structure. The principal types of membrane modules used for wastewater treatment are (1) tubular, (2) hollow fine-fiber, and (3) spiral wound. Plate and frame and pleated cartridge filters are also available, but are used more commonly in industrial applications.

Definition sketches for the various membranes and detailed descriptions are presented in Table 11–24. There are two basic flow patterns with membranes: (1) outside-in [see Fig. 11–28(a)] and (2) inside-out [see Fig. 11–28(b)]. In most wastewater treatment applications where hollow fiber and membrane sheets are used, the flow is pattern is outside-in. With an outside-in flow pattern, the membrane can be backwashed with air, water, or a combination of both. The outside-in flow pattern is also used for feed water solutions with higher TSS and turbidities.

Membrane Materials. Most commercial membranes are produced as tubular, fine hollow fibers, or flat sheets. In general, three types of membranes are produced: symmetric, asymmetric and thin film composite (TFC) (see Fig. 11–29). As shown on Fig. 11–29(a) and (b), symmetric membranes are the same throughout. Symmetric membranes can vary from microporous to nonporous (so called dense). Asymmetric membranes [see Fig. 11–29(c)] are cast in one process and consist of a very thin (less than 1 μm) layer and a thicker (up to 100 μm) porous layer that adds support and is capable of high water flux.

Thin-film composite membranes [see Fig. 11–29(d)] are made by bonding a thin cellulose acetate, polyamide, or other active layer (typically 0.15 to 0.25 μm thick) to a thicker porous substrate, which provides stability. As reported in Table 11–23 membranes can be made from a number of different organic and inorganic materials. The membranes used for wastewater treatment are typically organic, although some ceramic membranes have been used. The choice of membrane and system configuration is based on minimizing membrane clogging and deterioration, typically based on pilot plant studies.

Driving Force. The distinguishing characteristic of the first four membrane processes considered in Table 11–23 (MF, UF, NF, and RO) is the application of hydraulic pressure, or vacuum, to bring about the desired separation. Electrodialysis involves the use of an electromotive force and ion selective membranes to accomplish the separation of charged ionic species.

Table 11–23

General characteristics of membrane processes

Membrane process	Membrane driving force	Typical separation mechanism	Typical pore size, μm	Typical operating range, μm	Membrane details	
					Materials (arranged alphabetically)	**Configuration**
Microfiltration	Hydrostatic pressure difference or vacuum in open vessels	Sieve	Macropores (> 50 nm)	0.07–2.0	Acrylonitrile, ceramic (various materials), polypropylene (PP), polysulfone (PS), polytetrafluorethylene (PTFE), polyvinylidene fluoride (PVDF), nylon	Spiral wound, hollow fiber, plate and frame
Ultrafiltration	Hydrostatic pressure difference or vacuum in open vessels	Sieve	Mesopores (2–50 nm)	0.008–0.2	Aromatic polyamides, ceramic (various materials) cellulose acetate (CA), polypropylene (PP), polysulfone (PS), polyvinylidene fluoride (PVDF), Teflon	Spiral wound, hollow fiber, plate and frame
Nanofiltration	Hydrostatic pressure difference in closed vessels	Sieve + solution/diffusion + exclusion	Micropores (< 2 nm)	0.0009–0.01	Cellulosic, aromatic polyamide, polysulfone (PS), polyvinylidene fluoride (PVDF), thin-film composite (TFC)	Spiral wound, hollow fiber, thin film composit
Reverse osmosis	Hydrostatic pressure difference in closed vessels	Solution/diffusion + exclusion	Dense (< 2 nm)	0.0001–0.002	Cellulosic, aromatic polyamide, thin-film composite (TFC)	Spiral wound, hollow fiber, thin film composite
Electrodialysis	Electromotive force	Ion exchange	Ion exchange	0.0003–0.002	Ion exchange resin cast as a sheet	Plate and frame

Table 11–24

Description of commonly used membrane types

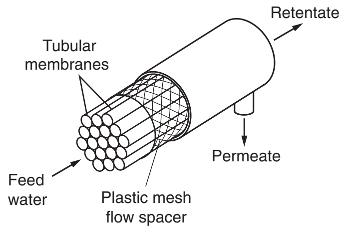
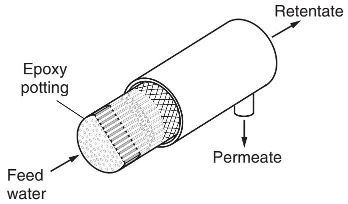
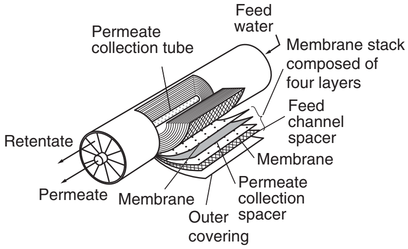
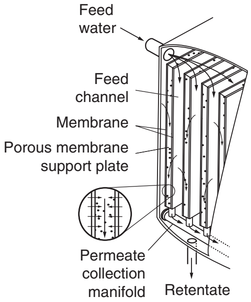
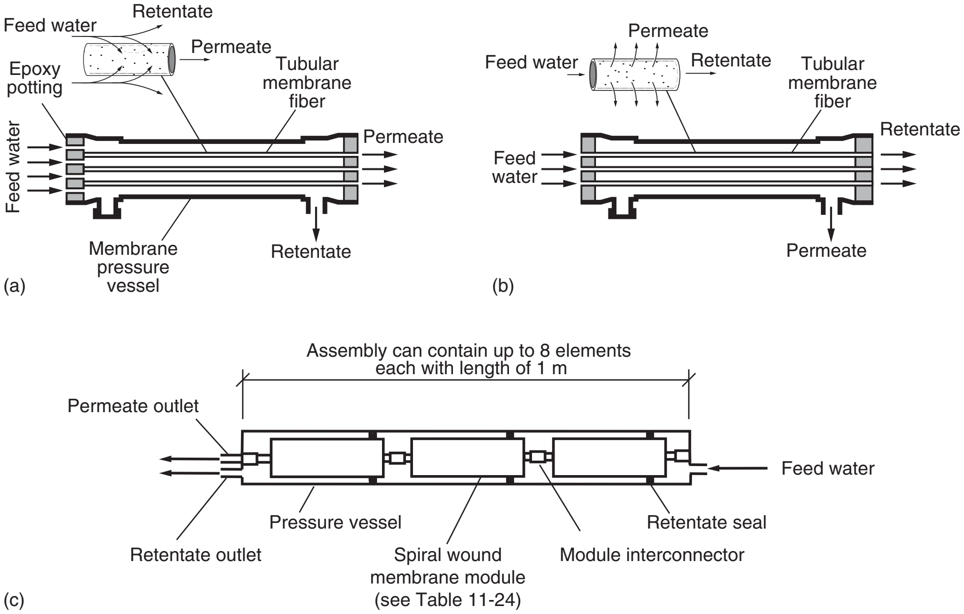
Type	Description
(a) Tubular 	In the tubular configuration the membrane is cast on the inside of a support tube. A number of tubes (either singly or in a bundle) is then placed in an appropriate pressure vessel. The feed water is pumped through the feed tube and product water is collected on the outside of the tubes. The retentate continues to flow through the feed tube. These units are used generally for water with high suspended solids or plugging potential. Tubular units are the easiest to clean, which is accomplished by circulating chemicals and pumping a "foamball" or "spongeball" through to mechanically wipe the membrane. Tubular units produce at a low product rate relative to their volume, and the membranes are generally expensive. The internal diameter of the Tubes will vary from 6 to 40 mm in diameter with lengths as long as 3.66 m (12 ft).
(b) Hollow fiber 	The hollow fiber membrane module consists of a bundle of hundreds to thousands of hollow fibers. The entire assembly is inserted into a pressure vessel. The feed can be applied to the inside of the fiber (inside-out flow) or the outside of the fiber (outside-in flow). Hollow fiber membrane modules are commonly used in membrane bioreactors (MBRs) as described in Chap. 7. Typical inside and outside diameters of the individual hollow fine fibers are about 35 to 45 and 90 to 100 μm, respectively. The typical length of a bundle of fibers is about 1.2 m (4 ft) long. A 100 mm (4 in.) diameter bundle may contain up to 650,000 individual fibers, although most contain fewer fibers per bundle. Fiber bundles can vary from 100 to 200 mm (4 to 8 in.) in diameter. Depending on the size of the bundle up to 7 bundles can be placed in a single pressure vessel [see Fig. 11–28(b)].
(c) Spiral wound 	In the spiral wound membrane, a flexible permeate spacer is placed between two flat membrane sheets. The membranes are sealed on three sides. The open side is attached to a perforated pipe. A flexible feed spacer is added and the flat sheets are rolled into a tight circular configuration. Thin film composites are used most commonly in spiral wound membrane modules. The term spiral derives from the fact that the flow in the rolled-up arrangement of membranes and support sheets follows a spiral flow pattern. The diameter of spiral wound elements will typically vary from 100 to 200 mm (4 to 8 in.), however, up to 300 mm (12 in.) diameter elements have been used. The active length of the membrane element is typically about 0.9 m (3 ft) between glue lines, although element lengths varying from 150 mm (6 in.) up to 1.5 m (5 ft) have been used. Operationally, from 2 to 6 membrane elements are included in a single pressure vessel [see Fig. 11–28(c)]. Six membrane elements are normally used for reverse osmosis. As an example, the membrane surface area in a pressure vessel containing four 100 mm (4 in.) diameter by 0.9 m (3 ft) long membrane elements is about 8.33 m^2 (90 ft^2).
(d) Plate and frame 	Plate and frame membrane modules are comprised of a series of flat membrane sheets and support plates. The water to be treated passes between the membranes of two adjacent membrane assemblies. The plate supports the membranes and provides a channel for the permeate to flow out of the unit. Typically, the dimensions of individual plates that comprise the plate and frame filter are about 20 by 40 mm (7.5 × 15 in.). The packing density of plate and frame units will vary between 100 to 400 m^2/m^3.

Figure 11–28

Definition sketch for types of membrane operation: (a) definition sketch for hollow fine-fiber membranes with flow from the outside to the inside of the fiber, (b) definition sketch for hollow fine fiber membranes with flow from the inside to the outside, and (c) spiral wound membranes in a containment vessel.



Removal Mechanisms. The separation of particles in MF and UF is accomplished primarily by straining (sieving, i.e., physical size exclusion), as shown on Fig. 11–30(a). In NF and RO, in addition to straining, small particles are rejected by the water layer adsorbed on the surface of the membrane which is known as a *dense* membrane [see Fig. 11–30(b)]. Some ionic species such as sodium (Na^+) and chloride (Cl^-) can be transported across the membrane by diffusion through the pores of the macromolecule comprising the membrane. Typically NF can be used to reject constituents as small as 0.001 μm whereas RO can reject particles as small as 0.0001 μm.

Size of Separation. The pore sizes in membranes are identified as macropores (>50 nm), mesopores (2 to 50 nm), and micropores (<2 nm). Because the pore sizes in RO membranes are so small, the membranes are defined as dense. The classification of membrane processes on the basis of the size of separation is shown on Fig. 11–31 and in Table 11–23. Referring to Fig.11–31, it can be seen that there is considerable overlap in the sizes of particles removed, especially between NF and RO. Nanofiltration is used most commonly in water softening operations in place of chemical precipitation.

Membrane Containment Vessels

Two types of containment vessels are used with membrane modules: (1) pressurized and (2) submerged.

Figure 11–29

Types of membrane construction: (a) microporous symmetric membrane, (b) nonporous (dense) symmetric membrane, (c) asymmetric membrane, and (d) thin film composite (TFC), sometimes identified as an asymmetric membrane.

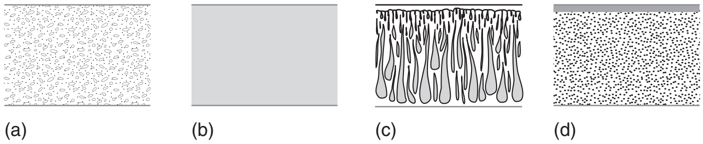
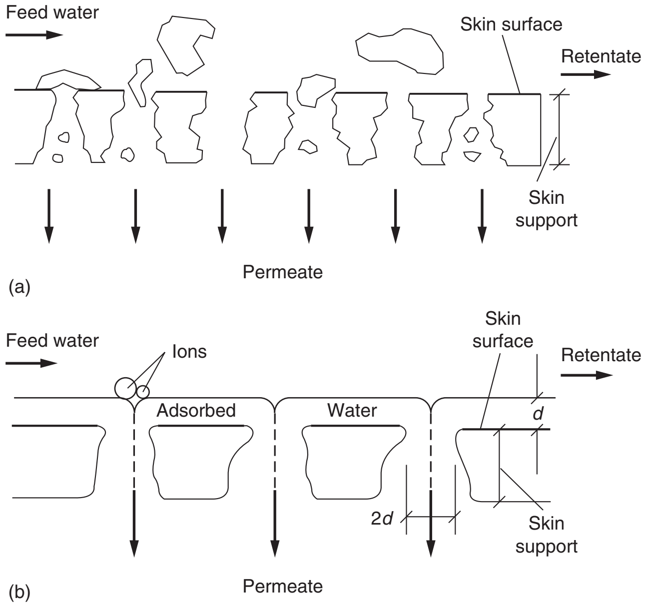


Figure 11–30

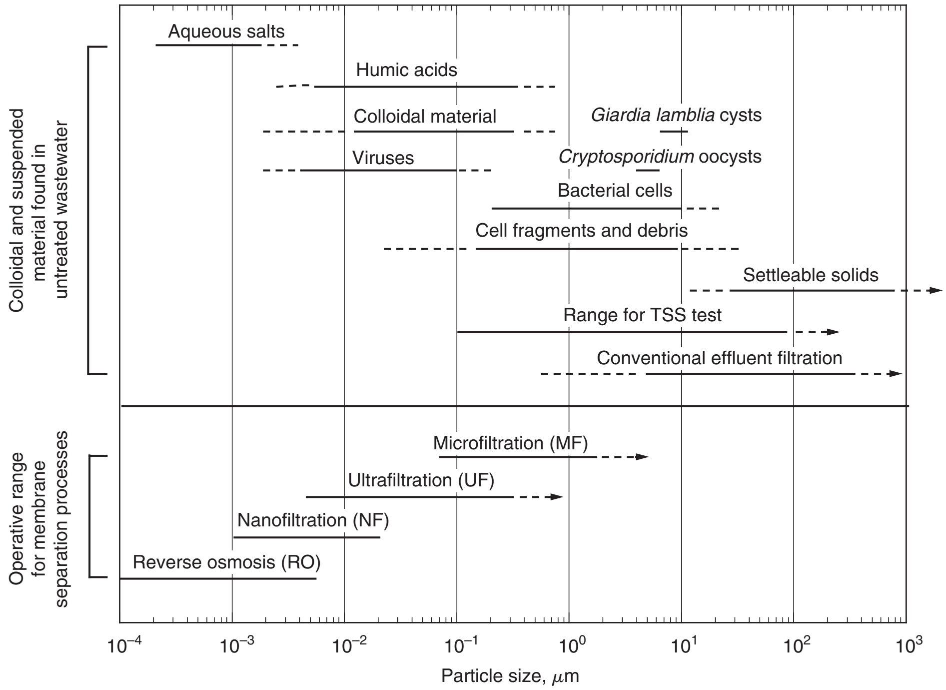
Definition sketch for the removal of wastewater constituents:
(a) removal of large molecules and particles by sieving (size exclusion) mechanism and
(b) rejection of ions by adsorbed water layer.



Pressurized. The primary purpose of the pressure vessel (or tube) is to support the membrane module or modules and keep the feed water and permeate (product water) isolated. The vessel must also be designed to prevent leaks and pressure losses to the outside, minimize the buildup of salt or fouling, and permit easy replacement of the membrane module. Microfiltration and ultrafiltration modules are generally 100 to 300 mm in

Figure 11–31

Comparison of the size of the constituents found in wastewater and the operating size ranges for membrane technologies. The operating size range for conventional depth filtration is also shown.



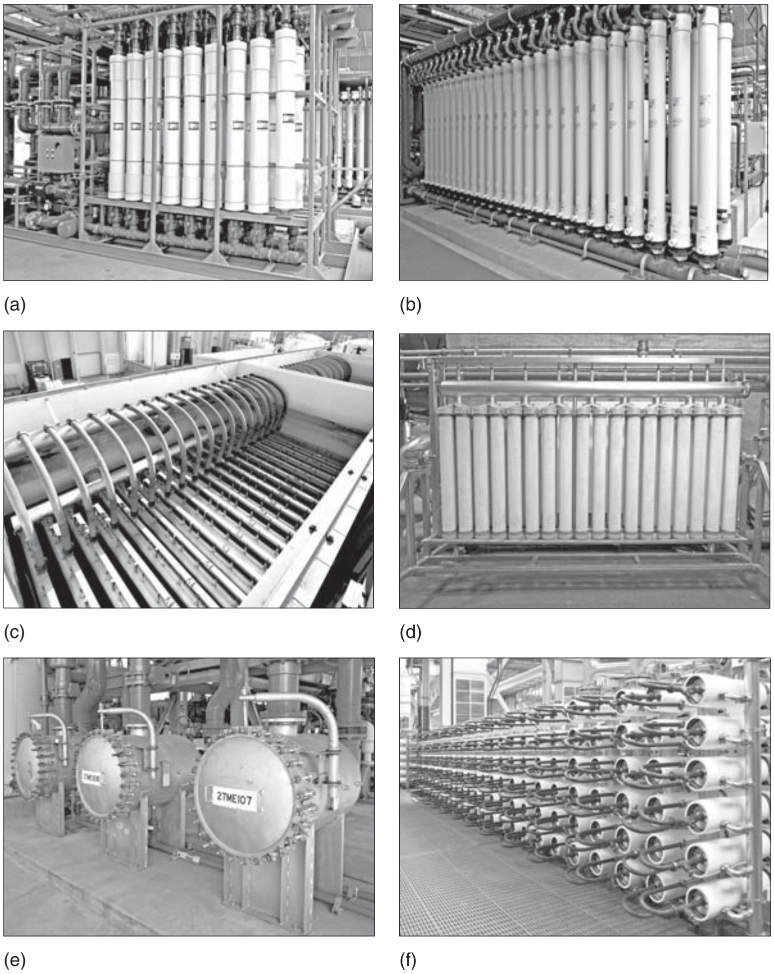


Figure 11–32

Views of various membrane installations: (a) pressurized microfiltration and (b) pressurized ultrafiltration membranes for the filtration of settled secondary effluent, (c) vacuum microfiltration membrane modules in open vessel, (d) membrane module used in open vessel shown in (c), (e) typical cartridge filter used before reverse osmosis, and (f) one bank of a large reverse osmosis installation used to treat activated sludge effluent following microfiltration, chemical addition, and cartridge filtration. Each bank of RO modules is designed to treat 19,000 m^3/d (5 Mgal/d). The capacity of the entire facility is 265,000 m^3/d (70 Mgal/d).

diameter, 0.9 to 5.5 m long, and a single module is placed in a pressurized vessel arranged in racks or skids. Each module must be piped individually for feed and permeate water. Typical pressurized MF membrane modules are shown on Figs. 11–32(a) and (b). The modules for NF and RO are 100 to 300 mm in diameter, 0.9 to 5.5 m long, and from two to eight modules are placed in a single pressurized vessel arranged in racks, either horizontally or vertically [see Figs. 11–32 (d) and (f)]. Vertical placement helps reduce the number of pipes and fittings and the total plant footprint.

In the pressurized configuration, pumps are used to pressurize the feed water and circulate it through the membrane [see Figs. 11–33(a) and (c)]. Centrifugal pumps can be used for MF, UF, and NF; positive displacement pumps or high-pressure turbine pumps are necessary for RO. Depending on the operating pressure and the characteristic of the feed water, a variety of materials have been used including plastic and fiberglass tubes and plumbing components. Steel pressure tubes are required for some reverse osmosis applications, and stainless steel is required for seawater and brackish water having high TDS.

Submerged (Vacuum) Type. In the submerged system, the membrane elements are immersed in a feed water tank as shown on Fig. 11–32(c). The permeate is withdrawn

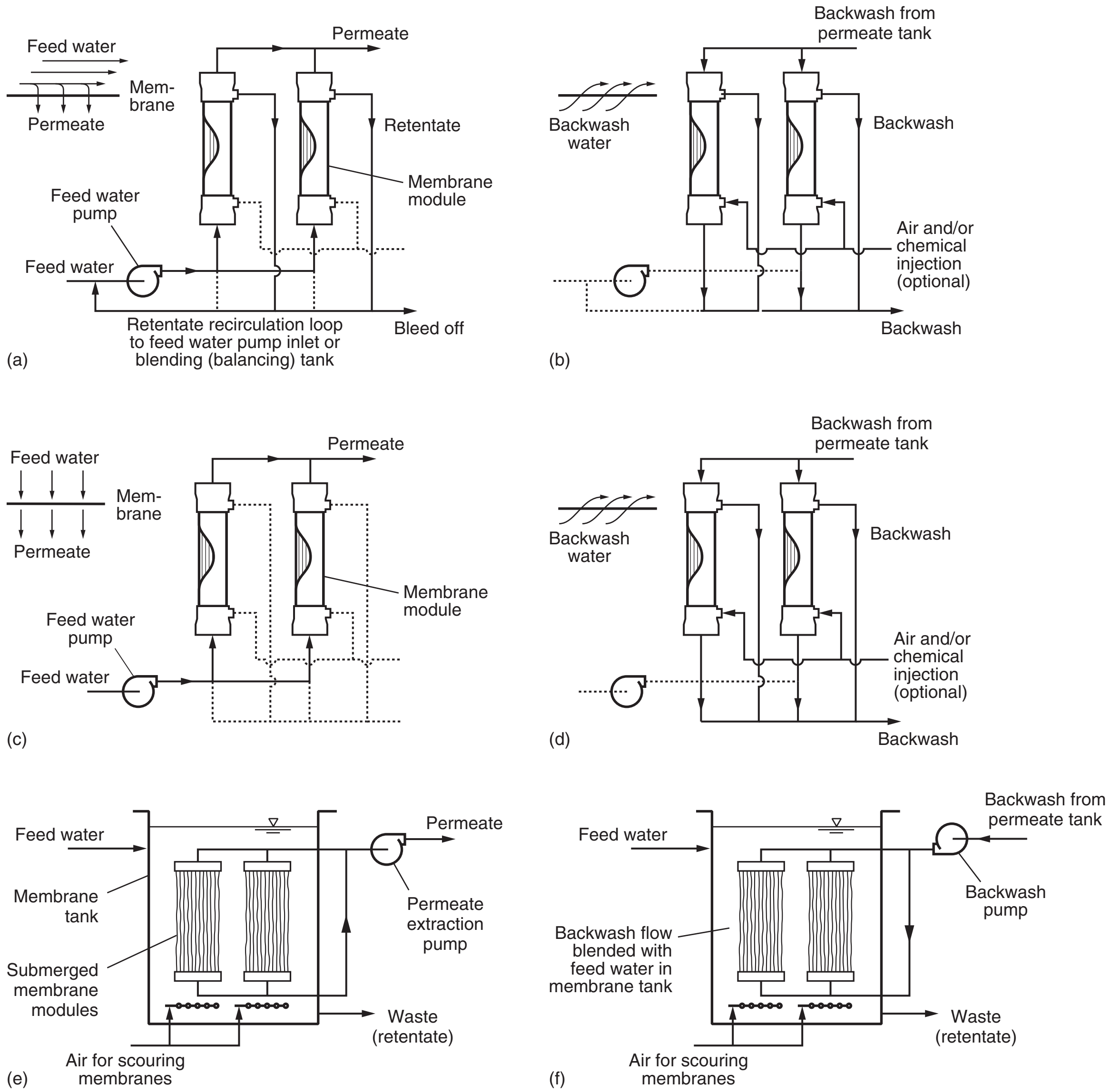


Figure 11–33

Definition sketch for membrane systems: (a) pressurized cross-flow membrane system (see insert), (b) backwashing pressurized cross-flow system, (c) pressurized dead-end flow [see insert, see also Fig. 8–32(a)] membrane system, (d) backwashing pressurized direct flow system, (e) submerged membrane with vacuum draw-off [see also Fig. 11-32(c)], and (f) backwashing submerged system.

through the membrane by applying a vacuum, usually from the suction of a centrifugal pump [see Fig. 11–33(e)]. Transmembrane pressure developed by the permeate pump causes clean water to be extracted through the membrane. Net positive suction head (NPSH) limitations of the permeate pump restrict the submerged membranes to a maximum transmembrane pressure of about 50 kPa, and they operate typically at a transmembrane pressure of 20 to 40 kPa (-28 to -100 kPa vacuum).

Operational Modes for Pressurized Configurations

Two different operational modes are used with pressurized microfiltration and ultrafiltration units: (1) cross-flow and (2) dead-end.

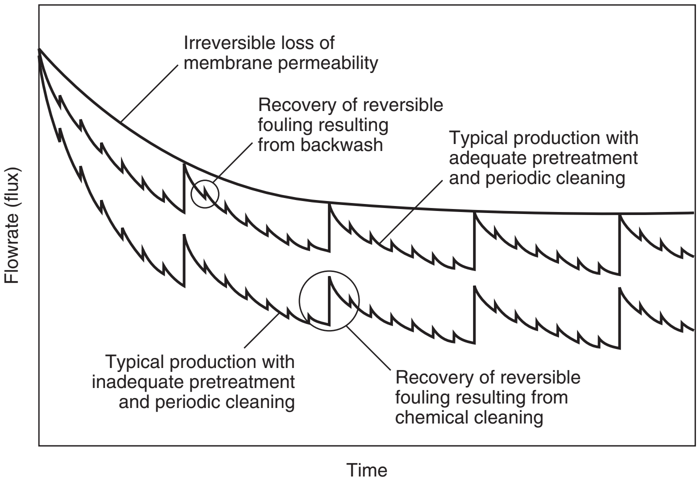
Cross-Flow Mode. In the *cross-flow* mode [see Fig. 11–33(a) and insert] the feed water is pumped more-or-less tangentially to the membrane. The accumulation of particulate matter on the surface of the membrane can be controlled by the shear force of the fluid velocity. The differential pressure across the membrane causes a portion of the feed water to pass through the membrane. Water that does not pass through the membrane is recirculated back to the membrane after blending with influent feed water or is recirculated to a blending (or balancing) tank. In addition, a portion of the water that did not pass through the membrane is bled off for separate processing and disposal [see Fig. 11–33(a)]. It should be noted that cross-flow is the flow pattern in spiral-wound membranes.

Dead-End Mode. In the second configuration, known as *dead-end* (also known as direct-feed or perpendicular feed) [see Fig. 11–33(c) and in the insert], there is no cross-flow (or liquid waste stream) during the permeate production mode. All of the water applied to the membrane passes through the membrane. Particulate matter that cannot pass through the membrane pores is retained on the membrane surface. Dead-end filtration is most effective when the concentration of particulate matter is low or where the accumulated material dose not cause a rapid headloss buildup. Dead-end filtration is used both for pretreatment and where the filtered water is to be used directly.

Membrane Cleaning. As constituents in the feed water accumulate on the membranes (often termed membrane fouling), the pressure builds up on the feed side, the membrane flux (i.e., flow through membrane) starts to decrease, and the percent rejection of certain water quality constituents [see Eq. (11–34)] may actually increase (see Fig. 11–34). When the flux has deteriorated to a given level, the membrane modules are taken out of service and backwashed and periodically cleaned chemically [see Figs. 11–33(b), (d), and (f)]. It should be noted that the quantity of waste washwater produced during membrane cleaning is typically less from pressurized as compared to submerged vacuum systems.

Figure 11–34

Definition sketch for the performance of a membrane filtration system as function of time with and without proper cleaning.



Chemical cleaning is used to restore the membrane performance to near its initial permeability. Two chemical cleaning methods are used typically: (1) clean-in-place (CIP) and (2) chemically enhanced backwash (CEB). In the CIP method, cleaning agents are used to soak and pass across the microfiber elements within the membrane module. In the CEB method, chemical cleaning agents are added to backwash water at selected backwash intervals (typically based on pressure buildup). Sometimes the CEB cleaning method is used for regular operation and when the membrane performance has deteriorated to a given level the CIP cleaning method is used.

A certain irreversible loss of membrane permeability will occur during process operation (see Fig. 11–34). The degree of irreversible permeability loss depends on the membrane material and operating conditions, including (1) long-term aging of the membrane material, (2) mechanical compaction and deformation from high operating pressures, (3) hydrolysis reactions related to solution pH, and (4) reactions with specific constituents in the feed water.

Process Analysis for MF and UF Membranes

Referring to Fig. 11–27, the process analysis for MF and UF membranes involves consideration of the operating pressure, permeate flow, the degree of recovery, and degree of rejection. Constituent and flowrate mass balances are used to assess the performance of the membranes.

Operating Pressure Cross-Flow Mode. For the cross-flow mode of operation, the transmembrane pressure is given by the following expression:

$$P_{tm} = \left(\frac{P_f + P_r}{2}\right) - P_p \tag{11–29}$$

where P_{tm} = transmembrane pressure gradient, bar (Note: 1 bar = 10^5 Pa)
P_f = inlet pressure of the feed water, bar
P_r = pressure of the retentate, bar
P_p = pressure of the permeate, bar

The overall pressure drop across the filter module for the cross-flow mode of operation is given by

$$P = P_f - P_p \tag{11–30}$$

where P = pressure drop across the module, bar
P_f and P_p as defined above

Operating Pressure Dead-End Mode. For the dead end pressurized and submerged modes of operation, the transmembrane pressure is given by the following expression:

$$P_{tm} = P_f - P_p \tag{11–31}$$

where P_{tm} = transmembrane pressure gradient, bar
P_f and P_p as defined above

Permeate Flowrate. The total permeate flowrate from a membrane system is given by

$$Q_p = F_w A \tag{11–32}$$

where Q_p = permeate flowrate, m^3/h
F_w = transmembrane water flux rate, m/h ($m^3/m^2 \cdot h$)
A = membrane area, m^2

As would be expected, the transmembrane water flux rate is a function of the quality and temperature of the feed water, the degree of pretreatment, the characteristics of the membrane, and the system operating parameters. Note that membrane area (A) is not the cross sectional area of the membrane module, but rather the active surface area of the membrane material. For example, a standard 200 mm diameter 1020 mm long (8-in. × 40-in.) RO module contains 37 m^2 (400 ft^2) of membrane surface area.

Recovery. Recovery, *r*, expressed as a percentage, is defined as the ratio of the net water produced to the total water applied during a filter run as follows:

$$r,\% = \frac{V_p}{V_f} \times 100 \tag{11–33}$$

where V_p = net volume of the permeate, m^3
V_f = volume of water fed to the membrane, m^3

In computing the net volume of permeate, the amount of backwash water used must also be taken into consideration.

Rejection. Rejection, *R*, expressed as a percentage or as a dimensionless fraction, is a measure of the amount of material removed from the feed water. It should be noted that there is a difference in the recovery, *r*, (which refers to the water) and rejection, *R*, (which refers to the solute). Rejection, *R*, is given by the following expression.

$$R,\% = \frac{C_f - C_p}{C_f} \times 100 = \left(1 - \frac{C_p}{C_f}\right) \times 100 \tag{11–34}$$

where C_f = feed water concentration, g/m^3, mg/L
C_p = permeate concentration, g/m^3, mg/L

Log Reduction. Another commonly used approach to express the rejection is as log rejection, *LR*, as given below.

$$LR = -\log(1 - R) = \log\left(\frac{C_f}{C_p}\right) \tag{11–35}$$

where *R* is the dimensionless form of Eq. (11–34).

Materials Mass Balance. The corresponding flowrate and constituent mass balance equations for the pressurized cross-flow membrane are:

Flowrate balance: $Q_f = Q_p + Q_r$ (11–36)

Constituent mass balance: $Q_f C_f = Q_p C_p + Q_r C_r$ (11–37)

where Q_f = feed water flowrate, m^3/h, m^3/s
Q_r = retentate flowrate, m^3/h, m^3/s
C_r = retentate concentration, g/m^3, mg/L

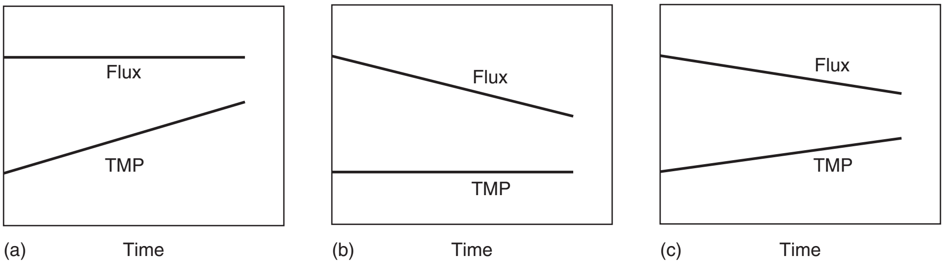


Figure 11–35

Three modes of membrane operation with respect to membrane flux and transmembrane pressure (TMP): (a) constant flux, (b) constant pressure, and (c) non restricted flux and pressure. (Adapted from Bourgeous et al., 1999.)

Operating Strategies for MF and UF Membranes

Operating strategies for membranes are developed based on a consideration of operating pressures and flux rates. Three different operating strategies can be used to control the operation of a membrane process with respect to flux and the transmembrane pressure (TMP). The three modes, illustrated on Fig. 11–35, are (a) constant flux in which the flux rate is fixed and the TMP is allowed to vary (increase) with time, (b) constant TMP in which the TMP is fixed and the flux rate is allowed to vary (decrease) with time, and (c) both the flux rate and the TMP are allowed to vary with time. Traditionally, the constant flux mode of operation has been used. However, based on the results of a study with various wastewater effluents (Bourgeous et al., 1999), the mode in which both the flux rate and the TMP are allowed to vary with time may be the most effective mode of operation. It should be noted that the diagrams on Fig. 11–35 do not reflect the irreversible permeability loss, as described previously. Regardless of the operating strategy an important issue with membranes systems is fiber breakage. The impact of fiber breakage is examined in Example 11–5.

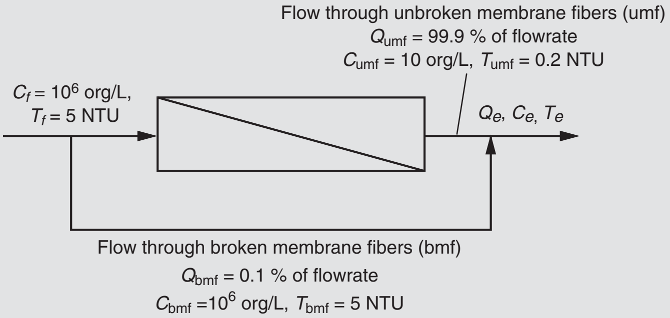
EXAMPLE 11–5 **Impact of Broken Fibers on Membrane Filter Effluent Quality** Membrane filtration is used to treat secondary effluent for reuse applications. The effluent from the wastewater treatment plant, which serves as the influent to the membrane filter installation, has an effluent turbidity of 5 NTU and contains a heterotrophic plate count (HPC) of 10^6 microorganisms/L. The effluent from the membrane filters typically contains less than 10 microorganisms/L and a turbidity of about 0.2 NTU. Using this information, what is the log rejection for microorganisms under normal operation with no broken fibers? If it is assumed that 6 out of 6000 (0.1 percent) membrane fibers have been broken during operation, determine the impact on the effluent microorganism count and turbidity. For the following analysis, neglect the water lost during the backwashing cycle.

Solution

1. Calculate the log rejection for microorganisms with no broken fibers using Eq. (11–35).

$$LR = \log\left(\frac{C_f}{C_p}\right) = \log\left[\frac{(10^6\ \text{org/L})}{(10\ \text{org/L})}\right] = 5.0$$

2. Determine the log rejection for microorganisms assuming that 6 fibers have been broken.
   a. Prepare a mass balance diagram for the condition with the broken fibers.



b. Write mass balance equation for microorganisms in the effluent from the membrane and solve for effluent microorganism concentration.

$$C_e = \frac{C_{\text{umf}}Q_{\text{umf}} + C_{\text{bmf}}Q_{\text{bmf}}}{Q_e}$$
$$= \frac{(10\text{ org/L})(0.999) + (10^6\text{ org/L})(0.001)}{1} = 1010\text{ org/L}$$

c. Calculate the log rejection for microorganisms for the condition with the broken fibers.

$$R_{\log} = \log\left(\frac{C_p}{C_f}\right) = \log\left[\frac{(10^6\text{ org/L})}{(1010\text{ org/L})}\right] = 3.0$$

3. Calculate the impact on turbidity assuming that 6 fibers have been broken. Use the mass balance equation developed in Step 2 and solve for the effluent turbidity.

$$T_e = \frac{T_{\text{umf}}Q_{\text{umf}} + T_{\text{bmf}}Q_{\text{bmf}}}{Q_e}$$
$$= \frac{(0.2\text{ NTU})(0.999) + (5\text{ NTU})(0.001)}{1} = 0.205\text{ NTU}$$

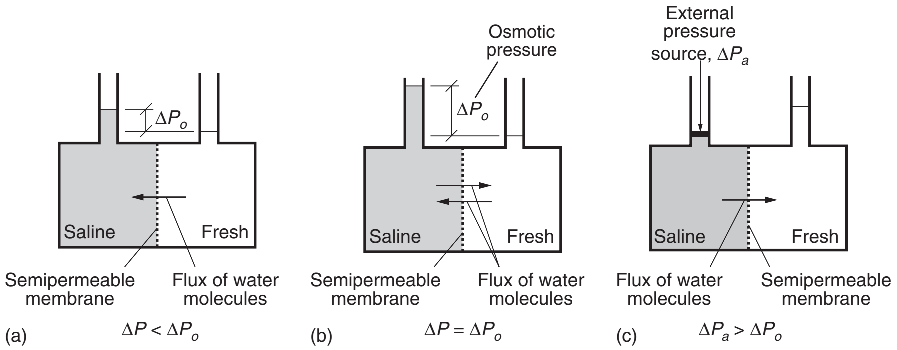
Comment This example is used to demonstrate that a few broken fibers can have a significant impact on the microorganism count in the effluent (1010 versus 10/L) and the log removal (5 versus 3.0 log) and essentially no impact on the effluent turbidity (0.2 versus 0.205 NTU, the difference is not measurable). For this reason, turbidity alone cannot be used as a surrogate measure for bacterial quality, and disinfection of microfiltration effluent will be required to protect public health in sensitive applications. The use of turbidity monitoring is often accompanied with the practice of pressure decay testing and particle counting for monitoring membrane integrity.

Process Analysis for Reverse Osmosis

Process analysis for reverse osmosis involves consideration of the membrane water and mass flux rate, permeate recovery ratio, rejection factor, and the corresponding mass balance analysis. To understand the details of the process analysis for reverse osmosis, it will be helpful to first review the fundamental basis for the reverse osmosis process.

Figure 11–36

Definition sketch for reverse osmosis: (a) osmosis (the differential pressure between solutions is less than the osmotic pressure), (b) osmotic equilibrium (the differential pressure between solutions is equal to the osmotic pressure), and (c) reverse osmosis (the applied pressure is greater than the osmotic pressure).



The fundamental basis is presented below followed by the expressions used for the analysis of the process.

Fundamental Basis for Reverse Osmosis Process. When two solutions having different solute concentrations are separated by a semipermeable membrane, a difference in chemical potential will exist across the membrane (see Fig. 11–36). Water will tend to diffuse through the membrane from the lower-concentration (higher-potential) side to the higher-concentration (lower-potential) side; this phenomenon is called forward osmosis [see Fig. 11–26(a)]. In a system having a finite volume, flow continues until the pressure difference balances the chemical potential difference. This balancing pressure difference is termed the osmotic pressure and is a function of the solute characteristics and concentration and temperature. If a pressure gradient, opposite in direction and greater than the osmotic pressure, is imposed across the membrane, flow from the more concentrated to the less concentrated region will occur; this phenomenon is termed *reverse osmosis* [see Fig. 11–36(c)].

Membrane Flux and Area Requirements. A number of different models have been developed to determine the membrane surface area and the number of stages (arrays) required (see Fig. 11–37). The basic equations used to develop the various models are as follows.

Feed Water Flux Rate. Referring to Fig. 11–27, the flux of water through the membrane is a function of the pressure gradient:

$$F_w = k_w(\Delta P_a - \Delta \Pi) = \frac{Q_p}{A} \tag{11–38}$$

where F_w = feed water flux rate, L/m^2·h

k_w = mass transfer coefficient for water flux (involving temperature, membrane characteristics, and solute characteristics), L/m^2·h·bar

ΔP_a = average applied pressure gradient, bar (Note: Number 1 bar = 10^5 Pa)

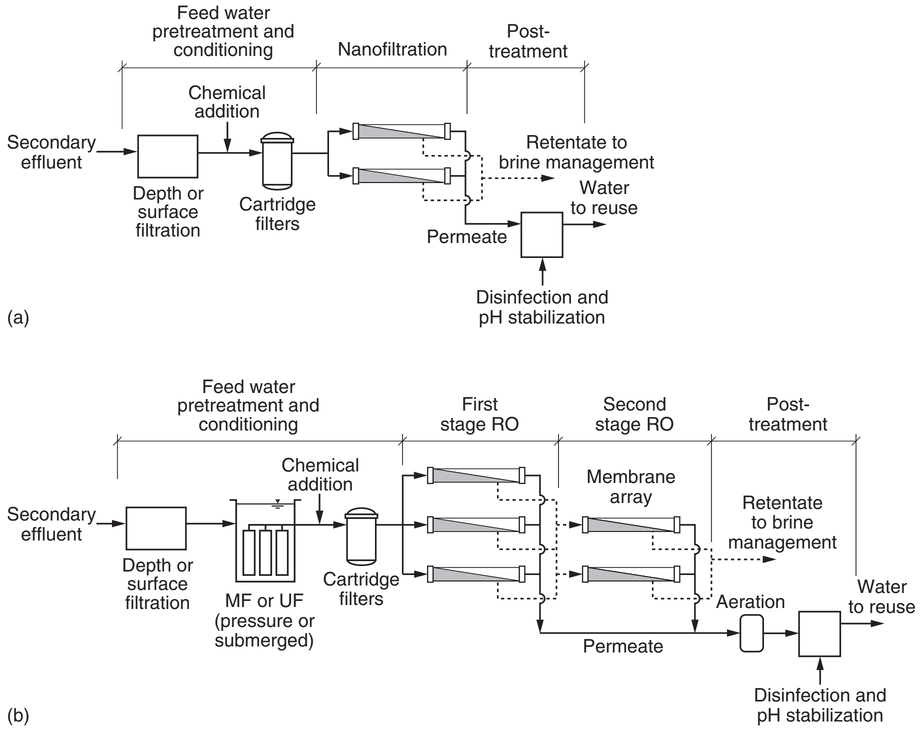
$$= \left(\frac{P_f + P_c}{2}\right) - P_p$$

$\Delta\Pi$ = osmotic pressure gradient, bar

$$= \left(\frac{\Pi_f - \Pi_c}{2}\right) - \Pi_p$$

Figure 11–37

Typical process flow diagrams: (a) depth or surface filtration with nanofiltration and (b) combined microfiltration or ultrafiltration with reverse osmosis.



P_f = inlet pressure of feed water, bar
P_r = pressure of retentate, bar
P_p = pressure of permeate, bar
Π_f = osmotic pressure of feed water, bar
Π_r = osmotic pressure of retentate, bar
Π_p = osmotic pressure of permeate, bar
Q_p = permeate flowrate, L/h
A = membrane area, m^2

Mass (Solute) Flux Rate. Some solute passes through the membrane in all cases. Solute flux can be described adequately by an expression of the form

$$F_s = k_s \, \Delta C_s = \frac{(Q_p)(10^{-3}\,\text{m}^3/\text{L})C_p}{A} \tag{11–39}$$

where F_s = mass flux of solute, $g/m^2 \cdot h$
k_s = mass transfer coefficient for solute, m/h
ΔC = solute concentration gradient across membrane, g/m^3
$= \left(\dfrac{C_f + C_r}{2}\right) - C_p$
C_f = solute concentration in feed water, g/m^3
C_r = solute concentration in retentate (concentrate), g/m^3
C_p = solute concentration in permeate, g/m^3
Q_p = permeate flowrate, L/h

Permeate Recovery Ratio. The permeate recovery ratio, *r*, expressed as a percentage, represents the conversion of feed water to permeate (product water), and is defined as

$$r, \% = \frac{Q_p}{Q_f} \times 100 \tag{11–40}$$

where Q_p = permeate flowrate, L/h, m^3/h, or m^3/s
Q_f = feed water flowrate, L/h, m^3/h, or m^3/s

The permeate recovery ratio affects the capital and operating cost of a membrane system. The volume of feed water required for a given permeate capacity is determined directly by the design recovery ratio. Also, the size of the feed water system, capacity of the pretreatment system, and size of the high pressure pumps and supply piping are also functions of the recovery ratio. With increased recovery, the feed water flowrate, is reduced, the pressure may increase somewhat, but the brine will be more concentrated which can make disposal more difficult.

An example of the effect of the permeate recovery ratio on feed pressure, power consumption and feed flow is shown on Fig. 11–38 for an RO system operating at recovery rates between 60 and 90 percent. The feed water flowrate depends only on the recovery ratio. The feed pressure is a complex function of recovery ratio, feed water salinity feed water temperature, and specific permeate flux of the membrane. The power requirement of the high pressure pump is proportional to the flowrate and pressure. In the usual range of operating parameters, for an increase in recovery ratio, the decrease in feed water flowrate will have a greater effect on power consumption than an increase in feed water pressure (Wilf, 1998). For RO, higher operating pressures are desirable because the degree of separation and the quality of the product are improved.

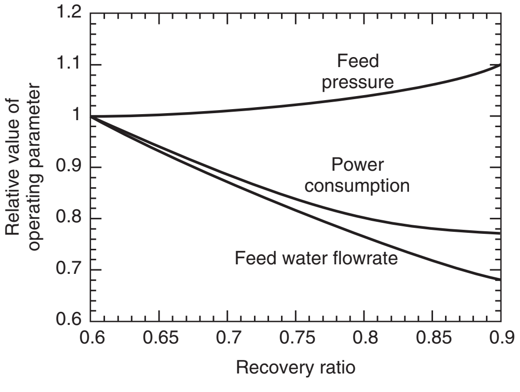
Rejection Factor. Rejection (or retention), R, expressed as percentage or as a dimensionless fraction, is a measure of the amount of solute or solid that is retained or does not pass through the membrane; it is calculated using the following expression.

$$R, \% = \left(\frac{C_f - C_p}{C_f}\right) \times 100 = \left(1 - \frac{C_p}{C_f}\right) \times 100 \tag{11–41}$$

where C_f = concentration in the feed water, g/m^3
C_p = concentration in the permeate, g/m^3

Figure 11–38

Effect of permeate recovery on feed pressure, feed flowrate, and power consumption.



The rejection efficiency of reverse osmosis membranes for specific species can range from 85 to 99.5 percent and is quoted by the manufacturer for a standard set of feed conditions. When the rejection of microorganisms is considered, it is more convenient to express the rejection as log rejection as given by Eq. (11–35), repeated here for convenience.

$$LR = -\log(1 - R) = \log\left(\frac{C_f}{C_p}\right) \qquad (11\text{–}42)$$

Materials Mass Balances. The flowrate and constituent mass balance equations given previously for microfiltration and ultrafiltrations also apply to NF and RO systems. The equations are

$$\text{Flowrate balance: } Q_f = Q_p + Q_r \qquad (11\text{–}43)$$

$$\text{Constituent mass balance: } Q_f\, C_f = Q_p\, C_p + Q_r\, C_r \qquad (11\text{–}44)$$

where Q_r = retentate flowrate, m^3/h, m^3/s
C_r = retentate concentration, g/m^3
Other terms are as defined previously

Use of the above equations to estimate the required surface area for TDS reduction is illustrated in Example 11–6.

EXAMPLE 11–6 **Determination of Membrane Area Required for Demineralization** A brackish water having a TDS concentration of 3000 g/m^3 is to be desalinized using a thin film composite membrane having a solvent (water) mass transfer coefficient k_w of 9×10^{-9} s/m (9×10^{-7} m/s·bar) and a solute (i.e., TDS) mass transfer coefficient k_i of 6×10^{-8} m/s. The product water is to have a TDS of no more than 200 g/m^3. The feed water flowrate is 0.010 m^3/s. The net operating pressure ($\Delta P_a - \Delta P$) will be 2500 kPa (2.5×10^6 kg/m·s^2). Assume the recovery rate, r, will be 90 percent and that all of the water is to be processed through the membrane unit to remove other constituents in addition to TDS. Estimate the rejection rate, R, and the concentration of the retentate.

Solution

1. The problem involves determination of the membrane area required to produce 0.009 m^3/s (0.9×0.010 m^3/s) of water with a TDS concentration equal to or less than 200 g/m^3. If the estimated permeate TDS concentration is well below 200 g/m^3 and TDS is the only constituent of concern, blending of feed water and permeate can be used to reduce the required membrane area.
2. Estimate the membrane area using Eq. (11–38) and the water mass transfer coefficient. The estimated area may need to be adjusted based on the solute mass transfer rate.

$$F_w = k_w(\Delta P_a - \Delta P)$$

$$= (9 \times 10^{-9}\ \text{s/m})(\,2.5 \times 10^6\ \text{kg/m·s}^2) = 2.25 \times 10^{-2}\ \text{kg/m}^2\text{·s}$$

$$Q_p = F_w \times A,\ Q_p = r\,Q_f,\ Q_p = 0.9\,Q_f$$

$$A = \frac{(0.9 \times 0.010\ \text{m}^3/\text{s})(10^3\ \text{kg/m}^3)}{(2.25 \times 10^2\,\text{kg/m}^2\text{·s})} = 400\ \text{m}^2$$

3. Estimate the permeate TDS concentration using Eq. (11–39) and the estimated area.

$$F_i = k_i\,\Delta C_i = \frac{Q_p C_p}{A}$$

Substituting for ΔC_i and solving for C_p yields

$$C_p = \frac{k_i[(C_f + C_r)/2]A}{Q_p + k_i A}$$

Assume $C_c \approx 10C_f$ (Note: If the estimated C_r value and the computed value of C_r, as determined below, are significantly different, the value of C_p must be recomputed)

$$C_p = \frac{(6 \times 10^{-8}\,\text{m/s})[(3\,\text{kg/m}^3 + 30\,\text{kg/m}^3)/2](400\,\text{m}^2)}{(0.01\,\text{m}^3/\text{s}) + (6 \times 10^{-8}\,\text{m/s})(400\,\text{m}^2)} = 0.044\,\text{kg/m}^3$$

4. Estimate the rejection rate using Eq. (11–41)

$$R,\% = \frac{C_f - C_p}{C_f} \times 100$$

$$R = \frac{(3.0\,\text{kg/m}^3) - (0.044\,\text{kg/m}^3)}{(3.0\,\text{kg/m}^3)} \times 100 = 98.5\%$$

5. Estimate the retentate TDS using Eq. (11–41).

$$C_r = \frac{Q_f C_f - Q_p C_p}{Q_r}$$

$$C_r = \frac{(1.0\,\text{L})(3.0\,\text{kg/m}^3) - (0.9\,\text{L})(0.044\,\text{kg/m}^3)}{(0.1\,\text{L})} = 29.6\,\text{kg/m}^3$$

The estimated value of C_r used in Step 3 (30 kg/m^3) is ok.

Comment If the permeate TDS concentration were significantly below 200 g/m^3, blending of feed and permeate could be used to reduce the required membrane area. In this example blending cannot be used.

Membrane Fouling

Membrane fouling is, perhaps, the most important consideration in the design and operation of membrane systems as it affects pretreatment needs, cleaning requirements, operating conditions, cost, and performance. Membrane fouling will occur depending on the site-specific physical, chemical, and biological characteristics of the feed water, the type of membrane, and operating conditions. As reported in Table 11–25, four general forms of fouling can occur: (1) particulate fouling, due to a buildup of the constituents in the feed water on the membrane surface, (2) precipitation of inorganic salts resulting in the formation of inorganic scales, (3) organic fouling due to the presence of organic matter, and (4) biological fouling due to the presence of microorganisms in the feed water. Any or all of these forms of fouling can occur simultaneously and over time. In addition, membranes can be damaged by the presence of chemical substances that can react with the membrane. Typical wastewater constituents in that can cause membrane fouling are also presented in Table 11–25.

Table 11–25

Typical constituents in wastewater that cause membrane fouling and other constituents that can cause damage to the membranes[a]

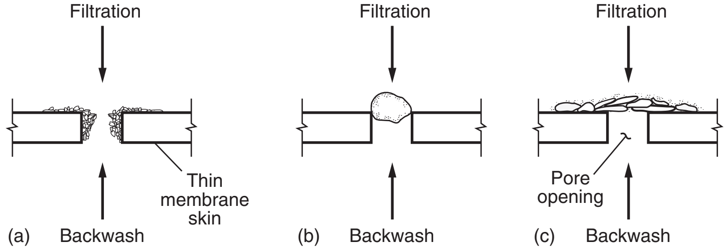
Type of fouling	Responsible wastewater constituents	Remarks
Particulate fouling	Organic and inorganic colloids Emulsified oils Clays and silts Silica Iron and manganese oxides Oxidized metals Metal salt coagulant products Powdered activated carbon	Particulate fouling can be reduced by cleaning the membrane at regular intervals
Scaling (precipitation of supersaturated salts)	Barium sulfate Calcium carbonate Calcium fluoride Calcium phosphate Strontium sulfate Silica	Scaling can be reduced by limiting salt content, by pH adjustment, and by other chemical treatments such as the addition of antiscalants
Organic fouling	Natural organic matter (NOM) including humic and fulvic acids, proteins and polysccharides Emulsified oils Polymers used in treatment process	Effective pretreatment can be used to limit organic fouling
Biofilm fouling	Dead microorganisms Living microorganisms Polymers produced by microorganisms	Biofilms are formed on membrane surface by colonizing bacteria
Damage to membrane	Acids Bases pH extremes Free chlorine Free oxygen	Membrane damage can be limited by controlling the amount of these substances in the feed water. The extent of the damage depends on the nature of the membrane selected

[a] In many cases all four types of fouling can occur simultaneously.

Particulate Fouling. Particulate fouling is caused by the presence of particulate matter in the feed water. To protect RO and NF membrane systems from large particulate fouling, cartridge filters [see Table 11–21(g)] are used ahead of the membrane feed pumps. These are normally 5 to 15 μm pore size woven filters designed to prevent fouling from relatively large particles. Nonetheless, much smaller particles can also damage and foul RO and NF modules. As noted in Table 11–25, particulate constituents can include organic and inorganic colloids, emulsified oils, clays and silts, silica, and metal oxides and salts. Of these constituents, silica has proven to be one of the more problematic in wastewater. Silica $(SiO_2)_n$ in wastewater can be found in a variety of different forms including reactive, colloidal, and particulate silica depending on the chemical characteristics of the feed water.

Figure 11–39

Modes of membrane fouling: (a) pore narrowing, (b) pore plugging, and (c) gel/cake formation caused by concentration polarization.



Three accepted mechanisms resulting in resistance to flow due to the accumulation of material (see Fig. 11–39) are (1) pore narrowing, (2) pore plugging, and (3) gel/cake formation caused by concentration polarization (Ahn et al., 1998). The mechanisms of pore plugging and pore narrowing will only occur when the particulate matter in the feed water is smaller than the pore size or the molecular weight cutoff. Pore plugging occurs when particles the size of the pores become stuck in the pores of the membrane. Pore narrowing consists of solid material attaching to the interior surface of the pores, which results in a narrowing of the pores. It has been hypothesized that once the pore size is reduced, concentration polarization is amplified further causing an increase in fouling (Crozes et al., 1997).

Gel/cake formation, caused by concentration polarization, occurs when the majority of the solid matter in the feed is larger than the pore sizes or molecular weight cutoff of the membrane. Concentration polarization can be described as the build up of matter close to or on the membrane surface that causes an increase in resistance to solvent transport across the membrane. Some degree of concentration polarization will always occur in the operation of a membrane system. The formation of a gel or cake layer, however, is an extreme case of concentration polarization where a large amount of matter has actually accumulated on the membrane surface forming a gel or cake layer.

Scaling. As chemical constituents in the feed water are removed at the surface of the membrane, their concentration increases locally. When the concentrations of the individual constituents increase beyond their solubility limits, a variety of different types of salts can be precipitated, depending on the chemical characteristics and temperature of the feed water. Chemical precipitation is especially critical in RO units used for desalination, because of the high initial salt concentration in seawater. The chemical scale that forms on the membrane surface is of importance because it can reduce the water permeability of the membrane and potentially cause irreversible damage to the membrane.

Organic Fouling. Most treated wastewater contains a variety of organic matter in varying concentrations. As noted in Table 11–25, organic foulants can include NOM that was present originally in the water supply, NOM produced during biological treatment, emulsified oils, and organic polymers that may have been used in the wastewater treatment process including polymers used as filter aids in tertiary treatment and polymers recycled to the treatment process from dewatering activities. Because these polymeric materials are sticky, they can accumulate on the membrane surface and accelerate fouling by forming stable organic/inorganic particulate matter, which can reduce the water permeability of the membrane.

Biological Fouling. Effluent from biological treatment systems presents a special problem, as the membranes are susceptible to fouling because of the biological activity that can occur. Because the concentration of organic matter and nutrients is elevated at the membrane surface, conditions are favorable for the growth of microorganisms. As microorganisms begin to colonize on the membrane surface, the water permeability of the membrane will be reduced. When a membrane process is operated intermittently, the water permeability of the membrane can be reduced further if the microorganisms start to grow into the membrane pores. The growth of microorganisms is also of concern because of the production of extracellular polymers which can interact with other foulants, as described previously.

Control of Membrane Fouling

Typically, three approaches are used to control membrane fouling: (1) pretreatment of the feed water, (2) membrane backflushing, and (3) chemical cleaning of the membranes. Pretreatment is used to reduce the TSS, colloidal material, and bacterial content of the feed water. Often the feed water will be conditioned chemically to limit chemical precipitation within the units. Many proprietary pretreatment chemicals such as anti-scalants, biocides, and scale inhibitors are available in the marketplace to control NF and RO fouling. In low pressure membrane operations (MF and UF), the method of eliminating the accumulated material from the membrane surface is backflushing with water and/or air. Chemical treatment is used to remove constituents that are not removed during conventional backwashing. Chemical precipitates can be removed by altering the chemistry of the feed water and by chemical treatment. Damage of the membrane due to deleterious constituents typically cannot be reversed. The need for pretreatment and pretreatment options for NF and RO are discussed below.

Assessing Need for Pretreatment for NF and RO. To assess the treatability of a given wastewater with NF and RO membranes, a variety of fouling indexes have been developed over the years. The three principal indices are the silt density index (SDI), the modified fouling index (MFI), and the mini plugging factor index (MPFI). Fouling indexes are determined from simple membrane tests. The sample must be passed through a 0.45 μm Millipore filter with a 47-mm internal diameter at 210 kPa (30 $\text{lb}_f/\text{in.}^2$) gauge to determine any of the indexes. The time to complete data collection for these tests varies from 15 min to 2 h, depending on the fouling nature of the water.

Silt Density Index. The most widely used index is the SDI (DuPont, 1977; ASTM, 2002). The SDI is defined as follows.

$$SDI = \frac{100[1 - (t_i/t_f)]}{t} \tag{11–45}$$

where t_i = time to collect the initial sample of 500 mL
t_f = time to collect final sample of 500 mL
t = total time for running the test

The silt density index is a static measurement of resistance which is determined by samples taken at the beginning and end of the test. The SDI does not measure the rate of change of resistance during the test. Recommended SDI values are reported in Table 11–26. The calculation of the SDI is demonstrated in Example 11–7.

Modified Fouling Index. The modified fouling index (MFI) is determined using the same equipment and procedure used for the SDI, but the volume is recorded every 30 s over a

Table 11–26

Recommended values for fouling indexes[a]

Membrane process	Fouling index		
	SDI	MFI, s/L^2	MPFI, L/s^2
Nanofiltration	0–2	0–10	0–1.5 × 10^{-4}
Reverse osmosis hollow fiber	0–2	0–2	0–3 × 10^{-5}
Reverse osmosis spiral wound	0–3[b]	0–2	0–3 × 10^{-5}

[a] Adapted in part from Taylor and Wiesner (1999) and AWWA (1996).

[b] Although a value of 3 is acceptable, the trend is to lower the upper limit to a value of 2 or less.

15-min filtration period (Schippers and Verdouw, 1980). Derived from a consideration of cake filtration, the MFI is defined as follows:

$$\frac{1}{Q} = a + \text{MFI} \times V \qquad (11\text{–}46)$$

where Q = average flowrate, L/s
a = constant (intercept of linear portion of curve)
MFI = modified fouling index, s/L^2
V = volume, L

The value of the MFI is obtained as the slope of the straight-line portion of the curve obtained by plotting the inverse of the flowrate versus the cumulative volume [see Fig. 11–40(a)].

Mini Plugging Factor Index. The mini-plugging factor index (MPFI) is a measure of the change in flowrate as a function of time, as illustrated on Fig. 11–40(b) (Taylor and Jacobs, 1996). The equipment used for the MPFI test is the same as that used for the SDI and MFI tests. The MPFI is defined as the slope of the linear portion of the flowrate versus time curve [see Fig. 11–40(b)], which is ascribed to cake fouling. In equation form, the MPFI is expressed as follows:

$$Q = (\text{MPFI})\,t + a \qquad (11\text{–}47)$$

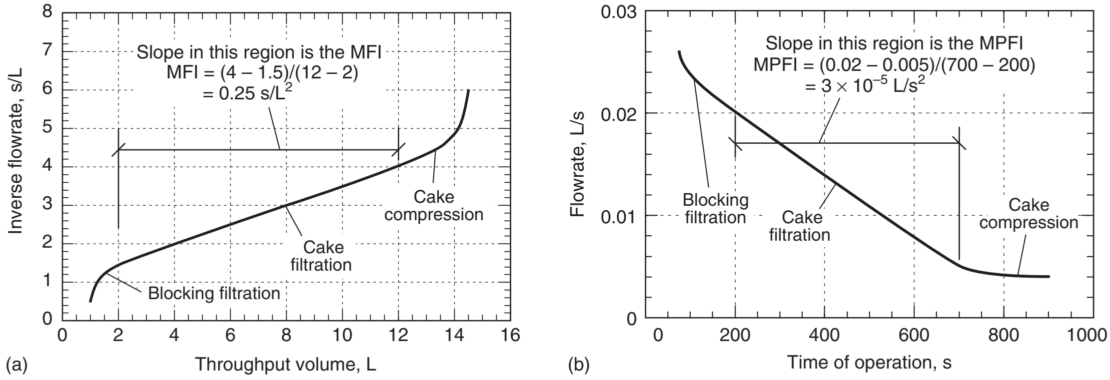


Figure 11–40

Typical plot to determine fouling indexes: (a) modified fouling factor (MFI) and (b) mini plugging factor index (MPFI).

where Q = average flowrate at 30 s intervals, L/s
MPFI = mini plugging factor index, L/s^2
t = time, s
a = constant (intercept of linear portion of curve)

Typical values for the MPFI are reported in Table 11–26. Because the MFI is based on throughput volume, it is thought to be a more sensitive index than the MPFI for characterization of fouling.

EXAMPLE 11–7 **Silt Density Index for Reverse Osmosis** Determine the silt density index for a proposed feed water from the following test data. If a spiral wound RO membrane is to be used, will pretreatment be required?

Test run time = 30 min
Initial 500 mL = 2 min
Final 500 mL = 10 min

Solution

1. Calculate the SDI using Eq. (11–45).

$$\text{SDI} = \frac{100[1 - (t_i/t_f)]}{t}$$

$$\text{SDI} = \frac{100[1 - (2/10)]}{30} = 2.67$$

2. Compare the SDI to the acceptable criteria.
   Calculated SDI value of 2.67 is less than 3 (see Table 11–26); therefore, further pretreatment would not be needed normally.

Comment As a practical matter, because the SDI value is close to 3.0 it may be prudent to consider some form of pretreatment to prolong the filtration cycle.

Limitations of Fouling Indexes. The SDI and MFI fouling indexes described above, and others currently in use, have serious limitations including (1) the fact that a dead-end test is used to gather data to predict the fouling performance of a cross-flow membrane, (2) the test is conducted with a 0.45 μm filter which does not capture the effect of smaller colloidal particles, (3) the test is not representative of cake filtration, which occurs in cross-flow, (4) the test does not measure the propensity for scale formation, and (5) the test is conducted under conditions of constant pressure with variable flux, where the opposite operational mode is normally used in practice. It should be noted that several other indexes, using MF or UF membranes in place of the Millipore filter, to reflect the effect of smaller colloidal material and large dissolved organic material on fouling, are currently under development.

Pretreatment for NF and RO. A very high quality feed is required for efficient operation of a nanofiltration or reverse osmosis unit. Membrane elements in the reverse osmosis unit can be fouled by colloidal matter and dissolved constituents in the feed water. The pretreatment options identified in Table 11–27 have been used singly and in combination. The effectiveness of the treatment options can be assessed with one or more of the indexes discussed previously. Regular chemical cleaning of the membrane elements (about once a month) is also necessary to restore and maintain the membrane flux.

Application and Performance of Membranes

With evolving health concerns and the development of new and lower cost membranes, the application of membrane technologies in the field of environmental engineering has increased dramatically within the past 5 years. The increased use of membranes is expected to continue well into the future. In fact, the use of conventional filtration technology, such as described in Sections 11–4 and 11–5, may be a thing of the past within 10 to 15 years, especially in light of the need to remove resistant organic constituents of concern. The principal applications of the various membrane technologies in wastewater treatment are reported in Table 11–28. Application of the membrane technologies for the removal of specific constituents from wastewater is given in Table 11–29. Each of the membrane technologies is considered further in the following discussion.

Microfiltration. Microfiltration membranes are the most numerous on the market and are the least expensive. The use of membranes for biological treatment is currently one of the most important uses of membranes in wastewater treatment. In advanced treatment applications, microfiltration has been used, most commonly as a replacement for depth filtration to reduce turbidity, remove residual suspended solids, and reduce microorganisms for effective disinfection and as a pretreatment step for reverse osmosis (see Fig. 11–41). Typical operating information for microfiltration including size range, operating pressures, and flux rate are presented in Table 11–30. Typical performance data are reported in Table 11–31. Corresponding variability data are presented in Table 11–32. Care should be used in applying the performance data reported in Table 11–30 as it has been found that the performance of MF is to a large extent site specific, especially with respect to fouling.

Ultrafiltration. Ultrafiltration (UF) membranes are used for many of the same applications as described above for microfiltration. Some UF membranes with small pore sizes have also been used to remove dissolved compounds with high molecular weight, such as colloids, proteins, and carbohydrates. The membranes do not remove sugar or salt. The major distinction between UF and MF is that UF can remove viruses whereas MF cannot.

Figure 11–41

Typical process flow diagram for the production of potable water employing filter screens, open vessel microfiltration, cartridge filters, revers osmosis, UV advanced oxidation, decarbonation, and lime stabilization. (Adapted from Orange County Water District, CA.)

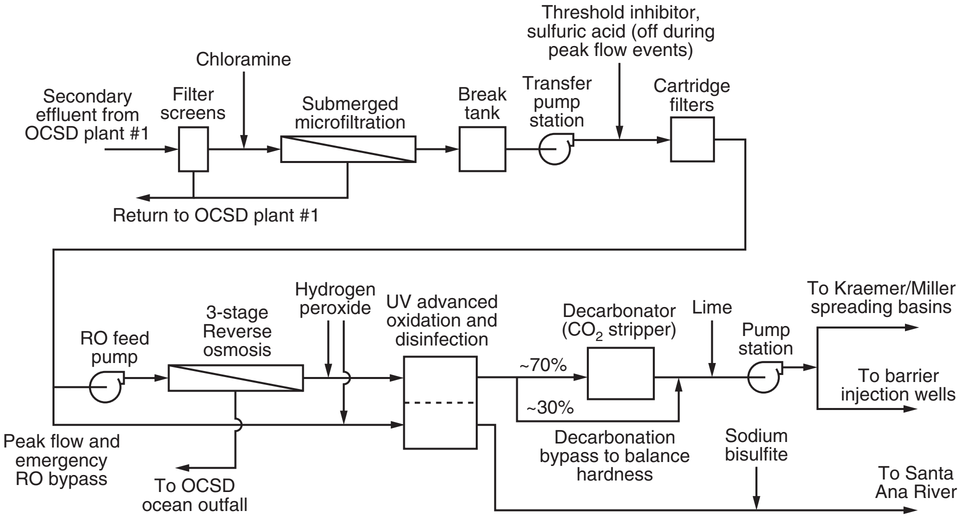


Table11–27

Methods of pretreatment for nanofiltration and reverse osmosis systems

Material to be removed	Method of pretreatment	Description or discussion
Particulate matter and colloidal	Depth or surface filtration, microfiltration, or ultrafiltration	Particulate matter can be removed by various methods of filtration. Fouling agents may pass through these filtration systems, thus the potential for membrane fouling should be verified by pilot testing.
Particulate matter	Cartridge filter	Cartridge filters are pressure-driven filters with pore sizes varying from 5 to 15 μm and are commonly installed ahead of RO membranes. Cartridge filters provide a final level of protection against the intrusion of relatively large solids into the reverse osmosis system. When antiscalants are used, cartridge filters are used to remove the colloidal material and impurities found in antiscalent chemicals. The filters do not remove dissolved substances. Generally, the pressure drop across a clean cartridge filter is between 0 and 35 kPa. As the solids accumulate and the pressure drop reaches a threshold range of 70 to 80 kPa, the cartridge has to be removed and replaced (Paranjape et al., 2003).
Microorganisms	Disinfection	Disinfection of the feed water may be accomplished using either chlorine, ozone, or UV irradiation to limit bacterial activity. Ultrafiltration can also be used to reduce the number of microorganisms.
Scale formation	pH adjustment	To inhibit scale formation, the pH of the feed water is adjusted (usually with sulfuric acid) within the range from 4.0 to 7.5. A low pH enhances conversion of carbonate into bicarbonate species, which are much more soluble. Cellulose acetate RO membranes have an optimum pH range of 5 to 7 as they are prone to hydrolysis below a pH of 5. Newer polyamide RO membranes can be used over a broader pH range of 2 to 11 (Paranjape et al., 2003).
	Antiscalants	Antiscalants are compounds that either prevent scale formation entirely or permit formation of scales that can be removed easily during cleaning. Certain antiscalants, however, may increase the fouling caused by humic acids (Richard et al., 2001).
Iron and manganese	Ion exchange or chemical treatment	Removal of iron and manganese will decrease scaling potential. The exclusion of oxygen may be necessary to prevent oxidation of iron and manganese.
Sparingly soluble salts	Chemical treatment	Sparingly soluble salts such as silica can be removed by chemical treatment for industrial purposes, i.e., removal of silica may be required to prevent precipitation on heat exchangers. Chemical treatment may include the addition of aluminum and iron oxides, zinc chloride, magnesium oxide, ozone (when ozone-resistant membranes are used), and ultra-high lime clarification. Lime clarification, however, may not be as effective as other pretreatment methods in removing materials that foul RO membranes thus resulting in more frequent cleaning of the membranes (Gagliardo, 2000).

Ultrafiltration is used typically in industrial applications for the production of high purity process rinse water. Typical operating and performance data are presented in Tables 11–30 and 11–31, respectively. Performance variability data are presented in Table 11–32.

Nanofiltration. Nanofiltration, also known as "loose" RO or low pressure RO, can reject particles as small as 0.001 μm. Nanofiltration is used for the removal of selected dissolved constituents from wastewater such as the multivalent metallic ions responsible for hardness (i.e., calcium and magnesium). For this reason, NF is the preferred membrane

Table 11–28

Typical applications for membrane technologies in wastewater treatment[a]

Applications	Description
Microfiltration and Ultrafiltration	
Aerobic biological treatment	Membranes are used to separate the treated wastewater from the active biomass in an activated sludge process. The membrane separation unit can be internal (immersed in the bioreactor) or external to the bioreactor (see Fig. 8–2 in Chap. 8). Such processes are known as membrane bioreactor (MBR) processes
Anaerobic biological treatment	Membrane is used to separate the treated wastewater from the active biomass in an anaerobic complete-mix reactor
Membrane aeration biological treatment	Plate and frame, tubular, and hollow membranes are used to transfer pure oxygen to the biomass attached to the outside of the membrane. Such processes are known as membrane aeration bioreactor (MABR) processes
Membrane extraction biological treatment	Membranes are used to extract degradable organic molecules from inorganic constituents such as acids, bases, and salts from the waste stream for subsequent biological treatment [see Fig. 11–47(b)]. Such processes are known as extractive membrane bioreactor (EMBR) processes
Pretreatment for effective disinfection	Membranes are used to remove residual suspended solids from settled secondary effluent or from the effluent from depth or surface filters to achieve effective disinfection with either chlorine or UV radiation for reuse applications.
Pretreatment for nanofiltration and reverse osmosis	Microfilters are used to remove residual colloidal and suspended solids as a pretreatment step for additional processing
Nanofiltration	
Effluent reuse	Used to treat prefiltered effluent (typically with microfiltration) for indirect potable reuse applications such as groundwater injection. Credit is also given for disinfection when using nanofiltration
Wastewater softening	Used to reduce the concentration of multivalent ions contributing to hardness for specific reuse applications
Reverse osmosis	
Effluent reuse	Used to treat prefiltered effluent (typically with microfiltration) for indirect potable reuse applications such as groundwater injection. Credit is also given for disinfection when using reverse osmosis
Effluent dispersal	Reverse osmosis processes have proven capable of removing sizable amounts of selected compounds such as NDMA.
Two-stage treatment for boiler use	Two-stages of reverse osmosis are used to produce water suitable for high pressure boilers.

[a] Adapted in part from Stephenson et al. (2000).

for membrane softening. The advantages of nanofiltration over lime softening include the production of a product water that meets the most stringent reuse water quality requirements. Because both inorganic and organic constituents and bacteria and viruses are removed, disinfection requirements are minimized. Typical operating and performance data are presented in Tables 11–30 and 11–33, respectively. Performance variability data are presented in Table 11–32.

Reverse Osmosis. Worldwide, reverse osmosis (RO) is used primarily for desalination (Voutchkov, 2013). In wastewater treatment, RO is used for the removal of dissolved constituents from wastewater, remaining after advanced treatment with depth filtration or

Table 11–29

Application of membrane technologies for the removal of specific constituents found in wastewater[a]

Constituent	Membrane technology MF	 UF	 NF	 RO	Comments
Biodegradable organics		✔	✔	✔	
Hardness			✔	✔	
Heavy metals			✔	✔	
Nitrate			✔	✔	
Priority organic pollutants		✔	✔	✔	
Synthetic organic compounds			✔	✔	
TDS			✔	✔	
TSS	✔	✔			TSS removed during pretreatment for NF and RO
Bacteria	✔[b]	✔[b]	✔	✔	Used for membrane disinfection. Removed as pretreatment for NF and RO with MF and UF
Protozoan cysts and oocysts and helminth ova	✔	✔	✔	✔	
Viruses			✔	✔	Used for membrane disinfection

[a] Specific removal rates will depend on the composition and constituent concentrations in the treated wastewater.

[b] Variable performance, depending on the membrane nominal pore size and operating conditions.

microfiltration. The membranes exclude ions, but require high pressures to produce deionized water. The Orange County Water District flow diagram involving the use of reverse osmosis for the production of potable water for groundwater recharge is shown on Fig. 11–41. Typical operating information for reverse osmosis used for wastewater including operating pressures and flux rate rates is reported in Table 11–30. Corresponding performance data and variability data are presented in Tables 11–34 and 11–32. As noted above, care should be used in applying the performance data reported in Table 11–34 as it has been found that the performance of RO is also site specific, especially with respect to fouling (see Table 11–25). Important process design considerations for NF and RO membranes are reported in Table 11–35.

Depending on the level of dissolved solids removal, the product water from NF and RO processes may be corrosive to equipment and piping. Typical postreatments will involve the addition of chemical to adjust the stability of the treated water, in some reuse applications, the removal or addition of gases (see Fig. 11–41), and the addition of chemicals to meet disinfection requirements and to control the growth of microorganisms in pipelines. The types and use of chemicals to stabilize NF and RO product water are considered in Chap. 6. In some cases, blending with other waters, especially in potable reuse applications, may be appropriate.

Membrane Energy Consumption. Typical product recovery and energy consumption values for various membrane systems are presented in Table 11–30 for the processing of wastewater. Corresponding data for seawater desalination are also given for the

Table 11–30

Typical operating characteristics of membrane technologies used in wastewater treatment applications and for desalination[a]

		Operating pressure[c]		Rate of flux		Energy consumption[c]	
Membrane technology	**Product recovery[b], %**	**lb/in.^{2}**	**kPa**	**gal/ft^2·d**	**L/m^2·h**	**kWh/10^3 gal**	**kWh/m^3**
Wastewater with TDS from 800 to 1200 mg/L							
Microfiltration (vacuum type)	85–95	−3–−14	−28–−100	15–25	25–42	0.75–1.1	0.2–0.3
Microfiltration (pressure type)	85–95	5–30	34–200	24–35	40–60	0.75–1.1	0.2–0.3
Ultrafiltration	85–95	10–35	68–350	24–35	40–60	0.75–1.1	0.2–0.3
Nanofiltration	85–90	100–200	700–1400	8–12	14–20	1.5–1.9	0.4–0.5
Reverse osmosis (without energy recovery)	80–85	125–230	800–1900	8–12	14–20	1.9–2.5	0.5–0.65
Reverse osmosis (with energy recovery)[d]	80–85	125–230	800–1900	8–12	14–20	1.7–2.3	0.46–0.6
Electrodialysis	75–95			20–25	33–42	4.2–8.4	1.1–2.2
Seawater with TDS of about 35,000 mg/L							
Ultrafiltration (pretreatment)	85–95	10–35	68–350	24–47	40–80	0.75–1.1	0.2–0.3
Reverse osmosis (without energy recovery)[e]	30–55	700–1000	4800–6900	8–12	14–20	34–45	9–12
Reverse osmosis (with turbine/pump energy recovery)	30–55	700–1000	4800–6900	8–12	14–20	19–26	5–7
Reverse osmosis (with pressure exchange energy recovery)	30–55	700–1000	4800–6900	8–12[f]	14–20	9.5–15	~2.5–4

[a] Adapted in part from Patel (2013), Voutchkov (2013), Wetterau (2013).

[b] Cross-flow mode [see Fig, 11–33(a)]. In dead end mode [see Fig, 11–33(c)] all of the water passes through the membrane.

[c] The operating pressure and energy consumption will vary with the influent water quality and temperature of the feed water.

[d] Overall total energy reduction will vary from 6 to 12 percent, depending on the energy recovery device (ERD) and process configuration.

[e] At 50 percent recovery, the minimum theoretical energy required is 1.06 kWh/m^3; the corresponding practical limit is about 1.56 kWh/m^3 (Elimelech and Phillip, 2007).

[f] Flux rate with open intake is in the range from 12–17 L/m^2·h (7–10 gal/ft^2·d).

Note:
kPa × 0.1450 = lb$_f$/in.^{2}
L/m^2·h × 0.5890 = gal/ft^2·d
kWh/m^3 × 3.785 = kWh/10^3 gal
Bar = 100 kPa.

Table 11–31

Expected performance of microfiltration and ultrafiltration membranes on secondary effluent

		Value	
Constituent	**Rejection**	**Microfiltration**	**Ultrafiltration**
TOC	%	45–65	50–75
BOD	%	75–90	80–90
COD	%	70–85	75–90
TSS	%	95–98	96–99.9
TDS	%	0–2	0–2
NH_3-N	%	5–15	5–15
NO_3-N	%	0–2	0–2
PO_4^-	%	0–2	0–2
SO_4^{2-}	%	0–1	0–1
Cl^-	%	0–1	0–1
Total coliform[a]	log	2–5	3–6
Fecal coliform[a]	log	2–5	3–6
Protozoa[a]	log	2–5	>6
Viruses[a]	log	0–2	2–7[b]

[a] The reported values reflect observed practice and integrity concerns (see Example 8–4 in Chap. 8) and also a wide range of performance differences between membranes, as given in following footnote.

[b] The low and corresponding mean removal values for four different UF membranes treating the same water were 2.5, 4.0, 5.3, and 6.1 and 3.8, 5.0, 6.5, and 7.5, respectively (Sakaji, R. H., 2006).

purpose of comparison. The impact of water quality on energy consumption can be seen by comparing the energy requirement for wastewater with TDS ~1000 mg/L (~0.6 kWh/m^3) versus seawater with TDS ~35,000 mg/L (~10.5 kWh/m^3, without energy recovery). The importance of energy recovery, as discussed below, especially in seawater desalination, is also clearly evident.

In reviewing the information presented in Table 11–30, it is important to note that the reported operating pressure values for all of the membrane processes are considerably lower than comparable values of ten years ago. It is anticipated that operating pressures will continue to go down as new membranes and operating techniques are developed, but at a considerably lower rate as compared to the last ten years. At the present time, where the use of membranes is being considered, special attention must be devoted to the characteristics of the wastewater to be processed.

Energy Recovery from Nanofiltration and Reverse Osmosis. Because NF and RO in particular produce a high-pressure retentate flowrate , especially in seawater desalination, various methods have been developed or are under development to recover the energy lost in depressurizing the retentate flowrate. Energy recovery devices (ERDs) are designed to recover energy from the retentate flowrate and transfer it to the feed water to reduce the overall process energy (see Fig. 11–42 on page 1213). Typical devices that have been used operate on the following principles.

- Reverse running pumps
- Pelton wheel turbines

Table 11–32

Typical effluent quality variability observed with processes used for the removal of from dissolved constituents from reclaimed wastewater

Particulate removal process	Unit	Range of effluent values[a]	Geometric standard deviation, s_g[b]	
			Range	Typical
Microfiltration				
Turbidity	NTU	0.1–0.4	1.1–1.4	1.3
TSS	mg/L	0–1	1.3–1.9	1.5
Ultrafiltration				
Turbidity	NTU	0.1–0.4	1.1–1.4	1.3
TSS	mg/L	0–1	1.3–1.9	1.5
Nanofiltration				
TDS	mg/L	50–100	1.3–1.5	1.4
TOC	mg/L	1–5	1.2–1.4	1.5
Turbidity	NTU	0.01–0.1	1.5–2.0	1.75
Reverse osmosis[c]				
TDS	mg/L	25–50	1.3–1.8	1.6
TOC	mg/L	0.1–1	1.2–2.0	1.8
Turbidity	NTU	0.01–0.1	1.2–2.2	1.8
Electrodialysis				
TDS	mg/L	na	1.2–1.75	1.5

[a] Typical effluent values are not given for the processes because they will vary widely and depend on the operating conditions and water quality requirements.

[b] s_g = geometric standard deviation; $s_g = P_{84.1}/P_{50}$.

[c] Because measured effluent values are typically near the constituent detection limits, the error in the detection method can contribute to the observed effluent variability.

- Hydraulic turbocharger
- Isobaric energy recovery – piston type
- Isobaric energy recovery – rotary type
- Pressure amplifying pump

Pumps, Turbines, and Hydraulic Turbochargers. The reverse running pumps (i.e., Francis turbines), Pelton wheel turbines, and hydraulic turbochargers are adaptations of well know hydraulic machinery applied for the recovery of energy from NF and RO installations. Functionally, as shown on Fig. 11–42(a), the Pelton wheel turbine shaft is coupled to the motor used to drive the pump that pressurizes the feed water. The operation of the hydraulic turbocharger is similar to the Pelton turbine, with the exception that a pump impeller is mounted on the same shaft as the turbine, and a motor is not used.

Isobaric Devices. Isobaric energy recovery devices (also known as flow work exchangers) [see Fig. 11–42(b)] utilize the principles of positive displacement and isobaric

Table 11–33

Typical rejection rates for NF and "loose" RO membranes used to treat wastewater

Constituent	Unit	Rejection rate Nanofiltration	Rejection rate Loose RO
Total dissolved solids	%	40–60	
Total organic carbon	%	90–98	
Color	%	90–96	
Hardness	%	80–85	
Sodium chloride	%	10–50	70–95
Sodium sulfate	%	80–95	80–95
Calcium chloride	%	10–50	80–95
Magnesium sulfate	%	80–95	95–98
Nitrate	%	80–85	85–90
Fluoride	%	10–50	
Arsenic (+5)	%	<40	
Atrazine	%	85–90	
Proteins	log	3–5	3–5
Bacteria[b]	log	3–6	3–6
Protozoa[b]	log	> 6	> 6
Viruses[b]	log	3–5	3–5

[a] Adapted in part from www.gewater.com and Wong (2003).

[b] Theoretically all microorganisms should be removed. The reported values reflect integrity concerns (see Example 8–4 in Chap. 8).

chambers to transfer energy from a high pressure stream (the brine stream in the case of NF and RO) to a low pressure incoming feed water (Stover, 2007). The dual work exchange energy recovery (DWEER)® is an alternating piston driven device with two isobaric chambers. The PX® technology combines an isobaric positive displacement device with a centrifugal ERD; the transfer of energy is accomplished without the use of a piston. Isobaric ERDs, because of their ease of operation and flexibility, are used extensively throughout the world in desalination installations and are replacing and/or reducing the use of centrifugal type ERDs.

Performance of Energy Recovery Devices. Because of the relatively low feed water pressures used in the RO treatment of wastewater, energy recovery efficiency is relatively low. Typical recoveries for Pelton wheels and isobaric devices vary from 25 to 45 and 45 to 65 percent, respectively. The overall process energy reduction is about 6 to 12 percent, depending on the device and process configuration. By comparison, in seawater desalination, recovery efficiencies as high as 95 percent have been achieved, depending on the ERD (Voutchkov, 2013). The overall total energy reductions that can be achieved in seawater desalination will vary from 30 to 75 percent. On a relatively small scale, the Clark® pressure amplifying pump employs two opposing cylinders and pistons on a single rod to pressurize the feed water in conjunction with a small feed pump.

Table 11–34

Typical performance for reverse osmosis treatment[a]

Constituent	Unit	Rejection rate
Total dissolved solids	%	90–98
Total organic carbon	%	90–98
Color	%	90–96
Hardness	%	90–98
Sodium chloride	%	90–99
Sodium sulfate	%	90–99
Calcium chloride	%	90–99
Magnesium sulfate	%	95–99
Nitrate	%	84–96
Fluoride	%	90–98
Arsenic (V)	%	85–95
Atrazine	%	90–96
Proteins	log	4–7
Bacteria[b]	log	4–7
Protozoa[b]	log	>7
Viruses[b]	log	4–7

[a] Adapted in part from www.gewater.com and Wong (2003).

[b] Theoretically all microorganisms should be removed. The reported values reflect integrity concerns (see Example 8–4 in Chap. 8).

Forward Osmosis: An Emerging Membrane Technology

The membrane processes discussed in this section are based on reversing the natural osmosis process through the addition of a driving force greater than the osmotic pressure to produce purified water. Although not used commonly for water purification, a number of alternative processes based on the utilization of the osmotic pressure are under development. Processes that utilize the natural osmotic pressure [see Fig. 11–43(a)] are termed forward osmosis (FO) or direct osmosis (DO). In the FO process, illustrated on Fig. 11–43(b), water from the feed solution permeates through the membrane to dilute a more concentrated solution, known by a variety of names including draw solution, osmotic agent, and driving agent. The draw solution is the name used most commonly for this solution.

The principal requirement for the draw solution is that its osmotic pressure must be greater than that of the feed solution. Another requirement for the draw solution is that it must be easy to reconcentrate after being diluted by the water from the feed solution. A solution of sodium chloride (NaCl) has been used as the draw solution because it can be reconstituted easily by reverse osmosis without the problems associated with scaling. A draw solution comprised of multivalent ions has been used where high rejection is required. Based on the results of preliminary testing, advantages of the FO process include minimal pressure requirements and high rejection for a variety of constituents. The FO process may also result in less membrane clogging, but more research is needed to define the controlling conditions that will minimize clogging.

Table 11–35

Process design considerations for NF and RO[a]

Design consideration	Discussion
Feed water characterization	Complete characterization of the feed water is essential for identifying constituents that produce a high potential for membrane fouling. The effect of residual suspended solids in the influent to the membranes especially should be evaluated
Pretreatment	Pretreatment must be evaluated to extend membrane life, and issues such as flow equalization, pH control, chemical treatment, and residual solids removal should be considered
Flux rate	Flux rate influences system costs by establishing the filter area, affecting polarization control, and affecting membrane life
Recovery	Recovery rate affects solute rejection, membrane performance, and brine generation volumes
Membrane fouling	Parameters should be developed based on pilot plant testing. Acid, antiscalants, and biocides are used to control membrane fouling, as are staging and operational conditions
Membrane cleaning	Cleaning procedures and frequency need to be established
Membrane life	The principal economic consideration that governs successful application of membrane technology
Operating and maintenance costs	High pressure systems require significant energy costs, high capital costs for high pressure pumps, and high maintenance costs associated with equipment wear. After membrane replacement, energy is the next major operating expense
Recycle flows	Provisions for recycling a portion of the product water should be included as an operating consideration to control membrane velocity, influent concentration, and equalizing influent flow variations
Retentate and backwash disposal	Retentate and backwash characteristics need to be considered especially, if chemicals are used in pretreatment or membrane cleaning and large volumes of waste require disposal

[a] Adapted in part from Celenza (2000).

For example, potential applications in the wastewater management field include concentration of dilute industrial wastewater, concentration of RO brines, concentration of digester supernatant, and the concentration of the return flows from sludge thickening processes. By reducing the volume of RO brines that must be processed,

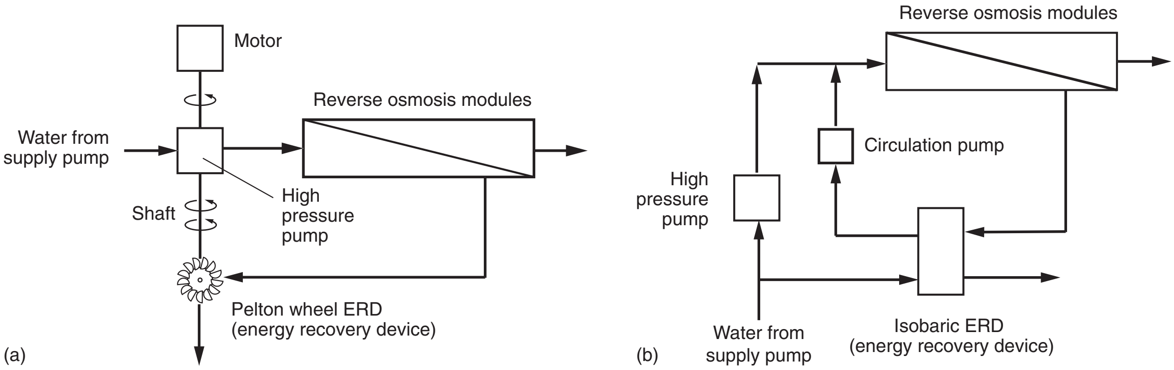
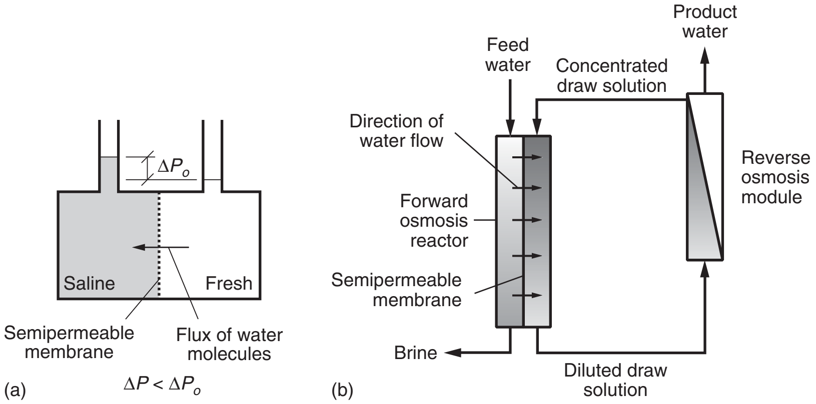


Figure 11–42

Application of energy recovery devices in conjunction with reverse osmosis: (a) Pelton wheel and (b) isobaric piston type.

Figure 11–43

Application of forward osmosis: (a) definition sketch for forward osmosis (the differential pressure between solutions is less than the osmotic pressure) and (b) flow diagram for the application of forward osmosis. Note: the draw solution is far more concentrated than the feed water.



technologies such as evaporation and crystallization become technically and economically feasible. An excellent review article has been prepared by Cath and his associates (Cath et al., 2006).

Pilot-Plant Studies for Membrane Applications

Because every wastewater is unique with respect to its constituent characteristics, it is difficult to predict *a priori* how a given membrane process will perform. As a result, the selection of the best membrane for a given application is based usually on the results of pilot studies. Membrane fouling indexes (see Table 11–26) can be used to assess the need for pretreatment. In some situations, manufacturers of membranes will provide a testing service to identify the most appropriate membrane for a specific feed water. Typical pilot scale facilities used to evaluate the performance of NF and RO treatment processes are shown on Fig. 11–44.

The elements that comprise a pilot plant include (1) the pretreatment system; (2) tankage for flow equalization and cleaning; (3) pumps for pressurizing the membrane, recirculation, and backflushing with appropriate controls; (4) the membrane test module; (5) facilities for monitoring the performance of the test module; and an appropriate membrane backflushing system. Typical membrane operating parameters and water quality

Figure 11–44

Views of membrane pilot plant test units: (a) ultrafiltration and (b) reverse osmosis. Note the membrane modules being tested are full scale.

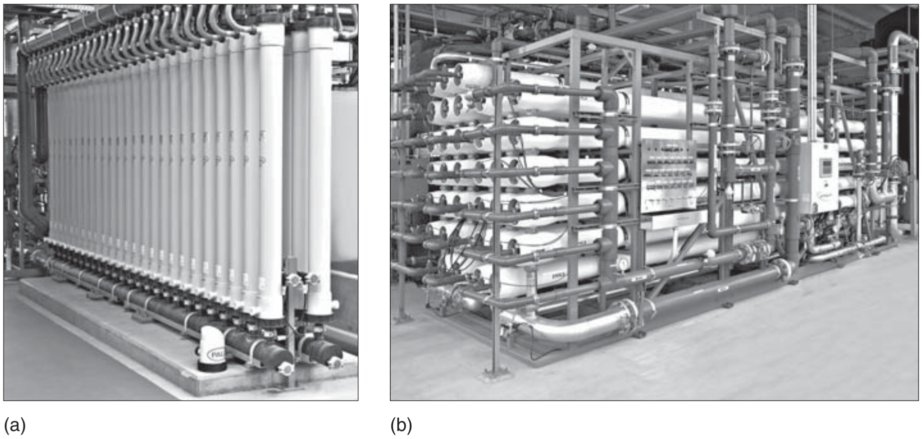

(a) (b)

Table 11–36

Typical operating parameters and water quality measurements used for pilot testing membrane facilities[a]

Membrane operating parameters	
Pretreatment requirements including chemical dosages	
Transmembrane flux rate correlated to operating time	
Transmembrane pressure	
Recovery	
Washwater requirements	
Recirculation ratio	
Cleaning frequency including protocol and chemical requirements	
Posttreatment requirements	
Typical water quality measurements	
Turbidity	Heterotrophic plate count
Particle counts	Other bacterial indicators
Total organic carbon	Specific constituents that can limit recovery such as silica, barium, calcium, fluoride, strontium, and sulfate
Nutrients	
Heavy metals	
Organic priority pollutants	Biotoxicity
Total dissolved solids	Fouling indexes
pH	
Temperature	

[a] Tchobanoglous et al. (2003).

measurements are presented in Table 11–36. Additional specific parameters selected for evaluation will depend on the final use of the product water.

Management of Retentate

Management of the retentate produced by membrane processes represents the major problem that must be dealt with in their applications. Methods that can be used for the treatment and disposal of the retentate are reported in Table 11–37. While small facilities can dispose of small quantities of retentate by blending with other wastewater flows, this approach is not suitable for large facilities. The retentate from NF and RO facilities will contain hardness, heavy metals, high molecular weight organics, microorganisms, and often hydrogen sulfide gas. The pH is usually high due to the concentration of alkalinity, which increases the likelihood of metal precipitation in disposal wells. As a result, most of the large-scale desalination facilities are located along coastal regions, both in the United States and in other parts of the world. For inland locations, long transmission lines to coastal regions are being considered. While controlled evaporation is technically feasible, because of high operating and maintenance costs, this approach is used where no other alternatives are available and the value of product water is high. The quality and quantity of the concentrated retentate produced from nanofiltration, reverse osmosis and electrodialysis can be estimated using simplified recovery and rejection computations as illustrated in Example 11–8.

Table 11–37

Treatment methods and disposal options for concentrated brine solutions from membrane processes

Disposal option	Description
Treatment options	
Concentration by using multistage membrane arrays	Concentration of brine stream
Falling film evaporators	Thicken and concentrate brine streams
Crystallizers	Concentration of brine stream into a crystalized form for processing or disposal
Forward osmosis	Concentration of brine stream
Membrane distillation	Concentration of brine stream
Solar evaporators	Thicken and concentrate brine streams
Spray dryers	Concentration of brine stream
Vapor compression evaporators	Concentration of brine stream
Disposal options	
Deep well injection	Depends on whether subsurface aquifer is brackish water or is otherwise unsuitable for domestic uses.
Discharge to wastewater collection system	This option is only suitable for very small discharges such that the increase in TDS is not significant (e.g., less than 20 mg/L).
Evaporation ponds	Large surface area required in most areas with the exception of some southern and western states.
Land application	Land application has been used for some low concentration brine solutions.
Ocean discharge	The disposal option of choice for facilities located in the coastal regions of the United States. Typically, a brine line, with a deep ocean discharge, is used by a number of dischargers. Combined discharge with power plant cooling water has been used in Florida. For inland locations, truck, rail hauling or pipeline is needed for transportation.
Surface water discharge	Discharge of brines to surface waters is the most common method of disposal for concentrated brine solutions.

EXAMPLE 11–8 **Estimate Quantity and Quality of Waste Streams from a Reverse Osmosis Facility** Estimate quantity and quality of the retentate and the total quantity of water that must be processed, from a reverse osmosis facility that is to produce 4000 m^3/d of water to be used for industrial cooling operations. Assume that both the recovery and rejection rates are equal to 90 percent and that the TDS concentration of the feed steam is 400 mg/L.

Solution

1. Determine the flowrate of the concentrated retentate and the total amount of water that must be processed.
   a. Combining Eqs. (11–36) and (11–40) results in the following expression for the retentate flowrate.

$$Q_r = \frac{Q_p(1 - r)}{r}$$

b. Determine the retentate flowrate.

$$Q_r = \frac{(4000\ \text{m}^3/\text{d})(1 - 0.9)}{0.9} = 444\ \text{m}^3/\text{d}$$

c. Determine the total amount of water that must be processed to produce 4000 m^3/d of RO water. Using Eq. (11–43) the required amount of water is

$$Q_f = Q_p + Q_r = 4000\ \text{m}^3/\text{d} + 444\ \text{m}^3/\text{d} = 4444\ \text{m}^3/\text{d}$$

2. Determine the concentration of the permeate. The permeate concentration is obtained by writing Eq. (11–41) in decimal form as follows:

$$C_p = C_f(1 - R) = 400\ \text{mg/L}\ (1 - 0.9) = 40\ \text{mg/L}$$

3. Determine the concentration of the retentate. The required value is obtained by solving Eq. (11–44).

$$C_r = \frac{Q_f\,C_f - Q_p\,C_p}{Q_r}$$

$$C_r = \frac{(4444\ \text{m}^3/\text{d})(400\ \text{mg/L})C_f - (4000\ \text{m}^3/\text{d})(40\ \text{mg/L})}{(444\ \text{m}^3/\text{d})}$$

$$C_r = 3643\ \text{mg/L}$$

Comment A variety of concentration methods are currently under investigation to reduce the volume of the retentate that must be treated.

11–8 ELECTRODIALYSIS

Electrodialysis (ED) is an electrochemical separation process in which mineral salts and other ionic species are transported through ion-selective membranes from one solution to another under the driving force of a direct current (DC) electric potential. As compared to NF and RO, which transport pure water through the membrane leaving the salts behind, with ED salt is gradually stripped from solution leaving a dilute solution behind containing particulate matter and neutral species not removed by the ED process. The salt transferred through the membrane then forms the concentrate. A typical flow diagram employing electrodialysis for the control of dissolved solids is shown on Fig. 11–45.

Description of the Electrodialysis Process

The key to the ED process is the ion selective membranes that are essentially ion exchange resins cast in sheet form. Ion exchange membranes that allow passage of positively charged ions such as sodium and potassium are called cation membranes. Membranes that allow passage of negatively charged ions such as chloride and phosphate are called anion membranes. To demineralize a solution using ED, cation and anion membranes are arranged alternately between plastic spacers in a stacked configuration with a positive electrode (anode) at one end and a negative electrode (cathode) at the other (see Fig. 11–46). When a DC voltage is applied, the electrical potential created becomes the driving force to move ions, with the membranes forming barriers to the ions of opposite charge. Therefore, anions attempting to migrate to the anode will pass through the adjacent anion membrane but will be stopped by the first cation membrane they encounter. Cations trying to

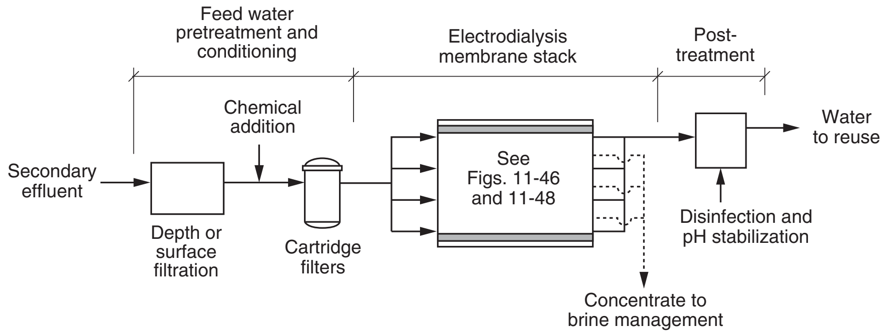


Figure 11–45

Typical process flow diagram employing electrodialysis for the removal of total dissolved solids (TDS) from secondary effluent.

migrate to the cathode will pass through the cation membrane but will be stopped by the anion membrane. The membranes, therefore, form ion diluting compartments and ion concentrating compartments (www.gewater.com).

An ED assembly, known as a *stack*, consists of multiple cell pairs located between an anode and a cathode. A set of adjacent components consisting of a diluting compartment spacer, an anion membrane, a concentrating compartment spacer, and a cation membrane is called a *cell pair*. Electrolysis stacks can contain as many as 600 cell pairs. Feed water (filtered wastewater) is pumped through the stack assembly. Typical flux rates are from 35 to 45 $L/m^2 \cdot h$. Dissolved solids removals vary with the (1) wastewater temperature, (2) amounts of electric current passed, (3) type and amount of ions, (4) permeability/selectivity of the membrane, (5) fouling and scaling potential of the feed water, (6) feed water flowrates, and (7) number and configuration of stages.

Electrodialysis Reversal

In the early 1970s, the electrodialysis reversal (EDR) process was introduced. An EDR unit operates on the same principle as ED technology, except that both the product and concentrate channels are identical in construction (see Fig. 11–47). The same membranes are used to provide a continuous self-cleaning ED process that uses periodic reversal of the DC polarity to allow systems to run at high recovery rates. Polarity reversal causes the concentrating

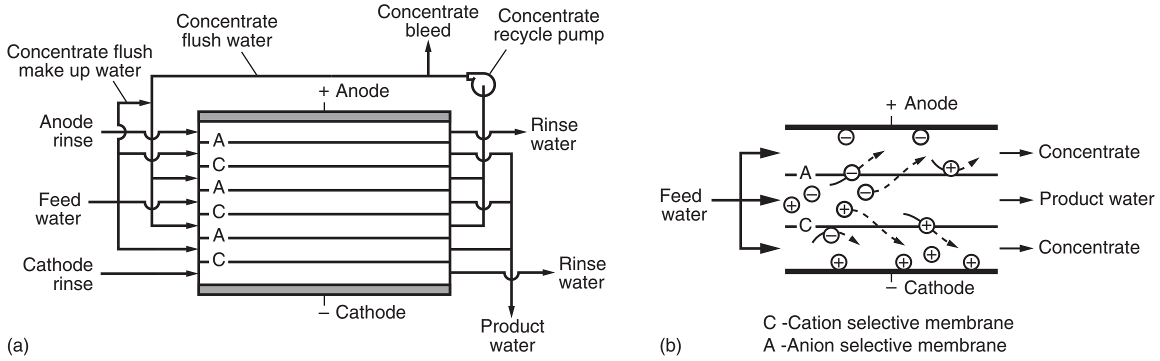
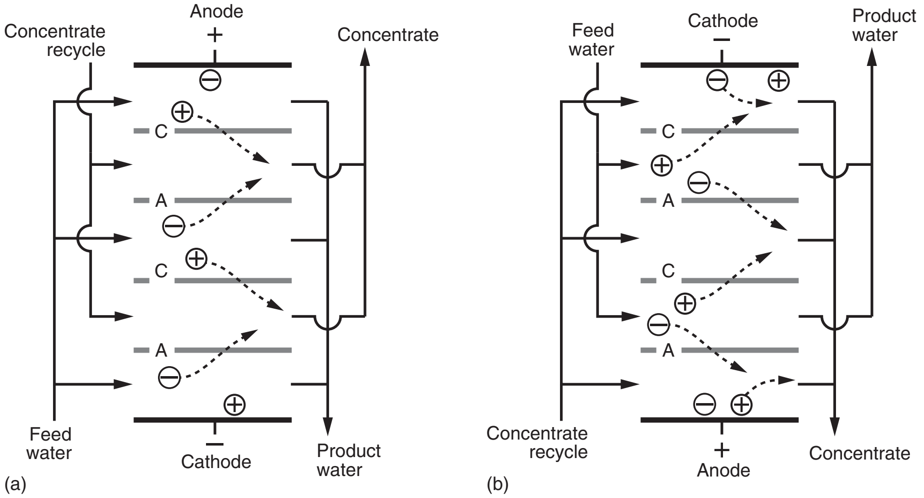


Figure 11–46

Conventional electrodialysis: (a) schematic of electrodialysis membrane stack with anode and cathode rinse and (b) schematic illustration of ion migration within the membrane stack. Note: The conventional electrodialysis process has been largely replaced the electrodialysis reversal (EDR) process (see Fig 11–47).

Figure 11–47

Schematic of electrodialysis reversal (EDR) process: (a) negative polarity and (b) positive polarity. Because the polarity is reversed, the anode and cathode rinse shown on Fig. 11–46 is not needed.



and diluting flow to switch after every cycle. Any fouling or scaling constituents are removed when the process reverses, sending fresh product water through the compartments filled previously with concentrated waste. The reversal process is useful in breaking up and flushing out scales, slimes, and other deposits in the cells before they buildup. Product water is not collected during a short interval of time following reversal.

EDR systems are able to reduce dissolved ions in feed water containing 10,000 to 12,000 mg/L of total dissolved solids, but because of energy requirements are ideally suited for the treatment of brackish water in the range from 800 to 5000 mg/L. As a rule of thumb, it takes about 1 to 1.2 kWh/m^3 to remove a kilogram of salt (see Table 11–38). Typical removal rates can range from 50 to 94 percent removal (www.gewater.com). A view of an EDR installation and an exposed membrane stack are shown on Fig. 11–48. The EDR facility shown on Fig. 11–48 is used to remove TDS from a portion (sidestream) of the reclaimed water produced at the North City plant in San Diego, CA. The treated water with a reduced TDS concentration is blended back into the main flow which has a TDS concentration that varies from 1200 to 1300 mg/L to produce a final reclaimed water with a TDS equal to or less than 1000 mg/L to meet contractual agreements with the users of the reclaimed water.

Table 11–38

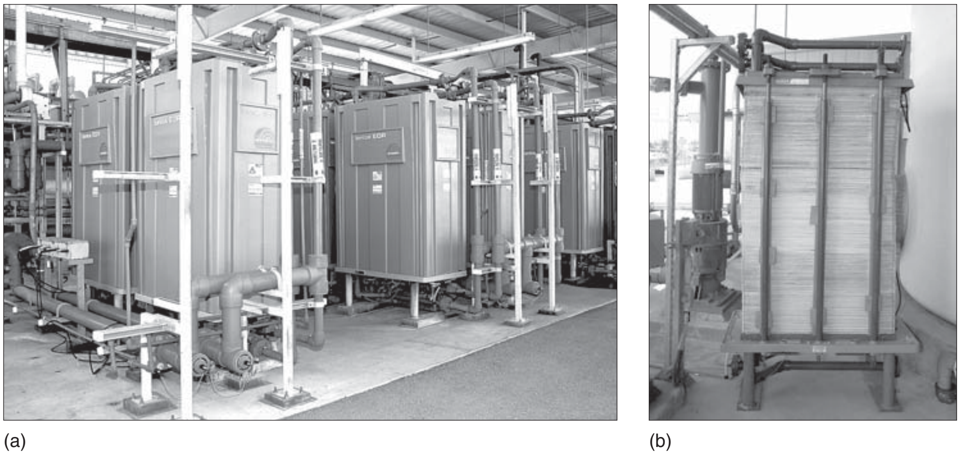
Typical operating parameters for electrodialysis units

Parameter	Unit	Range
Flux rate	m^3/m^2·d	0.8–1.0
Water recovery	%	75–90
Concentrate flowrate	% of feed	12–20
TDS removal	%	50–94
CD/N (current density to normality) ratio	(mA/cm^2)/(g-eq/L)	500–800
Membrane resistance, Ω	ohms	4–8
Current efficiency	%	85–95
Energy consumption[a]	kWh/m^3	1.5–2.6
Approximate energy per kg of salt removed	kWh/m^3·kg	1–1.2

[a] Based on treating reclaimed water with a TDS concentration in the range from 1000 to 2500 mg/L. Not recommended for TDS concentration values beyond 10,000 to 12,000 mg/L.

Figure 11–48

Electrodialysis reversal process used to remove TDS from reclaimed water at the North City plant in San Diego, CA: (a) view of full scale electrodialysis facility and (b) view of electrodialysis membrane stack with cover removed.



Power Consumption

The ED/EDR process uses electric power to transfer ions through the membranes and to pump water through the system. Two, or sometimes three, pumping stages are used typically.

Power Requirements for Ion Transfer. The current required for ED can be estimated using Faraday's laws of electrolysis. Because one Faraday of electricity will cause one gram equivalent of a substance to migrate from one electrode to another, the number of gram equivalents removed per unit time is given by:

$$\text{Gram-eq/unit time} = Q_p(N_{\text{inf}} - N_{\text{eff}}) = Q_p\Delta N = Q_pN_{\text{inf}}E_r \tag{11–48}$$

$$\text{where gram/eq} = \frac{\text{Mass of solute, g}}{\text{Equivalent weight of solute, g}}$$

Q_p = product water flowrate, L/s
N_{inf} = normality of influent (feed), g-eq/L
N_{eff} = normality of effluent (product), g-eq/L
ΔN = change in normality between the influent and effluent, g-eq/L
E_r = efficiency of salt removal, % (expressed as a decimal)

The corresponding expression for the current for a stack of membranes is given by:

$$i = \frac{FQ_p(N_{\text{inf}} - N_{\text{eff}})}{nE_c} = \frac{FQ_pN_{\text{inf}}E_r}{nE_c} \tag{11–49}$$

where i = current, A, ampere
F = Faraday's constant, 96,485 A·s/g-eq
n = number of cell pairs in the stack
E_c = current efficiency, % (expressed as a decimal)

In the analysis of the ED process, it has been found that the capacity of the membrane to pass an electrical current is related to the current density (CD) and the normality (N) of the feed solution. Current density is defined as the current in milliamperes that flows through a square centimeter of membrane perpendicular to the current direction. Normality corresponds to the concentration of a solution based on the number of gram equivalent weights of a solute per liter of solution. A solution containing one gram of equivalent weight per liter is referred to as *one normal* (1 N). The relationship between current density and the solution normality is known as the *current density to normality* (CD/N) ratio.

High values of the CD/N ratio are indicative that there is insufficient charge to carry the current. When high ratios exist, a localized deficiency of ions may occur on the surface of the membrane, causing a condition called *polarization*. Polarization should be avoided as it results in high electrical resistance leading to excessive power consumption. In practice, CD/N ratios will vary from 500 to 800 when the current density is expressed as mA/cm^2. The resistance of an ED unit used to treat a particular water must be determined experimentally. Once the resistance, R, and the current flow, i, are known, the power required can be computed using Ohm's law as follows:

$$P = E \times i = R(i)^2 \qquad (11\text{–}50)$$

where P = power, W
E = voltage, V
$= R \times i$
R = resistance, Ω
i = current, A

The application of the above relationships is considered in Example 11–9.

EXAMPLE 11–9 **Determine Power Requirements and Membrane Area for ED Treatment of Reclaimed Water** Determine the power and area required to reduce the TDS content of 4000 m^3/d of treated wastewater to be used for industrial cooling water. Assume the following data apply.

1. Number of cell pairs in stack = 500
2. Influent TDS concentration = 2500 mg/L (~ 0.05 g-eq/L)
3. TDS removal efficiency, E_r = 50%
4. Product water flowrate = 90% of feed water
5. Current efficiency, E_c = 90%
6. CD/N ratio = (500 mA/cm^2)/(g-eq/L)
7. Resistance = 5.0 Ω

Solution

1. Calculate the current using Eq. (11–49).

$$i = \frac{FQ_pN_{\text{inf}}E_r}{nE_c}$$

$$Q_p = (4000\text{ m}^3/\text{d})(10^3\text{ L}/1\text{ m}^3)/(86{,}400\text{ s/d}) = 46.3\text{ L/s}$$

$$i = \frac{(96{,}485\text{ A}\cdot\text{s/g-eq})(46.3\text{ L/s})(0.05\text{ g-eq/L})(0.5)}{(500)(0.90)}$$

$$i = 248\text{ A}$$

2. Determine the power required using Eq. (11–50).

$$P = R(i)^2$$
$$P = (5.0\ \Omega)(248\text{ A})^2 = 307{,}520\text{ W} = 308\text{ kW}$$

3. Determine the power requirement per m^3 of treated water.

$$\text{Power consumption} = \frac{(308\text{ kW})(24\text{ h/d})}{(4000\text{ m}^3/\text{d})(0.9)} = 2.05\text{ kWh/m}^3$$

4. Determine the required surface area per cell pair. The area is given by

$$A = \frac{i\text{, current}}{CD\text{, current density}}$$

a. Determine the current density from the CN/N ratio:

$CD = [(500\ \text{mA/cm}^2)/(\text{g-eq/L})](0.05\ \text{g-eq/L}) = 25\ \text{mA/cm}^2$

b. The required area is:

$$\text{Area} = \frac{i}{CD} = \frac{(248\ \text{A})(1000\ \text{mA/A})}{(25\ \text{mA/cm}^2)} = 9920\ \text{cm}^2 = 0.99\ \text{m}^2$$

Comment The actual performance will have to be determined from pilot tests. The computed value for the power required per unit volume, 2.05 kWh/m^3, is within the range of values reported in Table 11–38 (1.1 to 2.6 kWh/m^3) for water with 1000 to 2500 mg/L TDS.

Power Requirements for Pumping. For pumping, the power requirements depend on the concentrate recirculation rate, the need for both product and waste pumping for discharge, and the efficiency of the pumping equipment (USBR, 2003).

Operating Considerations

The ED process may be operated in either a continuous or a batch mode. The units can be arranged either in parallel to provide the necessary hydraulic capacity or in series to obtain the desired degree of demineralization. A typical three-stage, two-line ED flow diagram is shown on Fig. 11–49. The ED process should be protected from particulate fouling by a 10 micron cartridge filter [see Table 11–21(g) and Fig. 11–45].

A single electrodialysis stack can remove from 25 to 60 percent of the TDS, depending on the feed water characteristics. Further desalting requires that two or more stacks be used in series (USBR, 2003). A portion of the resulting concentrate is recycled to improve system performance. Makeup water, usually about 10 percent of the feed volume, is required to wash the membranes continuously. A portion of the concentrate flowrate is recycled to maintain nearly equal flowrates and pressures on both sides of each membrane. Typical operating parameters for the electrodialysis process are reported in Table 11–38.

Operating Issues. Problems associated with the ED process for wastewater treatment include chemical precipitation of salts with low solubility on the membrane surface and clogging of the membrane by the residual colloidal organic matter in wastewater treatment plant effluents. To reduce membrane fouling, some form of filtration may be necessary. With a properly designed plant, membrane cleaning should be infrequent. However, for both ED and EDR systems, clean-in-place (CIP) systems are provided normally to

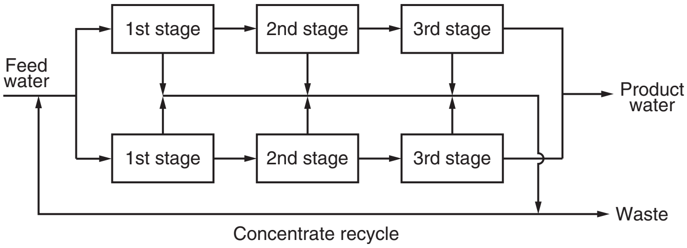


Figure 11–49

Schematic diagram for a three-stage, two-line electrodialysis process.

circulate either hydrochloric acid solution for mineral scale resolution or sodium chloride solution with pH adjustment for organics removal (USBR, 2003).

Membrane and Electrode Life. Membranes for ED and EDR applications have a life of about 10 years before they are replaced. Effective and timely cleaning-in place extends the membrane life and improves product quality and power consumption. Cation membranes typically last longer than anion membranes because anion membranes are particularly susceptible to oxidation by chlorine and other strong oxidants (USBR, 2003). With the development of the EDR process and new electrode design, the life of anode and cathode electrodes is typically 2 to 3 years. Anode life is typically less than cathode life. Electrodes can be reconditioned (USBR, 2003).

Electrodialysis Versus Reverse Osmosis

In a recently completed study, two advanced treatment processes were compared to reduce the salinity of reclaimed water from a TDS concentration of 750 ± 50 mg/L to 500 mg/L or less (Adham et al., 2004). The two advanced treatment processes evaluated were (1) MF followed by RO and (2) EDR. The study was conducted for a period of about six months. Based on the results of the side-by-side testing, it was found that the EDR process with cartridge prefiltration was more cost effective than the combined MF/RO process. Some of the advantages and disadvantages cited for each advanced treatment process are reported in Table 11–39. As more of the potential applications of EDR are currently under investigation, the current literature should be consulted.

Table 11–39

Comparison of advantages and disadvantages of electrodialysis and reverse osmosis for desalination[a]

Advantages	Disadvantages
Electrodialysis (EDR)	
• Minimal pretreatment may be required (cartridge filtration is recommended) • Operates at a low pressure • Process is much quieter because high pressure pumps are not required • Antiscalant is not required • Membrane life expectancy is longer because foulants are removed continuously during the reversal process • Requires less maintenance than RO due to reversal process	• Limited to 50 percent salt rejection for a single membrane stack (stage) • Requires larger footprint to produce similar quantity and quality of water if multiple staging is used • Electrical safety requirements • Less experience for wastewater demineralization in the U.S. • Not as effective at removing microorganisms and many anthropogenic organic contaminants
Reverse osmosis	
• RO membranes provide a barrier to microorganisms and many anthropogenic organic contaminants (for the treated portion of the water produced) • More demonstrated experience for wastewater demineralization • RO membranes can remove more than 90 percent of TDS • Source water blending will reduce size of systems • Flexibility to provide higher quality water, if desired	• Requires high pressure to achieve high salt rejection • Requires pretreatment processes to minimize scaling and fouling • Requires chemical addition for MF & RO fouling control • More routine maintenance may be required to maintain performance

[a] Adapted from Adham et al. (2004).

11–9 ADSORPTION

In wastewater treatment adsorption is used for the removal of substances that are in solution by accumulation of those substances on a solid phase. Adsorption is considered to be a mass transfer operation as a constituent is transferred from a liquid phase to a solid phase (see Table 11–2). The *adsorbate* is the substance that is being removed from the liquid phase at the interface. The *adsorbent* is the solid, liquid, or gas phase onto which the adsorbate accumulates. Although adsorption is used at the air-liquid interface in the flotation process (see Sec. 5–8), only the case of adsorption at the liquid-solid interface is considered in this section. Activated carbon is the primary adsorbent used in adsorption processes. The basic concepts of adsorption are presented in this section along with elements of design and limitations of the adsorption process.

Applications for Adsorption

Adsorption treatment of wastewater is usually thought of as a polishing process for water that has already received normal biological treatment. Adsorption has been used for the removal of refractory organic constituents; residual inorganic constituents such as nitrogen, sulfides, and heavy metals; and odor compounds from wastewater. Under optimum conditions, it appears that adsorption can be used to reduce the effluent COD to less than 10 mg/L. In water reclamation applications adsorption is used for (1) the continuous removal of organics and (2) as a barrier against the breakthrough of organics from other unit processes. In some cases, adsorption is used for the control of precursors that may form toxic compounds during disinfection.

Representative compounds that are readily and poorly adsorbed onto activated carbon are listed in Table 11–40. As shown in Table 11–40, activated carbon is known to have a low adsorption affinity for low molecular weight polar organic compounds. If biological activity is low in the carbon contactor or in other biological unit processes, low molecular weight polar organic compounds may be difficult to remove with activated carbon.

Types of Adsorbents

Treatment with adsorbent materials involves either (1) passing a liquid to be treated through a bed of adsorbent material held in a reactor/contactor (either fixed or fluidized) or (2) blending the adsorbent material into a unit process followed by sedimentation or filtration for removal of the spent adsorbent. The principal types of adsorbents include activated carbon, granular ferric hydroxide (GFH), and activated alumina. Carbon-based adsorbents are used most commonly for wastewater adsorption because of their relatively low cost. Other adsorbents that may prove to be effective with further research include manganese greensand, manganese dioxide, hydrous iron oxide particles, and iron oxide coated sand. Regardless of the adsorbent selected for a particular application, pilot testing will be necessary for determination of process performance and design parameters. The characteristics of materials used for adsorption are summarized in Table 11–41.

Activated Carbon. Activated carbon is derived by subjecting an organic base material, such as wood, coal, almond, coconut, or walnut hulls to a pyrolysis process followed with activation by exposure to oxidizing gases such as steam and CO_2 at high temperatures. The resulting carbon structure is porous, as illustrated on Fig. 11–50, on page 1226, with a large internal surface area. The resulting pore sizes are defined as follows:

Macropores > 500 nm
Mesopores > 20 nm and < 500 nm
Micropores < 20 nm

Table 11–40

Readily and poorly adsorbed organics on activated carbon[a]

Readily adsorbed organics	Poorly adsorbed organics
Aromatic solvents	Low-molecular weight ketones, acids, and aldehydes
Benzene	Sugars and starches
Toluene	Very high molecular weight or colloidal organics
Nitrobenzenes	Low–molecular weight aliphatics
Chlorinated aromatics	
PCBs	
Chlorophenols	
Polynuclear aromatics	
Acenaphthene	
Benzopyrenes	
Pesticides and herbicides	
DDT	
Aldrin	
Chlordane	
Atrazine	
Chlorinated non-aromatics	
Carbon tetrachloride	
Chloroalkyl ethers	
Trichloroethene	
Chloroform	
Bromoform	
High–molecular weight hydrocarbons	
Dyes	
Gasoline	
Amines	
Humics	

[a] From Froelich (1978).

The surface properties, pore size distribution, and regeneration characteristics that result are a function of both the initial material used and the preparation procedure, therefore many variations are possible. The two size classifications of activated carbon are granular activated carbon (GAC), which has a diameter greater than 0.1 mm (~140 sieve) and is used in pressure or gravity filtration, and powdered activated carbon (PAC), which typically has a diameter of less than 0.074 mm (200 sieve) and is added directly to the activated sludge process or solids contact processes.

Granular Ferric Hydroxide. Granular ferric hydroxide (GFH) is manufactured from a ferric chloride solution by neutralization and precipitation with sodium hydroxide. The adsorption capacity of GFH depends on water quality parameters, including pH, temperature, and other constituents in the water. Constituents that have been removed using GFH include arsenic, chromium, selenium, copper, and other metals. The process performance is reduced by suspended solids and precipitated iron and manganese, and by

Table 11–41

Comparison of various adsorbent materials[a]

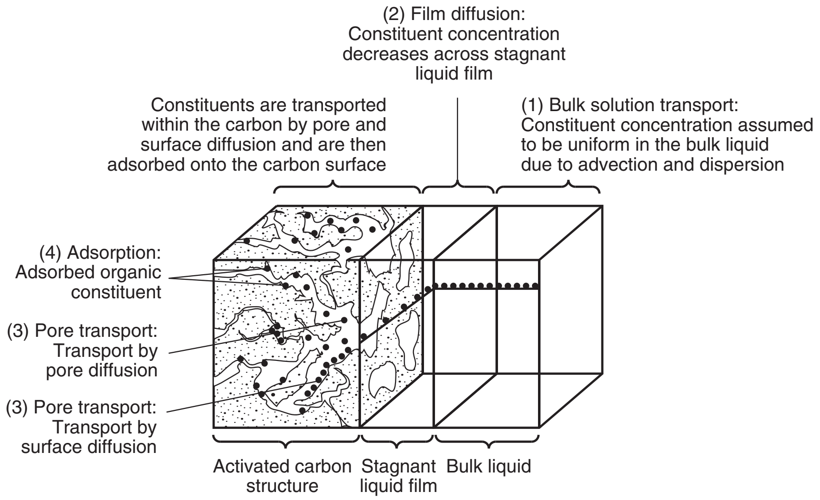
Parameter	Unit	Activated carbon Granular (GAC)	Activated carbon Powdered (PAC)	Activated alumina	Granular ferric hydroxide
Total surface area	m^2/g	700–1300	800–1800	280–380	250–300
Bulk density	kg/m^3	400–500	360–740	600–800	1200–1300
Particle density, wetted in water	kg/L	1.4–1.5	1.3–1.4	3.97	1.59
Particle size range	μm	100–2400	5–50	290–500	150–2000
Effective size	mm	0.6–0.9	na		
Uniformity coefficient	UC	≤ 1.9	na		
Mean pore radius	Â	16–30	20–40		
Iodine number		600–1100	800–1200		
Abrasion number	minimum	75–85	70–80		
Ash	%	≤ 10	≤ 6		
Moisture as packed	%	2–4	2–4		

[a] Specific values will depend on the source material used for the production of activated carbon.

constituents that compete for adsorption sites, including organic matter and other ions (e.g., phosphate, silicate, sulfate). While GFH adsorbents can be effective from a performance standpoint for removal of specific constituents (e.g., arsenic), the cost associated with GFH the GFH process is often prohibitive for large systems. The adsorption capacity of GFH media is reduced significantly following regeneration; thus after reaching capacity, GFH adsorbents are typically disposed of in a landfill and replaced with new media. However, because GFH is not regenerated, the costs associated with management of the waste regenerant can be avoided, making the process viable in some situations, especially where the waste regenerant must be handled as a hazardous waste.

Figure 11–50

Definition sketch for the adsorption of an organic constituent onto an activated carbon particle.



Activated Alumina. Activated alumina is derived from a naturally occurring mineral processed from bauxite that has been treated to remove molecules of water from its crystalline structure. Activated alumina is used in drinking water treatment for the removal of arsenic and fluoride (Clifford, 1999) and may have application in water reclamation for specific constituents. Activated alumina can be regenerated with a strong-base followed by a strong-acid. The regeneration of activated alumina and subsequent waste management issues result in significant operation and maintenance costs. As mentioned for GFH, pH (best performance at pH of 5.5 to 6), temperature, and competing constituents will affect the performance of activated alumina adsorption. The use of powdered activated alumina coupled with membranes (microfiltration and ultrafiltration) may also be a promising treatment process.

Fundamentals of Adsorption Processes

The adsorption process, as illustrated on Fig. 11–50, takes place in four, more or less definable steps: (1) bulk solution transport, (2) film diffusion transport, (3) pore and surface transport, and (4) adsorption (or sorption). The adsorption step involves the attachment of the material to be adsorbed to the adsorbent at an available adsorption site (Snoeyink and Summers, 1999). Additional details on the physical and chemical forces involved in the adsorption process may be found in Crittenden et al. (2012). Adsorption can occur on the outer surface of the adsorbent and in the macropores, mesopores, micropores, and submicropores, but the surface area of the macro and mesopores is small compared with the surface area of the micropores and submicropores and the amount of material adsorbed there is usually considered negligible.

Because the adsorption process occurs in a series of steps, the slowest step in the series is identified as the rate-limiting step. When the rate of adsorption equals the rate of desorption, equilibrium has been achieved, and the capacity of the adsorbent has been reached. The theoretical adsorption capacity for a given adsorbent for a particular contaminant can be determined by developing adsorption isotherms, as described below. Because activated carbon is the most common adsorbent used in advanced wastewater treatment applications, the focus of the following discussion is on activated carbon.

Development of Adsorption Isotherms

The quantity of adsorbate that can be taken up by an adsorbent is a function of both the characteristics and concentration of adsorbate and the temperature. The characteristics of the adsorbate that are of importance include: solubility, molecular structure, molecular weight, polarity, and hydrocarbon saturation. Generally, the amount of material adsorbed is determined as a function of the concentration at a constant temperature, and the resulting function is called an adsorption isotherm. Adsorption isotherms are developed by exposing a given amount of absorbate in a fixed volume of liquid to varying amounts of activated carbon. Typically, more than ten containers are used, and the minimum time allowed for the samples to equilibrate where powdered activated carbon is used is seven days. If activated carbon is used, it is usually in the powdered form (as opposed to granular) to minimize adsorption times.

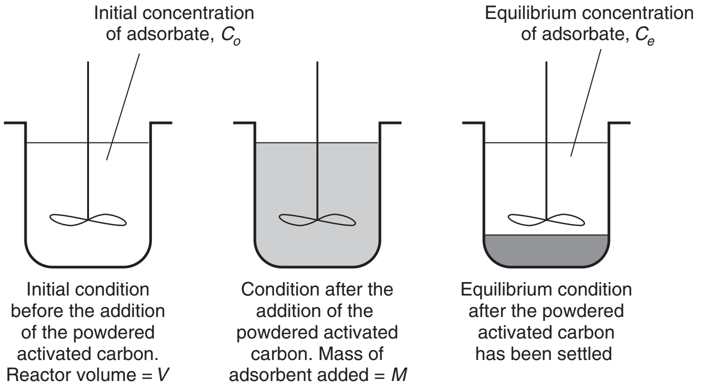
Mass Balance. If a mass balance is performed for a batch reactor into which a quantity of powdered activated carbon has been added (see Fig. 11–51), the resulting expression at equilibrium at the completion of the mass transfer process is given by

1. General word statement:

$$\begin{array}{c}\text{Amount of reactant}\\ \text{adsorbed within the}\\ \text{system boundary}\end{array} = \begin{array}{c}\text{initial amount of}\\ \text{reactant within the}\\ \text{system boundary}\end{array} - \begin{array}{c}\text{final amount of}\\ \text{reactant within the}\\ \text{system boundary}\end{array} \qquad (11\text{–}51)$$

Figure 11–51

Definition sketch for mass balance of carbon adsorption.



2. Simplified word statement:

$$\begin{array}{l}\text{Amount}\\\text{adsorbed}\end{array} = \begin{array}{l}\text{initial amount}\\\text{of adsorbate}\\\text{present}\end{array} - \begin{array}{l}\text{final amount}\\\text{of adsorbate}\\\text{present}\end{array} \tag{11–52}$$

3. Symbolic representation at equilibrium (refer to Fig. 11–51):

$$q_e M = VC_o - VC_e \tag{11–53}$$

where q_e = adsorbent phase concentration after equilibrium, mg adsorbate/g adsorbent
M = mass of adsorbent, g
V = volume of liquid in the reactor, L
C_o = initial solution concentration of adsorbate, mg/L
C_e = final solution equilibrium concentration of adsorbate after adsorption has occurred, mg/L

Equation (11–53) can be written as follows:

$$q_e = -\frac{V}{M}(C_e - C_o) \tag{11–54}$$

The adsorbent phase concentration data computed using Eq. (11–54) are then used to develop adsorption isotherms as described below.

Freundlich Isotherm. Equations used to describe the experimental isotherm data were developed by Freundlich, Langmuir, and Brunauer, Emmet, and Teller (BET isotherm) (Shaw, 1966). Of the three, the Freundlich isotherm is used most commonly to describe the adsorption characteristics of the activated carbon used in water and wastewater treatment. Derived empirically in 1912, the Freundlich isotherm is defined as follows:

$$\frac{x}{m} = K_f C_e^{1/n} \tag{11–55}$$

where x/m = mass of adsorbate adsorbed per unit mass of adsorbent, mg adsorbate/g activated carbon
K_f = Freundlich capacity factor, (mg absorbate/g activated carbon) × (L water/mg adsorbate)$^{1/n}$ = (mg/g)(L/mg)$^{1/n}$
C_e = equilibrium concentration of adsorbate in solution after adsorption, mg/L
$1/n$ = Freundlich intensity parameter

Table 11–42

Freundlich adsorption isotherm constants for selected organic compounds[a,b]

Compound	pH	$K(mg/g)(L/mg)^{1/n}$	1/n
Benzene	5.3	1.0	1.6–2.9
Bromoform	5.3	19.6	0.52
Carbon tetrachloride	5.3	11	0.83
Chlorobenzene	7.4	91	0.99
Chloroethane	5.3	0.59	0.95
Chloroform	5.3	2.6	0.73
DDT	5.3	322	0.50
Dibromochloromethane	5.3	4.8	0.34
Dichlorobromomethane	5.3	7.9	0.61
1, 2–Dichloroethane	5.3	3.6	0.83
Ethylbenzene	7.3	53	0.79
Heptachlor	5.3	1,220	0.95
Hexachloroethane	5.3	96.5	0.38
Methylene chloride	5.3	1.3	1.16
N-Dimethylnitrosamine	na	6.8×10^{-5}	6.60
N-Nitrosodi-n-propylamine	na	24	0.26
N-Nitrosodiphenylamine	3–9	220	0.37
PCB	5.3	14,100	1.03
PCB 1221	5.3	242	0.70
PCB 1232	5.3	630	0.73
Phenol	3–9	21	0.54
Tetrachloroethylene	5.3	51	0.56
Toluene	5.3	26.1	0.44
1, 1, 1–Trichloroethane	5.3	2–2.48	0.34
Trichloroethylene	5.3	28	0.62

[a] Adapted from Dobbs and Cohen (1980) and LaGrega et al. (2001).

[b] The adsorption isotherm constants reported in this table are meant to be illustrative of the wide range of values that will be encountered for various organic compounds. It is important to note that the characteristics of the activated carbon used as well as the analytical technique used for the analysis of the residual concentrations of the individual compounds will have a significant effect on the coefficient values obtained for specific organic compounds.

The constants in the Freundlich isotherm can be determined by plotting log (x/m) versus log C_e and making use of the linear form of Eq. (11–55) rewritten as

$$\log\left(\frac{x}{m}\right) = \log K_f + \frac{1}{n}\log C_e \tag{11–56}$$

The U.S. EPA (1980) has developed adsorption isotherms for a variety of toxic organic compounds, some of which are presented in Table 11–42. As shown in Table 11–42, the variation in the Freundlich capacity factor for the various compounds is extremely wide (e.g., 14,100 for PCB to 6.8×10^{-5} for N-Dimethylnitrosamine). Because of the wide variation, the Freundlich capacity factor must be determined for each new compound. Application of the Freundlich adsorption isotherm is illustrated in Example 11–10.

EXAMPLE 11–10 **Activated Carbon Required to Treat a Wastewater** As a result of effluent chlorination, the amount of chloroform formed was found to be 0.12 mg/L. How much powdered activated carbon will be required to treat an effluent flowrate of 4000 m^3/d to reduce the chloroform concentration to 0.05 mg/L? The Freundlich adsorption isotherm coefficients for chloroform are: $K_f = 2.6\ (\text{mg/g})(\text{L/mg})^{1/n}$ and $1/n = 0.73$.

Solution

1. Combine Eqs. (11–54) and (11–55) to obtain an expression for V/M as follows:

$$q_e = \frac{V}{M}(C_e - C_o)$$

$$q_e = \frac{x}{m} K_f C_e^{1/n}$$

$$-\frac{V}{M} = \frac{K_f C_e^{1/n}}{(C_e - C_o)}$$

2. Substitute the isotherm coefficients and solve for M/V:

$$-\frac{V}{M} = \frac{K_f C_e^{1/n}}{(C_e - C_o)} = \frac{2.6(0.05)^{0.73}}{0.05 - 0.12} = -4.17\ \text{L/g}$$

$$M/V = 1/4.17 = 0.24\ \text{g/L}$$

3. Determine the amount of carbon required to treat 4000 m^3/d.

$$\text{PAC required} = \frac{(0.24\ \text{g/L})(4000\ \text{m}^3/\text{d})(10^3\ \text{L}/1\ \text{m}^3)}{(10^3\ \text{g}/1\ \text{kg})} = 960\ \text{kg/d}$$

Comment Due to the cost and the amount of PAC required to treat the effluent to reduce the residual chloroform to 0.05 mg/L, carbon adsorption is a poor choice for the removal of residual chloroform.

Langmuir Isotherm. Derived from rational considerations, the Langmuir adsorption isotherm is defined as:

$$\frac{x}{m} = \frac{abC_e}{1 + bC_e} \qquad (11\text{–}57)$$

where x/m = mass of adsorbate adsorbed per unit mass of adsorbent, mg adsorbate/ g activated carbon
a, b = empirical constants
C_e = equilibrium concentration of adsorbate in solution after adsorption, mg/L

The Langmuir adsorption isotherm was developed by assuming (1) a fixed number of accessible sites are available on the adsorbent surface, all of which have the same energy, and

(2) adsorption is reversible. Equilibrium is reached when the rate of adsorption of molecules onto the surface is the same as the rate of desorption of molecules from the surface. The rate at which adsorption proceeds is proportional to the driving force, which is the difference between the amount adsorbed at a particular concentration and the amount that can be adsorbed at that concentration. At the equilibrium concentration, this difference is zero.

Correspondence of experimental data to the Langmuir equation does not mean that the stated assumptions are valid for the particular system being studied because deviations from the assumptions can have a canceling effect. The constants in the Langmuir isotherm can be determined by plotting $C_e/(x/m)$ versus C_e and making use of the linear form of Eq. (11–57) rewritten as:

$$\frac{C_e}{(x/m)} = \frac{1}{ab} + \frac{1}{a}C_e \tag{11–58}$$

For the case where the Langmuir adsorption isotherm best represents a set of experimental isotherm data, a plot of $C_e/(x/m)$ vs. C_e will be linear, with a slope of $1/a$ and a y-intercept of $1/(ab)$. Application of the Langmuir adsorption isotherm is illustrated in Example 11–11.

EXAMPLE 11–11 **Analysis of Activated Carbon Adsorption Data** Determine which isotherm equation (i.e., Freundlich and Langmuir) best fits the isotherm coefficients for the following GAC adsorption test data. Also determine the corresponding coefficients for the isotherm equation. The liquid volume used in the batch adsorption tests was 1 L. The initial concentration of the adsorbate in solution was 3.37 mg/L. Equilibrium was obtained after 7 d.

Mass of GAC, m, g	Equilibrium concentration of adsorbate in solution, C_e, mg/L
0.0	3.37
0.001	3.27
0.010	2.77
0.100	1.86
0.500	1.33

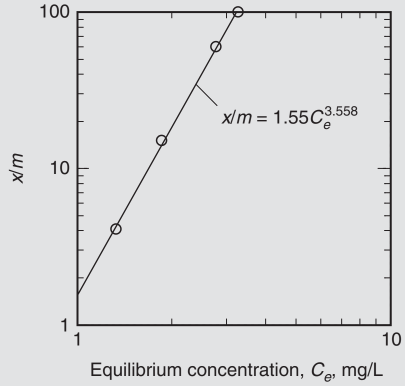
Solution

1. Derive the values needed to plot the Freundlich and Langmuir adsorption isotherms using the batch adsorption test data.

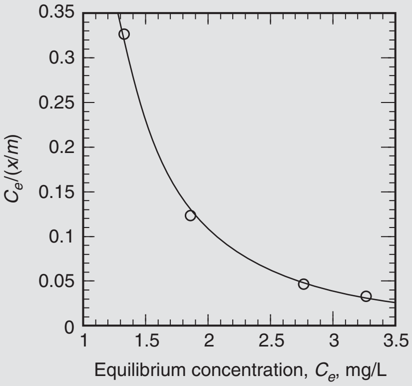
Adsorbate concentration, mg/L					
C_o	C_e	$C_o - C_e$	m, g	x/m,[a] mg/g	$C_e/(x/m)$
3.37	3.37	0.00	0.000	—	—
3.37	3.27	0.10	0.001	100	0.0327
3.37	2.77	0.60	0.010	60	0.0462
3.37	1.86	1.51	0.100	15.1	0.1232
3.37	1.33	2.04	0.500	4.08	0.3260

[a] $\frac{x}{m} = \frac{(C_o - C_e)V}{m}$

2. Plot the Freundlich and Langmuir adsorption isotherms using the data developed in Step 1 and determine which isotherm best fits the data.
   a. The required plots are given below.



Freundlich isotherm plot



Langmuir isotherm plot

   b. From the above plots, the experimental data are best represented by the Freundlich isotherm. Because the plot for the Langmuir isotherm is curvilinear, use of the Langmuir adsorption isotherm is inappropriate.
3. Determine the Freundlich adsorption isotherm coefficients.
   a. When x/m versus C_e is plotted on log-log paper, the intercept on the x/m axis when $C_e = 1.0$ is the value of K_f and the slope of the line is equal to $1/n$. Thus, $x/\mathrm{m} = 1.55$, and $K_f = 1.55$
   b. When $x/\mathrm{m} = 1.0$, $C_e = 0.89$, and $1/n = 3.6$
   c. The form of the resulting isotherm is $\frac{x}{\mathrm{m}} = 1.55C_e^{3.6}$
   d. The Freundlich adsorption isotherm equation may also be determined using a power-type best fit through the data.

Adsorption of Mixtures

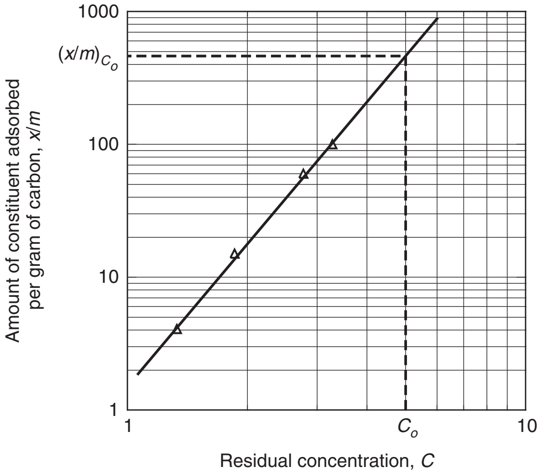
In the application of adsorption in water reclamation, mixtures of organic compounds in reclaimed water are always encountered. Typically, there is a depression of the adsorptive capacity of any individual compound in a solution of many compounds, but the total adsorptive capacity of the adsorbent may be larger than the adsorptive capacity with a single compound. The amount of inhibition due to competing compounds is related to the size of the molecules being adsorbed, their adsorptive affinities, and their relative concentrations. It is important to note that adsorption isotherms can be determined for a heterogeneous mixture of compounds including total organic carbon (TOC), dissolved organic carbon (DOC), chemical oxygen demand (COD), dissolved organic halogen (DOH), UV absorbance, and fluorescence (Snoeyink and Summers, 1999). The adsorption from mixtures is considered further in Crittenden et al. (1985, 1987a, 1987b, 1987c,) and Sontheimer and Crittenden (1988).

Adsorption Capacity

The adsorptive capacity of a given adsorbent is estimated from isotherm data as follows. If isotherm data are plotted, the resulting isotherm will be as shown on step 2 of Example 11–11. As shown on Fig. 11–52, the adsorptive capacity of the carbon can be estimated by

Figure 11–52

Plot of Freundlich isotherm used for determination of breakthrough adsorption capacity.



extending a vertical line from the point on the horizontal axis corresponding to the initial concentration C_o, and extrapolating the isotherm to intersect this line. The value at the point of intersection $[(x/m)_{C_o}]$ can be read from the vertical axis. The value represents the amount of constituent adsorbed per unit weight of carbon when the carbon is at equilibrium with the initial concentration of constituent, C_o. The equilibrium condition generally exists in the upper section of a carbon bed during column treatment, and it, therefore, represents the ultimate capacity of the carbon for a particular reclaimed water. The value of the breakthrough adsorption capacity $(x/m)_b$ can be determined using the small-scale column test described later in this section. Typically, breakthrough is said to have occurred when the effluent concentration reaches 5 percent of the influent value. Exhaustion of the adsorption bed is assumed to have occurred when the effluent concentration is equal to 95 percent of the influent concentration. A number of equations have been developed to describe the breakthrough curve including those by Bohart and Adams (1920) and Crittenden et al. (1987a).

Mass Transfer Zone. The area of the GAC bed in which sorption is occurring is called the mass transfer zone (MTZ), as shown on Fig. 11–53. After the water containing the constituent to be removed passes through a region of the bed whose depth is equal to the MTZ, the concentration of the contaminant in the water will have been reduced to its minimum value. No further adsorption will occur within the bed below the MTZ. As the top layers of carbon granules become saturated with organic material, the MTZ will move down in the bed until breakthrough occurs. The volume of a given water processed until breakthrough and exhaustion is designated as V_{BT} and V_E, respectively, as shown on Fig. 11–53. The length of the MTZ is typically a function of the hydraulic loading rate applied to the column and the characteristics of the activated carbon. In the extreme, if the loading rate is too great the length of the MTZ will be larger than the GAC bed depth, and the adsorbable constituents will not be removed completely by the carbon. At complete exhaustion, the effluent concentration is equal to the influent concentration.

Breakthrough Curve. In addition to the applied hydraulic loading rate, the shape of the breakthrough curve will also depend on whether the applied liquid contains nonadsorbable and biodegradable constituents. The impact of the presence of nonadsorbable and biodegradable organic constituents on the shape of the breakthrough curve is illustrated on

Figure 11–53

Typical breakthrough curve for activated carbon showing movement of mass transfer zone (MTZ) with throughput volume.

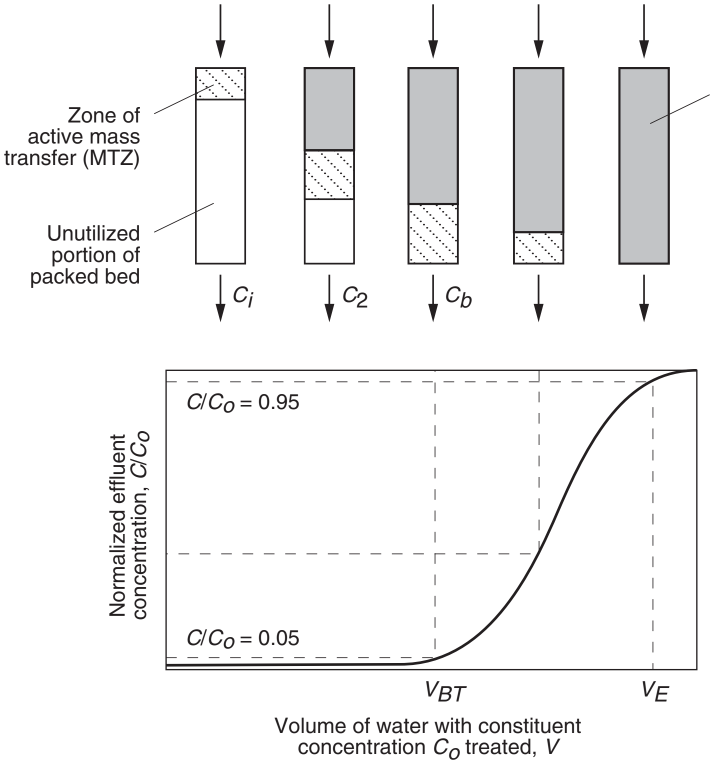


Fig. 11–54. As shown on Fig. 11–54, if the liquid contains nonadsorbable constituents, the nonadsorbable constituents will appear in the effluent as soon as the carbon column is put into operation (ignoring the short period of time for hydraulic conductivity). If adsorbable and biodegradable constituents are present in the applied liquid, the breakthrough curve will not reach a C/C_o value of 1.0 but will be depressed, and the observed C/C_o value will depend on the biodegradability of the influent constituents because biological activity continues even though the adsorption capacity has been utilized. If the liquid contains nonadsorbable and biodegradable constituents, the observed breakthrough curve will not

Figure 11–54

Impact of the presence of adsorbable, nonadsorbable, and biodegradable organic constituents on the shape of the activated carbon breakthrough curve. (Adapted from Snoeyink and Summers, 1999.)

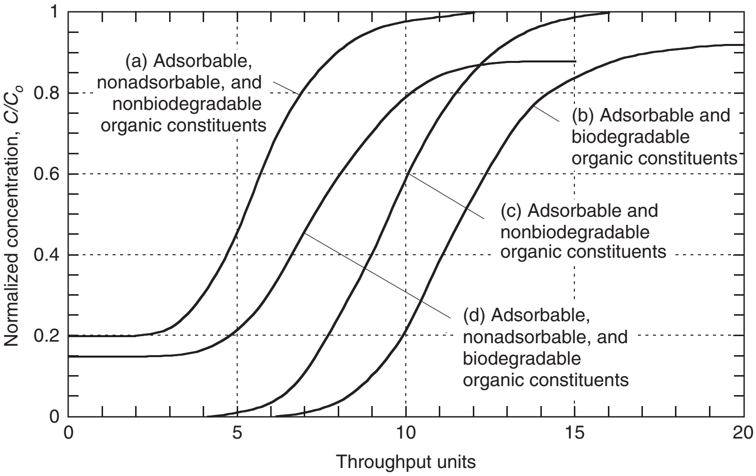
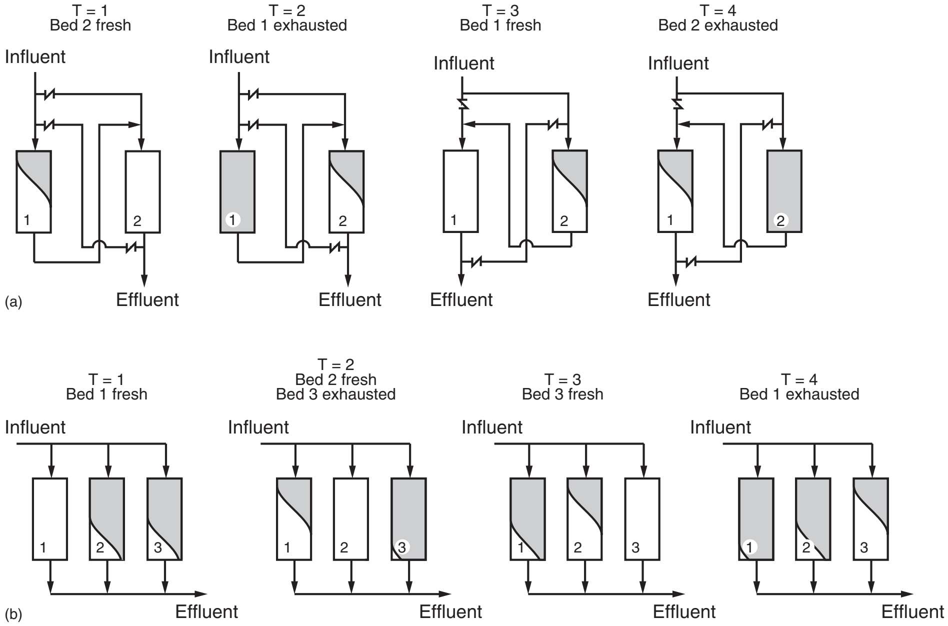


Figure 11–55

Activated carbon contactor configurations: (a) series and (b) parallel operation.



start at zero and will not terminate at a value of 1.0 (Snoeyink and Summers, 1999). The above effects are observed commonly in wastewater adsorption applications, especially with respect to the removal of COD.

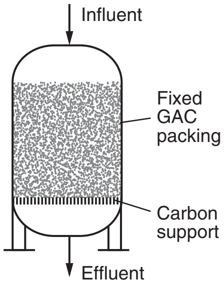
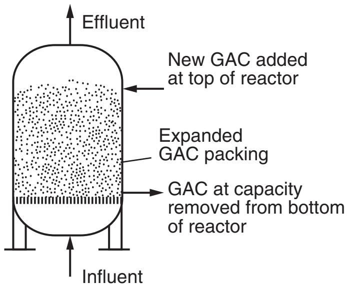
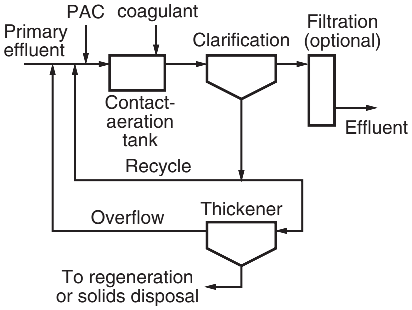
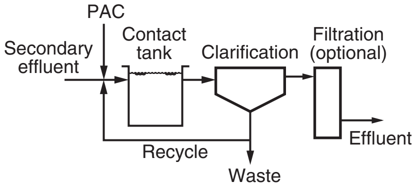
In practice, the only way to use the capacity at the bottom portion of the carbon adsorption column is to have two or more columns in series and switch them as they are exhausted (lead-lag), or to use multiple columns in parallel so that breakthrough in one column does not affect the combined effluent quality. This mode of operation is referred to as the carousel technique, where multiple columns can be rotated so that only one column reaches exhaustion at a given time. The arrangement of adsorption columns in series and parallel configurations is shown on Fig. 11–55(a) and (b), respectively. A minimum of two parallel or series carbon contactors is recommended for design. Multiple units permit one or more units to remain in operation while one unit is taken out of service for removal and regeneration of spent carbon, or for maintenance. The optimum flowrate and bed depth, as well as the operating capacity of the carbon, must be established to determine the dimensions and the number of columns necessary for continuous treatment. These parameters can be determined from dynamic column tests, as discussed below.

Adsorption Contactors. Several types of activated carbon contactors are used for trace constituent removal, including fixed and expanded beds, addition of PAC to the activated sludge process, and separate mixed systems with subsequent carbon separation, as summarized in Table 11–43. A typical pressurized, down-flow carbon contactor is shown on Fig. 11–56. The sizing of carbon contactors is based on a number of factors, as summarized in Table 11–44 for a downflow packed bed contactor. For the case where the mass transfer rate is fast and the mass transfer zone is a sharp wave front, a steady-state mass balance around a fixed bed carbon contactor may be written as:

Accumulation = inflow − outflow − amount adsorbed

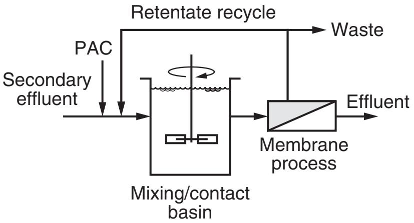
Table 11–43

Application of activated carbon for the removal of trace constituents from wastewater

Configuration	Description
(a) Fixed bed GAC column (downflow)  	Fixed-bed downflow columns can be operated singly, in series, or in parallel (see Fig. 11–55). Granular medium filters are used commonly upstream of the activated carbon contactors to remove the organics associated with the suspended solids present in secondary effluent, however the adsorption of organics and filtration of suspended solids can also be accomplished in a single step. In the downflow design, the water to be treated is applied to the top of the column and withdrawn at the bottom. The carbon is held in place with an underdrain system at the bottom of the column. Provision for backwashing and surface washing is often provided in wastewater applications to limit the headloss buildup due to the removal of particulate suspended solids within the carbon column. Unfortunately, backwashing has the effect of destroying the adsorption front. Although upflow fixed-bed reactors have been used, downflow beds are used more commonly to lessen the chance of accumulating particulate material in the bottom of the bed, where the particulate material would be difficult to remove by backwashing.
(b) Expanded bed GAC (upflow)  	In the expanded (or fluidized) bed system, the influent is introduced at the bottom of the column and the activated carbon is allowed to expand, much as a filter bed expands during backwash. When the adsorptive capacity of the carbon at the bottom of the column is exhausted, the bottom portion of carbon is removed, and an equivalent amount of regenerated or virgin carbon is added to the top of the column. In such a system, headloss does not build up with time after the operating point has been reached. In general, expanded bed upflow contactors may have more carbon fines in the effluent than downflow contactors because bed expansion leads to the creation of fines as the carbon particles collide and abrade, and allows the fines to escape through passageways created by the expanded bed. While not used commonly, continuous backwash moving-bed and pulsed-bed carbon contactors have been used (see Table 8–4 for filter configurations).
(c) Activated Sludge with PAC addition  	The use of powdered activated carbon with the activated sludge process, where activated carbon is added directly to the aeration tank, results in simultaneous biological oxidation and physical adsorption. A feature of this process is that it can be integrated into existing activated sludge systems at nominal capital cost. The addition of powdered activated carbon has several process advantages, including: (1) system stability during shock loads, (2) reduction of refractory priority pollutants, (3) color and ammonia removal, and (4) improved sludge settleability. In some industrial waste applications where nitrification is inhibited by toxic organics, the application of powdered activated carbon may reduce or limit this inhibition.
(d) Mixed PAC contactor with gravity separation  	Powdered activated carbon can be applied to the effluent from biological treatment processes in a separate contacting basin. The contactor can operate in a batch or continuous flow mode. In the batch mode, after a specified amount of time for contact, the carbon is allowed to settle to the bottom of the tank, and the treated water is then removed from the tank. The continuous flow operation consists of a basin divided for contacting and settling. The settled carbon may be recycled to the contact tank. Because carbon is very fine, a coagulant, such as a polyelectrolyte, may be needed to aid in the removal of the carbon particles, or filtration through rapid sand filters may be required. In some treatment processes, PAC is used in conjunction with chemicals used for the precipitation of specific constituents.

(*continued*)

Table 11–43 (*Continued*)

Configuration	Description
(e) Mixed PAC contactor with membrane separation  	The removal of trace constituents in a complete mix or plug flow contactor may be combined with separation by micro or ultrafiltration membranes. The PAC is added to the secondary effluent by continuous or pulse addition, followed by concentration of the PAC on the membrane. When the headloss across the membrane reaches a given value, a backwash cycle is initiated. The backwash containing the PAC retentate may be wasted or recycled to the contact basin. A number of full-scale plants have used this process (Snoeyink et al., 2000, Anselme et al., 1997).

$$0 = QC_o t - QC_e t - m_{\mathrm{GAC}} q_e \tag{11–59}$$

where Q = volumetric flowrate, L/h
C_o = initial concentration of adsorbate, mg/L
t = time, h
C_e = final equilibrium concentration of adsorbate, mg/L
m_{GAC} = mass of adsorbent, g
q_e = adsorbent phase concentration after equilibrium, mg adsorbate/g adsorbent

From Eq. (11–59), the adsorbent usage rate is defined as

$$\frac{m_{\mathrm{GAC}}}{Qt} = \frac{C_o - C_e}{q_e} \tag{11–60}$$

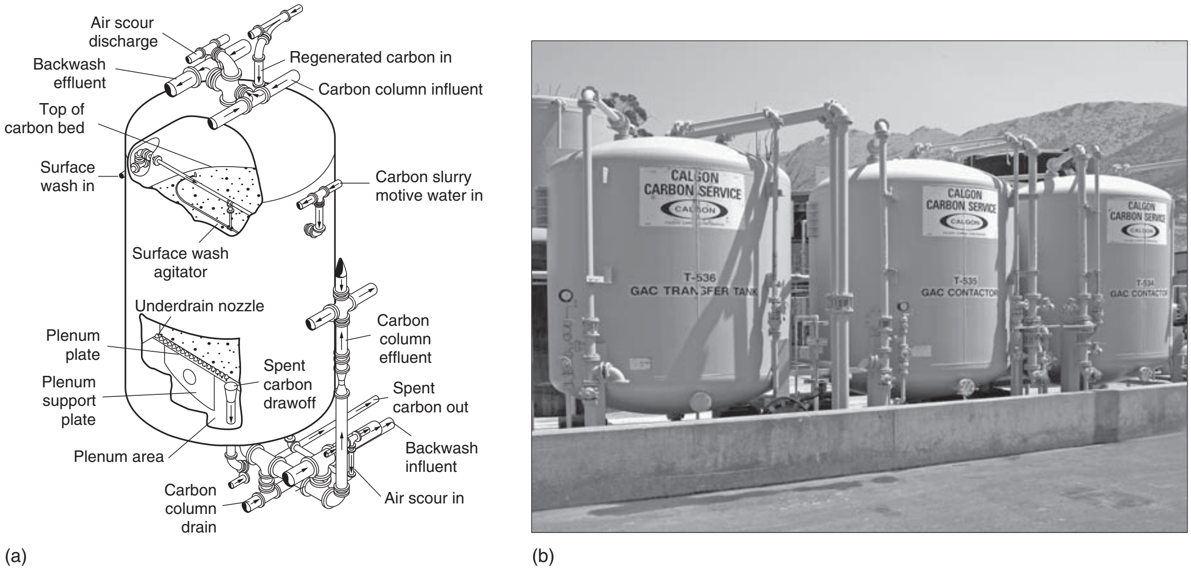


Figure 11–56

Activated carbon contactors: (a) illustration of typical pressure vessel contactor and (b) view of typical granular activated carbon contactors operated in parallel, used for the treatment of filtered secondary effluent.

Table 11–44

Typical design values for GAC contactors[a]

Parameter	Symbol	Unit	Value
Volumetric flowrate	V	m^3/h	50–400
Bed volume	V_b	m^3	10–50
Cross-sectional area	A_b	m^2	5–30
Carbon depth	D	m	1.8–4
Void fraction	α	m^3/m^3	0.38–0.42
GAC density	ρ	kg/m^3	350–550
Approach velocity	v_f	m/h	5–15
Effective contact time	t	min	2–10
Empty bed contact time	EBCT	min	5–30
Operation time	t	d	100–600
Throughput volume	V_L	m^3	10–100
Specific throughput	V_{sp}	m^3/kg	50–200
Bed volumes[b]	BV	m^3/m^3	2000–20,000

[a] Adapted from Sontheimer et al. (1988).

[b] Total volume of water processed expressed in terms of the reactor bed volume.

If it is assumed that the mass of the adsorbate in the pore space is small compared to the amount adsorbed, then the term $QC_e t$ in Eq. (11–60) can be neglected without serious error and the adsorbent usage rate is given by:

$$\frac{m_{\text{GAC}}}{Qt} \approx \frac{C_o}{q_e} \tag{11–61}$$

To quantify the operational performance of GAC contactors, the following terms have been developed and are used commonly.

1. Empty bed contact time (EBCT)

$$EBCT = \frac{V_b}{Q} = \frac{A_b D}{v_f A_b} = \frac{D}{v_f} \tag{11–62}$$

where EBCT = empty bed contact time, h
V_b = volume of contactor occupied by GAC, m^3
Q = volumetric flowrate, m^3/h
A_b = cross-sectional area of GAC filter bed, m^2
D = depth of GAC in contactor, m
v_f = linear approach velocity, m/h

2. Activated carbon density.
The density of the activated carbon is defined as

$$\rho_{\text{GAC}} = \frac{m_{\text{GAC}}}{V_b} \tag{11–63}$$

where ρ_{GAC} = density of granular activated carbon, g/L
m_{GAC} = mass of granular activated carbon, g
V_b = volume of contactor occupied by GAC, L

3. Specific throughput, expressed as m^3 of water treated per gram of carbon:

$$\text{Specific throughput, m}^3\text{/g} = \frac{Qt}{m_{\text{GAC}}} = \frac{V_b t}{\text{EBCT} \times m_{\text{GAC}}} \tag{11–64}$$

Using Eq. (11–63), Eq. (11–64) can be written as

$$\text{Specific throughput} = \frac{V_b t}{\text{EBCT}(\rho_{\text{GAC}} \times V_b)} = \frac{t}{\text{EBCT} \times \rho_{\text{GAC}}} \tag{11–65}$$

4. Carbon usage rate (CUR) expressed as gram of carbon per m^3 of water treated:

$$\text{CUR, g/m}^3 = \frac{m_{\text{GAC}}}{Qt} = \frac{1}{\text{Specific throughput}} \tag{11–66}$$

5. Volume of water treated for a given EBCT, expressed in liters, L:

$$\text{Volume of water treated, m}^3 = \frac{\text{Mass of GAC for given EBCT}}{\text{GAC usage rate}} \tag{11–67}$$

6. Bed life, expressed in days, d:

$$\text{Bed life, d} = \frac{\text{Volume of water treated for given EBCT}}{Q} \tag{11–68}$$

The application of these terms is illustrated in Example 11–12.

EXAMPLE 11–12 **Estimation of Activated Carbon Adsorption Breakthrough Time** A fixed-bed activated carbon adsorber has a fast mass transfer rate and the mass transfer zone is essentially a sharp wave front. Assuming the following data apply, determine the carbon requirements to treat a flowrate of 1000 L/min, and the corresponding bed life.

1. Compound to be treated = Trichloroethylene (TCE)
2. Initial concentration, $C_o = 1.0$ mg/L
3. Final concentration $C_e = 0.005$ mg/L
4. GAC density = 450 g/L
5. Freundlich capacity factor, $K_f = 28\ (\text{mg/g})(\text{L/mg})^{1/n}$ (see Table 11–42)
6. Freundlich intensity parameter, $1/n = 0.62$ (see Table 11–42)
7. EBCT = 10 min

Ignore the effects of biological activity within the column.

Solution

1. Estimate the GAC usage rate for TCE. The GAC usage rate is estimated using Eq. (11–60) and Eq. (11–55).

$$\frac{m_{\text{GAC}}}{Qt} = \frac{C_o - C_e}{q_e} = \frac{C_o - C_e}{K_f C_o^{1/n}}$$

$$= \frac{(1.0\ \text{mg/L}) - (0.005\ \text{mg/L})}{28(\text{mg/g})(\text{L/mg})^{0.62}\ (1.0\ \text{mg/L})^{0.62}}$$

$$= 0.036\ \text{g GAC/L}$$

2. Determine the mass of carbon required for a 10 min EBCT.

   The mass of GAC in the bed = $V_b \rho_{GAC}$ = (EBCT)(Q)(ρ_{GAC})

   Carbon required = 10 min (1000 L/min) (450 g/L) = 4.5×10^6 g

3. Determine the volume of water treated using a 10 min EBCT.

$$\text{Volume of water treated} = \frac{\text{Mass of GAC for given EBCT}}{\text{GAC usage rate}}$$

$$\text{Volume of water treated} = \frac{4.5 \times 10^6 \text{ g}}{(0.036 \text{ g GAC/L})} = 1.26 \times 10^8 \text{ L}$$

4. Determine the bed life.

$$\text{Bed life} = \frac{\text{Volume of water treated for given EBCT}}{Q}$$

$$\text{Bed life} = \frac{1.26 \times 10^8 \text{ L}}{(1000 \text{ L/min})(1440 \text{ min/d})} = 87.5 \text{ d}$$

Comment In this example, the full capacity of the carbon in the contactor was utilized based on the assumption that two columns in series are used. If a single column is used, then a breakthough curve must be used to arrive at the bed life. The Freundlich isotherm parameters, *K* and 1/*n*, are a function of the initial concentration, the actual carbon that is used, as well as the water quality (temperature, pH). Equilibrium isotherms are used to determine these parameters for the conditions of interest.

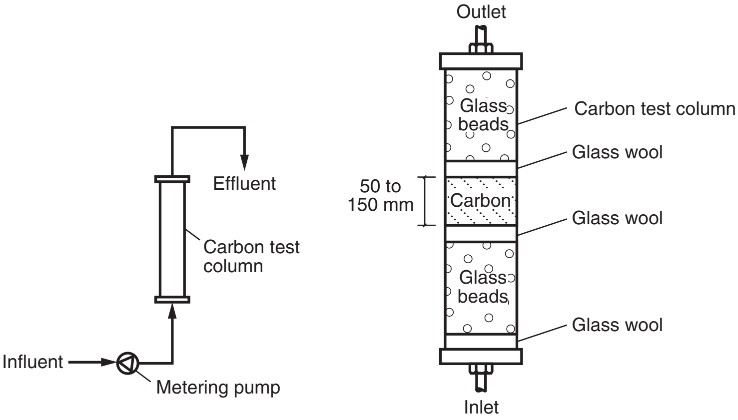
Small Scale Column Tests

Over the years, a number of small scale column tests have been developed to simulate the results obtained with full scale reactors. One of the early column tests was the high-pressure minicolumn (HPMC) technique developed by Rosene et al. (1980), and later modified by Bilello and Beaudet (1983). In the HPMC test procedure, a high-pressure liquid chromatography column loaded with activated carbon is used. Typically the HPMC test procedure is used to determine the capacity of activated carbon for the adsorption of volatile organic compounds. The principal advantage of the HPMC test procedure is that it allows for the rapid determination of the GAC adsorptive capacity under conditions similar to those encountered in the field.

An alternative procedure known as the *rapid small-scale column test* (RSSCT) has been developed by Crittenden et al. (1986, 1987d, 1991). The test procedure allows for the scaling of data obtained from small columns (see Fig. 11–57) to predict the performance of pilot or full-scale carbon columns. In developing the procedure, mathematical models were used to define the relationships between the breakthrough curve for small and large columns. In adsorption columns, the mass transfer mechanisms that are responsible for the spreading of the mass transfer zone are (1) dispersion, (2) film diffusion, and (3) intraparticle diffusion. Two different design relationships were developed, one for constant diffusivity and one for proportional, or non-constant, diffusivity. In the constant diffusivity model, it is assumed that dispersion is negligible because the hydraulic loading rate is high in the RSSCT, and that mass transfer occurs as a result of film diffusion. Further, it is

Figure 11–57

Schematic of column used for rapid small scale column testing (RSSCT) to develop data for pilot or full scale carbon columns.



Column to particle diameter = 20:1 (or greater)
Column diameter = 20 to 40 mm
Column length = 300 mm

assumed that the intraparticle diffusivity is the same for both the small and large columns. In the proportional diffusivity model, it is assumed that dispersion is negligible because the hydraulic loading rate is high in the RSSCT, and that mass transfer occurs as a result of intraparticle diffusion. The relationships for the two cases can be generalized as follows:

$$\frac{\mathrm{EBCT}_{SC}}{\mathrm{EBCT}_{LC}} = \left(\frac{d_{SC}}{d_{LC}}\right)^{2-x} = \frac{t_{SC}}{t_{LC}} \tag{11–69}$$

$$\frac{v_{SC}}{v_{LC}} = \frac{d_{LC}}{d_{SC}} \tag{11–70}$$

where d_{SC} = diameter of particle in small-scale column, mm
d_{LC} = diameter of particle in large-scale column, mm
t_{SC} = time in small-scale column, min
t_{LC} = time in large-scale column, min
v_{SC} = superficial velocity in small-scale column, m/h
v_{LC} = superficial velocity in large-scale column, m/h

For constant and proportional diffusivity, the value of x in the exponent in Eq. (11–69) is 0 and 1, respectively. The application of the above equations is illustrated in Example 11–13.

EXAMPLE 11–13 **Comparison of Rapid Small-scale Column Test Parameters to Pilot Scale Parameters** Determine the corresponding parameters for a RSSCT based on the following data proposed for a pilot scale column. Assume that film diffusion is the controlling mechanism.

Parameter	Unit	Pilot column (LC)	RSSCT (SC)
Particle diameter	mm	0.5	0.1
Carbon density	g/L	450	450
EBCT	min	10	

(continued)

(*Continued*)

Parameter	Unit	Pilot column (LC)	RSSCT (SC)
Loading rate	m/h	5.0	
Flowrate	mL/min	200	
Column diameter	mm	75	10[a]
Column length	mm	1000	
Mass adsorbent	g		
Time of operation	d	100	
Water volume	L	28,800	

[a] Assumed value for small column.

Solution

1. Estimate the EBCT for the RSSCT.

$$\mathrm{EBCT}_{SC} = \mathrm{EBCT}_{LC}\left(\frac{d_{SC}}{d_{LC}}\right)^2$$

$$\mathrm{EBCT}_{SC} = 10\ \mathrm{min}\left(\frac{0.1}{0.5}\right)^2 = 0.4\ \mathrm{min}$$

2. Estimate the loading rate for the RSSCT.

$$v_{SC} = v_{LC}\frac{d_{LC}}{d_{SC}}$$

$$v_{SC} = 5\ \mathrm{m/h}\,\frac{0.5}{0.1} = 25\ \mathrm{m/h}$$

3. Estimate flowrate for the RSSCT.

$$A = \frac{\pi}{4}d_{SC}^2 = \frac{\pi}{4}(10\ \mathrm{mm})^2 = 78.5\ \mathrm{mm}^2$$

$$Q_{SC} = (v_{SC})(A)$$

$$Q_{SC} = \frac{(25\ \mathrm{m/h})(10^3\ \mathrm{mm/1\ m})(78.5\ \mathrm{mm}^2)}{(60\ \mathrm{min/h})(10^3\ \mathrm{mm}^3/1\ \mathrm{mL})} = 32.7\ \mathrm{mL/min}$$

4. Estimate column length for the RSSCT.

$$L_{SC} = \frac{Q_{SC} \times \mathrm{EBCT}_{SC}}{A} = \frac{(32{,}700\ \mathrm{mm}^3/\mathrm{min})(0.4\ \mathrm{min})}{78.5\ \mathrm{mm}^2} = 166.7\ \mathrm{mm}$$

5. Estimate mass of adsorbent required for the RSSCT.

$$M_{SC} = \mathrm{EBCT}_{LC}\left(\frac{d_{SC}}{d_{LC}}\right)^2(Q_{SC})(\rho_{SC})$$

$$M_{SC} = 10\ \mathrm{min}\left(\frac{0.1\ \mathrm{mm}}{0.5\ \mathrm{mm}}\right)^2\left[\frac{(32.7\ \mathrm{mL/min})(450\ \mathrm{g/L})}{(10^3\ \mathrm{mL/1\ L})}\right] = 5.9\ \mathrm{g}$$

6. Estimate time of operation for the RSSCT.

$$t_{SC} = t_{LC}\frac{\mathrm{EBCT}_{SC}}{\mathrm{EBCT}_{LC}}$$

$$t_{SC} = 100\left(\frac{0.4\ \mathrm{min}}{10\ \mathrm{min}}\right) = 4\ \mathrm{d}$$

7. Estimate volume of water required for the RSSCT.

$$\upsilon_W = Q_{SC} \times t_{SC}$$

$$\upsilon_W = \frac{(32.7\ \text{mL/min})(4\ \text{d})(1440\ \text{min/d})}{(10^3\ \text{mL/1 L})} = 188.4\ \text{L}$$

8. Summarize the findings for the RSSCT.

Parameter	Unit	Pilot column	RSSCT
Particle radius	mm	0.5	0.1
Carbon density	g/L	450	450
EBCT	min	10	0.4
Loading rate	m/h	5.0	25.0
Flowrate	mL/min	200	32.7
Column diameter	mm	75	10[a]
Column length	mm	1000	166.7
Mass adsorbent	g		5.9
Time of operation	d	100	4
Water volume	L	28,800	188.4

[a] Assumed value for small column.

Comment The time savings in conducting the RSSCT versus the pilot column is apparent. Furthermore, many more tests can be conducted to test alternative configurations and carbon types. Often RSSCT's are performed in advance of piloting to narrow the list of the most appropriate carbon media for use in piloting.

Analysis of Powdered Activated Carbon Contactor

For a powdered activated carbon (PAC) application, the isotherm adsorption data can be used in conjunction with a materials mass balance analysis to obtain an approximate estimate of the amount of carbon that must be added as illustrated below. Here again, because of the many unknown factors involved, column and bench scale tests are recommended to develop the necessary design data. If a mass balance is written around the contactor (i.e., a batch reactor) after equilibrium has been reached, the resulting expression is given by Eq. (11–54), as derived previously. Estimation of the of powdered activated carbon (PAC) dose for adsorption is illustrated in Example 11–14.

EXAMPLE 11–14 **Estimation of Powdered Activated Carbon (PAC) Adsorption Dose and Cost** A treated wastewater with a flowrate of 1000 L/min is to be treated with PAC to reduce the concentration of residual organics measured as TOC from 5 to 1 mg/L. The Freundlich adsorption isotherm parameters were developed as discussed previously. Assuming the following data apply, determine the PAC requirements to treat the wastewater flow. If PAC costs \$0.50/kg, estimate the annual cost for treatment, assuming the PAC will not be regenerated.

1. Compound = mixed organics
2. Initial concentration, $C_o = 5.0$ mg/L
3. Final concentration, $C_e = 1.0$ mg/L
4. GAC density = 450 g/L
5. Freundlich capacity factor, $K_f = 150\ (\text{mg/g})(\text{L/mg})^{1/n}$
6. Freundlich intensity parameter, $1/n = 0.5$

Solution

1. Estimate the PAC dose based on the isotherm data. The PAC dose can be estimated by writing Eq. (11–54) as follows:

$$\frac{m}{V} = \frac{(C_o - C_e)}{q_e} = \frac{(C_o - C_e)}{K_f C_e^{1/n}}$$

Substituting the given values in the expression yields:

$$\frac{m}{V} = \frac{(5\ \text{mg/L} - 1\ \text{mg/L})}{150(\text{mg/g})(\text{L/mg})^{0.5}(1.0\ \text{mg/L})^{0.5}} = 0.0267\,\text{g/L}$$

2. Estimate the annual cost for the PAC treatment.

Annual cost =

$$= \frac{(0.0267\ \text{g/L})(1000\ \text{L/min})(1440\ \text{min/d})(365\ \text{d/y})(\$0.50/\text{kg})}{(10^3\ \text{g/1 kg})}$$

Annual cost = \$7008/y

Comment For small wastewater flows, it is not usually cost effective to plan for carbon regeneration.

Activated Sludge Powdered Activated Carbon Treatment

Powdered activated carbon treatment (PACT), a proprietary process, is described in Table 11–43 along with other applications. The dosage of powdered activated carbon and the mixed liquor-powdered activated carbon suspended solids concentration are related to the SRT as follows:

$$X_p = \frac{X_i\,\text{SRT}}{\tau} \tag{11–71}$$

where X_p = equilibrium powdered activated carbon-MLSS content, mg/L
X_i = powdered activated carbon dosage, mg/L
SRT = solids retention time, d
τ = hydraulic retention time, d

Carbon dosages typically range from 20 to 200 mg/L. With higher SRT values, the organic removal per unit of carbon is enhanced, thereby improving the process efficiency. Reasons cited for this phenomenon include (1) additional biodegradation due to decreased toxicity, (2) degradation of normally nondegradable substances due to increased exposure time to the biomass through adsorption on the carbon, and (3) replacement of low molecular weight compounds with high molecular weight compounds, resulting in improved adsorption efficiency and lower toxicity.

Carbon Regeneration

In many situations, the economical application of activated carbon depends on an efficient means of regenerating and reactivating the carbon after its adsorptive capacity has been reached. Regeneration is the term used to describe all of the processes that are used to recover the adsorptive capacity of the spent carbon, exclusive of reactivation. Typically, some of the adsorptive capacity of the carbon (about 4 to 10 percent) is lost in the regeneration process, while a loss of 2 to 5 percent is expected during the reactivation process, and a 4 to 8 percent loss of carbon is assumed due to attrition, abrasion, and mishandling. In general, regenerated activated carbon is not used in reclaimed water applications because of the potential for residual constituents, not removed in the regeneration process, to desorb and contaminate the reclaimed water. Additional details on carbon reactivation and regeneration may be found in Sontheimer and Crittenden (1988).

Adsorption Process Limitations

The adsorption process in water reuse applications is limited by (1) the logistics involved with transport of large volumes of adsorbent materials, (2) the area requirements for the carbon contactors, and (3) the production of waste adsorbent that can be difficult to regenerate and may need to be disposed of as hazardous waste due to the presence of toxic constituents. In particular, PAC contributes directly to the residuals solid loading and must be considered in terms of the impact on residuals handling. Further, the regeneration of some adsorbents is not feasible, resulting in potentially high media replacement costs. Process monitoring and control is essential, as the performance of carbon contactors will be affected by variations in pH, temperature, and flowrate.

11–10 GAS STRIPPING

Gas stripping involves the mass transfer of a gas from the liquid phase to the gas phase. The transfer is accomplished by contacting the liquid containing the gas that is to be stripped with a gas (usually air) which does not contain the gas initially. The removal of dissolved gases from wastewaters by gas (usually air) stripping has received considerable attention, especially for the removal of ammonia and odorous gases and volatile organic compounds (VOCs). Early work on the air stripping of ammonia from wastewater was conducted at Lake Tahoe, CA (Culp and Slechta, 1966; Slechta and Culp, 1967). The removal of VOCs by aeration is considered in Sec. 16–4 in Chap. 16.

The purpose of this section is to introduce the fundamental principles involved in gas stripping and to illustrate the general application of these principles. A design procedure is also presented. The material presented in this section is applicable to the removal of ammonia (NH_3), carbon dioxide (CO_2), oxygen (O_2), hydrogen sulfide (H_2S), and a variety of VOCs. The focus of the discussion in this section is on the analysis of facilities designed specifically for the removal of gaseous constituents as opposed to the removal of odorous gases (see Section 16–3, Chap. 16) and VOCs in aeration systems designed for the biological treatment of wastewater (see Sec. 16–4 in Chap. 16).

Analysis of Gas Stripping

Important factors that must be considered in the analysis of gas stripping include (1) the characteristics of the compound(s) to be stripped, (2) the type of contactor to be used and the required number of stages, (3) the materials mass balance analysis of the stripping tower, and (4) the required physical features and dimensions of the required stripping tower.

Characteristics of the Compound(s) to be Stripped. As noted above, the removal of volatile dissolved compounds by stripping involves contacting the liquid with a gas that does not contain the compound initially. The compound that is to be stripped will come out of solution and enter the gas phase to satisfy the Henry's law equilibrium as discussed in Chap. 2. Compounds such as benzene, toluene, and vinyl chloride which have Henry's law constants greater than 500 atm are especially amenable to stripping. Compounds with Henry's law constants values greater than 0.1 atm are classified as volatile and are considered amenable to stripping. Compounds with Henry's law constants between 0.001 and 0.1 atm are classified as semi-volatile and are marginally amenable to stripping. Compounds with Henry's law constants less than 0.001 atm are essentially not amenable to stripping.

The air stripping of ammonia from wastewater requires that the ammonia be present as a gas. Ammonium ions in wastewater exist in equilibrium with gaseous ammonia, as given by Eq. (2–38):

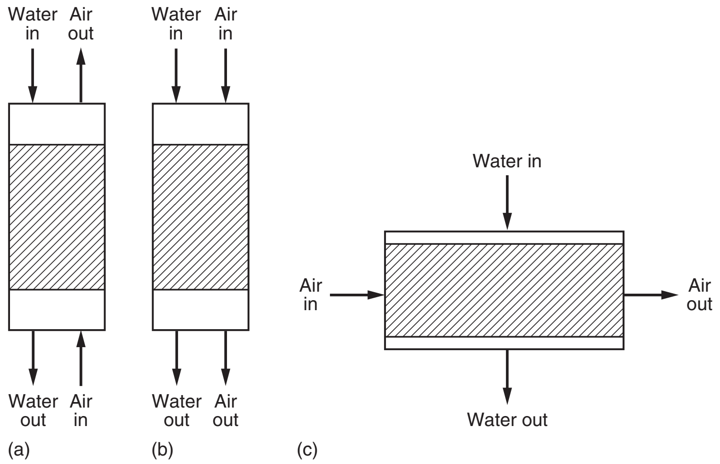
$$NH_4^+ \rightleftarrows NH_3 + H^+ \tag{2–38}$$

As the pH of the wastewater is increased above 7, the equilibrium is shifted to the right and the ammonium ion is converted to ammonia, which may be removed by gas stripping. The amount of lime required to raise the pH of wastewater to 11 as a function of the alkalinity is given on Fig. 6–12 in Chap. 6.

Methods Used to Contact Phases. In practice, two methods are used to achieve contact between phases so that mass transfer can occur: (1) continuous contact and (2) staged contact. As shown on Fig. 11–58, three flow patterns are used in practice: (1) cocurrent, (2) countercurrent, and (3) cross-flow. In addition, the contact packing may be fixed or mobile (Crittenden, 1999). The most common flow pattern in mass transfer operations is the countercurrent mode, in which the liquid to be stripped is pumped to the top of the tower and sprayed over the packing surface. Air is introduced into the bottom of the tower (i.e., counter current) and either blown or sucked up through the packing material. The packing material is used to distribute the applied liquid in a thin film to enhance the stripping process. In the cross-flow stripper, not used commonly, air is introduced along the side. One of the most critical issues in the design and operation of stripping towers is maintaining uniform airflow across the packing surface. To achieve a more uniform

Figure 11–58

Typical water and air flow patterns for gas stripping towers: (a) countercurrent flow, (b) cocurrent flow, and (c) cross-flow.



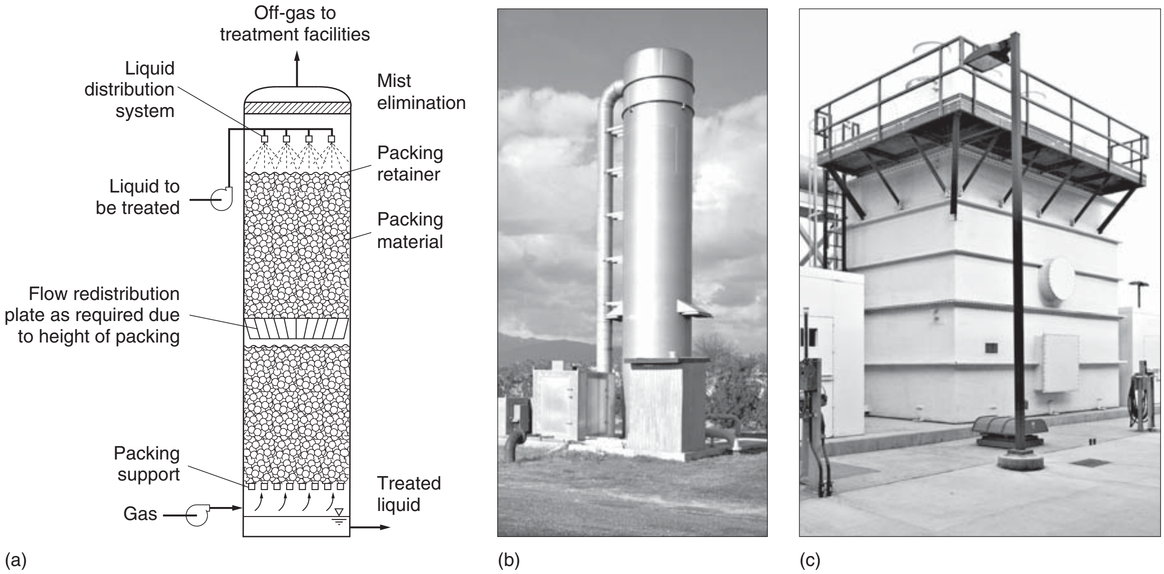


Figure 11–59

Typical examples of stripping towers: (a) schematic of packed bed stripping tower used for the removal of volatile gases from water, (b) view of stripping tower shown schematically in (a), and (c) typical stripping tower used for the removal of carbon dioxide from water following reverse osmosis treatment.

distribution of both air and water flow through the tower, packing is placed in individual stages on flow redistribution plates within the tower. A variety of packing materials are used in stripping towers including Raschig rings (cylinders), Berl saddles, or other proprietary plastic packings. Although a wide range of packing sizes is available, the most common size range is from 25 to 50 mm. A schematic and photograph of a typical gas stripping tower is shown on Fig. 11–59.

Mass Balance Analysis for a Continuous Stripping Tower. A steady-state materials balance for the lower portion of a countercurrent continuous stripping tower used for the removal of a dissolved gas from wastewater (see Fig. 11–60) is given by

1. General word statement:

$$\begin{array}{c} \text{Moles of solute} \\ \text{entering in} \\ \text{liquid stream} \end{array} + \begin{array}{c} \text{moles of solute} \\ \text{entering in} \\ \text{gas stream} \end{array} = \begin{array}{c} \text{moles of solute} \\ \text{leaving in} \\ \text{liquid stream} \end{array} + \begin{array}{c} \text{moles of solute} \\ \text{leaving in} \\ \text{gas stream} \end{array} \tag{11–72}$$

2. Simplified word statement:

$$\text{inflow} = \text{outflow} \tag{11–73}$$

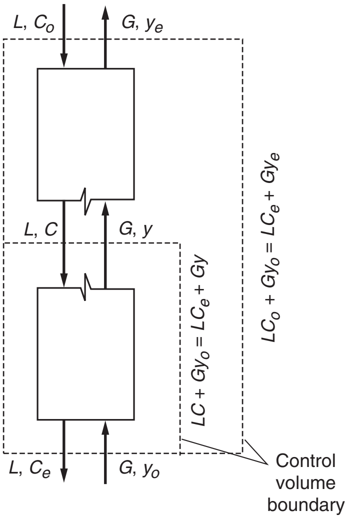
3. Symbolic representation (refer to Fig. 11–61):

$$LC + Gy_o = LC_e + Gy \tag{11–74}$$

where L = liquid flowrate, moles per unit time
C = concentration of solute in liquid at point within the tower, moles of solute per mole of liquid
G = gas flowrate, moles per unit time

Figure 11–60

Definition sketch for the analysis of a continuous countercurrent flow gas stripping tower.



y_o = concentration of solute in gas entering the bottom of the tower, moles of solute per mole of solute-free gas

C_e = concentration of solute in liquid leaving the bottom of the tower, moles of solute per mole of liquid

y = concentration of solute at a point within the tower, moles of solute per mole of solute-free gas

Figure 11–61

Operating lines for various gas stripping conditions: (a) general case, (b) condition when $y_o = 0$, (c) condition when $y_o = 0$ and y_e is in equilibrium with C_o, the constituent concentration in the incoming water, and (d) condition when $y_o = 0$, $C_e = 0$, and y_e is in equilibrium with C_o.

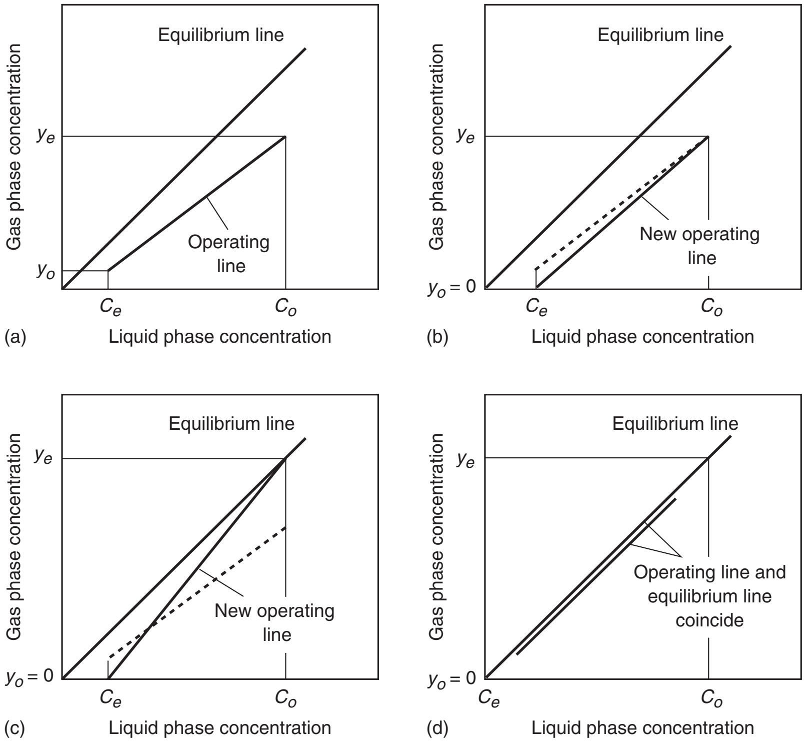
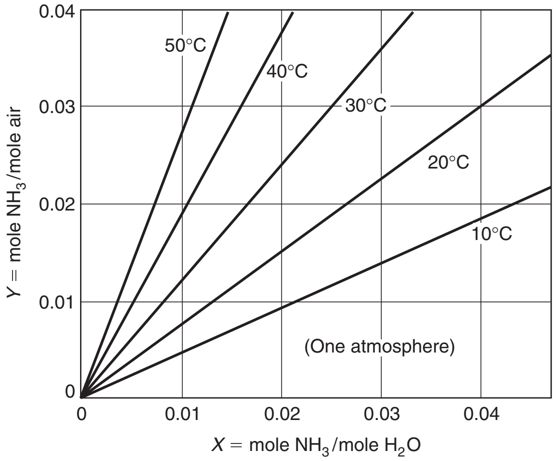


Figure 11–62

Equilibrium curves for ammonia in water as a function of temperature based on Henry's law.



Combining terms, Eq. (11–74) can be written as:

$$(y_o - y) = L/G(C_e - C) \tag{11–75}$$

If the overall tower is considered, Eq. (11–74) can be written as:

$$LC_o + Gy_o = LC_e + Gy_e \tag{11–76}$$

Combining terms, Eq. (11–76) can be written as:

$$(y_o - y_e) = L/G(C_e - C_o) \tag{11–77}$$

where C_o = concentration of solute in liquid entering at the top of the tower, moles of solute per mole of liquid

y_e = concentration of solute in gas leaving the top of the tower, moles of solute per mole of gas

Because Eq. (11–77) is derived solely from a consideration of the equality of input and output, it holds regardless of the internal equilibria that may control the mass transfer. Equation (11–77) represents the equation of a straight line with slope L/G which passes through the point (C_o, y_e) and point (C_e, y_o). The line passed through these two points [see Fig. 11–61(a)] is known as the *operating line* and represents the conditions at any point within the column. The equilibrium line is based on Henry's law. For example, equilibrium lines defined by Henry's law for ammonia as a function of temperature are presented on Fig. 11–62. It should be noted that when a gas is being stripped from solution, the operating line will lie below the equilibrium line. If a gas is being absorbed into solution the operating line will lie above the equilibrium line.

If it is assumed that the air entering the bottom of the tower contains no solute (i.e, $y_o = 0$), then Eq. (11–77) can be written as

$$y_e = L/G(C_o - C_e) \tag{11–78}$$

The new operating line for the condition defined by Eq. (11–78) is shown on Fig. 11–61(b). Using Henry's law [see Eq. (2–46)], y_e is defined as follows

$$y_e = \frac{H}{P_T} C'_o \tag{11–79}$$

where y_e = concentration of solute in gas leaving the top of the tower, moles of solute per mole of gas

H = Henry's law constant, $\dfrac{\text{atm (mole gas/mole air)}}{\text{(mole gas/mole water)}}$

P_T = total pressure, usually 1.0 atm

C'_o = the concentration of solute in liquid that is in equilibrium with the gas leaving the tower, moles of solute per mole of liquid

Using Eq. (11–79), Eq. (11–78) can be written as follows:

$$C'_o = \frac{P_T}{H} \times \frac{L}{G}(C_o - C_e) \tag{11–80}$$

If it is assumed that the concentration of solute in the liquid entering the tower is in equilibrium with the gas leaving the tower, Eq. (11–80) can be written as:

$$\frac{G}{L} = \frac{P_T}{H} \times \frac{(C_o - C_e)}{C_o} \tag{11–81}$$

The operating line for the condition defined by Eq. (11–81) is shown on Fig. 11–61(c). The value of G/L (air to liquid ratio) defined by Eq. (11–81) represents the minimum amount of air that can be used for stripping for the given conditions (i.e, $y_o = 0$ and $y_e = HC_o/P_T$). In practice, from one and a half to three times the theoretical minimum air to liquid ratio is used to achieve effective stripping of most constituents. The application of this relationship is illustrated in Example 11–15.

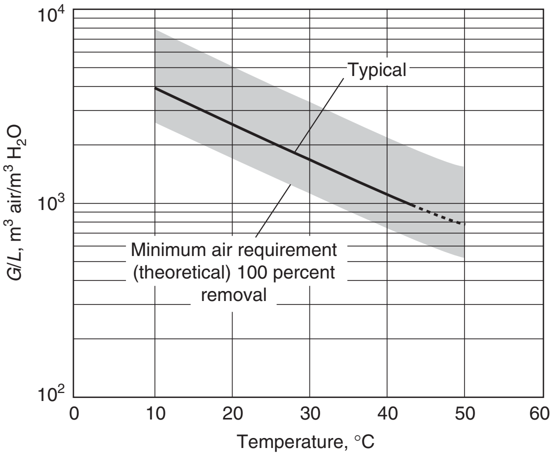
If it is assumed further that the liquid leaving and the air entering the bottom of the tower contains no solute, then Eq. (11–81) can be written as

$$\frac{G}{L} = \frac{P_T \times C_o}{H \times C_o} = \frac{P_T}{H} \tag{11–82}$$

The value of G/L for this condition corresponds to the equilibrium line defined by Henry's law [see Fig. 11–61(d)], and represents the theoretical minimum amount of air that can be used for stripping for the given conditions (i.e., $y_o = 0$, $C_e = 0$, and $y_e = HC_o/P_T$). The range of air-to-liquid ratios for stripping ammonia from wastewater as a function of temperature is plotted on Fig. 11–63. The theoretical ratio is derived by assuming the process

Figure 11–63

Air requirements for ammonia stripping as function of temperature.



to be 100 percent efficient with a stripping tower of infinite height—obviously unachievable in practice. Computation of the theoretical air to water ratio is illustrated in Example 11–15.

EXAMPLE 11–15 **Air Requirements for Ammonia Stripping** Determine the theoretical amount of air required at 20°C to reduce the ammonia concentration from 40 to 1 mg/L in a treated wastewater with a flowrate of 4000 m³/d. Assume that the Henry's constant for ammonia at 20°C is 0.75 atm (see Table 2–7 in Chap. 2) and the air entering the bottom of the tower does not contain any ammonia.

Solution

1. Determine the influent and effluent mole fractions of ammonia in the liquid using Eq. (2–3).

$$x_B = \frac{n_B}{n_A + n_B}$$

where x_B = mole fraction of solute B
n_B = number of moles of solute B
n_A = number of moles of solute A

$$C_o = \frac{[(40 \times 10^{-3})/17]}{[55.5 + (40 \times 10^{-3})/17]} = 4.24 \times 10^{-5} \text{ mole NH}_3/\text{mole H}_2\text{O}$$

$$C_e = \frac{[(1 \times 10^{-3})/17]}{[55.5 + (1 \times 10^{-3})/17]} = 1.06 \times 10^{-6} \text{ mole NH}_3/\text{mole H}_2\text{O}$$

2. Determine the mole fraction of ammonia in the air leaving the tower using Eq. (11–79).

$$y_e = \frac{H}{P_T} C_o$$

$$H = \left[\frac{(0.75 \text{ atm})(\text{mole NH}_3/\text{mole air})}{(\text{mole NH}_3/\text{mole H}_2\text{O})}\right] = (0.75 \text{ atm})\left(\frac{\text{mole H}_2\text{O}}{\text{mole air}}\right)$$

$$y_e = \frac{H}{P_T} \times C_o = \frac{0.75 \text{ atm}}{1.0 \text{ atm}}\left(\frac{\text{mole H}_2\text{O}}{\text{mole air}}\right) \times (4.24 \times 10^{-5}) \text{ mole NH}_3/\text{mole H}_2\text{O}$$

$$= 3.18 \times 10^{-5} \frac{\text{mole NH}_3}{\text{mole air}}$$

3. Determine the gas to liquid ratio using Eq. (11–81) rearranged as follows:

$$\frac{G}{L} = \frac{P_T}{H} \times \frac{(C_o - C_e)}{C_o} = \frac{(C_o - C_e)}{y_e}$$

$$\frac{G}{L} = \frac{(4.24 \times 10^{-5} - 0.106 \times 10^{-5})(\text{mole NH}_3/\text{mole H}_2\text{O})}{(3.18 \times 10^{-6} \text{ mole NH}_3/\text{mole air})} = 1.3 \frac{\text{mole air}}{\text{mole H}_2\text{O}}$$

4. Convert the moles of air and water to liters of air and water.

   For air at 20°C:

   $$1.3 \text{ mole} \times 24.1 \text{ L/mole} = 31.33 \text{ L}$$

   For water:

   $$(1.0 \text{ mole } H_2O)(18 \text{ g/mole})(1\text{L}/1000 \text{ g}) = 0.018 \text{ L}$$

   $$\frac{G}{L} = \frac{31.33 \text{ L air}}{0.018 \text{ L water}} = 1741 \text{ L/L} = 1741 \text{ m}^3/\text{m}^3$$

6. Determine the total quantity of air required based on ideal conditions.

   $$\text{Air required} = \frac{(1741 \text{ m}^3/\text{ m}^3)\ (4000 \text{ m}^3/\text{d})}{(1440 \text{ min/d})} = 4835 \text{ m}^3/\text{min}$$

Comment The procedure followed to determine the height of the stripping tower is illustrated in Example 11–14 presented later in this section. Also, it should be noted that the ammonium must first be converted to ammonia gas for this process to be effective. Steam stripping of ammonia is considered in Chap. 15.

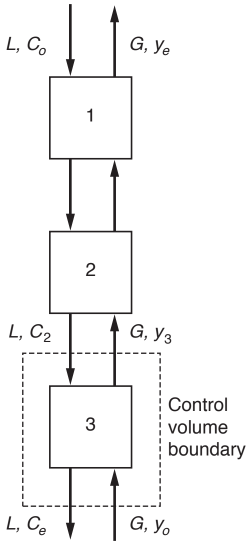


Figure 11–64

Definition sketch for the analysis of a three stage countercurrent flow gas stripping tower.

Mass Balance Analysis of a Multistage Stripping Tower. In the analysis of stripping towers, reference is often made to the number of ideal stages required for stripping. The analysis for the number of stages is analogous to the simulation of plug flow with a series of complete-mix reactors as detailed in Chap. 1. Separate stages are used to improve performance of stripping towers. Equilibrium conditions are assumed in each stage of the tower. A steady-state materials balance for the lower portion of a countercurrent staged stripping tower (see Fig. 11–64) is given by:

inflow = outflow

$$LC_e + Gy_3 = LC_2 + Gy_o \tag{11–83}$$

or

$$(y_3 - y_o) = L/G(C_2 - C_e) \tag{11–84}$$

If an overall mass balance is performed around the tower, the resulting equations are the same as derived above for the continuous stripping tower [see Eq. (11–76)].

In the 1920s McCabe and Theile (1925) developed a graphical procedure for determining the required number of ideal stages. The method is illustrated on Fig. 11–65 for a stripping tower comprised of three stages. The operating line for the three stages is shown on Fig. 11–65. The number of ideal stages required for stripping of a constituent is obtained as follows. The point C_o, y_e represents the air leaving and the water entering the top of the stripping column. The composition of the liquid in equilibrium with the constituent concentration in the air is found by extending a horizontal line from the

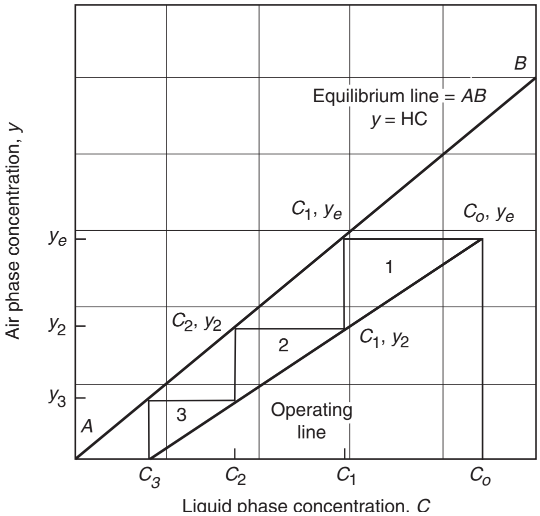


Figure 11–65

Operating line for three-stage countercurrent flow gas stripping tower.

point C_o, y_e to the point C_1, y_e. From the point C_1, y_e the value of y_2, the air entering stage 1 from stage 2, is obtained from the equation of the operating line. If a materials mass balance is performed between stages 1 and 2, the resulting expression for y_2 is:

$$y_2 = \frac{L}{G}C_1 + \frac{Gy_e - LC_o}{G} \tag{11–85}$$

The value of y_2 is obtained by drawing a vertical line from point C_1, y_e to the operating line at point C_1, y_2 as shown on Fig. 11–65. In a similar manner the value of C_2 is obtained by drawing a horizontal line from the point C_1, y_2 to the equilibrium line. This procedure is repeated until the point C_n, y_{n+1} is reached. The number of ideal stages is typically a fractional number (e.g., 4.2, 5.6, etc.). In practice, the number of stages is rounded to the next whole number.

Determination of Height of Stripping Tower Packing. The purpose of the following analysis is to illustrate how the height of a stripping tower packing is determined, based on an analysis of the mass transfer occurring within the tower. A mass balance performed on the liquid phase within the stripping tower shown on Fig. 11–66 is as follows.

Simplified word statement

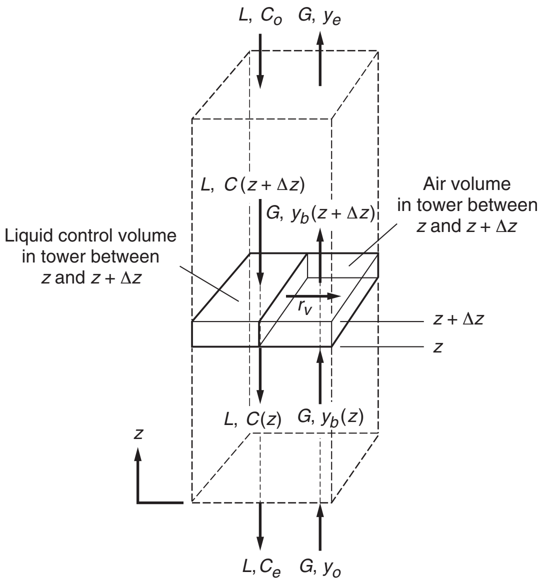
Accumulation = inflow − outflow + generation

$$\frac{\partial C}{\partial t}\Delta V = LC\big|_z - LC\big|_{z+\Delta z} + r_V \Delta V \tag{11–86}$$

where $\partial C/\partial t$ = change in concentration of constituent C with time, g/m^3·s
ΔV = differential volume, m^3
Δz = differential height, m
L = liquid volumetric flowrate, m^3/s
C = concentration of constituent C, g/m^3
r_V = rate of mass transfer of constituent C per unit volume per unit time, g/m^3·s

Figure 11–66

Definition sketch for the analysis of mass transfer within a stripping tower. Note: the packing material is not shown. (Adapted from Hand et al., 1999.) (LD)



Substituting area times the differential height ($A\Delta z$) for the differential volume (ΔV) and writing the differential form for the term $LC|_{Z+\Delta Z}$ in Eq. (11–86) results in the following expression:

$$\frac{\partial C}{\partial t}A\Delta z = LC - L\left(C + \frac{\Delta C}{\Delta z}\Delta z\right) + r_V A\Delta z \tag{11–87}$$

Simplifying Eq. (11–87) and taking the limit as Δz approaches zero yields

$$\frac{\partial C}{\partial t} = -\frac{L}{A}\frac{\partial C}{\partial z} + r_V \tag{11–88}$$

The rate of mass transfer as described in Chap. 5 [see Eq. (5–57)]:

$$r_V = K_L a(C_b - C_s) \tag{11–89}$$

where r_V = rate of mass transfer of constituent C per unit volume per unit time, g/m³·s
K_La = volumetric mass transfer coefficient which depends on water quality characteristics and temperature, 1/s
C_b = concentration of constituent C in liquid bulk phase at time t, g/m³
C_S = concentration of constituent C in liquid in equilibrium with gas as given by Henry's law, g/m³

Assuming steady-state conditions within the tower ($\partial C/\partial t = 0$) and substituting for r_V, Eq. (11–88) can now be written as:

$$\frac{dC_b}{dz} = \frac{K_L aA}{L}(C_b - C_s) \tag{11–90}$$

The height of the tower can be obtained by integrating the above expression:

$$\int_o^Z dz = \frac{L}{K_L aA} \int_{C_e}^{C_o} \frac{dC_b}{(C_b - C_s)} \tag{11–91}$$

To integrate the right hand side of the above equation, a relationship must be found between C_b and C_s because C_s is changing continuously throughout the height of the tower. From Henry's law, the value of C_s is given by:

$$C_s = \frac{P_T}{H} y \tag{11–92}$$

Substituting a modified form of Eq. (11–78) for y in Eq. (11–92) yields:

$$C_s = \frac{P_T}{H} \times \frac{L}{G}(C_b - C_e) \tag{11–93}$$

If Eq. (11–93) is substituted into Eq. (11–91) and the resulting expression is integrated, the following expression is obtained (Hand et al., 1999)

$$Z = \frac{L}{K_L a A}\left(\frac{C_o - C_e}{C_o - C_e - C_o'}\right) \text{In} \left(\frac{C_o - C_o'}{C_e}\right) \tag{11–94}$$

where Z = height of stripping tower packing, m

$C_o' = \frac{P_T}{H} \times \frac{L}{G}(C_o - C_e)$, given previously [see Eq. (11–80)]

It should be noted that if $C_o' = C_o$, Eq. (11–80) is the same as Eq. (11–81).

Design Equations for Stripping Towers. Utilizing the above equations, a number of process models and design equations have been developed. Equations that can be used to determine the height of a stripping tower are as follows:

$$Z = \text{HTU} \times \text{NTU} \tag{11–95}$$

where Z = height of stripping tower packing material, m
HTU = height of a transfer unit, m
NTU = number of transfer units

The height of a transfer unit is defined as:

$$\text{HTU} = \frac{L}{K_L a A} \tag{11–96}$$

where L = liquid volumetric flowrate, m^3/s
$K_L a$ = volumetric mass transfer coefficient, 1/s
A = cross-sectional area of tower, m^2

The HTU is a measure of the mass transfer characteristics of the packing material. The number of transfer units is defined as:

$$\text{NTU} = \left(\frac{C_o - C_e}{C_o - C_e - C_o'}\right) \text{In} \left(\frac{C_o - C_o'}{C_e}\right) \tag{11–97}$$

Substituting Eq. (11–93) in Eq (11–97) yields:

$$\text{NTU} = \left(\frac{S}{S - 1}\right) \text{In} \left[\frac{(C_o/C_e)(S - 1) + 1}{S}\right] \tag{11–98}$$

where S is known as the stripping factor and is defined as:

$$S = \frac{G}{L} \times \frac{H}{P_T} \tag{11–99}$$

A value of $S = 1$ corresponds to the minimum amount of air required for stripping. When $S > 1$ the amount of air is in excess and complete stripping is possible given a tower of infinite height. When $S < 1$, there is insufficient air for stripping. In practice, stripping factors vary from 1.5 to 5.0.

Values for $K_L a$ for specific compounds are best obtained from pilot plant studies or by using empirical correlations such as given in Chap. 16, and repeated here for convenience as Eq. (11–100). It should be noted that many other relationships have been proposed in the literature (Sherwood and Hollaway, 1940 and Onda et al., 1968).

$$K_L a_{\text{VOC}} = K_L a_{O_2} \left(\frac{D_{\text{VOC}}}{D_{O_2}}\right)^n \tag{11–100}$$

where $K_L a_{\text{VOC}}$ = system mass transfer coefficient, 1/h
$K_L a_{O_2}$ = system oxygen mass transfer coefficient, 1/h
D_{VOC} = diffusion coefficient of VOC in water, cm^2/s
D_{O_2} = diffusion coefficient of oxygen in water, cm^2/s
n = coefficient (0.5 for stripping towers)

Air and water temperature are significant factors in the design of stripping towers because of their effect on air and water viscosities, Henry's law constants, and volumetric mass transfer coefficients. The effect of temperature on the Henry's law constant is illustrated on Fig. 11–62. The value of $K_L a$ can be adjusted for temperature effects using Eq. (1–44) with a theta value of 1.024.

Design of Stripping Towers

In its simplest form a stripping tower consists of a tower (usually circular cross-section), a support plate for the packing material, a distribution system for the liquid to be stripped, located above the packing material, and an air supply located at the bottom of the stripping tower (see Fig. 11–59). The process design variables include (1) the type of packing material, (2) the stripping factor, (3) the cross-sectional area of the tower, and (4) the height of the packing material in the stripping tower. The cross-sectional area will depend on the pressure drop through the packing. Representative design values for stripping of VOCs and ammonia are presented in Table 11–45. The significant difference in the amount of air required for stripping is a clear illustration of the importance of the Henry's law constant.

Table 11–45

Typical design parameters for stripping towers for the removal of VOC and ammonia[a]

Item	Symbol	Unit	VOC removal[b]	Ammonia removal[c]
Liquid loading rate		$L/m^2 \cdot min$	600–1,800	40–80
Air to liquid ratio[d]	G/L	m^3/m^3	20–60:1	2,000–6,000:1
Stripping factor	S	unitless	1.5–5.0	1.5–5.0
Allowable air pressure drop,	ΔP	$(N/m^2)/m$	100–400	100–400
Height to diameter ratio	H/D	m/m	$\leq$10:1	$\leq$10:1
Packing depth[e]	D	m	1–6	2–6
Factor of safety	SF	%D, %H	20–50	20–50
Wastewater pH	pH	unitless	5.5–8.5	10.8–11.5
Approximate packing factors				
Pall rings, Intalox saddles				
12.5 mm[f]	C_f	1/m	180–240	180–240
25 mm[f]	C_f	1/m	30–60	30–60
50 mm[f]	C_f	1/m	20–25	20–25
Berl saddles, Raschig rings				
12.5 mm[f]	C_f	1/m	300–600	300–600
25 mm[f]	C_f	1/m	120–160	120–160
50 mm[f]	C_f	1/m	45–60	45–60

[a] Adapted in part from Eckert (1970, 1975), Kavanaugh and Trussell (1980), and Hand (1999).

[b] Typical data for VOCs with Henry's law constants greater than 500 atm (mole H_2O/mole air).

[c] Ammonia with a Henry's law constant of 0.75 atm (mole H_2O/mole air) is considered only marginally strippable, which accounts for the low loading rate and high air to liquid ratio.

[d] Ratio is highly temperature dependent.

[e] For packing depths greater than 5 to 6 m, redistribution of the liquid flow is recommended.

[f] Size of packing material.

The headloss through the packing is determined using a generalized gas pressure drop relationships such as plotted on Fig. 11–67 (Eckert, 1975). The pressure drop is expressed in Newton per square meter per meter of depth $(N/m^2)/m$. The upper line on Fig. 11–67 labeled *approximate flooding* represents the condition that occurs when the amount of water and air applied are so great that the pore spaces fill to the point where water starts to flood within the tower. The units for the x and y axis are as follows

X axis:

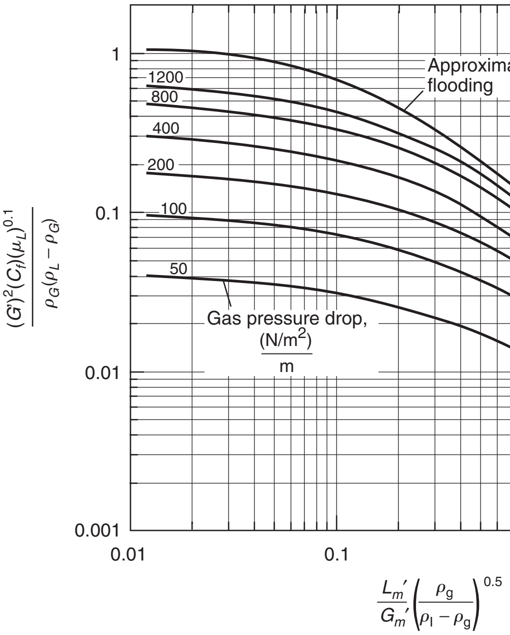
$$x = \frac{L'}{G'}\left(\frac{\rho_G}{\rho_L - \rho_G}\right)^{1/2} \approx \frac{L'}{G'}\left(\frac{\rho_G}{\rho_L}\right)^{1/2} \tag{11–101}$$

Y axis:

$$y = \frac{(G')^2(C_f)(\mu_L)^{0.1}}{(\rho_G)(\rho_L - \rho_G)} \tag{11–102}$$

Figure 11–67

Generalized pressure drop curves for packed bed stripping towers. Note the curves in this plot have been converted to metric units from US customary units in which the original curves were plotted. (Adapted from Eckert, 1975.)



Which can be rewritten as follows:

$$G' = \left[\frac{(\text{value from } y \text{ axis})(\rho_G)(\rho_L - \rho_G)}{(C_f)(\mu_L)^{0.1}}\right]^{1/2} \tag{11–103}$$

where L' = liquid loading rate, kg/m^2·s
G' = gas loading rate, kg/ m^2·s
ρ_G = density of gas, kg/m^3
ρ_L = density of liquid, kg/m^3
C_f = packing factor for packing material, 1/m
μ_L = viscosity of liquid, kg/m·s

The packing factor C_f depends on the type and size of the packing. Typical ranges for packing factors that can be used for preliminary assessments are reported in Table 11–35. For more detailed design calculations, current values should be obtained from manufacturers.

To use Fig. 11–67, select a value for G'/L' and compute the corresponding x value. Enter the plot at the computed value of x and move vertically upward to a preselected pressure drop line. Move horizontally from the point of intersection to the y axis and note the value on the y axis. Using the y-axis value, determine the gas loading rate, G', using Eq. (11–103) and the corresponding liquid loading rate, L'. To determine the required cross-sectional area, the liquid flowrate is divided by the liquid loading rate.

A generalized analysis procedure is as follows:

1. Select a packing material and its corresponding packing factor for use in Eq. (11–101).
2. Select several stripping factors for successive trials (e.g., 2.5, 3.0, 4.0, etc.).
3. Select acceptable pressure drop ΔP (typically a function of the packing material selected).
4. Determine the cross-sectional area of the tower, based on the allowable pressure drop, using the data presented on Fig. 11–67 or other appropriate relationships.

5. Determine the height of the transfer units using Eq. (11–96). To apply Eq. (11–96) the value of K_La must be known or estimated using Eq. (11–100).
6. Determine the number of transfer units using Eq. (11–98).
7. Determine the height of the stripping tower packing material using Eq. (11–95).
8. Determine the total height of the stripping tower. To account for the entrance plenum and exit gas collection system, an additional 2 to 3 m is added to the computed height of the packing to obtain the overall height of the stripping tower.

The design procedure outlined above is illustrated in Example 11–16. Representative design values for stripping towers are given in Table 11–45. To evaluate the stripping process more thoroughly, any of the commercially available software packages can be used.

EXAMPLE 11–16 **Determination of Height of Stripping Tower for the Removal of Ammonia** Determine the diameter and height of the stripping tower required to treat the wastewater in Example 11–15. The ammonia concentration in a treated wastewater from a flow of 4000 m^3/d is to be reduced from 40 to 1 mg/L. Assume that the Henry's law constant for ammonia at 20°C is 0.75 atm, and the air entering the bottom of the tower does not contain any ammonia. Assume the K_La value for ammonia is 0.0125 s^{-1}.

Solution

1. Select a packing material. Assume a packing factor of 20 for 50 mm Pall rings (see Table 11–45).
2. Select a stripping factor. Assume a stripping factor of 3.
3. Select an acceptable pressure drop. Assume a pressure drop of 400 $(N/m^2)/m$ (see Table 11–45).
4. Determine the cross-sectional area of the stripping tower using the pressure drop plot given on Fig. 11–67.
   a. Determine the value of the ordinate value for a stripping factor of 3.

$$S = \frac{G}{L} \times \frac{H}{P_T} = \frac{G \text{ mole air}}{L \text{ mole water}} \times \frac{0.75 \text{ atm}}{1.0 \text{ atm}} = \frac{G \text{ mole air}}{L \text{ mole water}} \times 0.75$$

$$S = 0.75 \times \left(\frac{G \text{ mole air}}{L \text{ mole water}}\right)\left(\frac{28.8 \text{ g}}{\text{mole air}}\right)\left(\frac{\text{mole water}}{18 \text{ g}}\right) = 1.2\frac{G \text{ g}}{L \text{ g}} = 1.2\frac{G' \text{ kg}}{L' \text{ kg}}$$

$$\frac{L'}{G'} = \frac{(1.2 \text{ kg/kg})}{3} = 0.4$$

$$\frac{L'}{G'}\left(\frac{\rho_G}{\rho_L - \rho_G}\right)^{1/2} \approx \frac{L'}{G'}\left(\frac{\rho_G}{\rho_L}\right)^{1/2} = (0.4 \text{ kg/kg})\left[\frac{(1.204 \text{ kg/m}^3)}{(998.2 \text{ kg/m}^3)}\right]^{1/2} = 0.0139$$

   b. For an abscissa value of 0.0139 and a pressure drop of 400 $(N/m^2)/m$, the ordinate value from Fig. 11–67 is 0.3.
   c. Using an ordinate value 0.3, determine the loading rate using Eq. (11–103). From Appendix *C*, $\mu_L = 1.002$ kg/m·s

$$G' = \left[\frac{(\text{value from } y \text{ axis})(\rho_G)(\rho_L - \rho_G)}{(C_f)(\mu_L)^{0.1}}\right]^{1/2}$$

$$G' = \left[\frac{(0.3)(1.204)(998.2 - 1.204)}{(20)(1.002)^{0.1}}\right]^{1/2} = 4.24 \text{ kg/m}^2\text{·s}$$

$L' = 0.4\,G' = 0.4 \times 4.24 \text{ kg/m}^2\text{·s} = 1.70 \text{ kg/m}^2\text{·s}$

d. Solve for the diameter of the tower, using the loading rate determined in Step 4.

$$D = \left[\frac{4}{3.14} \times \frac{(4000 \text{ m}^3/\text{d})(998.2 \text{ kg/m}^3)}{(4.24 \text{ kg/m}^2\text{·s})} \times \frac{1 \text{ d}}{86{,}400 \text{ s}}\right]^{1/2} = 3.73 \text{ m}$$

5. Determine the height of the transfer unit using Eq. (11–96).

$$\text{HTU} = \frac{L}{K_L a A}$$

$$\text{HTU} = \frac{L}{K_L a A} = \left[\frac{(4000 \text{ m}^3/\text{d})}{(0.0125/\text{s})[(3.14/4)(3.73)^2]} \times \frac{1 \text{ d}}{86{,}400 \text{ s}}\right] = 0.34 \text{ m}$$

6. Determine the number of transfer units using Eq. (11–98).

$$\text{NTU} = \left(\frac{S}{S - 1}\right) \ln\left[\frac{(C_o/C_e)(S - 1) + 1}{S}\right]$$

$$\text{NTU} = \left(\frac{3}{3 - 1}\right) \ln\left[\frac{(40/1)(3 - 1) + 1}{3}\right] = 4.94$$

7. Determine the theoretical height of the stripping tower packing using Eq. (11–95).

$Z = \text{HTU} \times \text{NTU} = 0.34 \times 4.94 = 1.68 \text{ m}$

8. Determine the total height of the stripping tower.

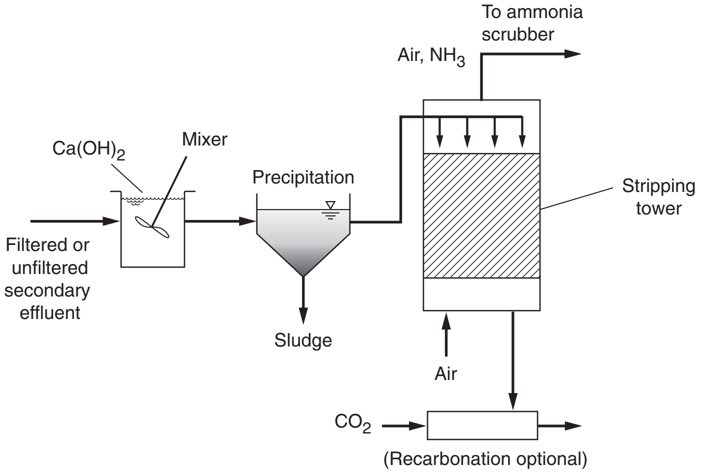
Add 3 m to obtain the total height of the stripper

Hstripper = Hpacking, m + 3 m = 1.68 m + 3 m = 4.68 m

Comment In this example, the value of $K_L a$ for ammonia was known. Quite often the required $K_L a$ value must be determined in the field, using pilot scale facilities. Alternatively Eq. (11–100) can be used to estimate a value for $K_L a$. In some cases, data from the literature or from manufacturers may be used to obtain preliminary sizing. Because of the relatively low Henry's law constant, a large surface area is required to achieve the desired ammonia removal. Further, because of the relatively large radius that is required, it is likely that two stripping towers, each with a radius and area of about 2.6 and 5.3 m^2, respectively would be used to optimize air flow. To optimize the design, various stripping ratios must be evaluated. Optimization is best accomplished using one of the commercially available stripping tower software packages. Because the Henry's law constant for ammonia is so low (0.75 atm), ammonia is often not considered amenable or suitable for stripping at ambient temperatures and is seldom done. Steam stripping of the concentrated ammonia present in return flows from sludge processing is considered in Chap. 15.

Figure 11–68

Typical flow diagram for the air stripping of ammonia from wastewater.



Air Stripping Applications

As noted previously, air stripping is used to remove a variety of gaseous constituents including VOCs, carbon dioxide (CO_2), oxygen (O_2), hydrogen sulfide (H_2S), and ammonia (NH_3). The removal and treatment of VOCs in aeration systems is considered in Sec. 16–4 in Chap. 16. The removal and treatment of odorous compounds (e.g., H_2S) is considered in Sec. 16–3 in Chap. 16. Air stripping has been used for the removal of ammonia from various wastewaters including untreated and treated wastewater, digester supernatant (also by steam stripping), and for the recovery of ammonia from return flows. The recovery of ammonia from return flows is considered in detail in Chap. 15. Ammonia stripping from wastewater is considered briefly below.

A typical flow diagram for the removal of ammonia from wastewater by air stripping is shown on Fig. 11–68. In most cases where ammonia stripping has been tried with wastewater, a number of operating problems have developed, the most serious being (1) maintaining the required pH for effective stripping, (2) calcium carbonate scaling within the tower and feed lines, and (3) poor performance during cold weather operation. Maintaining the required pH is a control problem that can be managed with multiple sensors. The amount and nature (soft to extremely hard) of the calcium carbonate scale formed varies with the characteristics of the wastewater and local environmental conditions and cannot be predicted a priori. Under conditions of icing, the liquid-air contact geometry in the tower is altered, which further reduces the overall efficiency. The best solution for cold weather conditions is to enclose the stripper. For the reasons cited above and cost, ammonia stripping from wastewater is seldom done. However, ammonia stripping is done on concentrated return flows resulting from the treatment of biosolids. The use of ammonia stripping for the recovery of nitrogen in the form of ammonium sulfate is considered in Chap. 15.

11–11 ION EXCHANGE

Ion exchange is a unit process in which ions of a given species are displaced from an insoluble exchange material by ions of a different species in solution. The most widespread use of this process is in domestic water softening, where sodium ions from a cationic exchange resin replace the calcium and magnesium ions in the treated water, thus reducing the hardness. Ion exchange has been used in wastewater applications for the removal of nitrogen, heavy metals, and total dissolved solids.

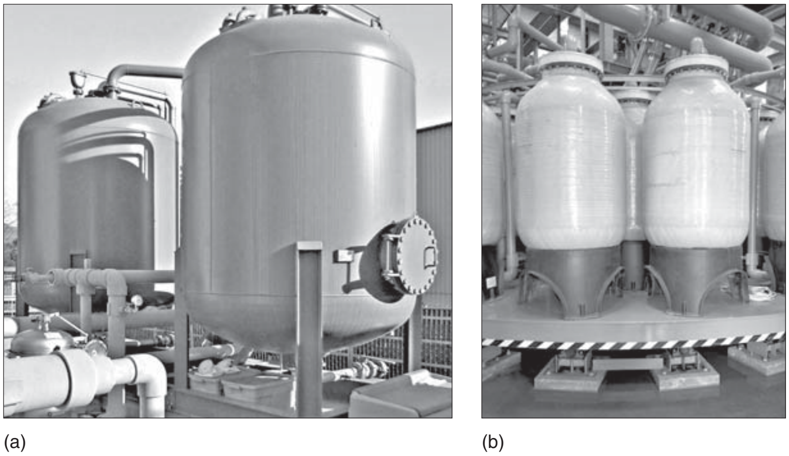


Figure 11–69

Two examples of full scale ion exchange installations: (a) large downflow packed-bed columns, and (b) ion exchange canisters on a rotating platform. The canisters are rotated so that one canister can be regenerated while the others are in operation.

(a) (b)

Ion exchange processes can be operated in a batch or continuous mode. In a batch process, the resin is stirred with the water to be treated in a reactor until the reaction is complete. The spent resin is removed by settling and subsequently is regenerated and reused. In a continuous process, the exchange material is placed in a bed or a packed column, similar to the one shown previously on Fig. 11–56(a), and the water to be treated is passed through it. Continuous ion exchangers are usually of the downflow, packed-bed column type. Wastewater enters the top of the column under pressure, passes downward through the resin bed, and is removed at the bottom. When the resin capacity is exhausted, the column is backwashed to remove trapped solids and is then regenerated. Two examples of commercial ion exchange reactors are shown on Fig. 11–69.

Ion Exchange Materials

Naturally occurring ion exchange materials, known as zeolites, are used for water softening and ammonium ion removal. Zeolites used for water softening are complex aluminosilicates with sodium as the mobile ion. Ammonium exchange is accomplished using a naturally occurring zeolite, clinoptilolite. Synthetic aluminosilicates are manufactured, but most synthetic ion exchange materials are resins or phenolic polymers. Five types of synthetic ion exchange resins are in use: (1) strong-acid cation, (2) weak-acid cation, (3) strong-base anion, (4) weak-base anion, and (5) heavy-metal selective chelating resins. The properties of these resins are summarized in Table 11–46.

Most synthetic ion exchange resins are manufactured by a process in which styrene and divinylbenzene are copolymerized. The styrene serves as the basic matrix of the resin and divinylbenzene is used to cross link the polymers to produce an insoluble tough resin. Important properties of ion exchange resins include exchange capacity, particle size, and stability. The exchange capacity of a resin is defined as the quantity of an exchangeable ion that can be taken up. The exchange capacity of resins is expressed as eq/L or eq/kg (meq/L or meq/g). The particle size of a resin is important with respect to the hydraulics of the ion exchange column and the kinetics of ion exchange. In general, the rate of exchange is proportional to the inverse of the square of the particle diameter. The stability of a resin is important to the long-term performance of the resin. Excessive osmotic swelling and shrinking, chemical degradation, and structural changes in the resin caused by physical stresses are important factors that may limit the useful life of a resin.

Table 11–46

Classification of ion exchange resins[a]

Type of resin	Characteristics
Strong-acid cation resins	Strong-acid resins behave in a manner similar to a strong-acid, and are highly ionized in both the acid (R–SO_3H) and salt (R–SO_3Na) form, over the entire pH range.
Weak-acid cation resins	Weak-acid resins have a weak-acid functional group (–COOH), typically a carboxylic group. These resins behave like weak organic acids that are weakly dissociated.
Strong-base anion resins	Strong-base resins are highly ionized having strong-base functional groups such as (OH^-); and can be used over the entire pH range. These resins are used in the hydroxide (OH^-) form for water deionization.
Weak-base anion resins	Weak-base resins have weak-base functional groups in which the degree of ionization is dependent on pH.
Heavy-metal selective chelating resins	Chelating resins behave like weak-acid cation resins, but exhibit a high degree of selectivity for heavy-metal cations. The functional group in most of these resins is EDTA, and the resin structure in the sodium form is R–EDTA–Na.

[a] Adapted in part from Ford (1992).

Typical Ion Exchange Reactions

Typical ion exchange reactions for natural and synthetic ion exchange materials are given below.

For natural zeolites used in water softening (Z):

$$ZNa_2 + \begin{bmatrix} Ca^{2+} \\ Mg^{2+} \\ Fe^{2+} \end{bmatrix} \rightleftarrows Z \begin{bmatrix} Ca^{2+} \\ Mg^{2+} \\ Fe^{2+} \end{bmatrix} + 2Na^+ \tag{11–104}$$

For synthetic resins (R):

Strong-acid cation exchange:

$$RSO_3H + Na^+ \rightleftarrows RSO_3Na + H^+ \tag{11–105}$$

$$2RSO_3Na + Ca^{2+} \rightleftarrows (RSO_3)_2Ca + 2Na^+ \tag{11–106}$$

Weak-acid cation exchange:

$$RCOOH + Na^+ \rightleftarrows RCOONa + H^+ \tag{11–107}$$

$$2RCOONa + Ca^{2+} \rightleftarrows (RCOO)_2Ca + 2Na^+ \tag{11–108}$$

Strong-base anion exchange:

$$RR'_3NOH + Cl^- \rightleftarrows RR'_3NCl + OH^- \tag{11–109}$$

Weak-base anion exchange:

$$RNH_3OH + Cl^- \rightleftarrows RNH_3Cl + OH^- \tag{11–110}$$

$$2RNH_3OCl + SO_4^{2-} \rightleftarrows (RNH_3)_2SO_4 + 2Cl^- \tag{11–111}$$

Exchange Capacity of Ion Exchange Resins

Reported exchange capacities vary with the type and concentration of regenerant used to restore the resin (see table 11–47.) Typical synthetic resin exchange capacities are in the range of 2 to 10 eq/kg of resin. Zeolite cation exchangers used for water softening have exchange capacities of 0.05 to 0.1 eq/kg. Exchange capacity is measured by placing the resin in a known form. A cationic resin would be washed with a strong-acid to place all of the exchange sites on the resin in the H^+ form or washed with a strong NaCl brine to place all of the exchange sites in the Na^+ form. A solution of known concentration of an exchangeable ion (e.g., Ca^{2+}) can then be added until exchange is complete and the amount of exchange capacity can be measured, or in the acid case, the resin is titrated with a strong- base. Determination of the capacity of an ion exchange resin based on titration is illustrated in Example 11–17.

Exchange capacities for resins often are expressed in terms of grams $CaCO_3$ per cubic meter of resin (g/m^3) or gram equivalents per cubic meter ($g\text{-eq}/m^3$). Conversion between these two units is accomplished using the following expression:

$$\frac{1\text{ g-eq}}{m^3} = \frac{(1\text{ g-eq})\left(\dfrac{100\text{ g CaCO}_3}{2\text{ g-eq}}\right)}{m^3} = 50\text{ g CaCO}_3/m^3 \tag{11–112}$$

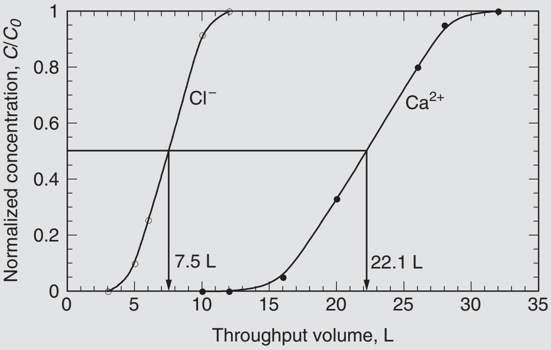
Calculation of the required resin volume for an ion-exchange process is also illustrated in Example 11–17.

EXAMPLE 11–17 **Determination of Ion Exchange Capacity for a New Resin** A column study was conducted to determine the capacity of a cation exchange resin. In conducting the study, 0.1 kg of resin was washed with NaCl until the resin was in the R-Na form. The column was then washed with distilled water to remove the chloride ion (Cl^-) from the intersticies of the resin. The resin was then titrated with a solution of calcium chloride ($CaCl_2$), and the concentrations of chloride and calcium were measured at various throughput volumes. The measured concentrations of Cl^- and Ca^{2+} and the corresponding throughput volumes are as given below. Using the data given below, determine the exchange capacity of the resin and the mass and volume of a resin required to treat 4000 m^3 of water containing 18 mg/liter of ammonium ion (NH_4^+). Assume the density of the resin is 700 kg/m^3.

Throughput volume, L	Constituent, mg/L	
	Cl^-	Ca^{2+}
2	0	0
3	trace	0
5	7	0
6	18	0
10	65	0
12	71	trace
20	71	13
26	71	32
28	71	38
32	$C_o = 71$	$C_o = 40$

Solution

1. Prepare a plot of the normalized concentrations of Cl^- and Ca^{2+} as a function of the throughput volume. The required plot is given below.



2. Determine the exchange capacity.
The exchange capacity (EC) of the resin in meq/kg is:

$$EC = \frac{VC_o}{R}$$

where V = throughput volume between the Cl^- and Ca^{2+} breakthrough curves at $C/C_o = 0.5$
C_o = initial calcium concentration in meq/L
R = amount of resin in kg

$$EC = \frac{(22.1\text{ L} - 7.5\text{ L})\left[\dfrac{(40\text{ mg/L})}{(20\text{ mg/meq})}\right]}{0.1\text{ kg of resin}} = 292\text{ meq/kg of resin}$$

3. Determine the mass and volume of resin required to treat 4000 m^3 of water containing 18 mg/liter of ammonium ion NH_4^+.
   a. Determine the meq of NH_4^+.

   $$NH_4^+\text{, meq/L} = \frac{(18\text{ mg/L as }NH_4^+)}{(18\text{ mg/meq})} = 1\text{ meq/L}$$

   b. The required exchange capacity is equal to

   $$(1.0\text{ meq/L})(4000\text{ m}^3)(10^3\text{ L/m}^3) = 4 \times 10^6\text{ meq}$$

   c. The required mass of resin is

   $$R_{\text{mass}}\text{, kg} = \frac{4 \times 10^6\text{ meq}}{(292\text{ meq/kg of resin})} = 13{,}700\text{ kg}$$

   d. The required volume of resin is

   $$R_{\text{vol}}\text{, m}^3 = \frac{13{,}700\text{ kg of resin}}{(700\text{ kg/m}^3)} = 19.6\text{ m}^3$$

Comment In practice, because of leakage and other operational and design limitations, the required volume of resin will usually be about 1.1 to 1.4 times that computed on the basis of exchange capacity. Also, the above computation is based on the assumption that the entire capacity of the resin is utilized.

Table 11–47

Characteristics of ion exchange resins used in wastewater treatment processes[a]

Resin Type	Acronym	Fundamental reaction[b]	Regenerant ions (X)	pK	Exchange capacity, meq/mL	Constituents removed
Strong-acid cation	SAC	$n[RSO_3^-]X^+ + M^{n+} \rightleftarrows$ $[nRSO_3^-]M^{n+} + nX^+$	H^+ or Na^+	<0	1.7 to 2.1	H^+ form: any cation; Na^+ form: divalent cations
Weak-acid cation	WAC	$n[RCOO^-]X^+ + M^{n+} \rightleftarrows$ $[nRCOO^-]M^{n+} + nX^+$	H^+	4 to 5	4 to 4.5	Divalent cations first, then monovalent cations until alkalinity is consumed
Strong-base anion (type 1)	SBA–1[c]	$n[R(CH_3)_3N^+]X^- + A^{n-} \rightleftarrows$ $[nR(CH_3)_3N^+]A^{n-} + nX^-$	OH^- or Cl^-	>13	1 to 1.4	OH^- form: any anion; Cl^- form: sulfate, nitrate, perchlorate, etc.
Stong-base anion (type 2)	SBA–2[d]	$n[R(CH_3)_2(CH_3CH_2OH)N^+]X^- + A^{n-} \rightleftarrows$ $[nR(CH_3)_2(CH_3CH_2OH)N^+]A^{n-} + nX^-$	OH^- or Cl^-	>13	2 to 2.5	OH^- form: any anion; Cl^- form: sulfate, nitrate, perchlorate, etc.
Weak-base anion	WBA	$[R(CH_3)_2N]HX + HA \rightleftarrows$ $[R(CH_3)_2N]HA + HX$	OH^-	5.7 to 7.3	2 to 3	Divalent anions first, then monovalent anions until strong-acid is consumed

[a] From Crittenden et al. (2005).

[b] Term within the brackets represents the solid phase of the resin.

[c] Greater chemical stability than SBA–1.

[d] Greater regeneration efficiency and capacity than SBA–2.

Ion Exchange Chemistry

The chemistry of the ion exchange process may be represented by the following equilibrium expression for the reaction of constituent A on a cation exchange resin and constituent B in solution.

$$nR^-A^+ + B^{n+} \rightleftarrows R_n^-B^{n+} + nA^+ \qquad (11\text{–}113)$$

where R^- is the anionic group attached to an ion exchange resin and A and B are cations in solution. The generalized form of the equilibrium expression for the above reaction is

$$\frac{[A^+]_S^n[R_n^-B^{n+}]_R}{[R^-A^+]_R^n[B^{n+}]_S} = K_{A^+ \rightarrow B^{n+}} \qquad (11\text{–}114)$$

where $K_{A^+ \rightarrow B^{n+}}$ = selectivity coefficient
$[A^+]_S$ = concentration of A in solution
$[R^-A^+]_R$ = concentration A on the exchange resin

The reactions for the removal of sodium (Na^+) and calcium (Ca^{2+}) ions from water using a strong-acid synthetic cationic exchange resin R, and the regeneration of the exhausted resins with hydrochloric acid (HCl) and sodium chloride (NaCl) are as follows:

Reaction:

$$R^-H^+ + Na^+ \rightleftarrows R^-Na^+ + H^+ \tag{11–115}$$

$$2R^-Na^+ + Ca^{2+} \rightleftarrows R_2^-Ca^{2+} + 2Na^+ \tag{11–116}$$

Regeneration:

$$R^-Na^+ + HCl \rightleftarrows R^-H^+ + NaCl \tag{11–117}$$

$$R_2^-Ca^{2+} + 2NaCl \rightleftarrows 2R^-Na^+ + CaCl_2 \tag{11–118}$$

The corresponding equilibrium expressions for sodium and calcium are as follows:

For sodium:

$$\frac{[H^+][R^-Na^+]}{[R^-H^+][Na^+]} = K_{H^+ \rightarrow Na^+} \tag{11–119}$$

For calcium:

$$\frac{[Na^+]^2[R^-Ca^{2+}]}{[R^-Na^+]^2[Ca^{2+}]} = K_{Na^+ \rightarrow Ca^{2+}} \tag{11–120}$$

The selectivity coefficient depends primarily on the nature and valence of the ion, the type of resin and its saturation, and the ion concentration in wastewater and typically is valid over a narrow pH range. In fact, for a given series of similar ions, exchange resins have been found to exhibit an order of selectivity or affinity for the ions. Approximate selectivity coefficients for cationic and anionic resins are given in Tables 11–48 and 11–49,

Table 11–48

Approximate selectivity coefficients scale for cations on 8 percent cross-linked strong-acid ion exchange resins[a]

Cation	Selectivity coefficient	Cation	Selectivity coefficient
Li^+	1.0	Co^{2+}	3.7
H^+	1.3	Cu^{2+}	3.8
Na^+	2.0	Cd^{2+}	3.9
NH_4^+	2.6	Be^{2+}	4.0
K^+	2.9	Mn^{2+}	4.1
Rb^+	3.2	Ni^{2+}	3.9
Cs^+	3.3	Ca^{2+}	5.2
Ag^+	8.5	Sr^{2+}	6.5
Mg^{2+}	3.3	Pb^{2+}	9.9
Zn^{2+}	3.5	Ba^{2+}	11.5

[a] Adapted from Bonner and Smith (1957), see also Slater (1991).

Table 11–49

Approximate selectivity coefficients for anions on strong-base ion exchange resins[a]

Anion	Selectivity coefficient	Anion	Selectivity coefficient
HPO_4^{2-}	0.01	BrO_3^-	1.0
CO_3^{2-}	0.03	Cl^-	1.0
OH^- (Type I)	0.06	CN^-	1.3
F^-	0.1	NO_2^-	1.3
SO_4^{2-}	0.15	HSO_4^-	1.6
CH_3COO^-	0.2	Br^-	3.0
HCO_3^-	0.4	NO_3^-	3.0–4.0
OH^- (Type II)	0.05–0.65	I^-	18.0

[a] Adapted from Peterson (1953) and Bard (1966).

respectively. The use of the selectivity coefficients given in these tables is illustrated in Example 11–18.

Typical selectivity series for synthetic cationic and anionic exchange resins are as follows.

$$Li^+ < H^+ < Na^+ < NH_4^+ < K^+ < Rb^+ < Ag^+ \tag{11–121}$$

$$Mg^{2+} < Zn^{2+} < Co^{2+} < Cu^{2+} < Ca^{2+} < Sr^{2+} < Ba^{2+} \tag{11–122}$$

$$OH^- < F^- < HCO_3^- < Cl^- < Br^- < NO_3^- < ClO_4^- \tag{11–123}$$

In practice, the selectivity coefficients are determined by measurement in the laboratory and are valid only for the conditions under which they were measured. At low concentrations, the value of the selectivity coefficient for the exchange of monovalent ions by divalent ions is, in general, larger than the exchange of monovalent ions by monovalent ions. This fact has, in many cases, limited the use of synthetic resins for the removal of certain substances in wastewater, such as ammonia in the form of the ammonium ion. There are, however, certain natural zeolites that favor NH_4^+ or Cu^{2+}.

Anderson (1975), in a classic paper, developed a method that can be used to evaluate the effectiveness of a proposed ion exchange process using strong ionic resins. In the development proposed by Anderson, it is assumed that at 100 percent leakage, the effluent concentration of a constituent is equal to the influent concentration (i.e., equilibrium has been reached). The equilibrium condition can be assumed to be either the limiting operating exchange capacity of the resin or the capacity corresponding to the maximum regeneration level that can be attained. Using this assumption, Eq. (11–114) is converted from concentration units to units of equivalent fractions by making the following substitutions:

$$X_{A^+} = \frac{[A^+]_S}{C} \text{ and } X_{B^+} = \frac{[B^+]_S}{C} \tag{11–124}$$

$$X_{A^+} + X_{B^+} = 1 \tag{11–125}$$

where X_A^+ and X_B^+ are the equivalent fractions of A and B in solution and C is the total cationic or anionic concentration in solution.

$$\overline{X}_{A^+} = \frac{[R^-A^+]_R}{\overline{C}} \text{ and } \overline{X}_{B^+} = \frac{[R^-B^+]_R}{\overline{C}} \tag{11–126}$$

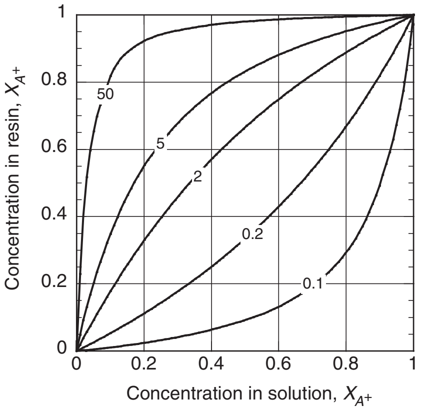


Figure 11–70

Distribution curves for a single monovalent ion A between the solution and the resin for different values of the selectivity coefficient.

$$\bar{X}_{A^+} + \bar{X}_{B^+} = 1 \tag{11–127}$$

Where $\bar{X}_A$ and $\bar{X}_B$ are the equivalent fractions of A and B in the resin and $\bar{C}$ is the total ionic concentration in the resin (i.e., the total resin capacity in eq/L). Substituting the above terms into Eq. (11–114) and simplifying results in the following expression:

$$\frac{\bar{X}_{B^+} X_{A^+}}{\bar{X}_{A^+} X_{B^+}} = K_{A^+ \rightarrow B^+} \tag{11–128}$$

Substituting for X_{A^+} and $\bar{X}_{A^+}$ in Eq. (11–128) results in:

$$\frac{\bar{X}_{B^+}}{1 - \bar{X}_{B^+}} = (K_{A^+ \rightarrow B^+})\left(\frac{X_{B^+}}{1 - X_{B^+}}\right) \tag{11–129}$$

It should be noted that Eq. (11–129) is only valid for exchanges between monovalent ions on fully ionized exchange resins. The distribution of a single monovalent ion A between the solution and the resin for different values of the selectivity coefficient is presented on Fig. 11–70. The distribution curves can be used to assess the effectiveness of a resin for the removal of a given ion, based on the selectivity coefficient.

The following three attributes of Eq. (11–129) were identified by Anderson (1975).

1. The term $\bar{X}_B/(1 - \bar{X}_B)$ corresponds to the state of the resin in an exchange column when the influent and effluent concentrations are the same.
2. The term $\bar{X}_B$ corresponds to the extent to which the resin can be converted to the B^+ form when the resin is in equilibrium with a solution of composition X_B.
3. The term $\bar{X}_B$ also corresponds to the maximum extent of regeneration that can be achieved with a regenerant composition of X_B.

The corresponding equation for exchanges between monovalent and divalent ions on a fully ionized exchange resin is:

$$\frac{\bar{X}_{B^{+2}}}{(1 - \bar{X}_{B^{+2}})^2} = (K_{A^+ \rightarrow B^{+2}})\left(\frac{\bar{C}}{C}\right)\frac{X_{B^{+2}}}{(1 - X_{B^{+2}})^2} \tag{11–130}$$

The application of these equations is illustrated in Example 11–18.

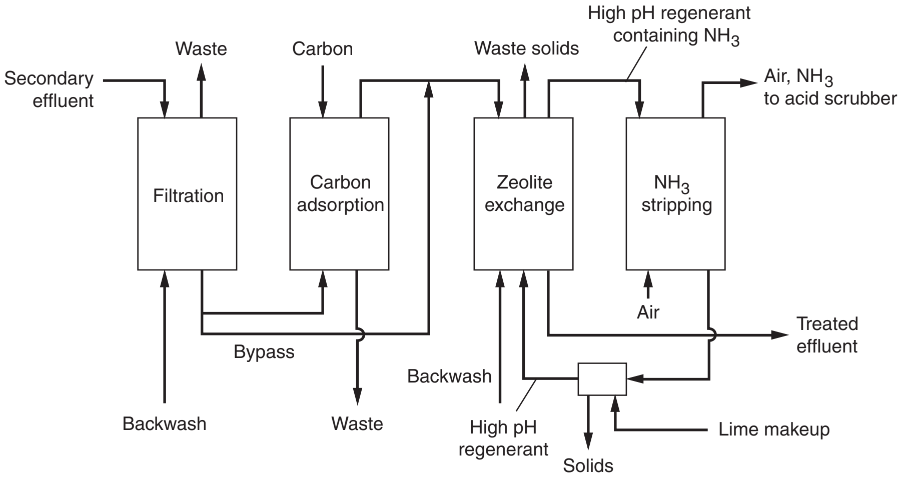


Figure 11–71

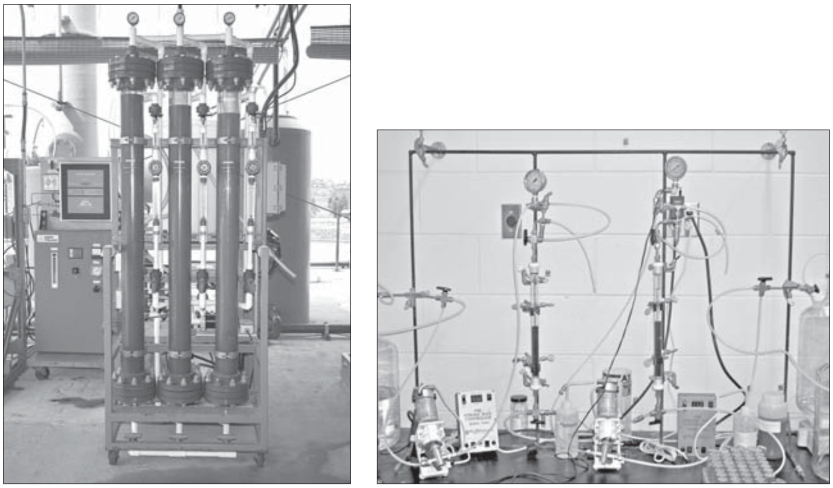
Typical flow diagram for the removal of ammonium by zeolite exchange. Note: the ammonium removed is recovered by high pH air stripping and acid scrubbing.

Application of Ion Exchange

As noted previously, ion exchange has been used in wastewater applications for the removal of nitrogen, heavy metals, and total dissolved solids.

For Nitrogen Control. For nitrogen control, the ions typically removed from the waste stream are ammonium, NH_4^+, and nitrate, NO_3^-. The ion that the ammonium displaces varies with the nature of the solution used to regenerate the bed. Although both natural and synthetic ion exchange resins are available, synthetic resins are used more widely because of their durability. Some natural resins (zeolites) have found application in the removal of ammonium from wastewater. Clinoptilolite, a naturally occurring zeolite, has proven to be one of the best natural exchange resins. In addition to having a greater affinity for ammonium ions than other ion exchange materials, it is relatively inexpensive when compared to synthetic media. One of the novel features of this zeolite is the regeneration system employed. Upon exhaustion, the zeolite is regenerated with lime [$Ca(OH)_2$] and the ammonium ion removed from the zeolite is converted to ammonia because of the high pH. A flow diagram for this process is shown on Fig. 11–71. The stripped liquid is collected in a storage tank for subsequent reuse. A problem that must be solved is the formation of calcium carbonate precipitates within the zeolite exchange bed and in the stripping tower and piping appurtenances. As indicated on Fig. 11–71, the zeolite bed is equipped with backwash facilities to remove the carbonate deposits that form within the filter.

When using conventional synthetic ion exchange resins for the removal of nitrate, two problems are encountered. First, while most resins have a greater affinity for nitrate over chloride or bicarbonate, they have a significantly lower affinity for nitrate as compared to sulfate, which limits the useful capacity of the resin for the removal of nitrate. The impact of the presence of sulfate on the nitrate removal capacity of conventional resins is illustrated in Example 11–18. Second, because of the lower affinity for nitrate over sulfate, a phenomenon known as *nitrate dumping* can occur. Nitrate dumping occurs when an ion exchange column is operated past the nitrate breakthrough, at which point the sulfate in the feed water will displace the nitrate on the resin causing a release of nitrate. To overcome the problems associated with low affinity and nitrate breakthrough, new types of resins have been developed within which the affinities for nitrate and sulfate have been exchanged. When significant amounts of sulfate are present (i.e., typically greater than 25 percent of the total of the sum of the sulfate and nitrate expressed in meq/L), the use of nitrate selective resins is advantageous. Because the performance of nitrate selective resins



(a) (b)

Figure 11–72

Typical ion exchange test columns (a) ion exchange columns used to study the removal of nitrate from water which has been processed with reverse osmosis and (b) bench scale ion exchange columns. (Courtesy of David Hand.)

will vary with the composition of the treated wastewater, pilot testing will usually be required (McGarvey et al., 1989; Dimotsis and McGarvey, 1995). Typical ion exchange test columns used to study the removal of nitrate from water which has been processed with reverse osmosis are shown on Fig. 11–72.

EXAMPLE 11–18 **Ion Exchange for the Removal of Nitrate without and with Sulfate Present in the Water** Nitrate is to be removed from two different treated wastewaters with the compositions given below. For the purpose of illustration, assume a conventional ion exchange resin will be used.

Wastewater A

Cation	Conc., mg/L	mg/meq	meq/L	Anion	Conc., mg/L	mg/meq	meq/L
Ca^{2+}	82.2	20.04	4.10	HCO_3^-	305.1	61.02	5.00
Mg^{2+}	17.9	12.15	1.47	SO_4^{2-}	0.00	48.03	0.00
Na^+	46.4	23.00	2.02	Cl^-	78.0	35.45	2.20
K^+	15.5	39.10	0.40	NO_3^-	50.0	62.01	0.81
		Σcations	7.99			Σanions	8.01

Wastewater B

Cation	Conc., mg/L	mg/meq	meq/L	Anion	Conc., mg/L	mg/meq	meq/L
Ca^{2+}	82.2	20.04	4.10	HCO_3^-	220.0	61.02	3.61
Mg^{2+}	17.9	12.15	1.47	SO_4^{2-}	79.2	48.03	1.65
Na^+	46.4	23.00	2.02	Cl^-	78.0	35.45	2.20
K^+	15.5	39.10	0.40	NO_3^-	50.0	62.01	0.81
		Σcations	7.99			Σanions	8.27

Determine the maximum amount of water that can be processed per liter of a strong-base anion exchange resin with an exchange capacity of 2.0 eq/L.

Solution: Wastewater A

1. Estimate the selectivity coefficient (see Table 11–49). To apply Eq. (11–129) the system must be reduced to two components. For this purpose, HCO_3^- and Cl^- are combined into a single component. Using a selectivity value of 4 for nitrate, the selectivity coefficient is estimated as follows:

$$K_{HCO_3^- \rightarrow NO_3^-} = \frac{4.0}{0.4} = 10.0$$

$$K_{Cl^- \rightarrow NO_3^-} = \frac{4.0}{1.0} = 4.0$$

$$K_{[(HCO_3^-)(Cl^-)] \rightarrow NO_3^-} = 7.0 \text{ (estimated)}$$

2. For the equilibrium condition ($C_e/C_o = 1.0$), estimate the nitrate equivalent fraction in solution.

$$X_{NO_3^-} = \frac{0.81}{8.01} = 0.101$$

3. Compute the equilibrium resin composition using Eq. (11–129).

$$\frac{\overline{X}_{B^+}}{1 - \overline{X}_{B^+}} = (K_{A^+ \rightarrow B^+})\left(\frac{X_{B^+}}{1 - X_{B^+}}\right)$$

$$\frac{\overline{X}_{NO_3^-}}{1 - \overline{X}_{NO_3^-}} = 7.0\left(\frac{0.101}{1 - 0.101}\right)$$

$$\overline{X}_{NO_3^-} = 0.44$$

Thus, 44 percent of the exchange sites on the resin can be used for the removal of nitrate.

4. Determine the limiting operating capacity of the resin for the removal of nitrate.

Limiting operating capacity = (2 eq/L of resin)(0.44) = 0.88 eq/L of resin

5. Determine the volume of water that can be treated during a service cycle.

$$\text{Vol} = \frac{\text{(nitrate removal capacity of resin, eq/L of resin)}}{\text{(nitrate in solution, eq/L of water)}}$$

$$= \frac{(0.88 \text{ eq/L of resin})}{(0.81 \times 10^{-3} \text{ eq/L of water})} = 1086 \frac{\text{L of water}}{\text{L of resin}}$$

Solution: Wastewater B

1. Estimate the selectivity coefficient (see Table 11–49). To apply Eq. (11–130) the system must be reduced to two components. For this purpose, HCO_3^-, Cl^-, and NO_3^- are combined into a single monovalent component. The selectivity coefficient is estimated as follows:

$$K_{HCO_3^- \rightarrow SO_4^{2-}} = \frac{0.15}{0.4} = 0.4$$

$$K_{Cl^- \rightarrow SO_4^{2-}} = \frac{0.15}{1.0} = 0.15$$

$$K_{\mathrm{NO_3^-} \rightarrow \mathrm{SO_4^{2-}}} = \frac{0.15}{4.0} = 0.04$$

$$K_{[(\mathrm{NO_3^-})(\mathrm{HCO_3^-})(\mathrm{Cl^-})] \rightarrow \mathrm{SO_4^{2-}}} = 0.2 \text{ (estimated)}$$

2. For the equilibrium condition ($C_e/C_o = 1.0$), estimate the sulfate equivalent fraction in solution.

$$X_{\mathrm{SO_4^{2-}}} = \frac{1.65}{8.27} = 0.2$$

3. Compute the equilibrium resin composition using Eq. (11–130).

$$\frac{\overline{X}_{B^{2-}}}{(1 - \overline{X}_{B^{2-}})^2} = (K_{A^- \rightarrow B^{2-}})\left(\frac{C}{C}\right)\left[\frac{X_{B^{2-}}}{(1 - X_{B^{2-}})^2}\right]$$

$$\frac{\overline{X}_{\mathrm{SO_4^{2-}}}}{(1 - \overline{X}_{\mathrm{SO_4^{2-}}})^2} = 0.2\frac{2}{0.00827}\left[\frac{0.2}{(1 - 0.2)^2}\right]$$

$\overline{X}_{\mathrm{SO_4^{2-}}} = 0.77$, determined by successive trials

Thus, 77 percent of the exchange sites on the resin will be in the divalent form at equilibrium. The relative amount of $\mathrm{NO_3^-}$ can be estimated by assuming that the remaining 23 percent of the resin sites are in equilibrium with a solution of $\mathrm{NO_3^-}$, $\mathrm{HCO_3^-}$, and $\mathrm{Cl^-}$ with the same relative concentration as the feed.

The equivalent fraction of nitrate in the solution will then be

$$X_{\mathrm{NO_3^-}} = \frac{0.81}{6.62} = 0.12$$

The selectivity coefficient for the monovalent system is estimated:

$$K_{\mathrm{HCO_3^-} \rightarrow \mathrm{NO_3^-}} = \frac{4.0}{0.4} = 10.0$$

$$K_{\mathrm{Cl^-} \rightarrow \mathrm{NO_3^-}} = \frac{4.0}{1.0} = 4.0$$

$$K_{[(\mathrm{HCO_3^-})(\mathrm{Cl^-})] \rightarrow \mathrm{NO_3^-}} = 7.0 \text{ (estimated)}$$

Compute the equilibrium resin composition using Eq. (11–129).

$$\frac{\overline{X}'_{B^+}}{1 - \overline{X}'_{B^+}} = (K_{A^+ \rightarrow B^+})\left(\frac{\overline{X}'_{B^+}}{1 - \overline{X}'_{B^+}}\right)$$

$$\frac{\overline{X}'_{\mathrm{NO_3^-}}}{1 - \overline{X}'_{\mathrm{NO_3^-}}} = 7.0\left(\frac{0.12}{1 - 0.12}\right)$$

$$\overline{X}'_{\mathrm{NO_3^-}} = 0.5$$

The fraction of the total resin capacity in the nitrate form is then computed.

$$\overline{X}_{\mathrm{NO_3^-}} = (1 - \overline{X}_{\mathrm{SO_4^{2-}}})(\overline{X}'_{\mathrm{NO_3^-}}) = (0.23)(0.5) = 0.115$$

4. Determine the limiting operating capacity of the resin for the removal of nitrate.

Limiting operating capacity = (2 eq/L of resin)(0.115) = 0.23 eq/L of resin.

5. Determine the volume of water that can be treated during a service cycle.

$$\text{Vol} = \frac{(\text{nitrate removal capacity of resin, eq/L of resin})}{(\text{nitrate in solution, eq/L of water})}$$

$$= \frac{(0.23\ \text{eq/L of resin})}{(0.81 \times 10^{-3}\ \text{eq/L of water})} = 284\ \frac{\text{L of water}}{\text{L of resin}}$$

Comment As illustrated in this problem, the ionic composition of the wastewater can have a significant effect on the amount of water that can be treated per unit volume of resin, especially where nitrate is to be removed. Because the sulfate is more than 25 percent of the sum of the sulfate and nitrate, the use of a nitrate selective resin would be advantageous in this application. The approximate nature of these calculations also demonstrates the importance of conducting pilot plant tests to establish actual throughput volumes.

Removal of Heavy Metals. Metal removal may be required as a pretreatment before discharge to a municipal sewer system. Because of the potential accumulation and toxicity of these metals, it is desirable to remove them from wastewater effluents before release to the environment. Ion exchange is one of the most common forms of treatment used for the removal of metals. Facilities and activities that may discharge wastewater containing high concentrations of metals include metal processing, electronics industries (semiconductors, printed circuit boards), metal plating and finishing, pharmaceuticals and laboratories, and vehicle service shops. High metal concentrations can also be found in leachate from landfills, and stormwater runoff.

Where industries produce effluents with widely fluctuating metal concentrations, flow equalization may be required to make ion exchange feasible. The economic feasibility of using ion exchange processes for metal removal greatly improves when the process is used for the removal and recovery of valuable metals. Because it is now possible to manufacture resins for specific applications, the use of resins that have a high selectivity for the desired metal(s) also improves the economics of ion exchange.

Materials used for the exchange of metals include zeolites, weak and strong anion and cation resins, chelating resins, and microbial and plant biomass. Biomass materials are generally more abundant, and therefore, less expensive when compared to other commercially available resins. Natural zeolites, clinoptilolite (selective for Cs), and chabazite (mixed metals background Cr, Ni, Cu, Zn, Cd, Pb) have been used to treat wastewater with mixed metal backgrounds (Ouki and Kavannagh, 1999). Chelating resins, such as aminophosphonic and iminodiacetic resins, have been manufactured to have a high selectivity for specific metals such as Cu, Ni, Cd, and Zn.

Ion exchange processes are highly pH dependent. Solution pH has a significant impact on the metal species present and the interaction between exchanging ions and the resin. Most metals bind better at higher pH due to less competition from protons for sites. Operating and wastewater conditions determine selectivity of the resin, pH, temperature, other ionic species, and chemical background. The presence of oxidants, particles, solvents, and polymers may affect the performance of ion exchange resins. The quantity and quality of regenerate produced and subsequently requiring management must also be considered.

Removal of Total Dissolved Solids. For the reduction of the total dissolved solids, both anionic and cationic exchange resins must be used (see Fig. 11–73). The wastewater is

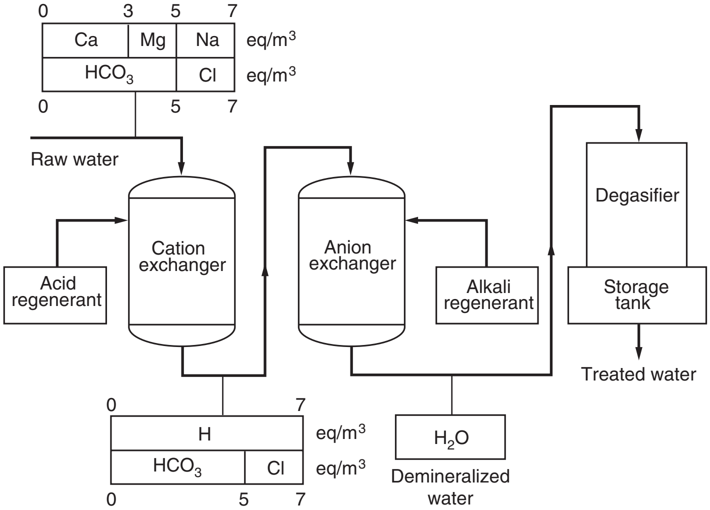


Figure 11–73

Typical flow diagram for the removal of hardness and for the complete demineralization of water using ion exchange resins.

first passed through a cation exchanger where the positively charged ions are replaced by hydrogen ions. The cation exchanger effluent is then passed over an anionic exchange resin where the anions are replaced by hydroxide ions. Thus, the dissolved solids are replaced by hydrogen and hydroxide ions that react to form water molecules.

Total dissolved solids removal can take place in separate exchange columns arranged in series, or both resins can be mixed in a single reactor. Wastewater application rates range from 0.20 to 0.40 $m^3/m^2 \cdot min$ (5 to 10 $gal/ft^2 \cdot min$). Typical bed depths are 0.75 to 2.0 m (2.5 to 6.5 ft). In reuse applications, treatment of a portion of the wastewater by ion exchange, followed by blending with wastewater not treated by ion exchange, would possibly reduce the dissolved solids to acceptable levels. In some situations, it appears that ion exchange may be as competitive as reverse osmosis.

Operational Considerations

To make ion exchange economical for advanced wastewater treatment, it would be desirable to use regenerants and restorants that would remove both the inorganic anions and the organic material from the spent resin. Chemical and physical restorants found to be successful in the removal of organic material from resins include sodium hydroxide, hydrochloric acid, methanol, and bentonite. To date, ion exchange has had limited application because of the extensive pretreatment required, concerns about the life of the ion exchange resins, and the complex regeneration system required.

High concentrations of influent TSS can plug the ion exchange beds, causing high headlosses and inefficient operation. Resin binding can be caused by residual organics found in biological treatment effluents. Some form of chemical treatment and clarification is required before ion exchange demineralization. This problem has been solved partially by prefiltering the wastewater or by using scavenger exchange resins before application to the exchange column.

11–12 DISTILLATION

Distillation is a unit process in which the components of a liquid solution are separated by vaporization and condensation. Along with reverse osmosis, distillation can be used to control the buildup of salts in critical reuse applications. Because distillation is expensive, its application is generally limited to applications where (1) a high degree of treatment is required, (2) contaminants cannot be removed by other methods, and (3) inexpensive heat is available. The purpose of this section is to introduce the basic concepts involved in distillation. As the use of distillation for wastewater reclamation is a recent development, the current literature must be consulted for the results of ongoing studies and more recent applications.

Distillation Processes

Over the past 20 years, a variety of distillation processes, employing a variety of evaporator types and methods of using and transferring heat energy, have been evaluated or used. The principal distillation processes are (1) boiling with submerged tube heating surface, (2) boiling with long-tube vertical evaporator, (3) flash evaporation, (4) forced circulation with vapor compression, (5) solar evaporation, (6) rotating-surface evaporation, (7) wiped-surface evaporation, (8) vapor reheating process, (9) direct heat transfer using an immiscible liquid, and (10) condensing-vapor-heat transfer by vapor other than steam. Of these types of distillation processes, multiple-effect evaporation, multistage flash evaporation, and vapor-compression distillation appear most feasible for the reclamation of municipal wastewater.

Multiple-Effect Evaporation Distillation. In multiple-effect evaporation distillation systems, several evaporators (boilers) are arranged in series, each operating at a lower pressure than the preceding one. In a three-stage, vertical-tube evaporator (see Fig. 11–74), preheated influent water to be demineralized is introduced into the first evaporation stage where it is evaporated with steam contained within heat exchange tubes. The vapor from the first stage enters the second stage where it is condensed within the evaporation tubes. Water from the first stage is the feed water for the second stage. The process is repeated in the next n^{th} stages. Heated vapor from the last stage is used to heat the influent feed water. In an alternative arrangement, the feed water for the second and subsequent stages is the preheated influent. Water which does not evaporate is taken off as brine at each stage. If air entrainment is kept low, almost all of the nonvolatile contaminants can be removed in a single evaporation step. Volatile contaminants, such as ammonia gas and low-molecular weight organic acids, may be removed in a preliminary evaporation step, but if their concentration is so small that their presence in the final product is not objectionable, this step with its added cost can be eliminated.

Multistage Flash Evaporation Distillation. Multistage flash evaporation distillation systems have been used commercially in desalination for many years. In the multistage flash process (see Fig. 11–75), the influent wastewater is first treated to remove TSS and deaerated before being pumped through heat transfer units in the several stages of the

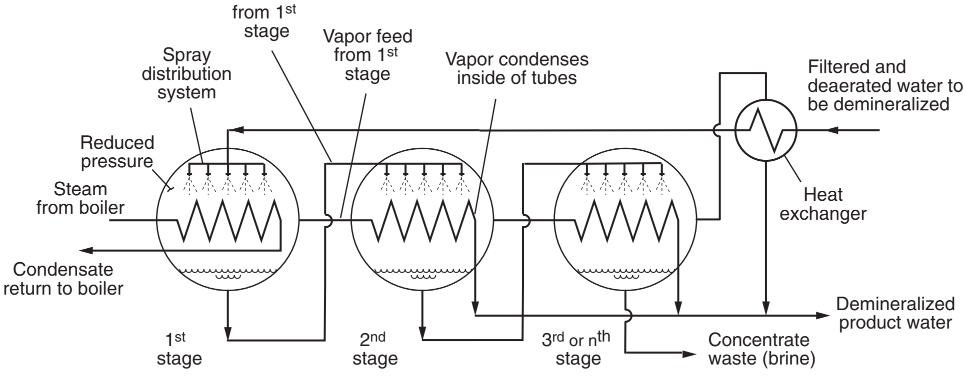
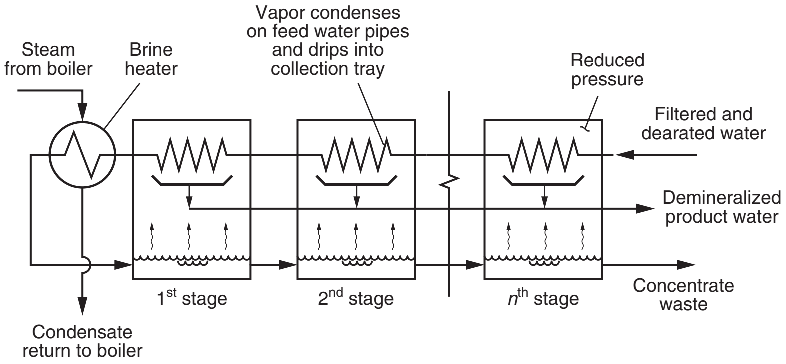


Figure 11–74

Schematic of multiple-effect evaporation distillation process.

Figure 11–75

Schematic of multistage flash evaporation distillation process.



distillation system, each of which is maintained at a lower pressure. Vapor generation or boiling caused by reduction in pressure is known as "*flashing*." As the water enters each stage through a pressure reducing nozzle, a portion of the water is flashed to form a vapor. In turn, the flashed water vapor condenses on the outside of the condenser tubes and is collected in trays (see Fig. 11–75). As the vapor condenses, its latent heat is used to preheat the wastewater that is being returned to the main heater where it will receive additional heat before being introduced to the first flashing stage. When the concentrated wastewater reaches the lowest pressure stage, it is pumped out. Thermodynamically, multistage flash evaporation is less efficient than ordinary evaporation. However, by combining a number of stages in a single reactor, external piping is eliminated and construction costs are reduced.

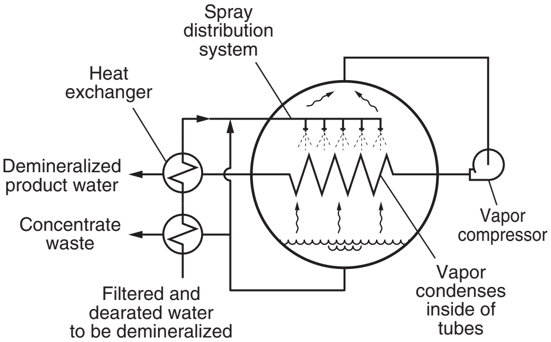
Vapor Compression Distillation. In the vapor compression process an increase in pressure of the vapor is used to establish the temperature difference for the transfer of heat. The basic schematic of a vapor compression distillation unit is shown on Fig. 11–76. After initial heating of the wastewater, the vapor pump is operated so that the vapor under higher pressure can condense in the condenser tubes, at the same time causing the release of an equivalent amount of vapor from the concentrated solution. Heat exchangers can conserve heat from both the condensate and the waste brine. The only energy input required during operation is the mechanical energy for the vapor pump. Hot concentrated wastewater must be discharged at intervals to prevent the buildup of excessive concentrations of salt in the boiler.

Performance Expectations in Reclamation Applications

The principal issues with the application of the distillation processes for wastewater reclamation are the carryover of volatile constituents found in treated wastewater and the degree of subsequent cooling and treatment that may be required to renovate the distilled

Figure 11–76

Schematic of vapor compression distillation process.



water. Typical water quality performance data for a multiple-effect distillation process have been reported for a pilot scale unit by Rose et al. (1999).

The theoretical thermodynamic minimum energy required to raise the temperature of wastewater and to provide the latent heat of vaporization is about 2260 kJ/kg. Typically, about 1.25 to 1.35 times the latent heat of vaporization will be required. Unfortunately, because of the many irreversibilities in an actual distillation processes, the thermodynamic minimum energy requirements are of little practical relevance in the practical evaluation of distillation processes. Typical energy requirements for the three distillation processes discussed above are (Voutchkov, 2013):

Multiple-effect evaporation: 5.7–7.8 kWh/m^3 (23–30 kWh/10^3 gal)

Multistage flash evaporation: 12.7–15.0 kWh/m^3 (48–57 kWh/10^3 gal)

Vapor compression: 8–12 kWh/m^3 (30–45 kWh/10^3 gal)

These values for distillation can be compared to the values given in Table 11–30 for reverse osmosis: 9–12 kWh/m^3 (34–45 kWh/10^3 gal) without energy recovery and $<$3–4 kWh/m^3 ($<$11–32 kWh/10^3 gal) with energy recovery. Based on this energy comparison it is clear why reverse osmosis used in seawater distillation has largely replaced distillation for seawater desalination.

Operating Problems

The most common operating problems encountered include scaling and corrosion. Due to temperature increases, inorganic salts come out of solution and precipitate on the inside walls of pipes and equipment. The control of scaling due to calcium carbonate, calcium sulfate, and magnesium hydroxide is one of the most important design and operational considerations in distillation desalination processes. Controlling the pH minimizes carbonate and hydroxide scales. Most inorganic solutions are corrosive. Cupronickel alloys are used most commonly in sea water desalination. Other metals that are used include aluminum, bronze, titanium, and monel.

Disposal of Concentrated Waste

All distillation processes reject part of the influent wastewater. Hence, all of these processes have concentrated wastewater disposal problems. The permissible maximum concentration in the wastewater depends on the solubility, corrosion, and vapor pressure characteristics of the wastewater. Therefore, the waste concentration is an important consideration in process optimization. Disposal of concentrated wastewater brines is essentially the same problem encountered with the membrane processes discussed in Sec. 11–5.

PROBLEMS AND DISCUSSION TOPICS

11–1 The following sieve analysis results were obtained for four different stock sands.

US sieve size designation[a]	Size of opening, mm	Cumulative weight passing, %. Sand sample 1	2	3	4
140	0.105	0.4	1.5	0.1	5.0
100	0.149	1.5	4.1	0.8	11.1
70	0.210	4.0	10.0	2.5	20.0

(*continued*)

(*Continued*)

US sieve size designation[a]	Size of opening, mm	Cumulative weight passing, %. Sand sample 1	2	3	4
50	0.297	9.5	21.0	8.2	32.0
40	0.420	18.5	40.6	18.5	49.5
30	0.590	31.0	61.0	32.0	62.3
20	0.840	49.0	78.3	58.1	78.3
16	1.190	63.2	90.0	76.3	88.5
12	1.680	82.8	96.0	90.0	94.4
8	2.380	89.0	99.0	96.7	97.8
6	3.360	98.0	99.9	99.0	99.0
4	4.760	100.0	100.0	100.0	100.0

[a] Note: sieve size number 18 has an opening size of 1.0 mm.

a. For sand sample 1, 2, 3, or 4 (to be selected by instructor) determine the geometric mean size, the geometric standard deviation, the effective size, and the uniformity coefficient for the stock sand.

b. It is desired to produce from the stock sand a filter sand with an effective size of 0.45 mm and a uniformity coefficient of 1.6. Estimate the amount of stock sand needed to obtain one ton of filter sand.

c. What U.S. Standard sieve size should be used to eliminate the excess coarse material?

d. If the sand remaining after sieving in Part c above is placed in a filter, what backwash rise rate would be needed to eliminate the excess fine material?

e. What depth of sieved material would have to be placed in the filter to produce 600 mm of usable filter sand?

f. On log-probability paper, plot the size distribution of the modified sand. Check against the required distribution and sizes.

g. Determine the headloss through 600 mm of the filter sand specified in part (b) for a filtration rate of 160 $L/m^2 \cdot min$ (4 $gal/ft^2 \cdot min$). Assume the sand is stratified and that the maximum and minimum sand sizes are 1.68 (sieve size 12) and 0.297 mm (sieve size 50), respectively. Assume also that $T = 20°C$, $\alpha = 0.4$ for all layers, and $\psi = 0.75$.

An excellent discussion of the procedures involved in developing a usable filter sand from a stock filter sand may be found in Fair et al. (1968).

11–2 Determine the sphericity and specific surface area of two of the following filter mediums (filter medium to be selected by instructor).

Item	Filter medium 1	2	3	4	5
Shape	Ellipsoid	Icosahedron	Cube	Rod	Isosceles tetrahedron
Dimensions	1 mm × 1.2 mm × 2 mm	Each face, 0.5 mm × 0.5 mm × 0.5 mm	1 mm × 1 mm × 1 mm	0.5 mm diameter × 2 mm in length	1.2 mm × 1.2 mm × 1.2 mm × 2.5 mm

11–3 Using the equations developed by Kozeny, Rose, and Ergun, given in Table 11–8, compare the headloss through a 600 mm sand bed. Assume that the sand bed is composed of spherical unisized sand with a diameter of 0.45, 0.55, or 0.6 mm (to be selected by instructor), the porosity of the sand is 0.40, the filtration rate is 240 L/m ^{2}•min (5 gal/ft^2•min) and the temperature is 15°C. The kinematic viscosity at 15°C is equal to 1.139×10^{-6} m^2/s (see Table C–1, Appendix C).

11–4 Using the equation developed by Rose, determine the headloss through a 750-mm sand bed for a filtration rate of 300 L/m^2•min. Assume that the sand bed is composed of spherical unisized sand with a diameter of 0.40, 0.45, 0.55, or 0.6 mm (size to be selected by instructor), the porosity of the sand is 0.40, and the temperature of the water is 10°C. The kinematic viscosity at 10°C is equal to 1.306×10^{-6} m^2/s (see Table C-1, Appendix C).

11–5 Solve Problem 11–4 assuming the bed is stratified. Assume that the given sand sizes correspond to the effective size (d_{10}) and that the uniformity coefficient for all of the sand sizes is equal to 1.5.

11–6 If a 0.3-m layer of uniform anthracite is placed on top of the sand bed in Prob. 11–3, determine the ratio of the headloss through the anthracite to that of the sand. Assume that the grain-size diameter of the anthracite is 2.0 mm and porosity is 0.50. Will intermixing occur?

11–7 Given the particle size distribution 1, 2, 3, or 4 (to be selected by instructor), determine the effective size (d_{10}) and uniformity coefficient UC, and clean-water headloss through a stratified bed 600 mm deep. If a layer of anthracite is to be added over 600 mm of sand determine the effective size required to minimize intermixing. Assume the filtration rate is 160 L/m^2•min, $\phi = 0.85$, and $\alpha = 0.4$.

	Percent of sand retained			
	Particle size distribution			
Sieve number	1	2	3	4
6–8	2	0	1	0.1
8–10	8	0.1	2	0.7
10–14	10	0.5	4	1.2
14–20	30	7.4	13	10
20–30	26	32	20	24
30–40	14	30	20	29
40–60	8	25	23	25
Pan	2	5	17	10

11–8 Four stratified sand beds with the size distribution given below are to be backwashed at a rate of 0.75 m^3/m^2•min. Determine the degree of expansion and whether the proposed backwash rate will expand all of the filter bed (to be selected by instructor). Assume the following data are applicable: sand specific gravity is 2.65, depth of the filer bed is 0.90 m, and the temperature is 20°C.

	Percent of sand retained			
	Stratified bed number			
Sieve size or number	1	2	3	4
8–10	10	2	0.1	0
10–12	10	4	0.5	0
12–18	30	14	4.4	0

(continued)

(*Continued*)

Sieve size or number	Percent of sand retained			
	Stratified bed number			
	1	2	3	4
18–20	10	8	7	1
20–30	34	40	48	28
30–40	5	22	30	41
40–50	1	9	9	27
Pan		1	1	3

11–9 Given the following granular medium filter effluent turbidity data collected at four different wastewater treatment plants, estimate the mean, the geometric standard deviation, s_g, and the probability of exceeding a turbidity reading of 2.5 NTU (treatment plant to be selected by instructor).

Turbidity, NTU			
Treatment plant			
1	2	3	4
1.7	1.7	1.0	1.2
1.8	1.1	1.8	1.4
2.2	0.9	1.5	1.5
2.0	1.4	1.1	1.6
	1.3	1.7	1.7
		1.3	1.9
			2.0
			2.1

11–10 Assuming the data for treatment plants 1 and 2 in Problem 11–9 were collected at the same treatment plant at different times, what is the impact of using all of the data given for 1 and 2 as one data set versus using the individual data sets? In general, what are the advantages or disadvantages of collecting more turbidity samples?

11–11 Gravity filters are to be used to treat 16,000, 20,000, or 24,000 m^3/d, to be selected by instructor, of settled effluent at a filtration rate of 200 L/m^2·min (5 gal/ft^2·min). The filtration rate with one filter taken out of service for backwashing is not to exceed 240 L/m^2·min (6 gal/ft^2·min). Determine the number of units and the area of each unit to satisfy these conditions. If each filter is backwashed for 30 min every 24 h at a wash rate of 960 L/m^2·min (24 gal/ft^2·min), determine the percentage of filter output used for washing if the filter is out of operation for a total of 30 min/d. What would be the total percentage of filter output used for backwashing if a surface washing system that requires 40 L/m^2·min (1 gal/ft^2·min) of filtered effluent is to be installed?

11–12 Using the performance data given in the following table for a microfiltration membrane, determine (water to be selected by instructor) the rejection and log rejection for each microorganism group.

	Microorganisms concentration, org/mL			
	Water 1		Water 2	
Microorganism	Feed water	Permeate	Feed water	Permeate
HPC	6.5×10^7	3.3×10^2	8.6×10^7	1.5×10^2
Total coliform	3.4×10^6	100	5×10^5	60
Enteric virus	7×10^3	6.6×10^3	2.0×10^3	9.1×10^2

11–13 A hollow-fiber membrane system with inside to outside flow is operated in a cross-flow arrangement. Each module contains 6000 fibers that have an inside diameter of 1.0 mm and a length of 1.25 m. Using this information determine

a. the feed water flowrate at the entrance to the module needed to achieve a cross-flow velocity of 1 m/s within the membrane fibers.
b. the permeate flowrate if the permeate flux of 100 L/m²·h is maintained.
c. the retentate cross-flow velocity at the exit from the membrane fibers.
d. the ratio of velocity of flow through the membrane surface to the average cross-flow velocity within an individual membrane fiber.
e. the ratio of permeate flowrate to feed water flowrate.

This problem was adapted from Crittenden et al., 2012.

11–14 Membrane filtration, operated in a dead-end mode, is used to treat secondary effluent. If the heterotrophic microorganism plate count (HPC) in the effluent increased from 5 org/L under normal operation to 200 org/L after an extended period of operation, estimate the number of broken fibers for the following conditions. The influent flowrate and organism count are 4000 m³/d and 6.7×10^7 org/L, respectively. The membrane bundle contains 5000 individual fibers. If the influent and effluent turbidity values under normal operation are 4 and 0.25 NTU respectively, estimate the increase in the effluent turbidity assuming the increase could be measured.

11–15 Contrast the advantages and disadvantages between depth filtration, surface filtration, and microfiltration. Cite a minimum of three recent articles (after 2000).

11–16 Four different waters are to be desalinized by reverse osmosis using a thin film composite membrane. For water 1, 2, 3, or 4 (water to be selected by instructor), determine the required membrane area, the rejection rate, and the concentration of the retentate.

		Water			
Item	Unit	1	2	3	4
Flowrate	m³/d	4000	5500	20,000	10,000
Influent TDS	g/m³	2850	3200	2000	2700
Permeate TDS	g/m³	200	500	400	225
Flux rate coefficient k_w	m/s·bar[a]	1.0×10^{-6}[a]	1.0×10^{-6}	1.0×10^{-6}	1.0×10^{-6}
Mass transfer rate coefficient, k_i	m/s	6.0×10^{-8}	6.0×10^{-8}	6.0×10^{-8}	6.0×10^{-8}
Net operating pressure	*kPa*	2750	2500	2800	3000
Recovery	%	88.0	90.0	89	86

[a] 1.0×10^{-6} m/s·bar = 1.0×10^{-8} s/m.

11–17 Using the data given below, determine the recovery and rejection rates for one of the following reverse osmosis units (unit to be selected by instructor).

		Reverse osmosis unit			
Item	Unit	1	2	3	4
Feed water flowrate	m³/d	4000	6000	8000	10,000
Retentate flowrate	m³/d	350	600	7500	9000
Permeate TDS	g/m³	65	88	125	175
Retentate TDS	g/m³	1500	2500	1850	2850

11–18 Using the data given below, determine the flux rate coefficient and the mass transfer rate coefficient.

		Reverse osmosis unit			
Item	Unit	1	2	3	1
Flowrate, Q_f	m^3/d	4000	5500	20,000	10,000
Influent TDS, C_f	g/m^3	2500	3300	5300	2700
Permeate TDS, C_p	g/m^3	20	50	40	23
Net operating pressure, ΔP	bar	28	25	28	30
Membrane area	m^2	1600	1700	9600	5500
Recovery, r	%	88.0	90.0	89	86

11–19 Estimate the SDI for the following filtered wastewater samples. If the water is to be treated with reverse osmosis will additional treatment be required?

	Volume filtered, mL			
	Wastewater sample number			
Test run time, min	1	2	3	4
2	315	480	180	500
5	575	895	395	700
10	905	1435	710	890
20	1425	2300	1280	1150

11–20 Calculate the modified fouling index (MFI) for the effluent from a microfiltration process (water sample to be selected by instructor) using the following experimental data:

	Volume filtered, L			Volume filtered, L	
	Water sample			Water sample	
Time, min	1	2	Time, min	1	2
0			3.5	6.78	7.17
0.5	1.50	1.50	4.0	7.48	8.03
1.0	2.50	2.50	4.5	8.08	8.87
1.5	3.45	3.48	5.0	8.57	9.67
2.0	4.36	4.40	5.5		10.34
2.5	5.22	5.37	6.0		10.97
3.0	6.03	6.28	6.5		11.47

11–21 Determine the cost (based on the current price of electricity) to treat a flowrate of 2500 m^3/d with a TDS concentration of 1300 g/m^3 and a cation and anion concentration of 0.13 g-eq/L using an electrodialysis unit. Assume the following typical values of operation for the electrodialysis unit.

Product flowrate = 90% of the feed water flowrate
Efficiency of salt removal = 50%
The current efficiency = 90%
Resistance = 5.0 ohms
Number of cell pairs in the stack = 350, 400, 450 (to be selected by instructor)
Assume an energy cost of \$0.13/kWh and 24 h/d operation.

11–22 Review and cite three current articles (within the last five years) dealing the disposal of nanofiltration, reverse osmosis, or electrodialysis brine. What types of process combinations are being proposed? What are the critical issues that stand out in your mind?

11–23 A wastewater is to be treated with activated carbon to remove residual COD. The following data were obtained from a laboratory adsorption study in which 1 g of activated carbon was added to a beaker containing 1 L of wastewater at selected COD values. Using these data, determine the more suitable isotherm (Langmuir or Fruendlich) to describe the data (sample to be selected by instructor).

Initial COD, mg/L	Equilibrium COD, mg/L			
	Wastewater sample number			
	1	2	3	4
140	5	10	0.4	5
250	12	30	0.9	18
300	17	50	2	28
340	23	70	4	36
370	29	90	6	42
400	36	110	10	50
450	50	150	35	63

11–24 Using the following isotherm test data, determine the type of model that best describes the data and the corresponding model parameters. Assume that a 1 L sample volume was used for each of the isotherm experiments.

Mass of GAC, mg	Equilibrium concentration of adsorbate in solution, C_e, μg/L			
	Test number			
	1	2	3	4
0	5.8	26	158.2	25.3
0.001	3.9	10.2	26.4	15.89
0.01	0.97	4.33	6.8	13.02
0.1	0.12	2.76	1.33	6.15
0.5	0.022	0.75	0.5	2.1

11–25 Using the results from Problem 11–23, determine the amount of activated carbon that would be required to treat a flowrate of 4800 m^3/d to a final COD concentration of 2 mg/L if the COD concentration after secondary treatment is equal to 30 mg/L.

11–26 Design a fixed-bed activated carbon process using the following data. Determine the number of contactors, mode of operation, carbon requirements, and corresponding bed life. Ignore the effects of biological activity within the column.

Parameter	Unit	Compound			
		Chloroform	Heptachlor	Methylene chloride	NDMA
Flowrate	m^3/d	4000	4500	5000	6000
C_o	ng/L	500	50	2000	200
C_e	ng/L	50	10	10	10
GAC density	g/L	450	450	450	450
EBCT	min	10	10	10	10

11–27 Referring to the data presented in Table 11–23, prepare a list of the top 5 most and least readily adsorbable substances.

11–28 Using the results from Prob. 11–13, determine the amount of activated carbon that would be required to treat a flowrate of 5000 m^3/d to a final COD concentration of 20 mg/L if the COD concentration after secondary treatment is equal to 120 mg/L.

11–29 Using the following carbon adsorption data (sample number to be selected by instructor) determine the Freundlich capacity factor (mg absorbate/g activated carbon) and Freundlich intensity parameter, $1/n$.

Carbon dose, mg/L	Residual concentration, mg/L — Sample number 1	2	3	4	5	6
0	25.9	9.20	9.89	27.5	20.4	9.88
5	17.4	7.36	9.39	24.8	19.3	7.95
10	13.2	6.86	8.96	24.2	18.6	7.02
25	10.2	3.86	7.83	18.9	16.1	3.66
50	3.6	1.13	5.81	11.8	12.2	0.98
100	2.5	0.22	4.45	2.3	6.7	0.25
150	2.1	0.18	2.98	1.1	3.1	0.09
200	1.4	0.11	2.01	0.9	1.1	0.04

11–30 Determine the theoretical air flowrate required to remove the following compounds in a stripping tower at the indicated concentrations (compound and water to be selected by instructor). Also estimate the height of the stripping tower for a water flowrate of 3000 m^3/d. Values of the Henry's law constant may be found in Table 16–12.

		Concentration μg			
		Water 1		Water 2	
Compound	K_La, s^{-1}	Influent	Effluent	Influent	Effluent
Clorobenzene	0.0163	100	5	120	7
Chloroethene	0.0141	100	5	150	5
TCE[a]	0,0176	100	5	180	10
Toluene	0.0206	100	5	200	15

[a] Henry's law constant for TCE is equal to 0.00553 m^3·atm/mole

11–31 A quantity of sodium-form ion exchange resin (5 g) is added to a water containing 2 meq of potassium chloride and 0.5 meq of sodium chloride. Calculate the residual concentration of potassium if the exchange capacity of the resin is 4.0 meq/g of dry weight and the selectivity coefficient is equal to 1.46.

11–32 Determine the exchange capacity for one of the following resins (resin to be selected by instructor). How much resin would be required to treat a flowrate of 4000 m^3/d to reduce the concentration of calcium (Ca^{2+}) from 125 to 45 mg/L? Assume the mass of resin used to obtain the data given in the table is 0.1 kg.

Throughput volume, L	Resin 1		Resin 2	
	Cl^-	Ca^{2+}	Cl^-	Ca^{2+}
0	0	0	0	0
5	2	0	2	0
10	8	0	13	0
15	44	0	29	0
20	65	0	45	0
25	70	0	60	1
30	71	0	69	8
35	71	6	71	17
40	71	20	71	27

(continued)

(*Continued*)

Throughput volume, L	Resin 1		Resin 2	
	Cl^-	Ca^{2+}	Cl^-	Ca^{2+}
45		34	71	35
50		39		39
55		40		40
60		40		40

11–33 Determine the exchange capacity for one of the resins given in Problem 11–32 (resin to be selected by instructor). How much resin would be required to treat a flowrate of 5500 m^3/d to reduce the concentration of magnesium, Mg^{2+}, from 115 to 15 mg/L?

11–34 Four different wastewaters have been reported to have the following ionic composition data. Estimate the selectivity coefficient and determine the amount of wastewater (1, 2, 3, or 4, to be selected by instructor) that can be treated by a strong-base ion exchange resin, per service cycle, for the removal of nitrate. Assume the resin has an ion exchange capacity of 1.8 eq/L.

			Concentration, mg/L			
			Wastewater sample number			
Cation	Conc., mg/L	Anion	1	2	3	4
Ca^{2+}	82.2	HCO_3^-	304.8	152	254	348
Mg^{2+}	17.9	SO_4^{2-}	0	0	0	0
Na^+	46.4	Cl^-	58.1	146.3	124	60
K^+	15.5	NO_3^-	82.5	90	21.5	42

11–35 Four different wastewaters have been reported to have the following ionic composition data. Estimate the selectivity coefficient and determine the amount of wastewater (1, 2, 3, or 4, to be selected by instructor) that can be treated by a strong-base ion exchange resin, per service cycle, for the removal of nitrate. Assume the resin has an ion exchange capacity of 2.5 eq/L.

			Concentration, mg/L			
			Wastewater sample number			
Cation	Conc., mg/L	Anion	1	2	3	4
Ca^{2+}	82.2	HCO_3^-	321	180	198.5	69
Mg^{2+}	17.9	SO_4^{2-}	65	36.5	124	136
Na^+	46.4	Cl^-	22	95	56	87
K^+	15.5	NO_3^-	46	93	34.5	97

11–36 For each compound in the following list, which of the treatment methods discussed in this chapter, if any, are suitable for use to reduce the concentration from 100 to 10 $\mu g/L$?

Benzene

Chloroform

Dieldrin

Heptachlor

N-Nitrosodimethylamine

Trichloroethylene (TCE)

Vinyl chloride

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12

Disinfection Processes

WORKING TERMINOLOGY

Term	Definition
Absorbance	A measure of the amount of light of a specified wavelength that is absorbed by a solution and the constituents in the solution.
Breakpoint chlorination	A process whereby enough chlorine is added to react with all oxidizable substances in water such that if additional chlorine is added it will remain as free chlorine (see below, HOCl + OCl^-).
Chlorine residual, total	The concentration of free or combined chlorine in water, measured after a specified time period following addition. Combined chlorine residual is measured most commonly amperometrically.
Combined chlorine	Chlorine combined with other compounds [e.g., monochloramine (NH_2Cl), dichloramine ($NHCl_2$), and nitrogen trichloride (NCl_3), among others].
Combined chlorine residual	Chlorine residual comprised of combined chlorine compounds [e.g., monochloramine (NH_2Cl), dichloramine ($NHCl_2$), and nitrogen trichloride (NCl_3) and others].
CT	The product of disinfectant residual, C, expressed in mg/L and contact time, T, expressed in min. The term CT is used to assess the effectiveness of the disinfection process.
Dechlorination	The removal of residual chlorine from solution by a reducing agent such as sulfur dioxide or by reacting it with activated carbon.
Disinfection	The partial destruction and inactivation of disease-causing organisms from exposure to chemical agents (e.g., chlorine) or physical processes (e.g., UV radiation).
Disinfection byproducts (DBPs)	Chemicals that are formed with the residual organic matter found in treated wastewater as a result of the addition of a strong oxidant (e.g., chlorine or ozone) for the purpose of disinfection.
Dose response curve	The relationship between the degree of microorganism inactivation and the dose of the disinfectant.
Free chlorine	The total quantity of hypochlorous acid (HOCl) and hypochlorite ion (OCl^-) in solution.
Inactivation	Rendering microorganisms incapable of reproducing, and thus their ability to cause disease.
Irradiation	Exposure to penetrating UV radiation.
Natural organic matter (NOM)	Dissolved or particulate organic constituents that are typically derived from three sources: (1) the terrestrial environment (mostly humic materials), (2) the aquatic environment (algae and other aquatic species and their byproducts), and (3) the microorganisms in the biological treatment process.
Pasteurization	The process of heating food or water at a specified temperature and time for the purpose of killing microorganisms.
Pathogens	Microorganisms capable of causing diseases of varying severity.
Photoreactivation/dark repair	The ability of microorganisms to repair the damage caused by exposure to UV irradiation.
Radiation	Energy such as light, heat, and sound that can be transmitted over large distances without conductors or special conduits.
Reduction equivalent dose (RED)	The inactivation observed through the UV disinfection system as compared to the UV dose response derived from a collimated beam dose response study.
Sterilization	The total destruction of disease-causing and other organisms.
Total chlorine	The sum of the free and combined chlorine.
Transmittance	The ability of a solution to transmit light. Transmittance is related to absorbance.
Ultraviolet (UV) light	Electromagnetic radiation with a wavelength less than that of visible light in the range from 100 to 400 nm.
Ultraviolet (UV) irradiation	A disinfection process in which the exposure to UV radiation (or light) is used to inactivate microorganisms.

Because of the critical importance of the disinfection process in wastewater treatment and/or reuse applications, the purpose of this chapter is to introduce the reader to the important issues that must be considered in the disinfection of treated water with various disinfectants to render it safe for dispersal to the environment or for reuse in a variety of applications. The four categories of human enteric organisms found in wastewater that are of the greatest consequence in producing disease are (1) bacteria, (2) protozoan oocysts and cysts, (3) viruses, and (4) helminth ova. Diseases caused by these waterborne microorganisms have been discussed previously in Chap. 2. Disinfection, the subject of this chapter, is the process used to achieve a given level of destruction or inactivation of pathogenic organisms. Because not all the organisms present are destroyed during the process, the term *disinfection* is differentiated from the term *sterilization*, which is the destruction of all organisms.

To delineate the issues involved in disinfection the following topics are considered: (1) an introduction to the disinfectants used in wastewater, (2) general considerations in wastewater disinfection, (3) disinfection with chlorine and related compounds, (4) disinfection with chlorine dioxide, (5) dechlorination, (6) design considerations for chlorination and dechlorination facilities, (7) disinfection with ozone, (8) disinfection with other chemicals and combination of chemicals, (9) disinfection with UV irradiation, and (10) disinfection by pasteurization.

12–1 INTRODUCTION TO DISINFECTANTS USED IN WASTEWATER

Before discussing the details of the individual disinfection technologies and the practical aspects of disinfection that follow, it is appropriate to consider the characteristics of an ideal disinfectant, the major types of disinfection agents used for wastewater, and to provide a general comparison between disinfectants.

Characteristics for an Ideal Disinfectant

To provide a perspective on the disinfection of wastewater, it is useful to consider the characteristics of an ideal disinfectant as given in Table 12–1. As reported, an ideal disinfectant would have to possess a wide range of characteristics such as safe to handle and apply, stable in storage, toxic to microorganisms, nontoxic to higher forms of life, and soluble in water or cell tissue. It is also important that the strength or concentration of the disinfectant be measurable. The latter consideration is an issue with the use of ozone, where little or no residual may remain after disinfection, and UV and pasteurization disinfection where no residual is measurable.

Disinfection Agents and Methods

Disinfection is most commonly accomplished by the use of (1) chemical agents and (2) nonionizing radiation. Each of these techniques is considered briefly in the following discussion. Other methods of disinfection and or inactivation are mentioned for completeness.

Chemical Agents. Chlorine and its compounds, and ozone, are the principal chemical compounds employed for the disinfection of wastewater. Other chemical agents that have been used as disinfectants in different applications include (1) bromine, (2) iodine, (3) phenol and phenolic compounds, (4) alcohols, (5) heavy metals and related compounds,

Table 12–1

Characteristics of an ideal disinfectant[a]

Characteristic	Properties/response
Alteration of solution characteristics	Should be effective with minimum alteration of the solution characteristics such as increasing the total dissolved solids (TDS)
Availability	Should be available in large quantities and reasonably priced
Deodorizing ability	Should deodorize while disinfecting
Homogeneity	Solution must be uniform in composition
Interaction with extraneous material	Should not be absorbed by organic matter other than bacterial cells
Noncorrosive and nonstaining	Should not disfigure metals or stain clothing
Nontoxic to higher forms of life	Should be toxic to microorganisms and nontoxic to humans and other animals
Penetration	Should have the capacity to penetrate through particle surfaces
Safety	Should be safe to transport, store, handle, and use
Solubility	Must be soluble in water or cell tissue
Stability	Should have low loss of germicidal action with time on standing
Toxicity to microorganisms	Should be effective at high dilutions
Toxicity at ambient temperatures	Should be effective in ambient temperature range

[a] Adapted from Tchobanoglous et al. (2003).

(6) dyes, (7) soaps and synthetic detergents, (8) quaternary ammonium compounds, (9) hydrogen peroxide, (10) peracetic acid, (11) various alkalies, and (12) various acids. Highly acidic or alkaline water will destroy pathogenic bacteria, because water with a pH greater than 11 or less than 3 is relatively toxic to most microorganisms.

Non-Ionizing Radiation. In general, energy in the form of electromagnetic waves, heat, and acoustic waves that can be transmitted over large distances without conductors or special conduits is termed *radiation*. Electromagnetic waves include visible light, infrared light, microwaves, and radio waves. Ultraviolet light (UV) is the most common form of electromagnetic radiation used for the disinfection of treated wastewater. Heating water to the boiling point, for example, will destroy the major disease producing non-spore forming bacteria. Commonly used in the food processing industry, pasteurization in the wastewater field has received greater interest recently because of the availability of new equipment, the opportunity to utilize waste heat, and energy concerns with other disinfectants. Pasteurization of sludge is used extensively in Europe.

Ionizing Radiation. Radiation with sufficient energy to ionize atoms is termed *ionizing radiation*. Alpha particles, beta particles, gamma rays, X-ray radiation, and neutrons are generally considered to be forms of ionizing radiation. For example, gamma rays emitted from radioisotopes, such as cobalt 60, have been used to disinfect (sterilize) both water and wastewater. Although the use of a high-energy electron-beam device for the irradiation of wastewater or sludge has been studied extensively, there are no commercial devices or full-scale installations in operation.

Removal by Mechanical Means. Incidental removal of bacteria and other organisms also occurs by mechanical means during wastewater treatment. The removals accomplished are byproducts of the primary function of the treatment process (e.g., screening, sedimentation, filtration, etc.). The use of membrane filtration (e.g., microfiltration and ultrafiltration) has been recognized as a means to reduce pathogenic organisms for water reuse applications. The level of reduction is assessed by spiking a known concentration of an indicator organism and measuring the inactivation achieved. In a full-scale operation, surrogate parameters, such as pressure decay across the membrane and turbidity, are used to monitor the integrity of the membrane. Pathogen removal by membrane processes is discussed further in Chap. 11.

Mechanisms Used to Explain Action of Disinfectants

The five principal mechanisms that have been proposed to explain the action of disinfectants are (1) damage to the cell wall, (2) alteration of cell permeability, (3) alteration of the colloidal nature of the protoplasm within the cell, (4) alteration of the organism's DNA or RNA, and (5) inhibition of enzyme activity within the protoplasm. A comparison of the mechanisms of disinfection using chlorine, ozone, UV irradiation, and pasteurization is presented in Table 12–2. To a large extent, observed performance differences for the various disinfectants can be explained on the basis of the operative inactivation mechanisms.

Damage, destruction, or alteration of the cell wall by oxidizing chemicals, such as chlorine and ozone, results in cell lysis and death. Oxidizing chemicals can also alter the chemical arrangement of enzymes and inactivate the enzymes. Some oxidants can inhibit the synthesis of the bacterial cell wall. Exposure to UV irradiation can cause the formation of double bonds in the DNA of microorganisms as well as rupturing some DNA strands. When UV photons are absorbed by the DNA in bacteria and protozoa and the DNA and RNA in viruses, covalent dimers can be formed from adjacent thymines in DNA or uracils in RNA. The formation of double bonds disrupts the replication process so that the organism can no longer reproduce and is thus inactivated. When heat is applied, both the

Table 12–2

Mechanisms of disinfection using chlorine, ozone, UV, and pasteurization

Chlorine	Ozone	UV radiation	Pasteurization
1. Direct oxidation of cell wall allowing cellular constituents to flow out of the cell 2. Modification of cell wall permeability 3. Alteration of the cell protoplasm 4. Inhibition of enzyme activity 5. Damage to the cell DNA and RNA	1. Direct oxidation of cell wall allowing cellular constituents to flow out of the cell 2. Reactions with radical byproducts of ozone decomposition 3. Damage to the constituents of the nucleic acids (purines and pyrimidines) 4. Breakage of carbon-nitrogen bonds leading to depolymerization	1. Photochemical damage to RNA and DNA (e.g., formation of double bonds) within the cells of an organism 2. The nucleic acids in microorganisms are the most important absorbers of the energy of light in the wavelength range of 240–280 nm 3. Because DNA and RNA carry genetic information for reproduction, damage of these substances can effectively inactivate the cell	1. The structure of the enzymes within the cell is altered by heat (e.g. denatured), rendering them inoperative 2. The structure of the proteins and fatty acids that make up the cell wall are damaged by heat, allowing contents of cell to escape 3. The fluids within the cell can expand and rupture the cell wall, releasing the contents of the cell

nature of the enzymes in the cell protoplasm and the structure of the cell wall are altered, rendering the microorganism incapable of reproducing.

Comparison of Disinfectants

Using the criteria defined in Table 12–1 as a frame of reference and the issues discussed above, the disinfectants that have been used in wastewater applications are compared in Table 12–3. Additional details on the relative performance of the various disinfection technologies are presented in the following sections. In reviewing Table 12–3, important comparisons that should be noted include safety (e.g., chlorine gas versus sodium hypochlorite) and the increase in TDS (e.g., chlorine gas versus UV irradiation). These issues are also addressed in the subsequent sections.

12–2 DISINFECTION PROCESS CONSIDERATIONS

The purpose of this section is to present background material that will serve as a basis for the discussion of individual disinfectants considered in the following sections. Topics to be discussed include (1) an introduction to the physical facilities used for disinfection, (2) the factors that affect the performance of the disinfection process, (3) development of CT values (residual disinfectant concentration times time) for predicting disinfection performance, (4) application of CT values, (5) a comparison of the performance of alternative disinfection technologies, and (6) a review of the advantages and disadvantages of each disinfection technology. Costs, both capital and operation and maintenance, have not been provided other than in a general context. Costs are influenced by many site-specific factors and must be evaluated on a case-by-case basis.

Physical Facilities Used for Disinfection

In general, the disinfection is accomplished as a separate unit process in specially designed reactors. The purpose of the reactors is to maximize contact between the disinfecting agent and the liquid to be disinfected. The specific design of the reactor depends on the nature and action of the disinfecting agent. The types of reactors used are illustrated on Figs. 12–1 and 12–2 and described below briefly.

Chlorine and Related Compounds. As shown on Figs. 12–1(a) and 12–1(b) baffled serpentine contact chambers or long pipelines are used for the application of diluted chlorine and related compounds. Both of these contact chambers are designed to perform as ideal plug-flow reactors. As will be discussed later, the efficacy of disinfection is affected by the degree to which the flow in these chambers is less than ideal. Views of full scale chlorine contact basins are shown on Figs. 12–2(a) and 12–2(b).

Ozone. Ozone is typically applied by bubbling ozone gas through the liquid to be disinfected in a contact chamber [see Fig. 12–1(c)] or in a sidestream [see Fig. 12–1(d)] and then injected into an ozone contactor [see Fig. 12–2(d)]. Fine bubble diffusers are used to improve ozone transfer to the liquid. Eductors and Venturi injectors are used in sidestream designs. To limit the amount of short circuiting that can occur in a single contact chamber a series of baffled chambers are used [see Fig. 12–1(c)].

Ultraviolet Light (UV). Both open [see Figs. 12–1(e) and (f)] and closed [see Fig. 12–1(g)] contact chambers (reactors) are used for UV disinfection. Open channel reactors are used commonly for low-pressure, low-intensity and low-pressure, high-intensity

Table12–3

Comparison of technologies used for the disinfection of treated wastewater[a]

Characteristic[b]	Chlorine gas[c]	Sodium hypochlorite[c]	Combined chlorine	Chlorine dioxide	Ozone	UV radiation	Pasteurization
Availability/cost	Low	Moderately low	Moderately low	Moderately low	Moderately high	Moderately high	Moderate
Deodorizing ability	High	Moderate	Moderate	High	High	na[d]	na
Interaction with organic matter	Oxidizes organic matter	Oxidizes organic matter	Oxidizes organic matter	Oxidizes organic matter	Oxidizes organic matter	Absorbance of UV radiation	na
Corrosiveness	Highly corrosive	Corrosive	Corrosive	Highly corrosive	Highly corrosive	na	na
Toxic to higher forms of life	Highly toxic	Highly toxic	Toxic	Toxic	Toxic	Toxic	Toxic
Penetration into particles	High	High	Moderate	High	High	Moderate	High
Safety concern	High	Moderate to low	High to moderate[e]	High	Moderate	Low	Low
Solubility	Moderate	High	High	High	Moderate	na	na
Stability	Stable	Slightly unstable	Slightly unstable	Unstable[f]	Unstable[f]	na	na
Effectiveness for							
Bacteria	Excellent	Excellent	Good	Excellent	Excellent	Good	Excellent
Protozoa	Fair to poor	Fair to poor	Poor	Good	Good	Excellent	Excellent
Viruses	Excellent	Excellent	Fair	Excellent	Excellent	Good	Good
Byproduct formation	THMs and HAAs[g]	THMs and HAAs[g]	Traces of THMs and HAAs, cyanogens, NDMA	Chlorite and chlorate	Bromate	None known in measurable concentrations	None known in measurable concentrations
Increases TDS	Yes	Yes	Yes	Yes	No	No	No
Use as a disinfectant	Common	Common	Common	Increasing slowly	Increasing slowly	Increasing rapidly	Increasing slowly

[a] Adapted in part from Tchobanoglous et al. (2003) and Crittenden et al. (2012).

[b] See Table 12–1 for a description of the characteristics of an ideal disinfectant.

[c] Free chlorine (HOCl and OCl^-).

[d] na = not applicable.

[e] Depends on whether chlorine gas or sodium hypchlorite is used to combine with nitrogenous compounds.

[f] Must be generated as used.

[g] THMs = trihalomethanes and HAAs = haloacetic acids.

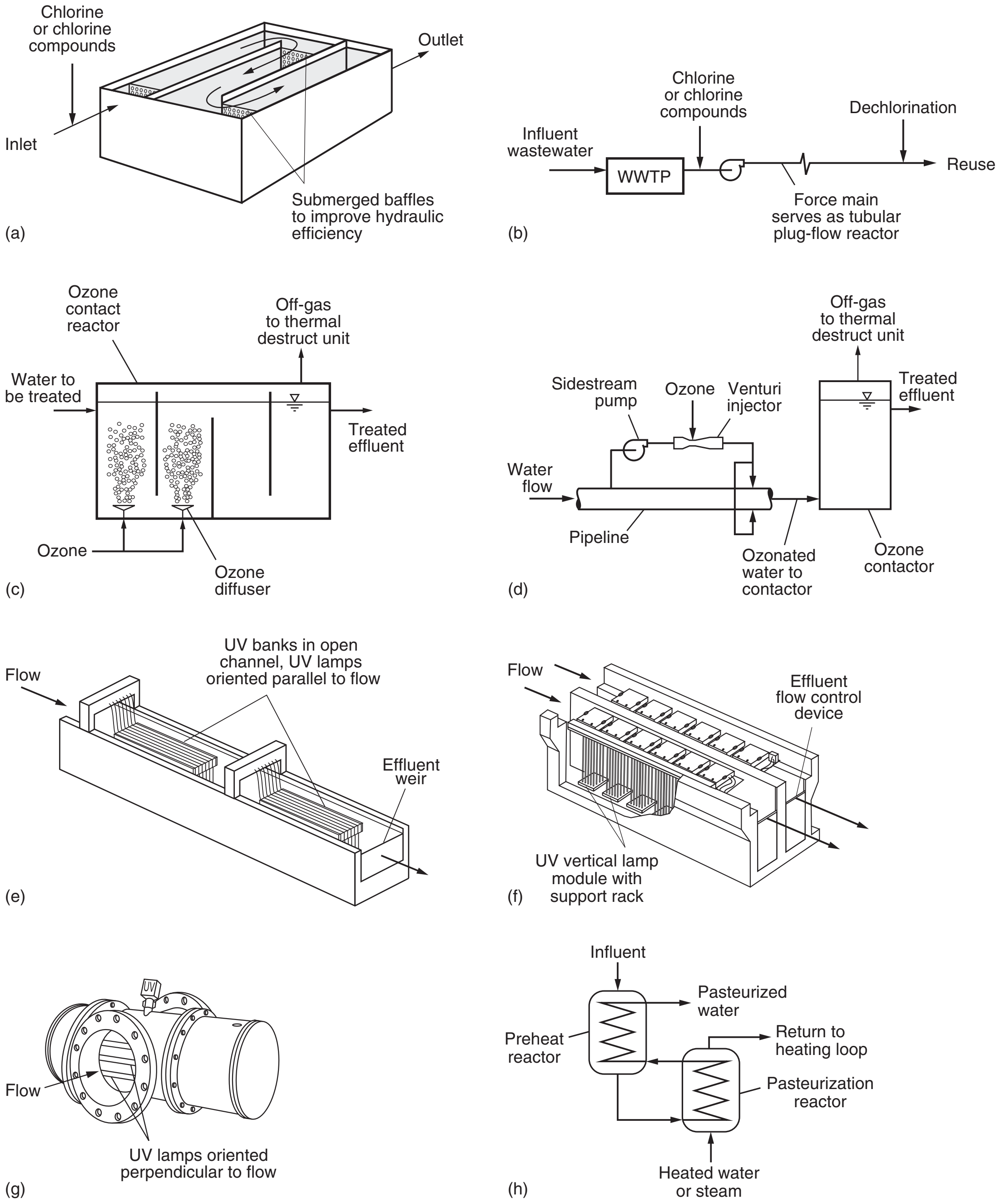
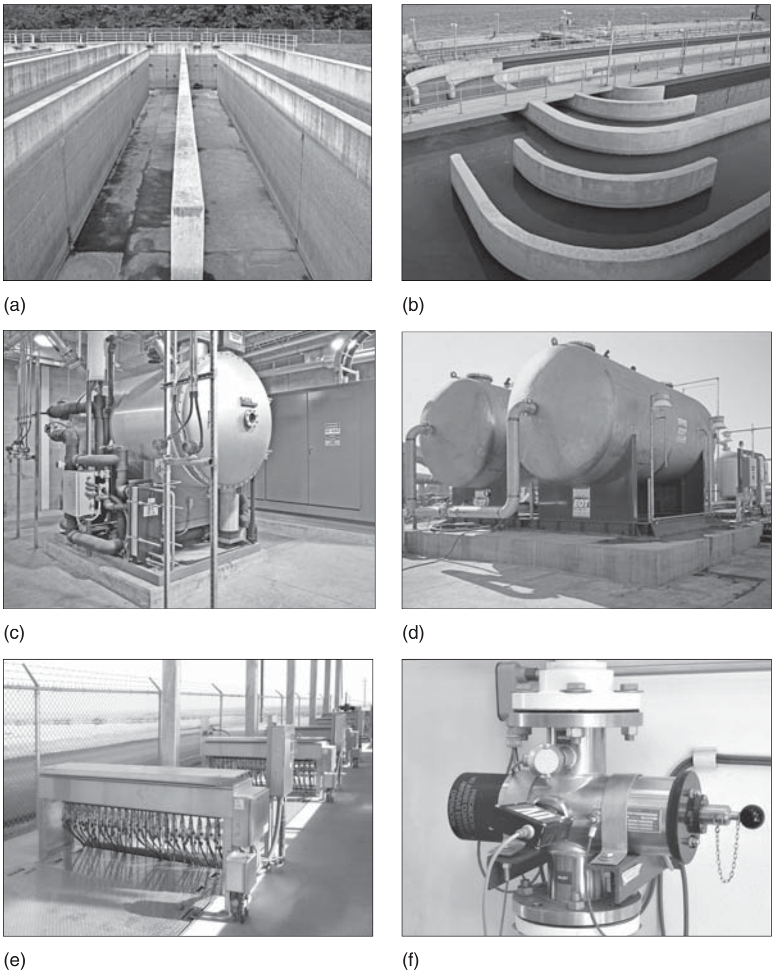


Figure 12–1

Types of reactors used to accomplish disinfection process: (a) plug-flow reactor in back-and-forth configuration, (b) force main which serves as a tubular plug-flow reactor, (c) multiple chamber inline ozone contactor, (d) sidestream ozone injection system, (e) UV irradiation in an open channel with two UV banks with flow parallel to UV lamps, (f) UV irradiation in an open channel with six UV banks with flow perpendicular to UV lamps, (g) UV irradiation in an closed reactor, and (h) reactors for pasteurization system.

Figure 12–2

Views of reactors used for disinfection: (a) serpentine plug-flow chlorine contact basin, (b) serpentine plug-flow chlorine contact basin with rounded corners and flow deflection baffles, (c) typical ozone generator, (d) view of ozone contactor used in conjunction with sidestream ozone injection, (e) view of open channel plug-flow UV reactor, and (f) view of closed channel UV reactor with manual lamp wiping device.



UV lamps. Closed proprietary reactors are used for low-pressure, high-intensity and medium pressure, high-intensity UV lamps. Because the contact time is short in UV reactors (seconds), the design of the open channel and closed reactors is of critical importance. Open plug-flow and closed channel UV reactors are shown on Figs. 12–2(e) and (f), respectively.

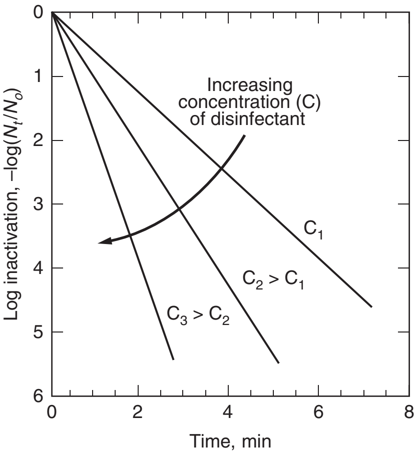
Pasteurization. The pasteurization process occurs in two reactors [see Fig. 12–1(h)]. In the first reactor, the liquid to be disinfected is preheated. Pasteurization occurs in the second reactor where the liquid to be disinfected is held for a specified period of time and temperature.

Factors Affecting Performance

In applying disinfection agents or physical processes, the following factors must be considered: (1) contact time and hydraulic efficiency of contact chambers, (2) concentration of the disinfectant, (3) intensity and nature of physical agent or means, (4) temperature, (5) types of organisms, (6) nature of suspending liquid (e.g., unfiltered or filtered secondary effluent), and (7) the upstream treatment processes. The subjects introduced in this

Figure 12–3

Log inactivation of dispersed microorganisms as a function of time in a batch reactor using increasing disinfectant dosages.



section are considered further in the subsequent sections dealing with the individual disinfectants.

Contact Time. Perhaps one of the most important factors in the disinfection process is contact time. Once the disinfectant has been added, the time of contact before the effluent is to be discharged or reused is of paramount importance. As shown on Fig. 12–1, disinfection reactors are designed to ensure that an adequate contact time is provided. The hydraulic efficiency of disinfection reactors is considered in Sec. 12–6.

Working in England in the early 1900s, Harriet Chick observed that for a given concentration of disinfectant, the longer the contact time, the greater the kill (see Fig. 12–3). This observation was first reported in the literature in 1908 (Chick, 1908). In differential form, Chick's law is

$$\frac{dN_t}{dt} = -KN_t \tag{12–1}$$

where dN_t/dt = the rate of change in the number (concentration) of organisms with time
K = inactivation rate constant, T^{-1}
N_t = number of organisms at time t
t = time

If N_o is the number of organisms when t equals 0, Eq. (12–1) can be integrated to

$$\ln\frac{N_t}{N_o} = -Kt \tag{12–2}$$

The value of the inactivation rate constant, K, in Eq. (12–2) can be obtained by plotting $-\ln(N_t/N_o)$ versus the contact time t, where K is the slope of the resulting line of best fit.

Concentration of Chemical Disinfectant. Also working in England in the early 1900s, Herbert Watson reported that the inactivation rate constant was related to the concentration as follows (Watson, 1908):

$$K = \Lambda C^n \tag{12–3}$$

where K = inactivation rate constant, T^{-1}, base e
Λ = coefficient of specific lethality, units vary with the value of n
C = concentration of disinfectant, mg/L
n = empirical constant related to dilution, dimensionless

The following explanation has been offered for various values of the dilution constant n:

$n = 1$, both the concentration and time are equally important
$n > 1$, concentration is more important than time
$n < 1$, time is more important than concentration

The value of n can be obtained by plotting C versus t on log-log paper for a given level of inactivation. When n is equal to one, the data are plotted on log-arithmetic paper.

Combining the expressions proposed by Chick and Watson in differential form yields (Haas and Karra, 1984a, b, c):

$$\frac{dN_t}{dt} = -\Lambda C^n N_t \tag{12–4}$$

The integrated form of Eq. (12–4) is:

$$\ln \frac{N_t}{N_o} = -\Lambda_{\text{base}\,e} C^n t \text{ or } \log \frac{N_t}{N_o} = -\Lambda_{\text{base}\,10} C^n t \tag{12–5}$$

If n is equal to one, a reasonable assumption based on past experience (Hall, 1973), Eq. (12–5) can be written as follows:

$$\log \frac{N_t}{N_o} = -\Lambda_{\text{base}\,10}(\mathrm{CT}) = -\Lambda_{\text{base}\,10}(D) \tag{12–6}$$

Where C = residual concentration of disinfectant, mg/L
T = contact time in the reactor, min
D = *germicidal dose* for a given degree of inactivation, mg·min/L

The concept of dose (concentration times time) is significant as the performance of the disinfectants, as discussed subsequently, is based on the concept (Morris, 1975). This concept has also been adopted by the U.S. EPA in establishing guidelines for the disinfection of public water supplies (see "Development of the CT Concept for Predicting Disinfection Performance" later in the chapter).

EXAMPLE 12–1 **Determination of the Coefficient of Specific Lethality Based on the Chick-Watson Expression** Using the microorganism inactivation data given below, determine the coefficient of specific lethality of the chemical disinfecting agent using Eq. (12–6).

C, mg/L	Time, min	Number of organisms, Number/100 mL
0	0	1.00×10^8
4.0	2	1.59×10^7
4.0	4.5	1.58×10^6
4.0	8	2.01×10^4
4.0	11.5	3.16×10^3

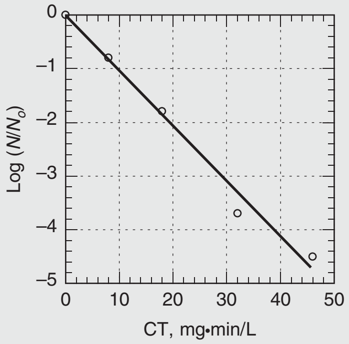
Solution

1. To determine the coefficient of lethality prepare a plot of $\log[N/N_o]$ as a function of CT and fit a linear trend line through the data.

a. Determine the values of log[N/N_o] and CT. The required data table is shown below.

C, mg/L	Time, min	Number of organisms, N/100 mL	CT, mg·min/L	log(N/N_o)
0	0	1.00×10^8	0	0
4.0	2	1.59×10^7	8	−0.8
4.0	4.5	1.58×10^6	18	−1.8
4.0	8	2.01×10^4	32	−3.7
4.0	11.5	3.16×10^3	46	−4.5

b. Prepare a plot of log(N/N_o) as a function of CT. The required plot is shown below.



2. Determine the coefficient of specific lethality. The slope of the line in the above plot corresponds to the coefficient of specific lethality, $-\Lambda_{CW}$ (base 10). From the plot

$$-\Lambda_{CW}\text{ (base 10)} = \frac{-5 - 0}{49 - 0}$$

$$\Lambda_{CW}\text{(base 10)} = 0.102\,\text{L/mg·min}$$

Check, when CT = 46,

$$\log\frac{N_t}{N_o} = -\Lambda_{\text{base }10}\text{CT} = -0.102(46) = -4.69\text{ versus } -4.5\text{ OK}$$

Temperature. The effect of temperature on the rate of kill with chemical disinfectants can be represented by a form of the van't Hoff-Arrhenius relationship. Increasing the temperature results in a more rapid kill. In terms of the coefficient of specific lethality, Λ, the effect of temperature is given by the following relationship, repeated here from Chap. 1.

$$\ln\frac{\Lambda_1}{\Lambda_2} = \frac{E(T_2 - T_1)}{RT_1T_2} \tag{12–7}$$

where Λ_1, Λ_2 = coefficient of specific lethality at temperatures T_1 and T_2, respectively
E = activation energy, J/mole
R = universal gas constant, 8.3144 J/mole·K

Typical values for the activation energy for various chlorine compounds at different pH values are given in Sec. 12–3. The effect of temperature is considered in Example 12–2.

EXAMPLE 12–2 **Effect of Temperature on Disinfection Times** Estimate the time required for a 99 percent kill for a chlorine dosage of 0.05 mg/L at a temperature of 20°C. Assume the activation energy is equal to 26,800 J/mole (from Table 12–12 in Sec. 12–3). The following coefficients were developed for Eq. (12–5) at 5°C using a batch reactor.

$\Lambda = 10.5\ \text{L/mg·min}$

$n = 1$

Solution

1. Estimate the time required at 5°C for a 99 percent kill using Eq. 12–5.

$$\log \frac{N_t}{N_o} = -10.5\,\text{CT}$$

$$\log \frac{10}{100} = -(10.5\ \text{L/mg·min})(0.05\ \text{mg/L})T$$

$$T = \frac{-6.91}{(-10.5)(0.05)} = 13.2\ \text{min at } 5°\text{C}$$

2. Estimate the time required at 20°C using the van't Hoff-Arrhenius equation [Eq. (12–7)].

$$\ln\frac{\Lambda_1}{\Lambda_2} = \frac{E(T_2 - T_1)}{RT_1T_2}$$

$$\ln\frac{10.5}{\Lambda_2} = \frac{(26{,}800\ \text{J/mole})(278 - 293)\text{K}}{(8.3144\ \text{J/mole·K})(293)(298)}$$

$$\ln\frac{10.5}{\Lambda_2} = -0.594$$

$$\ln\frac{10.5}{\Lambda_2} = e^{-0.594} = -0.552$$

$$\Lambda_2 = 19.0\ \text{L/mg·min}$$

$$T = \frac{-6.91}{(-19.0)(0.05)} = 7.27\ \text{min at } 20°\text{C}$$

Intensity and Nature of Non-Ionizing Radiation. As noted earlier, irradiation with ultraviolet light (UV) is used commonly for the disinfection of water. It has been found that the effectiveness of UV disinfection is a function of the average UV intensity, expressed as milliwatts per square centimeter (mW/cm^2). When the exposure time is considered, the dose of UV to which the microorganisms in the liquid are exposed to is given by the following expression.

$$D = I_{\text{avg}} \times t \tag{12–8}$$

Where D = UV dose, mJ/cm^2 (Note: $\text{mJ/cm}^2 = \text{mW·s/cm}^2$)
I_{avg} = average UV intensity, mW/cm^2
t = time, s

The UV dose is expressed in mJ/cm^2 (millijoule per square centimeter) which is equivalent to mW·s/cm^2. Thus, the concept of dose can also be used to define the effectiveness of UV light in a manner analogous to that used for chemical disinfectants, as well as when heat is used as in pasteurization.

Types of Organisms. The effectiveness of various disinfectants is influenced by the type, nature, and condition of the microorganisms. For example, viable growing (vegetative) bacteria cells are often killed or inactivated more easily than older cells that have developed a slime (polymer) coating. Bacteria that are able to form spores enter this protective state when stressors, such as increased temperature or a toxic agent, is applied. Bacterial spores are extremely resistant, and many of the chemical disinfectants normally used have little or no effect on them. Similarly, many of the viruses and protozoa of concern respond differently to each of the chemical disinfectants. In some cases, other disinfecting agents, such as heat or UV irradiation, may have to be used for effective disinfection. The inactivation of different types of microorganism groups is considered in the following sections.

Nature of Suspending Liquid. In reviewing the development of the relationships developed by Chick and Watson for the inactivation of microorganisms, as cited above, it is important to note that most of the tests were conducted in batch reactors using distilled or buffered water, under laboratory conditions. In practice, the nature of the suspending liquid must be evaluated carefully. Three constituents found in wastewater are significant: (1) inorganic constituents that can react with the disinfectant, (2) organic matter including both natural organic material (NOM) and other organic compounds, and (3) suspended material. The NOM found in treated wastewater will react with most oxidizing disinfectants and reduce their effectiveness or result in greater dosages to effect disinfection. The NOM is derived from three sources: (1) the terrestrial environment (mostly humic materials), (2) the aquatic environment (algae and other aquatic species and their byproducts), and (3) the microorganisms in the biological treatment process. The source of the other organic compounds is from the constituents discharged to the collection system. The presence of suspended matter will also reduce the effectiveness of disinfectants by absorption of the disinfectant and by shielding the entrapped bacteria.

Because of the interactions that can occur between the disinfecting agent and the wastewater constituents, departures from the Chick-Watson rate law [Eqs.(12–5) and (12–6)] are common as shown on Fig. 12–4. As shown on Fig. 12–4(a), there can be a lag or shoulder effect in which constituents in the suspending liquid react initially with the disinfectant rendering the disinfectant ineffective followed by a log-linear portion. The tailing effect in which large particles shield the organisms to be disinfected is shown on Fig. 12–4(b). The combined effects of lag and tailing are illustrated on Fig. 12–4(c). In general, Eq. (12–5) as applied to wastewater fails to account for the variable, heterogeneous characteristics of wastewater.

Effect of Upstream Treatment Processes. The extent to which upstream processes remove NOM, other organic matter, and suspended matter will greatly influence the disinfection process. Incidental removals of bacteria and other organisms are also achieved by mechanical and biological means during wastewater treatment. Typical removal efficiencies for various treatment operations and processes are reported in Table 12–4. The first and last four operations listed are essentially physical. The actual removal accomplished is a byproduct of the primary function of the process.

Another factor that impacts the performance of both chlorine and UV disinfection for unfiltered effluents (especially when coliform bacteria are used as the regulatory indicator) is the number of particles with associated coliform bacteria. It has been observed that for

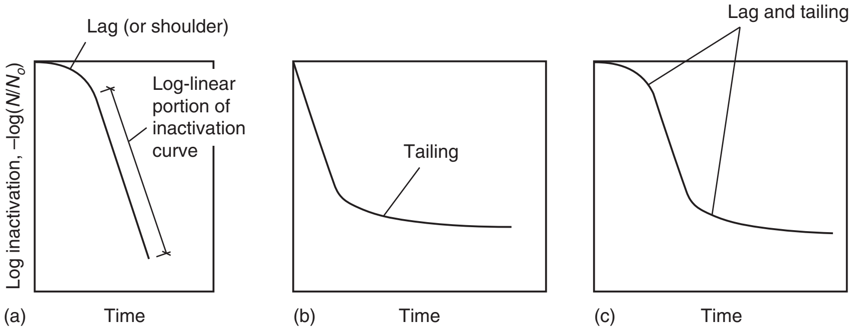


Figure 12–4

Departures observed from the Chicks' law: (a) lag or shoulder effect in which the disinfectant reacts first with constituents in the suspending liquid after which the response is log-linear (i.e., first order kinetics), (b) log-linear response followed by tailing effect in which large particles shield the organisms to be disinfected following the inactivation of dispersed organisms, and (c) combined lag, log-linear, and tailing effects.

activated sludge plants the number of particles with associated coliform organisms is a function of the solids retention time (SRT). The relationship between the fraction of wastewater particles with one or more associated coliform organisms and the SRT is illustrated on Fig. 12–5. As illustrated, longer SRTs result in a decrease in the fraction of particles containing coliform bacteria. The use of deep final clarifiers (or other filtration methods) reduces the number of large particles that may shield bacteria [see Fig. 7–7(b) in Chap. 7]. In general, without some form of filtration, it is difficult to achieve extremely low coliform concentrations in the settled effluent from activated sludge plants operated at low SRT values (e.g., 1 to 2 d).

Development of the CT Concept for Predicting Disinfection Performance

Although the disinfection models discussed above are useful for analyzing disinfection data, they are difficult to use to predict disinfection performance over a wide range of operating conditions. In the water treatment field, before the adoption of the Surface Water Treatment Rule (SWTR) (circa 1989) and before the importance of *Cryptosporidium* as a

Table 12–4

Removal or destruction of total coliform by different treatment processes

Process	Removal	
	Percent	log[a]
Coarse screens	0–5	~0
Fine screens	10–20	0–0.1
Grit chambers	10–25	0–0.1
Plain sedimentation	25–75	0.1–0.6
Chemical precipitation	40–80	0.2–0.7
Trickling filters	90–95	1–1.3
Activated sludge	90–98	1–1.7
Depth filtration	—	0.25–1
Microfiltration	—	2–4[b]
Ultrafiltration	—	2–5[b]
Reverse osmosis	—	2–6[b]

[a] The log-reduction credit allowed by regulatory agencies for these processes will vary from state to state.

[b] Depends on the characteristics and configuration of the membrane.

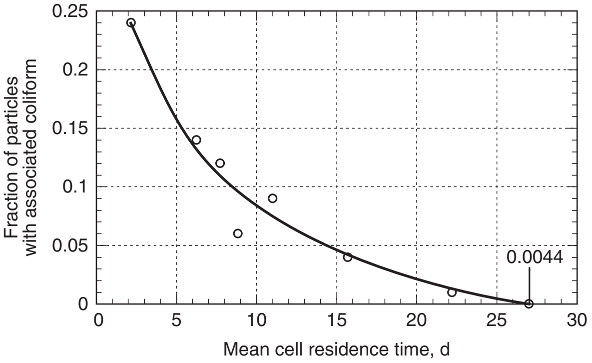


Figure 12–5

Fraction of particles in settled wastewater with one or more associated coliform organisms as function of the solids retention time. (From Emerick et al., 1999.)

causative agent in waterborne disease outbreaks was recognized, meeting water quality requirements was quite straightforward. Chlorine and its compounds were generally used to inactivate coliform bacteria to meet the drinking water standards in effect at that time.

In developing the rationale for the first SWTR, the U.S. EPA needed some way to ensure the safety of public water supplies that were unfiltered (e.g., New York City, San Francisco, Seattle). Based on ongoing research, the U.S. EPA determined that four logs of virus and three logs of *Giardia* reduction would be required by means of disinfection. Recognizing that guidance was required on how to achieve adequate disinfection, the U.S. EPA undertook an evaluation of the most commonly used disinfectants for the disinfection of viruses and *Giardia* cysts. In conducting their evaluation, the U.S. EPA adopted the CT concept (the product of the residual disinfectant concentration, C, in mg/L times and the contact time, T, in min), derived from the simplified Chick-Watson model [see Eq. (12–6)], as a measure of performance. The CT values obtained, typically in laboratory bench scale studies, are used as a surrogate measure of disinfection effectiveness. Thus, if a given CT value is achieved it could be assumed generally that disinfection requirements had been met. Bauman and Ludwig (1962) were among the first, if not the first, to suggest use of the CT concept in a paper published in 1962. The CT concept was not picked up again in a meaningful way until 1980 when it was used by the Safe Drinking Water Committee of the National Research Council in its evaluation of the disinfection literature (NRC, 1980; Hoff, 1986).

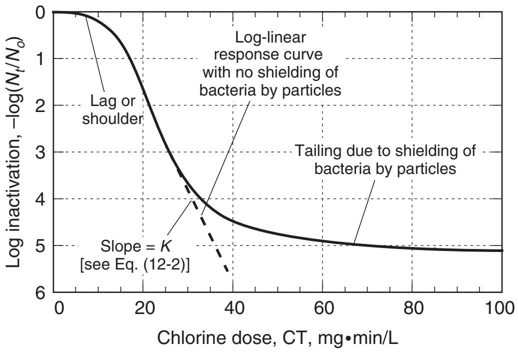
Although *Cryptosporidium* had been identified at the time the SWTR was adopted in 1989, CT values for *Cryptosporidium* were not included because it would have delayed adoption of the SWTR. It has since been found that many pathogens, including *Cryptosporidium,* remain intact and viable while in the presence of various disinfectants at concentrations that are sufficient to inactivate most other pathogens. Based on ongoing work, the U.S. EPA has now published extensive tables of CT values for a variety of disinfectants, microorganisms, and operating conditions (U.S. EPA, 2003a, 2006). In addition, corresponding UV dose values have also been published for *Cryptosporidium*, *Giardia,* and viruses. From a practical standpoint, the utility of the CT or UV dose approach can be appreciated as it is relatively easy to measure the residual concentration of the disinfectant or the UV intensity and the exposure contact time. With respect to the contact time, the t_{10} value (the contact time during which no more than 10 percent in the influent water has passed through the process–see discussion in Sec. 12–3) is used commonly in the field of water treatment for disinfectants other than UV irradiation.

Application of the CT Concept to Wastewater Disinfection

Use of the CT concept to control the disinfection process is now becoming more common in the wastewater field. In some states, the CT value and the chlorine contact time are

Figure 12–6

Typical disinfection dose-response curve obtained with wastewater containing oxidizable constituents and suspended solids. Both lag and tailing effects are evident.



specified in regulatory requirements. For example, the California Department of Public Health (CDPH) California requires a minimum CT value of 450 mg·min/L (based on combined chlorine residual) and a modal contact time of 90 min at peak flow for certain water reclamation applications. It is assumed, based on past testing, that a minimum CT value of 450 mg·min/L will produce a four-log inactivation of poliovirus. As the use of the CT concept becomes more common in the wastewater field, a number of past limitations must be considered in the application of this concept for regulatory purposes. In the past, most of the CT values reported in the literature are obtained using (1) complete-mix batch reactors (i.e., ideal plug flow conditions) in a laboratory setting under controlled conditions, (2) discrete organisms grown in the laboratory in pure culture, (3) a buffered fluid for the suspension of the discrete organisms, and (4) an absence of particulate matter.

Further, many of the CT values reported in the literature were based on older analytical techniques. As a consequence, CT values used for regulatory purposes often do not match what is observed in the field. Referring to Fig. 12–6, it can be seen that in the tailing region, the residual concentration of microorganisms is essentially independent of the CT value. In addition, some compounds present in treated wastewater will (1) react with the chlorine and its compounds, (2) be measured as combined residual, and (3) have no disinfection properties (see Sec. 12–3). In a similar manner, dissolved constituents, such as metals and humic acids, will reduce the effectiveness of UV disinfection. Thus, it is difficult to develop standardized CT or UV dose values suitable for all conditions encountered in wastewater treatment. Clearly, as discussed subsequently, site-specific testing is required to establish the appropriate disinfectant dose.

Performance Comparison of Disinfection Technologies

A general comparison of the germicidal effectiveness of the disinfection technologies based on Eq. (12–6) by classes of organisms is presented in Table 12–5. Additional information is presented in the sections dealing with the individual technologies. It is important to note that the values given in these tables are only meant to serve as a guide in assessing the effectiveness of these technologies. The CT values also vary with both temperature and pH. Because the characteristics of each wastewater and the degree of treatment will significantly impact the effectiveness of the various disinfection technologies, site-specific testing must be conducted to evaluate the effectiveness of alternative disinfection technologies and to establish appropriate dosing ranges.

Advantages and Disadvantages of Alternative Disinfection Technologies. The general advantages and disadvantages of using chlorine, chlorine dioxide, ozone, and UV for the disinfection of wastewater are summarized in Table 12–6. In most wastewater

Table 12–5

Relative CT values for various levels of inactivation of bacteria, viruses, *Cryptosporidium,* and *Giardia lamblia* cysts in filtered secondary effluent (pH ~7.5, ~20°C)[a,b]

		Inactivation			
Disinfectant	**Unit**	**1–log**	**2–log**	**3–log**	**4–log[c]**
Bacteria[d]					
Chlorine (free)	mg·min/L	0.4–0.6	0.8–1.2	1.2–1.8	1.6–2.4
Chloramine	mg·min/L	50–70	95–150	140–220	200–300
Chlorine dioxide	mg·min/L	0.4–0.6	0.8–1.2	1.2–1.8	1.6–2.4
Ozone	mg·min/L	0.005–0.01	0.01–0.02	0.015–0.03	0.02–0.04
UV radiation	mJ/cm^2	10–15	20–30	30–45	40–60
Virus					
Chlorine (free)	mg·min/L		1.5–1.8	2.2–2.6	3–3.5
Chloramine	mg·min/L		370–400	550–600	750–800
Chlorine dioxide	mg·min/L		5–5.5	9–10	12.5–13.5
Ozone	mg·min/L		0.25–0.3	0.35–0.45	0.5–0.6
UV radiation[e]	mJ/cm^2		40–50	60–75	80–100
Protozoa (*Cryptosporidium*)[f]					
Chlorine (free)	mg·min/L	2000–2600	4000–5000		
Chloramine	mg·min/L	4000–5000	8000–10,000		
Chlorine dioxide	mg·min/L	120–150	235–260	350–400	
Ozone	mg·min/L	4–4.5	8–8.5	12–13	
UV radiation	mJ/cm^2	2.5–3	6–7	12–13	
Protozoa (*Giardia lamblia* cysts)[g]					
Chlorine (free)	mg·min/L	20–30	45–55	70–80	
Chloramine	mg·min/L	400–450	800–900	1100–1300	
Chlorine dioxide	mg·min/L	5–5.5	9–11	15–16	
Ozone	mg·min/L	0.25–0.3	0.45–0.5	0.75–0.8	
UV radiation	mJ/cm^2	2–2.5	5.5–6.6	11–13	

[a] Adapted in part from AWWA (1991), Baumann and Ludwig (1962), Crittenden et al. (2012), Hoff (1986), Code of Federal Regulations – Title 40 (40 CFR 141.2), Maguin et al. (2009), Montgomery (1985), Roberts et al. (1980), Sung (1974), U.S. EPA (1999b).

[b] Reported CT values are highly temperature and pH sensitive. Disinfection rates will increase by a factor of 2 to 3 for each 10°C increase in temperature.

[c] The range of CT values for 4-log removal is for the linear portion of the dose-response curve (see Fig. 12–6). Depending on the particle size distribution resulting from the filtration of secondary effluent, much higher CT values may be needed to achieve a 4-log removal.

[d] The reported CT values are for total coliform. Significantly lower CT values have been reported for fecal coliform and *E coli*.

[e] With the exception of adenovirus which requires a much higher UV dose (as high as 160–200 mJ/cm^2 for 4-log inactivation).

[f] The data for *Cryptosporidium* inactivation with free or combined chlorine are extremely variable. Values of CT greater than 10,000 mg·min/L have been reported for 99 percent inactivation with chloramines. Clearly, free or combined chlorine is not an effective disinfectant for *Cryptosporidium*. Further, *Cryptosporidium* oocysts will, in general, require even higher CT values.

[g] Based primarily on the results of infectivity studies.

Note: Because there is such a wide variability in the susceptibility of different microorganism groups as well as within a microorganism group to the different disinfection technologies, a wide range of dosage values has been reported in the literature. Thus, the data presented in this table are only meant to serve as general guide to the relative effectiveness of the different disinfection technologies and are not for a specific microorganism.

Table 12–6

Advantages and disadvantages of chlorine, chlorine dioxide, ozone, and UV for the disinfection of treated wastewater[a]

Advantages	Disadvantages
Free and combined chlorine species	
1. Well established technology 2. Effective disinfectant 3. Chlorine residual can be monitored and maintained 4. Combined chlorine residual can also be provided by adding ammonia 5. Germicidal chlorine residual can be maintained in long transmission lines 6. Availability of chemical system for auxiliary uses such as odor control, dosing RAS, and disinfecting plant water systems 7. Oxidizes sulfides 8. Capital cost is relatively inexpensive, but cost increases considerably if conformance to Uniform Fire Code regulations is required 9. Available as calcium and sodium hypochlorite that are considered to be safer than chlorine gas 10. Hypochlorite can be generated onsite	1. Hazardous chemical that can be a threat to plant workers and the public; thus, strict safety measures must be employed especially in light of the Uniform Fire Code 2. Relatively long contact time required as compared to other disinfectants 3. Combined chlorine is less effective in inactivating some viruses, spores, and cysts at low dosages used for coliform organisms 4. Residual toxicity of treated effluent must be reduced through dechlorination 5. Forms trihalomethanes and other DBPs including NDMA[b] (see Table 12–16) 6. Releases volatile organic compounds from chlorine contact basins 7. Oxidizes iron, magnesium, and other inorganic compounds (consumes disinfectant) 8. Oxidizes a variety of organic compounds (consumes disinfectant) 9. Increases TDS level of treated effluent 10. Increases chloride content of treated effluent 11. Acid generation; pH of the wastewater can be reduced if alkalinity is insufficient 12. Chemical scrubbing facilities may be required to meet Uniform Fire Code regulations 13. Formal risk management plan may be required 14. Not effective disinfectant for *Cryptosporidium*
Chlorine dioxide	
1. Effective disinfectant for bacteria, *Giardia* and viruses 2. More effective than chlorine in inactivating most viruses, spores, cysts and oocysts 3. Biocidal properties not influenced by pH 4. Under proper generation conditions, halogen-substituted DBPs are not formed 5. Oxidizes sulfides 6. Provides residuals	1. Unstable, must be produced onsite 2. Oxidizes iron, magnesium, and other inorganic compounds (consumes disinfectant) 3. Oxidizes a variety of organic compounds 4. Forms DBPs (i.e., chlorite and chlorate), limiting applied dose 5. Potential for the formation of halogen-substituted DBPs 6. Decomposes in sunlight 7. Can lead to the formation of odors 8. Increases TDS level of treated effluent 9. Operating costs can be high (e.g., must test for chlorite and chlorate)

(*continued*)

Table 12–6 (*Continued*)

Advantages	Disadvantages
Ozone	
1. Effective disinfectant	1. Ozone residual monitoring and recording requires more operator time than chlorine residual monitoring and recording
2. More effective than chlorine in inactivating most viruses, spores, cysts and oocysts	2. No residual effect
3. Biocidal properties not influenced by pH	3. Less effective in inactivating some viruses, spores, cysts at low dosages used for coliform organisms
4. Shorter contact time than chlorine	4 Forms DBPs (see Table 12–15)
5. Oxidizes sulfides	5. Oxidizes iron, magnesium, and other inorganic compounds (consumes disinfectant)
6. Requires less space	6. Oxidizes a variety of organic compounds (consumes disinfectant)
7. Contributes dissolved oxygen	7. Off-gas requires treatment
8. At higher dosages than required for disinfection, ozone reduces the concentration of trace organic constituents	8. Safety concerns
	9. Highly corrosive and toxic
	10. Energy intensive
	11. Relatively expensive
	12. Highly operational and maintenance sensitive
	13. Has been shown to control the growth of filamentous microorganisms, but more expensive than chlorine
UV	
1. Effective disinfectant	1. No immediate measure of whether disinfection was successful
2. Requires no hazardous chemicals	2. No residual disinfectant
3. No residual toxicity	3. Less effective in inactivating some viruses, spores, and cysts at low dosages used for coliform organisms
4. More effective than chlorine in inactivating most viruses, spores, and cysts	4. Energy intensive
5. No formation of DBPs at dosages used for disinfection	5. Hydraulic design of UV system is critical
6. Does not increase TDS level of treated effluent	6. Capital cost is relatively expensive, but price is coming down as new and improved technology is brought to the market
7. At very high dosages, effective in the destruction of resistant organic constituents such as NDMA	7. Large number of UV lamps required where low-pressure, low-intensity systems are used
8. Improved safety as compared to use of chemical disinfectants	8. Acid washing to remove scale from quartz sleeves may be required for any technology
9. Requires less space than chlorine disinfection	9. Lacks a chemical system that can be adapted for auxiliary uses such as odor control, dosing RAS, and disinfecting plant water systems
10. At higher UV dosages than required for disinfection, UV radiation can be used to reduce the concentration of trace organic constituents of concern such as NDMA (see Sec. 10–8 in Chap. 10)	10. Fouling of UV lamps
	11. Lamps require routine periodic replacement
	12. Lamp disposal is problematic due to presence of mercury

[a] Adapted in part from Crites and Tchobanoglous (1998), U.S. EPA (1999b), and Hanzon et al. (2006).

[b] DBPs = disinfection byproducts.

treatment applications, the choice of disinfectant has usually between chlorine and UV. Recently, however, with concerns regarding trace constituents of concern, a renewed interest has developed in the use of ozone. Deciding factors in the selection of a disinfectant are commonly (1) economic evaluation, (2) public and operator safety, (3) environmental effects, and (4) ease of operation (Hanzon et al., 2006). Other treatment objectives are also important in the selection of a disinfectant. Potential concerns with pesticides, trace constituents of concern, endocrine disruptors, and similar compounds may influence the choice of disinfectants. Each disinfectant offers varying treatment performance with regard to these potential concerns.

12–3 DISINFECTION WITH CHLORINE

Chlorine, of all the chemical disinfectants, is the one used most commonly throughout the world. Specific topics considered in this section include a brief description of the characteristics of the various chlorine compounds, a review of chlorine chemistry and breakpoint chlorination, an analysis of the performance of chlorine as a disinfectant and the factors that may influence the effectiveness of the chlorination process, a discussion of the formation of disinfection byproducts (DBPs), and a consideration of the potential impacts of the discharge of DBPs to the environment. Disinfection with chlorine dioxide and dechlorination are considered in the following two sections, respectively. Chlorination facilities are considered in Sec. 12–6.

Characteristics of Chlorine Compounds

The principal chlorine compounds used at water reclamation plants are chlorine (Cl_2), sodium hypochlorite (NaOCl), and chlorine dioxide (ClO_2). Calcium hypochlorite [$Ca(OCl)_2$], another chlorine compound, is used in very small treatment plants due to its ease of operation. Many large cities have switched from chlorine gas to sodium hypochlorite because of the safety concerns and regulatory requirements related to the handling and storage of pressurized liquid chlorine (see Table 12–3). The characteristics of Cl_2, NaOCl, and $Ca(OCl)_2$ are considered below. The characteristics of chlorine dioxide and its use as a disinfectant are discussed in the following section.

Chlorine. The general properties of chlorine (Cl_2) are summarized in Table 12–7. Chlorine can be present as a gas or a pressurized liquid. Chlorine gas is greenish yellow in color and about 2.48 times as heavy as air. Liquid chlorine is amber colored and about 1.44 times as heavy as water. Unconfined liquid chlorine vaporizes rapidly to a gas at standard temperature and pressure with one liter of liquid yielding about 450 liters of gas. Chlorine is moderately soluble in water, with a maximum solubility of about 1 percent at 10°C (50°F).

Although the use of chlorine for the disinfection has been of great significance from a public health perspective in both potable water supplies and treated wastewater, serious concerns have been raised about its continued use. Important concerns include the following:

1. Chlorine is a highly toxic substance that is transported by rail and truck, both of which are prone to accidents.
2. Chlorine is a highly toxic substance that potentially poses health risks to treatment plant operators, and the general public, if released by accident.
3. Because chlorine is a highly toxic substance, stringent requirements for containment and neutralization must be implemented as specified in the Uniform Fire Code (UFC).

Table 12–7

Properties of chlorine, chlorine dioxide, and sulfur dioxide[a]

Property	Unit	Chlorine (Cl_2)	Chlorine dioxide (ClO_2)	Sulfur dioxide (SO_2)
Molecular weight	g	70.91	67.45	64.06
Boiling point (liquid)	°C	−33.97	11	−10
Melting point	°C	−100.98	−59	−72.7
Latent heat of vaporization	kJ/kg	253.6	27.28	376.0
Liquid density at 15.5°C	kg/m^3	1422.4	1640[b]	1396.8
Solubility in water at 15.5°C	g/L	7.0	70.0[b]	120
Specific gravity of liquid at 0°C (water = 1)	s.g.	1.468		1.486
Vapor density at 0°C and 1 atm	kg/m^3	3.213	2.4	2.927
Vapor density compared to dry air at 0°C and 1 atm	unitless	2.486	1.856	2.927
Specific volume of vapor at 0°C and 1 atm	m^3/kg	0.3112	0.417	0.342
Critical temperature	°C	143.9	153	157.0
Critical pressure	kPa	7811.8		7973.1

[a] Adapted in part from U.S. EPA (1986), White (1999).

[b] At 20°C.

4. Chlorine reacts with the organic constituents in wastewater to produce odorous compounds.
5. Chlorine reacts with the organic constituents in wastewater to produce byproducts, many of which are known to be carcinogenic and/or mutagenic.
6. Residual chlorine in treated effluent is toxic to aquatic life.
7. The discharge of chloro-organic compounds has long-term effects on the environment that are not known.

Sodium Hypochlorite. Sodium hypochlorite (NaOCl) (i.e., liquid bleach), is only available as an aqueous solution and usually contains 12.5 to 17 percent available chlorine at the time it is manufactured. Sodium hypochlorite can be purchased in bulk or manufactured onsite; however, the solution decomposes more readily at high concentrations and is affected by exposure to light and heat. A 16.7 percent solution stored at 26.7°C (80°F) will lose 10 percent of its strength in 10 d, 20 percent in 25 d, and 30 percent in 43 d. It must, therefore, be stored in a cool location in a corrosion resistant tank. Another disadvantage of sodium hypochlorite is the chemical cost. The purchase price may range from 150 to 200 percent of the cost of liquid chlorine. The handling of sodium hypochlorite requires special design considerations because of its corrosiveness, the presence of chlorine fumes, and gas binding and caking in chemical feed lines. Several proprietary systems are available for the generation of sodium hypochlorite from sodium chloride (NaCl) or seawater. These systems are electric power intensive and result in a very dilute solution, a maximum of 0.8 percent as chlorine. Onsite generation systems have been used only on a limited basis, typically at relatively large plants, due to their complexity and high power cost.

Calcium Hypochlorite. Calcium hypochlorite [$Ca(OCl)_2$], is available commercially in either a dry or a wet form. In dry form it is available as an off-white powder or as granules, compressed tablets, or pellets. Calcium hypochlorite granules or pellets are readily soluble in water, varying from about 21.5 g/100 mL at 0°C (32°F) to 23.4 g/100 mL at 40°C (104°F). Because of its oxidizing potential, calcium hypochlorite should be stored in a cool, dry location away from other chemicals in corrosion resistant containers. With proper storage conditions the granules are relatively stable. Calcium hypochlorite is more expensive than liquid chlorine, loses its available strength when stored, and because it must be dissolved before being used, it is difficult to handle for large installations. In addition, calcium hypochlorite can clog metering pumps, piping, and valves as it tends to crystallize readily. Calcium hypochlorite is used most commonly at small installations in a dry form as tablets, where handling is relatively easy for plant operators.

Chemistry of Chlorine Compounds

The reactions of chlorine in water and the reaction of chlorine with ammonia are presented below.

Chlorine Reactions in Water. When chlorine in the form of Cl_2 gas is added to water, two reactions take place: *hydrolysis* and *ionization*.

Hydrolysis may be defined as the reaction in which chlorine gas combines with water to form hypochlorous acid (HOCl).

$$Cl_2 + H_2O \rightarrow HOCl + H^+ + Cl^- \tag{12–9}$$

The equilibrium constant, K_H, for this reaction is

$$K_H = \frac{[HOCl][H^+][Cl^-]}{[Cl_2]} = 4.5 \times 10^{-4} \text{ at } 25°C \tag{12–10}$$

Because of the magnitude of the equilibrium constant, large quantities of chlorine can be dissolved in water.

Ionization of hypochlorous acid to hypochlorite ion (OCl^-) may be defined as

$$HOCl \rightleftarrows H^+ + OCl^- \tag{12–11}$$

The ionization constant, K_i, for this reaction is

$$K_i = \frac{[H^+][OCl^-]}{[HOCl]} = 3 \times 10^{-8} \text{ at } 25°C \tag{12–12}$$

The variation in the value of K_i with temperature is reported in Table 12–8.

The total quantity of HOCl and OCl^- present in water is called the *free chlorine*. The relative distribution of these two species (see Fig. 12–7) is very important because the killing efficiency of HOCl is many times that of OCl^-. The percentage distribution of HOCl at various temperatures can be computed using Eq. (12–13) and the data in Table 12–8.

$$\frac{[HOCl]}{[HOCl] + [OCl^-]} = \frac{1}{1 + [OCl^-]/[HOCl]} = \frac{1}{1 + [K_i]/[H^+]} = \frac{1}{1 + K_i 10^{pH}} \tag{12–13}$$

Hypochlorite Reactions in Water. Free chlorine can also be added to water in the form of hypochlorite salts. Both sodium and calcium hypochlorite hydrolyze to form hypochlorous acid (HOCl) as follows:

$$NaOCl + H_2O \rightarrow HOCl + NaOH \tag{12–14}$$

Table 12–8

Values of the ionization constant of hypochlorous acid at different temperatures[a]

Temperature, °C	$K_i \times 10^8$, mole/L
0	1.50
5	1.76
10	2.04
15	2.23
20	2.62
25	2.90
30	3.18

[a] Computed using equation from Morris (1966).

$$\mathrm{Ca(OCl)_2 + 2H_2O \rightarrow 2HOCl + Ca(OH)_2} \tag{12–15}$$

The ionization of hypochlorous acid was discussed previously [see Eq. (12–11)].

Chlorine Reactions with Ammonia. Untreated wastewater contains nitrogen in the form of ammonia, ammonium, and various combined organic forms (see Table 2–6 in Chap. 2). The effluent from most treatment plants also contains significant amounts of nitrogen, usually in the form of ammonia, ammonium, or nitrate, if the plant is designed to achieve nitrification. As noted in Chap. 2 the relative distribution between ammonia and ammonium will depend on the pH. Because hypochlorous acid is a very active oxidizing agent, it will react readily with ammonia (used here for the purpose of illustration) in water to form three types of chloramines in successive reactions:

$$\mathrm{NH_3 + HOCl \rightarrow NH_2Cl\ (monochloramine) + H_2O}\ \text{[see also Eq. (12–19)]} \tag{12–16}$$

$$\mathrm{NH_2Cl + HOCl \rightarrow NHCl_2\ (dichloramine) + H_2O} \tag{12–17}$$

$$\mathrm{NHCl_2 + HOCl \rightarrow NCl_3\ (nitrogen\ trichloride) + H_2O} \tag{12–18}$$

These reactions are very dependent on the pH, temperature, and contact time, and on the ratio of chlorine to ammonia (White, 1999). The two species that predominate, in most cases, are monochloramine (NH_2Cl) and dichloramine ($NHCl_2$). The ratio of dichloramine to monochloramine as a function of the ratio of chlorine to ammonia at various pH values is presented in Table 12–9. The amount of nitrogen trichloride present is negligible up to chlorine-to-nitrogen ratios of 2.0. As will be discussed subsequently, chloramines also

Figure 12–7

Distribution of hypochlorous (HOCl) acid and hypochlorite ion (OCl^-) in water as a function of pH at 0 and 20°C.

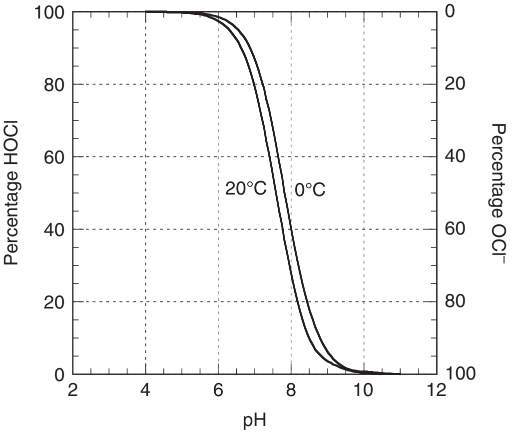


Table 12–9

Ratio of dichoramine to monochloramine under equilibrium conditions as a function of pH and applied molar dose ratio of chlorine to ammonium[a]

Molar Ratio $Cl_2:NH_4^+$	pH 6	pH 7	pH 8	pH 9
0.1	0.13	0.014	1E-03	0.000
0.3	0.389	0.053	5E-03	0.000
0.5	0.668	0.114	0.013	1E-03
0.7	0.992	0.213	0.029	3E-03
0.9	1.392	0.386	0.082	0.011
1.1	1.924	0.694	0.323	0.236
1.3	2.700	1.254	0.911	0.862
1.5	4.006	2.343	2.039	2.004
1.7	6.875	4.972	4.698	4.669
1.9	20.485	18.287	18.028	18.002

[a] From U.S. EPA (1986).

serve as disinfectants, although they are slow-reacting. When chloramines are the only disinfectants, the measured residual chlorine is defined as *combined chlorine residual* as opposed to free chlorine in the form of hypochlorous acid and hypochorite ion.

Breakpoint Reaction with Chlorine

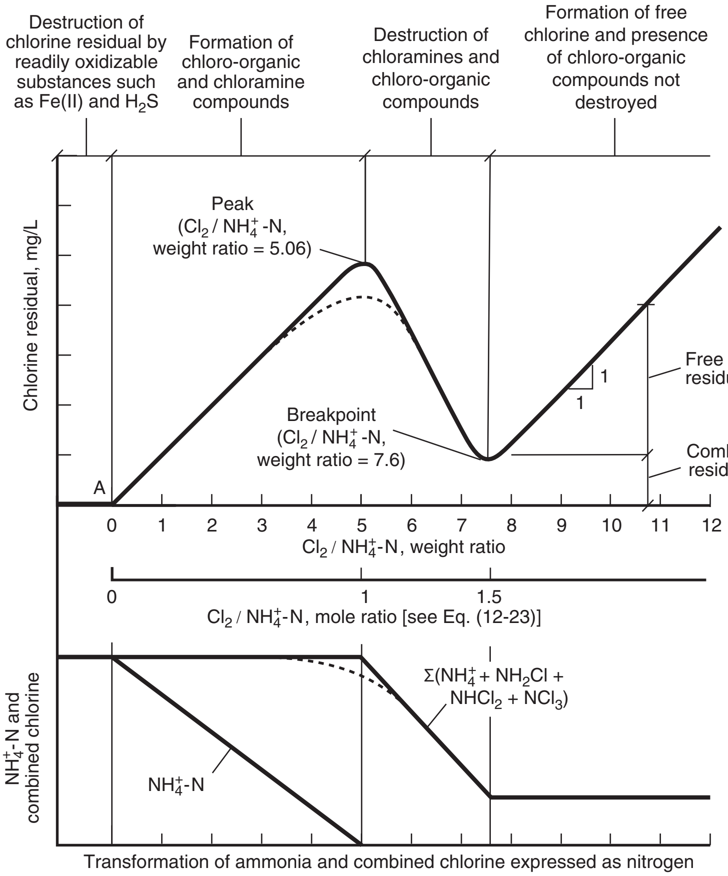
The maintenance of a residual (free or combined) for the purpose of disinfection is complicated because free chlorine not only reacts with ammonium, as noted previously, but also is a strong oxidizing agent. The term *breakpoint chlorination* is the term applied to the process whereby enough chlorine is added to react with all oxidizable substances such that if additional chlorine is added it will remain as free chlorine. The main reason for adding enough chlorine to obtain a free chlorine residual is that effective disinfection can usually then be assured. The amount of chlorine that must be added to reach a desired level of residual is called the *chlorine demand*. Breakpoint chlorination chemistry, acid generation, and the buildup of dissolved solids are considered in the following discussion.

Breakpoint Chlorination Chemistry. The stepwise phenomena that result when chlorine is added to water containing oxidizable substances and ammonium can be explained by referring to Fig. 12–8. As chlorine is added, readily oxidizable substances, such as Fe^{2+}, Mn^{2+}, H_2S, and organic matter, react with the chlorine and reduce most of it to the chloride ion (point A on Fig. 12–8). After meeting this immediate demand, the added chlorine continues to react with the ammonium to form chloramines between point A and the peak of the curve, as discussed above. For mole ratios of chlorine to ammonium less than 1, monochloramine and dichloramine are formed. At the peak of the curve, the mole ratio of chlorine (Cl_2) to ammonium (NH_4^+ as N) is equal to one [see Eq. (12–16)]. The corresponding Cl_2/NH_4^+ weight ratio is 5.06.

The distribution of the two chloramine forms is governed by their rates of formation, which are dependent on the pH and temperature. Between the peak and the breakpoint, some chloramines are converted to nitrogen trichloride [see Eq. (12–18)], the remaining chloramines are oxidized to nitrous oxide (N_2O) and nitrogen (N_2), and the chlorine is reduced to chloride ion. Most of the chloramines will be oxidized at the breakpoint. Continued addition of chlorine past the breakpoint, as shown on Fig. 12–8, will result in a directly proportional increase in the free chlorine. Theoretically, the weight ratio of

Figure 12–8

Generalized breakpoint chlorination curve. The upper portion of the diagram represents residual chlorine as a function of the amount of chlorine added to wastewater containing ammonium. The lower portion represents the fate of ammonium and chloramines during the breakpoint chlorination process. The dashed line reflects the fact that along with the formation of chloramines, some destruction of the chloramines occurs simultaneously before the peak is reached.



chlorine to ammonium nitrogen at the breakpoint is 7.6 to 1 (see Example 12–3) and the mole ratio is equal to 1.5 to 1 [see Eq. (12–23)].

Possible reactions to account for the appearance of N_2 and N_2O and the disappearance of chloramines during breakpoint chlorination are as follows (Saunier, 1976; Saunier and Selleck, 1976):

$$NH_4^+ + HOCl \rightarrow NH_2Cl + H_2O + H^+ \qquad (12\text{–}19)$$

$$NH_2Cl + HOCl \rightarrow NHCl_2 + H_2O \qquad (12\text{–}20)$$

$$NHCl_2 + H_2O \rightarrow NOH + 2HCl \qquad (12\text{–}21)$$

$$NHCl_2 + NOH \rightarrow N_2 + HOCl + HCl \qquad (12\text{–}22)$$

The overall reaction, obtained by summing Eqs. 12–19 through 12–22, is given as:

$$2NH_4^+ + 3HOCl \rightarrow N_2 + 3H_2O + 3HCl + 2H^+ \qquad (12\text{–}23)$$

Occasionally, serious odor problems have developed during breakpoint-chlorination operations because of the formation of nitrogen trichloride and related compounds. The presence of additional compounds that will react with chlorine, such as organic nitrogen, may greatly alter the shape of the breakpoint curve, as shown on Fig. 12–9. The formation of disinfection byproducts is considered later in this section.

Acid Generation. The addition of chlorine gas produces acid. When chlorine is added to water the hydrolysis reaction results in the formation of a strong acid (HCl) as given

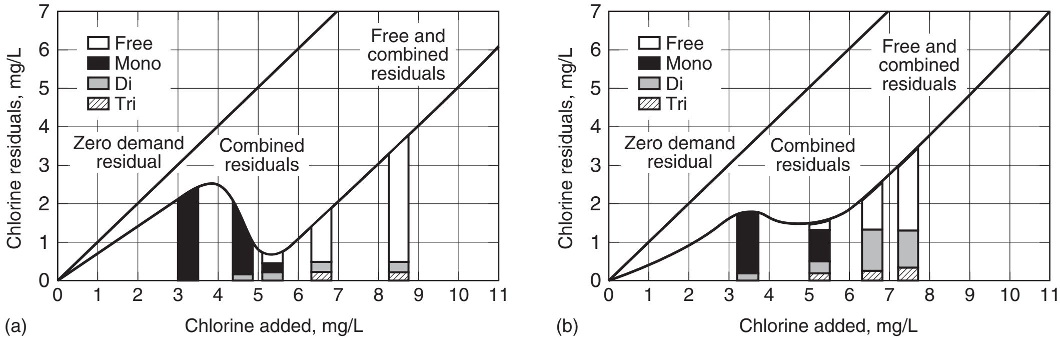


Figure 12–9

Curves of chlorine residual versus chlorine dosage for wastewater: (a) for wastewater containing ammonia nitrogen and (b) for wastewater containing nitrogen in the form of ammonia and organic nitrogen. (Adapted from White, 1999.)

by Eq. (12–9). The reaction of HOCl with ammonium also results in the formation of acid as given by Eq. (12–23). The total moles of hydrogen that must be neutralized can be determined by combining Eq. (12–9) with Eq. (12–23), which results in the following expression:

$$2NH_4^+ + 3Cl_2 \rightarrow N_2 + 6HCl + 2H^+ \qquad (12\text{–}24)$$

In practice, the hydrochloric acid formed during chlorination [see Eq. (12–16)] reacts with the alkalinity of the wastewater, and under most circumstances, there is a slight pH drop. Stoichiometrically, 14.3 mg/L of alkalinity, expressed as $CaCO_3$, are required for each 1.0 mg/L of ammonium nitrogen that is oxidized in the breakpoint-chlorination process (see Example 12–3).

Buildup of Total Dissolved Solids (TDS). In addition to the formation of hydrochloric acid, the chemicals added to achieve the breakpoint reaction also contribute an incremental increase in the TDS. As shown in Eq. (12–24), 6 moles of HCl and 2 moles of H^+ are formed, while 2 moles of NH_4^+ are removed from solution. In situations where the level of total dissolved solids may be critical with respect to water reuse applications, this incremental buildup from breakpoint chlorination should always be checked. The TDS contribution for each of several chemicals that may be used in the breakpoint reaction is summarized in Table 12–10. The magnitude of the possible buildup of TDS is illustrated

Table 12–10

Effects of chemical addition on total dissolved solids in breakpoint chlorination[a]

Chemical addition	Increase in total dissolved solids per unit of NH_4^+ consumed
Breakpoint with chlorine gas	6.2 : 1
Breakpoint with sodium hypochlorite	7.1 : 1
Breakpoint with chlorine gas—neutralization of all acidity with lime (CaO)	12.2 : 1
Breakpoint with chlorine gas—neutralization of all acidity with sodium hydroxide (NaOH)	14.8 : 1

[a] From U.S. EPA (1986).

in Example 12–3 in which the use of breakpoint chlorination is considered for the seasonal control of nitrogen.

EXAMPLE 12–3 **Analysis of Disinfection Process for Nitrified Secondary Effluent with Free Chlorine** Estimate the daily required chlorine dosage, the required alkalinity, if alkalinity needs to be added, and the resulting buildup of TDS when breakpoint chlorination is used achieve disinfection with free chlorine. Assume that the following data apply to this problem:

1. Plant flowrate = 3800 m^3/d
2. Secondary effluent characteristics
   a. BOD = 20 mg/L
   b. Total suspended solids = 25 mg/L
   c. Residual NH_3–N = 2 mg/L
   d. Alkalinity = 150 mg/L as $CaCO_3$
3. Required free chlorine residual concentration for disinfection = 0.5 mg/L
4. Any alkalinity added is in the form of lime (CaO)

Solution

1. Determine the molecular weight ratio of hypochlorous acid (HOCl), expressed as Cl_2, to ammonium (NH_4^+), expressed as N, using the overall reaction for the breakpoint reactions given by Eq. (12–23).

$$\underset{\substack{2(18)\\2(14)}}{2NH_4^+} + \underset{\substack{3(52.45)\\3(2\times35.45)}}{3HOCl} \rightarrow N_2 + 3H_2O + 3HCl + 2H^+$$

Molecular weight ratio is:

$$\frac{Cl_2}{NH_4^+\text{-N}} = \frac{3(2 \times 35.45)}{2(14)} = 7.60$$

2. Estimate the required Cl_2 dosage.
   a. Determine the Cl_2 dosage needed to reach the breakpoint using the molecular ratio developed in Step 1.

$$Cl_2 = (2\ g/m^3)(7.6\ g/g) = 15.2\ g/m^3$$

   b. Determine the required Cl_2 dosage including the free residual.

$$Cl_2/d = (3800\ m^3/d)[(15.2 + 0.5)\ g/m^3](1\ kg/10^3\ g) = 59.9\ kg/d$$

3. Determine the alkalinity required.
   a. The total number of moles of H^+ that must be neutralized per mole of NH_4^+ oxidized is given by Eq. (12–24), which has been divided by 2.

$$NH_4^+ + 1.5Cl_2 \rightarrow 0.5N_2 + 3HCl + H^+$$

   b. When using lime to neutralize the acidity, the required alkalinity ratio is computed as follows:

$$2CaO + 2H_2O \rightarrow 2Ca^{2+} + 4OH^-$$

$$\text{Required alkalinity ratio} = \frac{2(100\ \text{g/mole of } CaCO_3)}{(14\ \text{g/mole of } NH_4^+ \text{ as N})} = 14.3$$

c. The required alkalinity is

$$\text{Alk} = \frac{[(14.3\,\text{mg/L alk})/(\text{mg/L NH}_4^+)](2\,\text{mg/L NH}_4^+)(3800\,\text{m}^3/\text{d})}{(10^3\,\text{g/kg})}$$

$$= 108.7\ \text{mg/L as CaCO}_3$$

4. Determine whether sufficient alkalinity is available to neutralize the acid during breakpoint chlorination.
   Because the available alkalinity (150 mg/L) is greater than the required alkalinity (108.7 mg/L), alkalinity will not have to be added to complete the reaction.
5. Determine the increment of TDS added to the secondary effluent. Using the data reported in Table 12–10, the TDS increase per mg/L of ammonia consumed when CaO is used to neutralize the acid formed is equal to 12.2 to 1.

   TDS increment added = 12.2(2) mg/L = 24.4 mg/L

Comment The ratio computed in Step 1 will vary somewhat, depending on the actual reactions involved. In practice, the actual ratio typically has been found to vary from 8 : 1 to 10 : 1. Similarly, in Step 3, the stoichiometric coefficients will also depend on the actual reactions involved. In practice, it has been found that about 15 mg/L of alkalinity are required because of the hydrolysis of chlorine. In Step 5, it should be noted that although breakpoint chlorination can be used to control nitrogen, it may be counter productive if in the process the treated effluent is rendered unusable for other applications because of the buildup of total dissolved solids, and the potential formation of disinfection byproducts.

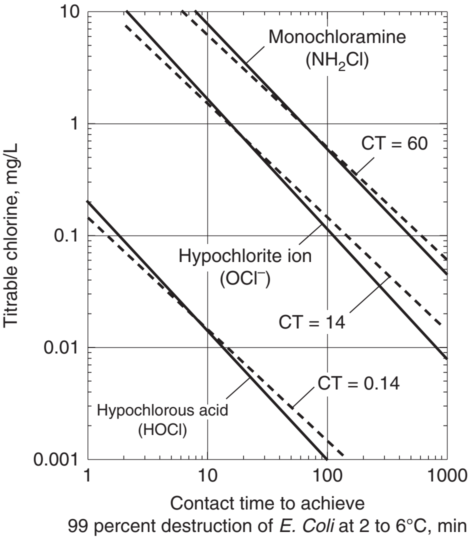
Effectiveness of Free and Combined Chlorine as Disinfectants

In view of the renewed interest in public health, environmental water quality, and water reclamation, the effectiveness of the chlorination process is of great concern. Numerous tests have shown that when all the physical parameters controlling the chlorination process are held constant, the germicidal efficiency of disinfection, as measured by the survival of "discrete bacteria," depends primarily on the form of the chlorine residual and time (i.e., CT).

Relative Effectiveness of Free and Combined Chlorine. Generalized data on the relative germicidal effectiveness of combined and free chlorine for the disinfection of different microorganisms were presented previously in Table 12–5 in terms of the required CT values to achieve various levels of inactivation. A comparison of the relative germicidal efficiency of hypochlorous acid (HOCl), hypochlorite ion (OCl^-), and monochloramine (NH_2Cl), based on the work of Butterfield et al. (1943), is presented on Fig. 12–10. As shown on Fig. 12–10, for a given contact time or chlorine residual, the germicidal efficiency of HOCl is 100 times that of OCl^- and more than 400 times that of NH_2Cl. However, because of the equilibrium relationship that exists between HOCl and OCl^- ion (see Fig. 12–7), maintenance of the proper pH is extremely important in achieving effective disinfection. It should be noted, however, that given an adequate contact time, monochloramine is nearly as effective as free chlorine in achieving disinfection. In addition to the data for the chlorine compounds given on Fig. 12–10, corresponding CT values have been added for the purpose of comparison. As shown, the

Figure 12–10

Comparison of the germicidal efficiency of hypochlorous (HOCl) acid and hypochlorite ion (OCl^-), and monochloramine for 99 percent destruction of *E. coli* at 2 to 6°C with CT values added for the purpose of comparison. (From Butterfield et al., 1943.)



disinfection data presented on Fig. 12–10 can be represented quite well with the CT relationship.

Coefficient of Specific Lethality. Another parameter that can be used to assess the relative effectiveness of the various forms of chorine is the coefficient of specific lethality, Λ. Utilizing the data from Table 12–5 and Fig. 12–10 as well as numerous literature sources, coefficients of specific lethality, computed for various microorganism groups and disinfecting agents, are summarized in Table 12–11. It is important to note that the data presented in Table 12–11 were derived primarily using batch reactors operated under

Table 12–11

Relative coefficients of specific lethality, Λ, for the inactivation of bacteria, viruses, *Cryptosporidium*, and *Giardia lamblia* cysts in filtered secondary effluent with various disinfectants (pH ~7.5, ~20°C)[a]

		Coefficient of specific lethality[b], $\Lambda_{(\text{base } 10)}$			
Disinfectant	**Unit**	**Total coliform[c]**	**Virus**	**Protozoa *Cryptosporidium***	**Protozoa *Giardia lamblia* cysts**
Chlorine (free)	L/mg·min	2	1.2	0.00044	0.04
Chloramine	L/mg·min	0.016	0.0052	0.00022	0.0024
Chlorine dioxide	L/mg·min	2	0.38	0.008	0.2
Ozone	L/mg·min	44	7.27	0.24	4.21
UV radiation	cm^2/mJ	5.7	0.0215	0.31	0.33

[a] Based on Eq. (12–6).

[b] The coefficient of specific lethality values are for the linear portion of the dose (CT) response curve.

[c] The reported coefficient of specific lethality values for fecal coliform and *E coli* are quit different.

Table 12–12

Activation energies for aqueous chlorine and chloramines at normal temperatures[a]

Compound	**pH**	***E*, Cal/mole**	***E*, J/mole**
Aqueous chlorine	8.5	6400	26,800
	9.8	12,000	50,250
	10.7	15,000	62,810
Chloramines	7.0	12,000	50,250
	8.5	14,000	58,630
	9.5	20,000	83,750

[a] Adapted from Fair et al. (1948) who developed the reported values using the data developed by Butterfield et al. (1943).

controlled conditions, and, as such, are of limited use other than for the purpose of illustrating the relative differences in the effectiveness of the different disinfectants for different organism groups. As shown, there are significant differences in the effectiveness of the various disinfectants for each organism group. For example, free chlorine is very effective for the inactivation of bacteria and viruses, but less so for *cryptosporidium* and *Giardia* cysts.

Effect of pH. The importance of pH and temperature on the disinfection process with chlorine and chloramines was investigated by Butterfield and his associates in 1943 (Butterfield et al., 1943; Wattie and Butterfield, 1944). Based on the results published by Butterfield et al. (1943), Fair and Geyer (1954) determined the activation energy values reported in Table 12–12 for the disinfection of *E. coli* in clean water. Reviewing the data in Table 12–12, it is important to note the magnitude of the activation energy as a function of pH. As the pH increases, the value of the activation energy increases which corresponds to a reduced effectiveness which is consistent with the data presented on Fig. 12–10.

Effect of Temperature. Temperature also has a significant impact on the coefficient of specific lethality, Λ. As a rule of thumb, it has been found that for each 10°C increase in temperature (identified as the Q_{10} temperature coefficient in biological and chemical engineering literature) there is a 2 to 2.5 times increase in the coefficient of specific lethality. Thus, when referring to the CT values given in Table 12–5 and the coefficient of specific lethality values in Table 12–11, it is important to note that the given values are for a pH and temperature of approximately 7.5 and 20°C, respectively.

Measurement and Reporting of Disinfection Process Performance

To provide a framework in which to consider the effectiveness of disinfection and the factors that affect the disinfection of treated wastewater, it is appropriate to consider how the effectiveness of the chlorination process is now assessed and how the results are analyzed. When using chlorine for the disinfection, the principal parameters that can be measured, apart from environmental variables such as pH and temperature, are the number of organisms and the form of the chlorine residual (i.e., combined or free or both) remaining after a specified period of time.

Number of Organisms Remaining. The coliform group of bacteria can be determined using a number of different techniques (Standard Methods, 2012). The membrane filter technique or the most probable number (MPN) procedure as discussed in

Chap. 2 are used commonly. The organisms remaining can also be determined by the plate-count procedure using an agar mixture as the plating medium. Either the standard "pour-plate" method or the "spread-plate" method can be used. The plates should be incubated at 37°C (98.6°F) because this temperature results in the optimum growth of *E. coli*, and the colonies should be counted after a 24-h incubation period.

Measurement of Chlorine Residual. The principal methods used to measure the free and combined chlorine residual include (1) the DPD (N,N-diethyl-p-phenylenediamine) colorimetric method, (2) DPD titration method, (3) the Iodometric titration method, and (4) the amperometric titration method. Of these methods, the DPD colorimetric method is currently the most widely used because it can be used to differentiate between free and combined chlorine species. Both field hand-held and continuous online residual analyzers are available. In the DPD method, appropriate chemicals, typically preformed in packets, are added to a sample containing chlorine. The red color resulting from the presence of chlorine is measured with a spectrophotometer or filter photometer. The initial color is due to free chlorine. Additional chemicals are added to obtain total residual chlorine (free and combined). Additional details on these chlorine analysis methods may be found in Harp (2002) and Standard Methods (2012).

Reporting of Results. Disinfection process results are reported in terms of the number of organisms and the chlorine residual remaining after a specified period of time. When the results are plotted it is common practice to plot the logs of removal versus the corresponding CT value as shown previously on Fig. 12–6.

Factors that Affect Disinfection of Wastewater with Chlorine Compounds

The purpose of the following discussion is to explore the important factors that affect the disinfection efficiency of chlorine compounds in actual wastewater applications. These include the following:

1. Initial mixing
2. Chemical characteristics of the water to be disinfected
3. NOM content
4. Impact of particles and particle associated microorganisms
5. Characteristics of the microorganisms
6. Contact time

Each of these factors are discussed in more detail below.

Issues related to the design of chlorine contact basins not included in this chapter include (1) basin configuration, (2) the use of baffles and guide vanes, (3) number of chlorine contact basins, (4) precipitation of solids in chlorine contact basins, (5) solids transport velocity, and (6) a procedure for predicting disinfection performance. These subjects are considered in detail elsewhere (Tchobanoglous et al., 2003).

Initial Mixing. The importance of initial mixing on the disinfection process cannot be overstressed. It has been shown that the application of chlorine in a highly turbulent regime ($N_R \geq 10^4$) results in kills two orders of magnitude greater than when chlorine is added separately to a conventional rapid-mix reactor under similar conditions. Although the importance of initial mixing is well delineated, the optimum level of turbulence is not known. Examples of mixing facilities designed to achieve the rapid mixing of chlorine with the water are presented later in Sec. 12–6 (see Fig. 12–22).

Based on recent findings, questions have now been raised about the form in which the chlorine compounds are added. In some plants where chlorine injectors are used, there is concern over the practice of using chlorinated wastewater for the chlorine injection water as opposed to clean water. The concern is that if nitrogenous compounds are present in the wastewater, a portion of the chlorine that is added reacts with these compounds, and by the time chlorine solution is injected, it is in the form of monochloramine or dichloramine. The formation of chloramines can be a problem if adequate retention time is not available in the chlorine contact basin as combined chlorine requires a longer contact time. Again, it should be remembered that although both HOCl and NH_2Cl are both effective as disinfecting compounds, the contact time required is significantly different for the same residual concentration (see Fig. 12–10).

The formation of disinfection byproducts (DBPs) is another major concern with the use of free chlorine. When wastewater is exposed to free chlorine, competing reactions, such as the formation of chloramines (free chlorine and ammonia), and DBPs can occur. The predominant reaction depends on the applicable kinetic rates for the various reactions. The formation and control of DBPs is discussed later in this section.

Chemical Characteristics of Wastewater. It has often been observed that, for treatment plants of similar design with exactly the same effluent characteristics measured in terms of BOD, COD, and nitrogen, the effectiveness of the chlorination process varies significantly from plant to plant. To investigate the reasons for this observed phenomenon and to assess the effects of the compounds present in the chlorination process, Sung (1974) studied the characteristics of the compounds in untreated and treated wastewater. Among the more important conclusions derived from Sung's study are the following:

1. In the presence of interfering organic compounds, the total chlorine residual cannot be used as a reliable measure for assessing the bactericidal efficiency of chlorine.
2. The degree of interference of the compounds studied depended on their functional groups and their chemical structure.
3. Saturated compounds and carbohydrates exert little or no chlorine demand and do not appear to interfere with the chlorination process.
4. Organic compounds with unsaturated bonds may exert an immediate chlorine demand, depending on their functional groups. In some cases, the resulting compounds may titrate as chlorine residual and yet may possess little or no disinfection potential.
5. Compounds with polycyclic rings containing hydroxyl groups and compounds containing sulfur groups react readily with chlorine to form compounds which have little or no bactericidal potential, but which still titrate as chlorine residual.
6. To achieve low bacterial counts in the presence of interfering organic compounds, additional chlorine and longer contact times are required.

From the results of Sung's work, it is easy to see why the efficiency of chlorination at plants with the same general effluent characteristics can be quite different. Clearly, it is not the value of the BOD or COD that is significant, but the nature of the organic compounds that make up the measured values. Thus, the nature of the treatment process used in any plant also has an effect on the chlorination process. The impact of wastewater characteristics on chlorine disinfection is presented in Table 12–13. The presence of oxidizable compounds such as humics and iron causes the inactivation curve to have a lag or shoulder affect as shown on Fig. 12–6. In effect, the added chlorine is being utilized in the oxidization of these substances and is not available for the inactivation of microorganisms.

Because more wastewater treatment plants are now removing nitrogen, operational problems with chlorine disinfection are now reported more frequently. In treatment plants

Table 12–13

Impact of wastewater constituents on the use of chlorine for wastewater disinfection

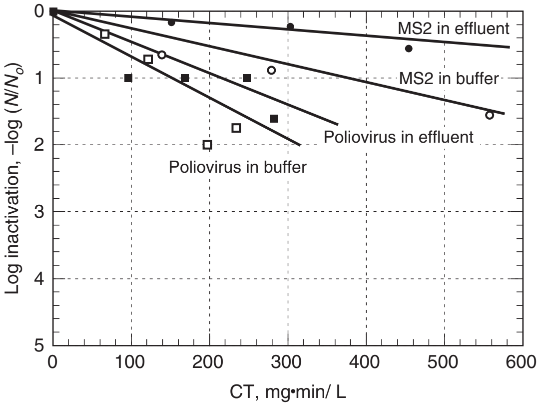
Constituent	Effect
BOD, COD, TOC, etc.	Organic compounds that comprise the BOD and COD can exert a chlorine demand. The degree of interference depends on their functional groups and their chemical structure
NOM (natural organic matter)	Reduces effectiveness of chlorine by forming chlorinated organic compounds that are measured as chlorine residual, but are not effective for disinfection
Oil and grease	Can exert a chlorine demand
TSS	Shields embedded bacteria
Alkalinity	No or minor effect
Hardness	No or minor effect
Ammonia	Combines with chlorine to form chloramines
Nitrite	Oxidized by chlorine, formation of N-nitrosodimethylamine (NDMA)
Nitrate	Chlorine dose is reduced because chloramines are not formed. Complete nitrification may lead to the formation of NDMA due the presence of free chlorine. Partial nitrification, especially diurnal swings in nitrification, may lead to difficulties in establishing the proper chlorine dose
Iron	Oxidized by chlorine
Manganese	Oxidized by chlorine
pH	Affects distribution between hypochlorous acid and hypochlorite ion
Industrial discharges	Depending on the constituents, may lead to a diurnal and seasonal variations in the chlorine demand

where the effluent is nitrified completely, the chlorine added to the water is present as free chlorine, after satisfying any immediate and nitrogenous (see Example 12–3) chlorine demand. In general, the presence of free chlorine will reduce significantly the required chlorine dosage. However, the presence of free chlorine may lead to the formation of disinfection byproducts including N-nitrosodimethylamine (NDMA). In treatment plants that do not nitrify completely, partially nitrify, or move in and out of nitrification diurnally, control of the chlorination process is especially difficult because of the variation in the effectiveness of the chlorine compounds. Some of the chlorine is used to satisfy the demand of the residual nitrite and/or ammonia. Because of the uncertainties involved in knowing to what degree the plant is nitrifying at any point in time, the chlorine dosage that is added is based on the dosage required if the disinfection is to be accomplished by combined chlorine compounds, resulting in excessive chlorine use.

Impact of Particles Found in Treated Wastewater. Another factor that must be considered is the presence of suspended solids in the water to be disinfected. As shown previously on Fig. 12–6, when suspended solids are present, the disinfection process is controlled by two different mechanisms. The log-linear bacterial inactivation that is observed initially, after the shoulder effect, is of individual free swimming bacteria and bacteria in small clumps. The straight line portion of the bacterial inactivation can be described using Eq. (12–2). In the curved portion of the curve the bacterial kill is controlled by the presence of suspended solids. The slope of the curved portion of the curve is a function of (1) the particle size distribution and (2) the number of particles with associated coliform organisms. Further, as noted previously, if particles contain

Figure 12–11

Inactivation of MS2 coliphage and polivirus in buffer and treated wastewater effluent with combined chlorine. (From BioVir Laboratories, 2001.)



significant numbers of organisms, the organisms can provide protection to other organisms embedded within the particle by limiting the penetration of chlorine through diffusion. Unfortunately, the observed variability caused by the presence of particles often is masked by the addition of excess chlorine to overcome both chemical and particle effects.

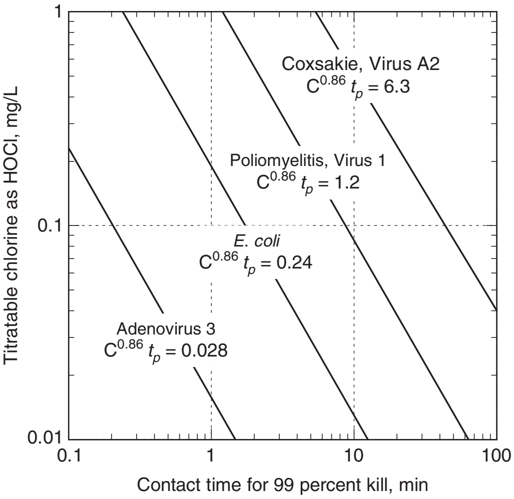
Characteristics of the Microorganisms. Other important variables in the chlorination process are the type, characteristics, and age of the microorganisms. For a young bacterial culture (1 d old or less) with a free chlorine dosage of 2 mg/L, only 1 min was needed to reach a low bacterial number. When the bacterial culture was 10 d old or more, approximately 30 min was required to achieve a comparable reduction for the same applied chlorine dosage. It is likely that the resistance offered by the polysaccharide sheath, which microorganisms develop as they age, accounts for this observation. In the activated sludge treatment process, the operating solids retention time (SRT), which to some extent is related to the age of the bacterial cells in the system, will, as discussed previously, affect the performance of the chlorination process. Some recent data on the disinfection of bacteriophage MS2 and poliovirus are shown on Fig. 12–11. As shown on Fig. 12–11, it is clear that a CT value of 450 mg·min/L, as used by the State of California, does not result in a four-log reduction of virus, when the measured residual chlorine is combined chlorine (i.e., mono- and dichloramine). Clearly, site-specific testing is required to establish the appropriate chlorine dose.

Some representative data on the effectiveness of chlorine for the inactivation of *E. coli* and three enteric viruses are reported on Fig. 12–12. Because of newer analytical techniques that have been developed, the data presented on Fig 12–12 are only meant to illustrate the differences in the resistances of different organisms. From the available evidence on the viricidal effectiveness of the chlorination process, it appears that chlorination beyond the breakpoint to obtain free chlorine is required to kill many of the viruses of concern. Where breakpoint chlorination is used, it is necessary to dechlorinate the treated water before discharge to the environment or reuse in sensitive applications to reduce any residual toxicity that may remain after chlorination. Based on the use of integrated cell culture-PCR techniques (see Chap. 2), it has been reported that the inactivation of poliovirus may require five times more chlorine than thought previously (Blackmer et al., 2000).

Contact Time. Along with the residual concentration of the disinfectant, contact time is of critical importance in the design and operation of chlorination facilities.

Figure 12–12

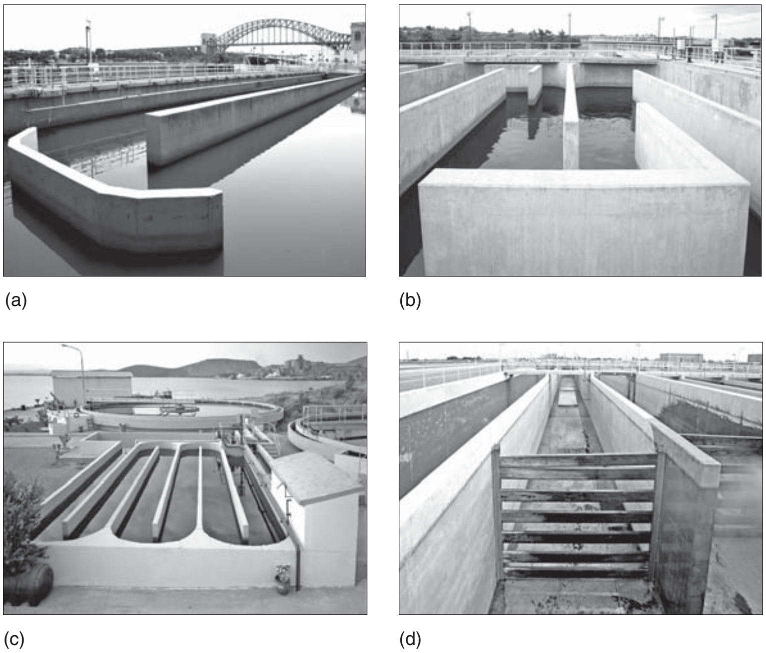
Concentration of chlorine as HOCl required for 99 percent kill of *E. coli* and three enteric viruses at 0 to 6°C (Butterfield et al., 1943).



The principal design objective for chlorine contact basins is to ensure that some defined percentage of the flow remains in the chlorine contact basin for the design contact time to ensure effective disinfection. The mean contact time is usually specified by the regulatory agency and may range from 30 to 120 min; contact times of 15 to 90 min at peak flow are common. To be assured that a given percentage of the flow remains in the chlorine contact basin for a given period of time, the most common approach is to use long plug-flow, around-the-end type of contact basins (see Fig. 12–13). For example, for water reuse applications the CDPH requires a CT value of 450 mg·min/L based on a modal contact time of 90 min at peak flow. In other states, the t_{10} is used in the

Figure 12–13

Views of chlorine contact basins: (a) and (b) serpentine plug-flow chlorine contact basins with flow deflection baffles, (c) plug-flow chlorine contact basin with rounded corners, and (d) plug-flow basin with inlet diffuser.


(a)
(b)
(c)
(d)

CT relationship (see subsequent discussion on assessing the performance of chlorine contact basins).

Modeling the Chlorine Disinfection Process

When considering the disinfection of both secondary and filtered secondary effluent, both the lag or shoulder effect and the effect of the residual particles (see Fig. 12–6) must be considered. As noted previously, depending on the constituents in the wastewater, a shoulder region may be observed in which there is no reduction in the number of organisms as the result of the addition of a disinfectant. As additional chlorine is added beyond some limiting value, a log linear reduction in the number of organisms is observed with increased chlorine dosages. If particles (typically greater than 20 μm) are present, the disinfection curve starts to diverge from the log linear form and a tailing region is observed due to particle shielding of the microorganisms. The tailing region is of importance as more restrictive standards are to be achieved (e.g., 23 MPN/100 mL). It is interesting that the tailing region was identified in an early report on the chlorination of treated wastewater (Enslow, 1938). Further, because large particles have little effect on turbidity (see Chap. 2), effluents with low measured turbidity values can still be difficult to disinfect, due the presence of undetected large particles (Ekster, 2001; see also discussion of turbidity in Chap. 8).

The Collins-Selleck Model. In the early 1970s, Collins conducted extensive experiments on the disinfection of various wastewaters (Collins, 1970; Collins and Selleck, 1972). Using the batch reactor whose contents were well stirred, Collins and Selleck found that the reduction of coliform organisms in a chlorinated primary treated effluent followed a linear relationship when plotted on log-log paper (see Fig. 12–14). The equation developed to describe the observed results is

$$\frac{N}{N_o} = \frac{1}{(1 + 0.23\,\mathrm{CT})^3} \tag{12–25}$$

Note that the form of the equation developed by Collins accounts for the shoulder effect and for tailing. A number of other models have been proposed including an empirical model proposed by Gard (1957), Hom (1972), which was subsequently rationalized by Haas and Joffe (1994) and Rennecker et al. (1999).

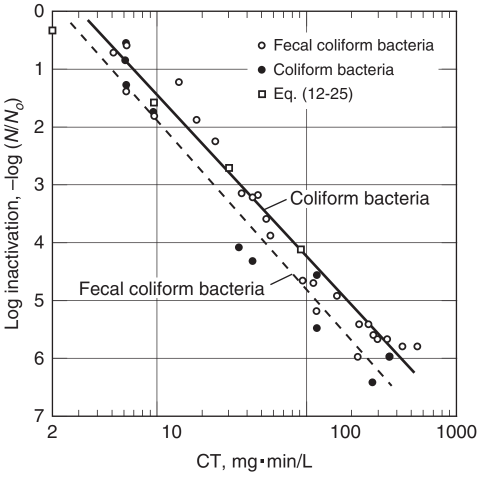
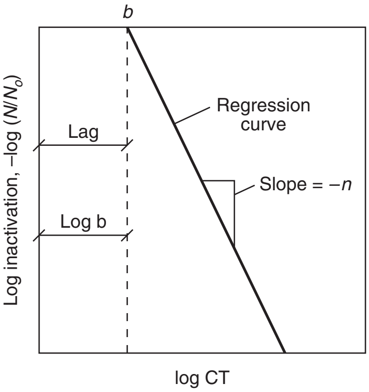


Figure 12–14

Coliform survival in a batch reactor as a function of amperometric chlorine residual and contact time (temperature range 11.5 to 18°C). (From Collins, 1970; Collins and Selleck, 1972.)

Figure 12–15

Definition sketch for the application of Eq. (12–27).



The Refined Collins-Selleck Model. A refinement of the original Collins model for the disinfection of secondary effluent in which a shoulder effect and tailing is observed, as proposed by White (1999), is

$$N/N_o = 1 \text{ for CT} < b \tag{12–26}$$

$$N/N_o = [(\text{CT})/b]^{-n} \text{ for CT} > b \tag{12–27}$$

Where C = residual concentration of chemical agent at the end of time t, mg/L
T = contact time, min
n = slope of inactivation curve
b = value of x-intercept when $N/N_o = 1$ or $\log (N/N_o) = 0$ (see Fig. 12–15)

Typical values for the coefficients n and b for secondary non-nitrified effluent for coliform and fecal coliform organisms are 2.8 and 4.0 and 2.8 and 3.0, respectively (Roberts et al., 1980; White, 1999; Black & Veatch Corporation, 2010). However, because of the variability of the chemical composition of wastewater and the variable particle size distribution, it is recommended that the constants be determined for the wastewater in question.

Effluent from Membrane Processes. The most important characteristic of these effluents is that they do not contain particles that can shield microorganisms. Depending on the type of membrane process used (microfiltration, ultrafiltration, nanofiltration, or reverse osmosis), moderate to significant reductions in the number of microorganisms present will also be observed (see discussion in Chaps. 8, 9, and 11). For these effluents, the Chick-Watson model, as given by Eq. (12–6), or, if a shoulder exists, the Collins-Selleck relationship, can be used to model the disinfection process with chlorine. Typically, the shoulder effect is reduced considerably, especially with reverse osmosis effluent.

Required Chorine Dosages for Disinfection

The required chemical dosage for disinfection can be estimated by considering (1) the initial chlorine demand of the treated wastewater, (2) the allowance needed for decay during the chlorine contact time, and (3) the required chlorine residual concentration determined using Eq. (12–27) for the organism under consideration (e.g., bacteria, virus, or protozoan oocysts and cysts).

Initial Chlorine Demand. The chlorine dosage required to meet the initial demand depends on the constituents in the water (see Fig. 12–15). It is important to remember that the chlorine added to meet the initial demand, due to inorganic compounds, is reduced to the chloride ion and will not be measured as chlorine residual. Also, chlorine that combines with humic and other organic material that may be present is not effective as a disinfectant, but is nevertheless measured as a chlorine residual contributing to the lag term, *b*, in Eq. (12–27).

Chlorine Decay. Two different cases of chlorine decay have to be considered: (1) chlorine decay, which occurs in chlorine contact basins and (2) chlorine decay, which occurs in long effluent transmission lines and in water reuse distribution systems.

In Chlorine Contact Basins. In chlorine contact basins the principal factors affecting chlorine decay are

1. Chemical reactions that occur in the bulk liquid.
2. Reactions that occur with the biofilm on the walls of the chlorine contact basin.
3. Photooxidation that occurs in uncovered chlorine contact basins.

In addition to the rapid reactions that occur initially, slower addition and substitution type reactions can occur with the residual organic matter in the bulk of the treated effluent as it moves through the chlorine contact basin (Gang et al., 2003). Similarly, addition and substitution chemical reactions can occur with the biofilms attached to the wall of the chlorine contact basin. The nature of the reactions is site specific, but biofilms are almost always present, especially in uncovered chlorine contact basins. To reduce the decay due to UV oxidation observed in open uncovered chlorine contact basins, a variety of floating and fixed covers have been added to existing contact basins (see Fig. 12–16). Typical decay values for chlorine residual are on the order of 2 to 4 mg/L for a contact time of about one hour.

In Transmission and Distribution Piping. The principal factors affecting chlorine decay in transmission and distribution piping are

1. Chemical reactions that occur in the bulk liquid flowing within the pipe.
2. Reactions that occur with the biofilm on the walls of the piping system.
3. Chemical reactions due to the release of constituents under anaerobic conditions that will react with chlorine.

The reactions that occur in the bulk fluid and in the biofilm on the pipe wall are as described above for the chlorine contact basin. Under anaerobic conditions that often occur in long transmission lines, bacteria on the pipe walls will convert the sulfate present in the treated effluent to form sulfide, which in turn forms hydrogen sulfide, which readily reacts with any chlorine that may be present. There are many articles in the literature that deal with the modeling of chlorine decay in pipelines. In general, the decay process is modeled

(a) (b)

Figure 12–16

Typical examples of covered back-and-forth chlorine contact basins: (a) basins covered with inexpensive floating tarp, and (b) basins covered with specially designed polypropylene cover.

Table 12–14

Typical chlorine dosages, based on combined chlorine unless otherwise indicated, required to achieve different total coliform disinfection standards for various wastewaters based on a 30 min contact time with a decay factor of 0.6[a]

Type of wastewater	Initial total coliform count, MPN/100 mL	Chlorine dose, mg/L			
		Effluent standard, MPN/100 mL			
		1000	200	23	≤2.2
Raw wastewater	10^7–10^9	16–30			
Primary effluent	10^7–10^9	8–12	18–24		
Trickling filter effluent	10^5–10^6	6–7.5	12–15	18–22	
Activated sludge effluent	10^5–10^6	5.5–7.5	10–13	13–17	
Filtered activated sludge effluent	10^4–10^6	2.5–3.5	5.5–7.5	10–13	13–17
Nitrified effluent[b]	10^4–10^6		0.02–0.03	0.03–0.04	0.04–0.05
Filtered nitified effluent[b]	10^4–10^6		0.02–0.03	0.03–0.04	0.04–0.05
Microfiltration effluent[b]	10^1–10^3			0.02–0.03	0.03–0.04
Reverse osmosis[b]	~ 0				0.01–0.02
Septic tank effluent	10^7–10^9	16–30	30–60		
Intermittent sand filter effluent	10^2–10^4	1–2	2–4	3–6	4–8

[a] The combined chlorine values are based on the assumption that the added chlorine only combines with ammonia to form monochloramine. The reported values are independent of the chlorine dose required to meet the immediate chlorine demand.

[b] Based on free chlorine. The reported values are independent of the chlorine dose required to reach the breakpoint (see Example 12–3).

using either first or second-order decay models. The computer model EPANET 2, developed by the U.S. EPA to simulate the hydraulic and water quality behavior within pressurized pipe networks, has also been used to study the decay of chlorine in pipelines (Rossman, 2000). It is critical that decay be considered in long transmission and distribution piping in determining the required residual that may be needed.

Required Chlorine Residual. Typical chlorine dosage values for various for residual total coliform concentrations, based on a contact time of 30 min, are reported in Table 12–14. It should be noted that the dosage values given in Table 12–14 are only meant to serve as a guide for the initial estimation of the required chlorine dose. As noted above, site-specific testing is required to establish the appropriate chlorine dose. Estimation of the required chlorine dose is illustrated in Example 12–4.

EXAMPLE 12–4 **Estimate the Required Chlorine Dose for a Typical Non-nitrified Secondary Effluent** Estimate the chlorine dose needed to disinfect a filtered non-nitrified secondary activated sludge effluent using the refined Collins-Selleck model assuming a shoulder effect exists and that the following conditions apply. Check the computed summer combined residual using Eq. (12–6).

1. Effluent total coliform count before disinfection = 10^7/100 mL
2. Required summer effluent total coliform count = 23/100 mL
3. Required winter effluent total coliform count = 240/100 mL
4. Immediate summer or winter effluent chlorine demand, not including the shoulder effect = 2 mg/L
5. Chlorine demand due to decay in chlorine contact tanks during the summer months (May–October) = 2.5 mg/L
6. Chlorine demand due to decay in chlorine contact tanks during the winter months (November–April) = 1.5 mg/L
7. Required chlorine contact time = 30 min
8. Use the typical values given in the above discussion for the coefficients $n = 2.8$ and $b = 4.0$
9. The coefficient of specific lethality for summer conditions = 0.024 L/mg·min (base 10)

Solution

1. Estimate the required combined chlorine residual using the refined Collins-Selleck Model, Eq. (12–27) and the given coefficients.

$$N/N_o = (\mathrm{CT}/b)^{-n}$$

   a. Summer

$$23/10^6 = (\mathrm{CT}/4.0)^{-2.8}$$

$$(23/10^6)^{-\frac{1}{2.8}} = (\mathrm{CT}/4.0)$$

$$(45.3)4 = \mathrm{C}(30)$$

$$\mathrm{C} = 6.0 \text{ mg/L}$$

   b. Winter

$$240/10^6 = (\mathrm{CT}/4.0)^{-2.8}$$

$$(19.6)4 = \mathrm{C}(30)$$

$$\mathrm{C} = 2.6 \text{ mg/L}$$

2. The required chlorine dosage is

   a. Summer

   Chlorine dosage = 2.0 mg/L + 2.5 mg/L + 6.0 mg/L = 10.5 mg/L

   b. Winter

   Chlorine dosage = 2.0 mg/L + 1.5 mg/L + 2.6 mg/L = 6.1 mg/L

3. Determine the required chlorine summer dose using Eq. (12–6).

   a. Solve Eq. (12–6) for the combined residual.

$$\log \frac{N_t}{N_o} = -\Lambda_{\text{base }10}\mathrm{CT} = \log \frac{23}{10^6} = (-0.024)(\mathrm{C})(30)$$

$$\mathrm{C} = \frac{-4.64}{-(0.024\,\mathrm{L/mg \cdot min})(30\,\mathrm{min})} = 6.4 \text{ mg/L}$$

   b. The required chlorine doses computed with the two methods are similar.

Comment The chlorine dosage increases significantly as the effluent standards become more stringent. In the above computation, it was assumed that the filtered effluent to be disinfected remained in the chlorine contact tank for the full 30 minutes. Thus, it is clear that the proper design of a plug-flow chlorine contact basin is critical to the effective use of chlorine as disinfectant. The design of chlorine contact basins is considered in Sec. 12–6.

Formation and Control of Disinfection Byproducts (DBPs)

In the early 1970s, it was found that the use of oxidants, such as chlorine and ozone, in water treatment for disinfection; for the control of tastes, odors, and color removal; and other in-plant uses resulted in the production of undesirable DBPs (Rook, 1974; Bellar and Lichtenberg, 1974). The DBPs measured most frequently and with the highest concentration are trihalomethanes (THMs) and haloacetic acids (HAAs), resulting from chlorination. In addition to trihalomethanes and haloacetic acids, a variety of other DBPs are also produced. The principal DBPs that have been identified are reported in Table 12–15. Many of the compounds identified in Table 12–15 have also been identified in treated effluent that has been disinfected using chlorine, chloramines, chlorine dioxide, and ozone.

Concerns with DBPs. Formation of DBPs is of great concern in effluent dispersal to the environment and for indirect and direct potable reuse because of the potential long-term (chronic) impact of these compounds on public health and the environment. Chloroform, for example, is a well-known animal carcinogen and many of the haloforms are also thought to be animal carcinogens. In addition, many of these compounds have been classified as probable human carcinogens. Still others of these compounds are known to cause chromosomal aberrations and sperm abnormalities. Recognizing the many unknowns and the potential public health and environmental risks associated with these compounds, the U.S. EPA has moved aggressively to control their formation in drinking water.

Formation of DBPs Using Chlorine for Disinfection. Trihalomethanes (THMs) and other DBPs are formed as a result of a series of complex reactions between free chlorine and a group of organic acids known collectively as humic acids. The reactions lead to the formation of single carbon molecules that are often designated as CHX_3, where X is either a chlorine (Cl^-) or bromine (Br^-) atom. For example, the chemical formula for chloroform is $CHCl_3$.

The rate of formation of DBPs is dependent on a number of factors, including

1. Presence of organic precursors
2. Free chlorine concentration
3. Bromide concentration
4. pH
5. Temperature
6. Time

The type and concentration of the organic precursor affects both the rate of the reaction and extent to which the reaction is completed.

The presence of free chlorine was thought to be necessary for the THM formation reaction to proceed, but it appears that THMs can form in the presence of combined chlorine (chloramines), but at a much reduced rate. It is important to note that initial mixing can affect the formation of THMs because of the competing reactions between chlorine and ammonia,

Table 12–15

Selected, known disinfection byproducts formed during application of chlorine, chloramine, ozone, and chlorine dioxide in natural waters[a]

Class	Byproduct	Chemical agent	Molecular formula
Trihalomethanes	Chloroform	Chlorine	$CHCl_3$
	Bromodichloromethane	Chlorine	$CHBrCl_2$
	Dibromochloromethane	Chlorine	$CHBr_2Cl$
	Bromoform	Chlorine, ozone	$CHBr_3$
	Dichloroiodomethane	Chlorine	$CHICl_2$
	Chlorodiiodomethane	Chlorine	CHI_2Cl
	Bromochloroiodomethane	Chlorine	$CHBrICl$
	Dibromoiodomethane	Chlorine	$CHBr_2I$
	Bromodiiodomethane	Chlorine	$CHBrI_2$
	Triiodomethane	Chlorine	CHI_3
Haloacetic acids	Monochloroacetic acid	Chlorine	$CH_2ClCOOH$
	Dichloroacetic acid	Chlorine	$CHCl_2COOH$
	Trichloroacetic acid	Chlorine	CCl_3COOH
	Bromochloroacetic acid	Chlorine	$CHBrClCOOH$
	Bromodichloroacetic acid	Chlorine	$CBrCl_2COOH$
	Dibromochloroacetic acid	Chlorine	$CBr_2ClCOOH$
	Monobromoacetic acid	Chlorine	$CH_2BrCOOH$
	Dibromoacetic acid	Chlorine	$CHBr_2COOH$
	Tribromoacetic acid	Chlorine	CBr_3COOH
Haloacetonitriles	Trichloroacetonitrile	Chlorine	$CCl_3C{\equiv}N$
	Dichloroacetonitrile	Chlorine	$CHCl_2C{\equiv}N$
	Bromochloroacetonitrile	Chlorine	$CHBrClC{\equiv}N$
	Dibromoacetonitrile	Chlorine	$CHBr_2C{\equiv}N$
Haloketones	1,1-Dichloroacetone	Chlorine	$CHCl_2COCH_3$
	1,1,1-Trichloroacetone	Chlorine	CCl_3COCH_3
Aldehydes	Formaldehyde	Ozone, chlorine	$HCHO$
	Acetaldehyde	Ozone, chlorine	CH_3CHO
	Glyoxal	Ozone, chlorine	$OHCCHO$
	Methyl glyoxal	Ozone, chlorine	CH_3COCHO
Aldoketoacids	Glyoxylic acid	Ozone	$OHCCOOH$
	Pyruvic acid	Ozone	$CH_3COCOOH$
	Ketomalonic acid	Ozone	$HOOCCOCOOH$
Carboxylic acids	Formate	Ozone	$HCOO^-$
	Acetate	Ozone	CH_3COO^-
	Oxalate	Ozone	$OOCCOO^{2-}$

(*continued*)

Table 12–15 (*Continued*)

Class	Byproduct	Chemical agent	Molecular formula
Oxyhalides	Chlorite	Chlorine dioxide	ClO_2^-
	Chlorate	Chlorine dioxide	ClO_3^-
	Bromate	Ozone	BrO_3^-
Nitrosamines	N-nitrosodimethylamine	Chloramines	$(CH_3)_2NNO$
Cyanogen Halides	Cyanogen chloride	Chloramines	ClCN
	Cyanogen bromide	Chloramines	BrCN
Misc.	Chloral hydrate	Chlorine	$CCl_3CH(OH)_2$
Trihalonitromethanes	Trichloronitromethane (Chloropicrin)	Chlorine	CCl_3NO_2
	Bromodichloronitromethane	Chlorine	$CBrCl_2NO_2$
	Dibromochloronitromethane	Chlorine	CBr_2ClNO_2
	Tribromonitromethane	Chlorine	CBr_3NO_2

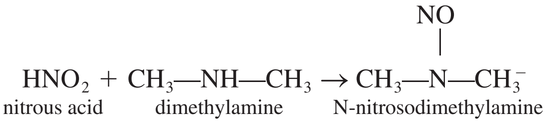
[a] Adapted from Krasner (1999), Krasner et al. (2001), and Thibaud et al. (1987).

and between chlorine and humic acids. If bromide is present, it can be oxidized to bromine by free chlorine. In turn the bromine ion can combine with the organic precursors to form THMs, including bromodichoromethane, dibromochoromethane, and bromoform. The rate of formation of THMs has been observed to increase with both pH and temperature. Additional details on the formation of THMs may be found in U.S. EPA (1999a).

Although chloramines, as discussed above, produce THMs at reduced rates, they can, nevertheless, produce other DBP compounds that are of concern. Other DBPs that are produced when treated wastewater is disinfected with chloramines include N-nitrosodimethylamine (NDMA), a member of a class of compounds known as nitrosoamines, cyanogen chloride, and cyanogen bromide (see Table 12–15). As a class of compounds nitrosoamines are among the most powerful carcinogens known (Snyder, 1995). The compounds in this class have been found to produce cancer in every species of laboratory animal tested.

One pathway leading to the formation of NDMA can be illustrated with the following two reactions:

$$\underset{\substack{\text{nitrite}\\\text{anion}}}{NO_2^-} + \underset{\substack{\text{hydrochloric}\\\text{acid}}}{HCl} \rightarrow \underset{\substack{\text{nitrous}\\\text{acid}}}{HNO_2} + \underset{\substack{\text{chloride}\\\text{ion}}}{Cl^-} \qquad (12\text{–}28)$$



$$\underset{\text{nitrous acid}}{HNO_2} + \underset{\text{dimethylamine}}{CH_3\text{—}NH\text{—}CH_3} \rightarrow \underset{\text{N-nitrosodimethylamine}}{CH_3\text{—}\overset{\overset{NO}{|}}{N}\text{—}CH_3^-} \qquad (12\text{–}29)$$

The concern in biological wastewater treatment facilities is that some nitrite may leak through the process. While the concentration of nitrite may be too low to measure by conventional means, concentrations of NDMA as low as 1 or 2 ng/L are being measured and the CDPH notification level for groundwater recharge is 10 ng/L. Based on a limited number of test locations, it has been observed that the concentrations of NDMA in the incoming wastewater can be quite variable, with concentrations as high as 6000 ng/L being measured.

In addition to the formation of NDMA as outlined above, it appears the addition of chloramines for disinfection can serve to amplify the concentration of any NDMA that may be present in the treated effluent before disinfection. In a series of studies conducted by the Los Angeles County Sanitation Districts (Jalali et al., 2005), it was found that chloramination increased the concentration of NDMA in treated effluent following disinfection by tenfold.

Other DBPs resulting from the use of chloramines as disinfectants in treated effluents include cyanogen chloride and cyanogen bromide, where bromides are present (see Table 12–15). In very large quantities, cyanogen chloride is used in tear gas, in fumigant gases, and as a reagent in the formation of other compounds. In the body, cyanogen chloride is metabolized rapidly to cyanide. Because there is limited information on the toxicity of low-level concentrations of cyanogen chloride, proposed guidelines are based on cyanide. The cyanogen compounds are of concern and they are now beginning to be regulated in effluent discharge permits. The current NPDES permit limit for cyanide is 5 mg/L.

Control of DBP Formation Using Chlorine for Disinfection. The principal means of controlling the formation of THMs and other related DBPs is to avoid the direct addition of free chlorine. Based on the evidence to date, it appears that the use of chloramines generally does not lead to the formation of THMs in amounts that would be of concern relative to current standards. As discussed previously, other DBPs may be produced that are of equal concern, but for other reasons (see following discussion). It is important to note that if chloramines are to be used for disinfection, the chloramine solution must be prepared with a potable water supply containing little or no ammonia (i.e., *treated plant effluent should not be used*). If the formation of DBPs is of concern due to the presence of specific organic precursors (i.e., humic materials), the practice of breakpoint chlorination cannot be used. Further, if humic materials are present consistently, it may be appropriate to investigate alternative means of disinfection such as UV irradiation.

The control of DBPs produced when chloramines are used (by reducing direct reactions of organics with residual free chlorine) can be more challenging as chloramination may form other DBPs. With respect to NDMA it appears that with proper control and operation of the biological treatment process, the potential for the formation or amplification of this compound can be reduced. Removals of 50 to 70 percent have been reported for NDMA when using reverse osmosis employing thin film composite membranes (see Chap. 11). The use of UV irradiation has also proven to be effective in the control of NDMA. Where the formation of NDMA and cyanogen chloride is a persistent concern, a number of wastewater agencies have switched to UV irradiation for disinfection. In the study cited above (Jalali et al., 2005), it was also found that there was no net change in the total cyanide (CN^-) concentration in the treated effluent due to UV irradiation. The use of sequential chlorination to control the formation of NDMA is considered in Sec. 12–8.

Environmental Impacts of Disinfection with Chlorine

The environmental impacts associated with the use of chlorine and chlorine compounds as a disinfectant in wastewater applications include the discharge of DBPs and the regrowth of microorganisms.

Discharge of DBPs. It has been shown that many of the DBPs can cause environmental impacts at very low concentrations. The occurrence of DBPs and compounds such as NDMA raises serious questions about the continued use of free chlorine for disinfection.

Regrowth of Microorganisms. In many locations, a regrowth of microorganisms has been observed in receiving water bodies and in long transmission pipelines following dechlorination of treated effluent disinfected with chlorine. The regrowth of microorganisms is not unexpected as it is well known that a number of microorganisms survive the disinfection process. It has been hypothesized that regrowth (also known as aftergrowth) results, in part, because (1) the amount of organic matter and available nutrients in treated wastewater is sufficient to sustain the limited number of organisms remaining after disinfection, (2) predators such as protozoa are absent, (3) there are favorable temperatures, and (4) disinfectant residuals are ineffective. Because regrowth is an especially important issue in transmission lines used for the transport of reclaimed water, a suitable chlorine residual (on the order of 1 to 2 mg/L, depending on local conditions) should be maintained in the pipeline to control regrowth (a common practice in water distribution systems). In very long pipelines, it may be necessary to add additional chlorine at intermediate points along the length of the pipeline.

12–4 DISINFECTION WITH CHLORINE DIOXIDE

Chloride dioxide (ClO_2), another bactericide, is equal to or greater than chlorine in disinfecting power. Chlorine dioxide has proven to be an effective virucide, being more effective in achieving inactivation of viruses than chlorine. A possible explanation is that because chlorine dioxide is absorbed by peptone (a protein), and that viruses have a protein coat, adsorption of ClO_2 onto this coating could cause inactivation of the virus. In the past, ClO_2 did not receive much consideration as a wastewater disinfectant due to its high costs; sodium chlorite feed stock is about ten times as expensive as chlorine on a weight basis.

Characteristics of Chlorine Dioxide

Chlorine dioxide (ClO_2) is, under atmospheric conditions, a yellow to red unpleasant smelling, irritating, unstable gas with a high specific gravity. Because chlorine dioxide is unstable and decomposes rapidly, it is usually generated onsite before its application. Chlorine dioxide is generated by mixing and reacting a chlorine solution in water with a solution of sodium chlorite ($NaClO_2$) according to the following reaction:

$$2NaClO_2 + Cl_2 \rightarrow 2ClO_2 + 2NaCl \tag{12–30}$$

Based on Eq. (12–30), 1.34 mg sodium chlorite reacts with 0.5 mg chlorine to yield 1.0 mg chlorine dioxide. Because technical grade sodium chlorite is only about 80 percent pure, about 1.68 mg of the technical grade sodium chlorite is required to produce 1.0 mg of chlorine dioxide. Sodium chlorite may be purchased and stored as a liquid (generally a 25 percent solution) in refrigerated storage facilities. The properties of chlorine dioxide were presented previously in Tables 12–3 and 12–7.

Chlorine Dioxide Chemistry

The active disinfecting agent in a chlorine dioxide system is free dissolved chlorine dioxide (ClO_2). At the present time, the complete chemistry of chlorine dioxide in an aqueous environment is not understood completely. Because ClO_2 does not hydrolyze in a manner similar to the chlorine compounds discussed in the previous section, the oxidizing power of ClO_2 is often referred to as *equivalent available chlorine*. The definition of the term equivalent available chlorine is based on a consideration of the following oxidation half reaction for ClO_2:

$$ClO_2 + 5e^- + 4H^+ \rightarrow Cl^- + 2H_2O \tag{12–31}$$

As shown in Eq. (12–31), the chlorine atom undergoes a 5 electron change in its conversion from chlorine dioxide to the chloride ion. Because the weight of chlorine in ClO_2 is 52.6 percent and there is a 5 election change, the equivalent available chlorine content is equal to 263 percent as compared to chlorine. Thus, ClO_2 has 2.63 times the oxidizing power of chlorine. The concentration of ClO_2 is usually expressed in g/m^3. On a molar basis, one mole of ClO_2 is equal to 67.45 g, which is equivalent to 177.25 g (5×35.45) of chlorine. Thus, 1 g/m^3 of ClO_2 is equivalent to 2.63 g/m^3 of chlorine.

Effectiveness of Chlorine Dioxide as a Disinfectant

Chlorine dioxide has an extremely high oxidation potential, which probably accounts for its potent germicidal powers. Because of its extremely high oxidizing potential, possible bactericidal mechanisms may include inactivation of critical enzyme systems or disruption of protein synthesis. It should be noted, however, that when ClO_2 is added to water it is often reduced to the chlorite ion (ClO_2^-), a weak disinfectant, according to the following reaction. The formation of ClO_2^- may help to explain the variability that is sometimes observed in the performance of ClO_2 as a disinfectant.

$$ClO_2 + e^- \rightarrow ClO_2^- \tag{12–32}$$

Based on the coefficient of specific lethality as reported in Table 12–11, the effectiveness of ClO_2 with respect to bacteria is similar to that of free chlorine. However, there are some differences, depending on the microorganism group and members within each group. Chlorine dioxide appears to be more effective than free chlorine in the inactivation of protozoan cysts.

Modeling the Chlorine Dioxide Disinfection Process

As discussed previously in Sec. 12–3, the models that have been developed to describe the disinfection process with chlorine can also be used, with appropriate caution, for chlorine dioxide. As with chlorine, the shoulder effect and the effect of the residual particles must be considered. Further, the differences between (1) secondary and filtered secondary effluent and (2) microfiltration and reverse osmosis effluent must also be considered.

Required Chlorine Dioxide Dosages for Disinfection

The required chlorine dioxide dosage will depend on the pH and the specific organism under investigation. Relative CT values for chlorine dioxide are given in Table 12–5, presented previously in Sec. 12–2, and values of the coefficient of specific lethality are given in Table 12–11. Because the data on chlorine dioxide in the literature are limited, site-specific testing is recommended to establish appropriate dosage ranges although the values given in Table 12–5 can be used as a starting point.

Byproduct Formation and Control

The formation of DBPs is of great concern with the use of chlorine dioxide. The formation and control of DBPs with chlorine dioxide is considered in the following discussion.

Formation of DBPs Using Chlorine Dioxide for Disinfection. The principal DBPs formed when chlorine dioxide is used as a disinfectant are chlorite (ClO_2^-) and chlorate (ClO_3^-), both of which are potentially toxic at low concentrations. The principal sources of the chlorite ion are from the process used to generate the chlorine dioxide and from the reduction of chlorine dioxide. As given by Eq. (12–30), all of the $NaClO_2$ reacts with chlorine to form ClO_2. Unfortunately, on occasion some unreacted chlorite ion can escape from the reactor where the chlorine dioxide is being generated and find its way into the water that is being treated. The second source of chlorite is from the reduction of chlorine dioxide as discussed above [see Eq. (12–32)]. The chlorate ion can be derived from the oxidation of chlorine dioxide, from the impurities in the sodium chlorite feed stock, and from the photolytic decomposition of chlorine dioxide.

The chlorine dioxide residuals and other end products are believed to degrade more quickly than chlorine residuals, and, therefore, may not pose as serious a threat to aquatic life as chlorine residuals. An advantage in using chlorine dioxide is that it does not react with ammonia to form the potentially toxic chlorinated DBPs. It has also been reported that halogenated organic compounds are not produced to any appreciable extent.

Control of DBP Formation Using Chlorine Dioxide for Disinfection. The formation of chlorite can be controlled by careful management of the feedstock or increasing the chlorine dose beyond the stoichiometric amount. Treatment methods for the removal of the chlorite ion involve reducing the chlorite ion to the chloride ion using either ferrous iron or sulfite. Granular activated carbon (GAC) can also be used to absorb trace amounts of chlorite. At the present time there are no cost-effective methods for the removal of the chlorate ion. The control of the chlorate ion depends primarily on the effective management of the facilities used for the production of chlorine dioxide (White, 1999; Black and Veatch Corporation, 2010).

Environmental Impacts

The environmental impacts associated with the use of chlorine dioxide as a wastewater disinfectant are not well known. It has been reported that the impacts are less adverse than those associated with chlorination. Chlorine dioxide does not dissociate or react with water as does chlorine. However, because chlorine dioxide is normally produced from chlorine and sodium chlorite, free chlorine may remain in the resultant chlorine dioxide solution (depending on the process) and impact the receiving aquatic environment, as does chlorine and its byproducts.

12–5 DECHLORINATION

Chlorination is one of the most commonly used methods for the destruction of pathogenic and other harmful organisms that may endanger human health. As noted in the previous sections, however, certain organic constituents in wastewater interfere with the chlorination process. Many of these organic compounds may react with the chlorine to form toxic compounds that can have long-term adverse effects on the beneficial uses of the waters to which they are discharged or reused. To minimize the effects of these potentially toxic chlorine residuals on the environment, dechlorination of treated effluent is necessary. Dechlorination may be accomplished by reacting the residual chlorine with a reducing agent such as sulfur dioxide or sodium bisulfite or by adsorption on and reaction with activated carbon.

Dechlorination of Treated Wastewater with Sulfur Dioxide

Where effluent toxicity requirements are applicable, or where dechlorination is used as a polishing step following the breakpoint chlorination process for the removal of ammonia nitrogen, sulfur dioxide (SO_2) is used most commonly for dechlorination. Sulfur dioxide is available commercially as a liquefied gas under pressure in steel containers. Sulfur dioxide is handled in equipment very similar to standard chlorine systems. When added to water, sulfur dioxide reacts to form sulfurous acid (H_2SO_3), a strong reducing agent. In turn, the sulfurous acid dissociates to form HSO_3^- that will react with free and combined chlorine, resulting in formation of chloride and sulfate ions. Sulfur dioxide gas successively removes free chlorine, monochloramine, dichloramine, nitrogen trichloride, and poly-n-chlor compounds as illustrated in Eqs. (12–33) through (12–38).

Reactions between sulfur dioxide and free chlorine:

$$SO_2 + H_2O \rightarrow H_2SO_3 \tag{12–33}$$

$$\underline{HOCl + H_2SO_3 \rightarrow HCl + H_2SO_4} \tag{12–34}$$

$$SO_2 + HOCl \rightarrow HCl + H_2SO_4 \tag{12–35}$$

Reactions between sulfur dioxide and monochloramine, dichloramine, and nitrogen trichloride are:

$$NH_2Cl + H_2SO_3 + H_2O \rightarrow NH_4Cl + H_2SO_4 \tag{12–36}$$

$$NHCl_2 + 2H_2SO_3 + 2H_2O \rightarrow NH_4Cl + 2H_2SO_4 + HCl \tag{12–37}$$

$$NCl_3 + 3H_2SO_3 + 3H_2O \rightarrow NH_4Cl + 3H_2SO_4 + 2HCl \tag{12–38}$$

For the overall reaction between SO_2 and chlorine [Eq. (12–35)], the stoichiometric amount of SO_2 required per mg/L of chlorine residual is 0.903 mg/L. In practice, as reported in Table 12–16, it has been found that about 1.0 to 1.2 mg/L of sulfur dioxide will be required for the dechlorination of 1.0 mg/L of chlorine residue (expressed as Cl_2). Because the reactions of sulfur dioxide with chlorine and chloramines are nearly instantaneous, contact time is not usually a factor and contact chambers are not used, but rapid and positive mixing at the point of application is an absolute requirement.

The ratio of free chlorine to the total combined chlorine residual before dechlorination will determine whether the dechlorination process is partial or proceeds to completion. If the ratio is less than 85 percent, it can be assumed that significant organic nitrogen is present and that it will interfere with the dechlorination of free residual chlorine.

Table 12–16

Typical information on the quantity of dechlorinating compound required for each mg/L of residual chlorine

Dechlorinating compound			Quantity, mg/(mg/L) residual	
Name	**Formula**	**Molecular weight**	**Stoichiometric amount**	**Range in use**
Hydrogen peroxide	H_2O_2	34.01	0.48	0.5–0.7
Sodium bisulfite	$NaHSO_3$	104.06	1.46	1.5–1.7
Sodium metabisulfite	$Na_2S_2O_5$	190.10	1.34	1.4–1.6
Sodium sulfite	Na_2SO_3	126.04	1.78	1.8–2.0
Sodium thiosulfate	$Na_2S_2O_3$	112.12	0.56	0.6–0.9
Sulfur dioxide	SO_2	64.09	0.903	1.0–1.2

In most situations, sulfur dioxide dechlorination is a very reliable unit process, provided that the precision of the combined chlorine residual monitoring service is adequate. Excess sulfur dioxide dosages should be avoided, not only because of the chemical wastage, but also because of the oxygen demand exerted by the excess sulfur dioxide. The relatively slow reaction between excess sulfur dioxide and dissolved oxygen is given by the following expression:

$$HSO_3^- + 0.5O_2 \rightarrow SO_4^{2-} + H^+ \qquad (12\text{–}39)$$

The result of this reaction is a reduction in the dissolved oxygen in the water, a corresponding increase in the measured BOD and COD, and a possible drop in the pH. All these effects can be eliminated by proper control of the dechlorination system.

Dechlorination of Treated Wastewater with Sodium Based Compounds

Sodium based chemicals that have been used for dechlorination include sodium sulfite (Na_2SO_3), sodium bisulfite ($NaHSO_3$), sodium metabisulfite ($Na_2S_2O_5$), sodium thiosulfate ($Na_2S_2O_3$), and hydrogen peroxide (H_2O_2). When these chemicals are used for dechlorination, the following reactions occur. The stoichiometric weight ratios of these compounds needed per mg/L of residual chlorine are given in Table 12–15, along with the range of values used in practice.

Sodium Sulfite. Reactions between sodium sulfite and free chlorine residual and combined chlorine residual, as represented by monochloramine:

$$Na_2SO_3 + Cl_2 + H_2O \rightarrow Na_2SO_4 + 2HCl \qquad (12\text{–}40)$$

$$Na_2SO_3 + NH_2Cl + H_2O \rightarrow Na_2SO_4 + Cl^- + NH_4^+ \qquad (12\text{–}41)$$

Sodium Bisulfite. Reaction between sodium bisulfite and free chlorine residual and combined chlorine residual, as represented by monochloramine:

$$NaHSO_3 + Cl_2 + H_2O \rightarrow NaHSO_4 + 2HCl \qquad (12\text{–}42)$$

$$NaHSO_3 + NH_2Cl + H_2O \rightarrow NaHSO_4 + Cl^- + NH_4^+ \qquad (12\text{–}43)$$

Sodium Metabisulfite. Reactions between sodium metabisulfite and free chlorine residual and combined chlorine residual, as represented by monochloramine:

$$Na_2S_2O_5 + Cl_2 + 3H_2O \rightarrow 2NaHSO_4 + 4HCl \qquad (12\text{–}44)$$

$$Na_2S_2O_5 + 2NH_2Cl + 3H_2O \rightarrow Na_2SO_4 + H_2SO_4 + 2Cl^- + 2NH_4^+ \qquad (12\text{–}45)$$

Sodium Thiosulfate and Related Compounds. Often used as a dechlorinating agent in analytical laboratories, the use of sodium thiosulfate ($Na_2S_2O_3$) in full scale water reclaimation treatment plants is limited for the following reasons. It appears the reaction of sodium thiosulfate with residual chlorine is stepwise, creating a problem with uniform mixing. The ability of sodium thiosulfate to remove residual chlorine is a function of the pH (White, 1999; Black and Veatch Corporation, 2010). The reaction with residual chlorine is only stoichiometric at a pH value of 2, making prediction of the required dose impossible in wastewater applications. As reported in Table 12–16, the stoichiometric weight ratio of sodium thiosulfate per mg/L of residual chlorine is 0.556. Although not in common use, calcium thiosulfate (CaS_2O_3) ascorbic acid ($C_6H_8O_6$), and sodium ascorbate ($C_6H_7NaO_6$) have all been used at full scale for dechlorination.

Dechlorination with Hydrogen Peroxide

Hydrogen peroxide (H_2O_2) has also been used for dechlorination. Unlike sulfur dioxide and the sodium-based compounds discussed above, hydrogen peroxide does not result in an increase in the total dissolved solids as it only adds oxygen to the water. When hydrogen peroxide is used for dechlorination, the following reaction occurs.

$$H_2O_2 + Cl_2 \rightarrow O_2 + 2HCl \tag{12–46}$$

The stoichiometric weight ratio of hydrogen peroxide needed per mg/L of residual chlorine is 0.48. Because the reaction between hydrogen peroxide and chlorine compounds is so rapid, other inorganic and organic compounds generally do not interfere with the reaction. The optimal pH range is about 8.5 at which the reaction occurs instantaneously, although there is no upper limit. In the past, hydrogen peroxide has not been used for dechlorination because it is difficult to handle.

Dechlorination with Activated Carbon

Both combined and free residual chlorine can be removed by means of adsorption on and reaction with activated carbon. When activated carbon is used for dechlorination, the following reactions occur once chlorine or chlorine compounds have been adsorbed.

Reactions with free chlorine residual:

$$C + 2Cl_2 + 2H_2O \rightarrow 4HCl + CO_2 \tag{12–47}$$

Reactions with combined residual as represented by mono- and dichloramine:

$$C + 2NH_2Cl + 2H_2O \rightarrow CO_2 + 2NH_4^+ + 2Cl^- \tag{12–48}$$

$$C + 4NHCl_2 + 2H_2O \rightarrow CO_2 + 2N_2 + 8H^+ + 8Cl^- \tag{12–49}$$

Granular activated carbon is used in either a gravity or pressure filter bed. If carbon is to be used solely for dechlorination, it must be preceded by an activated carbon process for the removal of other constituents susceptible to removal by activated carbon. In treatment plants where granular activated carbon is used to remove organics, either the same or separate beds can also be used for dechlorination.

Because granular carbon in column applications has proved to be very effective and reliable, activated carbon should be considered where dechlorination is required. However, this method is quite expensive. It is expected that the primary application of activated carbon for dechlorination will be in situations where high levels of organic removal are also required.

Dechlorination of Chlorine Dioxide with Sulfur Dioxide

Where treated wastewater is disinfected with chlorine dioxide, dechlorination can be achieved using sulfur dioxide. The reaction that takes place in the chlorine dioxide solution can be expressed as

$$SO_2 + H_2O \rightarrow H_2SO_3 \tag{12–50}$$

$$5H_2SO_3 + 2ClO_2 + H_2O \rightarrow 5H_2SO_4 + 2HCl \tag{12–51}$$

Based on Eq. (12–51), it can be seen that 2.5 mg of sulfur dioxide will be required for each mg of chlorine dioxide residual (expressed as ClO_2). In practice, 2.7 mg SO_2/mg ClO_2 would normally be used.

12–6 DESIGN OF CHLORINATION AND DECHLORINATION FACILITIES

The chemistry of chlorine in water and wastewater has been discussed in the previous sections, along with an analysis of how chlorine functions as a disinfectant. Important considerations in the implementation of chlorination and dechlorination facilities for a variety of purposes include (1) estimation of the chlorine dosage, (2) application flow diagrams, (3) dosage control, (4) injection and initial mixing, (5) chlorine contact basin design, (6) assessing the hydraulic performance of existing chlorine contact basins, (7) outlet control and chlorine residual measurement, (8) chlorine storage facilities, (9) chemical containment and neutralization facilities, and (10) dechlorination facilities. These topics are considered in the following discussion.

Sizing Chlorination Facilities

To aid in the design and selection of the required chlorination facilities and equipment, it is important to know the uses, including dosage ranges, to which chlorine and its compounds have been applied. Chlorination capacities for disinfection are generally selected to meet the specific design criteria of the state or other regulatory agencies controlling the receiving body of water. In any case, where the residual in the effluent is specified or the final number of coliform bacteria is limited, onsite testing is preferred to determine the dosage of chlorine required. Typical chlorine dosages for disinfection have been given previously in Table 12–14. Typical chlorine dosages for applications other than disinfection are given in Table 12–17. A range of dosage values is given because they will vary depending on the characteristics of the wastewater. In the absence of more specific data, the maximum values given in Tables 12–14 and 12–17 can be used as a guide in sizing chlorination equipment. The sizing of chlorination facilities is illustrated in Example 12–5.

Table 12–17

Typical dosages for various chlorination applications in wastewater collection and treatment

Application	Dosage range, mg/L
Collection:	
Corrosion control (H_2S)	2–9[a]
Odor control	2–9[a]
Slime growth control	1–10
Treatment:	
BOD reduction	0.5–2[b]
Digester and Imhoff tank foaming control	2–15
Digester supernatant oxidation	20–140
Ferrous sulfate oxidation	–[c]
Filter fly control	0.1–0.5
Filter ponding control	1–10
Grease removal	2–10
Sludge bulking control	1–10

[a] Per mg/L of H_2S.

[b] Per mg/L of BOD_5 destroyed.

[c] $6FeSO_4 \cdot 7H_2O + 3Cl_2 \rightarrow 2Fe_2(SO_4)_3 + 42H_2O$.

EXAMPLE 12–5 **Sizing of Chlorination Facilities** Determine the capacity of a chlorinator for a treatment plant with an average wastewater flowrate of 1000 m^3/d (0.26 Mgal/d). The peak daily factor for the treatment plant is 3.0 and the maximum required chlorine dosage (set by state regulations) is to be 20 mg/L.

Solution

1. Determine the capacity of the chlorinator at peak flow.

$$\begin{aligned} Cl_2, \text{kg/d} &= (20\ \text{g/m}^3)(1000\ \text{m}^3/\text{d})(3)(1\ \text{kg}/10^3\ \text{g}) \\ &= 60\ \text{kg/d} \end{aligned}$$

Use the next largest standard size chlorinator: two 90 kg/d (200 lb/d) units with one unit serving as a spare. Although the peak capacity will not be required during most of the day, it must be available to meet the chlorine requirements at peak flow. Best design practice calls for the availability of a standby chlorinator.

2. Estimate the daily consumption of chlorine. Assume an average dosage of 10 mg/L.

$$\begin{aligned} Cl_2, \text{kg/d} &= (10\ \text{g/m}^3)(1000\ \text{m}^3/\text{d})(1\ \text{kg}/10^3\ \text{g}) \\ &= 10\ \text{kg/d} \end{aligned}$$

Comment In sizing and designing chlorination systems, it is also important to consider the low flow/dosage requirements. The chlorination system should have sufficient turndown capability for these conditions so that excessive chlorine is not applied.

Disinfection Process Flow Diagrams

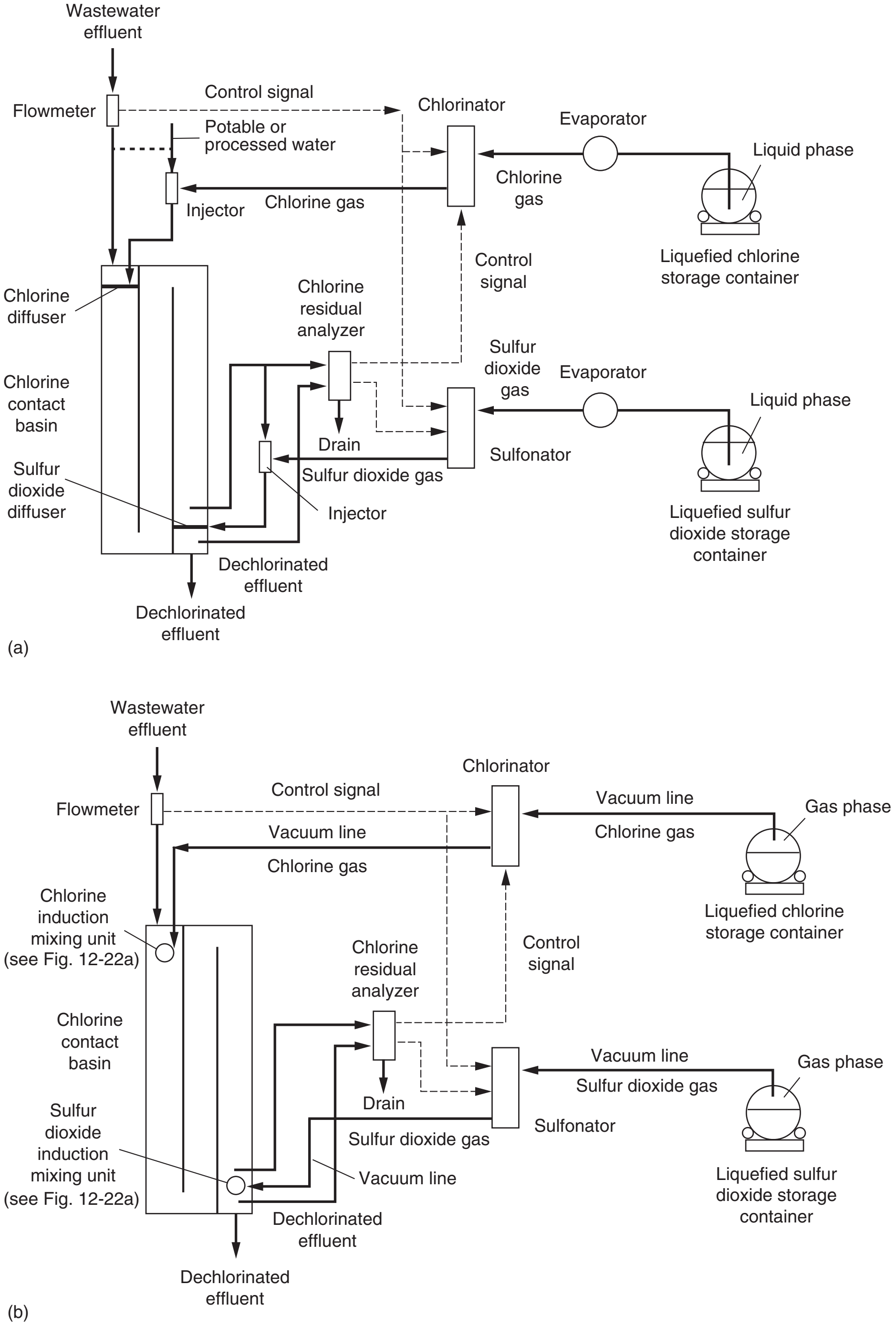
Process flow diagrams and equipment used to inject (feed) chlorine, hypochlorite, dry calcium hypochlorite, and chlorine dioxide into wastewater are illustrated and discussed below.

Flow Diagram for Chlorine. Chlorine may be applied directly as a gas or in an aqueous solution. Typical chlorine/sulfur dioxide chlorination/dechlorination process flow diagrams are shown on Fig. 12–17. The difference in the two diagrams shown on Fig. 12–17 is in the method of introducing and mixing the chlorine solution with the wastewater. Chlorine can be withdrawn from storage containers either in liquid or gas form. If withdrawn as a gas, the evaporation of the liquid in the container results in frost formation that restricts gas withdrawal rates to 18 kg/d (40 lb/d) for 68 kg (150 lb) cylinders and 205 kg/d (450 lb/d) for 0.9-tonne (1-ton) containers at 21°C (70°F). Evaporators are used normally where the maximum rate of chlorine gas withdrawal from a 0.9-tonne (1-ton) container must exceed approximately 180 kg/d (400 lb/d). Although multiple ton cylinders can be connected to provide more than 180 kg/d (400 lb/d), the use of an evaporator conserves space. Evaporators are almost always used when the total dosage exceeds 680 kg/d (1500 lb/d). Chlorine evaporators are available in sizes ranging from 1818 to 4545 kg/d (4000 to 10,000 lb/d) capacities; chlorinators are available normally in sizes ranging from 227 to 4545 kg/d (500 to 10,000 lb/d).

Flow Diagram for Liquid Hypochlorite Solutions. A typical sodium hypochlorite/sodium bisulfite chlorination/dechlorination process flow diagram is shown on Fig. 12–18. For small treatment plants, the most satisfactory means of feeding sodium or calcium hypochlorite is through the use of low capacity proportioning pumps. Generally, pumps are available in capacities up to 450 L/d (120 gal/d), with adjustable stroke for any reduced values. Large capacities or multiple units are available from some of

Figure 12–17

Schematic flow diagrams for chlorination/dechlorination: (a) using a chlorine injector system and (b) using a molecular chlorine vapor induction system.

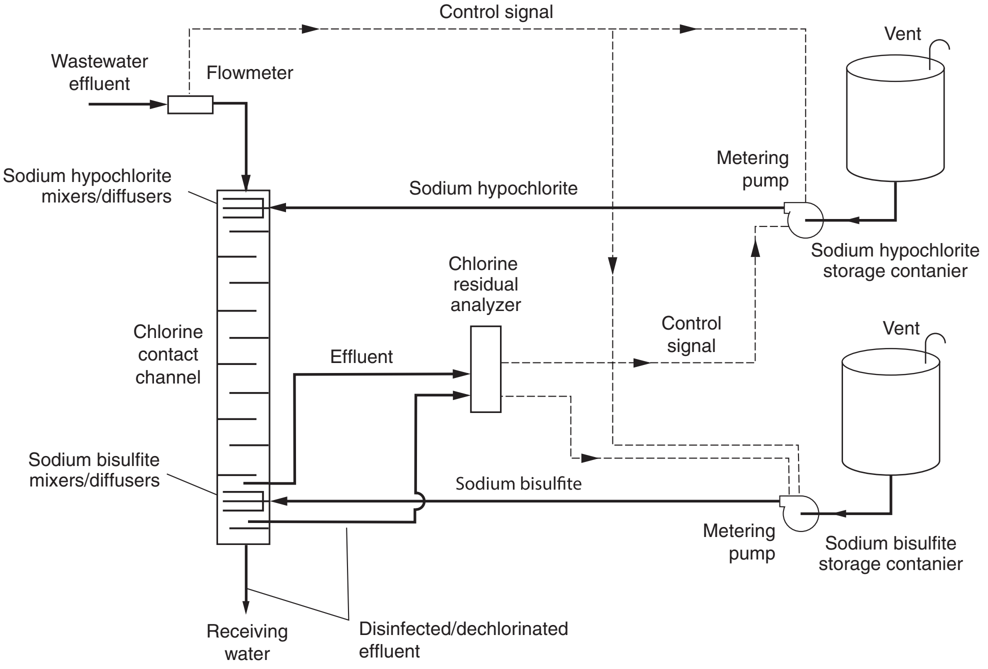


the manufacturers. The pumps can be arranged to feed at a constant rate, or they can be provided with variable speed and with analog signals for varying the feed rate. The stroke length can also be controlled.

Flow Diagram for Dry Calcium Hypochlorite Feed System. For small wastewater flowrates up to about 400 m^3/d (10^5 gal/d) chlorine in the form of dry calcium

Figure 12–18

Schematic flow diagrams for sodium hypochlorite chlorination with sodium bisulfite dechlorination.



hypochlorite tablets is used for disinfection. Two of the most common types of tablet chlorinators (nonpressurized and pressurized) are shown on Fig. 12–19. The schematic flow diagrams for the two tablet chlorinators are essentially the same; a sidestream of water is diverted from the main discharge line, chlorine at relatively high concentrations is added to the sidestream, and the chlorinated sidestream is discharged back into the main flow by means of a pump [see Fig. 12–19(a)] or by reducing the pressure in the main discharge line [see Fig. 12–19(b)]. As shown on Fig. 12–19(a), in the nonpressurized tablet chlorinator the sidestream contacts the bottom surface of the chlorine tablets that rest on a screen. The chlorine tablets have been designed to dissolve at a more or less

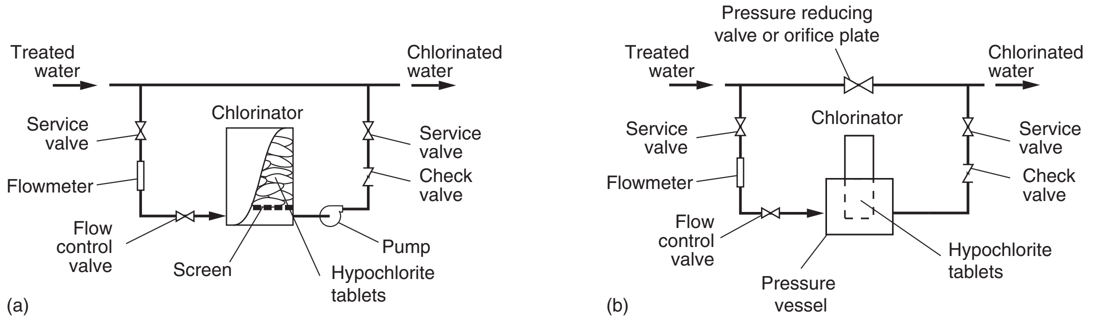
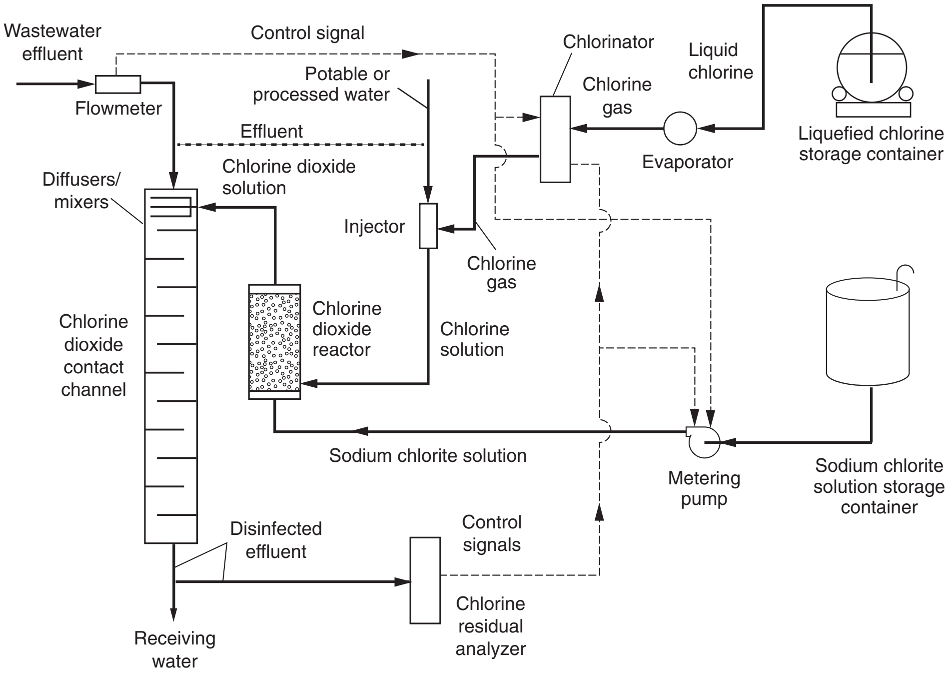


Figure 12–19

Schematic flow diagrams for calcium hypochlorite tablet chlorinators: (a) nonpressurized (adapted from PPG Industries, Inc.) and (b) pressurized (adapted from PPG Industries, Inc.).

Figure 12–20

Typical flow diagram for the addition of chlorine dioxide.



constant rate, releasing a controlled amount of chlorine. The amount of chlorine added is dependent on the flowrate through the tablet chlorinator. The tablet chlorinator shown on Fig. 12–19(b) is pressurized and hypochlorite is released as water flows over the hypochlorite tablets. Dry calcium hypochlorite tablets, typically 75 mm (3 in.) in diameter, contain about 65 to 70 percent available chlorine. For small treatment plants, the use of chlorine tablets eliminates the hazards associated with handling chlorine cylinders. Further, because there are no moving parts, tablet chlorinators are simple to operate and maintain.

Flow Diagram for Chlorine Dioxide. The chlorine dioxide, generated onsite, is present in an aqueous solution which is applied in same manner as that used for typical chlorination systems. A schematic process flow diagram of a typical chlorine dioxide installation is shown on Fig. 12–20.

Dosage Control

The control of the chlorine dosage can be accomplished in a number of different ways depending on the disinfection objective. The principal control methods are summarized in Table 12–18. The specific control method to be used will depend on the variability of the influent flowrate, the presence of unoxidized constituents that can react with chlorine, the pH of the wastewater, and whether combined or free or a combination of the two forms of chlorine will be used for disinfection. Dosage control is easiest where combined chlorine is used as the disinfectant. Dosage control has also become more difficult due to the impact of climate change resulting in short duration, high intensity rainfall events. Because of the rapid increase in the influent flowrate observed at some treatment plants, due to increased stormwater runoff, the carryover (washout) of solids from the secondary sedimentation basins, especially where shallow basins are used, has further complicated dosage control

Table 12–18

Methods used to control the chlorine dosage for disinfection[a]

Control method	Description
Manual control	Manual control, where the operator changes the feed rate to suit conditions, is the simplest method for controlling the chlorine dose. The required dosage is usually determined by measuring the free and/or the combined chlorine residual at the end of the chlorine contact basin and adjusting the chlorine dosage to obtain a the desired residual. This method works best where combined chlorine is used for disinfection, and the flowrate does not vary too rapidly, but can also be used where free chlorine residual is used.
Manual control with online effluent residual chlorine monitoring	An online chlorine analyzer is used to monitor the chlorine residual in the effluent from the chlorine contact basin. The chlorine dose is adjusted manually based on the plant flowrate and the residual chlorine concentrations. This method works best where combined chlorine is used for disinfection, and the flowrate does not vary too rapidly.
Flow pacing	The chlorine flowrate is paced proportional to the wastewater flowrate as measured by a primary meter such as a magnetic meter, Parshall flume, or flow tube. This method works best where combined chlorine is used for disinfection.
Flow pacing with online effluent residual chlorine monitoring	The chlorine dosage is controlled by automatic measurement of the chlorine residual and the wastewater flowrate. An automatic analyzer with signal transmitter and recorder is required.
Flow pacing with online effluent residual chlorine monitoring and automatic control	The control signals obtained from the wastewater flowmeter and the residual monitor are fed to a programmable logic controller (PLC) to provide more precise control of chlorine dosage and residual. This method works best where combined chlorine is used for disinfection.
Flow pacing with online residual chlorine monitoring after initial demand and automatic control	In this method, the chlorine residual is measured a short distance downstream from the point of chlorine addition. The readings from the wastewater flowmeter and the residual chlorine monitor are fed to a PLC to provide more precise control of chlorine dosage and residual. This method works best where combined chlorine is used for disinfection.
Flow pacing with online free and combined residual chlorine monitoring and automatic control	This approach is used for the disinfection of nitrified effluents with free chlorine where a variable residual ammonia must be removed to reach the breakpoint. The free and combined chlorine residual concentrations along with the readings from the wastewater flowmeter are fed to a PLC to provide more precise control of chlorine dosage. This approach is complex as the PLC must be programmed to recognize the difference between free and combined chlorine residuals and be able to interpret the data with respect to the chemistry of the breakpoint reaction. Data from an online ammonia analyzer, now available for field use, can also be integrated with the other data fed to the PLC to optimize the disinfection process with free chlorine.

[a] Adapted in part from Kobylinski et al. (2006).

because of the increase in chlorine demand needed to disinfect microorganisms embedded in floc particles. With manual control it is difficult, if not impossible, to maintain a constant CT, system monitoring and control is resource intensive, manual system is resource intensive; and chemical usage is higher due to residual variability. With automated systems, the ability to maintain the online analyzers is of critical importance if the benefits of automation are to be realized (Hurst, 2012).

Where free chorine is to be used as the disinfectant, dosage control is more difficult, especially if the concentration of residual ammonia in the effluent to be disinfected is somewhat variable. As noted in Table 12–18, both free and combined chlorine residual monitors can be used in conjunction with readings from the influent wastewater flowmeter(s) to provide input to a programmable logic controller (PLC). More recently, online ammonia analyzers have been developed that can be used in conjunction with a PLC and inputs from chlorine residual monitors and flowrate measuring devices. Because the three online monitors must be maintained for complete automated control of the disinfection process, such systems are best employed at larger treatment plants with adequate staff.

Injection and Initial Mixing

As pointed out previously in Section 12–3, other things being equal, effective mixing of the chlorine solution with the wastewater, the contact time, and the chlorine residual are the principal factors involved in achieving effective bacterial kill. The addition of chlorine solution is often accomplished with a diffuser, which may be a plastic pipe with drilled holes through which the chlorine solution can be distributed into the path of wastewater flow (see Fig. 12–21). Unfortunately, the use of diffusers for adding chlorine is not very effective. To optimize the performance of disinfection systems, the chlorine should be introduced and mixed as rapidly as possible (ideally in less than a second). Techniques that can be used to mix chlorine in a fraction of a second were introduced and discussed in Chap. 5. Effective devices for mixing chlorine with the wastewater within a fraction of a second are illustrated on Fig. 12–22.

Chlorine Contact Basin Design

The principal design objective for chlorine contact basins is to ensure that some defined percentage of the flow remains in the chlorine contact basin for the design contact time to ensure effective disinfection. The contact time is usually specified by the regulatory agency and may range from 30 to 120 min; contact times of 15 to 90 min at peak flow are common. For example, for reuse applications the CDPH of the State of California requires a CT value of 450 mg·min/L with a modal contact time of 90 min at peak flow. Issues related to the design and analysis of chlorine contact basins considered in the following discussion include (1) basin configuration, (2) the use of baffles and guide vanes, (3) number of chlorine contact basins, (4) precipitation of solids in chlorine contact basins, (5) solids transport velocity and (6) a procedure for predicting disinfection performance.

Chlorine Contact Basin Configuration. To be assured that a given percentage of the flow will remain in the chlorine contact basin for a given period of time, the most common approach is to use long plug-flow, around-the-end type of contact basins (see Fig. 12–13) or a series of interconnected basins or compartments. Plug-flow chlorine contact basins that are built in a serpentine fashion (e.g., folded back and forth) to conserve space require special attention in their design to eliminate the formation of hydraulic dead zones that will reduce the hydraulic detention times. Length-to-width ratios (L/W) of at least 20 to 1 (preferably 40 to 1) and the use of baffles and guide vanes, as described below,

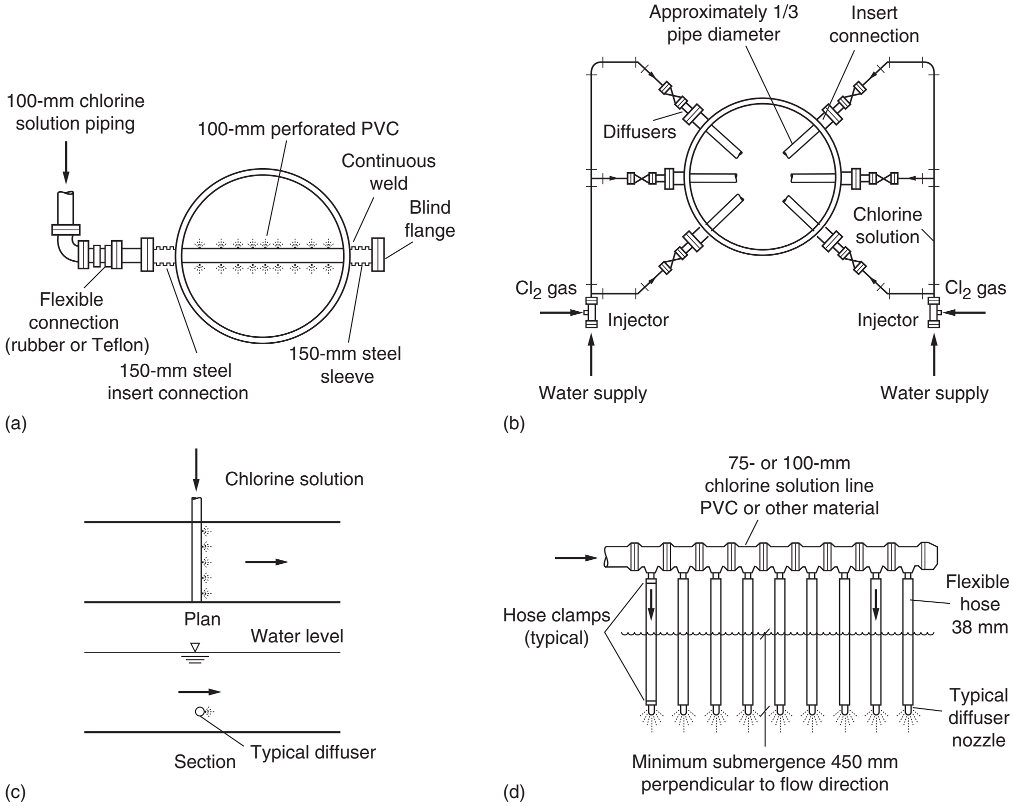


Figure 12–21

Typical diffusers used to inject chlorine solution: (a) across the pipe diffuser (b) diffuser system for large conduits, (c) single across the channel diffuser, and (d) typical hanging nozzle type chlorine diffuser for open channels. (Adapted from White, 1999.)

will help to minimize short circuiting. In some small plants, chlorine contact basins have been constructed of large diameter sewer pipe. The design of a chlorine contact basin based on dispersion is considered in Example 12–6.

Use of Baffles and Deflection Guide Vanes. To improve the hydraulic performance of chlorine contact basins, it has become common practice to use either submerged baffles, deflection guide vanes, or combinations of the two. Submerged baffles are used to break up density currents caused by temperature gradients, to limit short circuiting, and to minimize the effect of hydraulic dead spaces. The location of the baffles is critical in improving the performance of chlorine contact basins. A typical placement of baffles, and the effect on the corresponding tracer response curves, is illustrated on Fig. 12–23. As shown on Fig. 12–23, the addition of baffles improves the hydraulic performance of the chlorine contact basin significantly. The open area in submerged baffles will typically vary from 6 to 10 percent of the cross-sectional area of flow. The headloss through each baffle can be estimated using the following expression:

$$h = \frac{1}{2g}\left(\frac{Q}{Cna}\right)^2 \tag{12–52}$$

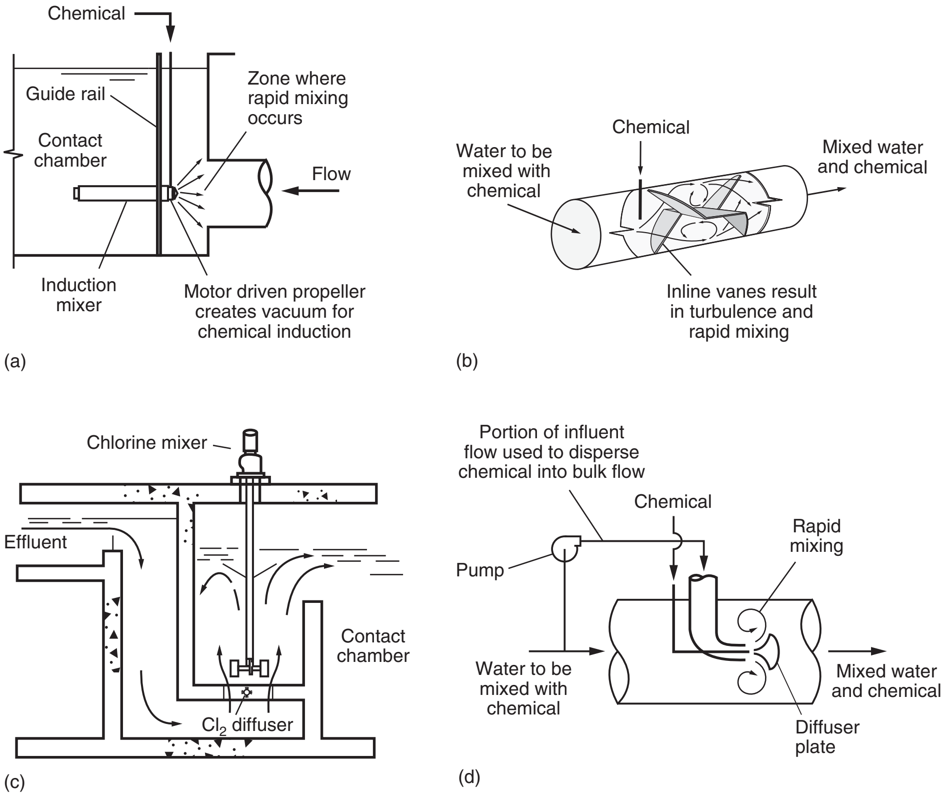


Figure 12–22

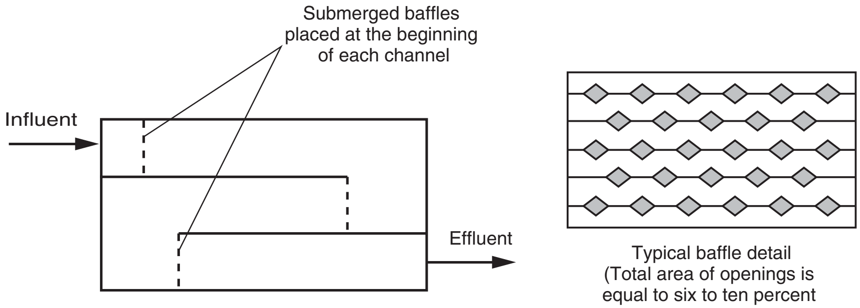
Typical mixers for the addition of chlorine: (a) Water Champ® induction mixer can be mounted horizontally, as shown, or vertically, depending on the basin configuration, (b) inline static mixer, (c) inline turbine mixer, and (d) inline injector pump type. For additional types of chlorine mixers see Fig. 5–12 in Chap. 5.

where h = headloss, m
g = acceleration due to gravity, 9.81 m/s^2
Q = discharge through chlorine contact basin channel, m^3/s
C = discharge coefficient, unitless (typically about 0.8)
n = number of openings
a = area of individual opening, m^2

An alternative approach that has been used to improve the performance of chlorine contact basins is through the addition of deflection guide vanes, as shown on Fig. 12–24. The placement and number of vanes will depend on the layout of the chlorine contact basin. Two or three guide vanes are used most commonly. The beneficial effect of adding guide vanes was studied extensively by Louie and Fohrman (1968).

Number of Chlorine Contact Basins. For most treatment plants, two or more contact basins should be used to meet reliability and redundancy requirements to facilitate maintenance and cleaning. Provisions should also be included for draining and scum removal. Vacuum type cleaning equipment may be used as an alternative to draining the basin for removal of accumulated solids. Bypassing the contact basin for maintenance should only be practiced on rare occasions, with the approval of regulatory agencies. If the time of travel in the outfall sewer at the maximum design flowrate is sufficient to equal or exceed the required contact time, it may be possible to eliminate the chlorine contact chambers, provided regulatory authorities agree.

Precipitation in Chlorine Contact Basins. A problem often encountered in the operation of chlorine contact basins is the formation and precipitation of a light flocculent



(a) Plan view of chlorine contact basin

(b) Typical baffle detail (Total area of openings is equal to six to ten percent of cross-sectional area)

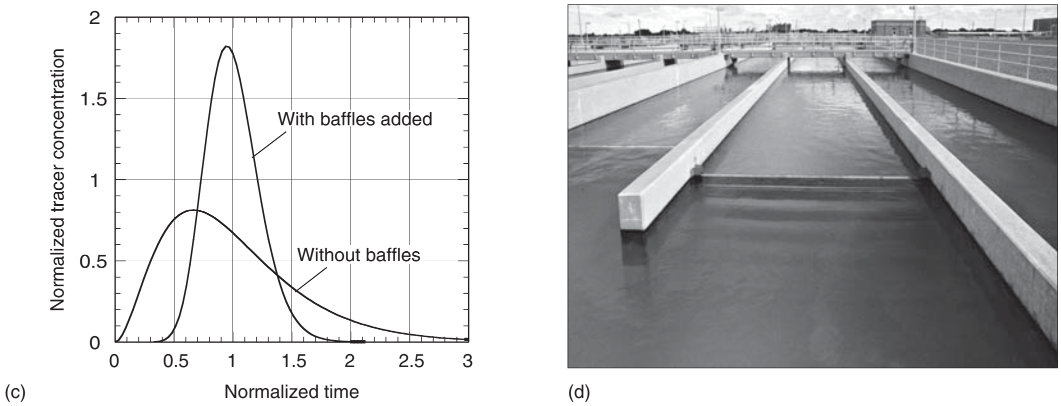
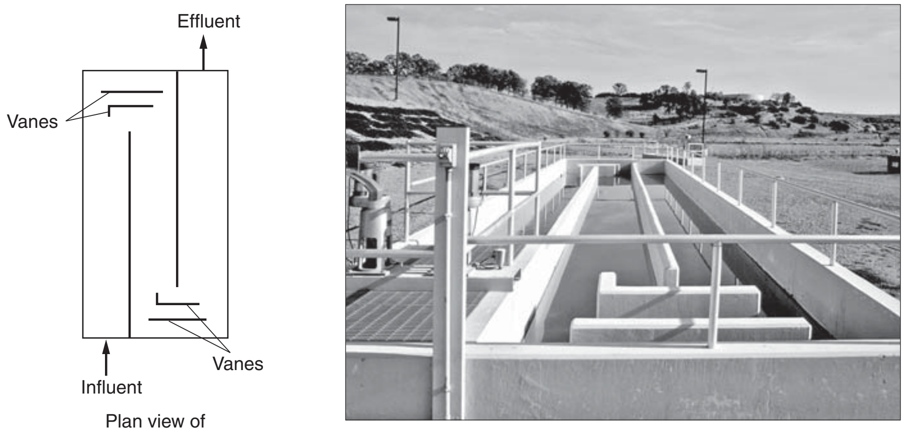


Figure 12–23

Baffling in chlorine contact basins: (a) placement of baffles in chlorine contact tank at the beginning of each channel (or pass) is critical (adapted from Crittenden et al., 2005), (b) typical submerged baffle detail (adapted from Kawamura, 2000), (c) effect of the use of baffles in chlorine contact basins (adapted from Hart, 1979), and (d) view of chlorine contact tank with submerged wooden baffles placed at the beginning and end of each channel.

Figure 12–24

Chlorine contact basin with flow deflection vanes: (a) schematic and (b) photograph of empty chlorine contact basin designed with guide vanes.



(a) Plan view of chlorine contact basin

(b)

material. The principal cause of the formation and precipitation of floc is the lowering of the pH that results from the addition of chlorine. The problem occurs most frequently where alum is used for phosphorus removal in separate precipitation facilities or is added before the effluent filters. For a variety of reasons including high pH and inadequate initial mixing, not all of the alum added will react completely to form a floc that can be removed by precipitation or filtration. However, when the pH is lowered in the chlorine contact basin due to the addition of chlorine, some of the unreacted alum may form a floc. Oxidation of organic material in the treated effluent can also result in the formation of precipitates. Thus, in addition to meeting reliability and redundancy requirements, a minimum of two chlorine contact basins is necessary to allow one basin to be removed from service so that the accumulated solids can be removed from the basins.

Solids Transport Velocity. The horizontal velocity at minimum flow in a chlorine contact basin should, in theory, be sufficient to scour the bottom or to limit the deposition of sludge solids that may have passed through the settling tank. To limit excessive deposition, horizontal velocities should be at least 2 to 4.5 m/min (6.5 to 15 ft/min). In general, it will be difficult to achieve such velocities and simultaneously meet stringent dispersion requirements (see Example 12–6). If floc particles form, it will generally be impossible to avoid the accumulation of a sludge layer in the chlorine contact basins, another reason at least two chlorine contact basins should be used.

EXAMPLE 12–6 **Design of a Chlorine Contact Basin Based on Dispersion** Design a chlorine contact basin for secondary effluent with an average flowrate of 4000 m^3/d. The estimated peaking factor is 2.0. The detention time at peak flow is to be 90 min. A minimum of two parallel channels must be used for redundancy requirements. The dimensions of the chlorine contact basin should be such to achieve a dispersion number equal to or less 0.015 at peak flow. Also check the dispersion number at average flow. What will happen if the low flow drops to 33 percent of the average flow in the early morning hours? Based on the resulting calculations will solids deposition occur, requiring periodic draining and cleaning of the chlorine contact basin?

Solution

1. Assume some trial cross-sectional dimensions for the chlorine contact basin and determine the corresponding length and flow velocity.
   a. Assumed dimensions

      Width = 2 m (6.6 ft)

      Depth = 3 m (9.8 ft)

      Number of parallel channels = 2

   b. Determine required length

$$L = \frac{(2 \times 4000\ \text{m}^3/\text{d})}{(2)(1440\ \text{min/d})} \times (90\ \text{min}) \times \frac{1}{(2\ \text{m} \times 3\ \text{m})} = 41.7\ \text{m}$$

   c. Check velocity at peak flow

$$v = \frac{(2 \times 4000\ \text{m}^3/\text{d})}{(2)(1440\ \text{min/d})(60\ \text{s/min})} \times \frac{1}{(2\ \text{m} \times 3\ \text{m})} = 0.0077\ \text{m/s}$$

2. Determine the coefficient of dispersion using Eq. (I–14) from Appendix I and the dispersion number using Eq. (I–9) from Appendix I for the chlorine contact basin.

a. Compute the coefficient of dispersion

$D = 1.01\nu N_R^{0.875}$

i. Compute the Reynolds number

$N_R = 4vR/\nu$

v = velocity in open channel, LT^{-1}, (m/s)

R = hydraulic radius = area/wetted perimeter, L, (m)

$v = 0.0077$ m/s

$\nu = 1.003 \times 10^{-6}$ m²/s (at 20°C)

$$N_R = \frac{(4)(0.0077 \text{ m/s})[(2.0 \text{ m} \times 3.0 \text{ m})/(2 \times 3.0 \text{ m} + 2.0 \text{ m})]}{(1.003 \times 10^{-6} \text{ m}^2/\text{s})} = 23{,}031$$

ii. Determine the coefficient of dispersion

$$D = 1.01(1.003 \times 10^{-6} \text{ m}^2/\text{s})(23{,}031)^{0.875} = 6.648 \times 10^{-3} \text{ m}^2/\text{s}$$

b. Determine the dispersion number

$$d = \frac{D}{vL} = \frac{Dt}{L^2} = \frac{(0.006648 \text{ m}^2/\text{s})(90 \text{ min} \times 60 \text{ s/min})}{(41.4 \text{ m}^2)} = 0.0206$$

Because the computed dispersion number (0.0206) is greater than the desired value (0.015), an alternative design must be evaluated. For the alternative design, assume three parallel channels will be used.

3. Assume new trial cross-sectional dimensions for the chlorine contact basin and determine the new length and flow velocity.

a. Assumed dimensions

Width = 1.25 m (5.0 ft)

Depth = 3 m (9.8 ft)

Number of parallel channels = 3

b. Determine required length

$$L = \frac{(2 \times 4000 \text{ m}^3/\text{d})}{(3)(1440 \text{ min/d})} \times (90 \text{ min}) \times \frac{1}{(1.25 \text{ m} \times 3 \text{ m})} = 44.4 \text{ m}$$

c. Check velocity at peak flow

$$v = \frac{(2 \times 4000 \text{ m}^3/\text{d})}{(3)(1440 \text{ min/d})(60 \text{ s/min})} \times \frac{1}{(1.25 \text{ m} \times 3 \text{ m})} = 0.0082 \text{ m/s}$$

4. Check the dispersion number for the chlorine contact basin.

a. Compute the Reynolds number

$N_R = 4vR/\nu$

$v = 0.0082$ m/s

$\nu = 1.003 \times 10^{-6}$ m²/s (at 20°C)

$$N_R = \frac{(4)(0.0082 \text{ m/s})[(1.25 \text{ m} \times 3.0 \text{ m})/(2 \times 3.0 \text{ m} + 1.25 \text{ m})]}{(1.003 \times 10^{-6} \text{ m}^2/\text{s})} = 16{,}915$$

b. Compute the coefficient of dispersion

$$D = 1.01\nu N_R^{0.875}$$

$$D = 1.01 \times 1.003 \times 10^{-6}\ \text{m}^2\text{/s}\ (16{,}915)^{0.875} = 5.07 \times 10^{-3}\ \text{m}^2\text{/s}$$

c. Determine the dispersion number

$$d = \frac{D}{vL} = \frac{Dt}{L^2} = \frac{(0.00507\ \text{m}^2\text{/s})(90\ \text{min} \times 60\ \text{s/min})}{(44.4\ \text{m})^2} = 0.0139$$

Because the computed dispersion number (0.0139) is smaller than the desired value (0.015), the proposed design is acceptable.

5. Check the dispersion number for the chlorine contact basin at average flow.

a. Compute the Reynolds number

$$N_R = 4vR/\nu$$

$$v = 0.0082/2 = 0.0041\ \text{m/s}$$

$$\nu = 1.003 \times 10^{-6}\ \text{m}^2\text{/s}$$

$$N_R = \frac{(4)(0.0041\ \text{m/s})[(1.25\ \text{m} \times 3.0\ \text{m})/(2 \times 3.0\ \text{m} + 1.25\ \text{m})]}{(1.003 \times 10^{-6}\ \text{m}^2\text{/s})} = 8{,}457$$

b. Determine the coefficient of dispersion

$$D = 1.01\ \nu N_R^{0.875}$$

$$D = 1.01(1.003 \times 10^{-6}\ \text{m}^2\text{/s})(8{,}457)^{0.875} = 2.77 \times 10^{-3}\ \text{m}^2\text{/s}$$

c. Determine the dispersion number

$$d = \frac{D}{vL} = \frac{Dt}{L^2} = \frac{(0.00277\ \text{m}^2\text{/s})(90\ \text{min} \times 60\ \text{s/min})}{(44.4\ \text{m})^2} = 0.0076$$

d. Because the velocity is reduced at average flow, the computed dispersion number is equivalent to about 66 complete-mix reactors in series.

Comment Under all flow conditions, deposition of residual suspended solids would be expected in the chlorine contact basin, especially so at low flow.

Predicting Disinfection Performance. An extremely important issue in the design of chlorine contact basins is being able to predict the performance of the proposed design. To predict performance, the actual residence time that a given molecule of the fluid spends in the reactor must be known. The residence time in the reactor can be determined using some of the analytical techniques presented previously in Chap. 1 and Appendix I. The pertinent equations from Appendix I are repeated here for convenience. In Appendix I, it was noted that the Peclet number divided by 2 is equal to the number of reactors in series. The relationship of the Peclet number to the dispersion number is

$$P_e = \frac{vL}{D} = \frac{1}{d} \tag{12–53}$$

For complete mix-reactors in series, the normalized residence time distribution curve, $E(\theta)$, where θ is equal to t/τ for n reactors in series, as derived in Appendix I, is given by

$$E(\theta) = \frac{n}{(n-1)!}(n\theta)^{n-1}e^{-n\theta} \tag{12–54}$$

Further, the fraction of tracer, $F(\theta)$, that has been in the reactor for less than time t is defined as follows:

$$F(\theta) = \int_0^t E(\theta)d\theta \approx \sum_0^n E(\theta)\Delta\theta \tag{12–55}$$

Thus, for a given dispersion number, the Peclet number can be used to determine the corresponding number of complete-mix reactors in series needed to achieve that dispersion number. Knowing the number of reactors in series, the value of $E(\theta)$ can be computed for various values of θ. The value of $F(\theta)$ can then be determined by summing the area under the $E(\theta)$ curve. The amount of flow that has been in the reactor for less than time θ can now be determined. Coupling normalized microorganism inactivation dose response data, obtained using a batch reactor, with the normalized detention time data, the actual performance for the chlorine contact basin can be estimated using what is known as a segregated flow model (SFM).

In the SFM approach, it is assumed that each block of fluid that enters a chlorine contact basin does not interact with other blocks of water. Thus, each block of water corresponds to an ideal plug flow reactor, each having a different residence time as defined by the value of $E(\theta)$, as given above. The reduction in organisms that would occur in each block of water can then be estimated for the period of time the block of water has remained in the chlorine contact basin. The overall performance is obtained by summing the results for each block of water. The SFM approach can be described as follows (Fogler, 1999):

Word statement

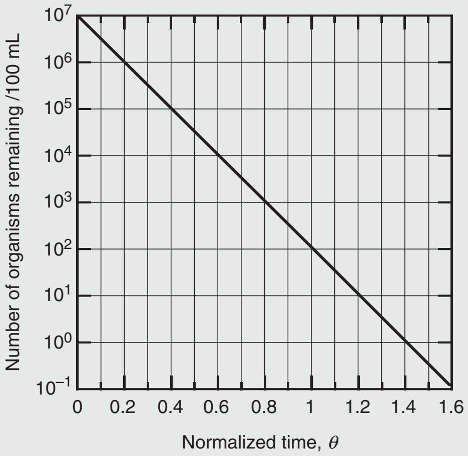
$$\begin{pmatrix}\text{Mean reduction in}\\\text{number of}\\\text{microorganisms}\\\text{spending between}\\\text{time } t \text{ and } t + dt \text{ in the}\\\text{chlorine contact basin}\end{pmatrix} = \begin{pmatrix}\text{number of}\\\text{microorganisms}\\\text{remaining after spending}\\\text{time } t \text{ in the chlorine}\\\text{contact basin based on}\\\text{batch test results}\end{pmatrix} \times \begin{pmatrix}\text{fraction of flow}\\\text{that remained}\\\text{in the chlorine}\\\text{contact basin}\\\text{between time } t\\\text{and } t + dt\end{pmatrix} \tag{12–56}$$

In equation form,

$$d\overline{N} = N(\theta) \times E(\theta)dt \tag{12–57}$$

The values of $N(\theta)$ and $E(\theta)$ are obtained from batch disinfection and tracer or dispersion prediction studies. Application of the above equations for predicting the hydraulic performance and the effluent microorganism concentration using the SFM are illustrated in Example 12–7 using the data from Example 12–6.

EXAMPLE 12–7 **Estimation of Performance of a Chlorine Contact Basin** Using the design information from Example 12–6, determine the fraction of flow that has not remained in the chlorine contact basin for the full hydraulic detention time. Determine how much larger the chlorine contact basin must be to be assured that 90 percent of the flow remains in the chlorine contact basin for the full design hydraulic detention time. Using the following normalized dose response data for an enteric virus, based on a τ value of 90 min and combined chlorine residual of 6 mg/L, estimate the performance of the chlorine contact basin in terms of the residual number of organisms remaining in the effluent.



Solution

1. Determine the number of complete mix reactors in series.
   a. From Example 12–6, the dispersion number at peak flow is

   $$d = 0.0139$$

   b. Using Eq. (12–53), the number of complete-mix reactors in series is

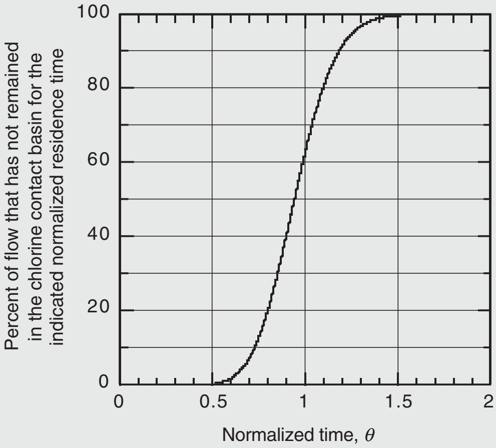
   $$\text{Number of reactors in series} = \frac{P_e}{2} = \frac{1}{2d} = \frac{1}{(2)\,0.0139} = 36$$

2. Determine the percentage of the flow that has been in the chlorine contact basin for less than the hydraulic detention time.
   a. Set up a computation table and compute $E(\theta)$ using Eq. (12–54) and the data given above.

   $$E(\theta) = \frac{n}{(n-1)!}(n\theta)^{n-1}e^{-n\theta}$$

Normalized time, θ	$E(\theta)$	$E(\theta) \times \Delta\theta \times 100$	Cumulative percent, $F(\theta)$
0.30	0.0000	0.000	0.000
0.40	0.0000	0.000	0.001
0.50	0.0046	0.046	0.046
0.60	0.0737	0.737	0.783
0.70	0.4435	4.435	5.218
0.80	1.2976	12.976	18.193
0.90	2.1878	21.878	40.071
1.00	2.3881	23.881	63.952
1.10	1.8337	18.337	82.290
1.20	1.0531	10.531	92.821
1.30	0.4739	4.739	97.560
1.40	0.1733	1.733	99.293
1.50	0.0530	0.530	99.822
1.60	0.0139	0.139	99.961
1.70	0.0031	0.032	99.992
1.80	0.0006	0.006	99.999
1.90	0.0001	0.001	100.00
2.00	0.0000	0.000	100.00

b. Plot the cumulative percent values from the above table.



c. From the computation table and graph given above, the percentage of the flow that has been in the chlorine contact basin for less than the hydraulic residence time is 64 percent. In fact, about 18 percent of the flow has left the chlorine contact basin before 80 percent of the nominal hydraulic detention time has elapsed.

3. Estimate how much larger the chlorine contact basin must be to be assured that 90 percent of the flow remains in the chlorine contact basin for the full hydraulic detention time. From the above graph, the size of the chlorine contact basin would have to be increased by a factor of 1.2.

4. Estimate the performance of the chlorine contact basin.

a. Set up a computation table to determine the number of organisms remaining in the effluent from the chlorine contact basin. The SFM approach, described above, will be used for this analysis. In effect, flow in each time period is treated as a batch reactor for the time interval it has remained in the reactor. The corresponding concentration of microorganisms leaving in any given volume of liquid is taken from the normalized dose response curve. The computation table for the application of the SFM is given below. The data in columns (1) and (3) are from the computation table prepared in Step 2 above, except that the data in column (3) are divided by 100. The data in column (2) are from the normalized dose response curve obtained as part of the process analysis for the design of the chlorine contact basin.

Normalized time, θ (1)	Number of organisms remaining, $N(\theta)$ MPN/100 mL (2)	$E(\theta) \times \Delta\theta$ (3)	Number of organisms remaining in effluent, ΔN MPN/100 mL (4)
0.30	300,000	0.00000	0.000
0.40	100,000	0.00000	0.00
0.50	30,000	0.00046	13.80
0.60	10,000	0.00737	73.70
0.70	3,000	0.04435	133.05
0.80	1,000	0.12976	129.76
0.90	300	0.21878	65.63

(*continued*)

(*Continued*)

Normalized time, θ (1)	Number of organisms remaining, $N(\theta)$ MPN/100 mL (2)	$E(\theta) \times \Delta\theta$ (3)	Number of organisms remaining in effluent, ΔN MPN/100 mL (4)
1.00	100	0.23881	23.88
1.10	30	0.18337	5.50
1.20	10	0.10531	1.05
1.30	3	0.04739	0.14
1.40	1	0.01733	0.02
1.50	0.3	0.00530	—
1.60	0.1	0.00139	—
1.70	0.03	0.00032	—
1.80	0.01	0.00006	—
1.90	0.003	0.00001	—
2.00	0.001	0.00000	—
Total		1.00000	446.53

b. The number of organisms in the effluent leaving the chlorine contact basin is:

$$\text{Organisms in effluent } N = \sum[N(\theta) \times E(\theta)\Delta\theta] = 447 \text{ MPN/100 mL}$$

c. By comparison, if it was assumed that the basin had performed as an ideal plug-flow reactor, then the organism concentration in the effluent would have been estimated to be 100 MPN/100 mL.

Comment The SFM method of analysis used to determine the number of organisms in the effluent is useful for estimating the performance of reactors with varying amounts of dispersion such as chlorine contact basins.

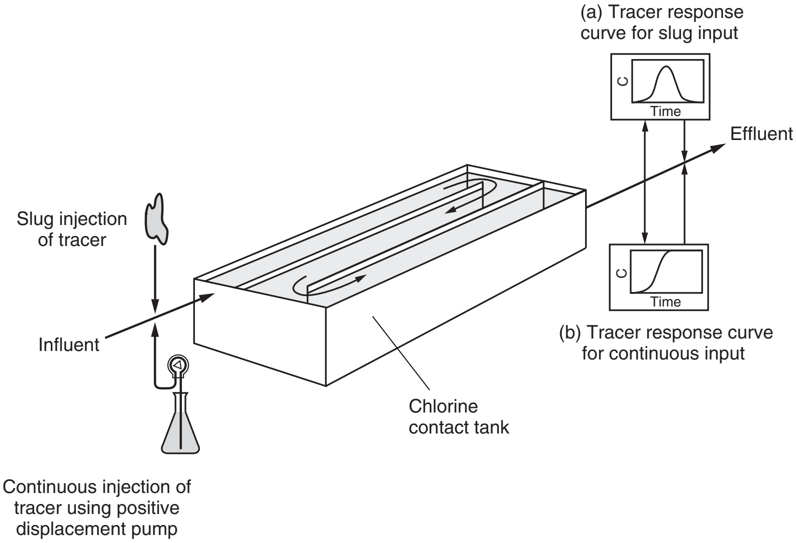
Assessing the Hydraulic Performance of Existing Chlorine Contact Basins

To be assured that a chlorine contact basin performs properly, most regulatory agencies request that tracer studies be conducted to determine the hydraulic characteristics of the chlorine contact basin. The types of tracers that have been used, the conduct of tracer tests, and analysis of tracer data are reviewed briefly below.

Compounds Used as Tracers. Tracers of various types are used commonly to assess the hydraulic performance of reactors used for wastewater disinfection. Dyes and chemicals that have been used successfully in tracer studies include congo red, fluorescein, fluosilicic acid (H_2SiF_6), hexafluoride gas (SF_6), lithium chloride (LiCl), Pontacyl Brilliant Pink B, potassium, potassium permanganate, rhodamine WT, sodium floride (NaF), and sodium chloride (NaCl). Pontacyl Brilliant Pink B (the acid form of rhodamine WT) is especially useful in the conduct of dispersion studies because it is not readily adsorbed onto surfaces. Because fluorescein, rhodamine WT, and Pontacyl Brilliant Pink B can be detected at very low concentrations using a flourometer, they are the dye tracers used most commonly in the evaluation of the performance of wastewater treatment facilities.

Figure 12–25

Schematic of setup for the conduct of a tracer study of a plug-flow chlorine contact basin using either a slug of tracer added to flow or a continuous input of tracer. The tracer response curve is measured continuously.



Conduct of Tracer Tests. In tracer studies, a tracer (i.e., a dye, most commonly) is introduced into the influent end of the reactor or basin to be studied (see Fig. 12–25). The time of its arrival at the effluent end is determined by collecting a series of grab samples for a given period of time or by measuring the arrival of a tracer using instrumental methods (see Fig. 12–25). The method used to introduce the tracer controls the type of response observed at the downstream end. Two types of tracer input methods are used, the choice depending on the reactor influent and effluent configurations. The first method involves the injection of a quantity of tracer (sometimes referred to a pulse or slug of tracer) over a short period of time. Initial mixing is usually accomplished with a static mixer or an auxiliary mixer. With the slug injection method it is important to keep the initial mixing time short relative to the detention time of the reactor being measured. The measured output is as described on Fig. 12–25(a). In the second method, a continuous step input of tracer is introduced until the effluent concentration matches the influent concentration. The measured response is as shown on Fig. 12–25(b). It should also be noted that another response curve can be measured after the dye injection has ceased and the dye in the reactor is flushed out.

Analysis of Tracer Test Response Curves. Tracer response curves, measured using a slug or continuous injection of a tracer, are known as C (concentration versus time) and F (fraction of tracer remaining in the reactor versus time) curves, respectively. The fraction remaining is based on the volume of water displaced from the reactor by the step input of tracer. The generalized results of three different dye tracer tests are shown on Fig. 12–26. As shown on Fig. 12–26, each of the three basins is subject to differing amounts of short circuiting. Length-to-width ratios (L/W) of at least 20 to 1 (preferably 40 to 1) and the use of baffles and guide vanes helps to minimize short circuiting. In some small plants, chlorine contact basins have been constructed of large diameter sewer pipe. The beneficial effect of using submerged baffles to improve the hydraulic efficiency of serpentine chlorine contact basins is illustrated on Fig. 12–23.

Tracer curves, such as shown on Figs. 12–25 and 12–26, are used to assess the hydraulic efficiency of chlorine contact basins. Parameters used to assess the hydraulic efficiency of chlorine contact basins are summarized in Table 12–19 and are illustrated on Fig. 12–27. As discussed previously, the mean, modal, and t_{10} times have been used to define the

Figure 12–26

Typical chlorine contact basin tracer response curves for three different basins with the same hydraulic detention time. The degree of short circuiting is illustrated clearly by the shape of the tracer curve.

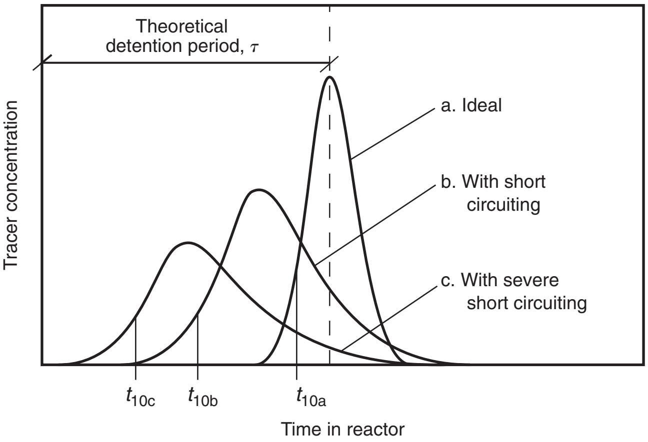


Table 12–19

Various terms used to describe the hydraulic performance of chlorine contact basins[a]

Term	Definition
τ	Theoretical hydraulic residence time (*V*, volume/*Q*, flowrate).
t_i	Time at which tracer first appears.
t_p	Time at which the peak concentration of the tracer is observed (mode).
t_g	Mean time to reach centroid of the residence time distribution (RTD) curve (see Appendix H).
t_{10}, t_{50}, t_{90}	Time at which 10, 50, and 90 percent of the tracer has passed through the reactor.
t_{90}/t_{10}	Morrill Dispersion Index, MDI (Morrill, 1932).
1/MDI	Volumetric efficiency as defined by Morrill (1932).
t_i/τ	Index of short circuiting. In an ideal plug-flow reactor, the ratio is 1, and approaches 0 with increased mixing.
t_p/τ	Index of modal retention time. Ratio will approach 1 in a plug-flow reactor, and 0 in a complete-mix reactor. For values of the ratio greater than or less than 1 the flow distribution in the reactor is not uniform.
t_g/τ	Index of average retention time. A value of 1 would indicate that full use is being made of the volume. A value of the ratio greater than or less than 1 indicates the flow distribution is not uniform.
t_{50}/τ	Index of mean retention time. The ratio t_{50}/τ is a measure of the skew of the RTD curve. A value of t_{50}/τ of less than 1 corresponds to an RTD curve that is skewed to the left. Similarly, for values greater than 1.0 the RTD curve is skewed to the right.
$t/\tau = \theta$	Normalized time, used in the development of the normalized RTD curve.
$\tau_{\Delta c} \approx \dfrac{\sum t_i C_i \Delta t_i}{\sum C_i \Delta t_i}$	Expression used to determine the mean hydraulic residence time, τ, if the concentration versus time tracer response curve is defined by a series of discrete time step measurements, where t_i is time at *i*th measurement C_i is concentration at *i*th measurement, and Δt_i is time increment about C_i.
$\sigma^2_{\Delta c} \approx \dfrac{\sum t_i^2 C_i \Delta t_i}{\sum C_i \Delta t_i} - (\tau_{\Delta c})^2$	Expression used to determine variance for a concentration versus time tracer response curve, which is defined by a series of discrete time step measurements.

[a] Adapted from Morrill (1932), Fair and Geyer (1954), and U.S. EPA (1986).

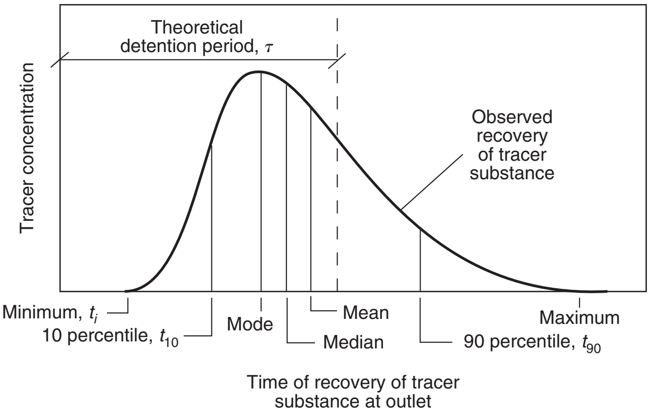


Figure 12–27

Definition sketch for the parameters used in the analysis of concentration versus time tracer response curves.

contact time in the CT relationship. The analysis of a tracer response curve is illustrated in Example 12–8. Additional details on the analysis of tracer response curves may be found in Appendix I and in Crittenden et al. (2012).

EXAMPLE 12–8 **Analysis of Tracer Data for an Existing Chlorine Contact Basin** The following tracer data have been gathered during a tracer test of a chlorine contact basin. During the tracer test, the total chlorine residual measured at the tank outlet was 4.0 mg/L. Using these data determine the mean hydraulic residence time (HRT), the variance, and the t_{10} time. Determine the CT values corresponding to the mean HRT and the t_{10} time. To further assess the performance of the chlorine contact basin, determine the Morrill Dispersion Index (MDI) and the corresponding volume efficiency (1/MDI) as defined in Table 12–19.

Time, min	Tracer concentration, μg	Time, min	Tracer concentration, μg
0.0	0.0000	144	9.333
16	0.000	152	16.167
40	0.000	160	20.778
56	0.000	168	19.944
72	0.000	176	14.111
88	0.000	184	8.056
96	0.056	192	4.333
104	0.333	200	1.556
112	0.556	208	0.889
120	0.833	216	0.278
128	1.278	224	0.000
136	3.722		

Solution

1. Determine the mean hydraulic residence time and variance for the tracer response data using equations given in Table 12–19.

a. Set up the required computation table. In setting up the computation table given below, the Δt value was omitted as it appears in both the numerator and in the denominator of the equations used to compute the residence time and the corresponding variance.

Time, t, min	Conc., C, μg	$t \times C$	$t^2 \times C$	Cumulative conc., μg	Cumulative percentage
88	0.000	0.000	0		
96	0.056	5.338	512.41	0.05	0.05
104	0.333	34.663	3604.97	0.39[a]	0.38[b]
112	0.556	62.227	6969.45	0.94	0.92
120	0.833	99.996	11,999.52	1.78	1.74
128	1.278	163.558	20,935.48	3.06	2.99
136	3.722	506.219	68,845.81	6.78	6.63
144	9.333	1343.995	193,535.31	16.11	15.75
152	16.167	2457.384	373,522.37	32.28	31.58
160	20.778	3324.480	531,916.80	53.06	51.91
168	19.944	3350.592	562,899.46	73.00	71.41
176	14.111	2483.536	437,102.34	87.11	85.22
184	8.056	1482.230	272,730.39	95.17	93.10
192	4.333	831.994	159,742.77	99.50	97.34
200	1.556	311.120	62,224.00	101.06	98.87
208	0.889	184.891	38,457.37	101.94	99.73
216	0.278	60.005	12,961.04	102.22	100.00
224	0.000	0.000			
Total	102.222	16,702.229	2,757,959.48		

[a] 0.056 + 0.333 = 0.39.

[b] (0.39/102.222) × 100 = 0.38.

b. Determine the mean hydraulic residence time.

$$\tau_{\Delta c} = \frac{\sum t_i C_i \Delta t_i}{\sum C_i \Delta t_i} = \frac{16{,}702.23}{102.22} = 163.4 \text{ min} = 2.7 \text{ h}$$

c. Determine the variance.

$$\sigma^2_{\Delta c} = \frac{\sum t_i^2 C_i \Delta t_i}{\sum C_i \Delta t_i} - (\tau_{\Delta c})^2 = \frac{2{,}757{,}959.48}{102.22} - (163.4)^2 = 280.5 \text{ min}^2$$

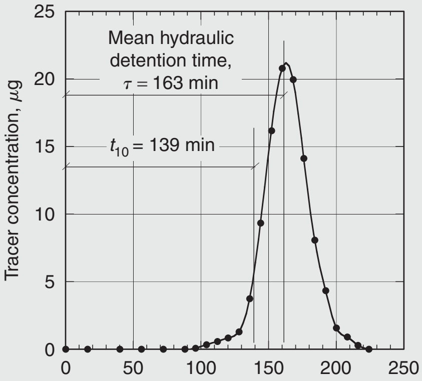
$$\sigma_{\Delta c} = 16.7 \text{ min}$$

d. Determine the t_{10} time using the cumulative percentage values. Because of the short time interval, a linear interpolation method can be used.

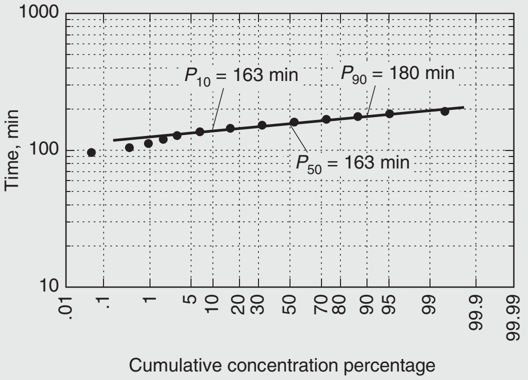
$$(15.75\% - 6.63\%)/(144 \text{ min} - 136 \text{ min}) = 1.14\%/\text{min}$$

$$t_{10} = 136 + (10\% - 6.63\%)/\,(1.14\%/\text{min}) = 139.0 \text{ min}$$

e. Identify the mean hydraulic residence and t_{10} times on the tracer curve.



2. Another technique that can be used to obtain the above times is to plot the cumulative concentration data on log-probability paper. Such a plot is also useful for determining the MDI. The required plot is given below.



The mean hydraulic retention and t_{10} times are read directly from the above plot.

t_{50} = 163 min

t_{10} = 139 min

3. Determine the corresponding CT values for the mean HRT and the t_{10} time determined above in Step 1.

CT (modal) = (4.0 mg/L)(163.4 min) = 654 mg·min/L

CT (t_{10}) = (4.0 mg/L)(139 min) = 556 mg·min/L

4. Determine the MDI and the corresponding volume efficiency using the expressions given in Table 12–19 and the values from the plot given in Step 2 above.

a. The Morrill Dispersion Index is

$$\text{Morrill Dispersion Index, MDI} = \frac{P_{90}}{P_{10}} = \frac{180}{139} = 1.30$$

c. The corresponding volumetric efficiency for the chlorine contact basin is

$$\text{Volumetric efficiency, \%} = \frac{1}{\text{MDI}} = \frac{1}{1.3} \times 100 = 77\%$$

Comment The variance computed in Step 1 is useful in assessing the dispersion in the chlorine contact basin (see Example I–1 in Appendix I; Crittenden et al., 2012). The CT values, based on the modal and t_{10} times, exceed the CT value of 450 mg·min/L required in California. It is important to note that if the tracer curve is very skewed, it may not be possible to achieve effective disinfection, especially if the t_{10} value is used. Thus, the design of a chlorine contact basin to achieve near plug flow is of critical importance. The MDI value (1.30) is characteristic of a chlorine contact basin with low dispersion. A MDI value below 2.0 has been established by the U.S. EPA as an effective design (U.S. EPA, 1986). Similarly, the volumetric efficiency is high, signifying near-ideal plug flow with a small amount of axial dispersion.

Outlet Control and Chlorine Residual Measurement

The flow at the end of the contact basin may be metered by means of a V-notch or rectangular weir or a Parshall flume. Control devices for chlorination in direct proportion to the flowrate may be operated from these meters or from the main plant flowmeter. Final determination of the success of a chlorine contact basin must be based on samples taken and analyzed to correlate chlorine residual and the MPN of coliform or other indicator organisms. When the chlorine residual is used for chlorinator control, chlorine residual sample pumps should be located at the front end of the first pass of the contact basin after rapid mixing to allow time for the initial demand to be met. Chlorine residual measurements should also be taken at the contact basin outlet to ensure compliance with the regulatory agency requirements. In the event that no chlorine contact basin is provided and the effluent pipeline is used for contact, the sample can be obtained at the point of chlorination, held for the theoretical detention time, and the residual determined. The sample is then dechlorinated and subsequently analyzed for bacteria using standard laboratory procedures.

Chlorine Storage Facilities

Storage and handling facilities for chlorine can be designed with the aid of information developed by The Chlorine Institute. Although all the safety devices and precautions that must be designed into the chlorine handling facilities are too numerous to mention, the following are fundamental:

1. Chlorine gas is toxic and very corrosive. Adequate exhaust ventilation with intakes at floor level should be provided because chlorine gas is heavier than air. The ventilation system should be capable of at least 60 air changes per hour with the exhaust directed vertically upwards.
2. Chlorine storage and chlorinator equipment rooms should be walled off from the rest of the plant and should be accessible only from the outdoors. A fixed glass viewing window should be included in an inside wall to check for leaks before entering the equipment rooms. Fan controls should be located at the room entrance. Air purifying respirators or self-contained breathing apparatus (SCBA) should also be located nearby in protected but readily accessible locations.
3. Temperatures in the scale and chlorinator areas should be controlled to avoid freezing.
4. Dry chlorine liquid and gas can be handled in black steel piping, but chlorine solution is highly corrosive and should be handled in Schedule 80 polyvinylchloride (PVC) piping.
5. Adequate storage of standby cylinders should be provided. The amount of storage should be based on the availability and dependability of the supply and the quantities used. Cylinders in use are set on scales and the loss of weight is used as a positive record of chlorine dosage.

6. Chlorine cylinders should be protected from direct sunlight in warm climates to prevent overheating of the full cylinders.
7. In larger systems, chlorine residual analyzers should be provided for monitoring and control purposes to prevent the under- or over-dosing of chlorine.
8. The chlorine storage and feed facilities should be protected from fire hazards. In addition, chlorine leak detection equipment should be provided and connected to an alarm system.

Chemical Containment Facilities

In 1991, the International Conference of Building Officials revised Article 80: Hazardous Materials of the Uniform Fire Code (UFC). The revisions were extensive and covered a variety of issues. The provisions of the new code apply to new facilities and to old facilities as well, if it is determined that they constitute a distinct hazard to life or property. The new code provisions are contained in the following divisions apply to the chemicals used for disinfection: I General Provisions, II Classification by Hazard, III Storage Requirements, and IV Dispensing, Use, and Handling. The classification of hazardous materials used for wastewater disinfection are summarized in Table 12–20. Storage requirements include provisions for spill control and containment, ventilation, treatment, and storage. Emergency scrubbing systems, usually using a caustic solution, are also required to neutralize leaking chlorine and sulfur dioxide gas. Many of the same topics contained in the storage requirements section also apply to the dispensing, use, and handling. Hazardous material management, provision for standby power, security, and alarms are among the additional topics covered. It is extremely important to review current UFC regulations in the design of new facilities and in the refurbishing of existing facilities. Furthermore, the U.S. EPA and many states, as well as OSHA, have implemented chemical safety regulations requiring formal hazard reviews, air dispersion modeling of release scenarios, and emergency response preparedness.

Dechlorination Facilities

Dechlorination of chlorinated effluents is accomplished most commonly using sulfur dioxide. Where granular activated carbon is used for the removal of residual organic material, the carbon can also be used for the dechlorination of chlorinated effluents.

Sulfur Dioxide. The principal elements of a sulfur dioxide dechlorination system include the sulfur dioxide containers, scales, sulfur dioxide feeders (sulfonators), solution

Table 12–20

Classification of hazardous materials used in wastewater disinfection

Category	Typical chemicals
Physical Hazards	
Compressed gases	Oxygen, ozone, chlorine, ammonia, sulfur dioxide
Oxidizers	Oxygen, ozone, chlorine, hydrogen peroxide, acids, chlorine
Health Hazards	
Highly toxic material	Chlorine, chlorine dioxide, ozone, acids, bases
Corrosives	Acids, bases, chlorine, sulfur dioxide, ammonia, hypochlorite, sodium bisulfite
Other health hazards—irritants, suffocating, etc.	Chlorine, sulfur dioxide, ammonia

injectors, diffuser, mixing chamber, and interconnecting piping. For facilities requiring large withdrawal rates of SO_2, evaporators are used because of the low vaporization pressure of 240 kN/m^2 at 21°C (35 lb$_f$/in.^{2} at 70°F). Common sulfonator sizes are 216, 864, and 3409 kg/d (475, 1900, and 7500 lb/d). The key control parameters of this process are (1) proper dosage based on precise (amperometric) monitoring of the combined chlorine residual and (2) adequate mixing at the point of application of sulfur dioxide.

Sodium Bisulfite. Sodium bisulfite is available as white powder, a granular material, or as a liquid. The liquid form is used most commonly for dechlorination at wastewater treatment facilities. Although available in solution strengths up to 44 percent, a 25 percent solution is most typical to minimize viscosity increases during cold weather. In most applications, a diaphragm type pump is used to meter the sodium bisulfite. The reaction between sodium bisulfite and chlorine residual was presented previously [see Eqs. (12–42 and 12–43)]. Based on Eq. (12–42), each mg/L of chlorine residual requires about 1.46 mg/L of sodium bisulfite and 1.38 mg/L of alkalinity as $CaCO_3$ will be consumed.

Granular Activated Carbon. The common method of activated carbon treatment used for dechlorination is downflow through either an open or enclosed vessel. The activated carbon system, while significantly more costly than other dechlorination approaches, may be appropriate when activated carbon is being used as an advanced wastewater treatment process. Typical hydraulic loading rates and contact times for activated carbon columns used for dechlorination are 3000 to 4000 L/m^2·d and 15 to 25 min, respectively.

12–7 DISINFECTION WITH OZONE

Although historically used primarily for the disinfection of water, recent advances in ozone generation and solution technology have made the use of ozone economically more competitive for the disinfection of treated wastewater. Further, interest in the use of ozone for disinfection has also been renewed because of its ability to reduce or eliminate trace constituents. Ozone can also be used in water reuse applications for the removal of soluble refractory organics, in lieu of the carbon adsorption process. The characteristics of ozone, the chemistry of ozone, the generation of ozone, an analysis of the performance of ozone as a disinfectant, and the application of the ozonation process are considered in the following discussion.

Ozone Properties

Ozone is an unstable gas produced when oxygen molecules dissociate into atomic oxygen. Ozone can be produced by electrolysis, photochemical reaction, and radiochemical reaction by electrical discharge. Ozone is often produced by ultraviolet light and lightning during a thunderstorm. The electrical discharge method is used for the generation of ozone in water and wastewater disinfection applications. Ozone is a blue gas at normal room temperatures, and has a distinct odor. Ozone can be detected at concentrations of 2×10^{-5} to 1×10^{-4} g/m^3 (0.01 to 0.05 ppm$_v$, by volume). Because ozone has an odor, it can usually be detected by the human olfactory system before health concerns develop. The stability of ozone in air is greater than it is in water, but in both cases is on the order of minutes. Gaseous ozone is explosive when the concentration reaches about 240 g/m^3 (20 percent weight in air). The properties of ozone are summarized in Table 12–21. The solubility of ozone in water is governed by Henry's law. Typical values of Henry's constant for ozone are presented in Table 12–22.

Table 12–21

Properties of ozone[a]

Property	Unit	Value
Molecular weight	g	48.0
Boiling point	°C	−111.9 ± 0.3
Freezing point	°C	−192.5 ± 0.4
Latent heat of vaporization at 111.9°C	kJ/kg	14.90
Liquid density at −183°C	kg/m^3	1574
Gaseous density at 0°C and 1 atm	g/mL	2.154
Solubility in water at 20.0°C	mg/L	12.07
Vapor pressure at −183°C	kPa	11
Vapor density compared to dry air at 0°C and 1 atm	unitless	1.666
Specific volume of vapor at 0°C and 1 atm	m^3/kg	0.464
Critical temperature	°C	−12.1
Critical pressure	kPa	5532.3

[a] Adapted in part from Rice (1996), U.S. EPA (1986), White (1999).

Ozone Chemistry

Some of the chemical properties displayed by ozone may be described by its decomposition reactions which are thought to proceed as follows:

$$O_3 + H_2O \rightarrow HO_3^+ + OH^- \qquad (12\text{–}58)$$

$$HO_3^+ + OH^- \rightarrow 2HO_2 \qquad (12\text{–}59)$$

$$O_3 + HO_2 \rightarrow HO\cdot + 2O_2 \qquad (12\text{–}60)$$

$$HO\cdot + HO_2 \rightarrow H_2O + O_2 \qquad (12\text{–}61)$$

The dot (·) that appears next to the hydroxyl and other radicals is used to denote the fact that these species have an unpaired electron. The free radicals formed, HO_2 and HO·, have great oxidizing powers and are active in the disinfection process. These free radicals also possess the oxidizing power to react with other impurities in aqueous solutions.

Table 12–22

Values of Henry's constant for ozone[a]

Temperature, °C	Henry's constant, atm/mole fraction
0	1940
5	2180
10	2480
15	2880
20	3760
25	4570
30	5980

[a] U.S. EPA (1986).

Table 12–23

Impact of wastewater constituents on the use of ozone for wastewater disinfection

Constituent	Effect
BOD, COD, TOC, etc.	Organic compounds that comprise the BOD and COD can exert an ozone demand. The degree of interference depends on their functional groups and their chemical structure
NOM (natural organic matter)	Affects the rate of ozone decomposition and the ozone demand
Oil and grease	Can exert an ozone demand
TSS	Increases ozone demand and shielding of embedded bacteria
Alkaliniy	No or minor effect
Hardness	No or minor effect
Ammonia	No or minor effect, can react at high pH
Nitrite	Oxidized by ozone
Nitrate	Can reduce effectiveness of ozone
Iron	Oxidized by ozone
Manganese	Oxidized by ozone
pH	Effects the rate of ozone decomposition
Industrial discharges	Depending on the constituents, may lead to a diurnal and seasonal variations in the ozone demand
Temperature	Affects the rate of ozone decomposition

Effectiveness of Ozone as a Disinfectant

Ozone is an extremely reactive oxidant and it is generally believed that bacterial kill through ozonation occurs directly because of cell wall disintegration (cell lysis). The impact of the wastewater characteristics on ozone disinfection is reported in Table 12–23. The presence of oxidizable compounds will cause the ozone inactivation curve to have a shoulder affect as discussed previously for chlorine (see Fig. 12–6). Tailing will also occur in the presence of residual floc particles.

Ozone is also a very effective viricide and is generally believed to be more effective than chlorine. The relative germicidal effectiveness of ozone for the disinfection of different microorganisms was presented previously in Table 12–5. Ozonation does not increase dissolved solids and disinfection effectiveness is not affected by ammonium ion. Although ozone is not necessarily impacted by water pH, ozone residual is more stable in acidic environments and less stable in waters with caustic pH. Therefore, it is typically easier to achieve disinfection when pH is reduced than when pH is greater than neutral. For these reasons, ozonation is considered as an alternative to either chlorination or hypochlorination, especially where dechlorination may be required and high purity oxygen facilities are available at the treatment plant.

Modeling the Ozone Disinfection Process

In practice, an ozone contactor will be comprised of three or more compartments or chambers (see Fig. 12–31 in discussion of ozone reactor characteristics). Water depth is typically 4.6 to 6 m (15 to 20 ft). Ozone is typically added to first or the first and second of the compartments and the remaining compartments serve as contact compartments. The detention time in the first compartment, used to meet the immediate ozone demand (i.e., peroxidation), is short, typically 2 to 4 min. Contact time in the subsequent compartments will vary from 3 to 10 min depending on the rate of ozone utilization.

Over the years a number of different mathematical relationships have been developed to model the disinfection process with ozone. The most common of these is Eq. (12–6) repeated here for convenience.

$$\log \frac{N_t}{N_o} = -\Lambda_{\text{base }10}(\text{CT}) \tag{12–6}$$

Values for the coefficient of specific lethality, Λ, are given in Table 12–11. Of the values given in Table 12–11, those for viruses, *Cryptosporidium,* and *Giardia* cysts are the most reliable as they can be derived from the published U.S. EPA CT tables (U.S. EPA, 2003a).

Because ozone is sparingly soluble, bench and/or pilot-scale studies (see Fig 12–28), using the same retention time as the full-scale reactor, are conducted to assess (1) the immediate ozone demand, (2) the amount of ozone that can be transferred to the liquid, and (3) the ozone decay profile along the reactor. The information gathered is used to determine the CT value and the level of inactivation that can be expected in the full-scale reactor. The amount of ozone that is utilized or transferred to the liquid is computed using the following expression:

$$\text{Ozone dose mg/L} = \frac{Q_g}{Q_l}(C_{g,\text{in}} - C_{g,\text{out}}) \tag{12–62}$$

where Q_g = gas flowrate, L/min
Q_l = liquid flowrate, L/min
$C_{g,\text{in}}$ = concentration of ozone in feed gas, mg/L
$C_{g,\text{out}}$ = concentration of ozone in off-gas, mg/L

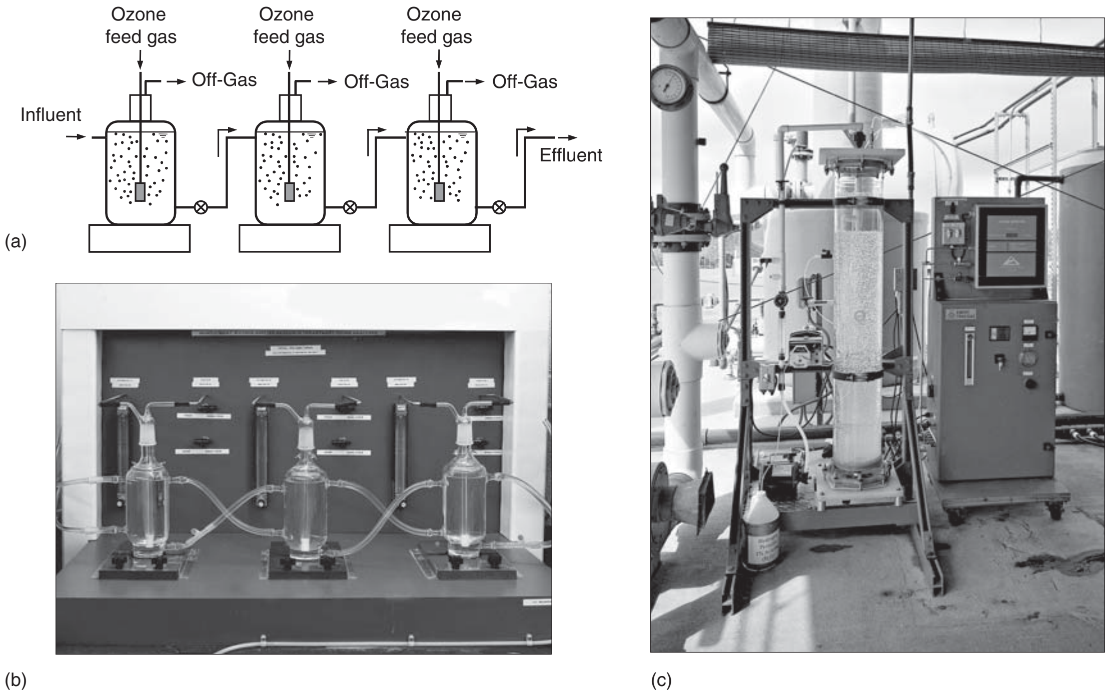


Figure 12–28

Typical ozone test reactors: (a) schematic of setup for bench-scale ozone testing, (b) view of bench-scale ozone test reactors, and (c) pilot-scale ozone test reactor.

The bench and/or pilot-scale reactors are operated in a continuous mode at various ozone concentrations. Once steady-state has been reached, both the water and ozone dosing are stopped, and the ozone decay is observed with time. The continuous operation simulates the compartments where ozone is being added and accounts for the immediate demand of the wastewater. The decay curve is used to estimate the residual ozone concentrations in the downstream compartments. Analysis of bench-scale ozone test data is illustrated in Example 12–9. Computation of the CT for an ozone contactor is illustrated in Example 12–10.

EXAMPLE 12–9 **Estimate the Immediate Ozone Requirement for a Typical Secondary Effluent** Estimate the immediate ozone demand from the following bench scale steady-state ozone test data collected at 20°C. Determine the first-order equation for the corresponding decay data.

The steady-state test results are

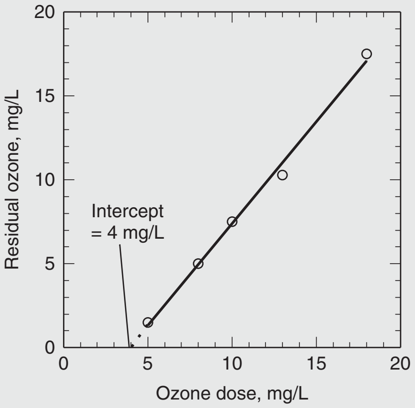
Test	Ozone dose, mg/L	Ozone residual, mg/L
1	5	1.5
2	8	5.0
3	10	7.5
4	13	10.3
5	18	17.5

The corresponding decay data are

Time	Ozone residual, mg/L
0	4.02
4	2.58
7	1.72
10	1.28

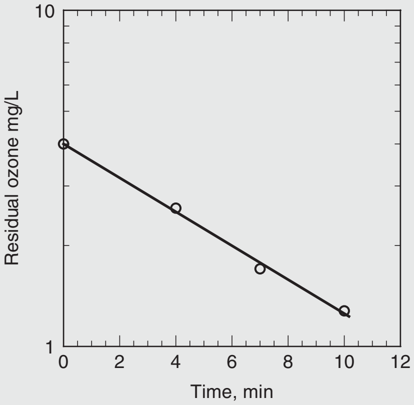
Solution

1. Plot the bench scale steady-state data and determine the immediate ozone demand.
   a. The required plot is given below.



b. Determine the immediate ozone demand.
The immediate ozone demand corresponds to the intercept on the *x*-axis. From the above plot the value is equal to 4 mg/L.

2. Plot the bench scale steady-state decay data and determine an appropriate first-order equation.
   a. The required plot is given below.



   b. The corresponding first-order equation is

$$C_{\text{residual ozone}} = (4.0\text{ mg/L})e^{-0.116t}$$

where t = contact time, min

Comment The existence of an immediate ozone demand is the reason why the first compartment where ozone is added is generally not considered in establishing the CT value for an ozone contactor. The CT value can be determined from the decay curve by considering the reactor as a whole or by considering each compartment individually.

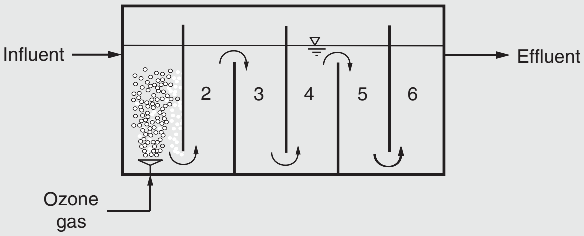
Required Ozone Dosages for Disinfection

The required ozone dosage for disinfection can be estimated by considering (1) the initial ozone demand, based on the results of a bench scale test as illustrated above, and (2) the corresponding decay curve. Computation of the CT value for an existing ozone reactor is illustrated in Example 12–10. The ozone dosage required to meet the initial demand depends on the constituents in the wastewater and is site specific and, in most cases, will be significantly greater than the dose required for disinfection of the coliform group of microorganisms. Based on the coefficient of lethality, Λ, values given in Table 12–11, it is clear that the ozone dosages required to meet total coliform standards are a fraction of those required for the inactivation of *Cryptosporidium* and *Giardia lamblia*. In most cases, bench and/or pilot-scale studies (see Fig. 12–28) will need to be conducted to establish the required dosage ranges.

Estimation of the CT Value

In water treatment, the CT value can be computed in four different ways as defined in the *LT2ESWTR Toolbox Guidance Manual* (U.S. EPA, 2010). The t_{10} approach is as follows. The CT value for an ozone contactor can be estimated as the summation of the average ozone concentration in each compartment, with the exception of the first compartment, times the detention time in each compartment. The time in each compartment is based on the t_{10} time (see Table 12–19) measured across all of the compartments divided by the proportional volume of each compartment. The first compartment is omitted, as noted above, as it is used to meet the immediate ozone demand and does not contribute to disinfection. Additional details may be found in the *LT2ESWTR Toolbox Guidance Manual* (U.S. EPA, 2010).

EXAMPLE 12–10 **Estimate the CT Value for an Ozone Contactor and Corresponding Log Reduction in Cryptosporidium** Estimate the CT value for the ozone contactor shown below and the log-reduction of *Cryptosporidium* that can be achieved. If the log reduction is greater than 2, estimate the number of reaction compartments that would be needed to achieve a 2-log reduction. The detention time of each compartment of the ozone contactor is 3 min. Assume the decay curve developed in Example 12–9 is applicable and that the t_{10}/t ratio is 0.6. The observed ozone concentration at the end of the first compartment is 4 mg/L.



Solution

1. Using the decay curve from Example 12–9, determine the residual ozone concentration at the end of each compartment starting with compartment 2.

Compartment no.	Ozone residual, mg/L
2	2.82[a]
3	1.99
4	1.41
5	0.99
6	0.88

[a] $C = (4\ \text{mg/L})e^{-0.116 \times t}$

$= (4\ \text{mg/L})e^{-0.116 \times 3} = 2.82\ \text{mg/L}$

2. Using the data from Step 1, determine the CT value for the ozone contactor, noting that the theoretical detention time in each reactor is 3 min and the t_{10}/t ratio is 0.6.

$$\text{CT} = \sum_{i=2}^{b} C_i T_i = [(2.82 + 1.99 + 1.41 + 0.99 + 0.88)\ \text{mg/L}](3\ \text{min} \times 0.6)$$

$$= 14.6\ \text{mg}\cdot\text{min/L}$$

3. Estimate the log reduction that can be achieved for *Cryptosporidium* using the information given in Table 12–5.
From Table 12–5, the estimated log reduction that can be achieved is 3 plus logs.

4. Check the log reduction that can be achieved using Eq. 12–6.
   a. The coefficient of specific lethality for cryptosporidium from Table 12–11 is 0.256 L/mg•min (base 10).
   b. The log reduction using Eq. (12–6) is

$$\log \frac{N_t}{N_o} = -\Lambda_{\text{base }10}\text{CT} = (0.256\ \text{L/mg}\cdot\text{min})(13.7\ \text{mg}\cdot\text{min/L}) = 3.74$$

5. Estimate the number of compartments that would be needed to achieve a 2-log reduction in *Cryptosporidium*.
   a. Assuming only one reaction compartment is needed, determine the CT value.
   CT = (2.82 mg/L)(3 min × 0.6) = 5.1 < 8.25 (from Table 12–5), hence an additional compartment is needed.
   b. Assuming two reaction compartments are needed, determine the CT value.
   CT = [(2.82 + 1.99) mg/L](3 min × 0.6) = 8.6 > 8.25, hence two compartments should be used.

Byproduct Formation and Control

As with chlorine, the formation of unwanted DBPs is one of the problems associated with the use of ozone as a disinfectant. The formation and control of DBPs when using ozone are considered in the following discussion.

Formation of DBPs Using Ozone for Disinfection. One advantage of ozone is that it does not form chlorinated DBPs such as THMs and HAAs (see Table 12–14). Ozone does, however, form other DBPs (see Table 12–24) including aldehydes, various acids, and aldo- and keto-acids when significant amounts of bromide are not present. In the presence of bromide, the following DBPs may also be produced: bromoform, brominated acetic acid, bromopicrin, brominated acetonitriles, cyanogen bromide, and bromate (see Table 12–24). On occasion, hydrogen peroxide can also be generated. The specific amounts and the relative distribution of compounds depend on the nature of the precursor compounds that are present. Because the chemical characteristics of wastewater vary from location to location, pilot testing will be required to assess the effectiveness of ozone as a disinfectant.

Control of DBP Formation Using Ozone for Disinfection. Because the nonbrominated compounds appear to be readily biodegradable, they can be removed by passage through a biologically active filter, by soil application, or by other biologically active processes. The removal of inorganic, brominated DBPs is more complex. Bench and pilot-scale testing is recommended to determine if brominated DBPs will be problematic. If it is expected that brominated DBPs will remain problematic, it may be appropriate to investigate an alternative means of disinfection such as by UV irradiation.

Environmental Impacts of Using Ozone

It has been reported that ozone residuals can be acutely toxic to aquatic life (Ward and DeGraeve, 1976). Several investigators have reported that ozonation can produce some toxic mutagenic and/or carcinogenic compounds. These compounds are usually unstable, however, and are present only for a matter of minutes in the ozonated water. White (1999) has reported that ozone destroys certain harmful refractory organic substances, such as humic

Table 12–24

Representative disinfection byproducts resulting from the ozonation of wastewater containing organic and selected inorganic constituents[a]

Class	Representative compounds
Acids	Acetic acids Formic acid Oxalic acid Succinic acid
Aldehydes	Acetaldehyde Formaldehyde Glyoxal Methyl glyoxal
Aldo-and ketoacids	Pyruvic acid
Brominated byproducts[b]	Bromate ion Bromoform Brominated acetic acids Bromopicrin Brominated acetonitriles Cyanogen bromide
Other	Hydrogen peroxide

[a] Adapted, in part, from U.S. EPA (1999a, 2002).

[b] The bromide ion must be present to form brominated byproducts.

acid (precursor of trihalomethane formation) and malathion. Whether toxic intermediates are formed during ozonation depends on the ozone dose, the contact time, and the nature of the precursor compounds. White (1999) has also reported that ozone treatment ahead of chlorination for disinfection purposes reduces the likelihood for the formation of THMs.

Ozone residual quenching is still required to meet OSHA indoor and outdoor ambient air quality standards. Ozone quenching of the off-gas is also required prevention or limit the corrosion of downstream piping and equipment. Where required, hydrogen peroxide, sodium bisulfite, and calcium thiosulfate have been used to quench residual ozone.

Other Benefits of Using Ozone

An additional benefit associated with the use of ozone for disinfection is that the dissolved oxygen concentration of the effluent will be elevated to near saturation levels as ozone rapidly decomposes to oxygen after application. The increase in oxygen concentration may eliminate the need for reaeration of the effluent to meet required dissolved oxygen water quality standards.

Ozone Disinfection Systems Components

A complete ozone disinfection system, as illustrated on Fig. 12–29, is comprised of the following components: (1) facilities for the preparation of the feed gas, (2) power supply, (3) the ozone generation facilities, (4) two alternative types of facilities for contacting the ozone with the liquid to be disinfected (inline or sidestream), and (5) facilities for the destruction of ozone in the off-gas (Rice, 1996; Rakness, 2005). Additional details on the design of ozone systems and related components may be found in a recently published book by Rakness (2005).

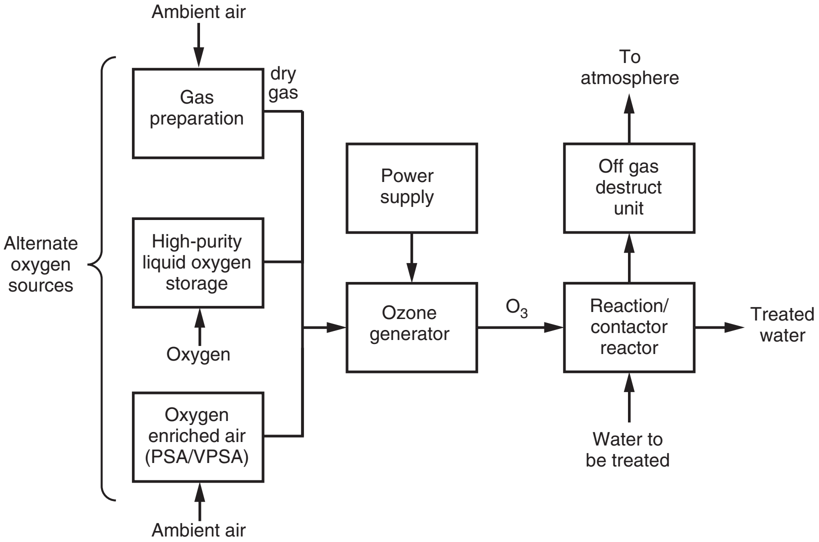


Figure 12–29

Schematic flow diagram for complete ozone disinfection system with alternative air sources. (Adapted from U.S. EPA, 1986.)

Preparation of Feed Gas. Ozone can be generated using air, high purity oxygen, or oxygen enriched air. If air is used for ozone generation, it must conditioned by removing the moisture and particulate matter before being introduced into the ozone generator. The following steps are involved in conditioning the air: (1) gas compression, (2) air cooling and drying, and (3) air filtration. If high purity oxygen is used, the conditioning steps are not required. The liquid oxygen (LOX) supply is stored onsite and is either generated onsite or trucked in as needed. In the oxygen enriched air system, high purity oxygen is generated onsite with a vacuum pressure swing adsorption (AVPAS) system or pressure swing adsorption (PSA) system for smaller treatment plants. Both oxygen generation systems have facilities for adsorbing moisture, which can damage the ozone generator dielectrics, and for the removal of hydrocarbons and nitrogen to enhance the purity of the oxygen. The choice of feed gas is influenced by the local cost of high purity oxygen.

Power Supply. The major requirement for power is for the production of ozone from oxygen. Additional power is required for preparation of the feed gas, contacting the ozone, destroying the residual ozone, and for the controls, instrumentation, and monitoring facilities. The energy requirements for the major components are reported in Table 12–25.

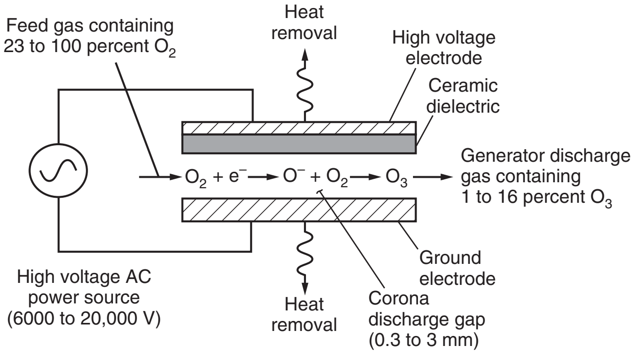
Table 12–25

Typical energy requirements for the application of ozone

Component	kWh/lb ozone	kWh/kg ozone
Air preparation (compressor and dryers)	2–3	4.4–6.6
Ozone generation		
Air feed	6–9	13.2–19.8
Pure oxygen	3–6	6.6–13.2
Ozone contacting	1–3	2.2–6.6
All other uses	0.5–1	1.2–2.2

Figure 12–30

Schematic detail of the generation of ozone. (Adapted from U.S. EPA, 1986.)



Ozone Generation. Because ozone is chemically unstable, it decomposes to oxygen very rapidly after generation, and thus must be generated onsite. The most efficient method of producing ozone today is by electrical discharge. Ozone is generated either from air or high purity oxygen when a high voltage is applied across the gap of narrowly spaced electrodes (see Fig. 12–30). The high energy corona created by this arrangement dissociates one oxygen molecule, which reforms with two other oxygen molecules to create two ozone molecules. The gas stream generated by this process from air will contain about 1 to 3 percent ozone by weight, and from pure oxygen about 8 to 12 percent ozone. Ozone concentrations up to 12 percent can now be generated with the latest medium frequency ozone generators.

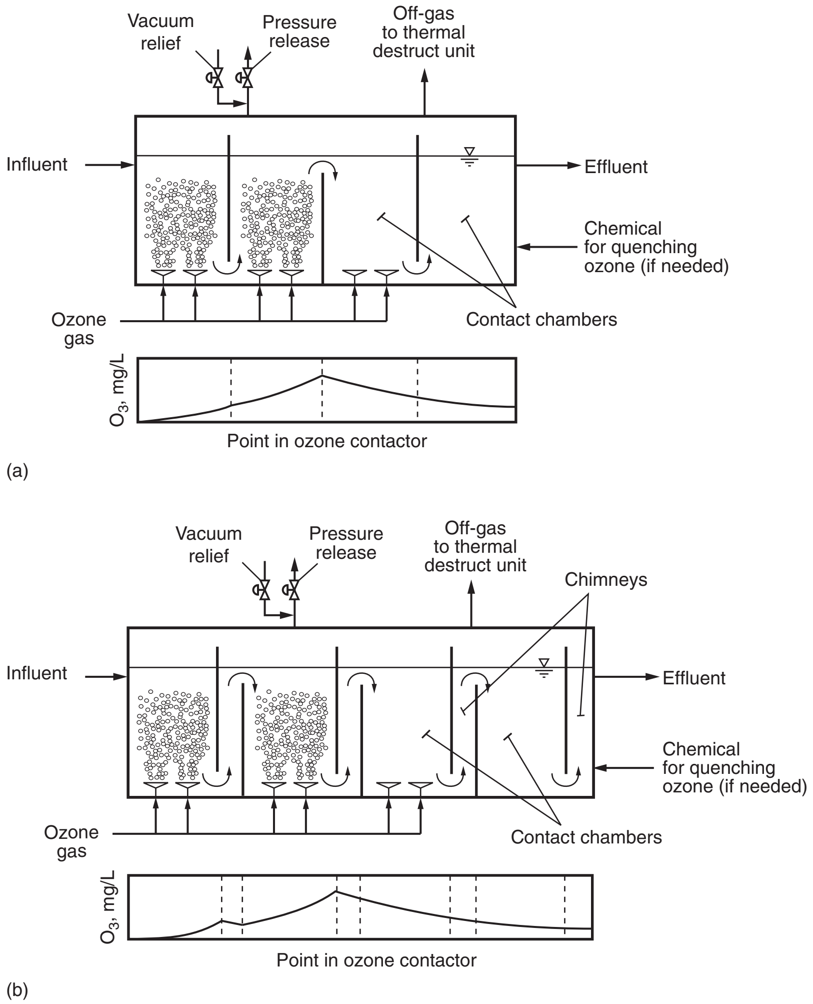
Inline Ozone Contact/Reaction Reactors. The concentration of ozone generated from either air or pure oxygen is so low that the transfer efficiency to the liquid phase is an extremely important economic consideration. To optimize ozone dissolution, deep and covered contact chambers are normally used. Two four-compartment ozone contact reactors are shown schematically on Fig. 12–31 without and with chimneys. The chimneys shown on Fig. 12–31(b) are used to enhance the countercurrent flow within the reactor. The chimneys also provide locations for ozone residual sampling.

Ozone is introduced by means of porous diffusers or injectors into the bottom of the first and second, and in some cases, the third chamber. Fast ozone reactions occur in the first chamber. The combined water-ozone mixture then enters the second chamber where slower reactions occur. Disinfection generally occurs in the second chamber. The third and fourth chambers are used to complete the slow reactions and to allow the ozone to decompose. The first and second chambers are identified as the *reaction* chambers. The third and fourth chambers, without ozone addition, are known as the *contact* chambers. The number of chambers used will depend on the treatment objectives.

Sidestream Ozone Contact/Reaction System. With the ability to generate higher concentrations of ozone (e.g., 10 to 12 percent), sidestream injection of ozone (see Fig. 12–32) is now a viable alternative to the use of porous diffusers in deep tanks as described above. As shown on Fig. 12–32, the ozone injection system is independent of the ozone contactor. The ozone is injected under pressure through a Venturi injector. Two sidestream configurations are used: (1) one with the inclusion of a degas vessel and (2) one without. The purpose of the degas vessel is to minimize the DO level in the water which has been ozonated and (2) to minimize the number of gas bubbles in the downstream pipe which serves as a reactor. The pipeline into which the ozonated water is injected also serves as a reactor prior to the discharge into the contactor (Rakness, 2005).

Figure 12–31

Schematic of typical four compartment ozone contactors: (a) without chimneys and (b) with chimneys. The chimneys in (b) are used to enhance the counter current flow through the reactor.



Destruction of Off-Gases. The off-gases from the contact chamber and the degas vessel must be treated to destroy any remaining ozone as it is an extremely irritating and toxic gas. Off-gas is destroyed to a concentration of <0.1 ppm$_v$. The product formed by destruction of the remaining ozone is pure oxygen which can be recycled if pure oxygen is being used to generate the ozone.

12–8 OTHER CHEMICAL DISINFECTION METHODS

Because of the concerns over the effectiveness of disinfection processes and concern over the formation of DBPs, ongoing research is continuing into the evaluation of alternative disinfection methods. The use of peracetic acid, peroxone, sequential chlorination, and combined disinfection processes are introduced and considered briefly in this section. Pasteurization, a physical method, is considered in Sec. 12–10. Because research on these and other disinfection methods is ongoing, current conference proceedings and literature must be consulted for the latest findings.

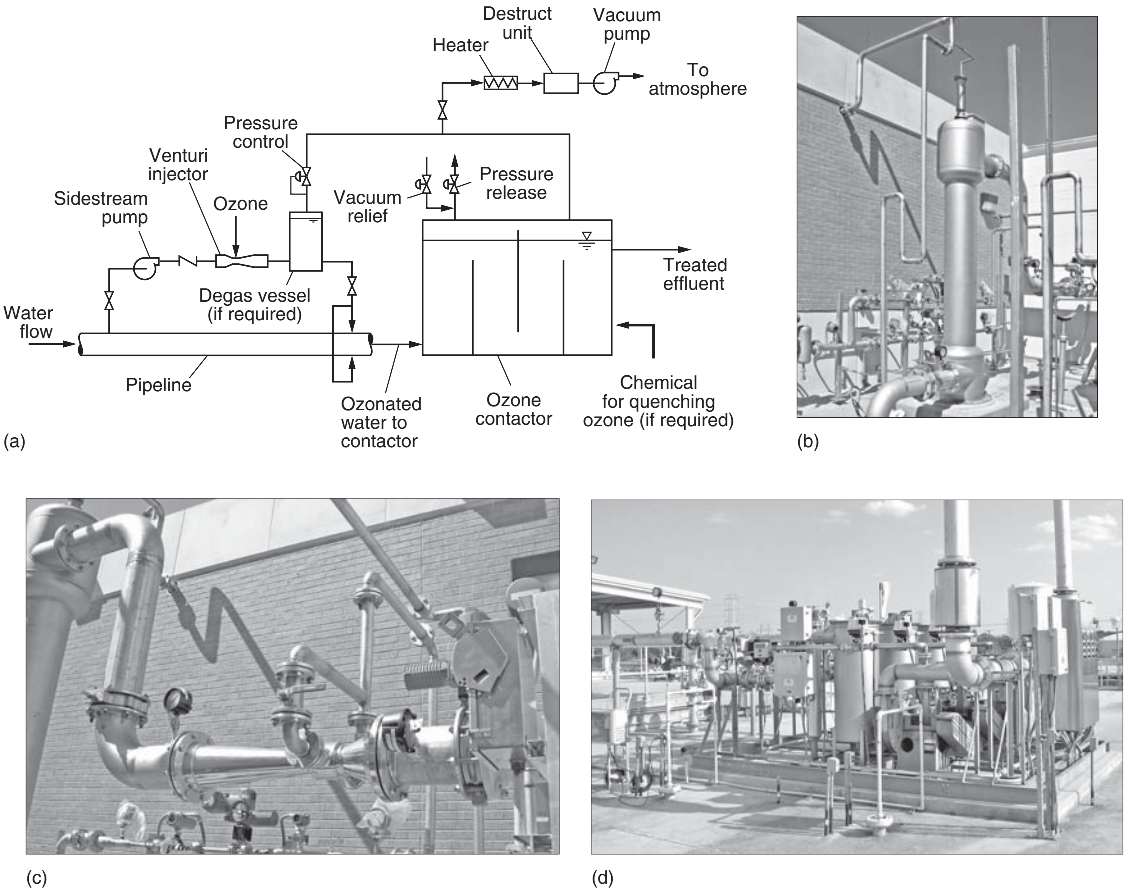


Figure 12–32

Sidestream ozone injection for disinfection: (a) typical schematic for sidestream injection system (adapted from Rakness, 2005), (b) view of degas vessel (venturi injector located on back right), (c) Venturi injector used in conjunction with degas vessel shown in (b) (photos (b) and (c) courtesy of Glenn Hunter, Process Applications, Inc.), and (d) view of sidestream injection system located above enclosed ozone contactor, including venturi injectors (left side), degas vessels (center), and destruct units (right).

Peracetic Acid

In the late 1980s, the use of peracetic acid (PAA) was proposed as a wastewater disinfectant. Peracetic acid, made up of acetic acid and hydrogen peroxide, has been used for many years as a disinfectant and sterilizing agent in hospitals. Peracetic acid is also used as a bactericide and fungicide, especially in food processing. Interest in the use of PAA as a wastewater disinfectant arises from considerations of safety and the possibility that its use will not result in the formation of DBPs. The use of PAA is considered briefly in this section as an example of the continuing search for alternative disinfectants to replace chlorine.

Peracetic Acid Chemistry and Properties. Commercially available PAA, also known as ethaneperoxide acid, peroxyacetic acid, or acetyl hydroxide, is only available as a

Table 12–26

Properties of various peracetic acid (PAA) formulations[a]

Property	Unit	PAA, % 1.0	PAA, % 5	PAA, % 15
Weight PAA	%	0.8–1.5	4.5–5.4	14–17
Weight hydrogen peroxide	%	min 6	19–22	13.5–16
Weight acetic acid	%	9	10	28
Weight available oxygen	Wt, %	3–3.1	9.9–11.5	9.3–11.1
Stabilizers	Yes/no	Yes	Yes	Yes
Specific gravity		1.10	1.10	1.12

[a] Adapted from Solvay Chemicals, Inc. (2013).

quaternary equilibrium solution containing acetic acid, hydrogen peroxide, peracetic acid, and water. The pertinent reaction is as follows.

$$\underset{\text{Acetic acid}}{CH_3CO_2H} + \underset{\text{Hydrogen peroxide}}{H_2O_2} \rightleftarrows \underset{\text{Peracetic acid}}{CH_3CO_3H} + H_2O \tag{12–63}$$

The undissociated PAA (CH_3CO_3H) is considered to be the biocidal form in the equilibrium mixture, however, the hydrogen peroxide may also contribute to the disinfection process. Hydrogen peroxide is also more stable than PAA. The properties of PAA are summarized in Table 12–26.

Effectiveness of Peracetic Acid as a Disinfectant. The effectiveness of PAA has been studied by Lefevre et al. (1992); Lazarova et al. (1998); Liberti et al. (1999); Gehr (2000, 2006); Koivunen (2005b); and Gehr et al. (2003), among others. A recent review was published by Kitis (2004). The findings to date are mixed concerning the bactericidal effectiveness of PAA, as well as the impact of wastewater characteristics on the effectiveness of PAA, especially when used alone. When combined with UV the effectiveness of PAA appears to be enhanced significantly (see discussion of combined disinfectants presented below). It has been hypothesized that the principal means by which disinfection is accomplished by PAA may be by the release of hydroxyl radicals ($HO^\bullet$) and the active oxygen resulting from secondary reactions (Caretti and Lubello, 2003). The current literature must be consulted for more information on the application of PAA.

In a report by the U.S. EPA (1999b), PAA was included among a total of 5 possible disinfectants for use on combined sewer overflows (CSOs). Based on data for disinfection of secondary treatment plant effluents, it was suggested that PAA be strongly considered for CSO disinfection. Among the desirable attributes listed are absence of persistent residuals and byproducts, not affected by pH, short contact time, and high effectiveness as a bactericide and viricide.

Formation of Disinfection Byproducts. Based on the limited data available, the principal end products identified were CH_3COOH (acetic acid or vinegar), O_2, CH_4, CO_2, and H_2O, none of which are considered toxic in the concentrations typically encountered.

Use of Peroxone as a Disinfectant

Peroxone is the combination of ozone (or ultraviolet light) with hydrogen peroxide. When hydrogen peroxide is added with ozone, it will quench ozone residual. In reacting with dissolved ozone (or reacting with photons of light from a UV process) it becomes an advanced oxidation process (AOP), which is characterized by the prolific generation of hydroxyl radicals (OH•). These OH• are stronger and less selective than ozone or other oxidants in regard to destroying target synthetic or naturally occurring micropollutants and microorganisms. The use of peroxone as an advanced oxidation process is discussed in Chap. 6.

Sequential Chlorination

Developed by the Sanitation Districts of Los Angeles County, *sequential chlorination* is a two-step disinfection process. In the first step, chlorine is added to nitrified filtered effluent to produce a free chlorine residual (FCR). As noted previously, free chorine, because of its high germicidal effectiveness, rapidly inactivates both bacteria and viruses. It also reacts with NDMA precursors to make them less available for subsequent NDMA formation. In the second step, ammonia and additional chlorine, if needed, are added to form chloramines, which provide additional bacterial and viral disinfection and minimize the formation of THMs. At the lowest free chlorine CT values tested (2 to 4 mg•min/L), it was possible to achieve an average of more than 6-log MS2 bacteriophage inactivation. With respect to DBP formation, the levels of THMs increased while the corresponding levels of NDMA decreased as compared to conventional chloramination. The sequential chlorination process has been developed to provide an alternative to the prescriptive CT value of 450 mg•min/L at a minimum modal contact time of 90 min required in the CDPH reclamation criteria [CCR, Section 60301.230(a)] (Maguin et al., 2009; Friess et al., 2013).

Combined Chemical Disinfection Processes

Interest in the sequential or simultaneous use of two or more disinfectants has increased within the last few years, especially in the water supply field. Reasons for the increased interest in the use of multiple disinfectants include (U.S. EPA, 1999a):

- The use of less reactive disinfectants, such as chloramines, has proven to be quite effective in reducing the formation of DBPs, and more effective for controlling biofilms in the distribution system.
- Regulatory and consumer pressure to produce water that has been disinfected to achieve high levels of inactivation for various pathogens has forced both the water and wastewater industry to search for more effective disinfectants. To meet more stringent disinfection standards, higher disinfectant doses have been used which, unfortunately, have resulted in the production of increased levels of DBPs.
- Based on the results of recent research it has been shown that the application of sequential disinfectants is more effective than the additive effect of the individual disinfectants. When two (or more) disinfectants are used to produce a synergistic effect by either simultaneous or sequential application to achieve more effective pathogen inactivation, the process is referred to as interactive disinfection (U.S. EPA, 1999a).

Currently, extensive research is being conducted on these processes. Some examples for the use of combined and/or sequential application of disinfectants are presented in Table 12–27. Because the application of multiple disinfectants is, at present, site specific, depending on the microorganism, the disinfection technologies employed, and other non-disinfection process objectives, the current literature must be reviewed to assess the suitability and effectiveness of combined disinfection technologies. For example, in Australia, combined use of chlorination and UV is becoming the norm in water-reuse applications.

Table 12–27

Effectiveness of combined disinfectants and processes for water and wastewater treatment[a,b]

Combined disinfectants	Response	Reference
In water treatment		
Ozone (O_3), UV, and chloramines replaced chlorination	Increase in CT credits by as much as 3 to 5 logs	Malley (2005)
Sequential sonification and chlorine	Increase in effectiveness over use of sonification or chlorine alone	Plummer and Long (2005)
Ozone and free chlorine, ozone and monochloramine, chlorine dioxide and free chlorine, chlorine dioxide, chlorine, and monochloramine	Synergistic response observed in the inactivation of *C. parvum* oocysts	Li et al. (2001)
Sequential UV and chlorine for inactivation of adenoviruses	Increase in effectiveness over use of UV or chlorine alone	Sirikanchana et al. (2005)
In wastewater treatment		
Free chlorine and combined chlorine	Reduced CT times for the inactivation of virus	Maguin et al. (2009), Friess et al. (2013)
Peracetic acid (PAA) and UV	Increase in effectiveness over use of UV or PAA alone	Chen et al. (2005), Lubello et al. (2002)
PAA and UV and PAA and ozone	Increase in effectiveness over use of PAA and UV alone	Caretti and Lubello (2005)
PAA and hydrogen peroxide (H_2O_2), H_2O_2 and UV, and H_2O_2 and O_3	No improved effectiveness	Caretti and Lubello (2005), Lubello et al. (2002)
Ozone, PAA, H_2O_2 and copper (Cu)	PAA and H_2O_2 alone had no effect, addition of 1 mg/L Cu had a dramatic effect	Orta de Velasque et al. (2005)
PAA/UV and H_2O_2/UV	PAA/UV had synergistic effects, whereas H_2O_2/UV did not	Koivunen (2005a)
Ultrasound and UV	Increase in effectiveness over use of UV alone	Blume et al. (2002) see also Blume and Neis (2004)

[a] Adapted in part from Gehr (2006).

[b] Additional combinations are reviewed in U.S. EPA (1999a).

12–9 ULTRAVIOLET (UV) RADIATION DISINFECTION

The germicidal properties of ultraviolet (UV) light sources have been used in a wide variety of applications since the use of UV was pioneered in the early 1900s, having been discovered first in the 1880s. First used on high quality water supplies, the use of UV light as a wastewater disinfectant evolved during the 1990s with the development of new lamps, ballasts, and ancillary equipment. With the proper combination of UV dose and water quality, UV irradiation has proven to be an effective disinfectant for bacteria, protozoa, and viruses in both unfiltered and filtered secondary effluent, while not contributing to the formation of toxic byproducts. In several cases, UV has even been proven to be effective at disinfecting primary effluents. To develop an understanding of the application of UV for the disinfection of wastewater the following topics are considered in this section: (1) source of UV radiation, (2) UV system configurations, (3) the germicidal effectiveness of UV irradiation, (4) modeling the UV disinfection process, (5) estimating the UV dose,

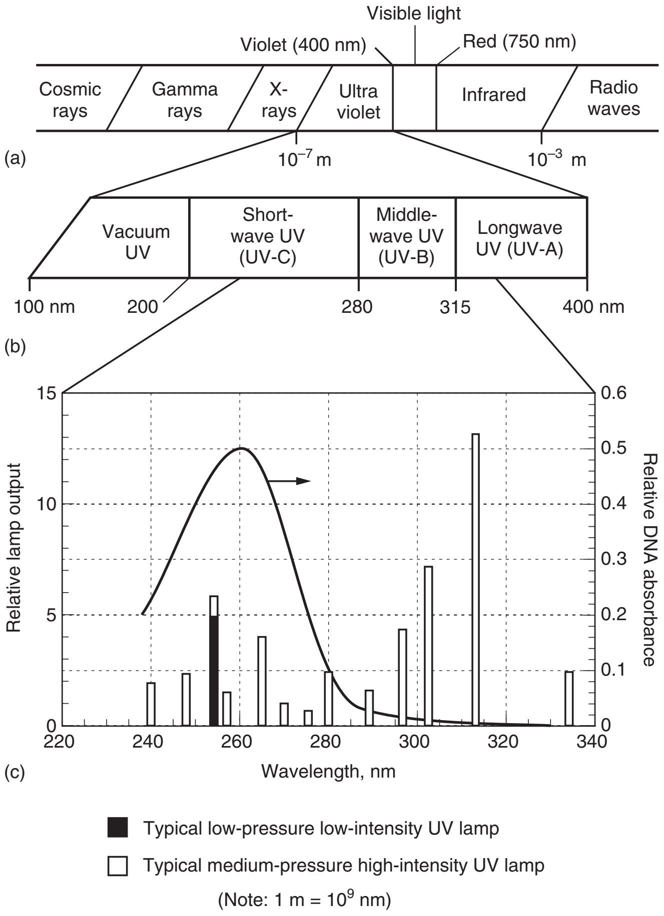


Figure 12–33

Definition sketch for ultraviolet (UV) radiation disinfection: (a) identification of the ultraviolet radiation portion of the electromagnetic spectrum, (b) identification of the germicidal portion of the UV radiation spectrum, and (c) UV radiation spectra for both low-pressure low-intensity and medium-pressure high-intensity UV lamps and the relative UV absorption for DNA superimposed over spectra of the UV lamps.

(6) UV disinfection guidelines, (7) analysis of a UV disinfection system, (8) operational issues with UV systems, and (9) the environmental impacts of disinfection with UV radiation.

Source of UV Radiation

The portion of the electromagnetic spectrum in which UV radiation occurs, is between 100 and 400 nm [see Fig. 12–33(a)]. The UV radiation range is characterized further according to wavelength as longwave (UV-A), also known as near-ultraviolet irradiation, middlewave (UV-B), and shortwave (UV-C), also known as far UV. The germicidal portion of the UV radiation band is between about 220 and 320 nm, principally in the UV-C range. The UV wavelengths between 255 to 265 nm are considered to be most effective for microbial inactivation [see Fig. 12–33(c)]. Most commonly, UV radiation is produced by striking an electric arc between two electrodes in specially designed lamps containing liquid mercury, as well as other gas mixtures. The energy generated by the excitation of the liquid mercury causes it to vaporize. Mercury in its gaseous form excites electrons in the lamps thus producing photons of UV light.

When used for water and wastewater disinfection, quartz sleeves are most often used to isolate the UV lamps from direct water contact and to control the lamp wall temperature by buffering the effluent temperature extremes to which the UV lamps are exposed, thereby maintaining a fairly uniform UV lamp output. In another less common configuration, the

Table 12–28

Typical operational characteristics for UV lamps

Item	Unit	Type of lamp Low-pressure low-intensity	 Low-pressure high-intensity	 Medium-pressure high-intensity
Power consumption	W	40–100	200–500[a]	1000–13,000
Lamp current	mA	350–550	Variable	Variable
Lamp voltage	V	220	Variable	Variable
Germicidal output/input	%	30–50	35–50	15–20[b]
Lamp output at 254 nm	W	25–27	60–400	100–2000
Lamp operating temperature	°C	35–50	100–150	600–800
Pressure	mm Hg	0.007	0.01–0.8	10^2–10^4
Lamp length	m	0.75–1.5	1.8–2.5	0.3–1.2
Lamp diameter	mm	15–20	Variable	Variable
Sleeve life	y	4–6	4–6	1–3
Ballast life	y	10–15	10–15	3–5
Estimated lamp life	h	8000–12,000	9000–15,000	3000–8000

[a] Up to 1200 W in very high output lamp.

[b] Output in the most effective germicidal range (~255 – 265 μm, see Fig. 12–33).

water to be disinfected is passed through proprietary plastic tubes that are themselves surrounded by UV lamps. The output of UV disinfection systems also decreases with time due to a reduction in the electron pool within the UV lamp, deterioration of the electrodes, and the aging of the quartz sleeve. Lamps with other gas mixtures and without electrodes, as described below, are also used to generate UV light.

Types of UV Lamps

The principal electrode-type lamps used to produce UV light fall into three categories based on the lamp's internal operating parameters: *low-pressure low-intensity, low-pressure high-intensity,* and *medium-pressure high-intensity* systems. Comparative information on the operational characteristics of these three types of UV lamps is presented in Table 12–28. In the brief discussion of these types of UV lamps presented below, it is important to note that UV lamp technology is changing rapidly. It is, therefore, imperative that current manufacturers' literature be consulted when designing a UV disinfection facility. The ballasts used in conjunction with UV lamps are also discussed briefly.

Low-Pressure Low-Intensity UV Lamps. Low-pressure low-intensity mercury-argon electrode type UV lamps [see Fig. 12–34(a)] are used to generate a broad spectrum of essentially monochromatic radiation in the UV-C region with an intense peak at a wavelength of 253.7 nm (essentially 254 nm) and a lesser peak at about 184.9 nm. The peak at 254 nm is close to the 260 nm wavelength considered to be most effective for microbial

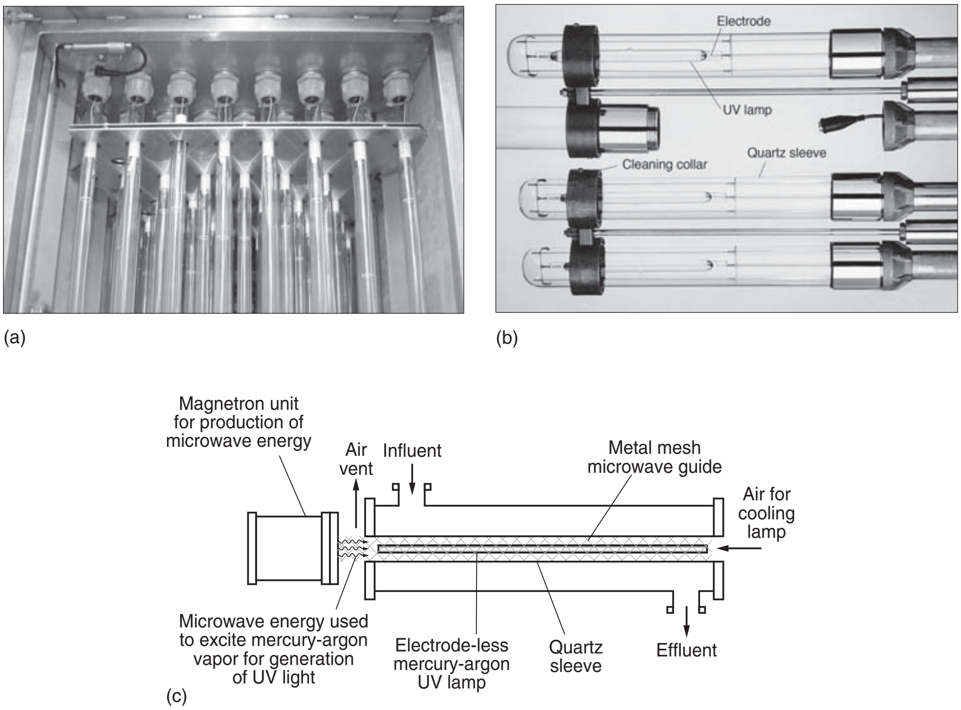


Figure 12–34

Typical examples of UV lamps: (a) low-pressure low-intensity, (b) medium-pressure high-intensity lamps with cleaning device, and (c) schematic illustration of the electrode-less microwave driven UV lamp [see Fig. 12–36(d)].

inactivation. Compared to power input, approximately 30 to 50 percent of the lamp's UV energy output is monochromatic at 254 nm, making it an efficient choice for disinfection processes. Also, approximately 85 to 88 percent of the lamp's output is monochromatic at 254 nm, making it an efficient choice for disinfection processes. Because there is an excess of liquid mercury in the low-pressure low-intensity UV lamp, the mercury vapor pressure is controlled by the coolest part of the lamp wall. If the lamp wall does not remain relatively near the optimum temperature of 40°C, some of the mercury will condense back to its liquid state thereby decreasing the number of mercury atoms available to release photons of UV; hence UV output declines.

Low-Pressure High-Intensity UV Lamps. Low-pressure high-intensity UV lamps are similar to the low-pressure low-intensity lamps [see Fig.12–34(a)] with the exception that a mercury-indium amalgam is used in place of liquid mercury. Use of the mercury amalgam allows greater UV-C output, typically from 2 to 10 times the output of conventional low-intensity lamps. Similar to low-pressure low-intensity lamps, low-pressure high-intensity lamps are very efficient at converting lamp input power into UV light. Compared to power input, approximately 35 to 50 percent of the lamp's UV energy output is monochromatic at 254 nm. Low-pressure high-intensity UV lamps operate at temperatures of 100 – 150°C. The amalgam in the low-pressure high-intensity UV lamps is used to maintain a constant level of mercury atoms, and, thus, provides greater stability over a broad temperature range. The UV output of low-pressure high-intensity lamps can be modulated between 30 and 100 percent. The range of modulation varies between different lamps. Current manufacturer's literature should be reviewed for lamp specifications as new low-pressure high-intensity lamps are being developed continuously.

Medium-Pressure High-Intensity UV Lamps. A number of medium-pressure high-intensity UV lamps have been developed over the last decade. Medium-pressure high-intensity UV lamps, which operate at temperatures of 600 to 800°C and vapor pressures of 10^2 to 10^4 mm Hg, generate polychromatic irradiation [see Fig. 12–33(c)]. Medium-pressure high-intensity UV lamps [see Fig. 12–34(b)] generate approximately 20 to 50 times the total UV-C output of low-pressure high-intensity UV lamps. Although the UV output of medium-pressure high-intensity lamps is significantly higher compared to low-pressure low-and high-intensity lamps, their efficiency is much lower. Compared to power input, only 15 to 20 percent of the lamp UV energy output is within the germicidal UV range. The use of medium-pressure high-intensity lamps is limited primarily to potable water supplies, large wastewater facilities, stormwater overflows, or on space-limited sites because fewer lamps are required and the footprint of the disinfection system is greatly reduced. The UV output of medium-pressure high-intensity lamps can be modulated across a range of power settings (typically 30 to 100 percent) without significantly changing the spectral distribution of the lamp. The particular UV lamp selected by UV system manufacturers is chosen on the basis of an integrated design approach in which the UV lamp, ballast, and reactor design are interdependent.

Alternative UV Lamp Technologies. A number of alternative technologies have been developed. Typically, they have not been used at the municipal scale, but this may change in the future. Some examples of the types of lamps that are being developed and applied include (1) the pulsed energy broad-band xenon lamp (pulsed UV), (2) the narrow band excimer UV lamp, (3) the mercury-argon electrode-less microwave powered high-intensity UV lamp, and (4) UV light emitting diodes (LED) lamp.

The pulsed UV lamp produces polychromatic light at high levels of radiation. It is estimated that the radiation produced by the pulsed UV lamp is 20,000 times as intense as sunlight at sea level. The disinfection effectiveness provided by pulsed UV lamps has been researched in some detail (O'Brien et al., 1996; EPRI, 1996; Mofidi et al., 2001). Narrow band excimer lamps produce essentially monochromatic light in three wavelengths: 172, 222, and 308 nm depending on the gas used in the lamp. Gases that have been used for the purpose include xenon (Xe), xenon chloride (XeCl), krypton (Kr), and krypton chloride (KrCl). In the microwave powered UV lamp, UV light is generated by striking a mercury-argon filled electrode-less UV lamp with microwave energy generated with a magnetron [see Fig. 12–34(c)]. Because the lamp does not contain electrodes, longer lamp life (3 to 5 y) is claimed, though no third-party certification has been completed. Based on preliminary results, it appears that the UV LED lamps currently under development will compete directly with conventional UV technologies. At present, there is no LED lamp technology that can compete with high output UV lamps.

As noted above, because developments in UV technology are occurring at such a rapid pace, it is essential that the current literature be consulted when designing UV disinfection systems. Note that in most cases, emerging technologies do not have a proven track record of cost-effective, reliable performance.

Ballasts for UV Lamps. A ballast is a type of transformer that is used to limit the current to a lamp. Because UV lamps are arc discharge devices similar to fluorescent lamps, the more current in the arc, the lower the resistance becomes. Without a ballast to limit current, the lamp would destroy itself. Thus, matching the lamp and ballast is of critical importance in the design of UV disinfection systems. Three types of ballasts are used: (1) standard (core coil), (2) energy efficient (core coil), and standard electronic (solid-state). In general, electronic ballasts are about 10 percent more energy efficient than magnetic ballasts. Electronic ballasts are now used most commonly for controlling the UV lamps used for disinfection.

UV Disinfection System Configurations

In addition to the type of lamp used, UV systems for the disinfection of wastewater can also be classified according to whether the flow occurs in open or closed channels. Each of these system configurations is described below.

Open Channel Disinfection Systems. The principal components of low-pressure low-and high-intensity open channel UV systems used for the disinfection of wastewater are illustrated on Fig. 12–35. As shown, lamp placement can be horizontal and parallel to the flow [see Fig. 12–35(a)], vertical and perpendicular to the flow [see Fig. 12–35(c)], or inclined (e.g. diagonal) to the flow [see Fig. 12–35(e)]. Each module contains a specified number of UV lamps encased in quartz sleeves. The total number of lamps is specific to each application, but the number of lamps in each module depends on the channel and overall system configuration and lamp manufacturer. The lamp spacing is manufacturer and lamp type specific and can range from 75 mm (3 in.) to 150 mm (6 in.). The inclined lamp UV systems are a relatively recent development. Stated advantages include the use of longer lamps with higher output, which reduces the total required number of lamps; improved system hydraulics and performance; and ease of installation, maintenance, and operation.

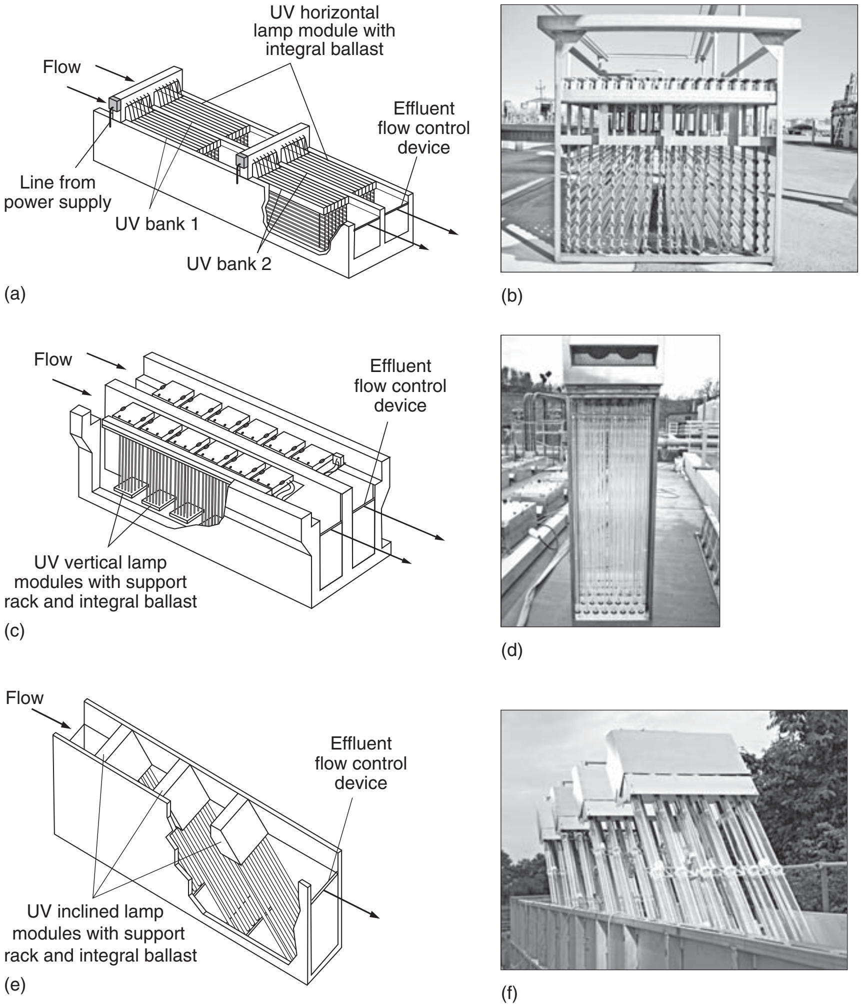


Figure 12–35

Isometric cut-away and photographic views of typical open channel UV disinfection systems: (a) horizontal lamp system parallel to flow (adapted from Trojan Technologies, Inc.), (b) view of one UV bank of a horizontal lamp system removed for cleaning, (c) vertical lamp system perpendicular to flow (adapted from Infilco Degremont, Inc.), (d) vertical lamp module removed from channel for cleaning, (e) inclined (45°) lamp system (adapted from Xylem, Inc.), and (f) view of inclined lamp UV system with lamps elevated out of the channel (courtesy of Xylem, Inc.).

An extended serpentine fixed sharp-crested weir, automatic level controlled adjustable weir, or a weighted flap gate (not recommended) is used to control the depth of flow through each disinfection channel. Proper level control is essential to (1) maintain submergence of the lamps at all times, (2) prevent short circuiting by ensuring that the water level above the top lamp is not too high, and (3) adequately seal the channel to prevent undisinfected water from bleeding through to the effluent channel when a bank of lamps is out of service. An inadequate level control device can often be the cause of poor UV disinfection performance.

Each channel typically contains two or more banks of UV lamps in series, and each bank is comprised of a number of modules (or racks of UV lamps). It is important to note that a standby bank or channel is normally provided for system reliability. The design flowrate is usually divided equally among a number of open channels. Typical examples of horizontal and vertical low-pressure low-high-intensity UV disinfection systems are shown on Fig. 12–35(c) through 12–35(f), respectively. A typical medium-pressure UV disinfection system is shown on Figs. 12–36(a) and 12–36(b). The lamps are arranged in

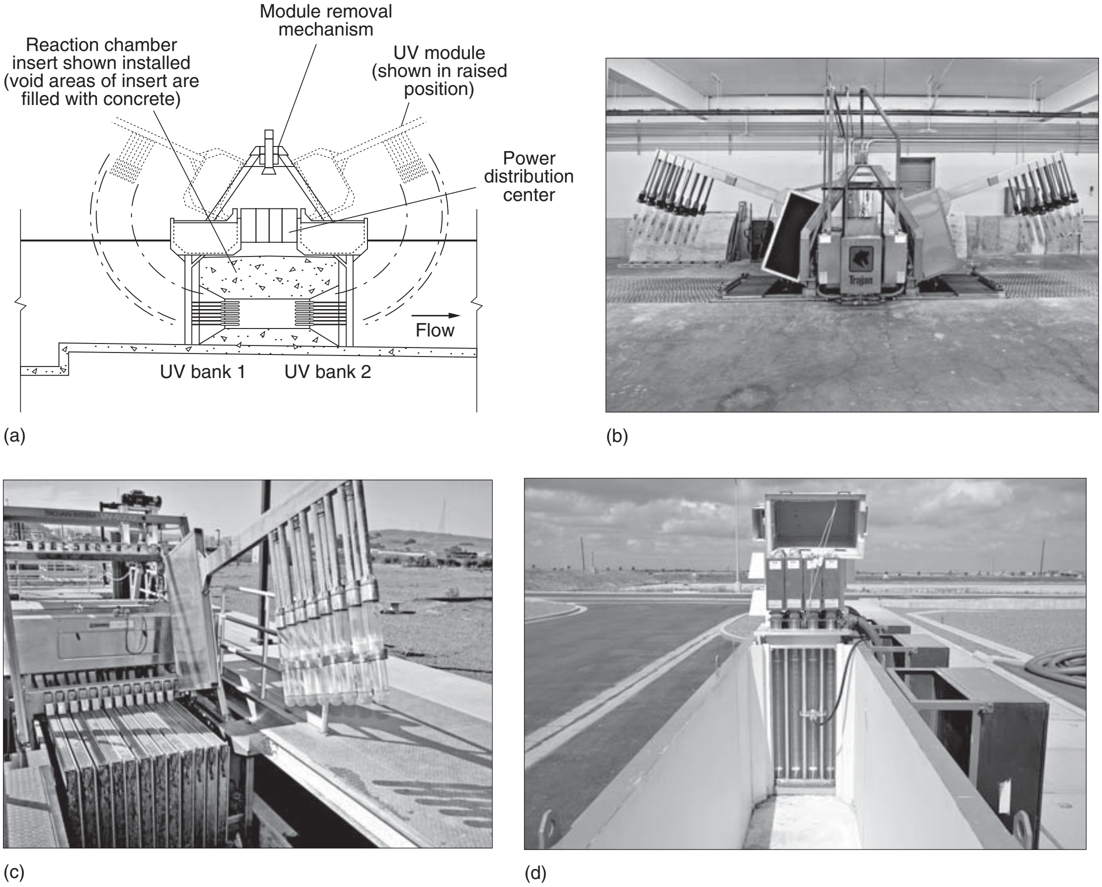


Figure 12–36

Typical examples of medium-pressure and microwave open channel UV disinfection systems: (a) schematic view through UV reactor (adapted from Trojan Technologies), (b) typical medium pressure UV system installed in open channel, (c) medium pressure UV system with one lamp module out of the reactor, and (d) microwave UV lamps with magnetron located above lamps [see Fig. 12–34(c)] in vertical orientation in open channel (adapted from Quay Technologies, Ltd).

modules and are positioned in a reactor with a fixed geometry [see Fig. 12–36(c)]. The lamp cleaning sleeves can be seen on Fig. 12–36(c). Vertical mercury-argon electrode-less microwave powered high-intensity UV lamps are shown on Fig. 12–36(d).

Closed Channel Disinfection Systems. A number of low- and medium-pressure high-intensity UV disinfection systems are designed to operate in closed channels or pipes. Two UV system configurations are used. In the first configuration, the direction of flow is perpendicular to the placement of the lamps, as shown on Fig. 12–37(a). In the second configuration, the direction of flow is parallel to the UV lamps [see Fig. 12–37(b)]. Because high-intensity UV lamps operate at a lamp wall temperature of between 600 to 800°C, the UV output of these lamps is unaffected by the effluent temperature. A typical medium-pressure UV disinfection reactor is shown on Figs. 12–37(c) and 12–37(d). A closed system pulsed UV reactor is shown on Fig. 12–37(f).

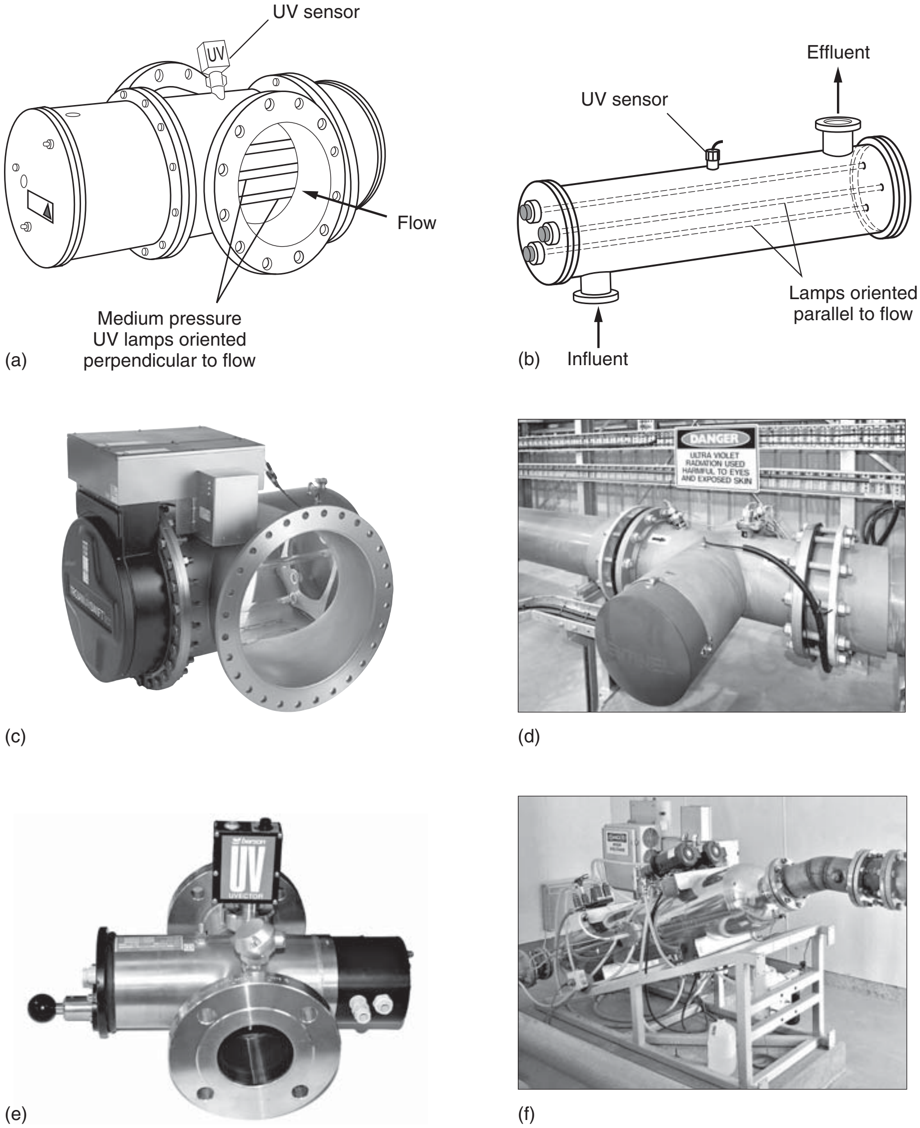


Figure 12–37

Views of medium-pressure high-intensity closed inline UV disinfection systems: (a) schematic of closed reactor with flow perpendicular to UV lamps, (b) schematic of closed reactor with flow parallel to UV lamps, (c) view through inline UV reactor (courtesy of Trojan Technologies, Inc.) (d) view of installed UV system, (e) close up of small inline UV system with manual cleaning device, and (f) view of pulsed UV reactor.

Quartz Sleeve Cleaning Systems

In UV disinfection systems, various physical and chemical characteristics of the water result in fouling of the quartz sleeves that encase each UV lamp. During operation of a UV lamp in wastewater, factors such as interfacial temperature, reactor hydraulics, and the quartz microstructure and topography allow attachment of inorganic debris and organic films or greases onto the protective quartz sleeve surrounding the lamp. These deposits absorb UV light and decrease the intensity of UV light penetration into the wastewater. The decrease in UV intensity leads to a decrease in UV dose resulting in reduction in disinfection performance. Fouling has been found to be complex and is difficult to predict. It also tends to be site-specific, owing mainly to the chemical and biological nature of the liquid matrix being treated by the UV system. To overcome quartz sleeve fouling, the majority of UV systems on the market have in-situ sleeve cleaning systems. These cleaning systems, as discussed below, can be divided into 2 categories: (1) mechanical, and (2) chemical-mechanical.

Mechanical Cleaning Systems. The first cleaning system category—mechanical cleaning—involves the use of a wiper that runs along the length of the quartz sleeve and removes large debris and scrapes off a degree of built up scaling. These types of systems are relatively effective in high quality effluents, but their performance can be compromised in high sleeve fouling effluents. Two items should be considered when designing a UV system with a mechanical wiper. The first is to apply an appropriate quartz sleeve fouling factor when sizing UV disinfection systems. Appropriate quartz sleeve fouling factors are considered subsequently. The second item that must be considered during design of the UV system is to include provisions for periodic (period of time is site specific) out-of-channel cleaning of the UV system. The external cleaning can be done manually with a manufacturer specified acid-based cleaning product for smaller UV systems or in an external acid bath (see Fig. 12–38) built for UV modules and with a crane for module channel removal and insertion into the bath.

Chemical-Mechanical Cleaning Systems. The second cleaning system category—chemical-mechanical—has been proven to be very effective at removing all

Figure 12–38

Cleaning UV disinfection system without integral wipers: (a) UV bank containing 18 lamp modules removed for external cleaning, positioned over cleaning solution bath and (b) UV bank placed in cleaning solution bath.


(a) (b)

scaling from quartz sleeves and maintaining a near 100 percent UV light output through the life of the system. Chemical-mechanical cleaning systems typically utilize two wipers that contain a small volume of an acidic gel (either phosphoric or citric acid based). This chemical gel remains in a canister surrounding each quartz sleeve and is replaced annually. Although a chemical-mechanical cleaning system has been shown to be more effective than mechanical cleaning alone, a quartz sleeve fouling factor should be applied to UV system sizing in any case.

Mechanism of Inactivation by UV Irradiation

Ultraviolet light is a physical rather than a chemical disinfecting agent. The mechanism of inactivation and photoreactivation are important concepts to understand as they are some of the fundamental principles of UV disinfection.

Inactivation Mechanisms. UV radiation penetrates the cell wall of the microorganism and is absorbed by the nucleic acids (DNA and RNA), which guide the development of all living organisms. Damage to the nucleic acid interferes with normal cell processes such as cell synthesis and cell division. Deoxyribonucleic acid (DNA) controls the structure, while ribonucleic acid (RNA) controls the metabolic processes. Typically, DNA is a double-stranded helical structure with four nucleotides: adenine, guanine, thymine, and cytosine, while RNA is a single-stranded structure with the nucleotides adenine, guanine, uracil, and cytosine.

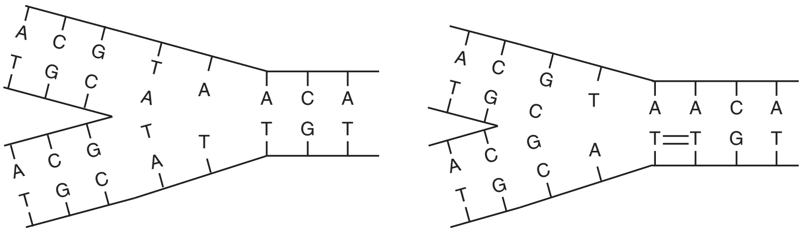
Exposure to UV radiation damages DNA by manipulating adjacent thymine molecules as illustrated on Fig. 12–39. The process of forming double bonds is known as dimerization. Cytosine-cytosine and cytosine-thymine dimers can also be formed. Thus, organisms rich in thymine such as protozoans *C. parvum* and *G. lamblia* tend to be more sensitive to UV radiation (see Table 12–5) (Mofidi et al., 2001; Mofidi et al., 2002). Uracil and cytosine are the corresponding molecules in RNA. Viruses contain either DNA or RNA, which is either single or double stranded. Adenovirus contains double-stranded DNA, which is considered as a possible explanation for its high sensitivity to UV light (Sommer et al., 2001). Exposure to UV radiation can also cause more severe damage, such as breaking chains, cross-linking DNA with itself, and cross-linking DNA with other proteins. (Crittenden et al., 2005). In general, UV irradiation must form a significant number of bonds or other damage to the cell to be effective, which is the case with the doses delivered in properly sized UV disinfection systems.

Figure 12–39

Formation of double bonds in microorganisms exposed to ultraviolet radiation.

Normal DNA

DNA with thymine dimerized



Normal DNA replicating

Replication stops at dimer

Microbial Growth Phase and Resistance to UV Irradiation. In addition to the mechanisms discussed above, it has been found that the growth phases of organisms can inherently provide protection against disinfectants. As noted in Chap. 7 there are four readily identifiable stages that cells undergo: lag growth phase, exponential growth phase, stationary growth phase, and death phase. The effect that different phases of growth cycle may have on UV susceptibility of *E. coli* has been evaluated by Modifi et al. (2002). It appears that when cell DNA is not actively dividing, bacteria may be more resistant to UV disinfection. Thus, if naturally occurring bacteria are not stressed by environmental factors, they may exhibit a similar spectrum of resistance to UV disinfection. Based on these findings, knowledge of the cell growth stage is of importance in establishing bacterial dose-response relationships.

Microbial Repair Following UV Irradiation. Because some organisms are able to maintain some metabolic activities after being exposed to UV radiation, they may be able to repair the damage caused by the exposure. Many organisms in nature have evolved mechanisms for reversing UV damage. Two different types of mechanisms are involved: (1) photoreactivation and (2) dark repair.

Photoreactivation. Photoreactivation involves specific enzymes that can repair sections of damaged DNA after being energized by exposure to light. The mechanism of photoreactivation, first discovered in 1949 for *Streptomyces griseus* by Kelner (1949) and for bacteriophage by Dulbecco (1949), was demonstrated to be enzyme-catalyzed (Rupert, 1960). The enzyme responsible for DNA repair is named *photolyase*. Photoreactivation can be described as the two-step enzymatic reaction between photolyase and its substrate, pyrimidine dimers (Friedberg et al., 1995). The first step is for photolyase to recognize any dimers (see Fig. 12–39) and specifically bind them to form an enzyme-substrate complex. The first step is light-independent and, therefore, can occur even under dark conditions. The enzyme-dimer complex is stable and goes through the second repair step in which the dimers are broken utilizing the energy of light at wavelengths between 310 and 490 nm. The second step is dependent only on light input.

For example, the *E. coli* photolyase has a round shape with a hole inside, which recognizes and structurally binds to the pyrimidine dimers sticking out from the genome DNA. Once the pyrimidine dimers are repaired (i.e., broken) and the structure is changed, the bind is loosened and the enzyme leaves the dimer (Friedberg et al., 1995). In the case of pathogenic parasites, the effects of photoreactivation are unclear. Based on infectivity studies, it was reported that the oocysts of *Cryptosporidium parvum* did not undergo photoreactivation (Rochelle et al., 2004). In another study, it has been reported that repair of the pyrimidine dimers did occur in oocysts of *Cryptosporidium parvum* (Oguma et al., 2001). What appears to be happening is that the repair of DNA following UV irradiation may not be sufficient for the organism to regain its infectivity. Although the necessary enzymes needed for repair are missing in viral DNA, the enzymes of the host cell can be used to accomplish the repair.

It should also be noted that the ability for an organism to repair itself appears to depend on a number of factors including UV dose (the effect is diminished at higher UV doses), UV wavelength, UV light intensity, and exposure time to photoreactivating light (Martin and Gehr, 2005). *Escherichia coli* exposed to monochromatic low-pressure UV light were able to repair themselves, whereas *Escherichia coli* exposed to polychromatic medium-pressure UV light were unable to repair themselves (Zimmer and Slawson, 2002; Oguma et al., 2002). However, *Legionella pneumophila* exhibited very high photoreactivation ability after exposure to either low-pressure or medium-pressure UV light (Oguma et al., 2004). From a review of some recent published findings, it appears that if effluent that has undergone UV disinfection is subsequently kept in the dark for approximately 3 hours, the regrowth potential is reduced significantly (Martin and Gehr, 2005). Clearly, more research needs to be done to understand what is causing the effect observed with medium-pressure UV light.

Dark Repair. In the early 1960s it was found that UV-induced DNA damage could be repaired without light (Hanawalt et al., 1979). Dark repair appears to be accomplished by two mechanisms: (1) excision repair and (2) recombination repair. In excision repair, enzymes remove the damaged section of DNA, and in recombination repair, the damaged DNA is regenerated using a complementary strand of DNA. Although the necessary enzymes needed for repair are missing in viral DNA, the enzymes of the host cell can be used to accomplish the repair. Contrary to photoreactivation, with high specificity to pyrimidine dimers, dark repair can act on various kinds of damage in the genome. Dark repair is a rather slow process compared to photoreactivation.

Germicidal Effectiveness of UV Irradiation

The overall effectiveness of the UV disinfection process depends on a number of factors including (1) the chemical characteristics of the wastewater to be irradiated, (2) the presence of particles, (3) the characteristics of the microorganisms, and (4) the physical characteristics of the UV disinfection system. Before considering these subjects, it is appropriate to consider the definition of UV dose to provide a frame of reference for the discussion of the factors affecting UV disinfection. The material presented below will also be useful in assessing the modeling of the UV process that is considered subsequently.

Definition of UV Dose. The effectiveness of UV disinfection is based on the UV dose to which the microorganisms are exposed. The UV dose, D, as defined previously, is given by Eq. (12–8), which is repeated here for convenience.

$$D = I_{\text{avg}} \times t \tag{12–8}$$

where D = UV dose, mJ/cm^2 (note mJ/cm^2 = mW·s/cm^2)
I_{avg} = average UV intensity, mW/cm^2
t = exposure time, s

Note that the UV dose term is analogous to the dose term used for chemical disinfectants (i.e., CT). As given by Eq. (12–8), the UV dose can be varied by changing either the intensity or exposure time. Additional details on the measurement of UV dose may be found in Linden and Mofidi (2003) and Jin et al. (2006).

Effect of Chemical Constituents in Wastewater. The constituents in wastewater can have a significant impact on the average UV intensity. The impact is measured in term of absorbance and transmittance. The reduction in UV intensity with distance is defined by Beers-Lambert Law, repeated here for convenience from Chap. 2.

$$\log\left(\frac{I}{I_o}\right) = -\varepsilon(\lambda)Cx = k(\lambda)x = [A(\lambda)/x]x \tag{2–19}$$

where I = light intensity at distance x from the light source, mW/cm^2
I_o = light intensity at light source, mW/cm^2

$\varepsilon(\lambda)$ = molar absorptivity (also known as the extinction coefficient) of the light-absorbing solute at wavelength λ, L/mole·cm
C = concentration of light-absorbing solute, mole/L
x = light path-length, cm
$k(\lambda)$ = the absorptivity, cm^{-1}
$A(\lambda)$ = absorbance, dimensionless

Although the absorbance, $A(\lambda)$, is dimensionless, it is often reported in units of cm^{-1}, which corresponds to absorptivity $k(\lambda)$. If the length of the light path is 1 cm, absorptivity is equal to the absorbance. In UV practice, it is more common to use transmittance, which is defined as

$$\text{Transmittance, } T, \% = \left(\frac{I}{I_o}\right) \times 100 \qquad (2\text{–}21)$$

Typical absorbance and transmittance values for wastewater after several different treatment processes are presented in Table 12–29.

Dissolved constituents impact UV disinfection either directly via absorbance (increasing absorbance serves to attenuate UV light to a larger degree) or via fouling of UV lamp sleeves such that a reduced intensity is applied to the bulk liquid medium. The effects of constituents found in effluent from different wastewater treatment processes are reported in Table 12–30. One of the most perplexing problems encountered in the application of UV disinfection for wastewater disinfection is the variation typically observed in the absorbance (or transmittance) at treatment plants. Often, the variations in transmittance are caused by industrial discharges, which can lead to diurnal as well as seasonal variations. Common industrial impacts are related to the discharge of inorganic and organic dyes, wastes containing metals, and complex organic compounds.

Of the inorganic compounds that affect transmittance, iron is considered to be the most important with respect to UV light absorbance because dissolved iron can absorb UV light directly. Organic compounds containing double bonds and aromatic functional groups can also absorb UV light. Absorbance values for a variety of compounds found in wastewater are given in Table 12–31. From a review of the information presented in Table 12–31, it is clear that the presence of iron in wastewater can have a significant impact on the use of UV. If iron salts are used within the treatment process, the economic benefits of changing to another chemical (e.g., alum) should be evaluated to determine whether the cost savings of having a smaller UV system are greater than the capital and operating costs of switching to the new chemical. It is also important to note that stormwater inflows can cause wide variations, especially when humic materials from terrestrial sources are present. In general, the solution to the problem of varying transmittance levels may require monitoring of industrial

Table 12–29

Absorbance and transmittance values for various wastewaters at a wavelength of 254 nm

Type of wastewater	Absorbance, a.u./cm	Transmittance[a],%
Primary	0.70 to 0.30	20 to 50
Secondary	0.35 to 0.15	45 to 70
Nitrified secondary	0.25 to 0.10	56 to 79
Filtered secondary	0.25 to 0.10	56 to 79
Microfiltration	0.10 to 0.04	79 to 91
Reverse osmosis	0.05 to 0.01	89 to 98

[a] $T, \% = 10^{-A(\lambda)} \times 100$.

Table 12–30

Impact of wastewater constituents on the use of UV radiation for wastewater disinfection

Constituent	Effect
BOD, COD, TOC, etc.	No or minor effect, unless humic materials comprise a large portion of the BOD
NOM (natural organic matter)	Strong absorbers of UV radiation
Oil and grease	Can accumulate on quartz sleeves of UV lamps, can absorb UV radiation
TSS	Absorption of UV radiation, can shield embedded bacteria
Alkalinity	Can impact scaling potential. Also affects solubility of metals that may absorb UV light
Hardness	Calcium, magnesium and other salts can form mineral deposits on quartz tubes, especially at elevated temperatures
Ammonia	No or minor effect
Nitrite	No or minor effect
Nitrate	No or minor effect
Iron	Strong absorber of UV radiation, can precipitate on quartz tubes, can become embedded in suspended solids and shield bacteria by absorption
Manganese	Strong absorber of UV radiation
pH	Can affect solubility of metals and carbonates
TDS	Can impact scaling potential and the formation of mineral deposits
Industrial discharges	Depending on the constituents (e.g., dyes), may lead to a diurnal and seasonal variations in the transmittance
Stormwater inflow	Depending on the constituents, may lead to short term as well as seasonal variations in the transmittance

Table 12–31

UV Absorbance of water and common chemicals found in wastewater

Compound	Form or designation	Molar absorption coefficient, L/mole·cm	Threshold concentration, mg/L
Ferric iron	Fe[III]	3069	0.057
Ferrous iron	Fe[II]	466	9.6
Hypochlorite ion	OCl^-	29.5	8.4
N-nitrosodimethylamine	NDMA	1974	
Nitrate	NO^{3-}	3.4	
Natural organic matter	NOM	80 to 350	
Ozone	O_3	3250	0.071
Zinc	Zn^{2+}	1.7	187
Water	H_2O	6.1×10^{-6}	

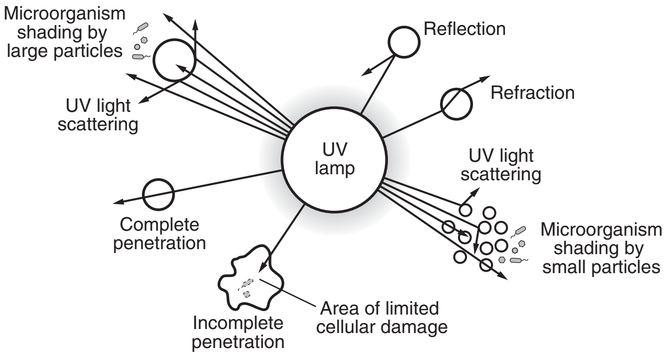


Figure 12–40

Particle interactions that effect the effectiveness of UV disinfection including microorganism shading; light scattering, reflection, and refraction; and incomplete penetration.

discharges, the implementation of source control programs, and correcting sources of infiltration. In some cases, biological treatment will mitigate the influent variations. In some extreme situations, the conclusion may be that UV disinfection is not practical.

Where the implementation of UV disinfection is being assessed, it is useful to install online transmittance monitoring equipment to document the variations that occur in the transmittance with time. Alternatively, a bench-top photometer may be used to sample UV transmittance manually. Sampling with a photometer will provide "snapshots" of UV transmittance as opposed to constant measurements from an online monitor. If enough "snapshots" are taken, fairly accurate UV transmittance trending data can be determined, which may be equal in accuracy to the online monitoring data.

Effect of Particles. The presence of particles in the wastewater to be irradiated can also impact the effectiveness of UV disinfection (Qualls et al., 1983; Parker and Darby, 1995; Emerick et al., 1999). The manner in which particles can affect UV performance is illustrated on Fig. 12–40. Many organisms of interest in wastewater (e.g., coliform bacteria) occur both in a dispersed state (i.e., not bound to other objects) and a particle-associated state (i.e., bound to other objects such as other bacteria or cellular debris). Coliform bacteria are of particular importance because of the central role they play in discharge permits [i.e., coliform bacteria are used as indicators for the presence of other pathogenic organisms (see Chap. 2) and their inactivation is assumed to correlate with the inactivation of other pathogenic organisms]. Dispersed coliform bacteria are inactivated readily because they are exposed fully to the average UV light intensity as compared to particle-embedded microorganisms (see Fig. 12–40). Treatment process related disinfection problems, when disinfecting unfiltered effluent, usually result from the influence of particle associated organisms (see also Fig. 12–5). In fact, coliform bacteria can associate with particles to such a degree that they are completely shielded from UV light resulting in a residual coliform bacteria concentration post UV irradiation.

It has been observed in activated sludge effluents that a minimum particle size (on the order of 10 μm) governs the ability to shield coliform bacteria from UV light (Emerick et al., 2000). Due to the inherent porous nature of activated sludge particles, particles smaller than that critical size are unable to reduce the applied intensity and thus embedded organisms are inactivated in a manner similar to dispersed organisms. Particles greater than the critical size can either reduce the applied UV intensity, leading to a reduced inactivation rate for organisms associated with the particle, or shield coliform bacteria. Particle size does not appear to be a governing factor once the critical size is exceeded because coliform bacteria are located randomly within particles, are not typically located in the most shielded regions within particles, and common enumeration techniques typically exclude the larger sized particles.

Table 12–32

Typical UV dosages required to achieve different effluent total coliform disinfection standards for various wastewaters

Type of wastewater	Initial coliform count, MPN/100 mL	UV dose, mJ/cm², Effluent standard, MPN/100 mL: 1000	200	23	≤2.2
Raw wastewater	10^7–10^9	20–50			
Primary effluent	10^7–10^9	20–50			
Trickling filter effluent	10^5–10^6	20–35	25–40	40–60	90–110
Activated sludge effluent	10^5–10^6	20–30	25–40	40–60	90–110
Filtered activated sludge effluent	10^4–10^6	20–30	25–40	40–60	80–100
Nitrified effluent	10^4–10^6	20–30	25–40	40–60	80–100
Filtered nitified effluent	10^4–10^6	20–30	25–40	40–60	80–100
Microfiltration effluent	10^1–10^3	5–10	10–15	15–30	40–50
Reverse osmosis	~0	—	—	—	5–10
Septic tank effluent	10^7–10^9	20–40	25–50		
Intermittent sand filter effluent	10^2–10^4	10–20	15–25	25–35	50–60

Characteristics of the Microorganisms. The effectiveness of the UV disinfection process depends on the characteristics of the microorganisms as well as the microorganism group. Typical values for the disinfection of coliform organisms with UV light for various wastewaters are reported in Table 12–32. It should be noted that the dosage values given in Table 12–32 are only meant to serve as a guide for the initial estimation of the required UV dose. The range of the reported values reflects the variable nature of wastewater. The relative effectiveness of UV irradiation for disinfection of representative microorganisms of concern is in wastewater is reported in Table 12–33. As with the values given in Table 12–5, the values given in Table 12–33 are only meant to serve as a guide in assessing the relative UV dose required for different microorganisms. Knowledge concerning the required UV dose for specific pathogen inactivation is changing continuously as improved methods of analysis are applied. For example, before infectivity studies were conducted, it was thought that UV irradiation at reasonable dosage values (i.e., less than 200 mJ/cm^2) was not effective for the inactivation of *Cryptosporidium parvum* and *Giardia lamblia*. However, based on infectivity studies, it has been found that both of these protozoans are inactivated with extremely low UV dosage values (typically in the range of 5 to 15 mJ/cm^2) (Linden et al., 2001; Mofidi et al., 2001; Mofidi et al., 2002). The current literature should be consulted to obtain the most contemporary information regarding required UV dosages for the inactivation of specific microorganisms.

Impact of System Characteristics. Problems with the application of Eq. (12–8) for use in the design of UV disinfection reactors are associated with (1) inaccurate knowledge of the average UV intensity and (2) the exposure time associated with all of the pathogens passing through a UV disinfection system. In practice, field-scale UV disinfection

Table 12–33

Estimated relative effectiveness of UV radiation for the disinfection of representative microorganisms of concern in wastewater

Organism	Dosage relative to total coliform dosage
Bacteria	
Escherichia coli (E.coli)	0.6–0.8
Fecal coliform	0.9–1.0
Pseudomonas aeruginosa	1.5–2.0
Salmonella typhosa	0.8–1.0
Streptococcus fecalis	1.3–1.4
Total coliform	1.0
Vibrio cholerae	0.8–0.9
Viruses	
Adenovirus	6–8
Coxsackie A2	1.2–1.2
MS-2 bacteriophage	2.2–2.4
Polio type 1	1.0–1.1
Rotavirus SA 11	1.4–1.6
Protozoa	
Acanthamoeba castellanii	6–8
Cryptosporidium parvum	0.4–0.5
Cryptosporidium parvum oocysts	1.3–1.5
Giardia lamblia	0.3–0.4
Giardia lamblia cysts	0.3–0.4

[a] Relative doses based on discrete non clumped single organisms in suspension. If the organisms are clumped or particle associated, the relative dosages have no meaning.

reactors have dose distributions resulting from both the internal intensity profiles and exposure time distribution. The internal intensity profiles are a reflection of the nonhomogeneous placement of lamps within the system, lack of ideal radial mixing within the system, the scattering/absorbing effects of particulate material, and the absorbance of the liquid medium. The distribution associated with exposure time is a reflection of non-ideal hydraulics leading to longitudinal mixing.

One of the most serious problems encountered with UV disinfection systems in both open and closed channel systems is achieving a uniform velocity field in the approach and exit to and from the UV banks. Achieving a uniform velocity field can be especially difficult when UV systems are retrofitted into existing open channels, such as converted chlorine contact basins. A second equally serious problem with UV system hydraulics is even flow distribution between channels whether the channels are new or are a retrofit of existing structures. An uneven flow split can lead to overdosing in one channel and most importantly, underdosing in the other(s) therefore compromising disinfection performance. Ensuring ideal flow distribution to multiple channels and uniform velocities within those channels is critical. To optimize the hydraulic performance of UV disinfection systems, computational fluid dynamics (CFD) modeling should be considered.

Estimating UV Dose

The first step in assessing the performance of a UV disinfection system is to determine the UV dose needed to inactivate the challenge microorganism to a level prescribed by the treatment plant discharge permit and/or is protective of public health in water reuse applications. Three methods have been used to estimate the UV dose. In the first method, an average UV dose is determined by assuming an average system UV intensity and exposure time. The average UV intensity is estimated using a computational procedure known as the point source summation (PSS) method (U.S. EPA, 1992). Over the past decade, the PSS method has been used less frequently by designers due to its failure to account for system-specific hydraulics (i.e., ideal hydraulic behavior is assumed in the PSS that never occurs in field-scale disinfection systems). At present, this method should not be used to determine UV dose.

The second method involves the use of CFD to integrate both the distribution of UV intensities and velocity profiles within the reactor to obtain a distribution of UV doses within a system (Batchley et al., 1995). Although the CFD method is promising, its use is limited at the present time (2013) because (1) the methodology is not standardized, (2) the methodology has been unable to predict disinfection performance adequately, and (3) the reporting of a distribution of UV doses, even if accurate, is problematic for UV disinfection system specification. In the third, and most widely used method, the UV dose is determined using a collimated beam bioassay. Use of the bioassay approach in designing UV disinfection systems is discussed below.

Determination of UV Dose by Collimated Beam Bioassay. The most common and industry accepted procedure for determining the required UV dose for the inactivation of challenge microorganisms involves the use of a collimated beam and a small reactor (i.e., a Petri dish) to which a known UV dose is applied. Typical collimated beam devices are shown on Fig. 12–41. Use of a monochromatic low-pressure low-intensity

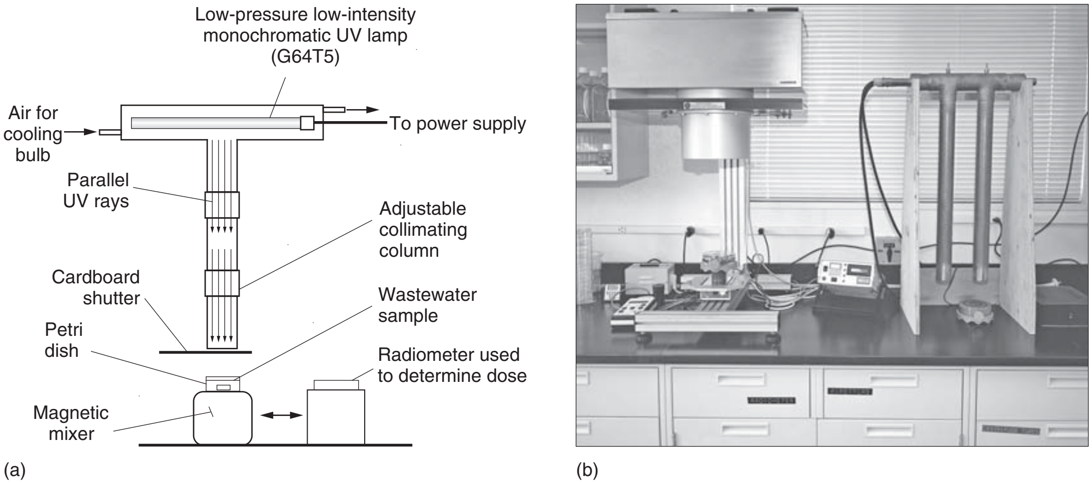


Figure 12–41

Collimated beam device used to develop dose-response curves for UV disinfection: (a) schematic and (b) view of two different types of collimated beam devices. The collimated beam on the left is of European design; the collimated beam on the right is of the type shown schematically in (a).

lamp in the collimated beam apparatus allows for accurate characterization of the applied UV intensity. Use of a batch reactor allows for accurate determination of exposure time. The applied UV dose, as defined by Eq. (12–8), can be controlled by simply varying the exposure time and maintaining a constant, known UV intensity. Because the geometry is fixed, the depth-averaged UV intensity within the Petri dish sample (i.e., the batch reactor) can be computed using the following relationship.

$$D = I_o t(1 - R)P_f\left[\frac{(1 - 10^{-k_{254}d})}{2.303(k_{254}d)}\right]\left(\frac{L}{L + d}\right) \tag{12–64}$$

$$D = I_o t(1 - R)P_f\left[\frac{(1 - e^{-2.303k_{254}d})}{2.303(k_{254}d)}\right]\left(\frac{L}{L + d}\right) \tag{12–65}$$

where D = average collimated beam UV dose ($I_o \times t$), mJ/cm^2
I_o = incident UV intensity averaged over the surface of the sample before and after irradiating sample, mW/cm^2
t = exposure time, s
R = reflectance at the air water interface at 254 nm
P_f = Petri dish factor
k_{254} = absorbance of sample, absorptivity, a.u./cm (base 10)
d = depth of sample, cm
L = distance from lamp centerline to liquid surface, cm

The term $(1 - R)$ on the right side of Eq. (12–64) accounts for the reflectance at the air water interface. The value of R is typically about 2.5 percent. The term P_f accounts for the fact that the UV intensity may not be uniform over the entire area of the Petri dish. The value of P_f is typically greater than 0.9. The term within the brackets is the depth averaged UV intensity within the Petri dish and is based on the Beers-Lambert Law (see Example 2–5, Chap. 2). The final term is a correction factor for the height of the UV light source above the sample. The application of Eq. (12–64) is illustrated in Example 12–11.

The uncertainty of the computed UV dose can be estimated using the sum of the variances as given by either of the following expressions:

Maximum Uncertainty

$$U_D = \pm \sum_{n=1}^{N}\left|U_{V_n}\frac{\partial D}{\partial V_n}\right| \tag{12–66}$$

Best Estimate of Uncertainty

$$U_D = \pm\left[\sum_{n=1}^{N}\left(U_{V_n}\frac{\partial D}{\partial V_n}\right)^2\right]^{1/2} \tag{12–67}$$

where U_D = uncertainty of UV dose value, mJ/cm^2
U_{V_n} = uncertainty or error in variable n
V_n = variable n
$\partial D/\partial V_n$ = partial derivative of the expression with respect to the variable V_n
N = number of variables

The maximum estimate of uncertainty as given by Eq. (12–66) represents the condition where every error will be a maximum value. The best estimate of uncertainty, as given by Eq. (12–67), is used most commonly because it is unlikely that every error will be a maximum at the same time and the fact that some errors may cancel each other. The application of Eq. (12–67) is illustrated in Example 12–11. Knowledge of the average UV intensity and exposure time allows calculation of the average applied UV dose using Eq. (12–8). The UV dose is then correlated to the microorganism inactivation results as discussed below.

EXAMPLE 12–11 **Determination of UV Dose Delivered in Collimated Beam Test** The following measurements were made to establish the UV dose using a collimated beam. Using these data determine the average UV dose delivered to the sample and best estimate of the uncertainty associated with the measurement.

$I_o = 5 \pm 0.35$ mW/cm^2 (accuracy of meter ±7%)

$t = 60 \pm 1$ s

$R = 0.025$ (assumed to be the correct value)

$P_f = 0.94 \pm 0.02$

$k_{254} = 0.065 \pm 0.005$ cm^{-1}

$d = 1 \pm 0.05$ cm

$L = 40 \pm 0.5$ cm

Solution

1. Using Eq. (12–64) estimate the UV dose delivered by the collimated beam.

$$D = I_o t(1 - R)P_f\left[\frac{(1 - 10^{-k_{254}d})}{2.303(k_{254}d)}\right]\left(\frac{L}{L + d}\right)$$

$$= (5 \times 60)(1 - 0.025)(0.94)P_f\left[\frac{(1 - 10^{-0.065 \times 1})}{2.303(0.065 \times 1)}\right]\left(\frac{40}{40 + 1}\right)$$

$$= (300)\,(0.975)\,(0.94)\,(0.928)\,(0.976) = 249 \text{ mJ/cm}^2$$

2. Determine the best estimate of uncertainty for the computed UV dose. The uncertainty of the computed dose can be estimated using Eq. (12–67). The procedure is illustrated for one of the variables and summarized for the remaining variables.

   a. Find the variability in the measured UV dose due to the variability of the measured time t. The partial derivative of the expression used in step one with respect to t is

$$U_t\frac{\partial D}{\partial t} = U_t\left\{I_o(1 - R)P_f\left[\frac{(1 - 10^{-k_{254}d})}{2.303(k_{254}d)}\right]\left(\frac{L}{L + d}\right)\right\}$$

$$U_t\frac{\partial D}{\partial t} = (1)\left\{5(1 - 0.025)(0.94)\left[\frac{(1 - 10^{-0.065 \times 1})}{2.303(0.065 \times 1)}\right]\left(\frac{40}{40 + 1}\right)\right\}$$

$$= 4.15 \text{ mJ/cm}^2$$

$$U_{D,t} = \pm\left[\left(U_t\frac{\partial D}{\partial t}\right)^2\right]^{1/2} = \pm[(4.15 \text{ mJ/cm}^2)^2]^{1/2} = \pm 4.15 \text{ mJ/cm}^2$$

   Percent = $100\, U_{D,t}/D = 100(4.15/249) = 1.67\%$

   b. Similarly, for the remaining variables, the corresponding values of the partial derivatives are as given below:

   $U_{D,I_o} = 17.44$ mJ/cm^2 and 7.0%

   $U_{D,P_f} = 5.30$ mJ/cm^2 and 2.13%

   $U_{D,k_{254}} = 1.40$ mJ/cm^2 and 0.56%

   $D_{D,d} = 1.21$ mJ/cm^2 and 0.49%

   $U_{D,L} = 0.076$ mJ/cm^2 and 0.03%

c. The best estimate of uncertainty using Eq. (12–67) is

$$U_D = \pm\,[(4.15)^2 + (17.44)^2 + (5.30)^2 + (1.40)^2 + (1.21)^2 + (0.076)^2]^{1/2}$$

$$= \pm 18.8 \text{ mJ/cm}^2$$

$$\text{Percent} = (100 \times 18.8)/249.0 = 7.55 \text{ percent}$$

3. Based on the above uncertainty computation the most likely UV dose is

$$D = 249 \pm 19 \text{ mJ/cm}^2$$

Comment Based on the best estimate of uncertainty, the most conservative estimate of the UV dose that can be delivered consistently, based on the collimated beam test, is 230 mJ/cm^2 (249 − 19). The maximum uncertainty would correspond to the summation of the individual errors and would be ± 30 mJ/cm^2.

Bioassay Testing. To assess the degree of microbial inactivation that can be achieved at a given UV dose, the concentration of microorganism is determined before and after exposure in a collimate beam (see Fig. 12–41). Microorganisms inactivation is measured using the most probable number (MPN) procedure or the membrane filtration test for bacteria, a plaque count procedure for viruses, or an animal infectivity procedure for protozoa. To verify the accuracy of the laboratory collimated beam dose-response test data, the collimated beam test must be repeated to obtain statistical significance. To be assured that stock solution of the challenge microorganisms is mono-dispersed, the laboratory inactivation test data must fall within an accepted set of quality control limits. Quality control limits proposed by the National Water Research Institute (NWRI, 2003) and the U.S. EPA (2003b) for Bacteriophage MS2 are as follows.

NWRI

$$\text{Upper bound: } -\log_{10}(N/N_o) = 0.040 \times D + 0.64 \tag{12–68a}$$

$$\text{Lower bound: } -\log_{10}(N/N_o) = 0.033 \times D + 0.20 \tag{12–68b}$$

U.S. EPA

$$\text{Upper bound: } -\log_{10}(N/N_o) = -9.6 \times 10^{-5} \times D^2 + 4.5 \times 10^{-2} \times D \tag{12–69a}$$

$$\text{Lower bound: } -\log_{10}(N/N_o) = -1.4 \times 10^{-4} \times D^2 + 7.6 \times 10^{-2} \times D \tag{12–69b}$$

where D = UV dose, mJ/cm^2

As will be illustrated in Example 12–12, the bounds proposed by the U.S. EPA are more lenient as compared to those used by NWRI. Similar bounding curves have been proposed for *B. subtilus* (U.S. EPA, 2003b). The NWRI guidelines are used for water reuse applications.

EXAMPLE 12–12 Verification of Laboratory Procedures for Bacteriophage MS2 Response The following collimated beam test results were obtained for a stock solution of bacteriophage MS2 which is to be used to test a UV reactor. These results are used to verify that the laboratory test results are acceptable and define the dose response equation.

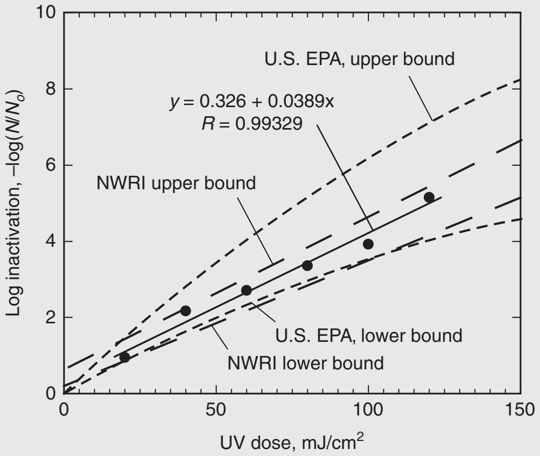
Dose, mJ/cm^2	Surviving concentration, phage/mL	Log survival,[a] log (phage/mL)	Log inactivation
0	1.00E + 07	7.000	0.000
20	1.12E + 06	6.049	0.951[b]
40	7.41 + 04	4.870	2.130
60	1.95E + 04	4.290	2.710
80	4.37E + 03	3.640	3.360
100	1.02E + 03	3.009	3.991
120	7.08E + 01	1.850	5.150

[a] The rule followed in the log transformation of a number is to retain in the mantissa the same number of significant figures as in the number that is being transformed.

[b] Sample calculation: Log inactivation = 7.000 − 6.049 = 0.951.

Solution

1. Plot the collimated beam test results and compare to the quality control range expressions provided in the NWRI [Eqs. 12–68(a) and 12–68(b)] and U.S. EPA [Eqs. 12–69(a) and 12–69(b]) UV Guidelines. The results are plotted on the figure given below.



2. As shown in the above plot all of data points fall within the acceptable range as defined by both NWRI and U.S. EPA.
3. Define the dose response relationship. Based on a linear regression analysis, the UV dose response relationship is

$$\text{UV dose} = \frac{\text{log inactivation} - 0.326}{0.0389}$$

Comment In general, when conducting bioassay testing, the initial concentration of MS2 should be 2-log higher than the number of logs of inactivation to be achieved. Irradiated samples should be diluted so that number of plaque forming units per plate is between 20 and 200 (NWRI, 2012).

Reporting and Using Bioassay Collimated Beam Test Results. The results of collimated beam bioassays are reported in the form of a dose response curve as developed in Example 12–12 and shown on Fig. 12–42. The inactivation curve shown on Fig. 12–42(a)

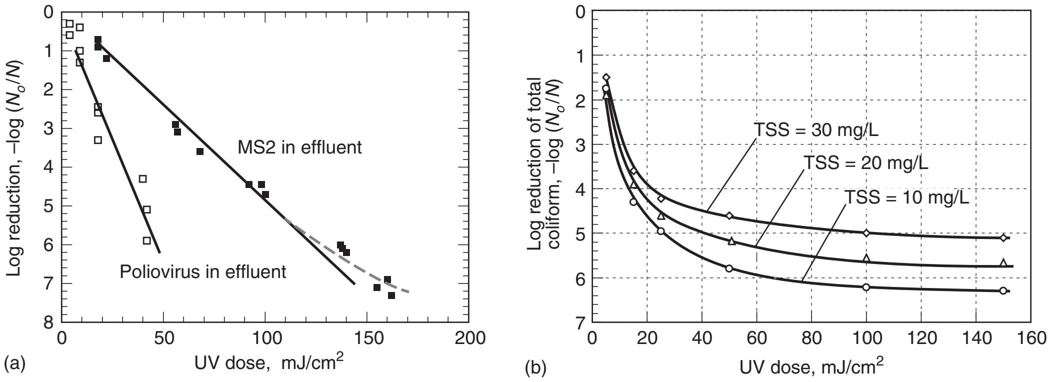


Figure 12–42

Typical dose response curves for UV disinfection developed from data obtained using a collimated beam device: (a) for dispersed microorganisms (Cooper et al., 2000) and (b) wastewater containing varying concentrations of TSS.

is for discrete organisms (MS2 and poliovirus) exposed to UV light, whereas the curve shown on Fig. 12–42(b) is for total coliform treated wastewater containing particulate material. In practice, the linear portion of the dose response curve for MS2 coliphage typically is between 20 and 120 mJ/cm^2. Below about 20 mJ/cm^2 there is uncertainty in the measurements and in the nature of the operative disinfection mechanism. Beyond about 120 mJ/cm^2 the presence of particles and/or particle clumping in wastewater samples causes a shoulder effect similar to the tailing effect observed with chlorine disinfection (see Fig. 12–6). In the literature, a polynomial curve passing through the origin is often used to fit all of the dose response data including the shoulder effect. The problem with a polynomial curve fit is that there is no theoretical basis for its use and the operative disinfection mechanisms are not the same at low and high UV doses. However, in the region where most UV reactors are tested, there is little difference between the linear and polynomial dose response curves. Additional details on the collimated beam protocol using MS2 coliphage can be found in the NWRI Guidelines (2012).

Ultraviolet Disinfection Guidelines

The National Water Research Institute and the American Water Works Association Research Foundation published "Ultraviolet Disinfection Guidelines for Drinking Water and Wastewater Reclamation" (NWRI, 1993; NWRI and AWWARF, 2000; NWRI, 2003; NWRI, 2012). The following elements are considered in the UV guidelines: (1) reactor design, (2) reliability design, (3) monitoring and alarm design, (4) the field commissioning test, (5) performance monitoring, and (6) an engineering report for unrestricted effluent reuse applications. Some of the items may not be applicable when utilizing UV disinfection for less demanding applications.

The guidelines that cover reclaimed water are similar to those that cover drinking water systems. The primary difference is that recommended (or mandatory) doses are provided for reclaimed water systems, whereas there is no mention of recommended doses for non-reclaimed wastewater applications. For reclaimed water systems, the recommended design UV doses for various effluents are 100 mJ/cm^2 for media filtration or equivalent effluent, 80 mJ/cm^2 for membrane filtration effluent, and 50 mJ/cm^2 for reverse osmosis effluent. The different dose requirements reflect the different virus density concentrations expected within each type of treatment process effluent. For example, the dosage of 100 mJ/cm^2 for media filtration effluent is intended to provide 5 logs of poliovirus inactivation with a factor of safety of about 2.

In addition to differing dose recommendations as a function of effluent quality, there are differing design transmittance recommendations. For granular medium and other types of filtration, microfiltration, and reverse osmosis effluents, the design transmittance's are 55, 65, and 90 percent, respectively. The differing transmittance values are based on field observations made to date, though site-specific variation does occur and should be accounted for. All UV disinfection systems installed for either drinking water or unrestricted reuse applications must undergo validation testing prior to their installation. Although the guidelines do not apply to the disinfection of non reclaimed wastewater, the general design issues addressed are applicable. The IUVA Manufacturer's Council has published a "low dose" bioassay approach (IUVA, 2011).

Relationship of UV Guidelines to UV System Design

The design of a UV disinfection system involves a number of issues including (1) determination of the UV dose required, based on bioassay testing, for adequate inactivation of the challenge (target) microorganism(s), (2) selection of manufacturer-specific validated UV disinfection reactors or systems, (3) determination of process operational parameters and UV system configuration (e.g., the number of lamps per module, modules per bank, banks per channel, and the overall number of channels) and, in some circumstances, (4) conduct of a spot-check bioassay test on the full-scale system to check compliance with the required UV design performance. For reclaimed wastewater applications, the first issue is addressed directly in the UV guidelines as discussed above. For the majority of applications which are not disinfecting for reclaiming water, an appropriate dose must be selected. Guidance on dose selection is provided in Table 12–33. Beyond this guidance, collimated beam tests should be performed on the actual wastewater to determine an appropriate dose to use for UV system design. A final resource for dose selection is information from UV equipment manufacturers, who maintain an extensive databases detailing dose requirements for varying disinfection limits, varying solids contents, and varying plant processes.

The general procedure for validating a UV reactor and some important guidance on design aspects are also included in the guidelines. Because of their fundamental importance in understanding the application of UV disinfection systems, these issues are discussed in the text and illustrated in the examples that follow.

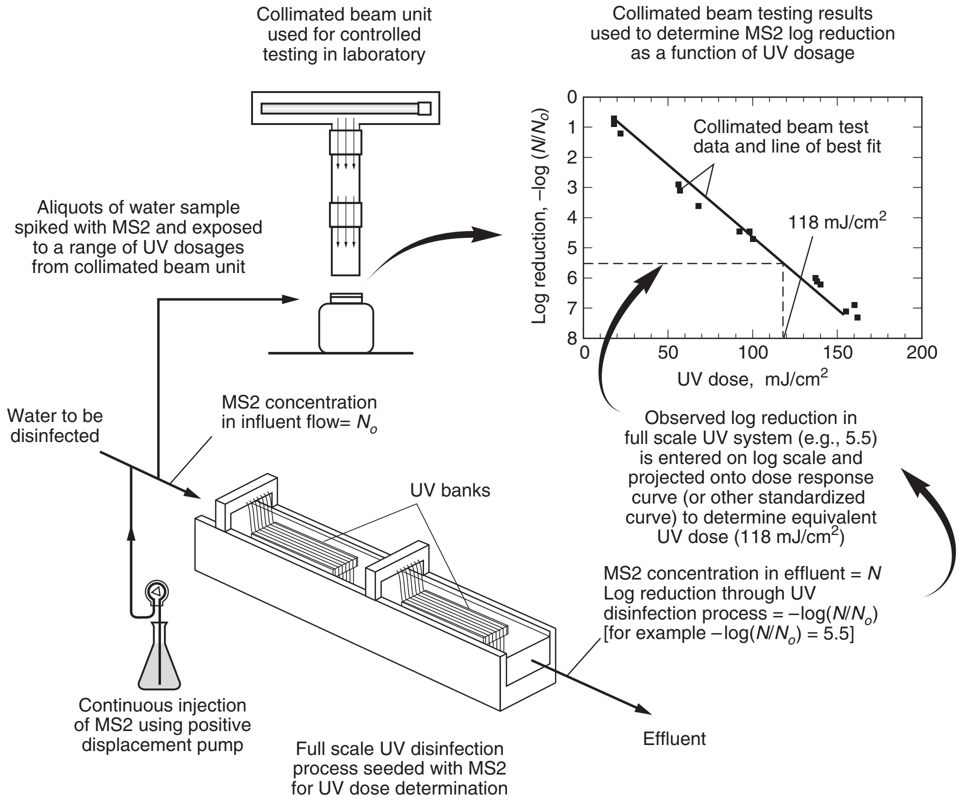
Validation of UV Reactor or System Performance

Validation testing consists of quantifying the level of inactivation of a virus surrogate (e.g., Bacteriophage MS2) by the UV disinfection reactor or system as a function of a number of process variables such as flowrate, transmittance, sensor settings, water level (where appropriate), and power settings. To quantify the inactivation achieved through the UV disinfection system, the UV dose response of the challenge microorganism to be used is determined using a collimated beam illustrated on Fig. 12–43. The inactivation observed through the UV disinfection reactor or system is compared to the UV dose response to establish a term called *reduction equivalent dose* (RED) or *delivered dose*, which corresponds to the UV dose delivered by the UV disinfection system. It should be noted that the RED is specific to the challenge organism and the test conditions.

In the past, validation testing was done once a UV system was installed and operational. To avoid unnecessary testing and the risk that an installed system does not perform adequately, validation testing is now typically completed by UV equipment manufacturers at test centers in the United States or at selected treatment facilities around the world. The manufacturers then provide design engineers with design information on which to base the

Figure 12–43

Schematic illustration of the application of biodosimetry as used to determine the performance of a test or full scale VU reactor.



design of a full-scale installation. The process flow diagram used for testing both open and closed UV reactors is illustrated on Fig. 12–44. In general, validation testing of UV disinfection equipment, using the setup shown on Fig. 12–44(a), consists of the following steps:

1. Selection of representative test water for use in the validation testing of the disinfection system.
2. Selection of the configuration of the UV disinfection system to be tested (i.e., 1, 2, 3, etc. UV banks in series). If the power to the UV lamps cannot be turned down to simulate the end of life lamp performance for a portion of the testing, then aged UV lamps must be used in the test.
3. Hydraulic performance testing of the UV disinfection system is done to verify the uniformity of the approach and exit velocities.
4. Quantification of the inactivation of the test organism (e.g., MS2) through the UV test reactor [see Fig. 12–44(b)] as a function of hydraulic loading rate and other variables.
5. Simultaneous with the field testing, a collimated beam test is conducted on the test water to determine the inactivation response of the viral test organism as a function of applied UV dose. The laboratory test data must fall within the area bounded by Eqs. (12–68a) and (12–68b) or Eqs. (12–69a) and (12–69b) given previously.
6. Assign UV doses to the pilot reactor or system based on the standardized dose response relationship (see NWRI, 2012). In the past, the dose response relationship developed from the collimated beam test was used.
7. Based on the assigned UV dose and the operative control parameters, manufacturers will develop design equations for the test reactor or system.

The steps required in conducting a validation test are illustrated in Example 12–13.

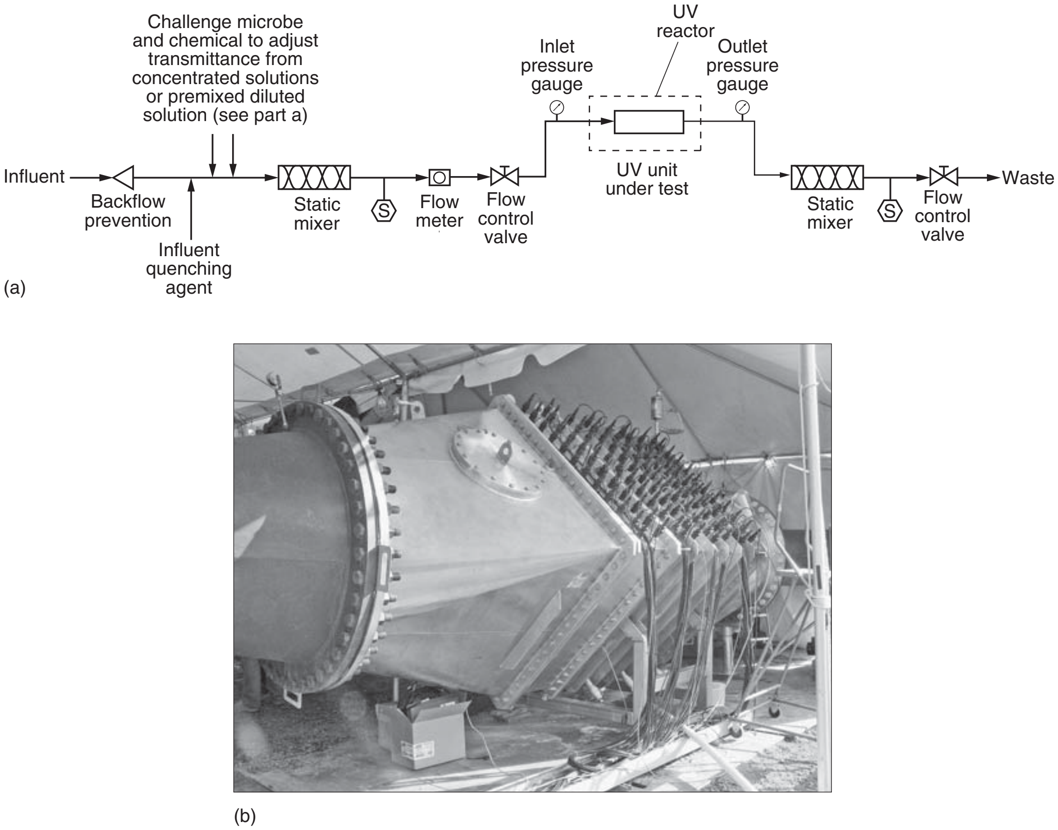


Figure 12–44

Validation testing of UV reactors: (a) schematic of the experimental test setup and (b) view of large closed UV reactor undergoing validation testing.

EXAMPLE 12–13 **Analysis of Pilot Test Results Used to Validate Performance of UV Reactor or Disinfection System** A manufacturer has supplied a pilot scale UV disinfection system whose performance is to be validated as a function of lamp hydraulic loading rate and water quality only. Other important variables such as power variation and water level variation are not included in this example. Operational curves are to be based on flowrate alone and flowrate and transmittance. For this test, the manufacturer chose to make use of a four-lamp per bank pilot facility with three banks in series to achieve the total applied dose. Each bank of lamps is hydraulically independent of subsequent banks. The engineer and owner are interested in knowing the range of flowrates and water quality over which the UV system can deliver a UV dose of 80 mJ/cm^2, before any design correction factors are applied. Assume the MS2 UV dose response curve given in Example 12–12 will be used for the analysis of the test results. The test program and the results of the field tests are as follows.

Solution

1. Develop test program.

   The testing was conducted on tertiary effluent from a local water reclamation facility. New lamps were placed in the pilot facility, as it would take more than a year to age the lamps. Normal transmittance of the tertiary effluent used is 75 percent. The first series of tests was conducted at a transmittance of 75 percent. In the second series of tests a transmittance reducing agent (e.g., SuperHume® or coffee) was injected into the effluent stream to lower the transmittance to 55 percent. The manufacturer has specified that the UV disinfection system should be tested for hydraulic loading rates ranging from 20 to 80 L/min·lamp, calculated as the flow in L/min·bank divided by the number of lamps in one bank. It should be noted that in a three-bank system, often each bank is tested separately to determine whether there are any inlet or outlet hydraulic conditions that would impact performance.

Because the titer of the virus indicator (i.e., MS2 bacteriophage) to be used for performance testing was approximately 1 × 10^{11} phage/mL, it was decided to test the system under the conditions outlined in the following table.

Hydraulic loading rate, L/min·lamp (1)	Flowrate, L/min·bank (2)	Virus titer Concentration, phage/mL (3)	Virus titer injection flowrate, L/min (4)	Approximate resulting virus concentration in process flow, phage/mL (5)
20	80	1E+11	0.008	1E+7
40	160	1E+11	0.016	1E+7
60	240	1E+11	0.024	1E+7
80	320	1E+11	0.032	1E+7

Notes on column entries:

(1) Desired range to be tested as specified by the manufacturer.

(2) The pilot system contained three banks with 12 lamps total; however, the hydraulic loading rate is only based upon the flowrate through one bank, which makes the calculation more similar to a velocity determination. Thus, at a hydraulic loading rate of 20 L/min·lamp, the process flowrate is equal to 80 L/min·bank [(20 L/min·lamp)(4 lamps/bank)].

(3) Provided by the laboratory.

(4) It was desired to obtain a virus titer in the process flow of about 1 × 10^7 phage/mL. Therefore, at 80 L/min, the solution containing the virus had to be injected at a rate of 0.008 L/min to obtain the desired initial titer.

2. Test results at 75 percent transmittance.

In conducting the test, each flowrate was tested randomly with respect to order. Three distinct replicate samples were collected per flowrate. An inlet and outlet sample (i.e., that contained the concentration of phage prior to any inactivation) was collected with each process replicate.

a. The inlet test results at 75 percent transmittance are as follows:

Flowrate, L/min·lamp	Replicate	Inlet concentration, phage/mL	Log-transformed inlet conc., log(phage/mL)[a]	Average log-transformed inlet conc., log(phage/mL)
20	1	5.25E+06	6.720	
20	2	1.00E+07	7.000	6.927
20	3	1.15E+07	7.061	
40	1	1.00E+07	7.000	
40	2	1.23E+07	7.090	7.067
40	3	1.29E+07	7.111	

(*continued*)

(*Continued*)

Flowrate, L/min·lamp	Replicate	Inlet concentration, phage/mL	Log-transformed inlet conc., log(phage/mL)[a]	Average log-transformed inlet conc., log(phage/mL)
60	1	1.23E+07	7.090	
60	2	1.05E+07	7.021	7.030
60	3	9.55E+06	6.980	
80	1	1.23E+07	7.090	
80	2	1.20E+07	7.079	7.023
80	3	7.94E+06	6.900	

[a] The rule followed in the log transformation of a number is to retain in the mantissa the same number of significant figures as in the number that is being transformed.

b. The outlet test results at 75 percent transmittance, based on triplicate samples, are as follows. Only the average log-transformed outlet concentration values from the 75 percent transmittance test are given. The procedure followed in obtaining these values was the same as illustrated above for the inlet test results.

Flowrate, L/min·lamp	Number of banks	Average log-transformed outlet conc., log(phage/mL)
20	2[a]	2.233
40	3	1.832
60	3	3.232
80	3	3.591

[a] Notice that at the low flowrate investigated (20 L/min·lamp), only two operational banks were investigated rather than 3. Only two banks were tested because three operational banks resulted in no detectable viruses in the effluent. Because the banks were hydraulically independent, it is allowed under the UV Guidelines to investigate the inactivation for only two banks and extrapolate to performance expected for additional banks of lamps.

3. Test results at 55 percent transmittance.
For the purposes of this example, assume the average log-transformed inlet concentration values from the 75 percent transmittance test apply to the 55 percent transmittance test. Only the average of the triplicate log-transformed outlet concentration values from the 55 percent transmittance test are given. The procedure followed in obtaining these values was the same as illustrated above for the 75 percent test.

Flowrate, L/min·lamp	Number of banks	Average log-transformed outlet conc., log(phage/mL)
20	3	1.703
40	3	3.987
60	3	4.662
80	3	4.997

4. Using the test data and the given information develop the necessary UV regression equation for 75 percent transmittance based on **flowrate** only.
   a. Set up computation table to determine the UV dose based on the test results. Using the measured phage data determine the corresponding UV dose based on the log-linear regression expression developed in Example 12–12.

Flowrate, L/min·lamp	Average phage concentration, log(phage/mL)			Assigned UV dose, mJ/cm²	Log transformed values	
	Inlet	Outlet	Diff.		Flowrate	UV dose
20	6.927	2.233	7.041[a]	172.6[b]	1.301	2.237
40	7.067	1.832	5.235	126.2	1.602	2.101
60	7.03	3.232	3.798	89.3	1.778	1.951
80	7.023	3.591	3.432	79.8	1.903	1.902

[a] The inactivation for this flowrate was extrapolated from the two-bank results. Because the system is a three-bank system, the inactivation for three banks is 150 percent greater than the inactivation observed with two operational banks [7.041 = (6.927 − 2.233) × 1.5].

[b] Sample calculation. Using the linear regression expression derived from the collimated beam test in Example 12–12, the equivalent UV dose at a flowrate of 20 L/min·lamp is:

$$\text{UV dose} = \frac{\text{log inactivation} - 0.326}{0.0389}$$

$$\text{UV dose, mJ/cm}^2 = \frac{7.041 - 0.326}{0.0389} = 172.6$$

   b. Develop the UV operational design equation.
      i. Use a linear regression analysis to develop a regression equation based on water flowrate. Other equations are possible, depending on the control strategy (e.g., flowrate and transmittance, flowrate, transmittance, and power setting).
      ii. To complete a regression analysis, the flowrate and UV dose data must first be log transformed. The data are log transformed to develop a linear relationship that can be used with the linear dose response curve developed using the collimated beam (see Example 12–12). The log transformed data are presented in columns 6 and 7 in the table developed in Step 4.
      iii. Using the UV dose (column 7) as the dependent variable and the flowrate (column 6) as the independent variable, the following results are obtained using the linear regression analysis program in Excel or other statistical analysis program.

Model Parameters

Source	Value
Intercept	2.997
X1	−0.577

      iv. The equation for UV dose as a function of flowrate, based on the regression analysis, is

$$\log(\text{UV dose}) = 2.997 - 0.577\,(\log \text{flowrate}) \text{ or}$$

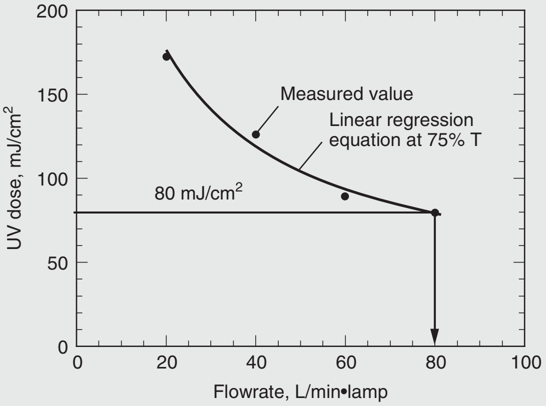
$$\text{UV dose, mJ/cm}^2 = (10^{2.997})[(\text{flowrate})^{-0.577}]$$

where the unit for flowrate is L/min·lamp

Note: The above dose equation is the delivered dose based upon three UV banks operating in series and with a UV transmittance (UVT) of 75 percent.

If the test had been conducted with one UV bank, the UV dose for two or three banks would be obtained by multiplying the regression equation for one UV bank by 2 or 3, respectively.

c. Plot the regression equation for UV dose versus the UV lamp hydraulic loading rate based on the results of the single variable (i.e., flowrate) linear regression analysis.



d. Determine the range of flowrates over which the UV disinfection system will deliver a UV dose of 80 mJ/cm^2. From the plot given above, the range of flows is up to 80 L/min·lamp.

5. Using the test data and the given information develop the necessary UV design curve based on **flowrate and transmittance**.

a. Set up computation table to determine the UV dose based on the test results. Using the measured phage data determine the corresponding UV dose based on the linear regression expression developed in Example 12–12.

Flowrate, L/min·lamp	T, %	Average phage concentration, log(phage/mL) Inlet	Outlet	Diff.	Assigned UV dose, mJ/cm^2	Log transformed values Flowrate	Transmittance	UV dose
20	75	6.927	2.233	7.041	172.6	1.301	1.875	2.237
40	75	7.067	1.832	5.235	126.2	1.602	1.875	2.101
60	75	7.03	3.232	3.798	89.3	1.778	1.875	1.951
80	75	7.023	3.591	3.432	79.8	1.903	1.875	1.902
20	55	6.927	1.703	5.224	125.9	1.301	1.740	2.100
40	55	7.067	3.987	3.08	70.8	1.602	1.740	1.850
60	55	7.03	4.662	2.368	52.5	1.778	1.740	1.720
80	55	7.023	4.997	2.026	43.7	1.903	1.740	1.640

b. Develop the operational design equation.

i. Use a linear regression analysis to develop an operational equation based on water flowrate and transmittance.

ii. To complete a regression analysis, the flowrate and UV dose data must first be log transformed. The log transformed data are presented in columns 7, 8, and 9 in the above table.

iii. Using the UV dose (column 9) as the dependent variable and the flowrate (column 7) and transmittance (column 8) as the independent variables, the following results are obtained using the linear regression analysis program in Excel or other statistical analysis program.

Model Parameters

Source	Value
Intercept	0.097
X1	−0.673
X2	1.631

iv. The equation for UV dose as a function of flowrate and transmittance, based on the linear regression analysis, is

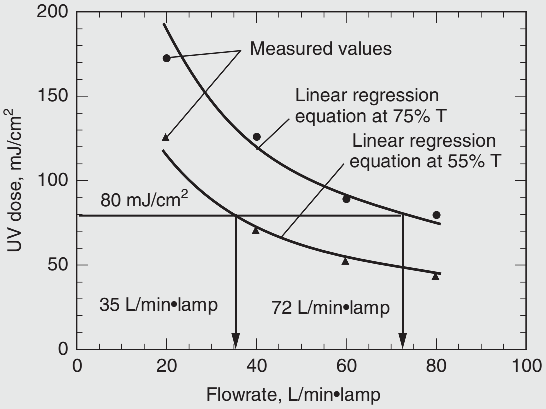
$$\log(\text{UV dose}) = 0.097 - 0.673\,(\log \text{flowrate}) + 1.631\,(\log \text{transmittance}) \text{ or}$$

$$\text{UV dose, mJ/cm}^2 = (10^{0.097})[(\text{flowrate})^{-0.673}][(\text{transmittance})^{1.631}]$$

where the units for flowrate and transmittance are L/min•lamp and percent, respectively.

Note: The above dose equation is the delivered UV dose based upon three banks operating in series and with UVT varying from 55 to 75 percent. If the UV validation test had been conducted with one bank, the UV dose for two or three banks would be obtained by multiplying the regression equation for one bank by 2 or 3, respectively.

c. Based on the results of the multiple variable (e.g., flowrate and transmittance) linear regression analysis, plot the curves of UV dose versus the UV lamp hydraulic loading rate for 75 and 55 transmittance, The required curves are shown on the following plot. It should be noted that the curve resulting from the multiple linear regression analysis for 75 percent transmittance is not exactly the same as that derived from the single variable (i.e., flowrate) linear regression analysis developed in Step 5. The reason for the difference is that the regression analysis with two variables must cover a significantly broader range of values as compared to a single variable regression analysis.



d. Determine the range of flowrates over which the UV disinfection system will deliver a UV dose of 80 mJ/cm². From the plot given above, the range of flowrates is up to 72 L/min·lamp at 75 percent transmittance and up to 35 L/min·lamp at 55 percent transmittance.

Comment When the lamps are new and the protective quartz sleeves are clean, it may not be necessary to operate all three banks, depending on the actual UV dose requirements of the full scale disinfection system.

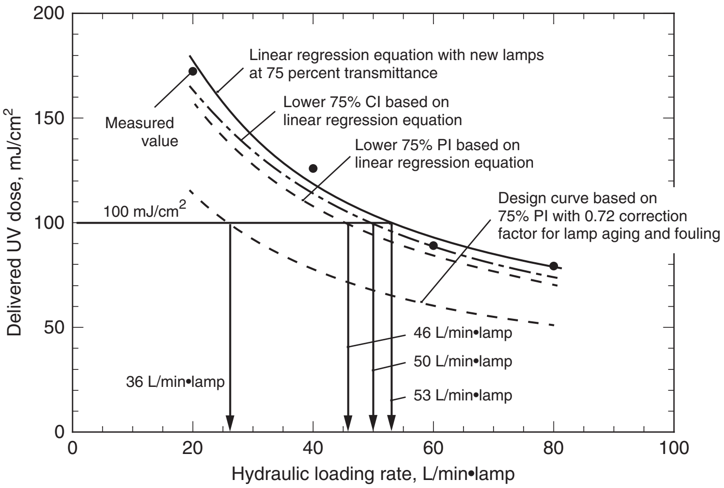
Factors Affecting UV System Design

Factors that affect the minimum number of UV lamps necessary for disinfection are (1) the UV lamp hydraulic loading rate based on the equipment validation test, (2) the level of confidence desired in meeting the permit requirements, and (3) the aging and fouling characteristics of the UV lamp/quartz sleeve assembly (discussed below). The validation of UV equipment has been considered in Example 12–13.

Confidence Level in Meeting Permit Limits. With respect to the level of confidence desired in the system performance it should be noted that the linear regression equations developed in Example 12–13 correspond to the line of best fit with half of the data points lying above and half lying below the predicted curve. Because some of the actual data points lie below the regression equation, a factor of safety must be used to account for the observed variability. One approach is to determine the confidence interval (CI) of the regression equation. Another is to develop a prediction interval (PI) based on the regression equation. The difference between the CI and the PI is as follows. The upper and lower CI for the regression analysis represents the interval in which the true average measurement is likely to lie if the procedure were repeated many times. Stated differently, a 75 percent confidence interval will contain the true mean value (not estimated value from data measurements) 75 percent of the time the interval is calculated. The upper and lower PI represents the interval in which a given percentage of new observations, independent of those used to develop the regression equation, will lie. Because there is more uncertainty in future measurements, the interval between the upper and lower PI limits is greater than that for the confidence interval, which is based on repeating the procedure an infinite number of times. The development of the PI is illustrated in Example 12–14. The relationship between the CI and PI and the regression equation are illustrated on Fig. 12–45.

Figure 12–45

Definition sketch illustrating the relationship between (1) the values measured in the UV reactor validation test, (2) the linear regression equation based on the measured values, (3) the lower 75 percent confidence interval (CI) based on the linear regression, (4) the lower 75 percent prediction interval (PI) based on the linear regression, and (5) the design curve based on the lower 75 percent PI with a combined correction factor for lamp aging and fouling.



Lamp Aging. As UV lamps age, the UV output from the lamps decreases. Aging factors range from 0.5 (NWRI default value in the absence of validation data to support a higher factor) to 0.98. However, manufacturers may gain approval of a higher (less conservative) factor if sufficient data are available to support an increase. As discussed previously, UV reactor validation is conducted with new lamps. Thus, when sizing a UV system, the validated performance must be de-rated by the lamp aging factor. For example, if 100 lamps are required to deliver a certain dose at a specific UVT, the UV system would need to be sized with 200 lamps so that when the lamps have aged and are producing 50 percent UV output compared to when they were new, the dose critical for disinfection is still being delivered. Lamp aging factors vary widely from vendor to vendor and from lamp to lamp. Aging factors range from 0.5 to 0.98.

Quartz Sleeve Fouling Factor. The correction factor for quartz sleeve fouling will vary between 0.7 and 0.95 depending on the cleaning system that is employed. For applications with high UVT, low solids and little iron present in the effluent, it is generally appropriate to accept a UV system vendor's validation of a higher factor (as high as 0.95 for mechanical cleaning systems). For chemical-mechanical cleaning system the fouling factor can be as high as 0.95 and should be validated independently of the UV system manufacturer. For any combination of low UVT, high solids, and high iron concentration in the effluent, a 0.8 (or lower) factor should be applied when sizing a UV system regardless of whether a UV system manufacturer has validation of a higher factor.

Application of Design Factors in UV System Sizing. With the variation of lamp aging factors from 0.5 to 0.98 and quartz sleeve fouling factors from 0.7 to 0.95, the combined design or correction factor can range from 0.35 to 0.94. The correction factor is applied to the PI to obtain the final design curve for the disinfection system. The importance of these correction factors can be assessed from a review of the plot given on Fig. 12–45. The design curve given on Fig. 12–45 is based on the lower 75 percent PI with a combined correction factor of 0.72 for lamp aging and fouling (0.72 factor based on an lamp aging factor of 0.9 and a quartz sleeve fouling factor of 0.8). Clearly, if the UV system had been designed on the basis of the manufacturer's design curve, the system would be undersized with respect to lamp aging and fouling. Determination of confidence and prediction intervals and the development of design equations taking into account lamp aging and fouling are illustrated in Example 12–14.

EXAMPLE 12–14 **Development of Operational UV Design Curves Taking into Account Variability and Aging and Fouling** Using the information from Example 12–13, develop design equations based on the PI alone and the PI with a factor for lamp aging and fouling. Also, determine the range of hydraulic loading rates over which the system can deliver a UV dose of 80 mJ/cm^2, based on the lower 75 percent PI with new lamps.

Solution Part A—Design equation based on flowrate alone

1. Define the lower 75 percent CI and PI limits for the regression equation developed in Example 12–13 based on flowrate alone. The regression equation is

   $$\text{UV dose, mJ/cm}^2 = (10^{2.997})[(\text{flowrate})^{-0.577}]$$

   where the unit for flowrate is L/min·lamp

   Based on the statistical analysis presented in Example 12–13, the assigned UV dose based on the field measurements, the predicted UV dose, the lower 75% CI, and the 75% PI values log transformed are given in the following table:

	Log-transformed values			
Flowrate, L/min·lamp	Assigned UV dose, mJ/cm^2	Predicted UV dose, mJ/cm^2	Predicted, 75% CI UV dose, mJ/cm^2	Predicted, 75% PI UV dose, mJ/cm^2
20	2.237	2.247	2.209	2.191
40	2.101	2.073	2.052	2.027
60	1.951	1.972	1.948	1.924
80	1.902	1.900	1.869	1.848

Although the lower 75 percent CI and PI values are obtained with a standard statistical program as given above, the procedure for determining these values is illustrated below for a linear regression expression with one variable–in this case, flowrate per lamp.

a. The lower 75 percent CI and PI intervals for the predicted mean response can be obtained using the following expressions:

   i. Confidence interval

$$\text{UV dose}_{75\%} = y_p - t_{\alpha/2} S \sqrt{\frac{1}{n} + \frac{(x - \bar{x})^2}{SS_{xx}}}$$

   ii. Prediction interval

$$\text{UV dose}_{75\%} = y_p - t_{\alpha/2} S \sqrt{1 + \frac{1}{n} + \frac{(x - \bar{x})^2}{SS_{xx}}}$$

where y_p = the predicted UV dose computed using the regression equation given above, mJ/cm^2
$t_{\alpha/2}$ = 1.706 which corresponds to the value of the t-distribution based on a 75% prediction level with $n - 2$ degrees of freedom
S = sample variance

$$S = \sqrt{\frac{\Sigma(y - y_p)^2}{n - 2}}$$

y = assigned UV dose from field measurements, mJ/cm^2
y_p = predicted UV dose, mJ/cm^2
n = number of sample pairs
x = flowrate, L/min·lamp
$\bar{x}$ = average flowrate, L/min·lamp
SS_{xx} = the sample corrected sum of squares

$$SS_{xx} = \sum_{1}^{n} (x - \bar{x})^2$$

b. Compute the values needed to determine the confidence intervals. Set up two computation tables, one for the UV dose and another for flowrate.

x	y	y_p	$(y - y_p)$	$(y - y_p)^2$
1.301	2.237	2.247	−0.010	0.000100
1.602	2.101	2.073	0.028	0.000784
1.778	1.951	1.972	−0.021	0.000441
1.903	1.902	1.900	0.002	0.000004
				0.001329

x	$(x - \bar{x})^a$	$(x - \bar{x})^2$
1.301	−0.345	0.119025
1.602	−0.044	0.001936
1.778	0.132	0.017424
1.903	0.257	0.066049
6.584		0.204434

[a] $\bar{x} = 6.584/4 = 1.646$

i. Solve for the sample variance, S.

$$S = \sqrt{\frac{0.001329}{4 - 2}} = 0.025778$$

ii. Solve for sample corrected sum of squares, SS_{xx}.

$$SS_{xx} = \sum_{1}^{n} (x - \bar{x})^2 = 0.204434$$

iii. Solve for the lower 75 percent CI at a flowrate of 40 L/min•lamp.

$$\text{UV dose}_{75\%\,\text{CI}} = y_p - t_{\alpha/2} S \sqrt{\frac{1}{n} + \frac{(x - \bar{x})^2}{SS_{xx}}}$$

The value of y_p, computed using the regression equation given above, is equal to 2.073. Thus,

$$\text{UV dose}_{75\%\,\text{CI}} = 2.073 - (1.706)(0.024434)\sqrt{\frac{1}{4} + \frac{0.001936}{0.204434}}$$

$$\text{UV dose}_{75\%\,\text{CI}} = 2.073 - (1.706)(0.024434)(0.509382) = 2.052$$

iv. Solve for the lower 75 percent PI at a flowrate of 40 L/min•lamp.

$$\text{UV dose}_{75\%\,\text{PI}} = y_p - t_{\alpha/2} S \sqrt{1 + \frac{1}{n} + \frac{(x - \bar{x})^2}{SS_{xx}}}$$

The value of y_p, computed using the regression equation given above, is equal to 2.073. Thus,

$$\text{UV dose}_{75\%\,\text{CI}} = 2.073 - (1.706)(0.024434)\sqrt{1 + \frac{1}{4} + \frac{0.001936}{0.204434}}$$

$$\text{UV dose}_{75\%\,\text{PI}} = 2.073 - (1.706)(0.024443)(1.122248) = 2.026$$

The CI and PI values computed manually are essentially the same as the values obtained from the linear regression analysis program as given in Step 1. Values computed manually may not be exact due to rounding errors that are magnified when dealing with log-transformed values.

2. Correct the lower 75 percent PI values for lamp aging and fouling.

   a. An overall correction factor of 0.72 is assumed for account for lamp aging and fouling. The correction factor for lamp aging is 0.9, based on the manufacturers recommendation. The corresponding fouling factor is 0.8. Note: the design engineer must decide if additional factors of safety may be required, depending on local conditions.

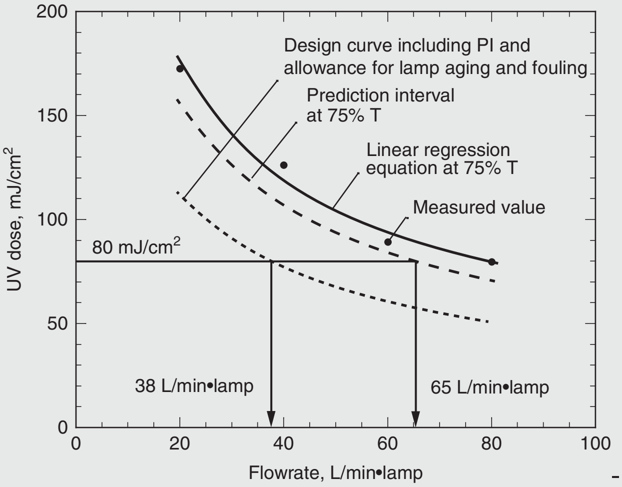
   The UV dose based on lamp aging and fouling is given in following table in which the log-transformed values have been transformed back to arithmetic form.

Flowrate, L/min·lamp	Assigned UV dose, mJ/cm^2	Predicted UV dose,[a] mJ/cm^2	Predicted UV dose at 75% PI, mJ/cm^2	Correction factor for lamp aging and fouling[b]	Design UVdose, mJ/cm^2
20	172.6	176.5	155.3	0.72	111.8
40	126.2	118.3	106.5	0.72	76.7
60	89.3	93.7	84.0	0.72	60.5
80	79.8	79.4	70.5	0.72	50.8

[a] From regression equation.

[b] Correction factor = Lamp aging factor (0.9) × fouling factor (0.8).

b. The measured values, the linear regression equation, the 75 percent PI curve, and the design curve based on the 75 percent PI and taking into account lamp aging and fouling are plotted on the following graph for 75 percent transmittance.



c. Determine the range of flowrates over which the UV disinfection system will deliver a UV dose of 80 mJ/cm^2, based on the lower 75 percent PI with new lamps. From the plot given above, the range of flowrates is up to 65 L/min·lamp at 75 percent transmittance.

3. Develop the design equations for the 75 percent PI curve, and the design curve based on the 75 percent PI and taking into account lamp aging and fouling for a transmittance value of 72 percent. The required equations can be be obtained by noting the ratio of the predicted PI UV dose to the predicted UV dose and the ratio of the design UV dose to the predicted UV dose as illustrated in the following table:

Predicted UV dose,[a] mJ/cm^2	Predicted UV dose at 75% PI, mJ/cm^2	Design UV dose,[b] mJ/cm^2	Ratio, PI/predicted UV dose	Ratio, Design/predicted UV dose
176.5	155.3	111.8	0.88	0.63
118.3	106.5	76.7	0.90	0.65
93.7	84.0	60.5	0.90	0.65
79.4	70.5	50.8	0.89	0.63

[a] From regression equation.

[b] Design equation based on PI with correction factor for lamp aging and fouling.

The ratios in the above table are not exact, because the prediction interval at the extremes of the range is greater than for the centermost values. Use of the ratio for the extremes of the range is conservative. Thus, the pertinent equations are

Design equation based on 75 percent PI

$$\text{UV dose, mJ/cm}^2 = (10^{2.997})[(\text{flowrate})^{-0.577}](0.88)$$

Design equation based on 75 percent PI and including lamp aging and fouling

$$\text{UV dose, mJ/cm}^2 = (10^{2.997})[(\text{flowrate})^{-0.577}](0.63)$$

where the unit for flowrate is L/min·lamp

Solution Part B—Design equation based on flowrate and transmittance

1. Define the lower 75 percent CI and PI limits for the regression equation developed in Example 12–13 based on flowrate and transmittance. The regression equation is

$$\text{UV dose, mJ/cm}^2 = (10^{0.097})[(\text{flowrate})^{-0.673}][(\text{transmittance})^{1.631}]$$

The computational procedure for the CI and PI values is similar to that illustrated above for a linear regression with one variable. In a multiple linear regression analysis, the computation of the CI and PI is more complicated because more three terms are involved. For this reason, the CI and PI values are usually determined using a standard statistical analysis program. The lower 75 percent CI and PI values for the regression equation are summarized in the following table for 75 and 55 percent transmittance.

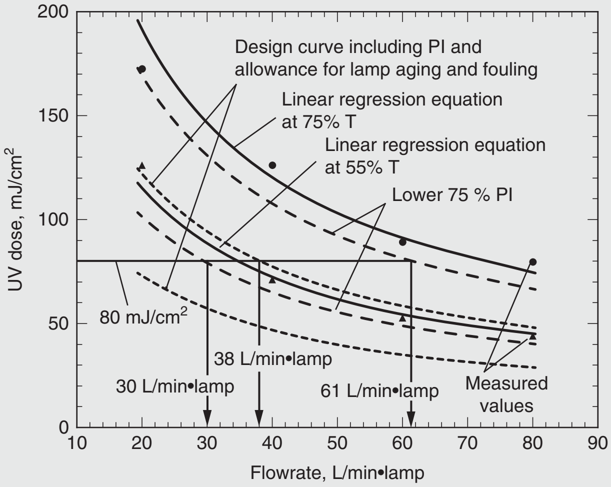
Log-transformed values				
Flowrate, L/min·lamp	Assigned UV dose, mJ/cm²	Predicted UV dose, mJ/cm²	Predicted, 75% CI UV dose, mJ/cm²	Predicted, 75% PI UV dose, mJ/cm²
75% transmittance				
1.30	2.237	2.280	2.248	2.227
1.60	2.101	2.077	2.056	2.029
1.78	1.951	1.959	1.936	1.910
1.90	1.902	1.875	1.847	1.824
55% transmittance				
1.30	2.100	2.060	2.028	2.006
1.60	1.850	1.857	1.835	1.809
1.78	1.720	1.739	1.715	1.690
1.90	1.640	1.654	1.627	1.604

2. Correct the lower 75 percent PI values for lamp aging and fouling.
   a. To account for lamp aging and fouling a correction factor of 0.72 will be applied. Note: the design engineer must decide if additional factors of safety may be required, depending on local conditions.

      The UV dose based on lamp aging and fouling is given in the following table in which the log-transformed values have been transformed back to arithmetic form.

Flowrate, L/min·lamp	Assigned UV dose, mJ/cm^2	Predicted UV dose, mJ/cm^2	Predicted UV dose at 75% PI, mJ/cm^2	Correction factor for lamp aging and fouling	Design UV dose, mJ/cm^2
75% transmittance					
20	172.6	190.5	168.5	0.72	121.4
40	126.2	119.5	107.0	0.72	77.0
60	89.3	91.0	81.3	0.72	58.5
80	79.8	74.9	66.6	0.72	48.0
55% transmittance					
20	125.9	114.7	101.5	0.72	73.1
40	70.8	72.0	64.4	0.72	46.4
60	52.5	54.8	49.0	0.72	35.3
80	43.7	45.1	40.1	0.72	28.9

b. The linear regression equation, the 75 percent PI curve, and the design curve taking into account lamp aging and fouling based on the lower 75 percent PI are plotted on the following graph for transmittance values of 75 and 55 percent.



c. Determine the range of flowrates over which the UV disinfection system will deliver a UV dose of 80 mJ/cm^2, based on the lower 75 percent PI with new lamps. From the plot given above the range of flowrates is up to 61 L/min·lamp at 75 percent transmittance and up to 30 L/min·lamp at 55 percent transmittance. The range of flowrates per lamp at 75 percent transmittance with aged and fouled lamps is up to 38 L/min·lamp.

3. Develop the design equations for the 75 percent PI curve, and the design curve based on the 75 percent PI and taking into account lamp aging and fouling for transmittance values varying from 55 to 75 percent. The required equations, developed as outlined above in Part 1, Step 4, are:

Design equation based on 75 percent PI.

$$\text{UV dose, mJ/cm}^2 = (10^{0.198})[(\text{flowrate})^{-0.674}][(\text{transmittance})^{1.5713}](0.88)$$

Design equation based on 75 percent PI and including lamp aging and fouling.

$$\text{UV dose, mJ/cm}^2 = (10^{0.198})[(\text{flowrate})^{-0.674}][(\text{transmittance})^{1.5713}](0.64)$$

where the units for flowrate and transmittance are L/min·lamp and percent, respectively.

Comment In both cases presented above and in previous sections discussing the impact of lamp aging and quartz sleeve fouling, the allowance made for these correction factors is significant as compared to the lower 75 percent PI. Thus, in evaluating UV disinfection systems, the selection of appropriate and validated lamp aging and fouling factors is of critical importance.

Selection and Sizing of a UV Disinfection System

Factors that affect the selection and sizing of a UV disinfection system include the selection and sizing of the UV disinfection reactor or system based on the UV design curve which takes into account the confidence and/or prediction interval associated with the hydraulic loading rate as determined in the equipment validation test and the lamp aging and fouling correction factors as illustrated in Example 12–13. The selection and sizing procedure for a UV disinfection system is illustrated in Example 12–15.

EXAMPLE 12–15 **Design of a UV Disinfection System for Secondary Effluent** Design a UV disinfection system for secondary effluent that will deliver a minimum design dose of 30 mJ/cm^2. Assume for the purpose of this example that the following data apply:

1. Wastewater characteristics
   a. Average design flowrate = 40,000 m^3/d = 27,778 L/min
   b. Maximum design flowrate = 100,000 m^3/d = 69,444 L/min (peak hour flow with recycle streams)
   c. Maximum total suspended solids = 20 mg/L
   d. Minimum transmittance = 65%
2. Fecal coliform discharge limit based on geometric mean
   200 FC/100 mL
3. System characteristics
   a. Horizontal lamp configuration
   b. From a validation study conducted on a single UV bank using the procedure described in Example 12–12, the following equation was developed based on the 75 percent PI with a lamp aging and fouling factor allowance of 72 percent.

   $$\text{UV dose, mJ/cm}^2 = (10^{-2.428})[(\text{flowrate})^{-0.650}][(\text{transmittance})^{3.126}](0.64)$$

   where the units for flowrate and transmittance are L/min·lamp and percent, respectively
   c. System headloss coefficient = 0.75 (manufacturer specific)
   d. Lamp/sleeve diameter = 23 mm
   e. Cross-sectional area of quartz sleeve = 4.15×10^{-4} m^2
   f. Lamp spacing = 75 mm (center to center)
   g. One standby UV bank will be required per channel

Solution

1. Determine the flowrate per lamp using the design equation based on the test conducted on a single UV bank. Based on the UV design equation for a dose of 30 mJ/cm^2, the corresponding flowrate per lamp is 258 L/min·lamp.

$$\text{Flowrate, L/min·lamp} = \left\{\frac{30}{(10^{-2.428})[(65)^{3.126}](0.64)}\right\}^{-(1/0.650)} = 258$$

2. Specify the flowrate range per UV channel assuming three channels will be in operation during peak flow conditions.
   i. Up to 24,000 L/min, use one channel
   ii. From 24,000 to 48,000 L/min, split the flow between two channels such that each channel receives up to 24,000 L/min.
   iii. From 48,000 to 72,000 L/min, split the flow between three channels such that each channel receives up to 24,000 L/min.
3. Determine the number of lamps required per bank.
   At 24,000 L/min, the total number of required lamps is:

$$\text{Lamps required, Lamps/bank} = \frac{(24{,}000\ \text{L/min·bank})}{(258\ \text{L/lamp·min})} = 93\ \text{lamps/bank}$$

4. Configure the UV disinfection system.
   Typically, 2, 4, 8, or 16 lamps per module are available. Using an 8 lamp module, 12 modules are required per bank for a total of 96 lamps per bank.

5. Determine the total number of lamps per channel including standby.

$$\begin{aligned}\text{Total number of lamps per channel} &= (2\ \text{banks/channel})(96\ \text{lamps/bank})\\ &= 192\ \text{lamps/channel}\end{aligned}$$

6. Determine total number of lamps.

$$\begin{aligned}\text{Total number of lamps per channel} &= (3\ \text{channel})(192\ \text{lamps/channel})\\ &= 586\ \text{lamps/channel}\end{aligned}$$

7. Check whether the headloss for the selected configuration is acceptable.
   a. Determine the channel cross-sectional area.

$$\begin{aligned}\text{Cross sectional area of channel} &= (12 \times 0.075\ \text{m})(8 \times 0.075\ \text{m})\\ &= 0.54\ \text{m}^2\end{aligned}$$

   b. Determine the net channel cross-sectional area by subtracting the cross sectional area of the quartz sleeves (4.15 × 10^{-4} m^2/lamp).

$$\begin{aligned}A_{\text{channel}} &= 0.54\ \text{m}^2 - [(12 \times 8)\ \text{lamps/bank}] \times (4.15 \times 10^{-4}\ \text{m}^2\text{/lamp})\\ &= 0.50\ \text{m}^2\end{aligned}$$

   c. Determine the maximum velocity in the channel.

$$v_{\text{channel}} = \frac{(24{,}000\ \text{L/min·channel})(0.001\ \text{m}^3\text{/L})(1\ \text{min/60s})}{0.5\ \text{m}^3} = 0.8\ \text{m/s}$$

   d. Determine the headloss per UV channel.

$$h_{\text{channel}} = 0.75\frac{v^2}{2g}$$

$$h_{\text{channel}} = \frac{(0.75)(0.80\text{ m/s})^2(1000\text{ mm/m})}{2(9.81\text{ m/s}^2)}(2\text{ banks}) = 49.0\text{ mm}$$

Note that 2 banks were used to determine system headloss. Use of two banks includes one redundant bank of lamps in each channel. The clear spacing between quartz sleeves is 52 mm (75 mm − 23 mm) and the headloss should not exceed this value.

8. Summarize the system configuration.
   System utilizes three channels, each channel containing two banks of lamps in series, one operational bank and one redundant bank. Each bank contains 12 modules, each of which contains 8 lamps.

Comment The majority of UV disinfection systems have the ability to turn banks of lamps on and off and vary power (and therefore UV output) to the banks that remain on. Turning UV lamps on and off is done automatically in response to varying flowrates and water quality (UVT). Varying the output of a UV system based on flow is accomplished by connecting a plant flowrate signal to the UV system's programmable logic controller (PLC). UVT can be manually entered based on readings taken from a bench-top photometer or based on continous readings from an online transmittance monitor.

Use of Spot-Check Bioassay to Validate UV System Performance

A spot-check-bioassay (SCB) test procedure has been developed to validate the performance of a newly installed and operational UV disinfection system. The test involves making a minimum of eight spot-check viral assays to demonstrate that the full-scale UV reactor performance complies with the design intent. Because new lamps are installed, the 75 percent PI is used as a reference. The CHPH has approved the use of the SCB test procedure to assess compliance of a full-scale disinfection reactor or system with the design intent. As implemented by CDPH, seven of the eight bioassay tests results must lie above the lower 75 percent PI predicted values. The rationale is that the percent ratio of seven out of eight is 87.5, which corresponds to the lower prediction interval. If more than one out of eight SCB test bioassays is below the PI curve, it is usually a clear indication that something may be wrong with the installation (e.g., poor inlet and outlet flow distribution, poor channel geometry, poor alignment, inappropriate weir placement, inappropriate flow control devices, inappropriate power settings, as well as other site conditions). If the installation site features can be corrected, they should be corrected, and the system should be retested. If the installation site features cannot be corrected, the UV system should be derated. The SCB test procedure along with the procedure for derating the UV system is illustrated in Example 12–16.

The SCB test procedure is similar to the procedure followed for UV reactor validation, as delineated in Example 12–14, Part A, with the exception that a wide range of operating conditions is evaluated. For example, consider a system comprised of two channels, each containing 4 banks of UV lamps. For such a system a typical test program might include four tests conducted under the following conditions:

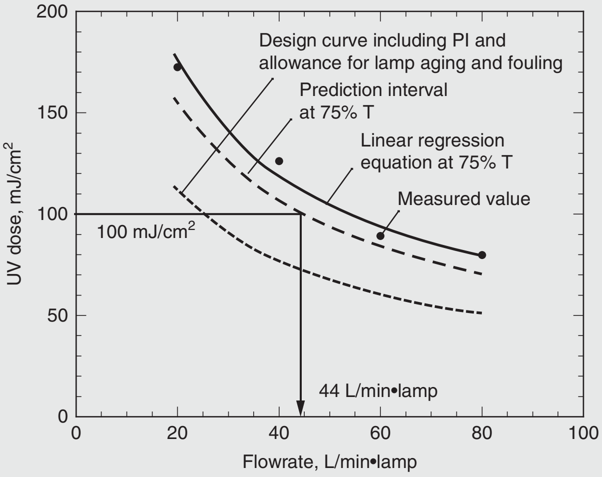
1. Maximum flowrate per lamp, minimum transmittance
2. Average flowrate per lamp, minimum transmittance
3. Maximum power setting, minimum transmittance
4. Minimum flowrate per lamp, minimum transmittance, first operational UV bank in sequence (i.e., 1, 2, 3, and 4)

Four additional tests could be conduced under the following conditions:

5. Ambient transmittance, maximum flowrate per lamp
6. Ambient transmittance, intermediate ballast output settings (60, 70, 80, or 90 percent)
7. Ambient transmittance, intermediate flowrates
8. Ambient transmittance, with the last operational UV bank in sequence (i.e., 1, 2, 3, and 4)

The goal of test 1 is to check performance under worst case conditions. The goal of test 2 is to check performance under typical flowrates and worst case water quality conditions. The goal of test 3 is to check performance under worst case water quality conditions at maximum power. The goal of tests 4 and 8 is to determine whether bank placement has an impact on operational performance. The goal of test 5 is to check performance under maximum flowrate per lamp and typical water quality conditions. The goal of test 6 is to check performance under different power settings. The goal of test 7 is to evaluate the performance of the UV system at various intermediate operating conditions. It should be noted that any number of test sequences can be used, as long as a wide range of operating conditions is evaluated.

EXAMPLE 12–16 **Conduct of Spot-check Bioassay to Validate Performance of Full-scale UV Disinfection System** A spot-check bioassay is to be conducted to validate the performance of a newly installed and operational UV disinfection system at a wastewater treatment plant. The UV system, validated in Example 12–14, is comprised of two channels, each containing four banks of lamps. Each bank contains 4 UV lamps oriented parallel to flow. The UV system was validated over a range of flowrates from 20 to 80 L/min·lamp at a transmittance of 75 percent. Based on the 75 percent PI, the UV system with new lamps will deliver a UV dose of 100 mJ/cm^2 up to a flowrate of 44 L/min·lamp, as shown in the following plot:



The regression equation is

$$\text{UV dose, mJ/cm}^2 = (10^{2.997})[(\text{flowrate})^{-0.577}]$$

The corresponding equation for the 75 percent PI is

$$\text{UV dose, mJ/cm}^2 = (10^{2.997})[(\text{flowrate})^{-0.577}](0.88)$$

Solution

1. Determine the maximum required flowrate for one channel.

   Max. flowrate = (4 lamp/bank)(44 L/min·lamp·bank) = 176 L/min

2. Determine the minimum required flowrate for one channel.

   Min. flowrate = (4 lamp/bank)(20 L/min·lamp·bank) = 80 L/min

3. Develop the test conditions.
   A minimum of eight spot-check viral assays must be conducted to demonstrate that full-scale UV reactor performance complies with the design intent.
   a. Bioassay test conditions.

Test No.	UVT, %	Operational banks	Flowrate, L/min	Hydraulic loading rate, L/min-lamp	Power setting, %
1	75	1, 2, 3	176	44	100
2	75	2, 3, 4	176	44	100
3	75	1, 3, 4	176	44	100
4	75	1, 2, 4	176	44	100
5	75	1, 2, 3	140	35	100
6	75	2, 3, 4	120	30	100
7	75	1, 3, 4	100	25	100
8	75	1, 2, 4	80	20	100

   b. Conduct spot-check bioassays.
      i. The first step is to conduct a quality assurance test to demonstrate that the laboratory procedures for the analysis of MS2 are valid (see Example 12–8).
      ii. The second step is to conduct the field spot-check bioassay tests. The log inactivation achieved from the field test and the assigned UV dose are as follows:

Test[a]	Log_{10} inactivation	UV Dose[b], mJ/cm^2
1	5.002	120.2
2	4.803	115.1
3	4.617	110.3
4	4.438	105.7
5	4.605	110.0
6	5.609	135.8
7	6.406	156.3
8	6.760	165.4

[a] See above table for operating conditions.

[b] The UV dose is based on the following equation:

$$\text{UV dose} = \frac{\text{log inactivation} - 0.326}{0.0389}$$

4. Compare the SCB test results to the values obtained from the linear regression equation and the PI equation.
   a. The two comparisons are presented in the following table:

Test	UV dose, mJ/cm^2 Predicted from regression equation	PI predicted from regression equation	Measured from spot check	Ratio spot-check/ predicted from regression equation	Ratio spot-check/ predicted PI from regression equation
1	111.9	98.4	120.2	1.07	1.22
2	111.9	98.4	115.1	1.03	1.17
3	111.9	98.4	110.3	0.99	1.12
4	111.9	98.4	105.7	0.94	1.07
5	127.7	112.3	110.0	0.86	0.98
6	139.5	122.8	135.8	0.97	1.11
7	155.0	136.4	156.3	1.01	1.15
8	176.3	155.2	165.4	0.94	1.07

   b. In comparing the SCB data to the regression equation it can be seen that the distribution of values is as would be expected, with a more or less equal distribution of values above and below the value obtained from the regression analysis.
   c. Based on the performance ratio, seven of the eight test results are above the predicted value of the PI, thus, the operation of the full-scale UV disinfection system is consistent with the design intent, as required by CDPH.
5. System adjustments for poor performance. In a situation in which more than one of the eight SCB test values lies below the PI curve, the following steps should be taken:
   a. Review the features of the installation, as discussed above, that may be leading to poor performance, correct any of the problems commonly encountered, and conduct a new SCB test.
   b. If the new test results are the same as the previous test, the UV system target UV dose set point must be adjusted or the system must be derated.
      i. Where the regulatory agency prefers not to modify the system dose equation, a site-specific target dose can be developed. The site-specific target dose can be computed using the following expression:

$$\text{Target UV dose} = \frac{\text{Design equation UV dose}}{\text{7th lowest 75\% PI spot check ratio}}$$

      The 7th lowest spot check ratio, based on the PI, is obtained as shown in column 5 in the above table. In this example the value is 0.94.
      ii. Alternatively, the system target UV dose can be derated using the following expression:

$$\text{UV dose}_{\text{Adj}} = \text{Design equation} \times \text{7th lowest 75\% PI spot check ratio}$$

      As above, 7th lowest spot check ratio, based on the PI, is 0.94.

Comment To avoid having to derate a UV system or change the target UV dose, it is imperative that careful attention be devoted to the design, installation, and operation of the UV disinfection system and its appurtenant facilities, especially the outlet control structure.

Troubleshooting UV Disinfection Systems

Problems associated with UV disinfection systems are related primarily to the inability to achieve permit limits. Some issues that must be considered when diagnosing problems associated with UV disinfection systems are discussed below.

UV Disinfection System Hydraulics. Perhaps one of the most serious problems encountered in the field is erratic or reduced inactivation performance due to poor system hydraulics. The most common hydraulic problems are related to (1) the creation of density currents that can cause the incoming water to move along the bottom or top of the UV lamp banks resulting in short circuiting, (2) inappropriate entry and exit conditions that can lead to the formation of eddy currents which ultimately create uneven velocity profiles that induce short circuiting, (3) the creation of dead spaces or zones within the reactor resulting in short circuiting, and (4) uneven flow distribution in systems with multiple channels leading to overloading in certain channels and underloading in others. The occurrence of short circuiting, channel overloading, and dead zones reduces the average contact time and leads to a decrease in UV dose and therefore compromises disinfection. When designing a UV system, the use of CFD modeling may be warranted to ensure hydraulic issues that could negatively impact disinfection performance are accounted for.

The principal hydraulic design features that can be used to improve system hydraulics in open channels include the use of (1) submerged perforated diffusers at the inlet of UV channel(s), (2) corner fillets in rectangular open channel systems with horizontal lamp placement, and (3) flow deflectors in open channel systems with vertical lamp placement. In rare cases, power input to mix the incoming flow may be necessary. Some of these corrective measures for open channel UV disinfection systems are illustrated on Fig. 12–46. Submerged perforated baffles should have an open area of about 4 to 6 percent of the cross-sectional area of the flow channel. Similar to open-channel systems, closed vessel UV systems may require similar design features to improve hydraulics through the reactors. Again, the use of CFD modeling may be of great value in studying the effect of various physical interventions in bringing about a more uniform approach velocity flow field (Sotirakos et al., 2013).

Biofilms on Walls of UV Channels and on UV Equipment. Another serious problem encountered with UV disinfection systems is the development of biofilms on the exposed surfaces of the UV reactor. The problem is especially serious in open channel systems covered with standard grating. It has been found that if the UV channels are exposed to any light, even very dim light, biofilms (typically fungal and filamentous bacteria) will develop on the exposed surfaces. The problem with biofilms is that they can harbor and effectively shield bacteria. When the clumped biofilms break away from the attachment surface, bacteria can be shielded as the clumps pass through the disinfection system. The best control measure is to completely cover the UV channels. Further, all concrete channels should be lined or coated to avoid the formation of bacterial colonies in the crevices and rough spots found in poured concrete. In addition, the channels can be cleaned and disinfected occasionally using hypochlorite, peracetic acid (see Sec. 12–8), or another suitable cleaning agent/disinfectant.

It should be noted that biofilm development can also occur in closed UV systems, but the severity is usually less, with the exception of UV systems in which medium-pressure high-intensity UV lamps are employed. Because medium-pressure high-intensity UV lamps emit some light in or near the visible light range (see Fig. 12–33) they can stimulate the growth of microorganisms on exposed surfaces. In some cases, growths approaching 300 mm in length have been found attached to the lamp support structure. The amount of

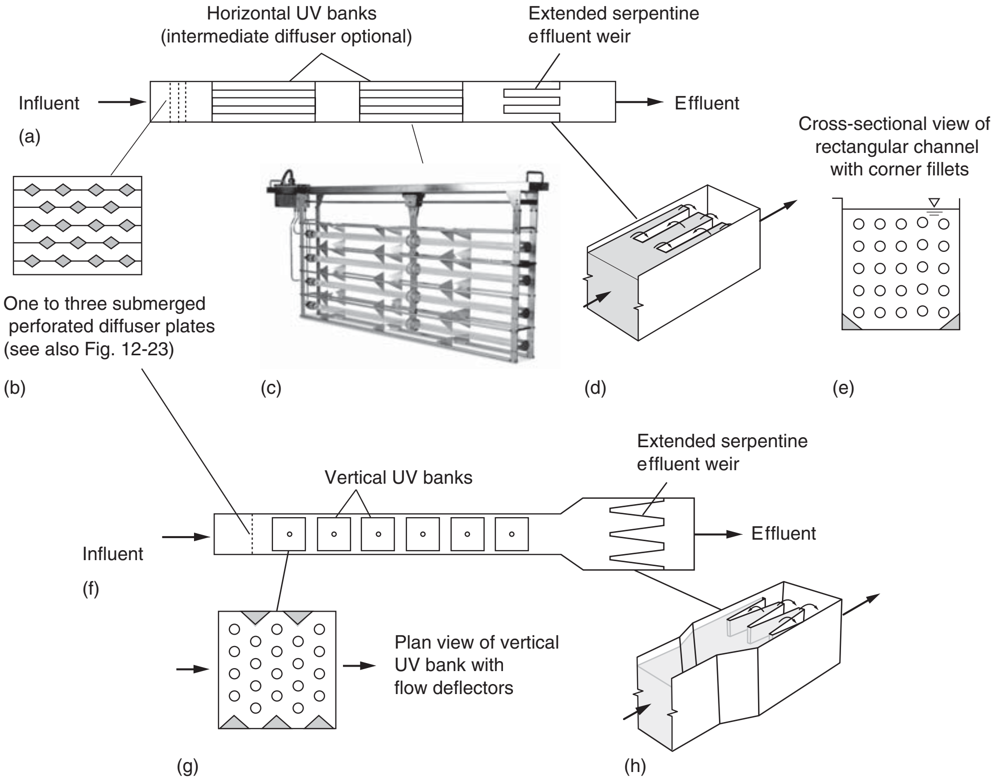


Figure 12–46

Typical examples of physical features that can be used to improve the performance of horizontal and vertical open UV reactors: (a) plan view of horizontal lamp UV system in channel with lining or coating, (b) perforated submerged diffuser plates to promote plug flow, (c) use of delta wings for enhanced internal mixing (courtesy of Calgon Carbon), (d) and (h) extended serpentine effluent weirs to promote plug flow, (e) elimination of dead space with corner fillets, (f) plan view of vertical lamp UV system in channel with lining or coating, and (g) use of baffle diffusers for enhanced internal mixing.

light emitted in the visible light range will vary with each type of lamp (i.e., manufacturer). Removal of these growths with a suitable disinfectant must be conducted on a periodic basis.

Overcoming the Impact of Particles by Increasing UV Intensity. It was thought at one time that the impact of particles on the performance of UV disinfection systems could be overcome by increasing the UV intensity. Unfortunately, it has been found that increasing the UV intensity tenfold has little effect on reducing the number of surviving particle-associated coliform bacteria because the absorption of UV irradiation by particles in wastewater is typically 10,000 times or more greater than the bulk liquid medium. Particles essentially block the transmission of UV light. Particles larger than a certain critical size (a function of the size of the target organism) will effectively shield the embedded microorganisms (Emerick et al., 1999; Emerick et al., 2000). Because the effectiveness of UV disinfection is governed primarily by the number of particles containing coliform bacteria, to improve the performance of a UV disinfection system, either the number of particles with associated coliform bacteria must be reduced (e.g., by modifying the treatment process mode of operation or by adding polymer to improve the performance of the secondary sedimentation facilities), or the particles themselves must be removed (e.g., by some form of filtration). Currently (ca. 2013), to meet the stringent total coliform bacteria requirements for body contact water reuse applications (i.e., equal to or less than 2.2 MPN/100 mL), some form of effluent filtration is required.

Effect of Treatment Processes on UV Performance. The number of particles with associated coliform bacteria, as well as their size, is another factor that will impact the performance of a UV disinfection system. As noted previously in Sec. 12–2, it has been observed that for activated sludge plants the number of particles with associated coliform organisms is a function of the SRT (see Fig. 12–5). Thus, both the mode of operation of the biological process and the design and operation of the secondary sedimentation facilities must be evaluated carefully, especially where an unfiltered effluent is to be UV irradiated. Even with effluent filtration, attention must be focused on the distribution of particle sizes in the filtered effluent (Darby et al., 1999; Emerick et al., 1999). Although the number of particle associated bacteria is reduced with long SRT values, it should be noted that the use of long SRT values leads to the formation of a larger number of particles for a given turbidity level. Dispersed particles are also difficult to filter and may pass through some filtration systems without violating turbidity limits, which, in turn, reduces the effectiveness of UV disinfection system.

Environmental Impacts of UV Radiation Disinfection

The environmental impacts associated with the use of UV disinfection for wastewater stem primarily from the fact that UV systems utilize much more electricity than other disinfection methods. To understand the carbon footprint of the system, the source of electricity for a UV system should be investigated. Because ultraviolet light is not a chemical agent, no toxic residuals are produced. However, certain chemical compounds may be altered by the ultraviolet irradiation. On the basis of the evidence to date, it appears that the compounds formed are harmless or are broken down into more innocuous forms at the dosages used for the disinfection of wastewater and reclaimed water (20 to 100 mJ/cm^2). Photooxidation, which does alter the structure of compounds, occurs above about 400 kJ/cm^2 range. Thus, the disinfection of wastewater with ultraviolet light is not considered to have any adverse environmental impacts. The impacts associated with some of the new very high-energy lamps, which may operate in the kilojoule range, is not known at present (ca. 2013).

12–10 DISINFECTION BY PASTEURIZATION

The process of heating food or water at a specified temperature and time for the purpose of killing microorganisms is known as *pasteurization*. The process was first demonstrated on April 20, 1862 in France by Pasteur and Bernard in response to a request by Emperor Napoleon III to save France's wine industry from what were called "diseases of wine" (Lewis and Heppell, 2000). The major contribution made by Pasteur was to define the exact time and temperature required to kill specific microorganisms, without affecting the taste of the wine. From the early beginnings, pasteurization is now used universally in the food industry to control pathogenic microorganisms. It is important to note that pasteurization is not intended to kill all microorganisms as compared to sterilization. Rather, it is intended to reduce the viable number of microorganisms present. Description of the pasteurization process, reported performance data, and regulatory requirements are discussed below.

Description of the Pasteurization Process

The operation of a pasteurization process is illustrated schematically on Fig. 12–47. As shown, effluent to be disinfected is introduced into the preheat reactor where heat from the

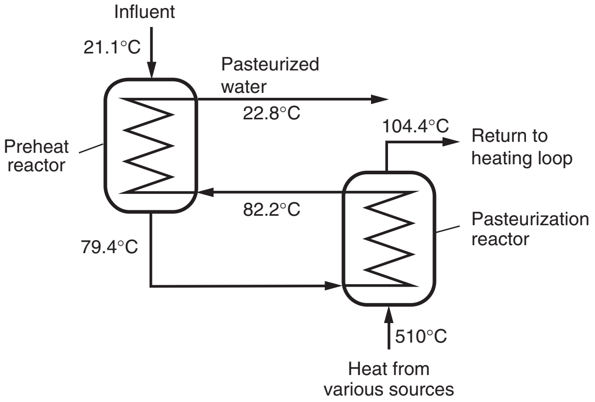


Figure 12–47

Definition sketch for the pasteurization process for the wastewater. (Adapted from Salveson et al., 2011.)

disinfected effluent is used to preheat the incoming effluent. The preheated effluent is then directed to the pasteurization reactor where heat from an external source is used to heat the preheated effluent to the desired temperature and where it is retained for a prescribed period of time. The external heat source can be from turbine exhaust, engine exhaust, waste gas burner exhaust, hot water, or from another suitable fluid. The reported temperature values shown on Fig. 12–47 are site specific and will depend on local conditions and the design of the heat exchange equipment.

Operationally, three different types of pasteurization are in use: (1) batch, (2) HTST (high-temperature short time) and (3) UHT (ultra-high temperature). Because of the large volumes required, batch pasteurization is only suitable for very small operations. Typical operational ranges for the three types of pasteurization are summarized in Table 12–34. The continuous flow HTST pasteurization process is used in most industrial operations and is the form used for the disinfection of treated wastewater. General operational data for the HTST process to achieve four-log inactivation of specific organism groups is presented in Table 12–35. The UHT pasteurization process, also known as flash pasteurization, is only used in more specialized applications.

Thermal Disinfection Kinetics

The disinfection performance of the pasteurization process depends on both temperature and holding time. As noted in Table 12–2, and as compared to other disinfection methods,

Table 12–34

General operating ranges for pasteurization technologies[a]

Pasteurization technology	Temperature		Time	Comments
	°C	°F		
Batch	62–64	144–147	30–35 min	Inactivates most vegetative bacterial calls including streptococci, staphylococci, and mycobacterium tuberculosis
High-temperature short time (HTST)	72–75	161–165	8–30s	Same effect as batch, but at much shorter times
Ultra-high temperature (UHT)	135–140	275–285	<1–5s	Lethal for most bacterial cells at even shorter times than HTST

[a] Adapted in part from Toder (2012), Hudson et al. (2003), Sorqvist (2003).

Table 12–35

General operating ranges for HTST pasteurization to achieve approximately 4-log inactivation of selected microorganisms

Microorganism	Temperature °C	Temperature °F	Time, s	Comments
Bacteria	72–77	161–170	6–16	
Protozoa	70–72	158–162	8–16	Essentially complete inactivation
Virus	80–85	176–185	10–30	
MS2 Coliphage	79–81	175–178	15–40	
Helminths	70–72	158–162	8–10	Essentially complete destruction

Various sources.

the high temperature is needed to denature the enzymes in the cell protoplasm and to alter the structure of the cell wall. The holding time is needed to complete the reaction within the cell and with the cell wall constituents. In a manner similar to chemical disinfection it has been observed that if the temperature is increased the time required to inactivate a given microorganism will decrease (Pflug et al., 2001).

First Order Kinetics. The microbial disinfection of microorganisms, at a specific temperature, can be modeled as a first order reaction as follows:

$$\frac{dN}{dt} = -kN \tag{12–70}$$

where N = the number of organisms surviving after time t
k = reaction rate constant, base e
t = the exposure time

The integrated form of Eq. (12–70) in base 10 is

$$N_t = N_o 10^{-Kt} \tag{12–71}$$

where N_t = the number of organisms remaining at time t
N_o = the initial number of organisms
K = reaction rate constant, base 10, Note K =0.4343k
t = time, s, min, h

The reaction rate constant, K, is given by

$$K = (\log N_o - \log N)/t \tag{12–72}$$

Heat Resistance Parameters. Two parameters, D and Z, are used commonly to describe the effectiveness of the pasteurization process (Goff, 2012; Pflug et al., 2001). The term D is a measure of the heat resistance of a given microorganism and corresponds to the time required to achieve 1-log of inactivation (i.e., 90 percent) at a given temperature T, as illustrated on Fig.12–48(a). Also known as the *decimal reduction time*, D, is given by the following expression:

$$D = 1/K = t/(\log N_o - \log N_t) \tag{12–73}$$

Using Eq. (12–73), Eq. (12–72) can be written as

$$N_t = N_o 10^{-t/D} \tag{12–74}$$

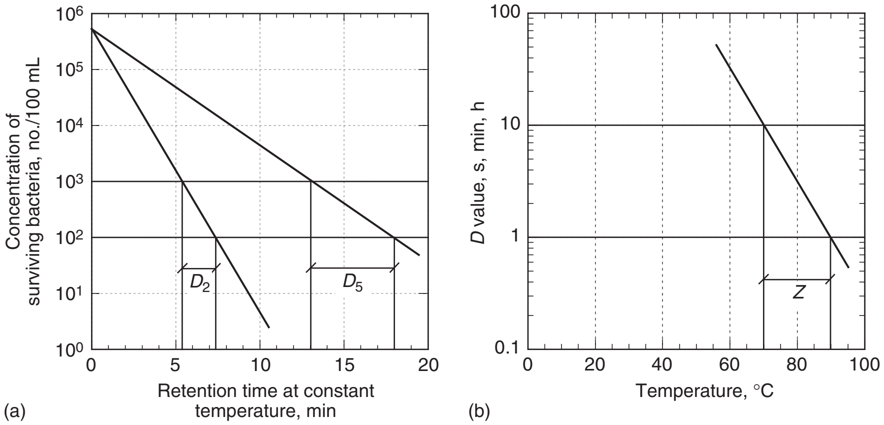


Figure 12–48

Definition sketches for the pasteurization process: (a) plot to determine the time, *D*, at constant temperature required to reduce the concentration of microorganisms by 1-log and (b) plot to determine, *Z*, the temperature increase required to reduce the *D* value by 1-log.

Each microorganism will have a specific value of D. It should be noted that the K and D values are only for a specific temperature. For example, at temperature T, an organism having a D value of 6 is less resistant than an organism with a D_T value of 10 [see Fig. 12–48(a)].

The term, Z, reflects the temperature dependence of D as illustrated on Fig. 12–48(b). The Z value corresponds to the temperature required to achieve 1-log change in the D value. The Z value is given by the following expression:

$$Z = (T_2 - T_1)/[\log(D_1) - \log(D_2)] \text{ or} \tag{12–75}$$

$$\log\left(\frac{D_1}{D_2}\right) = \frac{(T_2 - T_1)}{Z} \tag{12–76}$$

When D_1 and D_2 differ by one log, Eq. (12–76) reduces to

$$Z = (T_2 - T_1) \tag{12–77}$$

Using these two parameters, the heat resistance of a single microorganism can be quantified. Typical D and Z values for selected microorganisms are reported in Table 12–36. The values given in Table 12–36 are derived from a number of sources. The most extensive evaluation of D and Z values has been by Sorqvist (2003). Typical Z values for bacteria are in the range from 5 to 10. The application of the terms D and Z is illustrated in Example 12–17.

Non-Linear Inactivation. Although the equations for D and Z assume a more or less log-linear response, a number of researchers have observed both shoulder and tailing effects as described previously (see also Fig. 12–4). In general, the shoulder effect reflects the condition where the microorganism is less effected by temperature initially as compared to the subsequent linear inactivation phase. The tailing effect corresponds a period of reduced thermal destruction in which the response is less rapid than that that observed in the linear inactivation phase. In both cases, the effects are not understood completely (Hiatt, 1964). In addition, a number of other microorganisms such as spore forming bacteria *(B. anthracis and B. cereus)* exhibit a nonlinear response. Inactivation of spore-forming bacteria requires extremely high temperatures, especially if short contact times are to be used.

Table 12–36

Typical D and Z values for selected microorganisms[a]

	Temperature		*D*, s		*Z*, °C	
Microorganism	**°C**	**°F**	**Range**	**Typical**	**Range**	**Comp.[b]**
Campylobacter jejuni/coli	60	140	6.5–10	8.2	2.8–8.0	5.5
	70	158		0.12		
E. coli	60	140	35–42	40	3.2–9.2	5.0
	70	158		0.4		
E. coli 0157:H7	60	140	23–26	24	4.3–9.8	4.8
	70	158		0.2		
Enterococcus fecalis	60	140	360–480	415	2.2–14.2	6.0
	70	158		9.0		
Listeria monocytogenes	60	140	81–93	87	4.3–11.5	6.1
	70	158		2.0		
MS2 coliphage	70			14		10
	80			1.4		
Salmonella spp.	60	140	23–26	24	3.3–9.5	5.6
	70	158		0.4		
Staphylococcus aureus	60	140		54		10.5
	70	158		6		
Streptococcus	60	140		24		7.7
	70	158		1.2		
Total, coliform	60	140	42–60	50		7.9
	70	160		2.7		
Yersinia enterocolitica	60	140	24–37	30	4.0–13.7	6.6
	70	158		0.9		

[a] Adapted from Hudson et al. (2003), Sorqvist (2003), Salveson (2012).

[b] Computed *Z* values based on the typical *D* values given in column 5.

EXAMPLE 12–17 **Estimate Pasteurization Operating Conditions** Some new pasteurization equipment is to be installed to replace some existing equipment for the disinfection of an unknown strain of bacteria. The current equipment operates at a temperature of 65°C (150°F). The corresponding *D* and *Z* values for the bacteria are 10 s and 12°C, respectively. The initial bacterial count is 10^6 org/100 mL. If the new equipment is to operate at 77°C (170°F) for 4 s, what level of inactivation can be achieved? If the new equipment cannot be installed, what time would be required at a temperature of 65°C (150°F) to achieve 4 log removal?

Solution

1. Determine D_T at 77°C using Eq. (12–76).

$$\log\left(\frac{10.0\text{ s}}{D_2}\right) = \frac{(77 - 65)}{12} = 1.0$$

$$\left(\frac{10.0\text{ s}}{D_2}\right) = \text{inverse}\log(1.0) = 10.0$$

$D_2 = 10.0\text{ s}/10.0 = 1.0\text{ s}$

2. Determine the degree of inactivation using Eq. (12–73).

$$\log\left(\frac{N}{N_o}\right) = -4\text{s}/1.0\text{ s} = -4$$

Four-log inactivation can be achieved

3. Determine the time required to achieve a 4-log inactivation using the existing equipment. The required time can be estimated using Eq. (12–73).

$$\log\left(\frac{N}{N_o}\right) = -t/D_T$$

$$\log\left(\frac{N}{N_o}\right) = -4 = -t/10$$

$t = 4 \times 10 = 40\text{ s}$

Germicidal Effectiveness of Pasteurization

The literature on the thermal destruction of microorganisms in the food processing industry is vast. However, the literature on the disinfection of treated effluent is relatively limited. In a study completed recently, the use of UV and pasteurization were evaluated at pilot scale as possible replacements for the existing chlorine disinfection system. The secondary effluent was filtered. In the studies on pasteurization, conducted using proprietary equipment, contact times and temperatures were varied depending on the test organism (Salveson et al., 2011). Based on the test results for MS2 coliphage, it was found that within the range of times tested, contact time did not appear to be significant up to a temperature of about 73°C (163°F) (i.e., non-linear response). The temperature and time required to achieve a 4-log reduction of MS2 was on the order 80°C (176°F) and 7.7 s, respectively. These values are consistent with the values approved by the CDPH for pasteurization.

Regulatory Requirements

The CDPH has approved the use of the pasteurization process for the disinfection of treated effluent for Title 22 reuse applications. To achieve a 4-log virus reduction (based on MS2), the CDPH has set the temperature at 82°C (180°F) at a contact time of 10 s. For nonreuse applications where virus reduction is not needed, it may be possible to reduce the temperature to 74°C (165°F) and the time to 8 s. It is anticipated that these values will be revised as additional operational data are collected.

Application of Pasteurization for Disinfection

Pasteurization has been studied at a number of locations and has been found to have the lowest cost when compared to other disinfection technologies, and especially where waste heat is available (Salveson et al., 2011). A typical heat balance for the pasteurization process was shown previously on Fig. 12–47. The recovery of the heat from wastewater is another possibility that can be considered in the application of the pasteurization process. The survival of bacterial indicator species and bacteriophages in sludge and wastewater after thermal treatment has been studied (Moce-Uivina et al., 2003). The pasteurization of biosolids is considered further in Chap. 14.

PROBLEMS AND DISCUSSION TOPICS

12–1 Assuming Chick's law applies, determine the inactivation rate constant for total coliform for one of the following four treated effluents (sample to be selected by instructor). The effluent temperature was 20°C. If the activation energy for the disinfection reaction is 52 kJ/mole, determine the inactivation rate constant at 12°C.

	Time, min			
	Effluent sample			
Log of organisms remaining	1	2	3	4
8	0.0	0.0	0.0	0.0
7	1.8	5.5	3.8	2.6
6	3.6	11.5	8.0	5.5
5	5.6	17.5	12.3	8.0
4	7.4	23.5	16.5	11.0
3	9.2	20.9	20.9	13.9

12–2 Using the rate constant developed in Problem 12–1, determine the chlorine dose required to achieve a 99.99 percent inactivation of total coliform in 60 min at 15 and 25°C.

12–3 The following combined chlorine disinfection data were obtained in a series of laboratory tests performed on three different filtered activated sludge effluents:

	Residual fecal coliform count, no./100 mL		
	Test		
Combined chlorine CT, mg·min/L	1	2	3
0	10^6	10^6	10^6
50	10,000	199,500	316,000
100	10,200	31,600	63,000
200	126	800	4000
300	1	25	280
400		1	20
			1

a. Using these data, determine the value of the coefficient of specific lethality in Eq. (12–6) and the CT value to achieve a residual coliform count of 200/100 mL and 1000/100 mL.

b. Using the following data, determine the required volume in m^3 of a chlorine contact chamber designed to provide 60-min contact at the average winter flowrate. Using the equations developed in Part *a*, determine the minimum dosage required in mg/L to give the required kill for one of the test results given above (test condition to be selected by instructor). Assuming that the yearly chlorine requirement can be computed on the basis of the average flowrate for each of the two 6-mo periods, determine the minimum yearly chlorine requirement in kilograms.

Item	Unit	May-Oct	Nov-Apr
Average flowrate	m^3/d	20,000	26,000
Peak daily flowrate	m^3/d	40,000	52,000
Maximum permissible fecal coliform count in effluent	MPN/100 mL	200	1000

12–4 The following data were obtained for several filtered wastewater effluents. Using these data estimate the coefficients for the refined Collins-Selleck model [Eq. (12–27)] for wastewater number (to be selected by instructor).

	Time min			
	Wastewater number			
$-\log (N/N_o)$	1	2	3	4
1	2.1	6.9	2.9	3.5
2	4	15	5.9	8.1
3	7.1	36	12.3	18
4	13.6	80	24	40
5	21.5	190	55.5	90
6	42.3	430	115	200

Using the derived values, estimate the inactivation that could be achieved with a CT value of 30, 60, or 120 mg·min/L (value to be selected by instructor).

12–5 A consultant has proposed using chorine dosages of 15 and 8, 20 and 10, 30 and 20 mg/L during the summer and winter, respectively for effluent disinfection. If the effluent total coliform count before disinfection is 10^7/100 mL, estimate the final total coliform counts that can be achieved during the summer and winter with one of the dosage sets (to be selected by instructor).

1. Demand due to decay during chlorine contact = 2.0 mg/L
2. Required chlorine contact time = 45 min
3. Use the typical values given below for the coefficients.

   $b = 4.0$ and $n = 2.8$

12–6 The chlorine residuals measured when various dosages of chlorine were added to four different wastewater effluents are given below. For one of the effluents (to be selected by instructor), determine: (a) the breakpoint dosage and (b) the design dosage to obtain a free chlorine residual of 1, 2, or 3.5, mg/L (value to be selected by instructor).

	Residual, mg/L			
	Effluent number			
Dosage, mg/L	1	2	3	4
0	0	0	0	0
1	0	1	0	0
2	1	2	1	1
3	0.2	3	2	2
4	1	4	2.3	2.9
5	2	4.3	1.2	3.4
6	3	3.6	0.9	2.7
7		2.3	1.7	1.2

(*continued*)

(*Continued*)

	Residual, mg/L			
	Effluent number			
Dosage, mg/L	1	2	3	4
8		0.7	2.7	1.2
9		0.7	3.7	2.1
10		1.7		3.1
11		2.8		4.1

12–7 Estimate the daily required chlorine dosage, the required alkalinity, if alkalinity will have to be added, and the resulting buildup of total dissolved solids when breakpoint chlorination is used for the seasonal control of nitrogen. Assume that the following data apply to this problem:

1. Plant flowrate = 4800 m^3/d
2. Effluent characteristics
   a. BOD = 15 mg/L
   b. Total suspended solids = 15 mg/L
   c. NH_3-N = 1, 1.25, or 1.5 mg/L (value to be selected by instructor)
   d. Alkalinity = 125, 145, or 165 mg/L as $CaCO_3$ (value to be selected by instructor)

12–8 Review the current literature and prepare an assessment of the use of chlorine gas versus sodium hypochlorite for the disinfection of treated wastewater. A minimum of 3 recent (after 2000) articles and/or reports should be cited in your assessment.

12–9 The following data were obtained from dye tracer studies of five different chlorine contact basins. Using these data, determine the mean hydraulic residence time and the corresponding variance, the t_{10} time, and the Morrill Dispersion Index and the volumetric efficiency for one of the basins (to be selected by instructor). How would the performance of the basin selected for analysis be classified according to the U.S. EPA guidelines?

	Tracer concentration, μg				
	Basin number				
Time, min	1	2	3	4	5
0	0.0	0.0	0.0	0.0	0.0
10	0.0	0.0	0.0	0.0	0.0
20	3.5	0.1	0.1	0.0	0.0
30	7.6	2.1	2.1	0.0	0.7
40	7.8	7.5	10.0	0.3	4.0
50	6.9	10.1	12.0	1.8	9.0
60	5.9	10.2	10.2	4.5	12.5
70	4.8	9.7	8.0	8.0	11.5
80	3.8	8.1	6.0	11.0	8.8
90	3.0	6.0	4.3	11.0	5.5
100	2.4	4.4	3.0	9.0	3.0
110	1.9	3.0	2.1	4.3	1.8
120	1.5	1.9	1.5	2.0	0.8
130	1.0	1.0	1.0	1.0	0.4
140	0.6	0.4	0.5	0.2	0.1

(*continued*)

(*Continued*)

	Tracer concentration, μg				
	Basin number				
Time, min	1	2	3	4	5
150	0.3	0.1	0.1	0.0	0.0
160	0.1	0.0	0.0	0.0	0.0
170	0.0	0.0	0.0	0.0	0.0

12–10 Using the following dose response data for an enteric virus and the tracer data for four different chlorine contact basins, determine for one of the basins (to be selected by instructor) the expected effluent microorganism concentration based on the t_{10} and mean hydraulic residence times. Also estimate the chlorine residual that would be required to achieve 4 logs of removal with the existing basins.

Dose response data for enteric viruses

CT, mg/L·min[a]	Number of organisms remaining
0	10^7
100	$10^{6.2}$
200	$10^{5.4}$
400	$10^{3.8}$
600	$10^{2.1}$
800	$10^{0.6}$
1000	10^{-1}

[a] Combined chlorine residual = 6.0.

Tracer data for chlorine contact basins

	Tracer concentration, mg/L			
	Chlorine contact basin			
Time, min	1	2	3	4
0	0.0	0.0	0.0	0.0
10	0.0	0.0	0.0	0.0
20	0.0	0.0	0.0	0.0
30	0.1	0.0	0.0	0.0
40	2.0	0.0	0.0	0.0
50	7.3	1.1	0.1	0.0
60	7.0	7.0	1.3	0.1
70	5.2	7.3	8.0	1.5
80	3.3	5.7	8.5	7.5
90	1.7	4.2	6.2	8.0
100	0.7	2.9	2.9	5.5
110	0.2	1.7	1.3	3.5
120	0.0	0.9	0.4	1.8
130		0.3	0.0	0.9
140		0.1		0.3
150		0.0		0.1
160				0.0
τ, min	80	85	90	100

12–11 Determine the amount of sulfur dioxide (SO_2), sodium sulfite (Na_2SO_3), sodium bisulfite ($NaHSO_3$), sodium metabisulfite ($Na_2S_2O_5$), and activated carbon (C) that would be required per year to dechlorinate treated effluent containing a combined chlorine residual of 5.0, 6.5, 7.0, or 7.7 mg/L as Cl_2 (residual to be selected by instructor) from a plant with an average flowrate of 1500, 3300, 4600, or 7500 m^3/d (flowrate to be selected by instructor).

12–12 Estimate the immediate ozone demand and the first order decay equation for wastewater number (to be selected by instructor) using the following bench-scale steady-state and decay test data collected at 25°C. If the coefficient of specific lethality for the inactivation of a newly discovered microorganism with ozone is 0.15 L/mg·min, estimate the degree of inactivation that could be achieved at 15°C using an ozone contactor with 4 compartments following the injection of ozone. The theoretical detention time in each compartment is 3 min. Assume the activation energy for ozone for the new microorganism is 48 kJ/mole.

		Ozone residual, mg/L			
		Wastewater number			
Test	Ozone dose, mg/L	1	2	3	4
1	6		2.4	1.0	3.3
2	10	1.1	4.9	5.9	7.0
3	14	6.9	7.4	10.5	10.3
4	18	12.2	10.0	15.5	14.0
5	20	15.0	11.1	18.0	15.7

The corresponding decay data are:

	Ozone residual, mg/L			
	Wastewater number			
Time, min	1	2	3	4
0	3.8	2.8	2.0	3.25
5	2.25	1.4	1.37	2.3
10	1.35	0.72	0.95	1.65
15	0.82	0.37	0.67	1.19
20	0.50	0.19	0.46	0.84

12–13 Estimate the immediate ozone demand and the first order decay equation using the following bench-scale steady-state test data collected at 20°C. If a four-compartment ozone contactor, similar to the one shown on Fig. 12–31(a), is used, estimate the log reduction in *Cryptosporidium* that can be achieved at 5°C. Assume the activation energy for ozone for *Cryptosporidium* is 54 kJ/mole.

The steady-state test results are as follows.

Test	Ozone dose, mg/L	Ozone residual, mg/L
1	5	1.5
2	8	5
3	10	7.5
4	13	10.3
5	18	17.5

The corresponding decay data are:

Time, min	Ozone residual, mg/L
0	5
4	3
7	2.5
10	2

12–14 Given the following ozone decay test data, estimate the number of compartments that would be required in an ozone contactor to achieve 3-log reduction in *Cryptosporidium* based on test number (to be selected by instructor). Assume the theoretical detention time in each compartment is 3 min and the t_{10}/t for the reactor is 0.65.

	Ozone residual, mg/L			
	Test number			
Time, min	1	2	3	4
0	3.3	1.5	3.2	2.8
2	3.0		2.75	
4		1.0		2.1
6	2.0		1.8	
10		0.65		1.8
12	1.5		0.9	
16		0.3		1.6

12–15 Review the current literature and prepare an assessment of the use of ozone for the disinfection of treated wastewater. A minimum of three articles and/or reports dating back to 1995 should be cited in your assessment.

12–16 Review the current literature and prepare an assessment of the use of peracetic acid alone or in combination with other disinfectants. A minimum of three articles and/or reports dating back to 2000 should be cited in your assessment.

12–17 Given the following measurements and data for a collimated beam test, determine the average UV dose delivered to the sample and best estimate of the uncertainty associated with the measurement.

$I_m = 10 \pm 0.5$ mW/cm^2 (accuracy of meter ±7%)

$t = 30 \pm 1$ s

$R = 0.025$ (assumed to be the correct value)

$P_f = 0.94 \pm 0.02$

$\alpha = 0.065 \pm 0.005$ cm^{-1}

$d = 1 \pm 0.05$ cm

$L = 48 \pm 0.5$ cm

12–18 If the intensity of the UV irradiation measured at the water surface in a Petri dish is 12 mW/cm^2, determine the average UV intensity to which a sample will be exposed if the depth of water in the Petri dish is 10, 22, 14, 15, or 16 mm (water depth to be selected by instructor).

12–19 If the intensity of UV irradiation measured at the water surface in a Petri dish in Problem 12–18 is 8 mW/cm^2, and that the computed UV dose was based on a water depth of 10 mm. What would be the effect if the actual water depth in the Petri dish were 20 mm?

12–20 Determine the mean, the standard deviation, the 75 percent confidence interval, and the 75 percent prediction interval for following MS2 bacteriophage inactivation data, (test to be selected by instructor) obtained using a collimated beam device. What UV dose would be required to achieve a 4-log inactivation of MS2 based on the lower prediction interval of 75 percent?

	Log reduction, $-\log N/N_o$				
	Test number				
Applied UV dose, mJ/cm^2	1	2	3	4	5
20	0.9	1.7	1.4	1.1	1
40	1.7	3.3	2.6	2.2	1.8
60	2.4	5.2	4.1	3	2.8
80	3.5	6.5	5.1	4.3	3.7
100	4.3			5.5	4.7
120	4.9			6.2	5.4

12–21 In the latest edition of the NWRI UV Guidelines (NWRI, 2012), the standard dose response curve that should be used to evaluate reactor performance is given as

$$\text{UV dose} = \frac{\text{log inactivation} - 0.5464}{0.0368}$$

The dose response curve used in this chapter is:

$$\text{UV dose} = \frac{\text{log inactivation} - 0.326}{0.0389}$$

Compare the two curves for UV dosages of 20, 40, 60, 80, 100, and 120 mJ/cm^2 (dosages to be selected by instructor). Is the difference significant? Why is it reasonable to specify a standard curve that all equipment manufacturers should use to evaluate the performance of their equipment?

12–22 A UV reactor comprised of two banks with 4 lamps per bank was tested on two different reclaimed waters (1 and 2) at four flowrates using MS2 bacteriophage as the test organism. The transmittance for both wastewaters was 65 percent. The hydraulic loading rates were varied from 50 to 200 L/min·lamp. In conducting the test, each flowrate was tested randomly with respect to order. The measured inlet and outlet phage concentrations are as follows:

Flowrate, L/min	Replicate	Wastewater 1, phage/mL		Wastewater 2, phage/mL	
		Inlet	Outlet	Inlet	Outlet
200	1	9.65×10^6	3.80×10^1	1.05×10^7	2.19×10^2
200	2	1.00×10^7	3.98×10^1	6.98×10^6	1.54×10^2
200	3	1.15×10^7	3.72×10^2	1.15×10^7	1.70×10^2
400	1	1.00×10^7	1.95×10^3	1.00×10^7	3.75×10^2
400	2	1.29×10^7	1.55×10^3	1.23×10^7	3.62×10^2
400	3	9.55×10^6	1.77×10^3	1.12×10^7	3.08×10^2
600	1	1.23×10^7	1.12×10^3	1.20×10^7	1.32×10^4
600	2	1.05×10^7	9.33×10^3	1.05×10^7	1.05×10^4
600	3	1.25×10^6	8.91×10^3	9.55×10^6	9.95×10^3
800	1	1.13×10^7	4.79×10^4	1.03×10^7	5.95×10^4
800	2	1.08×10^7	8.35×10^4	1.19×10^7	1.00×10^5
800	3	8.95×10^6	6.61×10^4	1.11×10^7	7.68×10^4

Using the given data, for water 1 or 2 (water to be selected by instructor), develop design equations based on (1) the regression analysis, (2) the 75 percent prediction interval, and

(3) the 75 percent prediction interval taking into account lamp aging and fouling for a transmittance value of 72 percent. What is the maximum flowrate per lamp over which the UV system will deliver a dose of 50 mJ/cm^2? Assume the MS2 UV dose response curve given in Example 12–12 will be used for the analysis of the inactivation test results.

12–23 The following MS2 bacteriophage inactivation data were obtained for filtered wastewater with a transmittance of 55 percent at 254 nm with a UV pilot test unit comprised of 6 UV lamps in a single UV bank operated at various ballast settings. The reported inactivation test results are the average of triplicate samples. Determine the maximum flowrate, expressed as L/min·lamp, over which the UV disinfection system will deliver a dose of 100 mJ/cm^2 at a ballast setting of 100 percent, 80 mJ/cm^2 at a ballast setting of 80 percent, or 50 mJ/cm^2 at a ballast setting of 50 percent (ballast setting to be selected by instructor) taking into account lamp fouling and aging. Assume the lamp aging and fouling factor is 60 percent based on the regression equation developed for each ballast setting.

Flowrate, L/min	Ballast output, %	Log$_{10}$ MS2 bacteriophage inactivation
180	100	7.7559
180	80	6.7445
180	50	5.4219
400	100	6.3555
400	80	5.383
400	50	5.383
560	100	5.5775
560	80	4.7606
560	50	3.5547
732	100	5.0718
732	80	4.2549
732	50	3.2046

12–24 Using the data from Problem 12–23, develop the design curve based on the regression analysis taking into account flowrate and ballast settings. Plot the original data, the regression curve, and the 75 percent prediction interval.

12–25 Review the current literature and prepare an assessment of the use of low-pressure low-intensity versus low-pressure high-intensity UV disinfection systems for the disinfection of filtered secondary effluent. A minimum of 3 articles and/or reports dating back to 2005 should be cited in your assessment.

12–26 Using the D and Z values given in Table 12–36, determine whether the CDPH pasteurization requirements of 82°C for 10 s is sufficient to achieve a 4-log reduction in MS2 coliphage.

12–27 The following data were obtained from a pilot-plant pasteurization test. Using these data determine the D and Z values. If the temperature were increased to 68°C, how long would it take to achieve a 4-log reduction?

	Observed log reduction at indicated time		
Temp., °C	3 s	7 s	10 s
60	0.25	0.4	0.5
65	0.55	1.32	1.8
70	1.80	4.35	6.00

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13

Processing and Treatment of Sludges

WORKING TERMINOLOGY

Term	Definition
Acid-gas digestion	A modified process of anaerobic digestion where the acid phase hydrolysis is separated from the gas producing phase for increased volatile solids reduction.
Aerobic digestion	Biological stabilization process operated in the presence of oxygen in which the biodegradable matter in primary and secondary sludge is oxidized to carbon dioxide and other end products.
Anaerobic digestion	Biological stabilization process operated in the absence of oxygen in which the biodegradable matter in primary and secondary sludge is converted to methane, carbon dioxide, and other end products.
Autothermal thermophilic aerobic digestion (ATAD)	An aerobic digestion process in which the microbes generate enough heat to maintain temperatures in the thermophilic range. When maintained for enough time to meet 40CFR 503 requirements, the process results in biosolids that are relatively pathogen free and meet Class A standards.
Biosolids	Sludge from wastewater treatment processes that has been stabilized to meet the criteria in the U.S. EPA's 40 CFR 503 regulations and, therefore, can be used beneficially.
Class A biosolids	Biosolids that contain less than 1000 most probable number (MPN)/g of fecal coliforms and less than 3 MPN/4g of Salmonella bacteria and meet one of six stabilization alternatives given in 40 CFR 503. The material also must meet the pollutant limits and vector attraction reduction requirements set forth in 40 CFR 503.
Class B biosolids	Biosolids that contain less than 2 million colony-forming units (CFU) or most probable number (MPN) of fecal coliforms per gram of dry biosolids. The material also must meet the pollutant limits and vector attraction reduction requirements set forth in 40 CFR 503.
Digestion	The process of biologically degrading organic matter in sludge, thereby reducing the concentrations of volatile solids and pathogens.
Disposition	Disposition is a term used to reflect disposal of biosolids or sludge for either beneficial or non-beneficial use due to the value of the material.
Dissolved air flotation	A clarification process in which small air bubbles become attached to flocculated material, float to the surface, and are removed by skimming. Heavier solids which settle are removed by mechanical scrapers.
Dual digestion	A two-stage digestion process wherein the first stage is aerobic thermophilic digestion and the second stage is mesophilic anaerobic digestion. High-purity oxygen has also been used for the first stage.
Grit	Sand, gravel, cinders, other heavy inorganic materials and also organic matter such as eggshells, bone chips, seeds, and coffee grounds.

Term	Definition
Humus	Sludge removed from trickling filters.
Mesophilic anaerobic digestion	Anaerobic digestion that occurs in a temperature range of 30 to 38°C (85 to 100°F).
Methanogenesis	The metabolic conversion of organic acids or hydrogen and carbon dioxide to methane.
Screenings	The material removed from a screening device.
Scum	Buoyant materials (e.g., grease, food waste, paper, and foam) often found floating on the surface of primary and secondary clarifiers and thickeners.
Sidestream	A portion of the wastewater flow that has been diverted from the main treatment process flow for specialized treatment.
Solids	A term often used as a replacement for sludges that have not been stabilized by physical, chemical or biological treatment. The term solids is not used as a substitute for sludge in this chapter. The mass of dry material in sludge is referred to as the solids content.
Sludge	Any material (i.e., sludge) produced during primary, secondary, or advanced wastewater treatment that has not undergone any process to reduce pathogens or vector attraction.
Stabilization	A treatment process designed to reduce the number of pathogens in sludge and to reduce the attraction of vectors as defined in the requirements of 40 CFR 503.
Thermal hydrolysis	A thermal conditioning process, utilizing high-pressure steam for pretreating dewatered sludge prior to anaerobic digestion, that hydrolyzes and reduces the viscosity of the sludge.
Thermophilic anaerobic digestion	Anaerobic digestion that occurs in a temperature range of 50 to 57°C (122 to 135°F).

The constituents removed and/or produced in wastewater treatment plants include screenings, grit, scum, sludge, and biosolids. The sludge and biosolids (formerly collectively called sludge) resulting from wastewater treatment processes are usually in the form of a liquid or semisolid liquid, which typically contains from 0.25 to 12 percent solids by weight, depending on the operations and processes used. In the United States, the term *biosolids*, as defined by the Water Environment Federation (WEF 2010a), refers to any sludge that has been stabilized to meet the criteria in the U.S. Environmental Protection Agency's 40 CFR 503 regulations and, therefore, can be used beneficially. The term *sludge* is only used before beneficial use criteria (discussed in Sec. 14–2) have been achieved. The term *sludge* is generally used in conjunction with a process descriptor, such as *primary sludge*, *enhanced primary sludge, waste activated sludge*, and *secondary sludge*. Although the terms solids and has been used as a substitute for sludge, to avoid confusion only the terms *sludge*, as defined above, and *biosolids* are used in this chapter and book.

Of the constituents removed by treatment, sludge is by far the largest in volume, and its processing, reuse, and disposition present perhaps the most complex problem in the field of wastewater treatment. For this reason, two chapters have been devoted to this subject. The disposition of grit and screenings is discussed in Chap. 5. The problems of dealing with sludge are complex because (1) sludge is composed largely of the substances responsible for the offensive character of untreated wastewater; (2) the portion of sludge produced from biological treatment requiring disposition is composed of the organic matter contained in the wastewater but in another form, and it, too, will decompose and become offensive; and (3) only a small part of sludge is solid matter.

The purpose of this chapter is to describe the principal processes and methods used for sludge processing and treatment as identified in Table 13–1. Resource recovery

Table 13–1

Sludge handling and processing methods

Handling or processing method	Function	See Sec.
Pumping	Transport of sludge and biosolids	13–4
Preliminary operation		13–5
Grinding	Particle size reduction	13–5
Screening	Removal of fibrous material	13–5
Degritting	Grit removal	13–5
Blending	Homogenization of sludge	13–5
Storage	Flow equalization	
Thickening		
Gravity thickening	Volume reduction	13–6
Flotation thickening	Volume reduction	13–6
Centrifugation	Volume reduction	13–6
Gravity belt thickening	Volume reduction	13–6
Rotary drum thickening	Volume reduction	13–6
Stabilization		
Alkaline stabilization	Stabilization	13–8
Anaerobic digestion	Stabilization, mass reduction, resource recovery	13–9
Aerobic digestion	Stabilization, mass reduction	13–10
Composting	Stabilization, product recovery	14–5
Heat drying	Stabilization, volume reduction, resource recovery	14–3
Conditioning	Improve dewatering	14–1
Dewatering		
Centrifuge	Volume reduction	14–2
Belt filter press	Volume reduction	14–2
Rotary press	Volume reduction	14–2
Screw press	Volume reduction	14–2
Filter press	Volume reduction	14–2
Advanced dewatering	Volume reduction and stabilization	14–2
Drying beds	Volume reduction	14–2
Reed beds	Storage and volume reduction	14–2
Lagoons	Storage and volume reduction	14–2
Advanced Thermal Oxidation	Volume and mass reduction, resource recovery	14–4
Application of biosolids to land	Beneficial use and disposition	14–10
Conveyance and storage	Transport and storage of sludge and biosolids	14–6

methods, also identified in Table 13–1, and the beneficial use of the biosolids is discussed in Chapter 14. To understand the various sludge handling and processing methods, the first two sections of this chapter are devoted to a discussion of the sources, characteristics, and quantities of sludge; the current regulatory environment; and a presentation of representative sludge-treatment process flow diagrams. Because the pumping of sludge is a fundamental part of wastewater treatment plant design, a separate discussion (Sec. 13–4) is devoted to sludge and scum pumping. The preliminary processing of sludge is discussed in Secs. 13–5 and 13–6. Stabilization of sludge is introduced in Sec. 13–7 and is divided into three subsequent sections for more detailed discussion: alkaline stabilization, anaerobic digestion, and aerobic digestion (see Secs. 13–8 through 13–10). Composting, also used for sludge stabilization after dewatering, is considered in Chap. 14.

13–1 SLUDGE SOURCES, CHARACTERISTICS, AND QUANTITIES

To design sludge processing, treatment, and disposition facilities properly, the sources, characteristics, and quantities of the sludge to be handled must be known. The method of primary and secondary treatment of wastewater has a significant impact on quantity and quality of the sludge produced. For example, using membrane bioreactors in secondary treatment produces sludge that is difficult to dewater and digest anaerobically, as compared to using a conventional waste activated sludge process. Stringent regulations for producing high quality effluent have an impact on the process used for secondary treatment, which in turn impact the quantity and quality of biosolids produced from sludge. For example, using biological nutrient removal (BNR) systems to meet stringent nutrient effluent quality produces lesser amount of sludge, but a sludge that is more difficult to process downstream by dewatering or digestion. The purpose of this section is to present background data and information on these topics that will serve as a basis for the material to be presented in the subsequent sections of this chapter.

Sources

The sources of sludge in a treatment plant vary according to the type of plant and its method of operation. The principal sources of sludge and the types generated are reported in Table 13–2. For example, in a complete mix activated sludge process, if the wasting of sludge is accomplished from the mixed liquor line or aeration chamber, the activated sludge

Table 13–2

Sources of sludge from conventional wastewater treatment plants[a]

Unit operation or process	Types of sludge	Remarks
Preaeration	Grit and scum	In some plants, scum removal facilities are not provided in preaeration tanks. If the preaeration tanks are not preceded by grit removal facilities, grit deposition may occur in preaeration tanks.
Primary sedimentation	Primary and scum	Quantities of sludge and scum depend upon the nature of the collection system and whether industrial wastes are discharged to the system.
Biological treatment	Secondary and scum	Suspended solids are produced by the biological conversion of BOD. Some form of thickening may be required to concentrate the waste sludge stream from biological treatment. Provision for scum removal from secondary settling tanks is a requirement of the U.S. EPA.

[a] The coarse material removed by screening and grit during preliminary treatment are considered in Chap. 5.

settling tank is not a source of sludge. On the other hand, if wasting is accomplished from the activated sludge return line, the activated sludge settling tank constitutes a source of sludge. Processes used for thickening, digesting, conditioning, and dewatering of sludge produced from primary and secondary settling tanks also constitute sources.

Characteristics

To treat and reuse the sludge produced from wastewater treatment plants in the most effective manner, it is important to know the characteristics of the sludge that will be processed. The characteristics vary depending on the origin of the sludge, the amount of aging that has taken place, and the type of processing to which the sludge has been subjected (see Table 13–3).

General Composition. Typical data on the chemical composition of sludges are reported in Table 13–4. Many of the chemical constituents, including nutrients, are important in considering the ultimate disposition of the processed sludge and the liquid removed during processing. The measurement of pH, alkalinity, and organic acid content

Table 13–3

Characteristics of sludge and biosolids produced during wastewater treatment

Type	Description
Scum/grease	Scum consists of the floatable materials skimmed from the surface of primary and secondary settling tanks. Scum may contain grease, vegetable and mineral oils, animal fats, waxes, soaps, food wastes, vegetable and fruit skins, hair, paper and cotton, cigarette tips, plastic materials, condoms, grit particles, and similar materials. The specific gravity of scum is less than 1.0 and usually around 0.95.
Primary sludge	Sludge from primary settling tanks is usually gray and slimy and, in most cases, has an extremely offensive odor. Primary sludge can be readily digested under suitable conditions of operation.
Sludge from chemical precipitation	Sludge from chemical precipitation with metal salts is usually dark in color, though its surface may be red if it contains much iron. Lime sludge is grayish brown. The odor of chemical sludge may be objectionable, but is not as bad as primary sludge. While chemical sludge is somewhat slimy, the hydrate of iron or aluminum in it makes it gelatinous. If the sludge is left in the tank, it undergoes decomposition similar to primary sludge, but at a slower rate. Substantial quantities of gas may be given off and the sludge density increased by long residence times in storage.
Activated sludge	Activated sludge generally has a brown flocculant appearance. If the color is dark, the sludge may be approaching a septic condition. If the color is lighter than usual, there may have been underaeration with a tendency for the sludge to settle slowly. Sludge in good condition has an inoffensive "earthy" odor. The sludge tends to become septic rapidly and then has a disagreeable odor of putrefaction. Activated sludge digests well aerobically, but not anaerobically.
Trickling filter sludge	Humus sludge from trickling filters is brownish, flocculant, and relatively inoffensive when fresh. It generally undergoes decomposition more slowly than other undigested sludges. When trickling filter sludge contains many worms, it may become inoffensive quickly. Trickling filter sludge digests readily.
Aerobically digested biosolids	Aerobically digested biosolids are brown to dark brown and have a flocculant appearance. The odor of aerobically digested sludge is not offensive; it is often characterized as musty. Well digested aerobic sludge dewaters easily on drying beds.
Anaerobically digested biosolids	Anaerobically digested biosolids are dark brown to black and contain an exceptionally large quantity of gas. When thoroughly digested, they are not offensive, the odor being relatively faint and like that of hot tar, burnt rubber, or sealing wax.

[a] The characteristics of the coarse material removed by screening and grit during preliminary treatment are considered in Chap. 5.

Table 13–4

Typical chemical composition of untreated primary and activated sludge[a]

Item	Untreated primary sludge Range	Untreated primary sludge Typical	Untreated activated sludge Range	Untreated activated sludge Typical
Total dry solids (TS),%	1–6	3	0.4–1.2	0.8
Volatile solids (% of TS)	60–85	75	60–85	70
Grease and fats (% of TS)	5–8	6	5–12	8
Protein (% of TS)	20–30	25	32–41	36
Nitrogen (N, % of TS)	1.5–4	2.5	2.4–5	3.8
Phosphorus (P_2O_5, % of TS)	0.8–2.8	1.6	2.8–11	5.5
Potash (K_2O, % of TS)	0–1	0.4	0.5–0.7	0.6
Cellulose (% of TS)	8–15	10	—	
Iron (not as sulfide)	2–4	2.5	—	
Silica (SiO_2, % of TS)	15–20	—	—	
pH	5–8	6	6.5–8	7.1
Alkalinity (mg/L as $CaCO_3$)	500–1500	600	580–1100	790
Organic acids (mg/L as HAc)	200–2000	500	1100–1700	1350
Energy content, kJ/kg VSS	23,000–29,000	25,000	19,000–23,000	20,000

[a] Adapted, in part, from U.S. EPA (1979).

Note: kJ/kg × 0.4303 = Btu/lb.

is important in process control of anaerobic digestion. The content of heavy metals, pesticides, and hydrocarbons has to be determined when incineration and land application methods are contemplated. The thermal content of sludge is important where a thermal reduction process such as incineration or gasification is considered.

Specific Constituents. Biosolids characteristics that affect their suitability for application to land and for beneficial use include organic content (usually measured as volatile solids), nutrients, pathogens, metals, and toxic organics. The fertilizer value of the biosolids, which should be evaluated where they are to be used as a soil conditioner, is based primarily on the content of nitrogen, phosphorus, and potassium (potash). Typical nutrient values of wastewater biosolids as compared to commercial fertilizers are reported in Table 13–5. In most land application systems, biosolids provide sufficient nutrients for

Table 13–5

Comparison of nutrient levels in commercial fertilizers and wastewater biosolids

Product	Nutrients, % Nitrogen	Nutrients, % Phosphorus	Nutrients, % Potassium
Fertilizers for typical agricultural use[a]	5	10	10
Typical values for stabilized wastewater biosolids (based on TS)[b]	3.3	2.3	0.3

[a] The concentrations of nutrients may vary widely depending upon the soil and crop needs.

[b] The concentrations of nutrients may vary depending on wastewater nutrient removal requirements.

Table 13–6

Typical metal content in wastewater solids[a]

Metal	Range of dry solids, mg/kg[b]
Arsenic[c]	1.18–49.2
Cadmium[c]	0.21–11.8
Chromium[c]	6.74–1160
Cobalt	0.87–290
Copper[c]	115–2580
Iron	1575–299,000
Lead[c]	5.81–450
Manganese	34.8–14,900
Mercury[c]	0.17–8.3
Molybdenum[c]	2.51–132
Nickel	7.44–526
Selenium[c]	1.1–24.7
Tin	7.5–522
Zinc[c]	216–8550

[a] US EPA (2009).

[b] Because of the wide range of values there is no typical value.

[c] Metals currently regulated under 40 CFR 503.

plant growth. In some applications, the phosphorus and potassium content may be low and require augmentation.

Trace elements are those inorganic chemical elements that, in very small quantities, can be essential or detrimental to plants and animals. The term "heavy metals" is used to denote several of the trace elements present in sludge and biosolids. Concentrations of heavy metals may vary widely, as indicated in Table 13–6. Due to successful implementation of pretreatment programs, quality of sludge and biosolids in terms regulated heavy metals improved significantly. For the application of biosolids to land, concentrations of heavy metals may limit the application rate and the useful life of the application site (see Sec. 14–10).

Quantities

Data on the quantities of sludge produced from various processes and operations are presented in Table 13–7. Although the data in Table 13–7 are useful as presented, it should be noted that the quantity of sludge produced would vary widely. Corresponding data on expected sludge concentrations from various processes are given in Table 13–8.

Quantity Variations. The quantity of semi-solid and solid material entering the wastewater treatment plant daily may be expected to fluctuate over a wide range. To ensure capacity capable of handling these variations, the following items must be considered in the design of sludge processing and disposition facilities (1) the average and maximum rates of sludge production, and (2) the potential storage capacity of the treatment units within the plant. The variation in daily quantity of sludge that may be expected in large cities is shown

Table 13–7

Typical data for the physical characteristics and quantities of sludge produced from various wastewater treatment operations and processes

Treatment operation or process	Specific gravity of solids	Specific Gravity of Sludge	Dry solids lb/10^3 gal		Dry solids kg/10^3 m^3	
			Range	Typical	Range	Typical
Primary sedimentation	1.4	1.02	0.9–1.4	1.25	110–170	150
Activated sludge	1.25	1.05	0.6–0.8	0.7	70–100	80
Trickling filter	1.45	1.025	0.5–0.8	0.6	60–100	70
Extended aeration	1.3	1.015	0.7–1.0	0.8[a]	80–120	100[a]
Aerated lagoon	1.3	1.01	0.7–1.0	0.8[a]	80–120	100[a]
Filtration	1.2	1.005	0.1–0.2	0.15	12–24	20
Algae removal	1.2	1.005	0.1–0.2	0.15	12–24	20
Chemical addition to primary tanks for phosphorous removal						
Low lime (350–500 mg/L)	1.9	1.04	2.0–3.3	2.5[b]	240–400	300[b]
High lime (800–1600 mg/L)	2.2	1.05	5.0–11.0	6.6[b]	600–1300	800[b]
Suspended growth nitrification	—	—	—	—	—	—[c]
Suspended growth denitrification	1.2	1.005	0.1–0.25	0.15	12–30	18
Roughing filters	1.28	1.02	—	—[d]	—	—[d]

[a] Assuming no primary treatment.

[b] Solids in addition to that normally removed by primary sedimentation.

[c] Negligible.

[d] Included in biosolids production from secondary treatment processes.

on Fig. 13–1. The curve is characteristic of large cities having a number of large collection lines laid on flat slopes; even greater variations may be expected at small plants.

A limited quantity of sludge may be stored temporarily in the sedimentation and aeration tanks. Where digestion tanks with varying levels are used, their large storage capacity provides a substantial dampening effect on peak digested sludge loads. In sludge treatment systems where digestion is used, the design is usually based on maximum monthly loadings to provide a minimum of 15 d residence time during these maximum month loadings. Thus, based on average daily loadings, digesters have some sludge storage capacity. Where digestion is not used, the sludge treatment processes should be designed based on the inherent storage capacity available in the sludge handling system. For example, the mechanical dewatering system following gravity thickening could be based on the maximum 1 or 3 d sludge production. Certain components of the sludge processing system, such as sludge pumping and thickening, are sized to handle the maximum-day conditions.

Volume-Mass Relationships. The volume of sludge depends mainly on its water content and only slightly on the character of the solid matter. A 10 percent sludge, for example, contains 90 percent water by weight. If the solid matter is composed of fixed

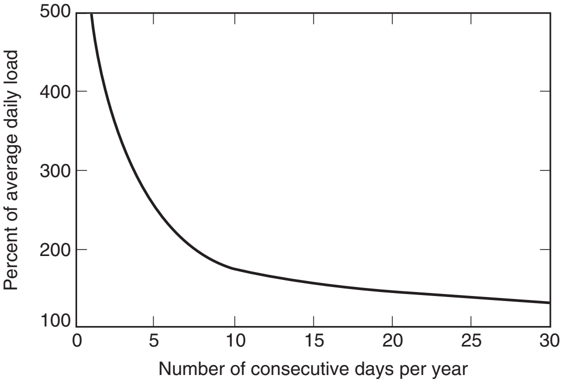
Table 13–8

Expected solids concentrations from various treatment operations and processes

Operation or process application	Solids concentration, % dry solids Range	Solids concentration, % dry solids Typical
Primary settling tank		
Primary sludge	1–6	3
Primary sludge to a cyclone degritter	0.5–3	1.5
Primary sludge and waste activated sludge	1–4	2
Primary sludge and trickling filter humus	4–10	5
Primary sludge with iron addition for phosphorus removal	0.5–3	2
Primary sludge with low lime addition for phosphorus removal	2–8	4
Primary sludge with high lime addition for phosphorus removal	4–16	10
Scum	3–10	5
Secondary settling tank		
Waste activated sludge with primary settling	0.5–1.5	0.8
Waste activated sludge without primary settling	0.8–2.5	1.3
High purity oxygen activated sludge with primary settling	1.3–3	2
High purity oxygen activated sludge without primary settling	1.4–4	2.5
Trickling filter humus	1–3	1.5
Rotating biological contactor waste sludge	1–3	1.5
Gravity thickener		
Primary sludge only	3–10	5
Primary sludge and waste activated sludge	2–6	3.5
Primary sludge and trickling filter humus	3–9	5
Dissolved air flotation thickener		
Waste activated sludge with polymer addition	4–6	5
Waste activated sludge without polymer addition	3–5	4
Centrifuge thickener (waste activated sludge only)	4–8	5
Gravity belt thickener (waste activated sludge with polymer addition)	3–6	5
Anaerobic digester		
Primary sludge	2–5	4
Primary sludge and waste activated sludge	1.5–4	2.5
Primary sludge and trickling filter humus	2–4	3
Aerobic digester		
Primary sludge only	2.5–7	3.5
Primary sludge and waste activated sludge	1.5–4	2.5
Waste activated sludge only	0.8–2.5	1.3

Figure 13–1

Peak sludge load as a function of the average daily load.



(mineral) and volatile (organic) solids, the specific gravity of all of the solid matter can be computed using Eq. (13–1).

$$\frac{W_s}{S_s\rho_w} = \frac{W_f}{S_f\rho_w} + \frac{W_v}{S_v\rho_w} \tag{13–1}$$

where W_s = weight of solids
S_s = specific gravity of solids
ρ_w = density of water
W_f = weight of fixed solids (mineral matter)
S_f = specific gravity of fixed solids
W_v = weight of volatile solids
S_v = specific gravity of volatile solids

Therefore, if one-third of the solid matter in a sludge containing 90 percent water is composed of fixed mineral solids with a specific gravity of 2.5, and two-thirds is composed of volatile solids with a specific gravity of 1.0, then the specific gravity of all solids S_s would be equal to 1.25, as follows:

$$\frac{1}{S_s} = \frac{0.33}{2.5} + \frac{0.67}{1.0} = 0.82$$

$$S_s = \frac{1}{0.82} = 1.25$$

If the specific gravity of the water is taken to be 1.0, the specific gravity of the sludge S_{sl} is 1.02, as follows:

$$\frac{1}{S_{sl}} = \frac{0.1}{1.25} + \frac{0.9}{1.0} = 0.98$$

$$S_{sl} = \frac{1}{0.98} = 1.02$$

The volume of sludge may be computed with the following expression:

$$V = \frac{M_s}{\rho_w S_s P_s} \tag{13–2}$$

where V = volume, m^3
M_s = mass of dry solids, kg
ρ_w = specific weight of water, 10^3 kg/m^3
S_{sl} = specific gravity of the sludge
P_s = percent solids expressed as a decimal

For approximate calculations for a given solids content, it is simple to remember that the volume varies inversely with the percent of solid matter contained in the sludge as given by

$$\frac{V_1}{V_2} = \frac{P_2}{P_1} \qquad \text{(approximate)}$$

Where V_1, V_2 = sludge volumes
P_1, P_2 = percent of solid matter

The application of these volume and weight relationships is illustrated in Example 13–1.

EXAMPLE 13–1 **Volume of Untreated and Digested Dewatered Sludge** Determine the liquid volume before and after digestion and dewatering and the percent reduction for 500 kg (dry basis) of primary sludge with the following characteristics:

	Primary	Digested and dewatered
Solids, %	5	20
Volatile matter, %	80	60 (destroyed)
Specific gravity of fixed solids	2.5	2.5
Specific gravity of volatile solids	≈1.0	≈1.0

Solution

1. Compute the average specific gravity of all the solids in the primary sludge using Eq. (13–1).

$$\frac{1}{S_s} = \frac{0.2}{2.5} + \frac{0.8}{1.0} = 0.88$$

$$S_s = \frac{1}{0.88} = 1.14 \quad \text{(primary solids)}$$

2. Compute the specific gravity of the primary sludge.

$$\frac{1}{S_{sl}} = \frac{0.05}{1.14} + \frac{0.95}{1} = 0.99$$

$$S_{sl} = \frac{1}{0.99} = 1.01$$

3. Compute the volume of the primary sludge using Eq. (13–2).

$$V = \frac{500\,\text{kg}}{(10^3\,\text{kg/m}^3)(1.01)(0.05)}$$
$$= 9.9\,\text{m}^3$$

4. Compute the percentage of volatile matter after digestion total volatile solids after digestion.

$$\text{Volatile matter, \%} = \frac{\text{total VS after digestion}}{\text{total TS after digestion}} \times 100$$

$$= \frac{(\text{VS Primary})M_s(1 - \text{VSR})}{M_s - M_s(\text{VS Primary})(\text{VSR})} \times 100$$

$$= \frac{(0.8)(500\text{ kg})(1 - 0.6)}{500\text{ kg} - 500\text{ kg}(0.8)(0.6)} \times 100 = 61.5\%$$

5. Compute the average specific gravity of all the solids in the digested sludge using Eq. (13–1).

$$\frac{1}{S_s} = \frac{0.385}{2.5} + \frac{0.615}{1.0} = 0.769$$

$$S_s = \frac{1}{0.769} = 1.30 \text{ (digested solids)}$$

6. Compute the specific gravity of the digested sludge (S_{ds}).

$$\frac{1}{S_{ds}} = \frac{0.20}{1.3} + \frac{0.80}{1} = 0.95$$

$$S_{ds} = \frac{1}{0.95} = 1.05$$

7. Compute the volume of digested sludge using Eq. (13–2).

$$V = \frac{500 \text{ kg} - 500\text{kg}(0.8)(0.6)}{(10^3 \text{ kg/m}^3)(1.05)(0.20)}$$
$$= 1.2\text{m}^3$$

8. Determine the percentage reduction in the sludge volume after digestion.

$$\text{Reduction} = \frac{(9.9 - 1.2) \text{ m}^3}{9.9 \text{ m}^3} \times 100 = 87.8\%$$

13–2 REGULATIONS FOR THE REUSE AND DISPOSITION OF SLUDGE IN THE UNITED STATES

In selecting the appropriate methods of sludge processing, reuse, and disposition, consideration must be given to the appropriate regulations. In the United States, regulations (40 CFR Part 503) were promulgated in 1993 by the U.S. Environmental Protection Agency (U.S. EPA) that established pollutant numerical limits and management practices for the reuse and disposition of sludge generated from the processing of municipal wastewater and septage (Federal Register, 1993). The regulations were designed to protect public health and the environment from any reasonably anticipated adverse effects of pollutants contained in the biosolids.

The regulations addressed by 40 CFR Part 503 cover specifically (1) land application of biosolids, (2) surface disposition of biosolids, (3) pathogen and vector reduction in treated biosolids, and (4) incineration. Each of these subjects is discussed below. The regulations directly affect selection of many of the processes used for sludge treatment, especially for sludge stabilization, i.e., alkaline stabilization, anaerobic digestion, aerobic digestion, and composting. In some cases, to achieve compliance, appropriate treatment requirements or methods are stipulated by the regulations. Additional discussion regarding regulations for applying biosolids on land is provided in Sec. 14–8.

Land Application

Land application relates to biosolids reuse and includes all forms of applying bulk or bagged biosolids to land for beneficial uses at agronomic rates, i.e., rates designed to provide the amount of nitrogen needed by crop or vegetation while minimizing the amount that passes below the root zone. The regulations establish two levels of biosolids quality with respect to heavy metals concentrations—pollutant ceiling and pollutant concentrations ("high" quality biosolids); two levels of quality with respect to pathogen densities—Class A and Class B; and two types of approaches for meeting vector attraction—biosolids processing or use of physical barriers. Vector attraction reduction decreases the potential for spreading infectious disease by vectors such as rodents, insects, and birds.

Surface Disposition

The surface disposition part of the Part 503 regulations applies to (1) dedicated surface disposition sites; (2) monofills, i.e., sludge-only landfills; (3) piles or mounds; and (4) impoundments or lagoons. Disposition sites and sludge placed on those sites for final disposition are addressed in the surface disposition rules. Surface disposition does not include placement of sludge for storage or treatment purposes. Where surface disposition sites do not have a liner or leachate collection system, limits are established for pollutants such as arsenic and nickel and vary based on the distance of the active surface disposition site boundary from the site property line (see Federal Register, 1993).

Pathogen and Vector Attraction Reduction

The 40 CFR Part 503 regulations divide the quality of biosolids into two categories, referred to as Class A and Class B (see requirements in Table 13–9,). Class A biosolids must meet specific criteria to ensure they are safe to be used by the general public and for nurseries, gardens, and golf courses. Class B biosolids have lesser treatment requirements than Class A, and typically are used for application to agricultural land and daily cover in a landfill.

In addition to meeting the requirements in Table 13–9, Class A biosolids must meet one of the following criteria:

- A fecal coliform density of less than 1000 MPN/g total dry solids
- Salmonella sp. density of less than 3 MPN/4 g total dry solids (3 MPN/4 g TS)

Bulk biosolids applied to lawns and home gardens or sold or given away in bags or other containers must meet the Class A criteria for pathogen reduction (see Table 13–9) and one of several vector attraction reduction processing options (see Table 13–10). Alternatively, biosolids can be treated by a prescribed process that reduces pathogens beyond detectable levels.

Class B pathogen requirements are the minimum level of pathogen reduction for land application and surface disposition. The only exception to achieving at least Class B level occurs when the sludge is placed in a surface disposition facility that is covered daily. Biosolids that do not qualify as Class B cannot be land applied. To meet Class B requirements, biosolids must be treated by a process that reduces but does not eliminate pathogens (see PSRP, also discussed below), or that must be tested to meet fecal coliform limits of less than 2.0×10^6 MPN/g TS or less than 2.0×10^6 CFU/g TS.

To meet pathogen and vector attraction reduction requirements, two levels of preapplication treatment are required and have been defined by the U.S. EPA as Processes to Further Reduce Pathogens (PFRP) and Processes to Significantly Reduce Pathogens (PSRP). These processes are defined in Tables 13–11 and 13–12. Because PSRPs reduce but do not eliminate pathogens, PSRP-treated biosolids still have the potential to transmit disease. Because PFRPs reduce pathogens below detectable levels, there are no pathogen-related restrictions for land application. Minimum frequency of monitoring, record-keeping, and reporting requirements must be met, however.

Table 13–9

Pathogen reduction alternatives[a]

Class A	Description
Alternative 1	Thermally Treated Sewage Sludge: Use one of four time-temperature regimes.
Alternative 2	Sewage Sludge Treated in a High pH-High Temperature Process: Specifies pH, temperature, and air-drying requirements.
Alternative 3	For Sewage Sludge Treated in Other Processes: Demonstrate that the process can reduce enteric viruses and viable helminth ova. Maintain operating conditions used in the demonstration.
Alternative 4	Sewage Sludge Treated in Unknown Processes: Demonstration of the process is unnecessary. Instead, test for pathogens—*Salmonella* sp. bacteria, enteric viruses, and viable helminth ova—at the time the sewage sludge is used or disposed, or is prepared for sale or give-away in a bag or other container for application to the land, or when prepared to meet the requirements in 503.10(b), (c), (e), or (f).
Alternative 5	Use of PFRP: Sewage sludge is treated in one of the processes to further reduce pathogens (PFRP).
Alternative 6	Use of a Process Equivalent to PFRP: Sewage sludge is treated in a process equivalent to one of the PFRPs, as determined by the permitting authority.
Class B	**Description**
Alternative 1	Monitoring of Indicator Organisms: Test for fecal coliform density as an indicator for all pathogens at the time of sewage sludge use or disposal.
Alternative 2	Use of PSRP: Sewage sludge is treated in one of the processes to significantly reduce pathogens (PSRP).
Alternative 3	Use of Processes Equivalent to PSRP: Sewage sludge is treated in a process equivalent to one of the PSRPs, as determined by the permitting authority.

[a] From US EPA (1992).

[b] In addition to meeting the requirements in one of the six alternatives listed below, fecal coliform or *Salmonella* spp. bacterial levels must meet specific densities at the time of sewage sludge use or disposal, when prepared for sale or give-away in a bag or other container for application to the land, or when prepared to meet the requirements in 503.10(b), (c), (e), or (f).

[c] The requirements in one of the three alternatives below must be met in addition to Class B site restrictions.

Incineration

Originally, the definition of a nonhazardous solid waste material included sludges from wastewater treatment and other secondary material being discarded. However, wastewater sludges were eventually exempted from being defined under the nonhazardous solid waste ruling and were instead covered under 40 *CFR* 503 regulations.

Resource Conservation and Recovery Act. Under the Resource Conservation and Recovery Act, in the Identification of Non-Hazardous Secondary Materials that are Solid Waste (published by the U.S. EPA on March 21, 2011), sludge is defined somewhat differently. If the secondary material (e.g., sludge) is being discarded, it falls under the solid waste definition. As such, the burning of sludges in sewage sludge incinerators (SSIs) would be considered discarding, and, thus, would fall under the “other discarded material” part of the solid waste definition. Accordingly, SSIs would be regulated under Sec. 129 of

Table 13–10

Vector attraction reduction[a]

Requirement	What is required?	Most appropriate for:
Option 1 503.33(b)(1)	At least 38 percent reduction in volatile solids during biosolids treatment	Biosolids processed by: Anaerobic biological treatment Aerobic biological treatment Chemical oxidation
Option 2 503.33(b)(2)	Less than 17 percent additional volatile solids loss during bench–scale anaerobic batch digestion of the biosolids for 40 additional d at 30 to 37°C (86 to 99°F)	Only for anaerobically digested biosolids
Option 3 503.33(b)(3)	Less than 15 percent additional volatile solids reduction during bench–scale aerobic batch digestion for 30 additional d at 20°C (68°F)	Only for aerobically digested biosolids with 2 percent or less solids—e.g., biosolids treated in extended aeration plants
Option 4 503.33(b)(4)	SOUR at 20°C (68°F) is 1.5 mg O_2/h·g total biosolids solids	Biosolids from aerobic processes (should not be used for composted sludges). Also for biosolids that has been deprived of oxygen for longer than 1 to 2 h
Option 5 503.33(b)(5)	Aerobic treatment of the biosolids for at least 14 d at over 40°C (104°F) with an average temperature of over 45°C (113°F)	Composted biosolids (Options 3 and 4 are likely to be easier to meet for biosolids from other aerobic processes)
Option 6 503.33(b)(6)	Addition of sufficient alkali to raise the pH to at least 12 at 25°C (77°F) and maintain a pH of 12 for 2 h and a pH of 11.5 for 22 more h	Alkali-treated biosolids (alkalies include lime, fly ash, kiln dust, and wood ash)
Option 7 503.33(b)(7)	Percent solids of 75 percent prior to mixing with other materials	Biosolids treated by an aerobic or anaerobic process (i.e., biosolids that do not contain unstabilized sludge generated in primary wastewater treatment)
Option 8 503.33(b)(8)	Percent solids of 90 percent prior to mixing with other materials	Biosolids that contain unstabilized sludge generated in primary wastewater treatment (e.g., any heat-dried sludges)
Option 9 503.33(b)(9)	Biosolids is injected into soil so that no significant amount of biosolids is present on the land surface 1 hour after injection, except Class A biosolids which must be injected within 8 h after the pathogen reduction process	Liquid biosolids applied to the land. Domestic septage applied to agricultural land, a forest, or a reclamation site
Option 10 503.33(b)(10)	Biosolids is incorporated into the soil within 6 h after application to land. Class A biosolids must be applied to the land surface within 8 h after the pathogen reduction process, and must be incorporated within 6 h after application	Biosolids applied to the land. Domestic septage applied to agricultural land, forest, or a reclamation site

[a] From U.S. EPA (1992).

the Clean Air Act, instead of how they have been regulated historically, under Sec. 112 of the Clean Air Act through Part 503 biosolids regulations.

Clean Air Act. Section 129 of the Clean Air Act requires the U.S. EPA to develop standards for solid waste combustion processes. As a result, the U.S. EPA was required to develop new source performance standards (NSPSs) and emission guidelines (EGs) for sewage sludge incineration units (SSIs). The new standards and emission guidelines were finalized by the U.S. EPA on February 21, 2011 and published in the Federal Register on March 21, 2011.

Table 13–11

Regulatory definition of processes to further reduce pathogens (PFRP)[a]

Process	Definition
Composting	Using either within-vessel or static aerated pile composting, the temperature of the biosolids is maintained at 55°C or higher for 3 d. Using windrow composting, the temperature of the wastewater sludge is maintained at 55°C or higher for 15 d or longer. During this period, a minimum of five windrow turnings is required.
Heat drying	Dewatered biosolids are dried by direct or indirect contact with hot gases to reduce the moisture content to 10 percent or lower. Either the temperature of biosolids particles exceed 80°C or the wet bulb temperature of the gas stream in contact with the biosolids as the biosolids leave the dryer exceeds 80°C.
Heat treatment	Liquid biosolids are heated to a temperature of 180°C or higher for 30 min.
Thermophilic aerobic digestion	Liquid biosolids are agitated with air or oxygen to maintain aerobic conditions, and the MCRT is 10 d at 55 to 60°C.
Beta ray irradiation	Biosolids are irradiated with beta rays from an accelerator at dosages of at least 1.0 megarad (Mrad) at room temperature (approximately 20°C).
Gamma ray irradiation	Biosolids are irradiated with gamma rays from certain isotopes such as 60 Cobalt or 135 Cesium at dosages of at least 1.0 Mrad at room temperature (approximately 20°C).
Pasteurization	The temperature of the biosolids is maintained at 70°C or higher for at least 30 min.

[a] Federal Register (1993).

Table 13–12

Regulatory definition of processes to significantly reduce pathogens (PSRP)[a]

Process	Definition
Aerobic digestion	Biosolids are agitated with air or oxygen to maintain aerobic conditions for a MCRT and temperature between 40 d at 20°C and 60 d at 15°C.
Air drying	Biosolids are dried on sand beds or on paved or unpaved basins for a minimum of 3 mo. During 2 of the 3 mo, the ambient average daily temperature exceeds 0°C.
Anaerobic digestion	Biosolids are treated in the absence of air between an MCRT of 15 d at temperatures of 35 to 55°C and an MCRT of 60 d at a temperature of 20°C. Times and temperatures between these endpoints may be calculated by linear interpolation.
Composting	Using either within-vessel, static aerated pile, or windrow composting, the temperature of the biosolids is raised to 40°C or higher for 5 d. For 4 h during the 5 d period, the temperature in the compost pile should exceed 55°C.
Lime stabilization	Sufficient lime is added to raise the pH of the biosolids to pH 12 and maintained for 2 h of contact.

[a] Federal Register (1993).

Emission Guidelines and New Source Performance Standards. The new rule requires facilities to meet the maximum achievable control technology (MACT) limits. The MACT standards for existing units were based on the best performing 12 percent of the existing units, while MACT standards for new or "modified" units are based on the "best controlled similar unit." MACT standards have been set for nine pollutants. These pollutants are: cadmium (Cd), lead (Pb), mercury (Hg), particulate matter (PM), carbon monoxide (CO), hydrogen chloride (HCl), sulfur dioxide (SO_2), nitrogen oxides (NOx), and dioxins and furans (PCDD/PCDF).

The new SSI rules and emission guidelines contain standards for existing and new multiple-hearth furnaces (MHFs) and fluidized bed incinerators (FBIs). This rule stipulates that all SSIs will require Title V operating permits, annual operator training, annual stack testing and/or continuous emissions monitoring systems, recordkeeping requirements, and establishment of operating limits. For new or modified SSIs, the proposed MACT limits are set as a composite of the best emissions performance from the best SSIs tested. For new SSIs, owners or operators are required to conduct a siting analysis prior to construction. This analysis would include site specific analysis of air pollution control alternatives to minimize the environmental and health impacts to the maximum extent practicable. Details regarding air pollution controls are discussed in Sec. 14–6 in Chap.14.

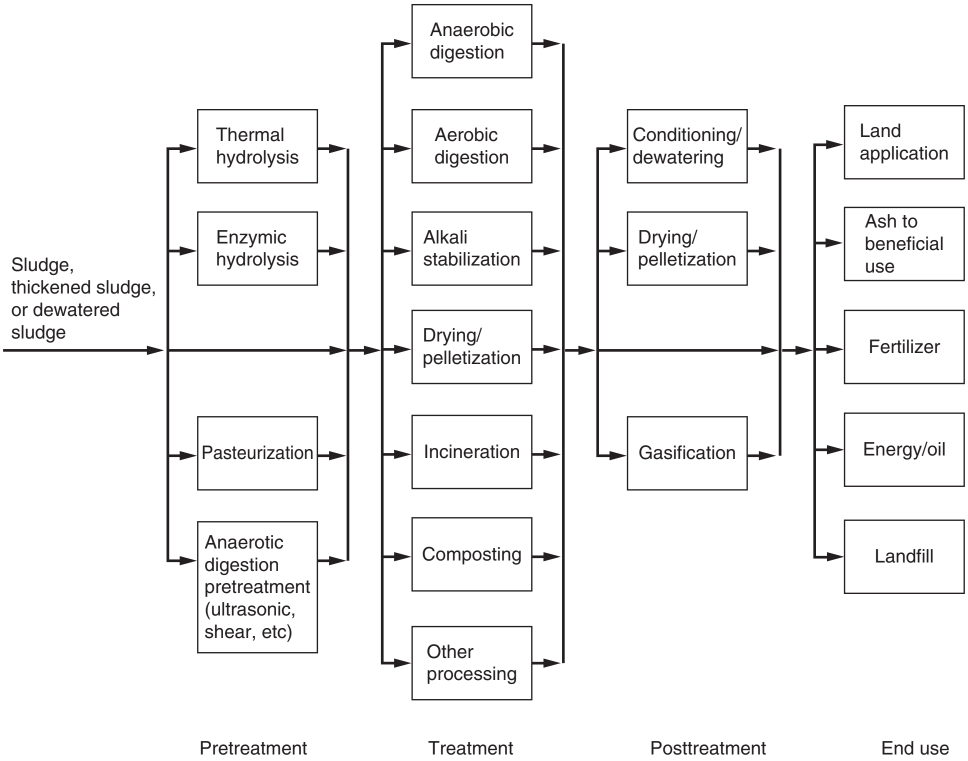
Clean Water Act. Under the 503 regulations, the requirements for biosolids incineration are still applicable. There is some overlap between the 503 regulations and the Clean Air Act Section 129 MACT based EGs and NSPSs, but they differ significantly due to the different approaches used to develop each set of rules. The 503 regulations are based on a risk based approach and aim to avoid adverse impacts. Limits are based partially on maximum allowable concentrations of pollutants within the feed biosolids coupled with stack monitoring of total hydrocarbons (or carbon monoxide) and operational standards to ensure good combustion and emissions performance. However, the MACT rules are technology based and set the limits relative to the best performing incinerator units within their class (ie. MHF or FBI). These limits are expressed as maximum concentrations of pollutants in the incinerator flue gases. The methods of measurement and media in which the concentrations are measured are entirely different and are not directly comparable to the 503 regulations. In practical application, the MACT emission requirements are much more stringent than previously required under Part 503 and they will generally dictate the required emission performance levels for both new and existing incinerators. For the time being, both sets of rules apply with overlapping requirements leading to duplicate sampling by operators of both the feed biosolids and the flue gases.

13–3 SLUDGE PROCESSING FLOW DIAGRAMS

A generalized flow diagram incorporating the unit operations and processes to be discussed in this chapter and chapter 14 is presented on Fig. 13–2. As shown, an almost infinite number of combinations are possible. In practice, the most commonly used process flow diagram for sludge processing involves biological treatment. Typical flow diagrams incorporating biological processing are presented on Fig. 13–3. Thickeners may be used depending upon the source of sludge and the method of sludge stabilization, dewatering, and disposition. Following biological digestion, any of the several methods shown may be used to dewater the sludge; the choice depends on economic evaluation, beneficial use requirements, and local conditions. In instances where biological stabilization is not used, dewatered sludge undergoes thermal decomposition, in either multiple-hearth or fluidized-bed incinerators. Furthermore, unstabilized dewatered cake can be dried, alkaline stabilized, or hauled to a landfill.

Figure 13–2

Generalized sludge processing flow diagram.



13–4 SLUDGE AND SCUM PUMPING

Sludge produced in wastewater treatment plants must be conveyed from point to point in the plant in conditions ranging from a watery sludge or scum to a thick sludge. Sludge may also be pumped off-site for long distances for treatment and disposition. For each type of sludge and pumping application, a different type of pump may be needed (see Table 13–13).

Pumps

Pumps used most frequently to convey sludge include the plunger, progressive cavity, hose, solids handling centrifugal (screw centrifugals and traditional “non-clog designs), recessed impeller, diaphragm, high-pressure piston diaphragm, and rotary lobe types. Other types of pumps such as hydraulic piston slurry pumps have also been used to pump sludge. Chopper pumps are used extensively for pumping scum containing rags, plastics, and other fibrous materials that require shredding. The advantages and disadvantages of each type of pump are summarized in Table 13–14.

Plunger Pumps. Plunger pumps [see Fig. 13–4(a)] have been used frequently for sludge applications, especially primary sludges, and have proved to be quite satisfactory. The advantages of plunger pumps are as follows:

1. Pulsating action of simplex and also duplex pumps tends to concentrate the sludge in the hoppers ahead of the pumps and resuspend solids in pipelines when pumping at low velocities.
2. They are suitable for suction lifts up to 3 m (10 ft) and are self-priming.

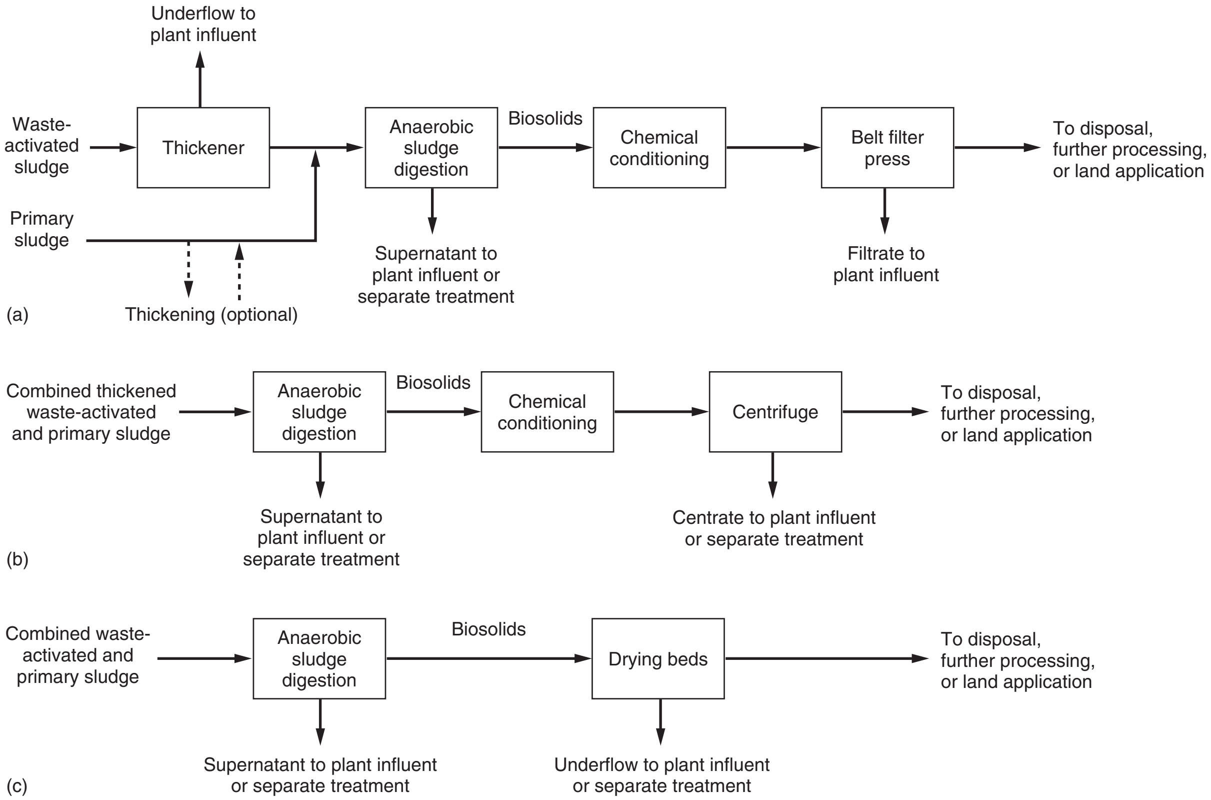


Figure 13–3

Typical sludge treatment flow diagrams with biological digestion and three different sludge dewatering processes: (a) belt filterpress, (b) centrifuge, (c) drying bed. In some plants, flows that are to be returned to the headworks are stored in equalization basins for return to the treatment process during the early morning hours when the plant load is reduced.

3. Low pumping rates can be used with large port openings.
4. Positive delivery is provided unless some object prevents the ball checks valves from seating.
5. They have constant but adjustable capacity, regardless of large variations in pumping head.
6. Discharge pressure limitations are approximately 10 to 11 bar (150 to 165 $lb_f/in.^2$).
7. Heavy sludge concentrations may be pumped if the equipment is designed for the load conditions.

Plunger pumps come with one, two, or three plungers (called simplex, duplex, or triplex units) with capacities of 2.5 to 3.8 L/s (40 to 60 gal/min) per plunger, and larger models are available. Pump speeds should be between 40 and 50 strokes per min. Because grease accumulations in sludge lines cause a progressive increase in head with use, heavier duty pumps should be designed for a minimum head of 6.9 bar (100 $lb_f/in.^2$). Capacity is decreased in constant speed pumps by shortening the stroke of the plunger; however, the pumps seem to operate more satisfactorily at or near full stroke. For this reason, many pumps are provided with variable-speed drives for speed control of capacity. A plunger pump differs from a centrifugal or recessed impeller pump in that its

Table 13–13

Application of pumps to types of sludge and biosolids[a]

Type of sludge or solids	Applicable pump	Comment
Ground screenings	Pumping screenings should be avoided	Pneumatic ejectors may be used.
Grit	Torque flow centrifugal	The abrasive character of grit and the presence of rags make grit difficult to handle. Hardened casings and impellers should be used for torque flow pumps. Pneumatic ejectors may also be used.
Scum	Plunger, progressive cavity, diaphragm, centrifugal, chopper	Scum is often pumped by the sludge pumps; valves are manipulated in the scum and sludge lines to permit this. In larger plants separate scum pumps are used. Scum mixers are often used to ensure homogeneity prior to pumping. Pneumatic ejectors may also be used.
Primary sludge	Plunger, centrifugal torque flow, diaphragm, progressive cavity, rotary lobe, chopper, hose	In most cases, it is desirable to obtain as concentrated a sludge as practicable from primary sedimentation tanks, usually by collecting the sludge in hoppers and pumping intermittently, allowing the sludge to collect and consolidate between pumping periods. The character of untreated primary sludge will vary considerably, depending on the characteristics of the solids in the wastewater, and the types of treatment units and their efficiency. Where biological treatment follows, the quantity of sludge from (1) waste activated sludge (2) humus sludge from settling tanks following trickling filters, (3) overflow liquors from digestion tanks, (4) and centrate or filtrate return from dewatering operations will also affect the sludge characteristics. In many cases, the character of the sludge is not suitable for the use of conventional nonclog centrifugal pumps. Where sludge contains rags, chopper pumps may be used.
Chemical precipitation	Same as for primary sludge	The precipitate may contain large amounts of inorganic constituents depending on the type and amount of chemicals used.
Trickling filter humus	Nonclog and torque flow centrifugal, progressive cavity, plunger, diaphragm	Humus is usually of homogenous character and can be easily pumped.
Return or waste activated sludge	Nonclog and torque flow centrifugal, progressive cavity, diaphragm	Sludge is dilute and contains only fine solids so that nonclog pumps may be used. For nonclog pumps, slow speeds are recommended to minimize the breakup of flocculent particles.
Thickened or concentrated sludge	Plunger, progressive cavity, diaphragm, high pressure piston, rotary lobe, hose	Positive displacement pumps are most applicable for concentrated sludge because of their ability to generate movement of the sludge mass. Torque flow pumps may be used but may require the addition of flushing or dilution facilities.
Digested biosolids	Plunger, torque flow centrifugal, progressive cavity, diaphragm, high pressure piston, rotary lobe	Well digested biosolids are homogenous, containing 2 to 5 percent total solids and a quantity of gas bubbles. Poorly digested biosolids may be difficult to handle. If good screening and grit removal are provided, nonclog centrifugal pumps may be considered.

[a] Adapted in part from U.S. EPA (1979).

Table 13–14

Advantages and disadvantages of various types of sludge pumps[a]

Type of Pump	Advantages	Disadvantages
Plunger	• Can pump heavy sludge concentrations (up to 15 percent) • Self-priming and can handle suction lifts up to 3 m (10 ft) • Constant but adjustable capacity regardless of variations in head • Cost-effective choice for flowrates up to 30 L/s (500 gal/min) and heads up to 60 m (200 ft) • Pulsating action of simplex and duplex pumps sometimes helps to concentrate sludge in hoppers ahead of pumps and resuspended solids in pipelines when pumping at low velocities • High Pressure Capacity	• Low efficiency • High maintenance if operated continuously • Depending on downstream processes, pulsating flow may not be acceptable
Progressing Cavity	• Provides a relatively smooth flow • Pumps greater than 3 L/s (50 gal/min) capacity can pass solids of about 20 mm (0.8 in.) in size • Easily controlled flowrates • Minimal pulsation • Relatively simple operation • Stator/rotor tends to act as a check valve, thus preventing backflow through pump. An external check valve may not be required	• Stator will burn out if pump is operated dry; needs a run dry protection system • Smaller pumps usually require grinders to prevent clogging • Power cost escalates when pumping heavy sludge • Grit in sludge may cause excessive stator wear • Seals and water required typically
Diaphragm	• Pulsating action may help to concentrate sludge in hoppers ahead of pumps and resuspend solids in pipelines when pumping at low velocities • Self-priming with suction lifts up to 3 m (10 ft) • Can pump grit with relatively minimum wear • Relatively simple operation	• Depending on downstream processes, pulsating flow may not be acceptable • Requires a source of compressed air • Operation may be excessively noisy • Low head and efficiency • High maintenance if operated continuously
Centrifugal Nonclog (mixed flow)	• Has high volume and excellent efficiency for activated sludge pumping applications • Relatively low cost	• Not recommended for other sludge pumping applications because of potential clogging due to rags and other debris
Recessed Impeller	• Because of recessed impeller design, pump can pass large solids and grit • Can pump digested sludges up to approximately 4%	• Low efficiency-about 5 to 20 percent lower than standard nonclog pumps • Limited to raw sludge with solid concentrations of 2.5 percent or less • Abrasion-resistant impellers cannot be trimmed to modify pumping characteristics
Chopper	• Reduces clogging of pump suction • May eliminate need for grinder or comminutor • Can handle higher sludge concentrations than nonclog pumps	• Relatively low efficiency-efficiency ranges from about 40 to 60 percent • Requires a level of maintenance similar to grinders
Rotary Lobe	• Provides a relatively smooth flow • Does not require a check valve in most applications with low to moderate discharge static heads • Able to run dry for short period of time without significant damage Low speed and low maintenance	• Because of close tolerances between rotating lobes, grit will cause excessive wear, thus reducing pumping efficiency • Fluid pumped must act as a lubricant • Cost of pumping increases with volume
Peristalic Hose	• Has self-priming capabilities • Because it is a positive-displacement pump, it is capable of metering flow • Relatively simple to maintain • Can pump sludge with abrasive grit	• Depending on downstream processes, pulsating flow may not be acceptable • High starting torque (two to three times running torque) • Replacement hoses may be expensive
High-pressure pistion	• Can be used to pump thickened sludge long distances • Can pump at rates of 30 L/s (500 gal/min) at pressures up to 13,800 kPa (2000 $\text{lb}_f/\text{in.}^2$) • Can run dry without major damage • Unobstructed internal flow path; can pass large solids	• High Capital Cost • Requires skilled maintenance personnel

[a] Adapted in part from WEF (2010a).

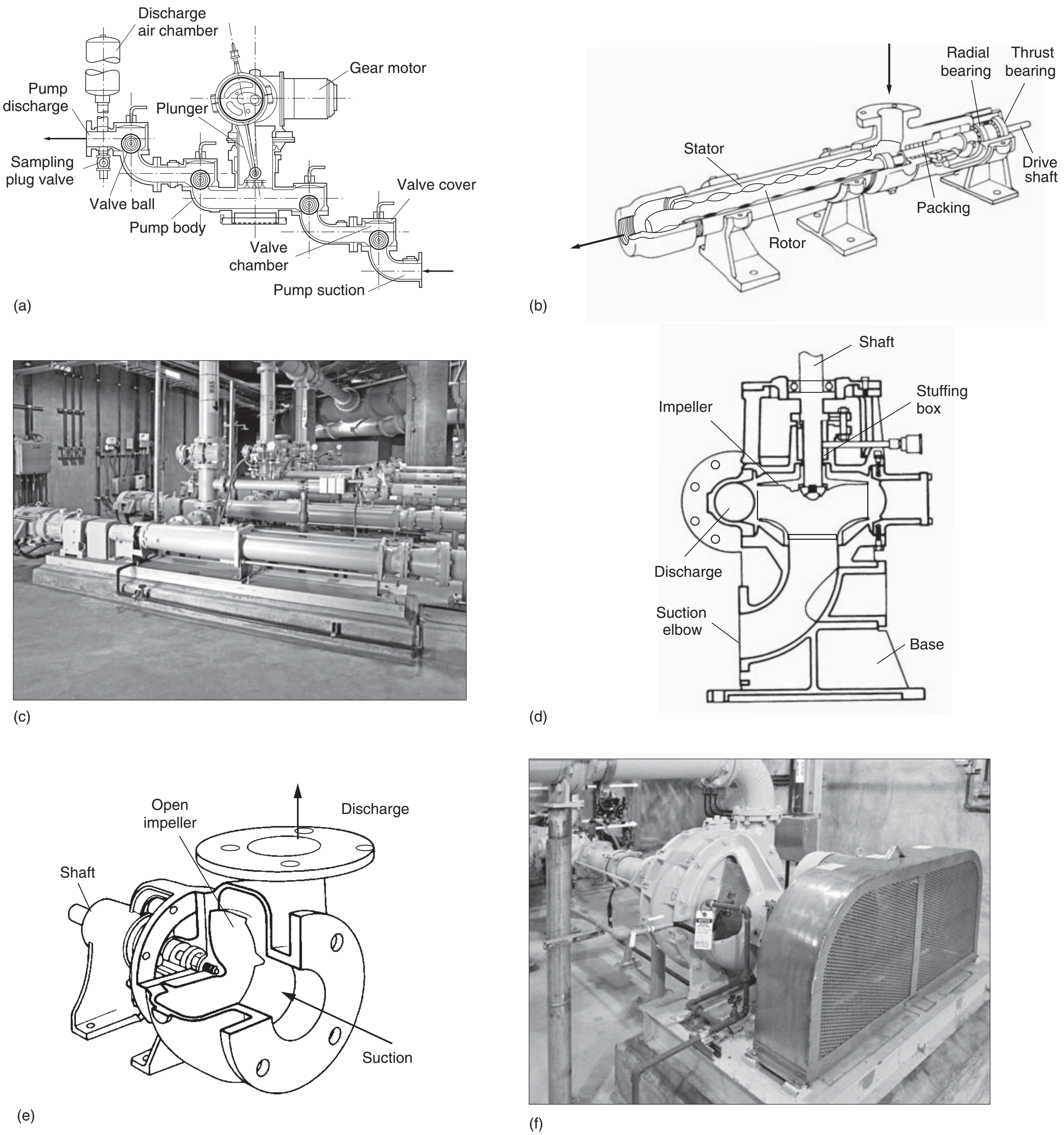
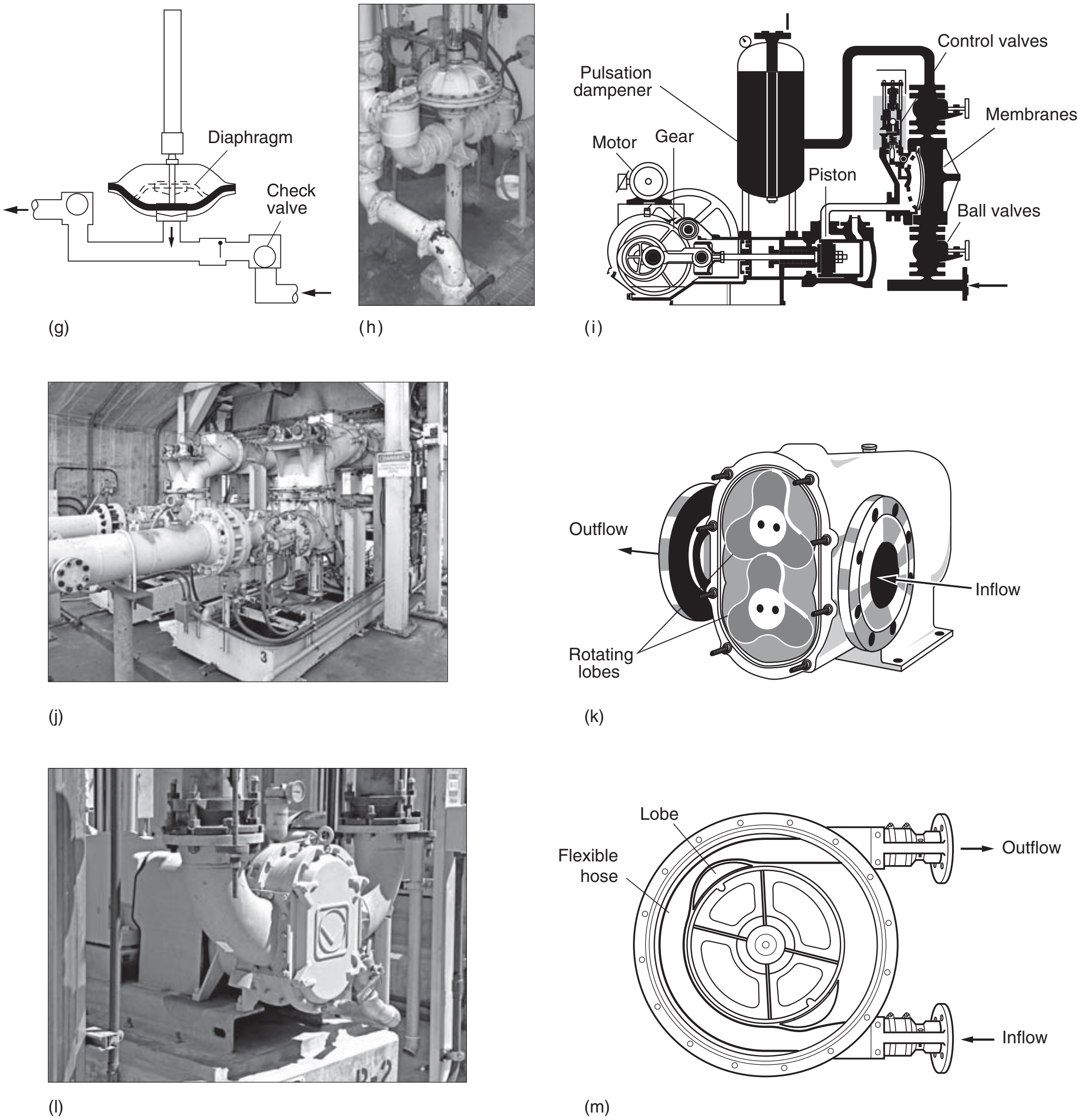


Figure 13–4

Typical sludge and scum pumps used in wastewater treatment plants: (a) plunger pump, (b) progressive cavity pump, (c) view of progressive cavity pump installation (d) section through nonclog centrifugal pump, (e) section through torque flow pump, (f) view of belt driven torque flow pump. *(Figure continues on next page.)*



(g) (h) (i)

(j) (k)

(l) (m)

Figure 13–4
(*Continued*)

Typical sludge and scum pumps used in wastewater treatment plants: (g) schematic of diaphragm pump, (h) view of diaphragm pump, (i) schematic of high pressure piston pump, (j) view of a duplex high pressure piston pump, (k) section through rotary lobe pump, (l) view of rotary lobe pump (m) section through peristaltic hose pump.

discharge is pulsing due to the action of a piston; consequently, the actual flow while sludge is moving in the pipeline is greater than average pumping capacity. The headloss calculations, therefore, must be based on the peak pulsating flow rather than the design flow. The factors given in Table 13–15 can be used to account for the actual peak pulsating or instantaneous flow.

Table 13–15

Factors for computing peak pulsating flowrate when using plunger pumps

Type of plunger pump	Actual pulsating peak flowrate
Simplex	3.1 × design flowrate
Duplex	1.55 × design flowrate
Triplex	1.2 × design flowrate

Progressive Cavity Pumps. The progressive cavity pump [see Fig. 13–4(b) and (c)] has been used successfully on almost all types of sludges. The pump is composed of a single-threaded rotor that operates with a minimum of clearance in a double-threaded helix elastomer stator. A volume or "cavity" moves progressively from suction to discharge when the rotor turns. The pump is self-priming at suction lifts up to 8.5 m (28 ft), but it must not be operated dry or it will burn out the elastomer stator. Progressive cavity pumps are available in capacities up to 126 L/s (2000 gal/min) and may be operated at discharge heads of 48 bar (720 $lb_f/in.^2$) with sludge. This type of pump requires oversizing to meet system conditions over the life of the of equipment. For example, if a 9.5 L/s (150 gal/min) pump is required the pump selection should be sized for an additional 50 percent or a pump sized for 14.25 L/s (225 gal/min). Speed for sludge applications should be limited to approximately 250 rev/min. For sludges and for systems feeding dewatering equipment, a grinder normally precedes these pumps. The pumps are expensive to maintain because of wear of the rotors and stators, particularly in primary sludge pumping applications where grit is present. For primary sludge applications consideration should be given to recessed impeller pumps. Advantages of the pumps are (1) the flowrates are controlled easily using variable speed drives, (2) pulsation is minimal, and (3) operation is relatively simple.

Centrifugal Pumps. Centrifugal pumps of solids handling or "non-clog" design [see Fig. 13–4(d)] are commonly used to pump activated sludge. In centrifugal pumping applications, the problem is choosing the proper number and capacity to accommodate the typical wide range of flowrates required. At any given speed, centrifugal pumps operate well only if the pumping head is within a relatively narrow range; the variable nature of sludge, however, causes pumping heads to change. The selected pumps must have sufficient clearance to pass the solids without clogging and have a small enough capacity to avoid pumping a sludge diluted by large quantities of wastewater overlying the sludge blanket. Throttling the discharge to reduce the capacity is impractical because of frequent stoppages; hence it is absolutely essential that these pumps be equipped with variable-speed drives. Centrifugal pumps of special design: recessed impeller and "chopper" type pumps have been used for pumping primary sludge.

Recessed impeller pumps [see Figs. 13–4(e) and (f)] have impellers that are fully recessed and are very effective in conveying sludge and higher sludge concentrations than the solids handling centrifugal pumps. The size of particles that can be handled is limited only by the diameter of the suction or discharge openings. The rotating impeller develops a vortex in the sludge so that the main propulsive force is the liquid itself. Most of the fluid does not actually pass through the vanes of the impeller, thereby minimizing abrasive contact; however, pumps used in sludge service are recommended to have nickel or chrome abrasion-resistant volute and impellers. The pumps can operate only over a narrow head range at a given speed, so the system operating conditions must be evaluated carefully. Variable speed control is recommended where the pumps are expected to operate over a wide range of head conditions. For high-pressure applications, multiple pumps may be used, connected together in series.

Chopper-type pumps have a cutter knife attached to a non-clog impeller that agitates and breaks up large solids that tend to block the pump suction. Incoming sludge is chopped by sharpened impeller blades that turn across the cutter bar. Chopper pumps are manufactured in sizes up to 380 L/s (6000 gal/min) in both horizontal and vertical dry pit configurations as well as submersible configurations.

Slow-speed centrifugal and mixed-flow pumps are commonly used for returning activated sludge to the aeration tanks. Screw pumps are also being used for this service especially where pumps are required to have a large turndown. Screw centrifugal pumps tend to have less clogging issues for these applications.

Diaphragm Pumps. Diaphragm pumps use a flexible membrane that is pushed and pulled to contract and enlarge an enclosed cavity [see Fig. 13–4(g) and (h)]. Flow is directed through this cavity by check valves, which may be either ball or flap type. The capacity of a diaphragm pump is altered by changing either the length of the diaphragm stroke or the number of strokes per minute. Pump capacity can be increased and flow pulsations smoothed out by providing two pump chambers and using both strokes of the diaphragm for pumping. Diaphragm pumps are relatively low capacity and low head; the largest available air diaphragm pump delivers 14 L/s (220 gal/min) against 15 m (50 ft) of head.

High-Pressure Piston Diaphragm Pumps. High-pressure piston pumps are used in high-pressure applications such as pumping sludge long distances. Several types of piston pumps have been developed for high-pressure applications and are similar in action to plunger pumps. The high-pressure piston pumps use separate power pistons or membranes or diaphragms to separate the drive mechanisms from contacting the sludge. A schematic of a piston pump is shown on Fig 13–4(i). A view of a duplex piston pump is shown on Fig. 13–4(j). Advantages of these types of pumps are (1) they can pump relatively small flowrates at high pressures, up to 13.8 bar (note 1 bar = 100 kPa) (200 $\text{lb}_f/\text{in.}^2$), (2) large solids up to the discharge pipe diameter can be passed, (3) a range of sludge concentrations can be handled, and (4) the pumping can be accomplished in a single stage. The pumps, however, are very expensive.

Rotary Lobe Pumps. Rotary lobe pumps [see Fig. 13–4(k) and (l)] are positive displacement pumps in which two rotating synchronous lobes push the fluid through the pump. Rotational speed and shearing stresses are low. For sludge pumping, lobes are made of hard metal or hard rubber. This type of pump requires oversizing to meet system conditions over the life of the of equipment. For example if a 9.5 L/s (150 gal/min) pump is required the pump selection should be sized for an additional 50 percent or a pump sized for 14.25 L/s (225 gal/min). Speed for sludge applications should be limited to approximately 250 to 300 rev/min depending on the abrasiveness of the sludge. An advantage cited for the rotary lobe pump is that lobe replacement is less costly than rotor, and stator replacement for progressive cavity pumps and the space required for installation is less. Rotary lobe pumps, like other positive-displacement pumps, must be protected against pipeline obstructions.

Hose Pumps. Peristaltic hose pumps [see Fig. 13–4(m)] have also been used for pumping sludge. The pump works by alternately compressing and relaxing a specially designed resilient but reinforced hose. The hose is compressed between the inner wall of the pump housing and the compression shoes on the rotor. A lubricant is used to reduce heat and wear on the hose. The pumped sludge only comes in contact with the inner wall of the hose, which cushions entrained abrasives during compression. The pumps are available in capacities ranging from 36 to 1250 L/min (10 to 330 gal/min). As a positive-displacement pump, the pump output is directly proportional to speed at either high or low discharge pressures. The primary disadvantages of the hose pump are the pulsating flow, hose wear, and the relatively high cost of hose replacement.

Headloss Determination

The headloss encountered in the pumping of sludge depends on the flow properties (rheology) of sludge, the pipe diameter, and the flow velocity. It has been observed that headlosses increase with increased solids content, increased volatile content, and lower temperatures. When the percent volatile matter multiplied by the percent solids exceeds 600, difficulties may be encountered in pumping sludge.

Water, oil, and most other fluids are "Newtonian," which means that the pressure drop is proportional to the velocity and viscosity under laminar flow conditions. As the velocity increases past a critical value, the flow becomes turbulent. Dilute sludges such as unconcentrated activated and trickling-filter sludges behave similar to water. Concentrated wastewater sludges, however, are non-Newtonian fluids. The pressure drop under laminar conditions for non-Newtonian fluids is not proportional to flow, so the viscosity is not a constant. Special procedures may be used to determine headloss under laminar-flow conditions, and the velocity at which turbulent flow begins. In this section both the simplified approach of calculating headloss and a method using the sludge rheology will be discussed.

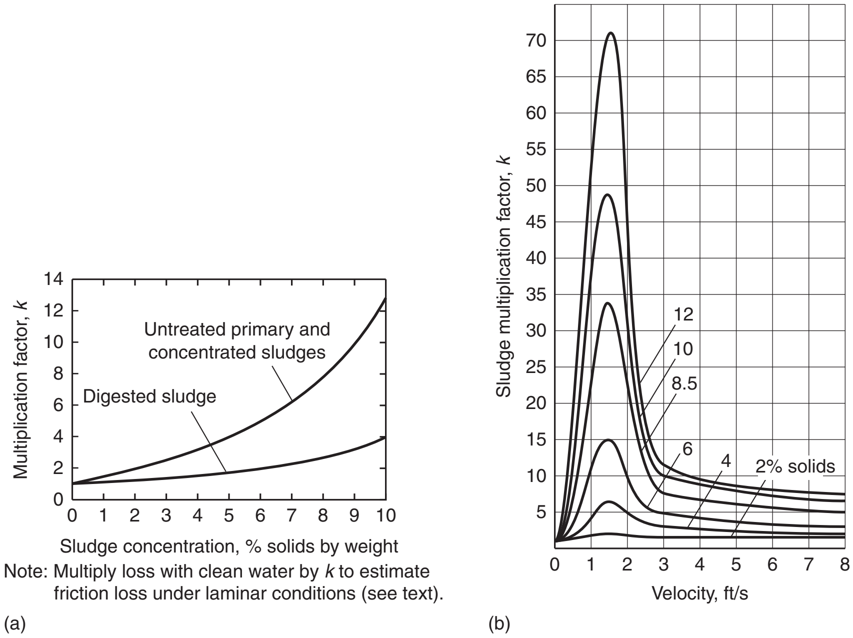
The headloss in pumping unconcentrated activated and trickling-filter sludges may be from 10 to 25 percent greater than for water. Primary, digested, and concentrated sludges at low velocities may exhibit a plastic-flow phenomenon in which a definite pressure is required to overcome resistance and start flow. The resistance then increases approximately with the first power of the velocity throughout the laminar range of flow, which extends to about 1.1 m/s (3.5 ft/s), the lower critical velocity. Above the higher critical velocity at about 1.4 m/s (4.5 ft/s), the flow may be considered turbulent. In the turbulent range, the losses for well-digested sludge may be more than two to three times the losses for water. The losses for primary and concentrated sludges, especially those conditioned with polymer, and scum may be considerably greater. The risk of underestimating the headloss also increases as the piping distance and sludge concentration increases. Where possible, particularly in long-distance sludge pumping, hydraulic studies should be conducted to confirm the ranges of headloss characteristics.

Simplified Headloss Computations. Relatively simple procedures are used to compute headloss for short sludge pipelines. The accuracy of these procedures may be adequate, especially at sludge solids concentrations less than 3 percent by weight. To determine the headloss, the factor k is obtained from Fig. 13–5(a) for a given solids content and type of sludge. The headloss when pumping sludge is computed by multiplying the headloss of water, determined by using the Darcy-Weisbach, Hazen-Williams, or Manning equations, by k. The values given on Fig. 13–5(a) should be used only when (1) velocities are at least 0.8 m/s (2.5 ft/s), (2) velocities do not exceed 2.4 m/s (8 ft/s), (3) thixotropic behavior is not considered, and (4) the pipe is not obstructed by grease or other materials.

Another approximate method makes use of empirical multiplication factor charts [see Fig. 13–5(b)]. The approximate method involves only velocity and percent solids consideration. Usually, the consistency of untreated primary sludge changes during pumping. At first, the most concentrated sludge is pumped. When most of the sludge has been pumped, the pump must handle a dilute sludge that has essentially the same hydraulic characteristics as water. The change in characteristics causes a centrifugal pump to operate farther out

Figure 13–5

Headloss multiplication factors: (a) for different sludge types and concentrations (b) for different pipeline velocities and sludge concentrations.



Note: Multiply loss with clean water by *k* to estimate friction loss under laminar conditions (see text).

on its head-capacity curve, beyond the areas of best efficiency. The pump motor should be sized for the additional load, and a variable-speed drive should be considered to reduce the flow under changing sludge characteristics. If the pump motor is not sized for the maximum load when pumping water at top speed, it is likely to be overloaded or damaged if the overload devices do not function or are set too high.

To determine the operating speeds and motor power required for a centrifugal pump handling sludge, system curves should be computed (1) for the most dense sludge anticipated with design friction factor, (2) for average conditions, and (3) for water with a new pipe friction factor to cover the full anticipated range of the pumping system. The curves should be plotted on a graph of the pump curves for a range of available speeds. The maximum and minimum speeds required of a particular pump are obtained from the intersection of the pump head-capacity curves with the system curves at the desired capacity. Where the maximum speed head-capacity curve intersects the system curve for water determines the power required. In constructing the system curves for sludge for velocities from 0 to 1.1 m/s (3.5 ft/s), the headloss can be considered constant at the figure computed for 1.1 m/s (3.5 ft/s). The intersection of the pump curves with the system curve for average conditions can be used to estimate hours of operations, average speed, and power costs.

Because the usual flow formulas cannot be used in the plastic and laminar range, judgment and experience must be relied upon. In this range, capacities will be small, and plunger, progressive cavity, or rotary-lobe pumps should be used with ample head and capacity as recommended previously.

Application of Rheology to Headloss Computations. For pumping sludge over long distances, an alternative method of computing headloss characteristics has been developed based on the flow properties of the sludge. A method of computing headloss for laminar flow conditions was derived originally by Babbitt and Caldwell (1939), based on the results of experimental and theoretical studies. Additional studies have been performed for the transition from laminar to turbulent flow (Mulbarger et al., 1981;

U.S. EPA, 1979) and are summarized in Sanks et al. (1998). Long-distance pumping of mixtures of untreated (raw) primary and secondary sludge is discussed by Carthew et al. (1983). The approach used in those studies for turbulent flow, which is of critical importance for long pipelines, is described below. For laminar and transitional flow, computational procedures described in Sanks et al. (1998) are recommended.

As stated previously, water, oil, and most other common fluids are "Newtonian," which means the pressure drop is directly proportional to the velocity and viscosity under laminar-flow conditions. As the velocity increases past a critical value, the flow becomes turbulent. The transition from laminar to turbulent flow depends on the Reynolds number, which is inversely proportional to the fluid viscosity. Wastewater sludge, however, is a non-Newtonian fluid. The pressure drop under laminar conditions is not proportional to flow, so the viscosity is not a constant. The precise Reynolds number at which turbulent-flow characteristics are encountered is uncertain for sludges.

Sludge has been found to behave much like a Bingham plastic, a substance with a straight-line relationship between shear stress and flow only after flow begins. A Bingham plastic is described by two constants: the yield stress s_y and the coefficient of rigidity η. Typical ranges of values for yield stress and coefficient of rigidity are shown on Figs. 13–6(a) and (b). If the two constants can be determined, the pressure drop over a wide range of velocities can be obtained using ordinary equations for water and the use of Fig. 13–6(c). As observed on Figs. 13–6(a) and (b), published data quantifying yield stress and the coefficient of rigidity values for wastewater sludges are highly variable. Pilot studies should be conducted to determine the rheological data for specific applications. Procedures for developing yield stress and the coefficient of rigidity using a pipeline viscometer and rotational viscometer are also given by Carthew et al. (1983).

Two dimensionless numbers can be used to determine the pressure drop due to friction for sludge: Reynolds number and Hedstrom number. Reynolds number is calculated by using the following expression:

$$N_R = \frac{\rho v D}{\eta} \qquad \text{SI units} \tag{13–3a}$$

$$N_R = \frac{\gamma v D}{\eta} \qquad \text{U.S. customary units} \tag{13–3b}$$

where N_R = Reynolds number, dimensionless

ρ = density of sludge, kg/m^3

γ = specific weight of sludge, lb/ft^3

v = average velocity, m/s (f/s)

D = diameter of pipe, m (ft)

η = coefficient of rigidity, kg/m·s (lb/ft·s)

Hedstrom number, which is reviewed by Hill et al. (1986), is calculated as follows:

$$H_e = \frac{D^2 s_y \rho}{\eta^2} \qquad \text{SI units} \tag{13–4a}$$

$$H_e = \frac{D^2 s_y g_c \gamma}{\eta^2} \qquad \text{U.S. customary units} \tag{13–4b}$$

where H_e = Hedstrom number, dimensionless

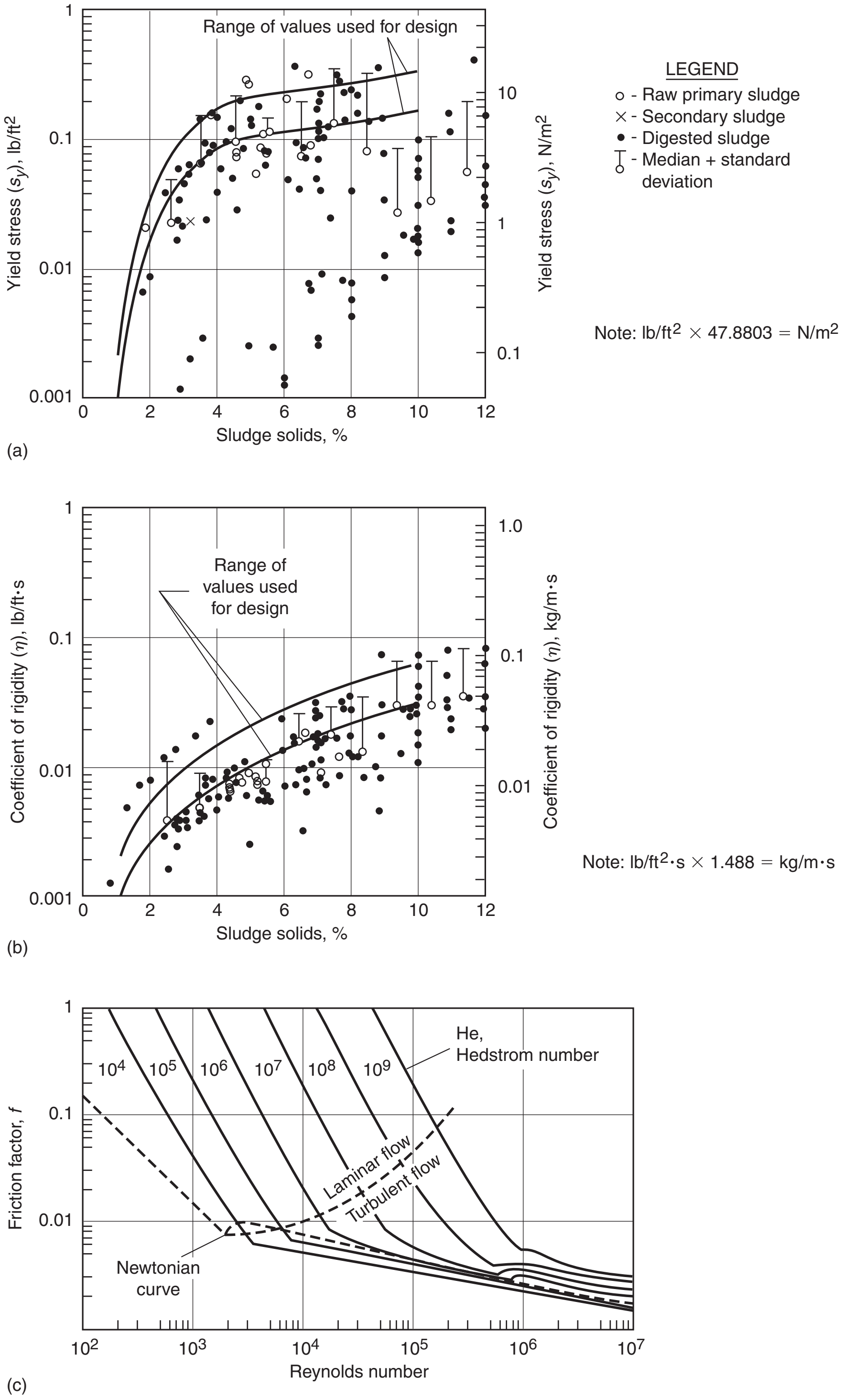
s_y = yield stress, N/m^2 (lb$_f$/ft^2)

g_c = 32.2 lb$_m$·ft/lb$_f$·s^2

Other terms are as defined previously.

Figure 13–6

Curves for computing pipeline headloss by the sludge rheology method: (a) yield stress vs. percent sludge solids, (b) coefficient of rigidity vs. percent sludge solids, and (c) friction factor for sludge analyzed as a Bingham plastic. (Adapted from Carthew et al., 1983.)



Note: lb/ft^2 × 47.8803 = N/m^2

Note: lb/ft^2·s × 1.488 = kg/m·s

Using the calculated Reynolds number and the Hedstrom number, the friction factor f can be determined from Fig. 13–6(c). The pressure drop for turbulent conditions can then be calculated from the following relationship:

$$\Delta p = \frac{2f\rho L v^2}{D} \qquad \text{SI units} \tag{13–5a}$$

$$\Delta p = \frac{2f\gamma L v^2}{g_c D} \qquad \text{U.S. customary units} \tag{13–5b}$$

where Δp = pressure drop due to friction, N/m^2 (lb$_f$/ft^2)
f = friction factor [from Fig. 13–6(c)]
L = length of pipeline, m (ft)
Other terms are as defined previously.

In using Eqs. (13–3), (13–4), and (13–5), it should be noted that the Reynolds number is not the same as the Reynolds number based on viscosity. In plastic flow, an effective viscosity may be defined, but it is variable and can be much greater than the coefficient of rigidity. Consequently, the two Reynolds numbers can differ greatly. The friction factor *f* will usually differ significantly from the *f* values reported in standard hydraulic texts for clear water, which may be four times the values used on Fig. 13–6(c). These equations apply to the entire range of laminar and turbulent flows, except that Fig. 13–6(c) does not allow for pipe roughness. To allow for pipe roughness, if customary water formulas for headloss result in a higher pressure drop than computed with Eq. (13–5), then roughness is dominant, the flow is fully turbulent, and the pressure drop given by the water headloss formula will be reasonably accurate. A safety factor on the order of 1.5 is recommended for worst-case design conditions (Mulbarger et al., 1981). The use of Eqs. (13–3), (13–4), and (13–5) is illustrated in Example 13–2.

EXAMPLE 13–2 **Computation of Headloss Using Sludge Rheology** Calculate the headloss in a 250-mm-diameter pipeline 10,000 m long conveying untreated (raw) sludge at an average flowrate 0.04 m^3/s. Determine also if the flow is turbulent. By testing, the following sludge rheology data were found:

Yield stress s_y = 1.3 N/m^2

Coefficient of rigidity η = 0.035 kg/m·s

Specific gravity = 1.01

Solution

1. Calculate the pipeflow velocity.
   a. Determine the pipe cross-sectional area.

$$A = \pi \times \frac{D^2}{4} = 3.14\frac{(0.25\text{ m})^2}{4} = 0.49\text{ m}^2$$

   b. Determine velocity.

$$v = \frac{Q}{A} = \frac{(0.04\text{ m}^3/\text{s})}{0.049\text{ m}^2} = 0.82\text{ m/s}$$

2. Compute sludge specific weight.

$$\rho = 1000\text{ kg/m}^3 \times 1.01 = 1010\text{ kg/m}^3$$

3. Compute Reynolds number using Eq. (13–3).

$$N_R = \frac{\rho v D}{\eta} = \frac{(1010\ \text{kg/m}^3)(0.82\ \text{m/s})(0.25\ \text{m})}{(0.035\ \text{kg/m·s})} = 5.92 \times 10^3$$

4. Compute Hedstrom number using Eq. (13–4).

$$H_e = \frac{D s_y \rho}{\eta^2} = \frac{(0.25\ \text{m})^2(1.3\ \text{N/m}^2)(1010\ \text{kg/m}^3)}{(0.035\ \text{kg/m·s})^2} = 6.70 \times 10^4$$

5. Determine friction factor f from Fig. 13–6(c) using the computed Reynolds and Hedstrom numbers.

$f = 0.007$

Note, on Fig. 13–6c, that the flow is in the turbulent zone.

6. Compute pressure drop using Eq. (13–5).

$$\Delta p = \frac{2 f \rho L v^2}{D} = \frac{2(0.007)(1010\ \text{kg/m}^3)(10{,}000\ \text{m})(0.82\ \text{m/s})^2}{0.25\ \text{m}}$$

$$= 380{,}309\ \text{kg/m·s}^2\ (\text{N/m}^2 \text{ or Pa})$$

Convert to meters of water.

$$\Delta p = \frac{380{,}309\ \text{kg/m·s}^2}{(10^3\ \text{kg/m}^3)(9.81\ \text{m/s}^2)} = 38.8\ \text{m}$$

Comment In this example, only one set of rheology data was used. In actual design, test data should be used for a range of probable conditions so that a family of headloss curves can be developed for the range of operating conditions. In addition, appropriate safety factors should be used for worst-case conditions. Comparison of the headloss to the headloss for water using the Hazen-Williams formula is left as a homework problem.

Sludge Piping

In wastewater treatment plants, conventional sludge piping should not be smaller than 150 mm (6 in.) in diameter although smaller-diameter glass-lined pipe has been used successfully. Sludge piping may not need to be larger than 200 mm (8 in.), unless the velocity exceeds 1.5 to 1.8 m/s (5 to 6 ft/s), in which case the pipe is sized to maintain that velocity. Gravity sludge withdrawal lines should not be less than 200 mm (8 in.) in diameter. It is common practice to install a number of cleanouts in the form of plugged tees or crosses instead of elbows so that the lines can be rodded if necessary. Pump connections should not be smaller than 100 mm (4 in.) in diameter.

A liberal number of hose gates should be installed in the piping, and an ample supply of high-pressure flushing water should be available for clearing stoppages. The flushing water should be plant effluent. The flushing water system should have a capacity of not less than 0.010 m^3/s (150 gal/min) at 500 kN/m^2 ($\sim$70 $\text{lb}_\text{f}/\text{in.}^2$). In large plants with larger piping, a greater capacity should be available, and the available pressure should be increased to 700 kN/m^2 (100 $\text{lb}_\text{f}/\text{in.}^2$).

Grease has a tendency to coat the inside of piping used for transporting primary sludge and scum. Grease accumulation is more of a problem in large plants than in small ones. The coating results in a decrease in effective diameter and a large increase in pumping head. For this reason, low capacity positive-displacement pumps are designed for heads greatly in excess of the theoretical head. Centrifugal pumps, with their larger capacity, usually pump a more dilute sludge, often containing some wastewater, and head buildup due to grease accumulations appears to occur more slowly. In some plants, provisions have been made for melting the grease by circulating hot water, steam, or digester supernatant through the main sludge lines.

In treatment plants, friction losses are low because the pipe runs are short; consequently, there is little difficulty in providing an ample safety factor. In the design of long sludge lines, however, special design features should be considered including (1) providing two pipes unless a single pipe can be shut down for several days without causing problems; (2) providing for external corrosion and pipe loads; (3) adding facilities for applying dilution water for flushing the line; (4) providing means to insert a pipe cleaner; (5) including provisions for steam injection, especially in cold climates and where excessive grease accumulation occurs; (6) providing air relief and blowoff valves for the high and low points, respectively, and (7) considering the potential effects of waterhammer. A discussion of waterhammer in force mains is provided in the companion volume to this text (Metcalf & Eddy, 1981).

13–5 PRELIMINARY SLUDGE PROCESSING OPERATIONS

Grinding, degritting, blending, and storage of sludge is necessary to provide a relatively constant, homogeneous feed to subsequent processing facilities. Blending and storage can be accomplished either in a single unit designed to do both or separately in other plant components. Screening of raw sludge or digested biosolids is sometimes required in reuse applications for the removal of plastics, rags, and other material. Each of these preliminary operations is discussed in this section.

Grinding

Sludge grinding is a process in which large and stringy material contained in sludge is cut or sheared into small particles to prevent clogging or wrapping around rotating equipment. A typical sludge grinder installation is shown on Fig. 13–7. Some of the processes that must be preceded by sludge grinders and the purposes of grinding are reported in Table 13–16. Grinders historically have required high maintenance, but newer designs of slow-speed grinders have been more durable and reliable. These designs include improved bearings and seals, hardened steel cutters, overload sensors, and mechanisms that reverse the cutter rotation to clear obstructions or shut down the unit if the obstruction cannot be cleared.

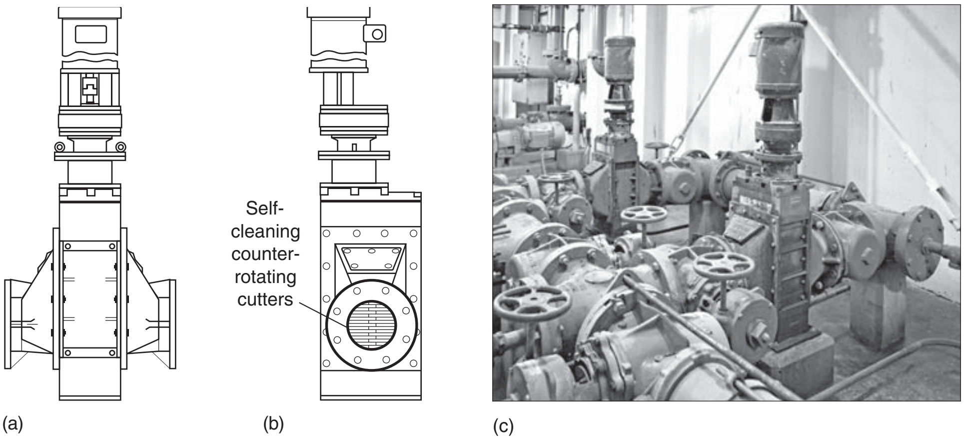


Figure 13–7

Typical inline sludge grinder: (a) side view, (b) end view, and (c) view of typical installation [(a) and (b) adapted from Franklin Miller].

Table 13–16

Operations or processes requiring the grinding of sludge and biosolids

Operation or process	Purpose of grinding
Pumping with progressive cavity pumps	Prevent clogging and reduce wear
Solid bowl centrifuges	Prevent clogging. Large solid bowl units generally can handle larger particles and may not require sludge or biosolids grinding
Belt filter press	Prevent clogging of the sludge or biosolids distribution system, prevent warping of rollers, and provide more uniform dewatering

Screening

Because raw wastewater screens can allow significant quantities of solid material to pass through, sludge screening is an alternative to grinding. Screening is advantageous in that nuisance material is removed from the sludge stream. Step screens, shown on Fig. 5–4(c) in Chap. 5, can be used for the removal of fine solids from septage and primary sludge. Screen openings normally range from 3 to 6 mm (0.12 to 0.24 in.), although openings up to 10 mm (0.4 in.) can be used.

Another type of sludge screen is an inline screen that can be installed in a pipeline (see Fig. 13–8). The screen removes material by passing the flow stream through a screen with 3 to 10-mm (0.12 to 0.4-in.) openings although 5 mm (0.2 in.) is the typical size for wastewater sludges. Material captured by the screen moves by a screw conveyor into a press or compaction zone where it is dewatered and compacted. Material is ejected from the press zone when sufficient solids build up to overcome the force on the unit's discharge cone. Screening solids concentrations range from 30 to 50 percent. Allowable operating pressure is reported to be 100 kPa (14 $lb_f/in.^2$) (Arakaki et al., 1998). The screened sludge is diluted and may require thickening.

Degritting

In some plants where separate grit removal facilities are not used ahead of the primary sedimentation tanks, or where the grit removal facilities are not adequate to handle peak

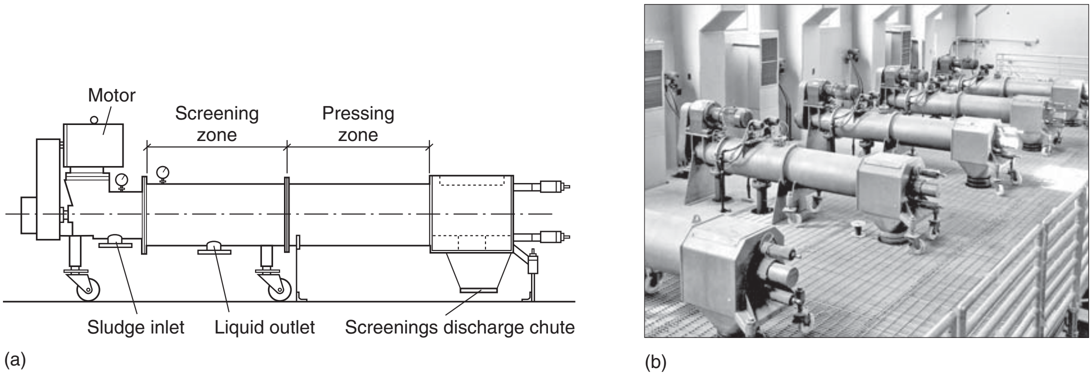


Figure 13–8

Sludge screenings press: (a) schematic and (b) view of a large installation (courtesy of the City of San Diego, CA).

Table 13–17

Grit removal efficiency using cyclone degritters for primary sludge[a]

Primary sludge concentration, % total solids	Mesh of removal[b]
1	150
2	100
3	65
4	28–35

[a] For a 300 mm (12 in.) hydrocyclone at 42 kN/m^2 (6 lb$_f$/in.^{2} gage) at 13 L/s (200 gal/min).

[b] About 95 percent or more of indicated particle size is removed.

Note: Normal design range is for 1 to 1.5 percent sludge.

flows and peak grit loads, it may be necessary to remove the grit before further processing of the sludge. Where further thickening of the primary sludge is desired, a practical consideration is sludge degritting. The most effective method of degritting sludge is through the application of centrifugal forces in a flowing system to achieve separation of the grit particles from the organic sludge. Such separation is achieved through the use of cyclone degritters, which have no moving parts. The sludge is applied tangential to a cylindrical feed section, thus imparting a centrifugal force. The heavier grit particles move to the outside of the cylinder section and are discharged through a conical feed section. The organic sludge is discharged through a separate outlet. The efficiency of the cyclone degritter is affected by pressure and by the concentration of the organics in the sludge. To obtain effective grit separation, the sludge must be relatively dilute, 1 to 2 percent TS. As the sludge concentration increases, the particle size that can be removed decreases. The general relationship between sludge concentration and effectiveness of removal for primary sludges is shown in Table 13–17.

Blending

Sludge is generated in primary, secondary, and advanced wastewater treatment processes. Primary sludge consists of settleable solids carried in the raw wastewater. Secondary sludge consists of biological solids as well as additional settleable solids. Sludge produced in the advanced wastewater may consist of biological and chemical solids. Sludge is blended to produce a uniform mixture to downstream operations and processes. Uniform mixtures are most important in short-detention-time systems, such as sludge dewatering, heat treatment, and incineration. Provision of well-blended sludge with consistent characteristics to these treatment units will enhance greatly plant operability and performance.

Sludge from primary, secondary, and advanced processes can be blended in several ways:

1. *In primary settling tanks*. Secondary or tertiary sludges can be returned to the primary settling tanks where they will mix and co-settle with the primary sludge.
2. *In pipes*. Blending in pipes requires careful control of sludge sources and feed rates to ensure the proper blend. Without careful control, wide variations in sludge consistency may be expected.
3. *In sludge-processing facilities with long detention times*. Aerobic and anaerobic digesters (complete-mix type) can blend the feed sludges uniformly.
4. *In a separate blending tank*. This practice provides the best opportunity to control the quality of the blended sludges.

In treatment plants of less than 0.05 m^3/s (1 Mgal/d) capacity, blending is accomplished usually in the primary settling tanks. In large facilities, optimum efficiency is achieved by separately thickening sludges before blending.

Storage

Storage should be provided to minimize fluctuations in the rate of sludge and biosolids production and to allow sludge to accumulate during periods when subsequent processing facilities are not operating, e.g., night shifts, weekends, and periods of unscheduled equipment downtime. Sludge and biosolids storage is particularly important in providing a uniform feed rate ahead of the following processes: mechanical dewatering, lime stabilization, heat drying, and thermal reduction.

Short-term sludge and biosolids storage may be accomplished in wastewater settling tanks or in thickening tanks. Long-term sludge and biosolids storage may be accomplished in stabilization processes with long detention times, e.g., aerobic and anaerobic digestion, or in specially designed separate tanks. In small installations, sludge is usually stored in the settling tanks and digesters. In large installations that do not use aerobic and anaerobic digestion, sludge is often stored in separate blending and storage tanks. Such tanks may be sized to retain the sludge for a period of several hours to a few days. If sludge or biosolids is stored longer than 2 to 3 d, it will deteriorate, become odorous, and be more difficult to dewater. The determination of the required storage volume is illustrated in Example 13–3. Sludge or biosolids is often aerated to prevent septicity and to promote mixing. Mechanical mixing may be necessary to assure complete blending of the sludge. Chlorine, iron salts, potassium permanganate, and hydrogen peroxide have been used with limited success to limit or control septicity and to control the odors from sludge storage and blending tanks. In cases where sludge storage occurs in enclosed tanks, ventilation should be provided along with appropriate odor-control technologies such as chemical scrubbers or biofilters (see Chap. 15).

EXAMPLE 13–3 **Determination of Volume Required for Sludge Storage** Assume that the yearly average rate of sludge production from an activated sludge treatment plant is 12,000 kg/d. Develop a curve of sustained sludge mass loading rates that can be used to determine the size of sludge-storage facilities required with various downstream sludge-processing units. Then, using the developed curve, determine the volume required for sludge storage, assuming that sludge accumulated for 7 d is to be processed in 5 working d, and that sludge accumulated for 14 d is to be processed in 10 working d. Note that the 5- and 10-d work periods correspond to 1 and 2 wk, respectively, assuming that certain sludge-processing facilities, such as belt-filter presses, will not be operated on the weekends.

Solution

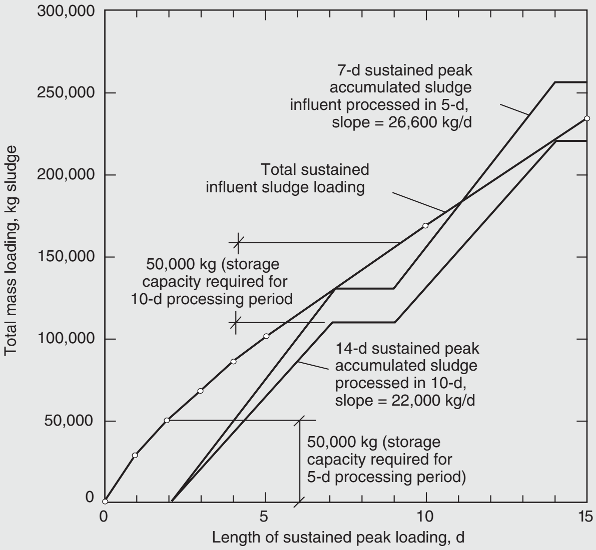
1. Develop a curve of sustained sludge mass loadings.
    a. Because no information is specified, it will be assumed that the sustained sludge production will mirror the sustained BOD plant loadings given on Fig. 3–13(a) and used in Example 3–7.
    b. Set up an appropriate computation table and compute the values necessary to plot the curve.

Length of sustained peak, d (1)	Peaking factor[a] (2)	Peak sludge mass loading, kg/d (3)	Total sustained loading, kg[b] (4)
1	2.4	28,800	28,800
2	2.1	25,200	50,400
3	1.9	22,800	68,400
4	1.8	21,600	86,400
5	1.7	20,400	102,000
10	1.4	16,800	168,000
15	1.3	15,600	234,000
365	1.0	12,000	

[a] From Fig. 3–13(a).

[b] Total mass produced for the corresponding sustained period given in col. 1.

c. Plot the sustained sludge loading curve (see following figure).



2. Determine the sludge storage volume required for the stated operating conditions.
   a. Determine the daily rate at which sludge must be processed to handle the 7-d sustained peak (from figure) in 5 working d.

$$\text{kg/d} = \frac{133{,}000}{5\,\text{d}} = 26{,}600\,\text{kg/d}$$

   b. Determine the daily rate at which sludge must be processed to handle the 14-d sustained peak (from figure) in 10 working d.

$$\text{kg/d} = \frac{220{,}000}{10\,\text{d}} = 22{,}000\,\text{kg/d}$$

c. Assuming that the sludge storage facilities are empty on Friday just before the weekend, plot on the figure the average daily rate at which sludge must be processed during the 5- and 10-d periods.

d. From the figure, the required storage capacity in pounds of sludge is
   i. Capacity based on 5 working d = 50,000 kg
   ii. Capacity based on 10 working d = 50,000 kg

Comment The downstream processing equipment can now be sized using the daily rate at which sludge must be processed. For example, if the number of kilograms per hour that can be processed with a belt-filter press is known, then the size and number of units can be computed from the number of shifts to be used per day and the assumed value of the actual working hours per shift. In sizing equipment, a trade-off analysis should always be performed between the cost of storage and processing facilities versus labor costs (for both one shift and two shifts) to determine the most cost-effective combination.

13–6 THICKENING

The solids content of primary, activated, trickling filter, or mixed sludge (i.e., primary plus waste activated) varies considerably, depending on the characteristics of the sludge, the sludge removal and pumping facilities, and the method of operation. Representative values of percent total solids from various treatment operations or processes were shown previously in Table 13–8. Thickening is a procedure used to increase the solids content of sludge by removing a portion of the liquid fraction. To illustrate, if waste activated sludge which is typically pumped from secondary settling tanks with a content of 0.8 percent, can be thickened to a content of 4 percent solids, then a five-fold decrease in sludge volume is achieved. Thickening is generally accomplished by physical means, including co-settling, gravity settling, flotation, centrifugation, gravity belt, and rotary drum. Typical sludge-thickening methods are described in Table 13–18.

Application

The volume reduction obtained by sludge concentration is beneficial to subsequent treatment processes, such as digestion, dewatering, and drying from the following standpoints: (1) capacity of tanks and equipment required, (2) quantity of chemicals required for sludge conditioning, and (3) amount of heat required by digesters and amount of auxiliary fuel required for heat drying.

For large facilities where sludge must be transported a significant distance, such as to a separate plant for processing, a reduction in sludge volume may result in a reduction of pipe size and pumping costs. For smaller facilities, the requirements of a minimum practicable pipe size and minimum velocity may necessitate pumping of significant volumes of wastewater in addition to sludge, thereby diminishing the value of volume reduction. Volume reduction is very desirable when liquid sludge is transported by tank trucks for direct application to land as a soil conditioner.

Sludge thickening is achieved at all wastewater treatment plants in some manner—in the primary clarifiers, in sludge digestion facilities, or in specially designed separate units. If separate units are used, the recycled flows are returned normally to the wastewater treatment facilities. In treatment plants of less than 4000 m^3/d (~1 Mgal/d) capacity, separate sludge thickening is seldom practiced. In small plants, gravity thickening is accomplished in the primary settling tank or in the sludge-digestion units, or both. In larger treatment

Table 13–18
Occurrence of thickening methods in sludge processing

Method	Type of sludge	Frequency of use and relative success
Gravity, co-settling in clarifier	Primary and waste activated	Occasional use; may negatively impact the effectiveness of the primary clarifier
Gravity, thickening in separate tank	Untreated primary	Commonly used with excellent results. Sometimes used with hydrocyclone degritting of sludge
	Untreated primary and waste activated	Often used. For small plants, generally satisfactory results with sludge concentrations in the range of 4 to 6 percent. For large plants, results are marginal. Can be odorous in warm weather
	Waste activated	Seldom used; poor solids concentration (2 to 3 percent)
Dissolved air flotation	Untreated primary and waste activated	Limited use; results similar to gravity thickeners
	Waste activated	Commonly used, but use is decreasing because of high operating cost; good results (3.5 to 5 percent solids concentration)
Solid bowl centrifuge	Waste activated	Often used in medium to large plants; good results (4 to 6 percent solids concentration)
Gravity belt thickener	Waste activated	Often used; good results (3 to 6 percent solids concentration)
Rotary drum thickener	Waste activated	Limited use; good results (5 to 9 percent solids concentration)

facilities, the additional costs of separate sludge thickening are often justified by the improved control over the thickening process and the higher concentrations attainable.

Description and Design of Thickeners

The following discussion is intended to introduce the reader to the operations used for the thickening of sludges. Because most of the equipment is mechanical, the primary concern is with its proper application to meet a given treatment objective rather than with the theory of mechanical design. In designing thickening facilities, it is important to (1) provide adequate capacity to meet peak demands and (2) prevent septicity, with its attendant odor problems, during the thickening process. The six methods of thickening discussed in this section are (1) co-settling thickening, (2) gravity, (3) dissolved air flotation, (4) centrifugal, (5) gravity belt, and (6) rotary drum.

Co-settling Thickening. Primary clarifiers are often used to thicken sludge for downstream processing. To thicken sludge, a sludge blanket must be created to consolidate the sludge without allowing the clarified water to be pulled through. Often, sludge retention times of 12 to 24 h or more are maintained in clarifiers to achieve thickened sludge concentration levels in the clarifier underflow. Excessive retention of sludge in the clarifier can cause septic conditions and gasification, and reduce the levels of TSS and BOD removal. Typical effects of sludge blanket retention on TSS removal are illustrated on Fig. 13–9.

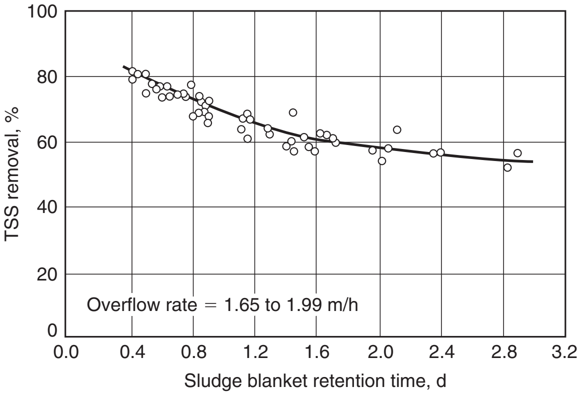


Figure 13–9

Effect of sludge blanket retention time on TSS removal for co-thickening of primary sludge (Albertson and Walz, 1997).

Successful thickening of sludge in primary clarifiers has been achieved by a combination of the following: (1) using one clarifier in a bank of clarifiers for co-settling thickening; dilute sludge underflow (less than 1 percent solids) from the other clarifiers is discharged to the thickening clarifier, (2) maintaining the sludge inventory for about 6 to 12 h, and (3) providing for the addition of coagulating chemicals such as polymer and ferric chloride to condition the sludge to enhance settling. The need for chemical addition depends upon the clarifier overflow rates. Underflow sludge concentrations on the order of 3 to over 5 percent have been reported (Albertson and Walz, 1997). By controlling the sludge blanket within the above sludge retention parameters, clarifier removal rates are enhanced, and sludge thickening is achieved. A schematic diagram of the co-settling thickening system is shown on Fig. 13–10.

Gravity Thickening. Gravity thickening is one of the most common methods used and is accomplished in a tank similar in design to a conventional sedimentation tank. Normally, a circular tank is used, and dilute sludge is fed to a center feed well. The feed sludge is allowed to settle and compact, and the thickened sludge is withdrawn from the conical tank bottom. Conventional sludge collecting mechanisms with deep trusses (see Fig. 13–11) or vertical pickets stir the sludge gently, thereby opening up channels for water to escape and promoting densification. The supernatant flow that results is drawn off and returned to either the primary settling tank, the influent of the treatment plant, or a return flow treatment process. The thickened sludge is pumped to the digesters or dewatering equipment as required; thus, storage space must be provided for the sludge. As indicated in Table 13–18,

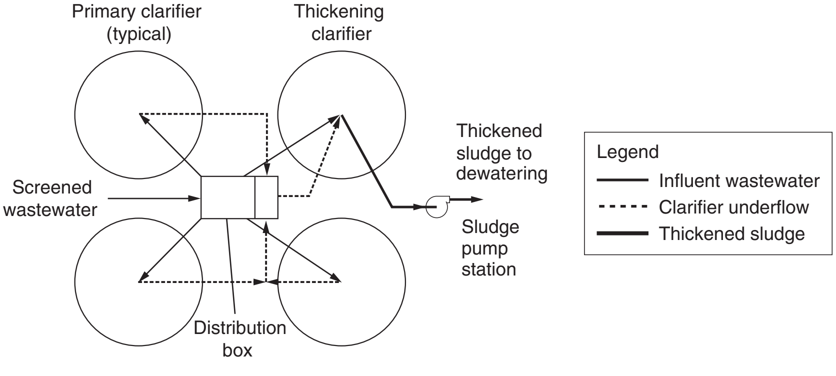
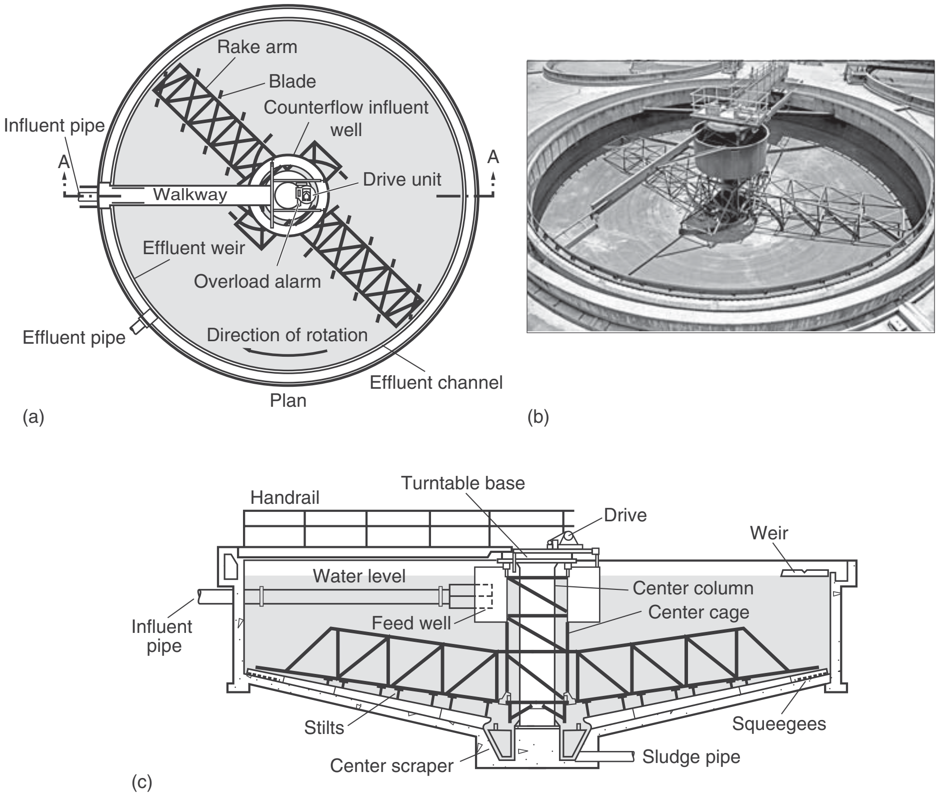


Figure 13–10

Schematic diagram of a sludge co-thickening system.

Figure 13–11

Schematic diagram of a gravity thickener: (a) plan and (b) section.



gravity thickening is most effective on primary sludge. Gravity thickeners are designed on the basis of solids loading and thickener overflow rate. Typical solids loadings based on existing data are reported in Table 13–19. Recommended maximum hydraulic overflow rates range from 15.5 to 31 $m^3/m^2{\cdot}d$ (380 to 760 $gal/ft^2{\cdot}d$) for primary sludges, 4 to 8 $m^3/m^2{\cdot}d$ (100 to 200 $gal/ft^2{\cdot}d$) for waste activated sludge, and 6 to 12 $m^3/m^2{\cdot}d$ (150 to 300 $gal/ft^2{\cdot}d$) for combined primary and waste activated sludge (WEF, 1980). High hydraulic loadings can cause excessive solids carryover. Conversely, low hydraulic loadings can cause septic conditions and odors, and floating sludge can result.

Provisions for dilution water and occasional chlorine addition are frequently included to improve process performance by maintaining the hydraulic loading. Polymer addition is frequently provided. To maintain aerobic conditions in gravity thickeners, especially when wastewater is warm (22 to 28°C), provisions should be included for adding up to 24 to 30 $m^3/m^2{\cdot}d$ (600 to 750 $gal/ft^2{\cdot}d$) of dilution water (final effluent) to the thickening tank. The dilution water may also remove certain soluble organic and inorganic compounds that consume large amounts of conditioning chemicals used in dewatering. Dilution water that is part of supernatant returned and recycled to the liquid process must be considered in process design.

Because the thickening characteristics of wastewater sludge can vary considerably, it is desirable to design a thickening facility using criteria based on a testing program. Testing programs that can be used include batch settling tests, bench-scale settling tests, and pilot-scale testing. The latter method is recommended wherever possible because data can be obtained from a variety of operating parameters. Test methods are described in WEF (2010a).

Table 13–19

Typical concentrations of unthickened and thickened sludges and solids loadings for gravity thickeners[a]

Type of sludge or biosolids	Solids concentration, %		Solids loading	
	Unthickened	Thickened	lb/ft²·d	kg/m²·d
Separate				
Primary sludge	1–6	3–10	20–30	100–150
Trickling filter humus sludge	1–4	3–6	8–10	40–50
Rotating biological contactor	1–3.5	2–5	7–10	35–50
Air–activated sludge	0.5–1.5	2–3	4–8	20–40
High–purity oxygen activated sludge	0.5–1.5	2–3	4–8	20–40
Extended aeration activated sludge	0.2–1.0	2–3	5–8	25–40
Anaerobically digested primary sludge from primary digester	8	12	25	120
Combined				
Primary and trickling filter humus sludge	1–6	3–9	12–20	60–100
Primary and rotating biological contactor	1–6	3–8	10–18	50–90
Primary and waste activated sludge	0.5–1.5	2–6	5–14	25–70
	2.5–4.0	4–7	8–16	40–80
Waste activated sludge and trickling filter humus sludge	0.5–2.5	2–4	4–8	20–40
Anaerobically digested primary and waste activated sludge	4	8	14	70
Chemical sludge:				
High lime	3–4.5	12–15	24–60	120–300
Low lime	3–4.5	10–12	10–30	50–150
Iron	0.5–1.5	3–4	2–10	10–50

[a] Adapted from WEF (2010a).

In operation, a sludge blanket is maintained on the bottom of the thickener to aid in concentrating the sludge. An operating variable is the sludge volume ratio, which is the volume of the sludge blanket held in the thickener divided by the volume of the thickened sludge removed daily. Values of the sludge volume ratio normally range between 0.5 and 20 d; the lower values are required during warm weather. Alternatively, sludge blanket depth should be measured. Blanket depths may range from 0.5 to 2.5 m (2 to 8 ft); shallower depths are maintained in the warmer months.

EXAMPLE 13–4 **Design a Gravity Thickener for Combined Primary and Waste Activated Sludge** Design a gravity thickener for a wastewater treatment plant having primary and waste activated sludge with the following characteristics:

Type of sludge	Specific gravity	Solids, %	Flowrate, m^3/d
Average design conditions:			
Primary sludge	1.03	3.3	400
Waste activated	1.005	0.2	2250
Peak design conditions:			
Primary sludge	1.03	3.4	420
Waste activated	1.005	0.23	2500

Solution

1. Compute the dry solids at peak design conditions.
   a. Primary sludge

$$\text{kg/d dry solids} = (420\ \text{m}^3/\text{d})(1.03)(0.034\ \text{g/g})(10^3\ \text{kg/m}^3)$$
$$= 14{,}708\ \text{kg/d}$$

   b. Waste activated sludge

$$\text{kg/d dry solids} = (2500\ \text{m}^3/\text{d})(1.005)(0.0023\ \text{g/g})(10^3\ \text{kg/m}^3)$$
$$= 5779\ \text{kg/d}$$

   c. Combined sludge mass = 14,708 + 5779 = 20,487 kg/d
   d. Combined sludge flowrate = 2,500 + 420 = 2,920 m^3/d

2 Compute solids concentration of the combined sludge, assuming the specific gravity of the combined sludge is 1.02.

$$\%\ \text{solids} = \frac{(20{,}487\ \text{kg/d})}{(2920\ \text{m}^3/\text{d})(1.02)(10^3\ \text{kg/m}^3)} \times 100 = 0.69\%$$

3. Compute surface area based on solids loading rate. Because the sludge concentration is between 0.5 and 1.5%, select a solids loading rate of 50 $kg/m^2 \cdot d$ from Table 13–19.

$$\text{Area} = \frac{(20{,}487\ \text{kg/d})}{(50\ \text{kg/m}^2 \cdot \text{d})} = 409.7\ \text{m}^2$$

4. Compute hydraulic loading rate.

$$\text{Hydraulic loading} = \frac{(2920\ \text{m}^3/\text{d})}{409.7\ \text{m}^2} = 7.13\ \text{m}^3/\text{m}^2 \cdot \text{d}$$

5. Compute diameter of thickener; assume two thickeners.

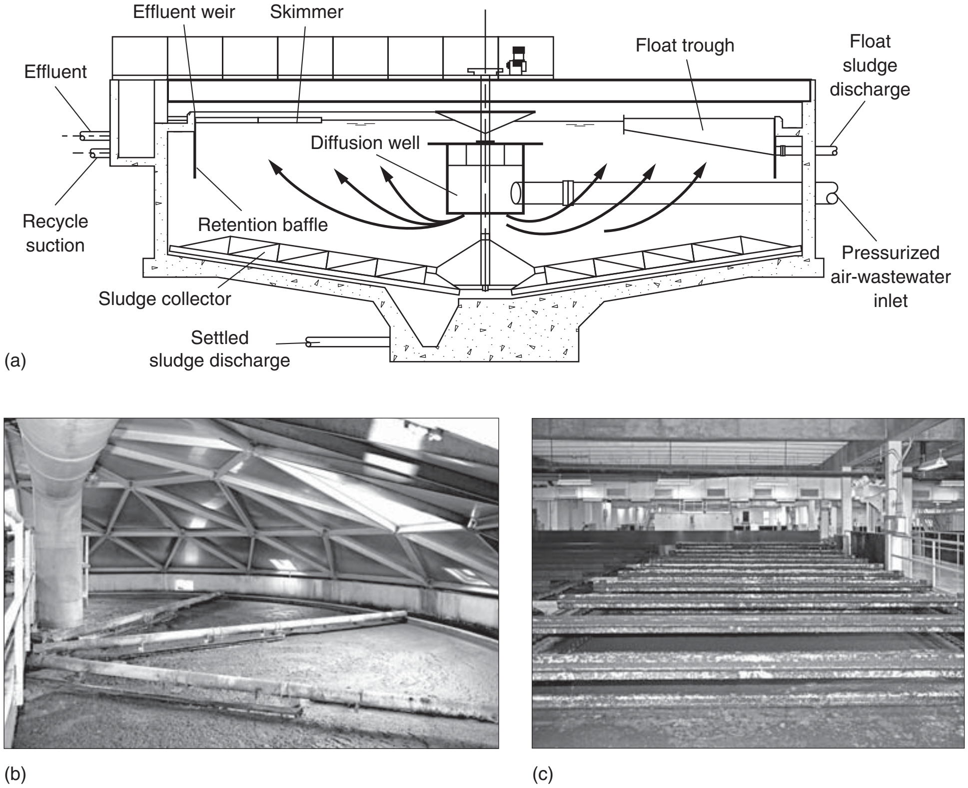
$$\text{Diameter} = \sqrt{\frac{4 \times 409.7\ \text{m}^2}{2 \times \pi}} = 16.15\ \text{m}$$

Comment The hydraulic loading rate of 7.13 $m^3/m^2 \cdot d$ at peak design flow is at the lower end of the recommended rate. To prevent septicity and odors, dilution water should be provided. Calculation of the dilution water requirements for average design flow is a homework problem. The thickener size of 16.15 m is within the maximum size of 20 m customarily recommended by thickener equipment manufacturers for use in municipal wastewater treatment. In actual design, round the thickener diameter to the nearest 0.5 m, or, in this case, 16 m.

Flotation Thickening. In dissolved air flotation, air is introduced into a solution that is being held at an elevated pressure. A typical unit used for thickening waste activated sludge is shown on Fig. 13–12. When the solution is depressurized, the dissolved air is released as finely divided bubbles carrying the sludge to the top, where it is removed.

Figure 13–12

Typical dissolved air flotation unit used for thickening waste activated sludge: (a) cross-section through typical circular flotation unit, (b) view inside covered circular flotation unit, and (c) view inside building containing rectangular flotation units.



Flotation thickening is used most efficiently for waste sludges from suspended-growth biological treatment processes, such as the activated sludge process or the suspended growth nitrification process. Other sludges such as primary sludge, trickling filter humus, aerobically digested sludge, and sludges containing metal salts from chemical treatment have been flotation thickened. In locations where freezing is a problem or where odor control is of concern, flotation thickeners are normally enclosed in a building.

The float solids concentration that can be obtained by flotation thickening of waste activated sludge is influenced primarily by the air-to-solids ratio, sludge characteristics (in particular the sludge volume index, SVI), solids loading rate, and polymer application. Although float solids concentrations have ranged historically between 3 and 6 percent by weight, float solids concentration is difficult to predict during the design stage without bench-scale or pilot-plant testing. The air-to-solids ratio is probably the most important factor affecting performance of the flotation thickener, and is defined as the weight ratio of air available for flotation to the solids to be floated in the feed stream. The air-to-solids ratio at which float solids are maximized varies from 2 to 4 percent. The SVI is also important because better thickening performance has been reported when the SVI is less than 200, using nominal polymer dosages. At high SVIs, the float concentration deteriorates and high polymer dosages are required.

Higher loadings can be used with dissolved air flotation thickeners than are permissible with gravity thickeners, because of the rapid separation of solids from the wastewater. Flotation thickeners typically are designed for the solids loadings given in Table 13–20. For design without the benefit of pilot studies, the minimum loadings should be used. The higher solids loadings generally result in lower concentrations of thickened sludge.

Table 13–20

Typical solids loadings for dissolved air flotation units[a,b]

	Loading, lb/ft²·h		Loading, kg/m²·h	
Type of sludge	**Without chemical addition**	**With chemicals**	**Without chemical addition**	**With chemicals**
Air activated sludge:				
Mixed Liquor	0.25–0.6	Up to 2	1.2–3	Up to 10
Settled	0.5–0.8	Up to 2	2.4–4	Up to 10
High purity oxygen activated sludge	0.6–0.8	Up to 2	3–4	Up to 10
Trickling filter humus sludge	0.6–0.8	Up to 2	3–4	Up to 10
Primary + air activated sludge	0.6–0.8	Up to 2	3–6	Up to 10
Primary + trickling filter humus sludge	0.83–1.25	Up to 2	4–6	Up to 10
Primary sludge only	0.83–1.25	Up to 2.5	4–6	Up to 12.5

[a] Adapted, in part, from U.S. EPA (1979) and WEF (2010a).

[b] Loading rates necessary to produce a minimum 4 percent solids concentration in the float.

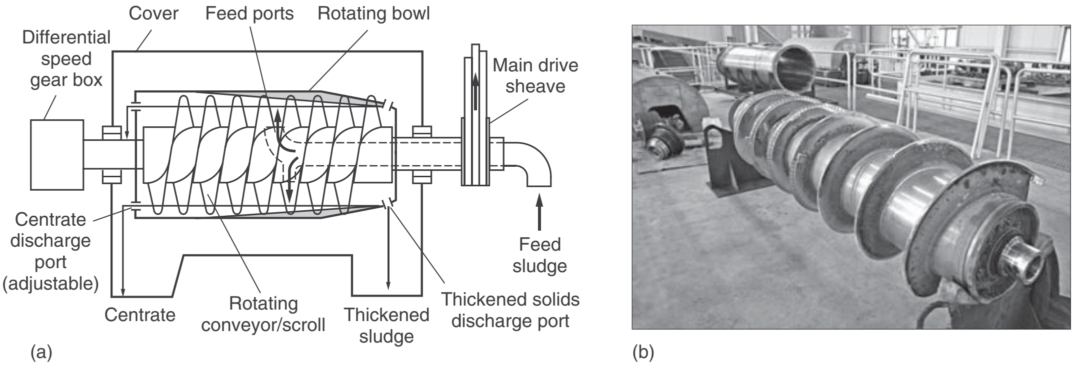
Operational difficulties may arise when the solids loading rate exceeds approximately 10 kg/m²·h (2.0 lb/ft²·h). The increased amount of float created at high solids loading necessitates continuous skimming, often at high skimming speeds.

Primary tank effluent or plant effluent is recommended as the source of air-charged water rather than flotation tank effluent, except when chemical aids are used, because of the possibility of fouling the air-pressure system with solids. The use of polymers as flotation aids is effective in increasing the solids recovery in the floated sludge from 85 to 98 or 99 percent, and in reducing the recycle loads. Polymer dosages for thickening waste activated sludge are 2 to 5 kg of dry polymer per tonne of dry solids (4 to 10 lb/ton).

Centrifugal Thickening. Centrifuges are used both to thicken and to dewater sludges. As indicated in Table 13–18, their application in thickening is limited normally to waste activated sludge. Thickening by centrifugation involves the settling of sludge particles under the influence of centrifugal forces. The basic type of centrifuge used for sludge thickening is the solid bowl centrifuge (see Fig. 13–13).

The solid-bowl centrifuge consists of a long bowl, normally mounted horizontally and tapered at one end. Sludge is introduced into the unit continuously, and the solids concentrate on the periphery. An internal helical scroll, spinning at a slightly different speed, moves the accumulated sludge toward the tapered end where additional solids concentration occurs and the thickened sludge is discharged.

Under normal conditions, thickening can be accomplished by centrifugal thickening without polymer addition. Maintenance and power costs for the centrifugal thickening process, however, can be substantial. Therefore, the process is usually attractive only at facilities larger than 0.2 m³/s (5 Mgal/d), where space is limited and skilled operators are available, or for sludges that are difficult to thicken by more conventional means. Many systems are designed with standby polymer systems for use to improve system performance. Polymer dosages for thickening waste activated sludge range from 0 to 4 kg of dry polymer per 1 tonne of dry solids (0 to 8 lb/ton).



(a)

(b)

Figure 13–13

Centrifuge used for sludge thickening: (a) schematic, and (b) scroll rotor removed for maintenance.

The performance of a centrifuge is often quantified by the concentration achieved in the thickened sludge product and the TSS recovery (sometimes termed "capture"). The recovery is calculated as the thickened dry solids as a percentage of the feed solids. Using the commonly measured solids concentrations, the recovery is calculated by the following expression (WEF, 2010a):

$$R = \frac{\text{TSS}_P(\text{TSS}_F - \text{TSS}_C)}{\text{TSS}_F(\text{TSS}_P - \text{TSS}_C)} \times 100 \tag{13–6}$$

where R = recovery, percent
TSS_P = total suspended solids concentration in thickened product, percent by weight
TSS_F = total suspended solids concentration in feed, percent by weight
TSS_C = total suspended solids concentration in centrate, percent by weight

For a constant feed concentration, the percent recovery increases as the concentration of solids in the centrate decreases. In concentrating sludge solids, recovery is important because with a higher recovery lesser amounts of biodegradable solids are returned to the treatment process for further treatment. In developing a mass balance for the treatment plant, return flows (also termed sidestream flows) from thickening, stabilization, and dewatering processes must be taken into account (see Sec. 14–7).

The principal operational variables include the following: (1) characteristics of the feed sludge (its water holding structure and the sludge volume index); (2) rotational speed; (3) hydraulic loading rate; (4) depth of the liquid pool in the bowl; (5) differential speed of the screw conveyor; and (6) polymer conditioning to improve the performance. Because the interrelationships of these variables will be different in each location, specific design recommendations are not available; in fact, bench-scale or pilot-plant tests are recommended.

Gravity Belt Thickening. The development of gravity belt thickeners stemmed from the application of belt presses for sludge dewatering. In belt-press dewatering, particularly for sludges having solids concentrations less than 2 percent, effective thickening occurred in the gravity drainage section of the press. The equipment developed for thickening consists of a gravity belt that moves over rollers driven by a variable-speed drive unit (see Fig. 13–14). The sludge is conditioned with polymer and fed into a feed/distribution box at one end, where the sludge is distributed evenly across the width of the moving belt. The water drains through the belt as the concentrating sludge is carried toward the discharge

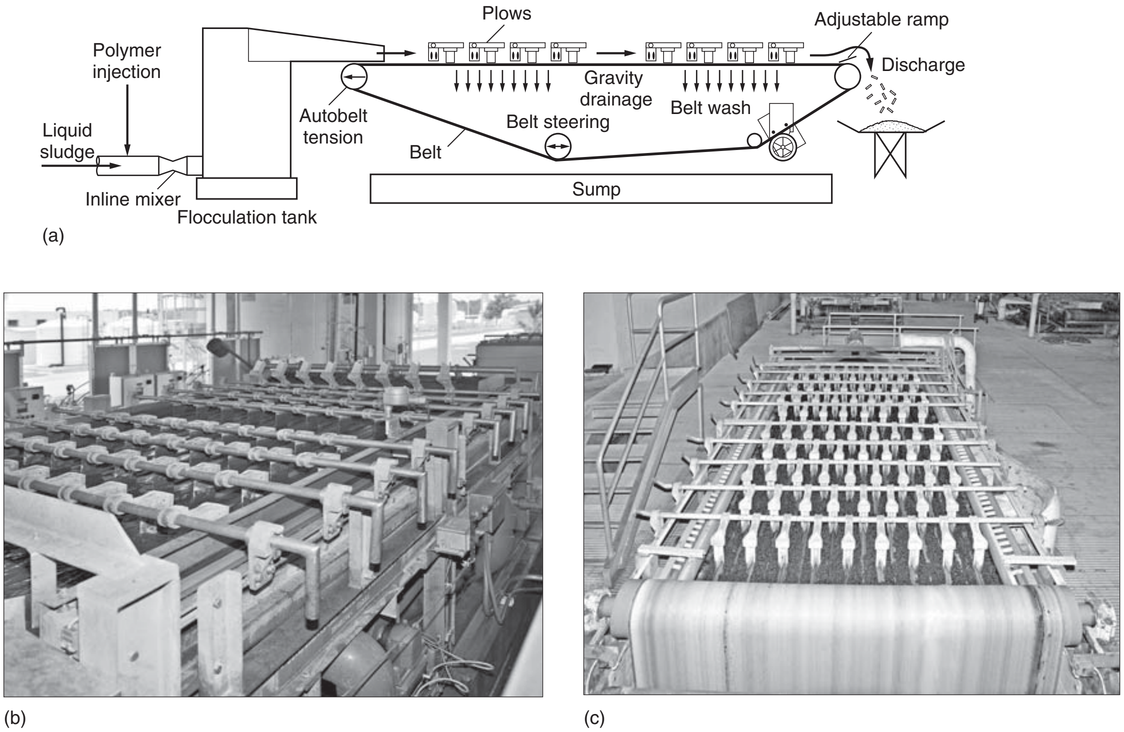


Figure 13–14

Gravity belt thickener: (a) schematic diagram (courtesy of Ashbrook Corporation), (b) top view of sludge plows used to aid the dewatering process, and (c) top view of gravity belt thickener viewed from the discharge end.

end of the thickener. The sludge is ridged and furrowed by a series of plow blades placed along the travel of the belt [see Fig. 13–14(b)], allowing the water released from the sludge to pass through the belt. After the thickened sludge is removed [see Fig. 13–14(c)], the belt travels through a wash cycle. The gravity-belt thickener has been used for thickening waste activated sludge, anaerobically and aerobically digested sludge, and some industrial sludges. Polymer addition is required. Testing is recommended to verify that the sludge can be thickened at typical polymer dosages.

Typical hydraulic loading rates for gravity-belt thickeners are given in Table 13–21. In lieu of pilot-plant data, a value of 800 L/m·min (64 gal/ft·min) is suggested as a design value;

Table 13–21

Typical hydraulic loading rates for gravity belt thickeners[a,b]

Belt size (effective dewatering width), m	Hydraulic loading range	
	gal/min	L/s
1.0	100–250	6.7–16
1.5	150–375	9.5–24
2.0	200–500	12.7–32
3.0	300–750	18–47

[a] Assumes 0.5 to 1.0 percent feed solids for municipal sludges. Variations in sludge density, belt porosity, polymer reaction rate, and belt speed will act to increase or decrease the rates of flow for any given size belt.

[b] Adapted from WEF (2010a).

Figure 13–15

Rotary drum thickener. (Courtesy of Parkson Corporation.)

the higher the feed rate, the greater the operator attention required to maintain stable operation. Solids loading rates range on the order of 200 to 600 kg/m·h (135 to 400 lb/ft^2·h). Systems are often designed for a maximum of 4 to 7 percent thickened solids. Solids capture typically ranges between 90 and 98 percent (WEF, 2010a). Polymer dosages for thickening waste activated sludge range from 3 to 7 kg of dry polymer per tonne of dry solids (6 to 14 lb/ton).

Rotary Drum Thickening. Rotary media-covered drums are also used to thicken sludges. A rotary drum thickening system consists of a conditioning system (including a polymer feed system) and rotating cylindrical screens (see Fig. 13–15). Polymer is mixed with dilute sludge in the mixing and conditioning drum. The conditioned sludge is then passed to rotating screen drums, which separate the flocculated solids from the water. Thickened sludge rolls out the end of the drums, while separated water decants through the screens. Some designs also allow coupling of the rotary drum unit to a belt filter press for combination thickening and dewatering.

Rotary drum thickeners can be used as a prethickening step before belt-press dewatering and are typically used in small- to medium-sized plants for waste activated sludge thickening. The addition of large amounts of polymer for conditioning can be of concern because of floc sensitivity and shear potential in the rotating drum (WEF, 2010a). Rotary drum thickeners are available in capacities up to 24 L/s (400 gal/min). Typical performance data for rotary drum thickeners are given in Table 13–22.

Table 13–22

Typical Performance ranges for rotary drum thickeners for sludge and biosolids[a]

Type of feed	Feed, % TS	Water removed, %	Thickened solids, %	Solids recovery, %
Untreated sludge				
Primary	3.0–6.0	40–75	7–9	93–98
WAS[b]	0.5–1.0	70–90	4–9	93–99
Primary + WAS	2.0–4.0	50	5–9	93–98
Anaerobically digested biosolids	2.5–5.0	50	5–9	90–98
Aerobically digested biosolids	0.8–2.0	70–80	4–6	90–98

[a] WEF (2010a).

[b] WAS = waste activated sludge.

13–7 INTRODUCTION TO SLUDGE STABILIZATION

Sludge is stabilized to (1) reduce pathogens, (2) eliminate offensive odors, and (3) inhibit, reduce, or eliminate the potential for putrefaction. The success in achieving these objectives is related to the effects of the stabilization operation or process on the volatile or organic fraction of the sludge. Survival of pathogens, release of odors, and putrefaction occur when microorganisms are allowed to flourish in the organic fraction of the sludge. The means to eliminate these nuisance conditions is mainly related to the biological reduction of the volatile content and the addition of chemicals to the sludge or biosolids to render them unsuitable for the survival of microorganisms.

Stabilization is not practiced at all wastewater treatment plants, but it is used by an overwhelming majority of plants ranging in size from small to very large. In addition to the health and aesthetic reasons cited above, stabilization can result for volume reduction, production of usable gas (methane), and improved sludge dewaterability.

The principal methods used for stabilization of sludge are (1) alkaline stabilization, usually with lime; (2) anaerobic digestion; (3) aerobic digestion; and (4) composting. These processes are generally defined in Table 13–23. Each of the processes, with the exception of

Table 13–23

Description of sludge stabilization processes

Process	Description	Comments
Alkaline stabilization	Addition of an alkaline material, usually lime, to maintain a high pH level to effect the destruction of pathogenic organisms.	An advantage of alkaline stabilization is that a rich soil-like product results with substantially reduced pathogens. A disadvantage is that the product mass is increased by the addition of the alkaline material. Some alkaline stabilization processes are capable of producing a Class A sludge.
Anaerobic digestion	The biological conversion of organic matter by fermentation in a heated reactor to produce methane gas and carbon dioxide. Fermentation occurs in the absence of oxygen.	Methane gas can be used beneficially for the generation of heat or electricity. The resulting biosolids may be suitable for land application. The process requires skilled operation as it may be susceptible to upsets and recovery is slow.
Aerobic digestion	The biological conversion of organic matter in the presence of air (or oxygen), usually in an open-top tank.	Process is much simpler to operate than an anaerobic digester, but no usable gas is produced. The process is energy-intensive because of the power requirements necessary for mixing and oxygen transfer.
Autothermal thermophilic digestion	Process is similar to aerobic digestion except higher amounts of oxygen are added to accelerate the conversion of organic matter. Process operates at temperatures of 40 to 80 °C, autothermally in an insulated tank.	Process is capable of producing a Class A sludge. Skilled operators are required and the process is a high energy user (to produce air or oxygen).
Composting	The biological conversion of solid organic matter in an enclosed reactor or in windrows or piles.	A variety of sludge or biosolids can be composted. Composting requires the addition of a bulking agent to provide an environment suitable for biological activity. Volume of compost produced is usually greater than the volume of wastewater sludge being composted. Class A or Class B sludge can be produced. Odor control is very important as process is odorous.

Table 13–24

Relative degree of attenuation achieved with various sludge stabilization processes[a]

Process	Degree of attenuation		
	Pathogens	Putrefaction	Odor potential
Alkaline stabilization	Good	Fair	Fair
Anaerobic digestion	Fair	Good	Good
Advanced anaerobic digestion	Excellent	Good	Good
Aerobic digestion	Fair	Good	Good
Autothermal thermophilic digestion (ATAD)	Excellent	Good	Good
Composting	Good	Good	Fair to good

[a] Adapted in part from WEF (2010a).

composting which is considered in Chap. 14, is discussed in more detail in the following sections, and their ability to mitigate or stabilize the effects related to pathogens, putrefaction, and odors is given in Table 13–24. Heat treatment and the addition of oxidizing chemicals, processes that seldom are used in the United States for stabilization, are not included in this text. For information about these processes please refer to Metcalf & Eddy (1991).

When designing a stabilization process, it is important to consider the sludge quantity to be treated, the integration of the stabilization process with the other treatment units, and the objectives of the stabilization process. The objectives of the stabilization process are often affected by existing or pending regulations. If sludge is to be applied on land, pathogen reduction has to be considered. The effect of regulations on application of biosolids to land is discussed in Sec. 14–8.

13–8 ALKALINE STABILIZATION

A method used to eliminate nuisance conditions in sludge involves the use of an alkaline material to render the sludge unsuitable for the survival of microorganisms. In the lime stabilization process, lime is added to untreated sludge in sufficient quantity to raise the pH to 12 or higher. The high pH creates an environment that halts or substantially retards the microbial reactions that can otherwise lead to odor production and vector attraction. The sludge will not putrefy, create odors, or pose a health hazard so long as the pH is maintained at this level. However, high ammonia odor levels have been observed during lime stabilization. The process can also inactivate virus, bacteria, and other microorganisms present. Advantages and disadvantages of alkaline stabilization are summarized in Table 13–25.

Chemical Reactions in Lime Stabilization

The lime stabilization process involves a variety of chemical reactions that alter the chemical composition of the sludge. The following simplified equations are illustrative of the types of reactions that may occur (WEF, 2010a):
Calcium

$$\mathrm{Ca^{2+} + 2HCO_3^- + CaO \rightarrow 2CaCO_3 + H_2O} \qquad (13\text{–}7)$$

Phosphorus

$$\mathrm{2PO_4^{3+} + 6H^+ + 3CaO \rightarrow Ca_3(PO_4)_2 + 3H_2O} \qquad (13\text{–}8)$$

Table 13–25

Advantages and disadvantages of alkaline stabilization[a]

Advantages	Disadvantages
1. Well proven process 2. Product is suitable for a variety of uses that are consistent with the EPA's national beneficial reuse policy 3. Simple technology requiring few special skills for reliable operation 4. Easy to construct of readily available parts 5. Small footprint 6. Flexible operation, easily started and stopped 7. Can produce Class A or Class B biosolids	1. The resulting product is not suitable for use on all soil, especially high alkaline soils 2. The volume of material to be managed and moved off-site is increased by approximately 15 to 50 percent in comparison with other stabilization techniques, such as digestion. The increased volume results in higher transportation costs when material is moved off-site 3. Potential for odor generation both at the processing and end use site due to ammonia and TMA release 4. Potential for dust production 5. The nitrogen content in the final product is lower than that in several other biosolids products because of ammonia volatilization. In addition, available phosphorous can be reduced through the formation of calcium phosphate

[a] Adapted, in part, from EPA (2000).

Carbon dioxide

$$CO_2 + CaO \rightarrow CaCO_3 \tag{13–9}$$

Reactions with organic contaminants:
Acids:

$$R\text{COOH} + \text{CaO} \rightarrow R\text{COOCaOH} \tag{13–10}$$

Fats:

$$\text{Fat} + \text{Ca(OH)}_2 \rightarrow \text{glycerol} + \text{fatty acids} \tag{13–11}$$

Other reactions also occur, such as the hydrolysis of polymers, especially polymeric carbohydrates and proteins, and the hydrolysis of ammonia from amino acids.

Initially, lime addition raises the pH of the sludge. Then, reactions occur such as those in the above equations. If insufficient lime is added, the pH decreases as the reactions take place. Therefore, excess lime is required.

Biological activity produces compounds, such as carbon dioxide and organic acids, that react with lime. If biological activity in the sludge being stabilized is not sufficiently inhibited, these compounds will be produced, reducing the pH and resulting in inadequate stabilization. Many odorous, volatile off-gases are also produced, especially ammonia, which require collection and treatment in odor-control systems such as chemical scrubbers or biofilters (see Chap. 16). Other odorous material such as trimethyl amine (TMA) is generated from the degradation of conditioning polymers during lime stabilization (Dentel et al., 2005).

Heat Generation

If quicklime, CaO (or any compound high in quicklime), is added to sludge, it initially reacts with water to form hydrated lime. This reaction is exothermic and releases approximately

64 kJ/g·mole (2.75×10^4 Btu/lb·mole) (WEF, 2010a). The reaction between quicklime and carbon dioxide is also exothermic, releasing approximately 180 kJ/g·mole (7.8×10^4 Btu/lb·mole). These reactions can result in substantial temperature rise (see discussion of lime posttreatment).

Application of Alkaline Stabilization Processes

Three methods of alkaline stabilization are commonly used: (1) addition of lime to sludge prior to dewatering, termed "lime pretreatment," (2) the addition of lime to sludge after dewatering, or "lime posttreatment," and (3) advanced alkaline stabilization technologies. Either hydrated lime, $Ca(OH)_2$, or quicklime is used most commonly for lime stabilization. Fly ash, cement kiln dust, and carbide lime have also been used as a substitute for lime in some cases.

Lime Pretreatment. Pretreatment (before dewatering) of liquid sludge with lime has been used for either (1) the direct application of liquid sludge to land, or (2) combining benefits of sludge conditioning and stabilization prior to dewatering. In the former case, large quantities of liquid sludge have to be transported to land disposition sites, which limits utilization of lime pretreatment of sludge to small treatment plants. When pretreatment is used prior to dewatering, dewatering has been accomplished using a pressure-type filter press and/or screw press. Lime pretreatment is seldom used with centrifuges or belt filter presses because of abrasive wear and scaling problems.

Lime pretreatment of liquid sludge requires more lime per unit weight of sludge processed than that necessary for dewatering. The higher lime dose is needed to attain the required pH because of the chemical demand of the liquid. In addition, sufficient contact time must be provided before dewatering so as to provide a high level of pathogen kill. The recommended design objective is to maintain the pH above 12 for about 2 h to ensure pathogen destruction (the minimum U.S. EPA criterion for lime stabilization), and to provide enough residual alkalinity so that the pH does not drop below 11 for several days. The lime dosage required varies with the type of sludge and solids concentration. Typical dosages are reported in Table 13–26. Generally, as the percent solids concentration increases, the required lime dose decreases. Testing should be performed for specific applications to determine the actual dosage requirements.

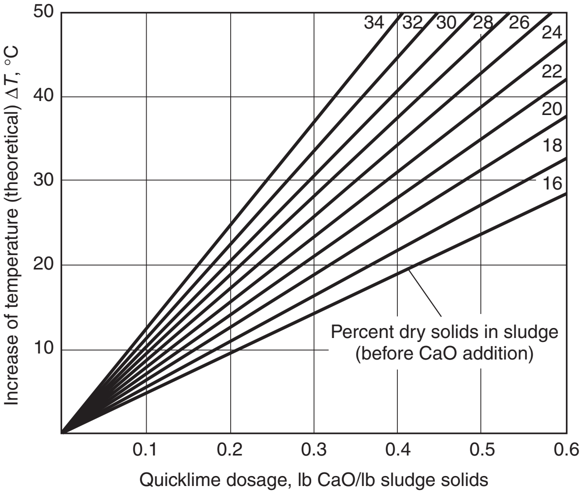
Table 13–26

Typical lime dosages for pretreatment sludge stabilization[a]

	Solids concentration, %		Lime dosage[b] lb $Ca(OH)_2$/ton dry solids		Lime dosage[b] g $Ca(OH)_2$/kg dry solids	
Type of sludge	**Range**	**Average**	**Range**	**Average**	**Range**	**Average**
Primary	3–6	4.3	120–340	240	60–170	120
Waste activated	1–1.5	1.3	420–860	600	210–430	300
Septage	1–4.5	2.7	180–1020	400	90–510	200

[a] Adapted from WEF, (1995a).

[b] Amount of $Ca(OH)_2$ required to maintain a pH of 12 for 30 min.



Note: In practice, higher temperature values are expected. 1.8 (°C) + 32 = °F

Figure 13–16

Theoretical temperature increase in post-lime stabilized sludge using quicklime (Roediger, 1987).

Because lime stabilization does not destroy the organics necessary for bacterial growth, the sludge must be treated with an excess of lime or used beneficially before the pH drops significantly. An excess dosage of lime may range up to 1.5 times the amount needed to maintain the initial pH of 12. Reasons for pH drop could be due to generation of carbon dioxide and organic acids from degradation of organic material in the sludge. The dissolution of carbon dioxide in the air could be another reason for pH decay. For additional details about pH decay following lime stabilization, WEF (1995a) is recommended.

Lime Posttreatment. In lime posttreatment, quicklime is mixed with dewatered sludge in a pugmill, paddle mixer, or screw conveyor to raise the pH of the mixture. Quicklime is used because the exothermic reaction of quicklime and water can raise the temperature of the mixture above 50°C, sufficient to inactivate worm eggs. The theoretical temperature increase by the addition of quicklime is illustrated on Fig. 13–16.

Lime posttreatment is more common than lime pretreatment and has several significant advantages when compared to lime pretreatment: (1) dry lime can be used; therefore, no additional water and equipment is needed for hydrated lime; (2) there are no special requirements for dewatering; and (3) scaling problems and associated maintenance problems of lime-sludge dewatering equipment are eliminated. Adequate mixing is critical for a posttreatment stabilization system so as to avoid pockets of putrescible material. A lime posttreatment stabilization system consists typically of a dry lime feed system, dewatered sludge cake conveyor, and a lime-sludge mixer (see Fig. 13–17). Good mixing is especially important to ensure contact between lime and small particles of sludge. When the lime and sludge are well mixed, the resulting mixture has a crumbly texture, which allows it to be stored for long periods or easily distributed on land by a conventional manure spreader. A potential disadvantage of lime posttreatment may be the release of odorous gases, specifically trimethyl amine (Novak, 2001; Dentel, 2005).

Advanced Alkaline Stabilization Technologies. Alkaline stabilization using materials other than lime is used by a number of municipalities. Most of the technologies that rely on additives, such as cement kiln dust, lime kiln dust, or fly ash, are modifications of conventional dry lime stabilization. The most common modifications include the addition

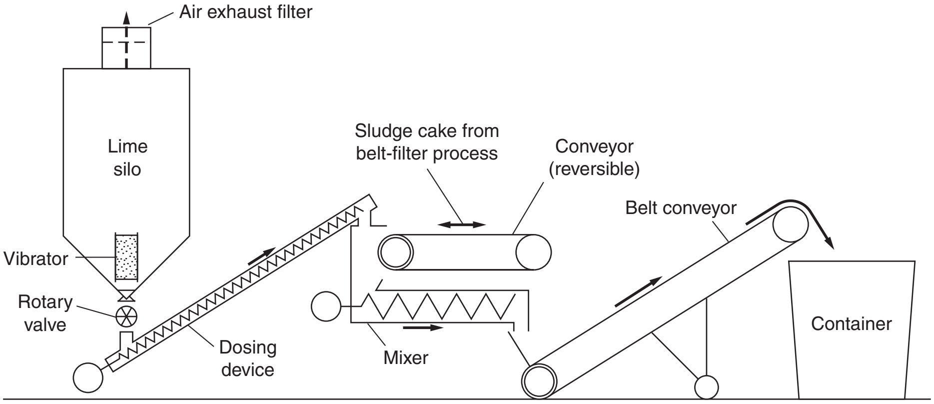


Figure 13–17

Typical lime posttreatment system. (From Roediger Pittsburgh.)

of other chemicals, a higher chemical dose, and supplemental drying. These processes alter the characteristics of the feed material and, depending on the process, may increase product stability, decrease odor potential, and provide product enhancement. To utilize these technologies, dewatered sludge is required.

Pasteurization may be accomplished by the exothermic reaction of quicklime with water to achieve a process temperature of 70°C and maintain it for more than 30 min. Other sources of energy can be used to aid in increasing the temperature generated from the exothermic chemical reaction. For example, the pasteurization system marketed by RDP Company uses electricity to generate heat for raising the temperature to the required degree. N-Viro International Corporation markets an advanced alkaline stabilization system combined with drying. To meet Class A biosolids criteria, the pasteurization reaction must be carried out under carefully controlled and monitored mixing and temperature conditions to ensure uniform treatment and inactivation of pathogens by the heat generated during the reaction. The process produces a soil-like material that is not subject to liquefaction under mechanical stress. Several other process variations of advanced alkaline stabilization are available, some of which are proprietary. Additional information may be found in WEF (2010a) and WEF (2012).

13–9 ANAEROBIC DIGESTION

Anaerobic digestion is among the oldest processes used for the stabilization of sludge. As described in Chap. 10, anaerobic digestion involves the decomposition of organic matter and reduction of inorganic matter (principally sulfate) in the absence of molecular oxygen. The major applications of anaerobic digestion are in the stabilization of concentrated sludges produced from the treatment of municipal and industrial wastewater. Great progress has been made in the fundamental understanding and control of the process, the sizing of tanks, and the design and application of equipment. Because of the emphasis on energy conservation and recovery and the desire to obtain beneficial use of wastewater biosolids, anaerobic digestion continues to be the dominant process for stabilizing sludge. Furthermore, anaerobic digestion of municipal wastewater sludge can, in many cases, produce sufficient digester gas to meet most of the energy needs for plant operation. An aerial view of a large digester installation is shown on Fig. 13–18.

In this section, a brief review is provided of process fundamentals followed by discussions of mesophilic anaerobic digestion, the most common basic process used; thermophilic digestion; and phased digestion. Phased digestion covers many of the new developments in anaerobic digestion.

Figure 13–18

Aerial view of several large anaerobic digesters at Boston, MA.

Process Fundamentals

As described in Chap. 7, the three types of chemical and biochemical reactions that occur in anaerobic digestion are hydrolysis; fermentation, also called acidogenesis (the formation of soluble organic compounds and short-chain organic acids); and methanogenesis (the bacterial conversion of organic acids into methane and carbon dioxide). Important environmental factors in the anaerobic digestion process are (1) solids retention time, (2) hydraulic retention time, (3) temperature, (4) alkalinity, (5) pH, (6) the presence of inhibitory substances, i.e., toxic materials, and (7) the bioavailability of nutrients and trace metals. The first three factors are important in process selection and are discussed in this section. Alkalinity is a function of feed solids and is important in controlling the digestion process. The effects of pH and inhibitory substances are discussed in Chaps. 7 and 10. The presence of nutrients and trace metals necessary for biological growth is described in Sec. 10–2 in Chap. 10.

Solids and Hydraulic Retention Times. Anaerobic digester sizing is based on providing sufficient residence time in well-mixed reactors to allow significant destruction of volatile suspended solids (VSS) to occur. Sizing criteria that have been used are (1) solids retention time SRT, the average time the solids are held in the digestion process, and (2) the hydraulic retention time τ, the average time the liquid is held in the digestion process. For soluble substrates, the SRT can be determined by dividing the mass of solids in the reactor (M) by the mass of solids removed daily (M/d). The hydraulic retention time τ is equal to the volume of liquid in the reactor (m^3) divided by the quantity of biosolids removed (m^3/d). For digestion systems without recycle, SRT $= \tau$.

The three reactions (hydrolysis, fermentation, and methanogenesis) are directly related to SRT (or τ). An increase or decrease in SRT results in an increase or decrease in the extent of each reaction. There is a minimum SRT for each reaction. If the SRT is less than the minimum SRT, bacteria cannot grow rapidly enough and the digestion process will fail eventually (WEF, 2010a).

Temperature. As discussed in Sec. 7–5, temperature not only influences the metabolic activities of the microbial population but also has a profound effect on such factors as gas transfer rates and the settling characteristics of biological sludges. In anaerobic digestion, temperature is important in determining the rate of digestion, particularly the rates of hydrolysis and methane formation. The minimum SRT required to achieve a given amount of VSS destruction is based on the design operating temperature. Most anaerobic digestion systems are designed to operate in the mesophilic temperature range, between 30 and 38°C (85 and 100°F). Other systems are designed for operation in the thermophilic temperature range of 50 to 57°C (122 to 135°F). Newly developed systems, as discussed in a latter part of this section, use a combination of mesophilic and thermophilic digestion in separate stages.

While selection of the design operating temperatures is important, maintaining a stable operating temperature is more important because the bacteria, especially the methane formers, are sensitive to temperature changes. Generally, temperature changes greater than 1°C/d affect process performance, and thus changes less than 0.5°C/d are recommended (WEF, 2010a).

Alkalinity. Calcium, magnesium, and ammonium bicarbonates are examples of buffering substances found in a digester. The digestion process produces ammonium bicarbonate from the breakdown of protein in the raw sludge feed; the others are found in the feed sludge. The concentration of alkalinity in a digester is, to a great extent, proportional to the solids feed concentration. A well-established digester has a total alkalinity of 2000 to 5000 mg/L.

The principal consumer of alkalinity in a digester is carbon dioxide, and not volatile fatty acids as is commonly believed (Speece, 2001). Carbon dioxide is produced in the fermentation and methanogenesis phases of the digestion process (see Sec. 7–12 in Chap. 7). Due to the partial pressure of gas in a digester, the carbon dioxide solubilizes and forms carbonic acid, which consumes alkalinity. The carbon dioxide concentration in the digester gas is, therefore, reflective of the alkalinity requirements. Volatile fatty acids are intermediate products from the acid phase of digestion and consume alkalinity. Volatile acids in digesters range from 50 to 300 mg/L. The ratio of volatile acids to the alkalinity is a parameter that is used to monitor the health of the digestion process and should be monitored closely. The volatile acids to alkalinity ratio for well-established digesters should fall between 0.05 to 0.25 with a 0.1 value indicating a good buffering capacity. Supplemental alkalinity can be supplied by the addition of sodium bicarbonate, lime, or sodium carbonate.

Description of Mesophilic Anaerobic Digestion Processes

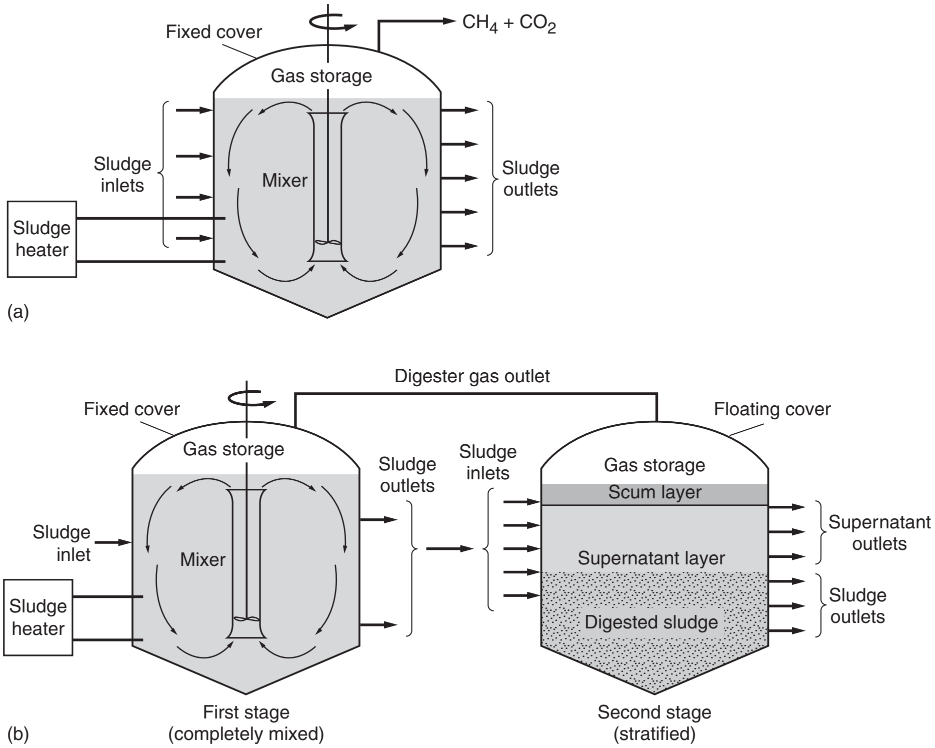
The operation and physical facilities for mesophilic anaerobic digestion in single-stage high-rate, two-stage, and separate digesters for primary sludge and waste activated sludge are described in this section. Standard-rate, sometimes called low-rate, digestion is seldom used for digester design (because of the large tank volume required and the lack of adequate mixing) and is not covered in this text. For information about standard-rate digestion, the reader is referred to the third edition of this text (Metcalf & Eddy, 1991) and WEF (1998). The processes described below normally operate in the mesophilic range; high-rate digesters also operate in the thermophilic range. Thermophilic digestion is discussed at the end of the section.

Single-Stage High-Rate Digestion. Heating, auxiliary mixing, uniform feeding, and thickening of the feed stream characterize the single-stage high-rate digestion process. The sludge is mixed by one of many systems such as gas recirculation, pumping, or draft-tube mixers (separation of scum and supernatant does not take place), and sludge is heated to achieve optimum digestion rates [see Fig. 13–19(a)].

Uniform feeding is very important, and sludge should be pumped to the digester continuously or on a 30-min to 2-h time cycle to help maintain constant conditions in

Figure 13–19

Schematic d diagram of typical anaerobic digesters (a) high-rate and (b) two-stage.



the reactor. In digesters fed on a daily cycle of 8 or 24 h, it is important to withdraw digested sludge from the digester before adding the feed sludge, because the pathogen kill is significantly greater when compared to using the feed sludge to displace the waste sludge (Speece, 2001). Because there is no supernatant separation in the high-rate digester, and the total solids are reduced by 45 to 50 percent and given off as gas, the digested sludge is about half as concentrated as the untreated sludge feed. Digestion tanks may have fixed roofs or floating covers (see subsequent discussion of digester types). Any or all of the floating covers may be of the gas holder type, which provides excess gas storage capacity. Alternatively, gas may be stored in a separate low-pressure gas holder or compressed and stored under pressure.

Two-Stage Digestion. Two-stage digestion, which was frequently used in the past, is seldom used in modern digester design. In two-stage digestion, a high-rate digester is coupled in series with a second tank [see Fig. 13–19(b)]. The first tank is used for digestion and is heated and equipped with mixing facilities. The second tank is usually unheated and used principally for storage. The tanks may be identical, in which case either one may be the primary. Tanks may have fixed roofs or floating covers, the same as single-stage digestion. In other cases, the second tank may be an open tank or a sludge lagoon. In the case of an open second digester some methane would escape if digestion continued leading to increased carbon footprint of the processes. Two-stage digestion of the type described above is seldom used, mainly because of the expense of building a large tank that is not fully utilized and because the second tank was of negligible benefit, operationally.

Because anaerobically digested biosolids may not settle well, the supernatant withdrawn from the second-stage tank may contain high concentrations of suspended solids. Reasons for poor settling characteristics include incomplete digestion in the primary digester (which generates gases in the secondary digester and causes floating solids) and fine-sized solids that have poor settling characteristics. Supernatant returned to the liquid processing system could cause upset conditions and might require separate treatment. Where two-stage digestion is used, return flows from the second tank must be accounted for in the solids mass balance. Less than 10 percent of the gas generated comes from the second stage.

In some installations, the second stage is a heated and mixed reactor to achieve further stabilization prior to dewatering or other subsequent processing. Additional discussion is provided later in this section on two-phase mesophilic digestion that provides more effective utilization of tank capacity.

Separate Sludge Digestion. Most wastewater treatment plants employing anaerobic digestion use a single digester for the digestion of a mixture of primary and biological sludge. The solid-liquid separation of digested primary sludge, however, is downgraded by even small additions of biological sludge, particularly activated sludge. The rate of reaction under anaerobic conditions is also slowed slightly. In separate sludge digestion, the digestion of primary and biological sludges is accomplished in separate tanks. Reasons cited for separate digestion include (1) the excellent dewatering characteristics of the digested primary sludge are maintained, (2) the digestion process is specifically tailored to the sludge being treated, and (3) optimum process control conditions can be maintained. Design criteria and performance data for the separate anaerobic digestion of biological sludges, however, are very limited. In some cases, especially where biological phosphorus removal is practiced, biological sludge is digested aerobically instead of anaerobically to prevent resolubilization of the phosphorus under anaerobic conditions. Separate sludge digestion is not currently a common practice at most plants.

Process Design for Mesophilic Anaerobic Digestion

Ideally, the design of anaerobic sludge digestion processes should be based on an understanding of the fundamental principles of biochemistry and microbiology discussed in Chap. 7 in Sec. 7–12. Because these principles have not been appreciated fully in the past, a number of empirical methods have also been used in the design of digesters. The purpose of the following discussion is to illustrate the various methods that have been used to design single-stage, high-rate digesters in terms of size. These methods are based on (1) solids retention time, (2) the use of volumetric loading factors, (3) volatile solids destruction, (4) observed volume reduction, and (5) loading factors based on population.

Solids Retention Time. Digester design based on SRT involves application of the principles discussed in Chaps. 7 and 10. To review briefly, the respiration and oxidation end products of anaerobic digestion are methane gas and carbon dioxide. The quantity of methane gas can be calculated using Eq. (13–12):

$$V_{CH_4} = (0.35)[(S_o - S)(Q)(1\,\text{kg}/10^3\,\text{g}) - 1.42P_x] \tag{13–12}$$

where V_{CH_4} = volume of methane produced at standard conditions (0°C and 1 atm), m^3/d

0.35 = theoretical conversion factor for the amount of methane produced, m^3, from the conversion of 1 kg of bCOD at 0°C (conversion factor at 35°C = 0.40, see Example 7–10 in Chap. 7)

Q = flowrate, m^3/d

S_o = bCOD in influent, g/m^3

S = bCOD in effluent, g/m^3

P_x = net mass of cell tissue produced per day, kg/d

Table 13–27

Suggested solids retention times for use in the design of complete-mix anaerobic digesters[a]

Operating temperature, °C	SRT (minimum)	SRT_{des}
18	11	28
24	8	20
30	6	14
35	4	10
40	4	10

[a] From McCarty (1964) and (1968).

Note: 1.8 (°C) + 32 = °F.

The theoretical conversion factor for the amount of methane produced from the conversion of 1 g of bCOD is derived in Sec. 7–12 in Chap. 7. For a complete-mix high-rate digester without recycle, the mass of biological solids synthesized daily, P_x, can be estimated using Eq. (13–13).

$$P_x = \frac{YQ(S_o - S)(1\ \text{kg}/10^3\ \text{g})}{1 + b(\text{SRT})} \tag{13–13}$$

where Y = yield coefficient, g VSS/g bCOD
b = endogenous coefficient, d^{-1} (typical values range from 0.02 to 0.04)
SRT = solids retention time, d
other terms as defined previously

For a complete-mix digester, the SRT is the same as the hydraulic retention time τ.

Typical anaerobic reaction values for Y and b are given in Table 10–13 in Chap. 10 and range from 0.05 to 0.10 and 0.01 to 0.04, respectively. Typical values for SRT at various temperatures are reported in Table 13–27. In practice for high-rate digestion, however, values for SRTs range from 15 to 20 d. Grady, Daigger, and Lim (1999) observed that (1) a lower SRT limit of 10 days at a temperature of 35°C is sufficient to ensure an adequate safety factor against a washout of the menthanogenic population, and (2) incremental changes in volatile solids destruction are relatively small for SRT values above 15 d at 35°C. In selecting the design SRT for anaerobic digestion, peak hydraulic loading must be considered. The peak loading can be estimated by combining poor thickener performance with the maximum sustained plant loading expected during seven continuous days during the design period (U.S. EPA, 1979). The application of Eqs. (13–12) and (13–13) in the process design of a high-rate digester is illustrated in Example 13–5.

EXAMPLE 13–5 Estimating Single-stage, High-rate Digester Volume and Performance

Estimate the size of digester required for primary sludge from a primary clarifier designed for 38,000 m^3/d (10 Mgal/d) of wastewater. Check the volumetric loading and the amount of gas produced. The influent wastewater BOD and TSS concentrations are 400 and 300 mg/L, respectively. The primary clarifier achieves 35 percent BOD removal and 50 percent TSS removal. Assume that the primary sludge contains about 95 percent moisture and has a specific gravity of 1.02. Other pertinent design assumptions are as follows:

1. The hydraulic regime of the reactor is complete-mix.
2. τ = SRT = 15 d at 35°C (see Table 13–27).

3. Efficiency of waste utilization (solids conversion) E = 0.70.
4. The sludge contains adequate nitrogen and phosphorus for biological growth.
5. $Y = 0.08$ kg VSS/kg bCOD utilized and $b = 0.03\ \text{d}^{-1}$.
6. Constants are for a temperature of 35°C.
7. Digester gas is 65 percent methane.

Solution

1. Determine the daily sludge mass and volume using Eq. (13–2).

$$\text{Sludge mass} = \frac{(38{,}000\ \text{m}^3/\text{d})(300\ \text{g/m}^3)(0.5)}{(10^3\ \text{g}/1\ \text{kg})} = 5700\ \text{kg/d}$$

$$\text{Sludge volume} = \frac{(5700\ \text{kg/d})}{1.02(10^3\ \text{kg/m}^3)(0.05)} = 111.8\ \text{m}^3/\text{d}$$

2. Determine the bCOD loading.

$$\text{bCOD loading} = (0.35)(400\ \text{g/m}^3)(38{,}000\ \text{m}^3/\text{d})(1\ \text{kg}/10^3\ \text{g}) = 5320\ \text{kg/d}$$

3. Compute the digester volume.

$$\tau = \frac{V}{Q}$$

$$V = Q\tau = (111.8\ \text{m}^3/\text{d})(15\text{d}) = 1677\text{m}^3$$

4. Compute the volumetric loading.

$$\frac{(\text{kg bCOD/d})}{\text{m}^3} = \frac{(5320\ \text{kg/d})}{1677\ \text{m}^3} = 3.17\ \text{kg/m}^3{\cdot}\text{d}$$

5. Compute the quantity of volatile solids produced per day using Eq. (13–13).

$$P_x = \frac{YQ(S_o - S)(10^3\ \text{g/kg})^{-1}}{1 + b(\text{SRT})}$$

$$S_o = 5320\ \text{kg/d}$$

$$S = 5320(1 - 0.70) = 1596\ \text{kg/d}$$

$$S_o - S = 5320 - 1596 = 3724\ \text{kg/d}$$

$$P_x = \frac{(0.08)[(5320 - 1596)\text{kg/d}]}{1 + (0.03\text{d}^{-1})(15\text{d})} = 205.5$$

6. Compute the volume of methane produced per day at 35°C using Eq. (13–12) (conversion factor at 35°C = 0.40).

$$V_{\text{CH}_4} = (0.40)[(S_o - S)(Q)(10^3\ \text{g/kg})^{-1} - 1.42\ P_x]$$

$$V_{\text{CH}_4} = (0.4\ \text{m}^3/\text{kg})[(5320 - 1596)\text{kg/d} - 1.42(205.5\ \text{kg/d})]$$

$$= 1373\ \text{m}^3\text{d}$$

7. Estimate the total gas production.

$$\text{Total gas volume} = \frac{1373}{0.65} = 2112\ \text{m}^3/\text{d}$$

Table 13–28

Typical design criteria for sizing mesophilic high-rate complete-mix anaerobic sludge digesters[a]

	U.S. customary units		SI units	
Parameter	**Units**	**Value**	**Units**	**Value**
Volume criteria				
Primary sludge	ft^3/capita	1.3–2.0	m^3/capita	0.03–0.06
Primary sludge + trickling filter humus sludge	ft^3/capita	2.6–3.3	m^3/capita	0.07–0.09
Primary sludge + activated sludge	ft^3/capita	2.6–4.0	m^3/capita	0.07–0.11
Solids loading rate[b]	lb VSS/10^3 $ft^3 \cdot d$	100–300	kg VSS/$m^3 \cdot d$	1.6–4.8
Solids retention time[b]	d	15–20	d	15–20

[a] Adapted, in part, from U.S. EPA (1979).

[b] Based on combined primary and secondary sludges digestion without any pretreatment methods.

Loading Factors. One of the most common methods used to size digesters is to determine the required volume based on a loading factor. Although a number of different factors have been proposed, the two most favored are based on (1) the mass of volatile solids added per day per unit volume of digester capacity and (2) the mass of volatile solids added to the digester each day per mass of volatile solids in the digester. Of the two, the first method is preferred. Loading criteria are based generally on sustained loading conditions (see Chap. 3), typically peak 2-wk or peak mo sludge production with provisions for avoiding excessive loadings during shorter periods. Typical design criteria for sizing mesophilic high-rate anaerobic digesters are given in Table 13–28. The upper limit of volatile solids loading rates is typically determined by the rate of accumulation of toxic materials, particularly ammonia, or washout of methane formers (WEF, 2010a).

Excessively low volatile solids loading rates can result in designs that are costly to build and are troublesome to operate. In a survey conducted by Speece (1988) of 30 digester installations in the United States, one of the most significant observations was the relatively low solids content in the sludge feed to the digesters. The average TSS in the sludge feed was 4.7 ± 1.6 percent and the average volatile solids content was 70 percent. The average VSS value in the digesters was a dilute 1.6 percent. Dilute sludge feed causes low volatile solids loading leading to starving conditions within the digester, resulting in the following adverse effects in digester operation: (1) reduced τ, (2) reduced VS destruction, (3) reduced methane generation, (4) reduced alkalinity, (5) increased volumes of digested biosolids and supernatant, (6) increased heating requirements, (7) increased dewatering capacity, and (8) increased hauling cost for liquid biosolids. As a cautionary note, a potential problem with ammonia toxicity could occur if the waste activated sludge is thickened too much. Thus, in planning the design and operation of anaerobic digesters, consideration should be given to optimizing volatile solids loading to effectively utilize digester capacity. The effect of solids concentration and hydraulic detention time on volatile solids loading is reported in Table 13–29.

Estimating Volatile Solids Destruction. The degree of stabilization obtained is often measured by the percent reduction in volatile solids. The reduction in volatile solids can be related either to the SRT or to the detention time based on the untreated sludge feed.

Table 13–29

Effect of sludge concentration and hydraulic detention time on volatile solids loading factors[a]

Sludge concentration, %	Volatile solids loading factor							
	lb/ft³·d				kg/m³·d			
	10 d[b]	12 d	15 d	20 d	10 d	12 d	15 d	20 d
2	0.09	0.07	0.06	0.04	1.4	1.2	0.95	0.70
3	0.13	0.11	0.09	0.07	2.1	1.8	1.4	1.1
4	0.18	0.15	0.12	0.09	2.9	2.4	1.9	1.4
5	0.22	0.19	0.15	0.11	3.6	3.0	2.4	1.8
6	0.27	0.22	0.18	0.13	4.3	3.6	2.9	2.1
7	0.31	0.26	0.21	0.16	5.0	4.2	3.3	2.5
8	0.36	0.30	0.24	0.18	5.7	4.8	3.8	2.9

[a] Based on 70 percent volatile content of sludge, and a sludge specific gravity of 1.02 (concentration effects neglected).

[b] Hydraulic detention time, d.

The amount of volatile solids destroyed in a high-rate complete-mix digester can be roughly estimated by the following empirical equation (Liptak, 1974):

$$V_d = 13.7 \ln(\text{SRT}_{\text{des}}) + 18.9 \tag{13–14}$$

where V_d = volatile solids destruction, %

SRT_{des} = time of digestion, d (range 15 to 20 d)

The equation does not account for variation in the sludge feed to digestion and the digestion mixing and other operating conditions and should be used to obtain a rough estimate only, and it appears that the equation overestimates volatile solids destruction. Typical volatile solids destruction ranges as a function of SRT are provided in Table 13–30. Because the untreated sludge feed can be measured easily, this method is also used commonly. In plant operation, calculation of volatile solids reduction should be made routinely as a matter of record whenever sludge is drawn to processing equipment or drying beds. Alkalinity and volatile acids content should also be checked daily as a measure of the stability of the digestion process.

In calculating the volatile solids reduction, the ash content of the sludge is assumed to be conservative; that is, the number of pounds of ash going into the digester is equal to that being removed. Digester VSR can be calculated based on two different methods. The first method is the mass balance method, which is shown below.

$$R_{\text{VSS}} = \frac{M_{\text{VS in feed}} - M_{\text{VS in digested sludge}} - M_{\text{VS in surpernatent}}}{M_{\text{VS in feed}}} \times 100 \tag{13–15}$$

Table 13–30

Estimated volatile solids destruction in high-rate complete-mix mesophilic anaerobic digestion

Digestion time, d	Volatile solids destruction, %
30	50–65
20	50–60
15	45–50

where R_{VSS} = volatile solids destruction, %
$M_{\text{VS in feed}}$ = mass flowrate of volatiles in digester feed, kg/d
$M_{\text{VS in digested sludge}}$ = mass flowrate of volatiles out of digester, kg/d
$M_{\text{VS in surpernatent}}$ = mass flowrate of volatiles in digester decant stream, kg/d

It should be noted that in modern high rate digesters, there is no decant so the $M_{\text{VS in surpernatent}}$ number goes to zero. Digester VSR can also be calculated with the simplified Van Kleeck formula given below.

$$R_{\text{VSS}} = \frac{W_{\text{VS in feed}} - W_{\text{VS in digested sludge}}}{W_{\text{VS in feed}} - (W_{\text{VS in digested sludge}})(W_{\text{VS in feed}})} \times 100 \tag{13–16}$$

where $W_{\text{VS in feed}}$ = Weight fraction of digested sludge volatile content per total dry solids
$W_{\text{VS in digested sludge}}$ = Weight fraction of volatiles out of digester per total dry solids

It should be noted that the Van Kleeck formula assumes that there is no supernatant withdrawal or accumulation of grit inside the digester so in practice the results may not be 100 percent accurate. A typical example calculation of volatile solids reduction is presented in Example 13–6.

EXAMPLE 13–6 **Determination of Volatile Solids Reduction** From the following analysis of untreated and digested biosolids, determine the total volatile solids reduction achieved during digestion. It is assumed that (1) the weight of fixed solids in the digested biosolids equals the weight of fixed solids in the untreated sludge and (2) the volatile solids are the only constituents of the untreated sludge lost during digestion.

	Volatile solids, %	Fixed solids, %
Untreated sludge	68	32
Digested sludge	50	50

Solution

1. Determine the weight of the digested solids. Because the quantity of fixed solids remains the same, the weight of the digested solids based on 1.0 kg of dry untreated sludge, as computed below, is 0.64 kg.

$$\text{Fixed solids in untreated sludge} = \frac{0.32\text{ kg}}{(0.32 + 0.68)\text{ kg}} 100 = 32\%$$

Let X equal the weight of volatile solids after digestion. Then

$$\text{Fixed solids after digestion} = \frac{0.32\text{ kg}}{(0.32 + X)\text{ kg}} 100 = 50\%$$

$$\text{Weight of volatile solids after digestion, } X\text{ kg} = \frac{0.32\text{ kg}}{0.5} - 0.32 = 0.32\text{ kg}$$

$$\text{Weight of digested solids} = 0.32\text{ kg} + 0.32\text{ kg} = 0.64\text{ kg}$$

2. Determine the percent reduction in total and volatile suspended solids.
   a. Percent reduction of total suspended solids

$$R_{\text{TSS}} = \frac{(1.0 - 0.64)\text{ kg}}{1.0\text{ kg}} 100 = 36\%$$

b. Percent reduction in volatile suspended solids using both methods
Using the mass balance method [Eq. (13–15)]

$$R_{\text{VSS}} = \frac{(0.68 - 0.32)\text{ kg}}{0.68\text{ kg}} 100 = 52.9\%$$

Using the Van Kleeck method [Eq. (13–16)]

$$R_{\text{VSS}} = \frac{0.68 - 0.5}{0.68 - 0.5(0.68)} 100 = 52.9\%$$

Population Basis. Digestion tanks are also designed on a volumetric basis by allowing a certain number of cubic meters per capita (cubic feet per capita). Detention times range from 10 to 20 d for high-rate digesters (U.S. EPA, 1979). These detention times are recommended for design based on total tank volume, plus additional storage volume if sludge is dried on beds and weekly sludge withdrawals are curtailed because of inclement weather.

Typical design criteria for heated anaerobic digesters based on population are shown in Table 13–28. The criteria are applied only where analyses and volumes of sludge to be digested are not available. The capacities shown in Table 13–28 should be increased 60 percent in a municipality where the use of food-waste grinders is universal and should be increased on a population-equivalent basis to allow for the effect of industrial wastes.

Selection of Tank Design and Mixing System

Most anaerobic digestion tanks are either cylindrical, conventional German design, or egg-shaped (see Fig. 13–20). The most common shape used in the United States is a shallow, vertical cylinder with a floating cover [see Fig. 13–20(a)] or fixed cover [see Fig. 13–20(b)]. Rectangular tanks were used in the past, but they experienced great difficulty in mixing the tank contents uniformly. German designers have worked on optimizing the shape of digesters, and two basic types have emerged: the conventional German digester and the egg-shaped digester. The conventional German digester [see Fig. 13–20(c)] is a deep cylindrical vessel with steeply sloped top and bottom cones (Stukenberg et al., 1992). The egg-shaped digester, shown on Fig. 13–20(d), is similar in appearance to an upright egg, and the design is sometimes modified to a sphere-cone shape. Egg-shaped tanks have been used extensively in Europe, especially in Germany, and are growing in popularity in the United States. Essentially all of the modern digester designs in the United States are of either the cylindrical or egg-shaped type. Cylindrical and egg-shaped digesters and the mixing systems used for each type of tank are discussed in the following paragraphs. Advantages and disadvantages of each type of digester are summarized in Table 13–31.

Proper mixing is one of the most important considerations in achieving optimum process performance. Various systems for mixing the contents of the digesters have been used; the most common types involve the use of (1) gas injection, (2) mechanical stirring, and (3) mechanical pumping. Some digester installations use a combination of gas mixing and recirculation by pumping. The advantages and disadvantages of the various mixing systems are summarized in Table 13–32; typical design parameters are shown in Table 13–33.

Cylindrical Tanks. Cylindrical sludge digesters are seldom less than 6 m (20 ft) or more than 38 m (125 ft) in diameter. The water depth should not be less than 7.5 m (25 ft) at the sidewall because of the difficulty in mixing shallow tanks, and the depth may be as

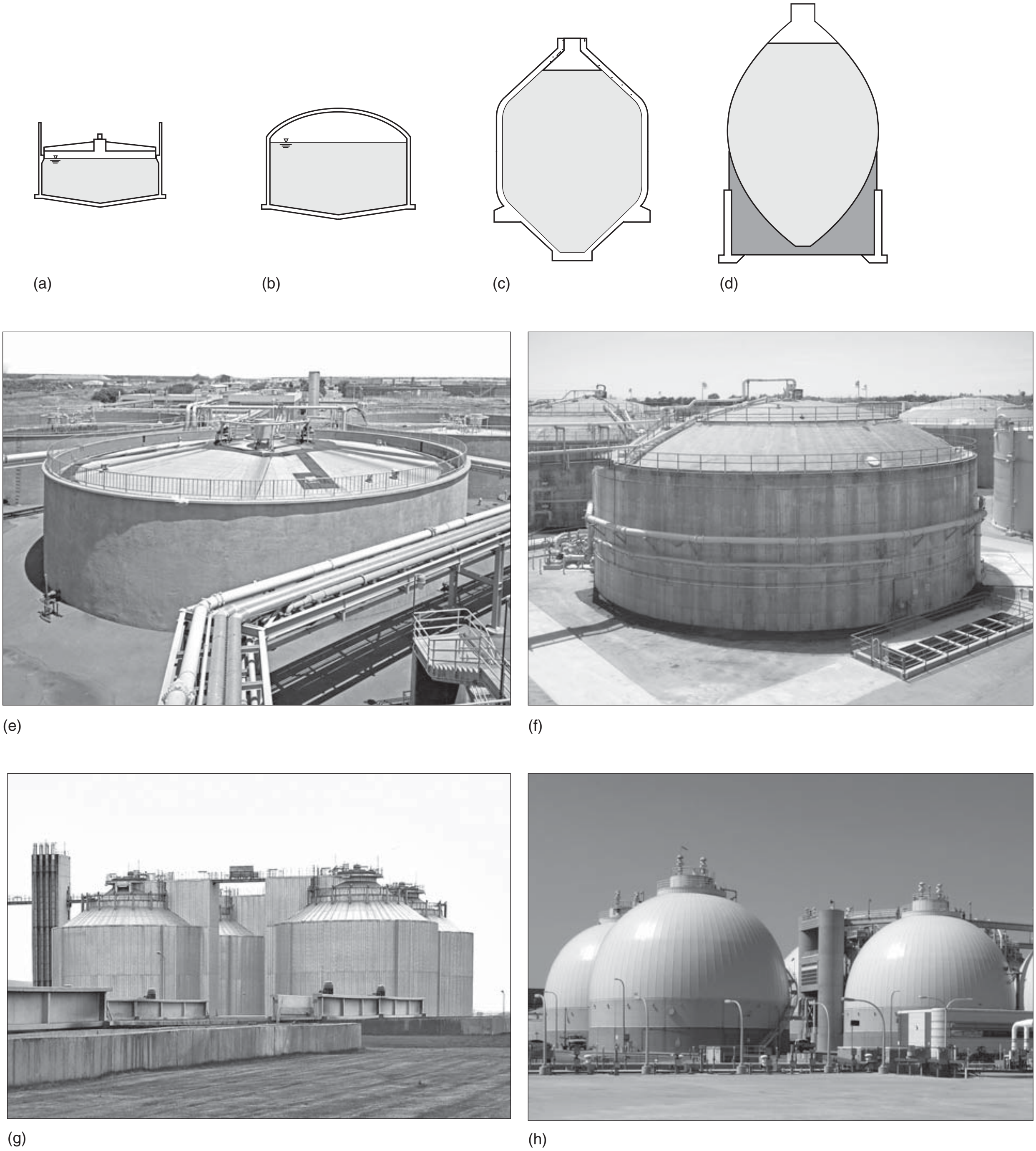


Figure 13–20

Typical shapes of anaerobic digesters: (a) and (e) cylindrical with floating cover, (b) and (f) cylindrical with fixed cover, (c), (g), and (h) conventional German design with reinforced concrete construction [note digesters in (g) and (h) are clad in a metal sheath], and (d) egg-shaped with steel shell (see also Figs. 13–24 and 13–25).

Table 13–31
Comparison of cylindrical and egg-shaped digesters[a]

Type of digester	Advantages	Disadvantages
Conventional	• Reactor shape results in large volume for gas storage • Reactor can be equipped with gas holder covers • Low profile • Conventional construction techniques can be applied; construction costs can be competitive	• Reactor shape results in inefficient mixing and dead spaces • Poor mixing results in grit accumulation • Large surface area provides space for scum accumulation and foam formation • Cleaning is required for removal of grit and scum accumulation; digester may be required to be taken out of service
Egg-shaped	• Minimum grit accumulation • Reduced scum formation • Higher mixing efficiency • More homogeneous biomass is obtained • Lower operating and maintenance costs; cleaning frequency significantly reduced • Smaller footprint; less land area is required • Foaming is minimized (except for gas mixing)	• Very little gas storage volume; external gas storage is required if as is recovered • High profile structures; may be aesthetically objectionable • Difficult access to top-mounted equipment; installation requires a high stair tower or an elevator • Greater foundation requirements and seismic considerations • Foaming of gas-mixed digester may be a problem in collecting gas • Higher construction costs • Construction limited to specialty contractors

[a] Adapted, in part, from Brinkman and Voss (1998).

much as 15 m (50 ft). The floor of the digester is usually conical with the bottom sloping to the center, with a minimum slope of 1 vertical to 6 horizontal where the sludge is drawn off (see Fig. 13–21). An alternative design uses a “waffle” bottom to minimize grit accumulation and to reduce the need for frequent digester cleaning (see Fig. 13–22).

Gas-injection systems used in cylindrical tanks are classified as unconfined or confined [see Figs. 13–23(a) and (b) on page 1518]. Unconfined gas systems collect gas at the top of the digesters, compress the gas, and then discharge the gas through a pattern of bottom diffusers or through a series of radially placed top-mounted lances. Unconfined gas systems mix the digester contents by releasing gas bubbles that rise to the surface, carrying and moving the sludge. These systems are suitable for digesters with fixed, floating, or gas holder covers. In confined gas systems, gas is collected at the top of the digesters, compressed, and discharged through confined tubes. Two major types of confined systems are the gas lifter and the gas piston [see Figs. 13–23(c) and (d)]. The gas lifter system consists of submerged gas pipes or lances inserted into an eductor tube or gas lifter. Compressed gas is released from the lances or pipes, and the gas bubbles rise, creating an air-lift effect. In the gas piston system, gas bubbles are released intermittently at the bottom of a cylindrical tube or piston. The bubbles rise and act like a piston, pushing the sludge to the surface. These systems are suitable for fixed, floating, or gas holder covers.

Mechanical stirring systems commonly use low-speed turbines or mixers [see Figs. 13–23(e) and (f)]. In both systems, the rotating impeller(s) displaces the sludge, mixing

Table 13–32

Summary of advantages and disadvantages of various anaerobic digester mixing systems

Type of mixer	Advantages	Disadvantages
All systems	• Increased rate of biosolids stabilization	• Corrosion and tear of ferrous metal piping and supports • Equipment wear by grit • Equipment plugging and operational interference by rags
Gas injection:		
Unconfined:		
Cover-mounted lances	• Lower maintenance and less hindrance to cleaning than bottom-mounted diffusers. Effective against scum buildup	• Corrosion of gas piping and equipment • High maintenance for compressor • Potential gas-seal problem • Compressor problems if foam gets inside, solids deposition • Plugging of gas lances • Entire tank contents are not mixed
Bottom-mounted diffusers	• Better movement of bottom deposits than cover-mounted lances	• Corrosion of gas piping and equipment • High maintenance for compressor • Potential gas-seal problem • Foaming. Incomplete mixing • Scum formation • Diffuser plugging • Bottom deposits can alter mixing patterns • Requires digester dewatering for maintenance
Confined:		
Gas lifters	• Better mixing and gas production and better movement of bottom deposits than cover mounted lances. Lower power requirements than cover mounted lances	• Corrosion of gas piping and equipment • High maintenance for compressor • Potential gas-seal problem • Corrosion of gas lifter • Lifter interferes with digester cleaning • Scum buildup • Does not provide good top mixing • Requires digester dewatering for maintenance if bottom mounted
Gas pistons	• Good mixing efficiency • Less susceptible to plugging due to rags or fibrous material • Provides surface agitation for management of scum layer • Can contain optional heating jacket	• Corrosion of gas piping and equipment • High maintenance for compressor • Potential gas-seal problem • Equipment internally mounted • Pistons interfere with digester cleaning • Requires digester dewatering for maintenance • Cannot operate at varying liquid levels

(*continued*)

Table 13–32 (*Continued*)

Type of mixer	Advantages	Disadvantages
Mechanical stirring: Low-speed turbines	• Good mixing efficiency	• Wear of impellers and shafts • Bearing failures. Interferences of impellers with rags • Requires oversized gear boxes • Gas leaks at shaft seal • Long overhung loads
Low-speed mixers	• Breaks up scum layers	• Not designed to mix entire tank contents • Bearing and gear box failures • Impeller wear. Interference of impellers by rags
Linear motion mixing	• Lower energy consumption compared to other technologies • High mixing efficiency • Capable of operating at various liquid levels • Suitable for retrofits • Equipment maintenance can be completed without taking digester out of service • Minimal mechanical complexity	• Limited number of installations in the United States • May require maintenance platform and equipment hoist • No redundancy • Sole supplier
Mechanical pumping:		
Internal draft tubes	• Good top-to-bottom mixing • Reversibility provides variable mixing dynamics • Provides surface agitation for management of scum layer	• Sensitive to liquid level • Corrosion and wear of impeller • Bearing and gear box failures • Requires oversized gear box • Structural modifications required for retrofit
External draft tubes	• Same as internal draft tube • Draft tube maintenance easier than internal type • Can contain optional heating jacket	• Same as internal draft tube
Pumps	• Good mixing control. Scum layer and sludge deposits can be recirculated. Pumps easier to maintain than compressors • Conductive for FOG addition	• Impeller wear • Plugging of pumps by rags[a] • Bearing failures • High electrical consumption • Decreased efficiency at higher solids concentrations • High mixing energy may contribute to foaming

[a] Impact from plugging with rags can be minimized or eliminated by using chopper pumps.

the digester contents. Low-speed turbine systems usually have one cover-mounted motor with two turbine impellers located at different sludge depths. A low-speed mixer system usually has one cover-mounted mixer. Mechanical stirring systems are most suitable for digesters with fixed or floating covers. Another type of mechanical mixer that is new to the market (as of writing this book) is a linear motion mixer that consists of a ring-shaped disc

Table 13–33

Typical design parameters for anaerobic digester mixing systems[a]

Parameter	Type of mixing system	Typical values[b]	
		U.S. customary units	SI units
Unit power	Mechanical systems	0.025–0.04 hp/10^3 gal of digester volume	0.005–0.008 kW/m^3 of digester volume
Unit gas flow[c]	Gas mixing		
	Unconfined	4.5–5 $ft^3/10^3\ ft^3\cdot min$	0.0045–0.005 $m^3/m^3\cdot min$
	Confined	5–7 $ft^3/10^3\ ft^3\cdot min$	0.005–0.007 $m^3/m^3\cdot min$
Velocity gradient, G[d]	All	50–80 s^{-1}	50–80 s^{-1}
Turnover time of tank contents	Confined gas mixing and mechanical systems	20–30 min	20–30 min

[a] Adapted from U.S. EPA (1987).

[b] Actual design values may differ depending on the type of mixing system, manufacturer, and digestion process or function.

[c] Quantity of gas delivered by the gas injection system divided by the digester gas volume.

[d] See Eq. (5–3) in Chap. 5.

Figure 13–21

Typical cross section through a high-rate, gas mixed cylindrical digester.

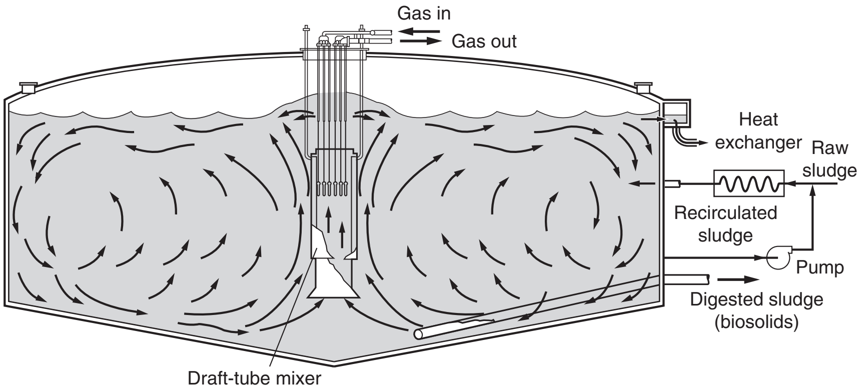
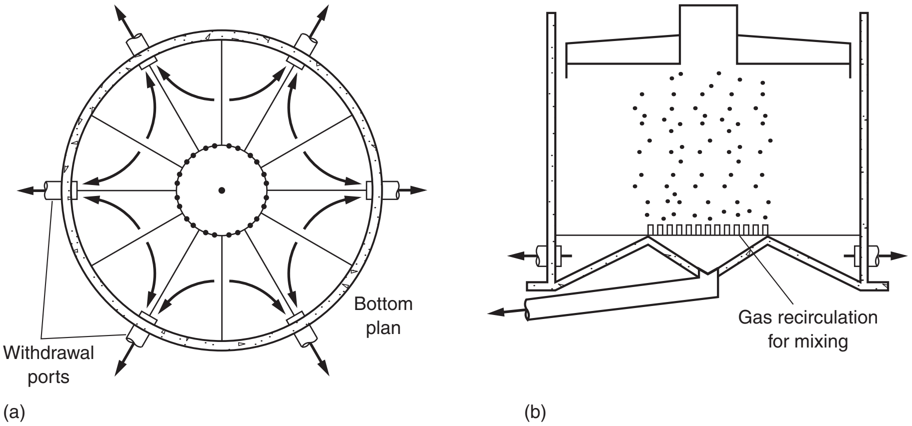


Figure 13–22

Typical waffle bottom anaerobic digester: (a) plan view (b) section.



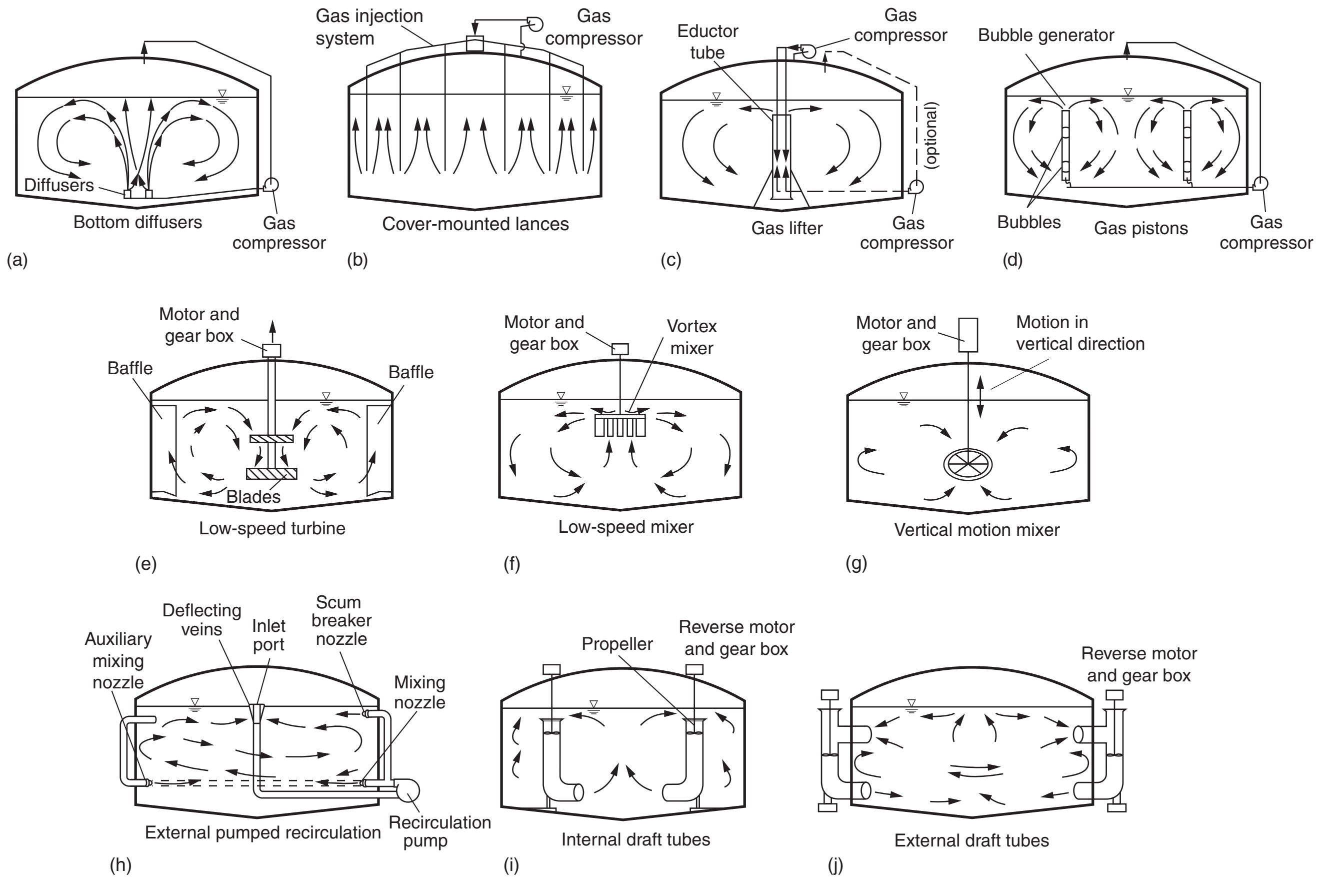


Figure 13–23

Devices used for mixing contents of anaerobic high rate digesters (a) and (b) unconfined gas injection systems; (c) and (d) confined gas injection systems, (e), (f), and (g) mechanical mixing systems; and (h), (i), and (j) mechanical pumping systems.

that oscillates up and down creating strong axial and lateral agitation needed for mixing of the digester contents [see Fig. 13–23(g)].

Most mechanical pumping systems consist of propeller-type pumps mounted in internal or external draft tubes, or axial-flow or centrifugal pumps and piping installed externally [see Figs. 13–23(h), (i), and (j)]. Mixing is promoted by the circulation of sludge. Mechanical pumping systems are suitable for digesters with fixed covers.

Egg-Shaped Tanks. The purpose of the egg-shaped design is to enhance mixing and to eliminate the need for cleaning. The digester sides form a steep cone at the bottom so that grit accumulation is minimized [see Figs. 13–20(d) and 13–24(a)]. Other advantages cited for the egg-shaped design include better control of the scum layer and smaller land-area requirements. Steel construction is more common for egg-shaped tanks in the United States; reinforced concrete construction requires complex formwork and special construction techniques. The structures are relatively high as compared to other treatment plant structures [see Fig. 13–24(b)], and may require an elevator for access to the top of the structure. In Boston, MA (see Fig. 13–18), and Baltimore, MD, the heights of the digesters were over 40 m (130 ft).

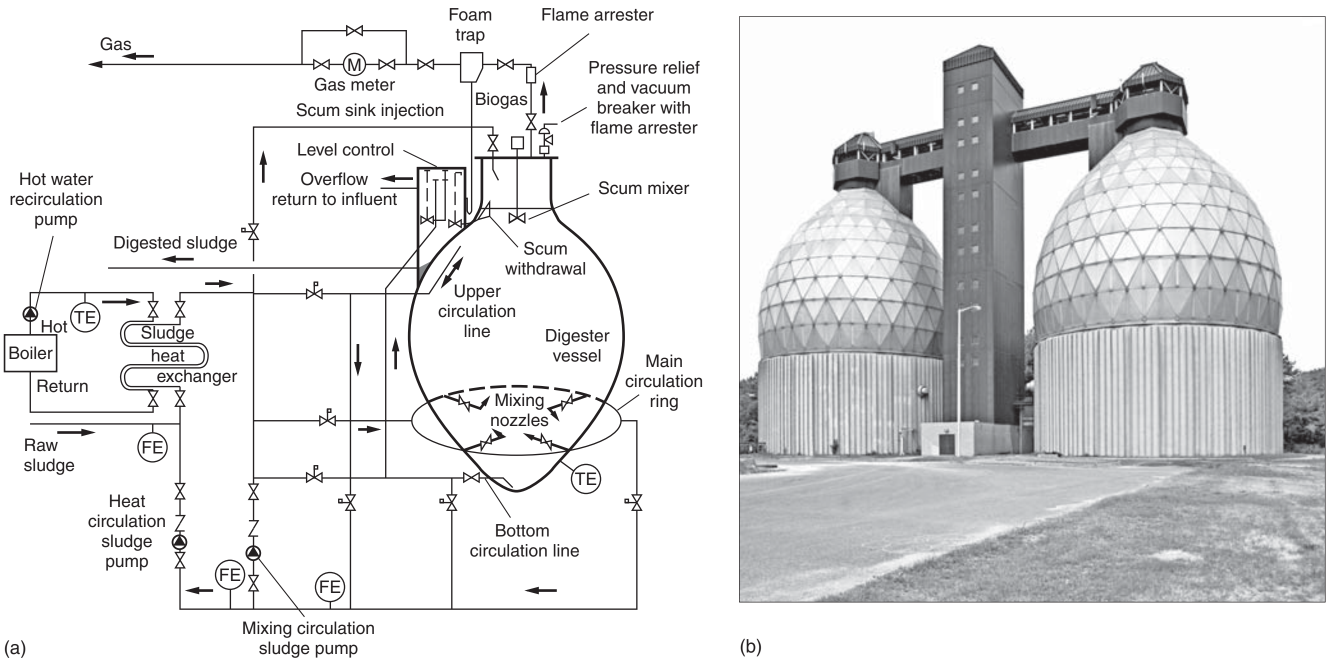


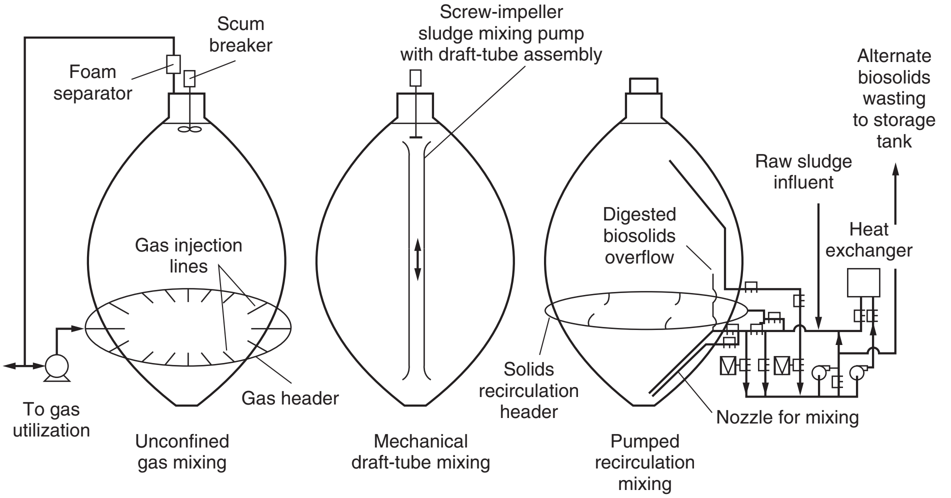
Figure 13–24

Egg-shaped anaerobic digester: (a) schematic diagram from Walker Process catalogue (b) pictorial view.

Egg-shaped digester mixing systems are similar to those for cylindrical tanks and consist of unconfined gas mixing, mechanical draft-tube mixing, or pumped recirculation mixing (see Fig. 13–25). Gas mixing is considered by some to be relatively ineffective in mixing the digester contents below the level of the injection nozzles. The mechanical draft tube and pumped recirculation mixing systems, however, are considered able to provide sufficient energy to mix even the sludge in the bottom cone of the digester. The mechanical draft-tube

Figure 13–25

Mixing systems for egg-shaped anaerobic digesters. (From Stukenberg et al., 1992.)



mixer, which can be operated in either an up- or down-pumping mode, also provides a positive means of mixing at the surface to control scum and foam (Stukenberg et al., 1992).

Recirculation mixing is generally more effective when the sludge is taken from the bottom and discharged near the gas-liquid interface or above the gas-liquid interface to break up scum that may have accumulated. Recirculation mixing is also effective for foam control in gas-mixed digesters.

Any or all of the mixing systems may be used, and all may be operated during any one day, although gas and mechanical draft-tube mixing are seldom used at the same time. Most digesters are fitted with a gas lance or hydraulic jet near the bottom of the cone to stir any accumulated grit.

A combination jet-pump draft-tube mixing system is also used that permits mixing in three zones of the digester. One jet pump is attached to the bottom of the centrally located vertical draft tube, and a second jet pump is attached to the top. In this configuration, the draft tube can function for pumping sludge upward or downward for periodic blending of bottom sludge and scum with the tank contents. A third pump is located at the vessel perimeter to create a swirling action. External recirculation pumps are also provided for sludge heating and additional circulation of the tank contents. The system is designed to circulate the tank volume 10 times per d (Clark and Ruehrwein, 1992).

Methods for Enhancing Sludge Loading and Digester Performance

Opportunities for enhancing the performance of anaerobic digesters include thickening the digester feed sludge or thickening a portion of the digesting sludge to increase the SRT. Recirculating a portion of the digested sludge and cothickening with untreated primary and waste sludge was reported originally by Torpey and Melbinger (1967). The solids concentration in the feed sludge improved and the performance of the digester, as measured by volatile solids destruction, increased significantly. The thickening system was installed at wastewater plants in New York City. In a study by Maco et al. (1998), the effects of thickening digested biosolids, either thickened separately or combined with prethickening of untreated sludge, increased the SRT of the digestion process and the production of biogas and decreased the hydraulic retention time.

The value of thickening the feed sludge to the digester is indicated by data presented in Table 13–29. For example, for a 15-d hydraulic retention time and an average TSS of 3 percent, the volatile solids loading factor in Table 13–29 is 1.4 $kg/m^3·d$. By improving the feed sludge TSS to 6 percent, the VSS loading can be increased to 2.9 $kg/m^3·d$, near the middle of the sludge loading range given in Table 13–28. In this hypothetical example, a doubling in digester capacity is achieved. In evaluating digested biosolids recycling to reduce the size and number of digesters or increase the sludge processing capacity of existing digesters, sludge rheology and sludge handling equipment require evaluation. While most digesters can accommodate increases in solids concentrations, the limits imposed by pumping and mixing systems require careful evaluation (Maco et al., 1998).

Sludge pretreatment prior to anaerobic digestion can enhance the solids loading and the digester performance. Pretreatment is practiced through adding energy in the form of mechanical, electrical, or ultrasonic. Combining heat with pressure to cause thermal hydrolysis has been practiced successfully. These methods of sludge pretreatment are presented later in this section.

Gas Production, Collection, and Use

Gas from anaerobic digestion contains about 65 to 70 percent CH_4 by volume, 25 to 30 percent CO_2, and small amounts of N_2, H_2, H_2S, water vapor, and other gases. Digester gas has a specific gravity of approximately 0.86 relative to air. Because production of gas is one of the best measures of the progress of digestion and because digester gas can be used as fuel, the designer should be familiar with its production, collection, and use.

Gas Production. The volume of methane gas produced during the digestion process can be estimated using Eq. (13–12), discussed previously. Total gas production is usually estimated from the percentage of volatile solids reduction. Typical values vary from 0.75 to 1.12 m^3/kg (12 to 18 ft^3/lb) of volatile solids destroyed. Gas production can fluctuate over a wide range, depending on the volatile solids content of the sludge feed and the biological activity in the digester. Excessive gas production rates sometimes occur during startup and may cause foaming and escape of foam and gas from around the edges of floating digester covers. In egg-shaped and shallow cylindrical digesters, foaming can clog the gas outlet unless foam control is provided. If stable operating conditions have been achieved and the foregoing gas production rates are being maintained, a well-digested sludge can be obtained.

Gas production can also be estimated crudely on a per capita basis. The normal yield is 15 to 22 $m^3/10^3$ persons·d (0.6 to 0.8 ft^3/person·d) in primary plants treating normal domestic wastewater. In secondary treatment plants, the gas production is increased to about 28 $m^3/10^3$ persons·d (1.0 ft^3/person·d).

Gas Collection. In cylindrical digesters, gas is collected under the cover of the digester. Three principal types of covers are used: (1) floating, (2) fixed, and (3) membrane. Floating covers fit on the surface of the digester contents and allow the volume of the digester to change without allowing air to enter the digester [see Fig. 13–26(a)]. Gas and air must not be allowed to mix, or an explosive mixture may result. Explosions have occurred in wastewater treatment plants. Gas piping and pressure-relief valves must include adequate flame traps. The covers may also be installed to act as gas holders for a limited storage of gas. High-rate digesters produce about two volumes of gas per volume of digester capacity/d (Speece 2001). Floating covers can be used for single-stage digesters or in the second stage of two-stage digesters.

Fixed covers provide a free space between the roof of the digester and the liquid surface [see Fig. 13–26(b)]. Gas storage must be provided so that (1) when the liquid volume is changed, gas, and not air, will be drawn into the digester; otherwise an overflow weir with a U-shaped trap needs to be provided to maintain a liquid seal, and (2) gas will not be lost by displacement. Gas can be stored either at low pressure in external gas holders that use floating covers or at high pressure in pressure vessels if gas compressors are used. Gas not used should be burned in a flare. Gas meters should be installed to measure gas produced and gas used or wasted.

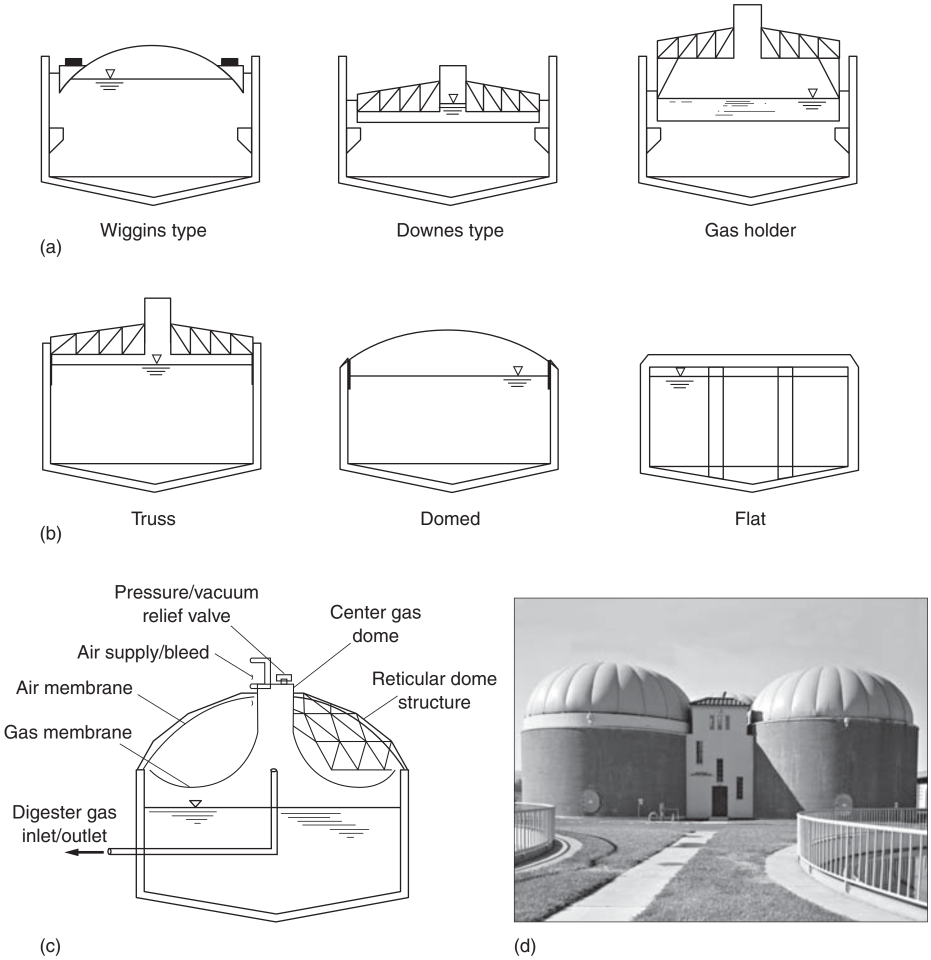
Another development in gas holder covers for cylindrical tanks is the membrane cover [see Fig. 13–26(c)]. This cover consists of a support structure for a small center gas dome and flexible air and gas membranes. An air-blower system is provided to pressurize the air space between the two membranes and vary the air space volume. Only the gas membrane and the center gas dome are in contact with the digester contents. The gas membrane is made from a flexible polyester fabric.

In egg-shaped digesters, the volume available for gas storage is small. For efficient utilization of digester gas, supplemental external storage may be required.

Gas Pretreatment and Use. Methane gas at standard temperature and pressure (20°C and 1 atm) has a lower heating value of 35,800 kJ/m^3 (960 Btu/ft^3). Lower heating value is the heat of combustion less the heat of vaporization of any water vapor present. Because digester gas is only 65 percent methane, the lower heating value of digester gas

Figure 13–26

Types of anaerobic digester covers: (a) floating, (b) fixed, and (c) and (d) schematic and view of membrane gas cover.



is approximately 22,400 kJ/m^3 (600 Btu/ft^3). By comparison, natural gas, which is a mixture of methane, propane, and butane, has a heating value of 37,300 kJ/m^3 (1000 Btu/ft^3). Digester gas can be used as a fuel for boilers to provide heat for digestion and other parts of the plant if surplus heat is available. Digester gas can also be used in cogeneration and when purified can be used as a natural gas substitute.

Because digester gas contains hydrogen sulfide, nitrogen, particulates, and water vapor, the gas frequently has to be cleaned before use. These impurities can significantly impact the operation and performance of equipment that utilizes the gas, especially cogeneration equipment. Proper design of a digester gas use system requires careful consideration of possible pretreatment requirements. The primary constituents of digester gas that require pretreatment are moisture, hydrogen sulfide, mercaptans, and siloxanes.

Pretreatment for Removing Moisture. Digester gas is typically saturated with water vapor. Common practice and regulatory codes require designing for removal of the moisture that condenses in the digester gas piping system using sediment traps and drip traps. Piping should be provided with a minimal slope of 10 mm/m (1/8-in./ft) to low points at which sediment traps or drip traps are located. Consideration should be given to access these devices for removal of the collected condensate. Automatic drip traps are available but are not allowed in some jurisdictions. A significant amount of the moisture in the gas will be condensed simply through the cooling in the piping system at ambient temperatures, which can then be removed by the appropriate traps. Long runs of piping or travel through other pretreatment systems can be advantageous for reducing the moisture content of the gas.

Some gas use systems may require additional removal of moisture from the gas. Chilling of the gas (reducing the dew point) is commonly used in systems where very dry gas is required. This treatment step is usually associated with systems where activated carbon is supplied for removal of other contaminants or where the gas use system requires a very dry gas. As cooling of the gas is costly to operate, this system should only be included after careful consideration of the costs and benefits obtained through that level of treatment.

Pretreatment for Removing Hydrogen Sulfide. Hydrogen sulfide can damage piping systems and digester gas use equipment. The hydrogen sulfide combines with the condensate and forms a weak sulfuric acid. Piping materials such as stainless steel or lined ductile iron pipe should be considered to resist the corrosion and erosion that can be caused by the slightly acidic condensate. Cement lined ductile iron pipe should not be used as the cement lining is readily destroyed by the condensate. Additional wall thickness ductile iron pipe has been used successfully providing an allowance for some corrosion and erosion.

A significant amount of the hydrogen sulfide is removed along with the condensate. This removal may be adequate for some digester gas uses like boilers, although even then the boiler may incur additional maintenance due to the presence of the hydrogen sulfide. Several systems are commonly used for removal of hydrogen sulfide from digester gas. Historically, iron sponge impregnated into wood chips has been used. The sulfide combines with the iron to form iron sulfide, a solid. These systems can also include methods to regenerate the iron in place.

There are several other commercially available processes for removal of hydrogen sulfide that include a liquid phase oxidation process, biological scrubbers or treatment systems, chemical systems, activated carbon, adsorptive resins, and others. Most, aside from the biological systems, resins, and activated carbon, use some form of iron chemistry. Systems that remove moisture through chilling will also remove a significant amount of the hydrogen sulfide in the condensate. However, consideration of condensate handling must be assessed. If returned to the treatment plant, consideration needs to be given to the ultimate fate of the hydrogen sulfide.

Pretreatment for Removing Siloxanes. Siloxanes are silicon containing volatile organic compounds. Siloxanes are used as carriers or conditioners in antiperspirants, skin care products, deodorants, liquid soaps, and hair care products and are now ubiquitous in the environment and wastewater. They are hydrophobic and tend to attach to the sludge produced during treatment. Once in an anaerobic digester, the conditions of mixing and heating tend to cause the siloxanes to volatilize into the digester gas. Wherever digester gas is burned, such as in a boiler, engine, or turbine, silicon dioxide (SiO_2), an inert white powder, is formed. Silica dioxide is the base material for glass, sandpaper, and grinding tools. The silicon dioxide can build up in the combustion equipment causing significantly increased maintenance or total equipment failure.

Sampling and testing procedures for siloxanes has not been standardized. Also, the allowable levels of siloxanes for gas use equipment vary by type and manufacturer of equipment. A proper design requires careful assessment and consideration of the siloxane composition of the gas and required cleaning for use. Sampling and testing program would be best for obtaining design information.

Siloxanes can be removed primarily in condensate through chilling or adsorption on a media or activated carbon. Return of the condensate to the treatment plant has been

Table 13–34

Typical electricity and heat generation efficiency from various co-generation systems[a]

Co–generation system	Electricity generation efficiency, %	Heat recovery efficiency,%
Internal combustion engine	37–42	35–43
Lean burn internal combustion engine	30–38	41–49
Conventional turbine	26–34	40–52
Recuperated Turbine	36–37	30–45
Microturbine	26–30	30–37
Molten carbonate fuel cell	40–45	30–40
Phosphoric acid fuel cell	36–40	NA

[a] Adapted from U.S. EPA (2010).

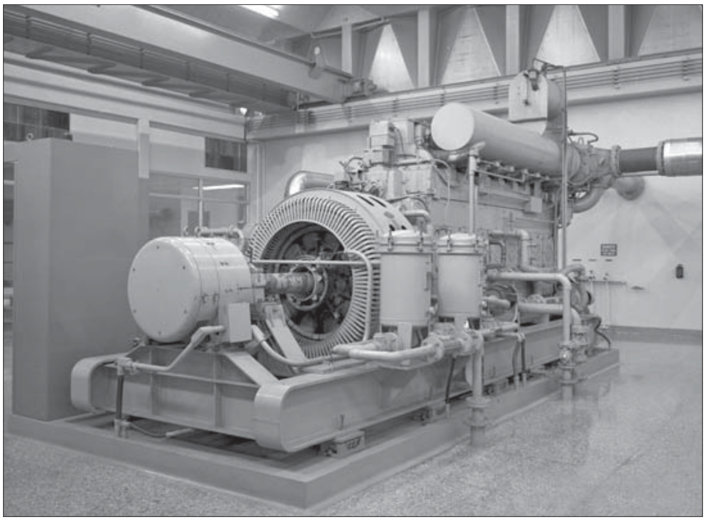
determined to be acceptable as a significant amount of the siloxane will be removed from the plant through volatilization from an aeration tank and in the final solids disposition. Media are regenerated through steam or other processes.

Digester Gas Use in Cogeneration. Cogeneration is generally defined as a system for generating electricity and producing another form of energy (usually heat in the form of steam or hot water). Cogeneration systems are also known as combined heat and power (CHP) systems. Most common at wastewater facilities are internal combustion engines or microturbines connected to generators. Several larger facilities also use turbines. Fuel cells are also used to create electricity with the heat recovered for process uses. Typical efficiencies for these processes are provided in Table 13–34. A typical internal combustion engine is shown on Fig. 13–27.

Design of a cogeneration system must consider variations in gas production, use of all of the gas or allowing occasional flaring of excess gas, and redundancy. These considerations are also impacted by the cost of electricity. It will be cost-effective to provide a more robust system when the electrical savings are greater due to higher electrical rates. Gas storage may also be considered when large variations in gas production are anticipated

Figure 13–27

Internal combustion engines for cogeneration at the Back River Wastewater Treatment facility in Baltimore, MD.



(due to industrial loadings or feed of supplemental organic material to the anaerobic digesters) or to maximize electricity generation at facilities where time of day electrical charges are significantly higher than off peak rates. The level of gas cleaning required for the cogeneration systems can vary widely and also needs to be considered in the system design.

The design of a cogeneration system must also consider air pollution impacts. Several jurisdictions have significant restrictions on combustion of fuels. Permitting aspects for cogeneration systems can impact the cost-effectiveness or ability to implement a specific system or device. Fuel cells have the lowest air emissions but the highest cost of the systems listed in Table 13–34.

Digester Gas Use as a Natural Gas. Digester gas is purified to enable its use as a substitute for natural gas. Besides the removal of H_2S, siloxanes, and water vapor, the additional removal of carbon dioxide (CO_2) is required to upgrade the digester gas to natural gas pipeline quality. Purified digester gas can have a methane content of 95 percent plus and can be modified slightly to meet pipeline quality methane for sale or use as natural gas. Purified digester gas can also be compressed to become compressed natural gas (CNG) for use as vehicle fleet fuel in natural gas burning engines. CNG plants can be useful in areas where air pollution is an issue and natural gas fleet vehicles are already in operation. Use of CNG can also be a substantial cost savings in areas where gasoline prices are higher. Several technologies are available for gas purification. The common technologies for gas purification are water adsorption, chemical adsorption, pressure swing adsorption (PSA), and cryogenic separation.

Digester Heating

The heat requirements of digesters consist of the amount needed (1) to raise the incoming sludge to digestion tank temperatures, (2) to compensate for the heat losses through walls, floor, and roof of the digester, and (3) to make up the losses that might occur in the piping between the source of heat and the tank. The sludge in digestion tanks is heated by pumping the sludge through external heat exchangers and back to the tank.

Analysis of Heat Requirements. In computing the energy required to heat the incoming sludge to the temperature of the digester, it is assumed that the specific heat of most sludges is essentially the same as that of water. The assumption that the specific heats of sludge and water are essentially the same has proved to be acceptable for engineering computations. The heat loss through the digester sides, top, and bottom is computed using the following expression:

$$q = UA\Delta T \tag{13–17}$$

where q = heat loss, J/s (Btu/h)
U = overall coefficient of heat transfer, J/m^2·s·°C (Btu/ft^2·h·°F)
A = cross-sectional area through which the heat loss is occurring, m^2 (ft^2)
ΔT = temperature drop across the surface in question, °C (°F)

In computing the heat losses from a digester using Eq. (13–17), it is common practice to consider the characteristics of the various heat transfer surfaces separately and to develop transfer coefficients for each one. The application of Eq. (13–17) in the computation of digester heating requirements is illustrated in Example 13–7.

Heat-Transfer Coefficients. Typical overall heat-transfer coefficients are reported in Table 13–35. As shown, separate entries are included for the walls, bottom, and top of the digester. Digestion tank walls may be surrounded by earth embankments that serve as

Table 13–35

Typical values for the overall coefficients of heat transfer for computing digester heat losses[a]

Item	U.S. customary, Btu/ft²·°F·h	SI units, W/m²·°C
Plain concrete walls (above ground)		
300 mm (12 in.) thick, not insulated	0.83–0.90	4.7–5.1
300 mm (12 in.) thick with air space plus brick facing	0.32–0.42	1.8–2.4
300 mm (12 in.) thick wall with insulation	0.11–0.14	0.6–0.8
Plain concrete walls (below ground)		
Surrounded by dry earth	0.10–0.12	0.57–0.68
Surrounded by moist earth	0.19–0.25	1.1–1.4
Plain concrete floors		
300 mm (12 in.) thick in contact with moist earth	0.5	2.85
300 mm (12 in.) thick in contact with dry earth	0.3	1.7
Floating covers		
With 35 mm (1.5 in.) wood deck, built–up roofing, and no insulation	0.32–0.35	1.8–2.0
With 25 mm (1 in.) insulating board installed under roofing	0.16–0.18	0.9–1.0
Fixed concrete covers		
100 mm (4 in.) thick and covered with built–up roofing, not insulated	0.70–0.88	4.0–5.0
100 mm (4 in.) thick and covered, but insulated with 25 mm (1 in.) insulating board	0.21–0.28	1.2–1.6
225 mm (9 in.) thick, not insulated	0.53–0.63	3.0–3.6
Fixed steel covers 6 mm (0.25 in.) thick	0.70–0.95	4.0–5.4

[a] Adapted in part from U.S. EPA (1979).

insulation, or they may be of compound construction consisting of approximately 300 mm (12 in.) of concrete, insulation, or an insulating air space, plus brick facing or corrugated aluminum facing over rigid insulation. The heat transfer from plain concrete walls below ground level and from floors depends on whether they are below the groundwater level. If the groundwater level is not known, it may be assumed that the sides of the tank are surrounded by dry earth and that the bottom is saturated earth. Because the heat losses from the tank warm up the adjacent earth, it is assumed that the earth forms an insulating blanket 1.5 to 3 m (5 to 10 ft) thick before stable ambient earth temperatures are reached. In northern climates, frost may penetrate to a depth of 1.2 m (4 ft). Therefore, the ground temperature can be assumed to be 0°C (32°F) at this depth and to vary uniformly above this depth to the design air temperatures at the surface. Below the frost depth, normal winter ground temperatures can be assumed, which are 5 to 10°C (10 to 20°F) higher at the base of the wall. Alternatively, an average temperature may be assumed for the entire wall below grade.

The loss of heat through the roof depends on the type of construction, the absence or presence of insulation and its thickness, the presence of air space (as with floating covers between the skin plate and the roofing), and whether the underside of the roof is in contact with sludge liquor or gas.

Radiation from roofs and aboveground walls also contributes to heat losses. At the temperatures involved, the effect is small and is included in the coefficients normally used, such as those given in the foregoing discussion. For the theory of radiant-heat transmission, the reader is referred to McAdams (1954). Heat requirements for a digester are determined in Example 13–7.

When external heaters are installed, the sludge is pumped at high velocity through the tubes while water circulates at high velocity around the outside of the tubes. The circulation promotes high turbulence on both sides of the heat transfer surface and results in higher heat transfer coefficients and better heat transfer. Another advantage of external heaters is that untreated cold sludge on its way into the digesters can be warmed, intimately blended, and seeded with sludge liquor before entering the tank. Heat exchangers require cleaning periodically to maintain heat transfer efficiency.

Digestion tanks have also been heated using internal heating systems. Some arrangements have included pipes mounted to the interior face of the digester wall and mixing tubes equipped with hot-water jackets. Because of inherent operating and maintenance problems with this type of heating system, internal heating is not recommended. Reported problems include caking of sludge on the heating surface and the inability to inspect or service the equipment unless the tank is dewatered (WEF, 1987).

EXAMPLE 13–7 **Estimation of Digester Heating Requirements** A digester with a capacity of 90,700 wet kg/d (200,000 lb/d) of thickened sludge is to be treated by circulation of sludge through an external hot water heat exchanger. Assuming that the following conditions apply, find the heat required to maintain the required digester temperature. If all heat were shut off for 24 h, what would be the average drop in temperature of the tank contents?

1. Concrete digester dimensions:
   Diameter = 20 m
   Side depth = 7 m
   Middepth = 10 m
2. Heat-transfer coefficients:
   Dry earth embanked for entire depth, $U = 0.68\ \text{W/m}^2\cdot{}^\circ\text{C}$
   Floor of digester in moist earth, $U = 2.85\ \text{W/m}^2\cdot{}^\circ\text{C}$
   Fixed Concrete Insulated Roof exposed to air, $U = 1.5\ \text{W/m}^2\cdot{}^\circ\text{C}$
3. Temperatures:
   Air = −5°C
   Earth next to wall = 0°C
   Incoming sludge = 10°C
   Earth below floor = 5°C
   Sludge contents in digester = 35°C
4. Specific heat of sludge 4200 J/kg·°C

Solution

1. Compute the heat requirement for the sludge.

$$q = (90{,}700\ \text{kg/d})[(35 - 10)^\circ\text{C}](4200\ \text{J/kg}\cdot{}^\circ\text{C})$$

$$= 95.2 \times 10^8\ \text{J/d}$$

2. Compute the area of the walls, roof, and floor.

   Wall area $= \pi(20)\,(7) = 439.6\ \text{m}^2$

   Floor area $= \pi(10)\,[10^2 + (10 - 7)^2]^{1/2} = 327.8\ \text{m}^2$

   Roof area $= \pi(10^2) = 314\ \text{m}^2$

3. Compute the heat loss by conduction using Eq. (13–17).

   $q = UA\Delta T$

   a. Walls:

   $$q = 0.68\ \text{W/m}^2\cdot{}^\circ\text{C}\ (439.6\ \text{m}^2)(35 - 0^\circ\text{C})(86{,}400\ \text{s/d}) = 9.0 \times 10^8\ \text{J/d}$$

   b. Floor:

   $$q = 0.85\ \text{W/m}^2\cdot{}^\circ\text{C}\ (268.2\ \text{m}^2)(32 - 5^\circ\text{C})(86{,}400\ \text{s/d}) = 5.32 \times 10^8\ \text{J/d}$$

   $$q = 2.85\ \text{W/m}^2\cdot{}^\circ\text{C}\ (327.8\ \text{m}^2)(35 - 5^\circ\text{C})(86{,}400\ \text{s/d}) = 24.2 \times 10^8\ \text{J/d}$$

   c. Roof:

   $$q = 1.5\ \text{W/m}^2\cdot{}^\circ\text{C}\ (314\ \text{m}^2)[35 - (-5^\circ\text{C})](86{,}400\ \text{s/d}) = 16.2 \times 10^8\ \text{J/d}$$

   d. Total losses:

   $$q_t = (9.0 + 24.2 + 16.2) \times 10^8\ \text{J/d} = 49.4 \times 10^8\ \text{J/d}$$

4. Compute the required heat-exchanger capacity.

   Capacity = heat required for sludge and heat required for digester

   $$= (95.2 + 49.4) \times 10^8\ \text{J/d} = 144.6 \times 10^8\ \text{J/d}$$

5. Determine the effect of heat shutoff.

   a. Digester volume $= \pi\left(\frac{D^2}{4}\right)h_s + \pi\left(\frac{D^2}{12}\right)h_c$

   $$= \pi\left(\frac{20^2}{4}\right)(7) + \pi\left(\frac{20^2}{12}\right)(10 - 7) = 2198 + 314$$

   $$= 2512\ \text{m}^3$$

   b. Weight of sludge $= (2512\ \text{m}^3)(10^3\ \text{kg/m}^3)$

   $$= 2.51 \times 10^6\ \text{kg}$$

   c. Drop in temperature $= \dfrac{(144.6 \times 10^8\ \text{J/d})(1\text{d})}{(2.51 \times 10^6\ \text{kg})(4200\ \text{J/kg}\cdot{}^\circ\text{C})} = 1.37^\circ\text{C/d}$

Heating Equipment. The contents of the digester can be heated by tube-in-tube, spiral-plate, or water-bath external heat exchangers. The tube-in-tube and spiral-heat exchangers are similar in design. A tube-in-tube exchanger consists of two concentric pipes, one containing the circulating sludge and the other containing hot water. Flow through the pipes is countercurrent. Spiral-plate heat exchangers [see Fig. 13–28(a) and (b)] are composed of two long strips of plate that are wrapped to form a pair of

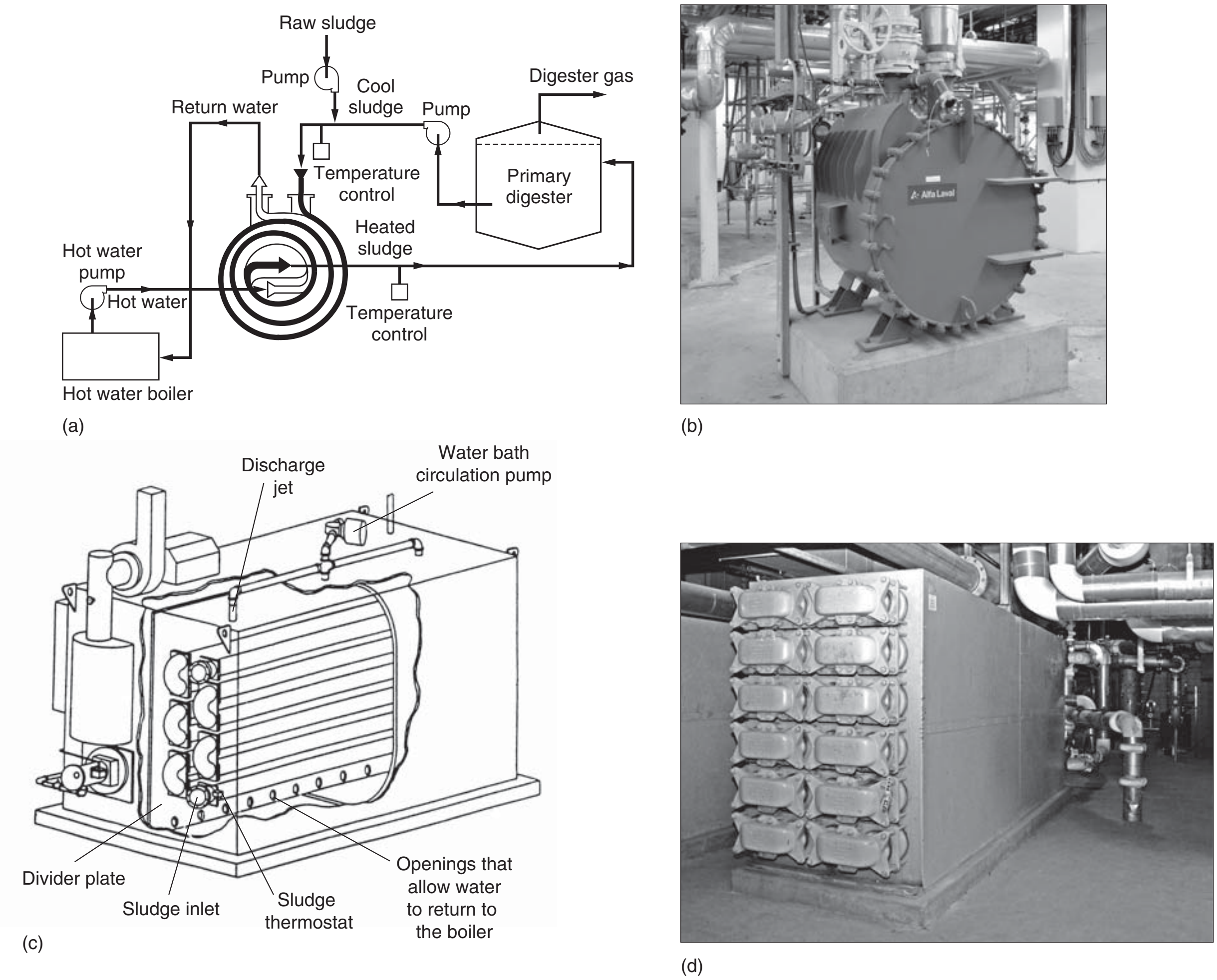


Figure 13–28

Heat exchangers used for heating digesting sludge: (a) schematic diagram of a spiral type, (b) view of a spiral type and (c) schematic of a water bath type heat exchanger, and (d) view of a water bath type heat exchanger.

concentric passages. The flow regime is also countercurrent. Water temperatures are kept generally below 68°C (154°F) to prevent caking of the sludge. Heat-transfer coefficients for external heat exchangers range from 0.9 to 1.6 $W/m^2 \cdot °C$ (WEF 2010a).

Operation of a water-bath heat exchanger involves circulation of the sludge through a heated water bath [see Fig. 13–28(c) and (d)]. The heat transfer rate is increased by pumping hot water in and out of the bath. Recirculation pumps allow the sludge feed to be heated before introduction to the digester.

Boilers and cogeneration systems are used typically to supply heat to the circulating water in the heat exchangers. Boilers can be fueled by digester gas; however, natural gas or fuel oil may be used as auxiliary fuel for times when sufficient digester gas is not available, such as for digester startup. If a cogeneration system is provided that uses digester gas to fuel an internal-combustion engine for generating electricity or powering pumps or blowers, heat from the engine jacket water can be used in the heat exchanger.

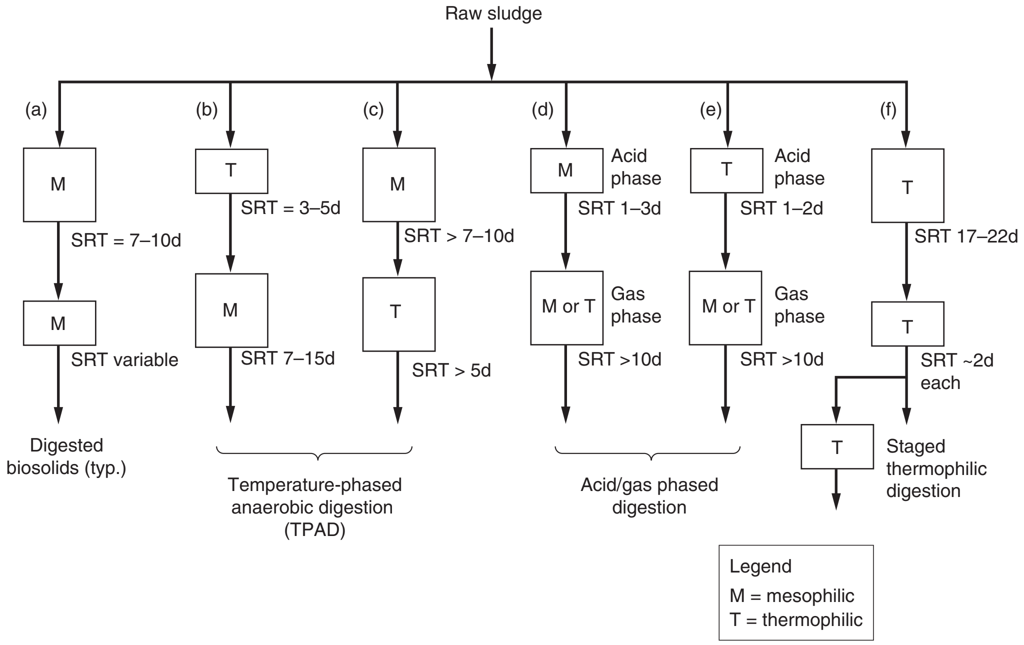


Figure 13–29

Options for staged anaerobic digestion: (a) staged mesophilic digestion, (b) temperature phased thermophilic-mesophilic digestion, (c) temperature phased mesophilic-thermophilic digestion, (d) acid/gas phased digestion with mesophilic acid-phase, (e) acid/gas phased digestion with thermophilic acid phase, and (f) staged thermophilic digestion. (Adapted from Schafer and Farrellm 2000 and Moen, 2000.)

Advanced Anaerobic Digestion

Advanced anaerobic digestion processes were developed to increase the volatile solids reduction during anaerobic digestion and/or produce high quality Class A biosolids for beneficial use of the generated biosolids. Among the advanced anaerobic digestion processes are the thermophilic digestion, staged thermophilic digestion, staged mesophilic digestion, acid/gas phase digestion and temperature phased digestion. These digestion options are shown on Fig. 13–29 and are discussed below. Typical SRTs are also noted on Fig. 13–29.

Thermophilic Anaerobic Digestion. Thermophilic digestion occurs at temperatures between 50 and 57°C (120 and 135°F), conditions suitable for thermophilic bacteria. Because biochemical reaction rates increase with temperature, doubling with every 10°C (18°F) rise in temperature until a limiting temperature is reached, thermophilic digestion is much faster than mesophilic digestion. Advantages and disadvantages of thermophilic digestion when compared to mesophilic digestion are provided in Table 13–36. Single-stage thermophilic digesters have been used only in limited applications; for municipal sludge treatment, they have been mainly used as the first stage of a temperature-phased anaerobic digestion process (Moen, 2000).

Although there may be greater reductions in pathogens in thermophilic digestion than in mesophilic digestion, U.S. federal regulations controlling land application of biosolids do not classify thermophilic digestion as a process to further reduce pathogens (PFRP). Both mesophilic and thermophilic digestion are classified as processes to significantly

Table 13–36

Advantages and disadvantages of thermophilic anaerobic digestion as compared to. mesophilic anaerobic digestion

Advantages	Disadvantages
1. Improved pathogen destruction, Class A sludge production is possible[a]	1. If not batch treatment process will require EPA certification for Class A PFRP
2. Reaction rate is increased which can reduce volume requirements (capital savings)	2. Increased thermal energy requirements[b]
3. May improve overall VSR and increase digester gas production	3. Biosolids may not dewater as well[c]
4. Components and design are essentially the same as conventional mesophilic digesters	4. Higher odor potential in dewatered cake[c]
	5. Increased ammonia concentration in dewatering sidestream
	6. Process may not be as stable
	7. More complex system due to heat recovery requirements
	8. May be more susceptible to foaming

[a] To meet Class A with thermophilic digestion, the process must incorporate batch thermophilic tanks that can meet the time/temperature requirements under Alternative 1. Otherwise, the process will require site specific testing under Alternative 3.

[b] Many plants incorporate a heat recovery loop where thermophilic sludge is cooled and the energy is utilized to preheat the raw sludge. Incorporating the heat recovery loop reduces thermal energy requirements but increases the cost and complexity of the digestion system.

[C] To improve dewatering and reduce odor potential many plants incorporate a mesophilic digestion stage prior to dewatering.

reduce pathogens (PSRP). Therefore, single-stage thermophilic digestion has significant limitations, as cited above.

Staged Thermophilic Digestion. A staged thermophilic digestion process [see Fig. 13–29(f)] uses a large reactor followed by one or more smaller reactors to reduce pathogen short circuiting and achieve a Class A sludge. At the Annacis Island Wastewater Treatment Plant in Vancouver, BC, the first stage is followed by three subsequent stages. Volatile solids reductions for the digestion system are reported to be on the order of 63 percent (Schafer and Farrell, 2000b).

Staged Mesophilic Digestion. Although digestion performed in two tanks coupled in series has been done in the past, little information is available about the operation of two-stage heated and mixed high-rate digesters. Researchers Torpey and Garber found that there were few benefits in volatile solids reduction and gas production in two series tanks as compared to a single-stage high-rate process (Torpey and Melbinger, 1967; Garber, 1982). More recent testing indicates that two-stage mesophilic digestion may produce more stable, less odorous biosolids that are easier to dewater (Schafer and Farrell, 2000a). Staged mesophilic digestion is shown on Fig. 13–29(a).

Acid/Gas Phased Digestion. In the acid/gas (AG) digestion process, anaerobic digestion proceeds through the three distinct phases of digestion described earlier—hydrolysis, fermentation (acidogenesis), and methanogenesis—but the process is divided into two separate steps. In the first stage, known as the acid phase digester, solubilization of particulate matter occurs (hydrolysis), and volatile acids are formed (acidogenesis). The first stage is conducted at a pH of 6 or less and at a short SRT conducive to the production of high concentrations of volatile acids (> 6000 mg/L). The second stage, known as the gas phase, is conducted at a neutral pH and a longer SRT to suit the environmental conditions for the methane-generating bacteria and maximize gas production. Advantages of this method of digestion are (1) greater volatile solids reduction can be achieved, (2) digester foaming can be controlled, and (3) either stage can be operated at mesophilic or thermophilic temperatures [see Figs. 13–29(d) and (e)]. More than 30 full-scale plants using the AG process are in operation at the time of writing of this text (2012) (Wilson et al., 2008). Total volatile solids reductions range from 50 to 60 percent. Most acid/gas systems produce Class B biosolids and operate in the mesophilic range for both phases and termed AGMM. However, at the Belmont Wastewater Plant in Indianapolis, IN, where a thermophilic acid phase and mesophilic gas phase system was pilot tested, it was found that the process was effective in meeting Class A requirements for pathogen reduction (Schafer and Farrell, 2000a).

The design of an AG digestion system requires control of the organic loading to the acid phase to prevent the formation of methanogens. Control is provided through control of the detention time. Ideal detention time is between 1 to 2 d. Due to the low SRT, volatile solids leading rates for the acid phase digestion are an order of magnitude greater than conventional digestion and range from 24 to 40 kg VS/m^3·d (1.5 to 2.5 lb/d·ft^3) (WEF, 2012). The methane phase can then be a little as 10 d. Regulatory approval of this short a detention time maybe required to retain Class B digested sludge quality. The original concept for this process recommended a plug flow acid phase reactor. Current designs have used a tall cylindrical tank with level control to enable good mixing while maintaining the high organic loading. There will be very little gas produced in this process and the gas will contain little, if any, methane. This gas may be burned separately or combined with the methane phase digester gas, but is often wasted and not otherwise used. Counterintuitive to typical biological processes, operation requires increasing the loading (by decreasing the detention time) when the process begins to become less acidic and/or methane begins to be produced.

Temperature-Phased Digestion. Temperature-phased anaerobic digestion (TPAD), shown on Figs. 13–29(b) and (c), was developed in Germany and is an approach that incorporates the advantages of thermophilic digestion and mitigates the disadvantages through the addition of a mesophilic phase that enhances stabilization. The design of the temperature-phased process utilizes the advantage of the greater thermophilic digestion rate, which generally is four times faster than mesophilic digestion. The TPAD process has shown the capability for absorbing shock loadings better, as compared to single-stage mesophilic or thermophilic digestion. The process can operate in either of two modes, thermophilic-mesophilic or mesophilic-thermophilic. In the thermophilic-mesophilic mode, shown on Fig. 13–29(b), the thermophilic phase is designed to operate at 55°C (130°F) with a 3 to 5 d detention time. The mesophilic phase is designed to operate at 35°C (95°F) with a 10 d or greater detention. The total average detention time of 15 d compares to the typical 10 to 20 d range of the single-stage high-rate mesophilic digestion process. The volatile suspended solids (VSS) destruction efficiencies of the TPAD process are on the order of 15 to 25 percent greater than single-stage mesophilic digestion (Schafer and Farrell, 2000b).

Through greater hydrolysis and biological activity in the thermophilic phase, the system tends to have greater VSS destruction and gas production. Foaming is also reduced. The mesophilic phase provides additional VSS destruction and conditions the sludge for further processing. The main advantages of the mesophilic phase are (1) the destruction of odorous compounds (mostly fatty acids) that are common to the thermophilic digestion process and (2) the improved stability of the digestion operation. The process is also reported to be capable of meeting Class A sludge requirements (WEF, 2010a).

A second temperature-phased digestion process shown on Fig. 13–29(c) has a mesophilic stage that precedes the thermophilic stage. Limited results from full-scale and pilot testing show that the volatile solids reduction is greater than that from single-stage mesophilic digestion (Schafer and Farrell, 2000b). Design considerations for the temperature-phased anaerobic digestion process include selection of the heating and mixing systems to ensure proper temperature control of each stage, sizing of the gas-handling equipment to meet the greater gas production rates, and control of the pumping systems for digester feed and heating (WEF, 2010a).

Sludge Pretreatment for Anaerobic Digestion

Pretreatment of sludge prior to anaerobic digestion is used to increase the solids loading, increase the volatile solids reduction, increase biogas production and in some cases produce Class A biosolids. Sludge pretreatment results in increased hydrolysis through the application of some form of energy to the sludge. The form of pretreatment can be chemical, physical, electrical, or thermal. This section discusses two main categories of pretreatment: thermal hydrolysis and physical-chemical and electrical pretreatment.

Thermal Hydrolysis Pretreatment. Thermal hydrolysis (TH) is a thermal conditioning process that operates at lower temperatures in the range 150–200°C and functions as a pretreatment step before anaerobic digestion. The cited benefits of the process include (1) break down of longer organic polymer chains to shorter chain organic matter to increase digestion and gas production, (2) production of Class A product under the time and temperature stipulations of the EPA Part 503 biosolids rule, (3) enhancement of digested biosolids dewatering characteristics achieving in many cases greater than 30 percent cake solids with mechanical dewatering, (4) produce good quality product in terms of odor and texture, and (5) significant reduction of digestion volume through reducing treated sludge viscosity allowing higher solids concentrations to be pumped and mixed during digestion.

At present (ca. 2013), two commercially available TH processes are available; one is called Cambi™ and is provided by Cambi AS, Norway; the other is called Exelys™ and is provided by Veolia Water Systems, France. The two systems operate under the same treatment parameters; however, the first system is batch based, and the second system is a plug flow design that is a modification of Veolia's batch TH process (Biothelys™). The Cambi™ system is the most widely used system for thermal hydrolysis and is considered fully developed system; however, comparison with other systems may be warranted (Abu-Orf and Goss, 2012). The focus of the information in this section is mainly on the batch TH system as provided by Cambi™.

Description of the TH Batch System as Provided by Cambi™. The Batch TH process is a well established process outside North America with more than two dozen installations (2012) processing sludge from 3.3 to 250 dry tonne/d (3.6 DT/d to 275 DT/d). The process is gaining interest in North America with the first installation expected at Blue Plains Advanced Wastewater Treatment Plant in 2014. The process appears to be cost effective and meets the future needs of Blue Plains (Abu-Orf et al., 2009). As illustrated on Fig. 13–30, the thermal hydrolysis step consists of three basic units: the pulper, the reactor and the flash tank that defines a process train. In reality there is more to the process train than those three units. The sludge needs to pass through some type of screening process ahead of the TH to remove damaging materials from the sludge stream. Furthermore, the sludge must be pre-dewatered to slightly higher than 16 percent total solids. The dewatered sludge is then transferred to a silo or bin that is large enough to provide equalization,

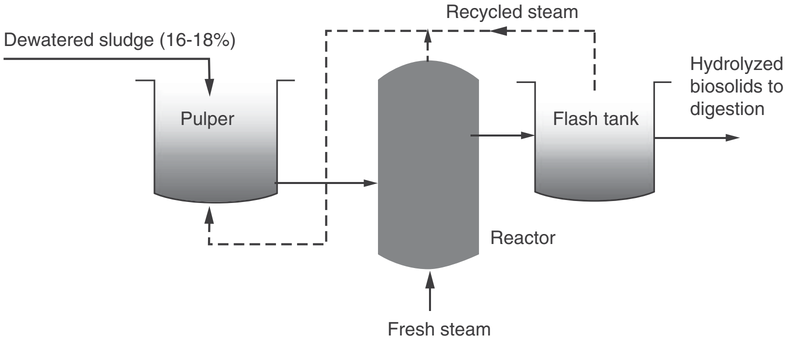


Figure 13–30

Main components of batch thermal hydrolysis system.

allowing the TH process to operate at a uniform flowrate. The sludge can then be transferred to the pulper tank via augers in the bottom of the storage bin and pumps. The concentration of solids to the pulper is diluted to a concentration between 14.5 and 16.5 percent TS using dilution water (typically plant effluent). The sludge is mixed in the pulper and preheated using steam that is recycled back from the flash tank. The preheated sludge is pumped to the reactor vessel where it will be heated using fresh stream to about 165°C (329°F) and a pressure of approximately 8–9 bar (120–130 $\text{lb}_f/\text{in.}^2$) gauge After the appropriate hydrolyzing time, the sludge is transferred from the reactor to the flash tank. The flash tank receives treated sludge with a solids concentration about 3 points less due to the steam injection and dilution.

The treated sludge temperature at this point is too hot to feed a mesophilic digester and requires cooling and dilution if needed. Again, the water used for dilution is required to lower the concentration of the treated sludge to between 8 and 12 percent for feeding into the digester. The temperature must also be reduced to about 42–44°C by heat transfer. The sludge at 8 to 12 percent TS is then transferred to the digestion facility. Finally, the digested biosolids are transferred to final dewatering before being distributed for beneficial reuse. The final product has very good stacking properties and is low in odor, making it very suitable for soil blending and for land application. In addition to the process train, there must be a source of fresh steam for heating the reactors, so there will be at least a boiler as part of the train. Usually, the biogas generated from the digestion process is processed via a combined heat and power (CHP) facility for producing electricity and a majority of the necessary steam for the TH process. The amount of waste heat from the CHP system typically provides 75–95 percent of the total steam energy requirements so a supplemental standby boiler that operates on digester gas or an auxiliary fuel source is required.

Reactor Operation Scheme. The TH reactors operate in a batch mode with a typical volume of 12 m^3 (424 ft^3) and receive approximately 7.6 m^3 (268 ft^3) ($\sim$70 percent of total volume) of sludge in each batch. Once the sludge is transferred, approximately one tonne of fresh steam is injected per one dry tonne of sludge (from Cambi™ Specifications). After the reaction period elapses, the sludge is released to the flash tank. When the steam is released from the reactors, it passes to the pulper and preheats the incoming sludge. The pulper and the flash tanks are usually sized twice the size of one reactor. The batch steps for a typical 90 min cycle for each reactor are shown in Table 13–37. The theoretical amount of steam required can be estimated by the following equation:

$$\frac{M_{\text{steam}}}{M_s} = \frac{\left(C_{\text{PS}} + \dfrac{C_{\text{PW}}}{W_s} - C_{\text{PW}}\right)(T_H - T_{\text{raw}})}{H - C_{\text{PW}}(T_H - T_{\text{ref}})} \tag{13–18}$$

Table 13–37

Steps for 90 min batch cycle operation

Step	Action	Time, min	Description
1	Fill	15	Fill reactor with 7.6 m^3 of sludge
2	Steam injection	15	Inject steam in the Reactor
3	React	30	Hold reactor at 160°C and 620 kP (90 $lb/in.^2$)
4	Steam out	15	Release steam to pulper
5	Empty	15	Transfer sludge to flash tank by pressure release.

where

M_{steam} = mass of live steam fed to the process

C_{PS} = specific heat of dry sludge fraction, 1.5 kJ/kg°C (0.36 Btu/lb·°F)

C_{PW} = specific heat of water, 4.18 kJ/kg·°C (1 Btu/lb·°F)

T_H = temperature out of the thermal hydrolysis system. For Cambi™, this temperature corresponds to the temperature out of the flash tank which is approximately 105–110°C (220–230°F)

T_{raw} = raw sludge temperature, typically 10–25°C (50–77°F)

H = enthalpy of steam which is approximately 2785 kJ/kg (1200 Btu/lb) at 12 bar (175 $lb_f/in.^2$)

T_{ref} = reference temperature, typically 0°C (32°F)

For example, sludge at 10°C that is dewatered to 16 percent TS by weight would require the following amount of steam for the hydrolyzed sludge to reach a temperature of 110°C:

$$\frac{M_{\text{stream}}}{M_s} = \frac{\left[(1.5\ \text{kJ/kg}\cdot{}^\circ\text{C}) + \dfrac{(4.18\ \text{kJ/kg}\cdot{}^\circ\text{C})}{0.16} - (4.18\ \text{kJ/kg}\cdot{}^\circ\text{C})\right][(110 - 10)^\circ\text{C}]}{(2785\ \text{kJ/kg}) - (4.18\ \text{kJ/kg}\cdot{}^\circ\text{C})[(110 - 0)^\circ\text{C}]}$$

$$= 1.0\ \text{kg steam/kg sludge}$$

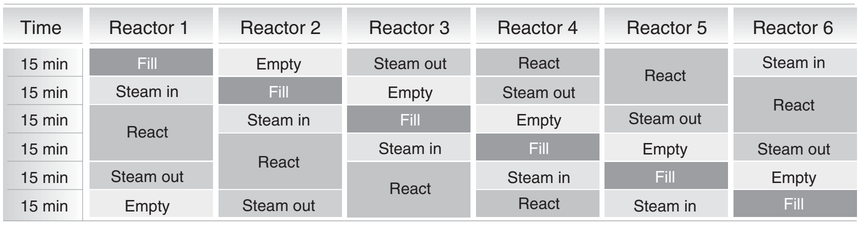
It should be noted that Eq. (13–18) is based on some simplifying assumptions such as no heat loss and that all flash steam is recovered and condensed successfully in the pulper. In reality some losses would be expected, and sizing for a design should be conducted carefully with the vendor.

Reactor System Sizing. The overall system sizing is based on a single reactor capacity. Usually it is assumed that sludge is delivered to the reactor at 14.7 percent. The reactor capacity is based on the cycle time and the concentration of the sludge in the reactor. For a standard 90 min cycle as described in Table 13–37 and 7.6 m^3/batch of sludge at a 14.7 percent solids concentration, the capacity of a single reactor is about 17.88 dry tonne/d (19.66 dry ton/d). Design size of the reactor is usually based on 95 percent reactor availability. The throughput capacity of train is dependent on the number of reactors within each train. Single reactor capacity can be increased by increasing the solids concentration (maximum would be 17 to 17.5 percent) or shortening the cycle time without compromising the reaction time that is necessary to achieve Class A biosolids according to Part 503 Regulations. In addition, Cambi™ is currently (2012) developing alternative reactor sizes allowing the system to be tailored to sizes at small to medium sized plants. An aerial view of a Cambi™ process installation in Davyhulme, Manchester UK, in which the thermal hydrolysis reactors are visible is shown on Fig. 13–31(a).

Figure 13–31

The Cambi™ process (a) Aerial view of installation in Davyhulme, Manchester, UK. The Cambi™ reactors are visible in the center of the photograph. The 8 × 7600 m^3 digesters, shown on the lower right, which originally processed 40,000 tonne/y of dry solids now process 92,000 tonne/y with the installation of the Cambi™ thermal hydrolysis pretreatment process (courtesy of Cambi™) and (b) schematic of the operation of a six reactor process, based on a 90 min cycle time.

(a)



Time	Reactor 1	Reactor 2	Reactor 3	Reactor 4	Reactor 5	Reactor 6
15 min	Fill	Empty	Steam out	React	React	Steam in
15 min	Steam in	Fill	Empty	Steam out		React
15 min	React	Steam in	Fill	Empty	Steam out	
15 min		React	Steam in	Fill	Empty	Steam out
15 min	Steam out		React	Steam in	Fill	Empty
15 min	Empty	Steam out		React	Steam in	Fill

(b)

Reactor Sequence Operation. The preferred smallest Cambi process train will have two reactors; otherwise it will be a batch operation. With two reactors the train works as a continuous operation. Based on solids residence of 90 min within each reactor, a sketch portraying the sequential status of a six reactor Cambi™ process, as selected by the Blue Plains Advanced Wastewater Treatment Plant, is shown on Fig. 13-31(b) (Abu-Orf et al., 2009). Process staggering allows continuous operation of all major mechanical equipment.

As shown on Fig. 13-31(b), one reactor would always be filling, one would always be receiving steam, one would always be releasing steam, and one would always be emptying treated sludge. Based on this operating sequence the sludge feed pumps and the steam plant would always be operating and that only valves would be opening and closing to direct the flow to and from the reactors.

Physical, Chemical, and Electrical Pretreatment. These pretreatment processes are generally applied to sludge produced from secondary treatment process as these sludges typically do not digest well anaerobically. Pretreatment of these sludges is accomplished through application of ultrasonic waves, mechanical shear, electrical pulse, pressure drop, or electrical field. The application of these different treatment methods has resulted in various degrees of success in enhance sludge digestion. For the application of the pretreatment process to be practically effective the amount of sludge entering the digester from secondary treatment must be more or at least same as the sludge produced from primary treatment.

A description of six sludge pretreatment technologies is presented in Table 13–38. Pretreatment technologies use pulse power, pressure drop combined with mechanical

Table 13–38

Description of commercially available physical, chemical and electrical technologies for pretreatment of secondary sludges prior to anaerobic digestion[a]

Technology/ manufacturer	Description	Advantages	Disadvantages
OpenCel/OpenCel, USA	Pulsed Power Technology, as offered by Open Cel®, exposes biological sludge to high voltage bursts between 20 and 100 microseconds to lyses cell membranes	Relatively low energy requirements Imparts usable heat to secondary sludge Low space requirements Low pressure operation	Relatively new Limited number of installations in North America Offered by only one manufacturer
Crown Disintegration/ Siemens, USA	Pretreatment is applied to only a portion of the secondary sludge, which includes grinding and mixing followed by pressurization. The high pressure drop causes cavitation of the sludge and rupturing of the cell membrane	Similar technology offered by other manufacturers Low space requirements Approximately 20 installations worldwide, all in Europe. Small footprint	Relatively high energy requirements. No operating facilities in North America. High pressure system 1200 kPa (175 $lb_f/in.^2$) with wear parts
Sludge Squeezer/ Huber, USA	Technology imparts a high pressure drop to a portion of the secondary sludge in a two stage process. In the first stage the sludge flocs are mechanically ruptured. In the second stage the flocs are mixed into the sewage sludge via a hydrodynamic flow field and homogenized	Similar technology offered by other manufacturers Low space requirements Approximately 3 installations, all in Europe	Relatively high energy requirements No operating facilities in North America High pressure system 1200 kPa (175 $lb_f/in.^2$) with wear parts
MicroSludge/ MicroSludge, Canada	A portion of the secondary sludge is pretreated with lime to soften cell membrane, then undergoes grinding and mixing followed by pressurization up to 1200 kPa (175 $lb_f/in.^2$). When the pressure is released, the biological cells are exposed to high shear forces which is hypothesized to rupture the cell membranes. The process includes course and fine screens for the thickened sludge and conditioned sludge, respectively, as well as a gas liquid separator to release ammonia gas formed at high pH	Small footprint Reduction in dewatering cost (polymer and electrical) Benefit for digester heating from 45°F rise of processed sludge Reduced digester mixing energy due to reduced viscosity and volume	Relatively high energy requirements. Requires lime addition High pressure system 1200 kPa (175 $lb_f/in.^2$) with wear parts. No operating facilities in North America
Sonolyzer/ Ovivo, USA	Based on ultrasonic treatment of secondary sludges, which consists of applying high frequency sound waves to the sludge matrix, causing cavitation and disintegration of the cell membranes	Intensively studied over the last 15 y. >25 installations worldwide	Relatively high energy requirements. Only one manufacturer currently in the North America
Electrokinetic Disintegration/ Sud-Chemie AG, Germany	The sludge is run through a series of pipes containing an internal electrical high voltage field and as it moves, the cellular structure is weakened and cracked which allows the bacteria to more effectively digest the sludge	Optimizes digestion Increase gas production Increase settling of sludge Several installations in Europe	High energy requirement Operation with high voltage

[a] The information in this table is current ca. 2013.

Table 13–39

Factors favoring direct co-digestion of organic feedstocks (WEF, 2010b)[a]

Category	Description
Technical	• Remove nuisance wastes from the collection system, especially if a waste is causing stoppage, odor or damage. • Remove organic loadings and nuisance factors from headworks and liquid treatment train. • Increase use of existing digester capacity, especially with co-digestion of wastes that are synergistic with wastewater sludge in terms of increasing the volatile solids loading rate. • Improve knowledge of how to handle organic wastes. • Provide a reliable outlet for organic wastes.
Economical	• Develop a new revenue stream from tipping fees for organic wastes. • Produce more biogas for combined heat and power systems, or thermal dryer systems, or other beneficial uses. • Reduce cost of operation, maintenance, and odor control in the liquid treatment train, from headworks to final clarifiers. • Avoid or defer construction of additional liquid train treatment capacity. • Increase the throughput rate of the sludge processing train.
Environmental	• Earn carbon credits, where applicable. • Reduce land application of organic wastes that contribute to methane production rather than carbon sequestration. • Reduce emission of greenhouse gases, particularly methane, coincidental to increasing energy recovery from waste materials.

[a] Adapted from WEF (2010b).

disintegration or chemical treatment, ultrasonic waves, and electrokinetic disintegration. These technologies have different advantages and disadvantages (see Table 13–38). One important observation is that these pretreatment technologies are used more widely in Europe than in North America. The main reason for the increased usage of pretreatment options is that treatment facilities in Europe benefit greatly from the small to moderate increase in biogas production from applying these technologies through green energy credit and other incentives for generating additional green energy from biosolids.

Co-digestion with Other Organic Waste Material

Anaerobic digestion has been applied traditionally as a single substrate, single purpose treatment process and is commonly used in municipal, industrial and agricultural treatment facilities. Most municipal wastewater treatment plants have reported an excess digestion capacity of 15 to 30 percent (Hansen, 2006). These facilities may be able to process with their existing digester capacity a wide range of organic material with municipal sludge and increase their biogas production. The process of digesting more than one substrate is called co-digestion. The technical, economical, and environmental drivers for co-digesting organic wastes with municipal sludge are outlined in Table 13–39 (adapted from WEF, 2010b).

Co-digestion or "co-fermentation" is the simultaneous digestion of a mixture of two or more organic substrates, usually a primary substrate such as wastewater sludge together with lesser amounts of one or more secondary substrate, such as organic municipal solid waste (MSW), source separated organic waste, flotation scum layers, glycerin, and brown

Table 13–40

Biogas unit production and methane content[a]

Feedstock type	Gas yield per unit solids destroyed, m^3/kg	Methane content
Fat	1.2–1.6	62–72
Scum	0.9–1.0	70–75
Grease	1.1	68
Protein	0.7	73

[a] Adapted from WEF (2010a).

grease collected from grease traps. Yellow grease, having other commercial uses, is considered too valuable to be used as a feedstock for co-digestion. When blending organic substrates, the possible outcomes can be synergistic, antagonistic, or neutral based upon methane production that is greater than, less than, or the same as that observed when each material is digested alone. The successful co-digestion of any organic feedstock requires careful management (Zitomer et al., 2008).

Anaerobic digestion process appears to become more stable when a variety of substrates are co-digested causing improved digester performance (Braun and Wellinger, 2003; Schafer and Lekven, 2008; STOWA, 2006). The addition of certain organic substrate to the anaerobic digestion system can stimulate biological activity and improve digestion performance in terms of biogas production as illustrated on Fig. 13–32.

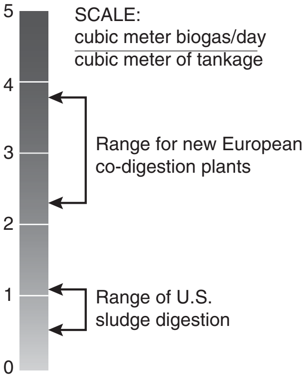


Figure 13–32

Gas production efficiency comparison. (Adapted from Schafer and Lekven, 2008.)

Co-digestion of Liquid, High Strength Organic Wastes. Co-digestion of fats, oils, and grease (FOG) is the most common among high strength liquid organic wastes due to its observed high biogas yield. Other liquid organic wastes are suitable for co-digestion including whey from cheese production or residual glycerin from biodiesel production. Biogas production and methane content from the degradation of some organic wastes are presented in Table 13–40. Detailed waste characterization data for selected high-energy organic feedstocks for co-digestion is presented in Table 13–41.

FOG is mostly referred to as material that is collected in interceptors or grease traps of restaurants and cafeterias. Separately collected unprocessed FOG generally has a high residual water content. Brown grease (often also referred to as trap grease) is obtained after the residual water content of the collected FOG is removed. In general, unprocessed FOG is comprised of a brown grease content of around 10 percent by volume; the remaining 90 percent is water (NREL, 2008).

Table 13–41

Characteristics of selected high-energy organic feedstocks[a]

Component	Unit	Restaurant interceptor grease	Biodiesel glycerin	Polymer dewatered FOG	Lime dewatered FOG
Total solids	% solids	1.8–21.9	14.7	42.4	49.1
Volatile solids	% solids	1.2–21.6	14.0	40.9	37.4
Volatile solids/total solids	%	88.9–98.6	95.2	96.5	76.5
pH	—	4.3–4.8	8.4	4.0	6.5

[a] Adapted from WEF (2010a).

Figure 13–33

City of San Francisco waste water treatment plant FOG receiving station.

Considerations for FOG Co-digestion. WWTPs that co-digest pretreated FOG and other high strength (liquid) organic wastes had to make modifications to their existing facilities design to accommodate delivery, storage, dosing, and mixing of the delivered material (WEF, 2010b). The actual co-digestion program and its process design may vary from plant to plant. While some facilities feed the pretreated FOG directly into the digester, most co-digestion programs use one or more separate holding tanks for treated FOG storage. In this case the received FOG (see Fig. 13–33) may be stored temporarily and added to the digester through combining with the influent sludge or combined with the digested sludge through existing heat exchange recycling system.

Additional components of the pretreatment train may include a rock trap or screen for contaminate removal, chopper pump for tank mixing, tank air ventilation, or activated carbon for odor control. Some co-digestion programs may also be designed to receive only substrate that has already been pretreated off-site. A dosing pump automatically and steadily feeds the FOG into the anaerobic digester. The treated FOG addition should be increased gradually to minimize lag effects and increase process stability. The gradual increase in FOG addition should also prevent any sudden surge in gas production that may result in solids flotation or digester stratification (WEF, 2010b).

Once the biomass in the digester has been acclimated to the FOG addition, an increase in both overall biogas production as well as methane content has been reported. For a successful FOG co-digestion program and plant operation, adequate injection, effective digester heating and mixing (to avoid dead zones) were found to be important process factors. To take full advantage of the anticipated increase in digester gas production extra co-generation capacity for power and heat production may be required.

Co-digestion of Organic Solid Wastes. Anaerobic digestion technologies were commercially developed for source-separated organic waste (SSO) and the organic fraction of MSW (OFMSW) in Europe around 1990 and are now in use worldwide. As the technologies for the treatment of SSO and OFMSW become more developed, the number of plants that co-digest these material will increase. SSO and OFMSW have been found suitable for co-digestion with municipal solids with promising gas yields. The amount of biogas that can be produced from co-digesting SSO and other organic substrates can be found in Braun and Wellinger (2003).

East Bay Municipal Utility District (EBMUD) in California is leading the effort on co-digestion of organic solid wastes. Funded in part through an U.S. EPA Region 9 grant, bench-scale studies were conducted to determine biodegradability, methane gas production,

Table 13–42

Inputs and outputs of the Co-digestion Economic Analysis Tool (CoEAT) of EPA

Inputs (including financial data)	Outputs
1. Feedstock type and generation	1. Fixed and recurring costs
2. Collection and Transportation	2. Savings from solids waste diversion
3. Processing	3. Capital investment
4. Digestion infrastructure	4. Biogas production and energy values
5. Disposal of biosolids	5. Methane reduction from landfills

and the required minimum mean residence time when feeding pulped SSO to the digesters. The results show that three to three and a half times more methane was produced from co-digestion when compared to digesting only sludge for a given digester volume. The EBMUD has developed and patented its in-house food-waste recycling process that has been operating since early 2000 and is accepting approximately 36 tonne/d (79,400 lb/d) of pretreated (crushed and screened) food scraps for co-digestion (Peck, 2008; Gray et al., 2008a; Gray et al., 2008b).

Cost-Effectiveness of Co-digestion. Cost effectiveness of co-digestion of organic wastes depends on many factors. The important factors for cost consideration include waste type, location and distance from plant, tipping fee, required on-site pretreatment, digestion capacity, method of beneficial use of the generated biogas, and electricity prices. Region 9 of the U.S. EPA developed a co-digestion economic analysis tool (CoEAT) based on the extensive research and experience of EBMUD. The CoEAT is designed for use by decision makers and is considered the initial step in assessing the economic feasibility of food waste co-digestion at wastewater treatment plants for the purpose of biogas production. The input and output information for the CoEAT program is presented in Table 13–42. The types of organic wastes considered includes residential food waste, commercial food waste, FOG, food processing waste (fruit, vegetables, breads, rendering byproducts), dairy waste—milk solids, and agricultural waste (fruit and vegetable trimmings).

13–10 AEROBIC DIGESTION

Aerobic digestion may be used to treat (1) waste activated sludge only, (2) mixtures of waste activated sludge or trickling-filter sludge and primary sludge, or (3) waste sludge from extended aeration plants. Aerobic digestion has been used primarily in plants of a size less than 0.2 m^3/s (5 Mgal/d), but in recent years the process has been employed in larger wastewater treatment plants with capacities up to 2 m^3/s (50 Mgal/d) (WEF, 2010a). In cases where separate sludge digestion is considered, aerobic digestion of biological sludge may be an attractive application. Advantages and disadvantages of conventional aerobic digestion as compared to anaerobic digestion are provided in Table 13–43.

As discussed in Sec. 13–2 and Table 13–12, aerobic digestion is one of the processes defined to meet PSRP requirements for Class B biosolids. To meet Class B requirements for pathogen reduction, the regulations state the solids retention times must be at least 40 d at 20°C and 60 d at 15°C. In many instances, plants that have facilities designed for SRTs less than 40 d and wish to meet the Class B requirements for pathogen reduction have had to add additional storage capacity or thickeners. If the design engineer uses aerobic digestion for stabilization and does not meet the above SRTs, it will be necessary to monitor the performance of the process to demonstrate that the pathogen reduction

Table 13–43

Advantages and disadvantages of aerobic digestion

Advantages	Disadvantages
1. Volatile solids reduction in a well-operated aerobic digester is approximately equal to that obtained anaerobically	1. High power cost is associated with supplying the required oxygen
2. Lower BOD concentrations in sidestreams than anaerobic digestion	2. Does not produce methane for energy recovery
3. Produces an odorless, humuslike, biologically stable end product	3. Aerobically digested biosolids produced have poorer mechanical dewatering characteristics than anaerobically digested biosolids
4. Allows recovery of the basic fertilizer values in the biosolids	4. The process is affected significantly by temperature, location, tank geometry, concentration of feed solids, type of mixing/aeration device, and type of tank material
5. Simple technology requiring few special skills for reliable operation	5. Process consumes alkalinity
6. Low capital cost for small facilities	
7. Easy to construct of readily available parts	
8. Suitable for digesting nutrient-rich waste activated sludges	
9. No risk for explosions	

[a] Adapted, in part, from WEF (2012).

criterion has been met. Monitoring is also required to demonstrate that the volatile solids reduction requirements are met for compliance to the vector attraction criterion (U.S. EPA, 2003).

Process Description

Aerobic digestion is similar to the activated sludge process. As the supply of available substrate (food) is depleted, the microorganisms begin to consume their own protoplasm to obtain energy for cell maintenance reactions. When energy is obtained from cell tissue, the microorganisms are said to be in the endogenous phase. Cell tissue is oxidized aerobically to carbon dioxide, water, and ammonia. In actuality, only about 75 to 80 percent of the cell tissue can be oxidized; the remaining 20 to 25 percent is composed of inert components and organic compounds that are not biodegradable. The ammonia is subsequently oxidized to nitrate as digestion proceeds. Nonbiodegradable volatile suspended solids will remain in final product from aerobic digestion. Considering the biomass wasted to a digester and the formula $C_5H_7NO_2$ is representative for cell mass of a microorganism, the biochemical changes in an aerobic digester can be described by the following equations:

Biomass destruction:

$$C_5H_7NO_2 + 5O_2 \rightarrow 4CO_2 + H_2O + NH_4HCO_3 \tag{13–19}$$

Nitrification of released ammonia nitrogen:

$$NH_4^+ + 2O_2 \rightarrow NO_3^- + 2H^+ + H_2O \tag{13–20}$$

Overall equation with complete nitrification:

$$C_5H_7NO_2 + 7O_2 \rightarrow 5CO_2 + 3H_2O + HNO_3 \tag{13–21}$$

Using nitrate nitrogen as electron acceptor (denitrification):

$$C_5H_7NO_2 + 4NO_3^- \rightarrow 5CO_2 + 2N_2 + NH_3 + 4OH^- \tag{13–22}$$

With complete nitrification/denitrification:

$$2C_5H_7NO_2 + 11.5O_2 \rightarrow 10CO_2 + N_2 + 7H_2O \tag{13–23}$$

As given by Eqs. (13–19) through (13–21), the conversion of organic nitrogen to nitrate results in an increase in the concentration of hydrogen ions and subsequently a decrease in pH if sufficient buffering capacity is not available in the sludge. Approximately 7 kg of alkalinity, expressed as $CaCO_3$, are destroyed per each kg of ammonia oxidized. Theoretically, approximately 50 percent of the alkalinity consumed by nitrification can be recovered by denitrification. If the dissolved oxygen is kept very low (less than 1 mg/L), however, nitrification will not occur. In practice, cycling of the aerobic digester between aeration and mixing has been found to be effective in maximizing denitrification while maintaining pH control. In situations where the buffering capacity is insufficient, resulting in pH values below 5.5, it may be necessary to install alkalinity feed equipment to maintain the desired pH.

Where activated or trickling-filter sludge is mixed with primary sludge and the combination is to be digested aerobically, direct oxidation of the organic matter in the primary sludge and oxidation of the cell tissue will both occur. Aerobic digesters can be operated as batch or continuous flow reactors (see Fig. 13–34). Three proven variations of the process are most commonly used: (1) conventional aerobic digestion, (2) high-purity oxygen aerobic digestion, and (3) autothermal aerobic digestion (ATAD). Aerobic digestion accomplished

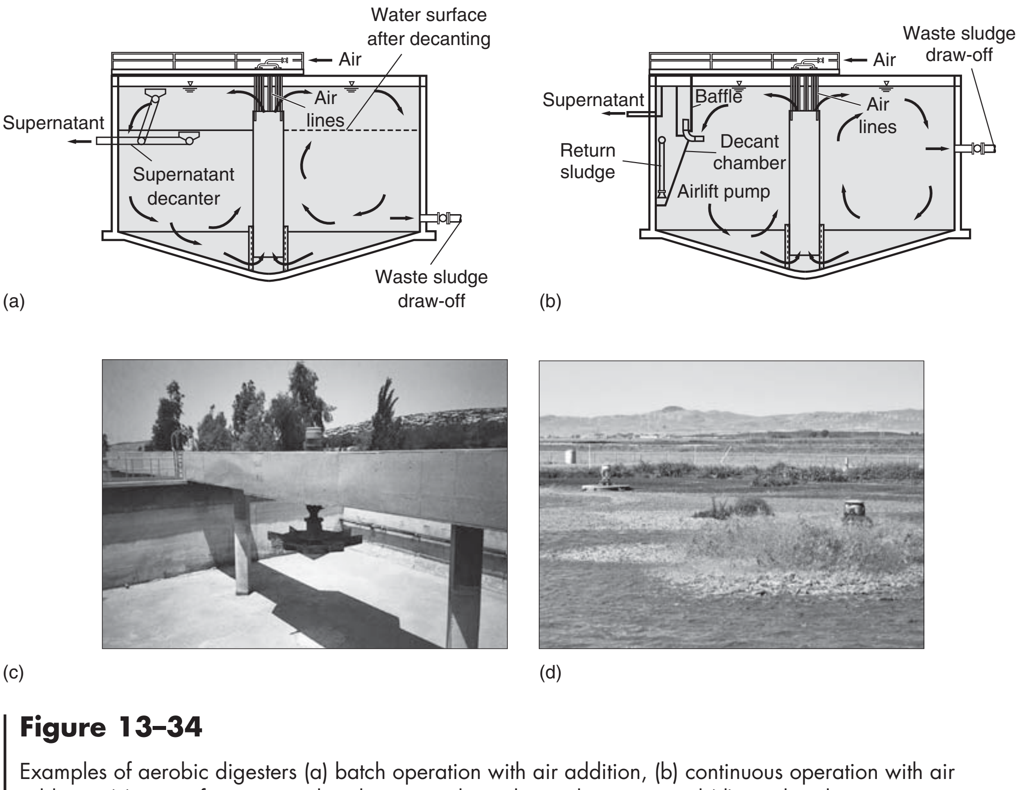


Figure 13–34

Examples of aerobic digesters (a) batch operation with air addition, (b) continuous operation with air addition, (c) view of empty aerobic digester with mechanical aerator, and (d) aerobic digester in lined earthen basin.

with air is the most commonly used process, so it is considered in greater detail in the following discussion.

Conventional Air Aerobic Digestion

Factors that must be considered in designing conventional aerobic digesters include temperature, solids reduction, tank volume, feed solids concentration, oxygen requirements, energy requirements for mixing, and process operation. Typical design criteria for aerobic digestion are presented in Table 13–44.

Temperature. Because the majority of aerobic digesters are open tanks, digester liquid temperatures are dependent on weather conditions and can fluctuate extensively. As with all biological systems, lower temperatures retard the process while higher temperatures accelerate it. In considering the temperature effects, heat losses should be minimized by using concrete instead of steel tanks, placing the tanks below grade instead of above grade or providing insulation for above-grade tanks, and using subsurface instead of surface aeration. In extremely cold climates, consideration should be given to heating the sludge or the air supply, covering the tanks, or both. The design should provide for the necessary degree of sludge stabilization at the lowest expected liquid operating temperature and should provide the maximum oxygen requirements at the maximum expected liquid operating temperature.

Volatile Solids Reduction. A major objective of aerobic digestion is to reduce the mass of the solids for disposition. This reduction is assumed to take place only with the

Table 13–44

Design criteria for aerobic digesters[a]

	U.S. customary units		SI units	
Parameter	**Units**	**Value**	**Units**	**Value**
SRT[b]	d		d	
At 20°C		40		40
At 15°C		60		60
Volatile solids loading	lb/ft^3·d	0.1–0.3	kg/m^3·d	1.6–4.8
Oxygen Requirements:				
Cell tissue[c]	lb O$_2$/lb VSS destroyed	~ 2.3 1.6–1.9	kg O$_2$/kg VSS destroyed	~ 2.3 1.6–1.9
BOD in primary sludge				
Energy requirements for mixing				
Mechanical aerators	hp/10^3 ft^3	0.75–1.5	kW/10^3 m^3	20–40
Diffused air mixing	ft^3/10^3 ft^3·min	20–40	m^3/m^3·min	0.02–0.040
Dissolved oxygen residual in liquid	mg/L	1–2	mg/L	1–2
Reduction volatile suspended solids	%	38–50	%	38–50

[a] Adapted, in part, from WEF (1995a); Federal Register (1993).

[b] To meet pathogen reduction requirements (PSRP) of 40 CFR Part 503 regulations.

[c] Ammonia produced during carbonaceous oxidation oxidized to nitrate.

biodegradable content of the sludge, although there may be some destruction of the nonorganics as well. Volatile solids reductions ranging from 35 to 50 percent are achievable by aerobic digestion. Optional criteria for meeting vector attraction requirements of 40 CFR Part 503 are (1) a minimum of 38 percent reduction in volatile solids during biosolids treatment or (2) less than a specific oxygen uptake rate (SOUR) of (1.5 mg O_2/h)/g of total sludge solids at 20°C (U.S. EPA, 1999).

The change in biodegradable volatile solids in a completely mixed digester can be represented by a first-order biochemical reaction at constant-volume conditions:

$$r_M = -k_d M \tag{13–24}$$

where r_M = rate of change of biodegradable volatile solids (M) per unit of time (Δmass/time), MT^{-1}

k_d = reaction rate constant, T^{-1}

M = mass of biodegradable volatile solids remaining at time t in the aerobic digester, M

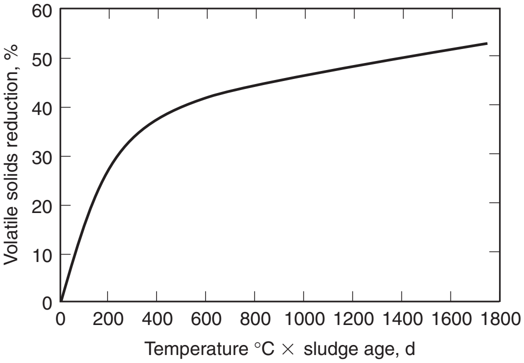
The time factor in Eq. (13–24) is the solid's retention time (SRT) in the aerobic digester. Depending on how the aerobic digester is being operated, time t can be equal to or considerably greater than the theoretical hydraulic residence time (t). Use of the biodegradable portion of the volatile solids is based on the fact that approximately 20 to 35 percent of the waste activated sludge from wastewater treatment plants with primary treatment is not biodegradable. The percentage of nonbiodegradable volatile solids in waste activated sludge from contact stabilization processes (no primary tanks) ranges from 25 to 35 percent (WEF, 2010a).

The reaction rate term k_d is a function of the sludge type, temperature, and solids concentration. Representative values for k_d may range from 0.05 d^{-1} at 15°C to 0.14 d^{-1} at 25°C for waste activated sludge. Because the reaction rate is influenced by several factors, it may be necessary to confirm decay coefficient values by bench-scale or pilot-scale studies.

Solids destruction is primarily a direct function of both basin liquid temperature and the SRT (sometimes referred to as sludge age), as indicated on Fig. 13–35. The data were derived from both pilot-and full-scale studies. The plot on Fig. 13–35 relates volatile solids reduction to degree-days (temperature times sludge age). Initially, as the degree-days increase, the rate of volatile solids reduction increases rapidly. As the degree-days approach 500, the curve begins to flatten. To produce well-stabilized biosolids, at least 550 degree-days are recommended for the aerobic digestion system (Enviroquip, 2000). The use of Fig. 13–35 is demonstrated in Example 13–8, the design of an aerobic digester.

Figure 13–35

Volatile solids reduction in an aerobic digester as a function of digester liquid temperature and digester sludge age.



Tank Volume and Detention Time Requirements. The tank volume is governed by the detention time necessary to achieve the desired volatile solids reduction. In the past, SRTs of 10 to 20 d were the norm for the design of aerobic digestion systems (Metcalf & Eddy, 1991). To meet the pathogen reduction requirements of 40 CFR Part 503 regulations, the SRT criteria (see Table 13–44) in conventional aerobic digesters take precedence over the vector attraction criteria of 38 percent solids reduction for sizing the tank volume.

The digester tank volume can be calculated by Eq. (13–25) (WEF, 2010a):

$$V = \frac{Q_i(X_i + YS_i)}{X(k_d P_v + 1/\text{SRT})} \tag{13–25}$$

where V = volume of aerobic digester, m^3 (ft^3)
Q_i = influent average flowrate to digester, m^3/d (ft^3/d)
X_i = influent suspended solids, mg/L
Y = fraction of the influent BOD consisting of raw primary solids, (expressed as a decimal)
S_i = influent BOD, mg/L
X = digester suspended solids, mg/L
k_d = reaction rate constant, d^{-1}
P_v = volatile fraction of digester suspended solids (expressed as a decimal)
SRT = solids retention time, d

The term YS_i can be neglected if primary sludge is not included in the sludge load to the aerobic digester.

If the aerobic digestion process is operated in a complete-mix, staged configuration (two or three stages), the total SRT should be divided approximately equally among the stages. For more information on staging aerobic digestion, Enviroquip (2000) should be consulted.

Feed Solids Concentration. The concentration of the digester feed solids is important in the design and operation of the aerobic digester. If thickening precedes aerobic digestion, higher feed solids concentrations will result in higher oxygen input levels per digester volume, longer SRTs, smaller digester volume requirements, easier process control (less decanting in batch-operated systems), and subsequently increased levels of volatile solids destruction (WEF 2010a). However, feed solids concentrations greater than 3.5 to 4 percent may affect the ability of the mixing and aeration system in maintaining well-mixed tank contents with adequate dissolved oxygen levels necessary to support the biological process. At feed solids concentrations greater than 4 percent, the aeration equipment must be evaluated carefully to ensure adequate mixing and aeration are achieved. Also at feed solids concentration greater than 4 percent, provisions for removing the heat from the digesters should be practiced.

Oxygen Requirements. The oxygen requirements that must be satisfied during aerobic digestion are those of the cell tissue and, with mixed sludges, the BOD in the primary sludge. The oxygen requirement for the complete oxidation of cell tissue (including nitrification), computed using Eqs. (13–19) and (13–20), is equal to 7 mole/mole of cells, or about 2.3 kg/kg of cells. The oxygen requirement for the complete oxidation of the BOD contained in primary sludge varies from about 1.6 to 1.9 kg/kg destroyed. The oxygen residual should be maintained at 1 mg/L or above under all operating conditions.

Energy Requirements for Mixing. To ensure proper operation, the contents of the aerobic digester should be well mixed. In large tanks, multiple mixing devices should be installed to ensure good distribution of the mixing energy. Typical energy requirements for mixing are given in Table 13–44. In general, because of the large amount of air that must be supplied to meet the oxygen requirement, adequate mixing should be achieved; nevertheless, mixing power requirements should be checked, particularly when feed solids concentrations are greater than 3.5 percent. If polymers are used in the thickening process, especially for centrifuge thickening, a greater amount of unit energy may be required for mixing.

If fine-pore diffused air mixing is used, considerations for selecting the aeration system should include limitations of feed solids concentration on achieving good mixing. Recommendations on feed solids limitations should be obtained from manufacturers of aeration equipment. In addition, the potential for diffuser fouling should be evaluated, especially if the process operation requires decanting.

Process Operation. Depending on the buffering capacity of the system, the pH may drop to a low value of about 5.5 at long hydraulic detention times. The potential drop in pH is due to the increased presence of nitrate ions in solution and the lowering of the buffering capacity due to air stripping. Filamentous growths may also develop at low pH values. The pH should be checked periodically and adjusted if found to be excessively low. Dissolved oxygen levels and respiration rates should also be checked to ensure proper process performance.

Aerobic digesters that do not include prethickening should be equipped with decanting facilities for thickening the digested biosolids before discharge to subsequent operations. Operator control and visibility of the decanting operation are important design considerations. If the digester is operated so that the incoming sludge is used to displace supernatant and the biosolids are allowed to build up, the solids retention time will not be equal to the hydraulic retention time.

EXAMPLE 13–8 **Aerobic Digester Design** Design an aerobic digester to treat the waste sludge produced by the activated sludge treatment plant. Assume that the following conditions apply:

1. The amount of waste sludge to be digested is 2100 kg TSS/d.
2. The minimum and maximum liquid temperatures are 15°C for winter operation and 25°C for summer operation.
3. The system must achieve 40 percent volatile solids reduction in the winter.
4. The minimum SRT for winter conditions is 60 d.
5. Waste activated sludge is concentrated to 3 percent, using a dissolved air flotation thickener.
6. The specific gravity of the waste sludge is 1.03.
7. Sludge concentration in the digester is 70 percent of the incoming thickened sludge concentration.
8. The reaction rate coefficient k_d is 0.06 d^{-1} at 15°C.
9. Volatile fraction of digester TSS is 0.65.
10. No primary sludge is included in the influent to the digester.
11. Diffused-air mixing is used.
12. Air temperature in diffused air system = 20°C.

Solution

1. Compute the volatile solids reduction for winter conditions using Fig. 13–35 and compute the percent volatile solids reduction under summer (maximum) conditions.
   a. For winter conditions, the degree-days from Fig. 13–35 are 15°C × 60 d = 900 degree-days. From Fig. 13–35, the volatile solids reduction is 45 percent, which exceeds the winter requirements of 40 percent.
      To meet the pathogen reduction requirements, the SRT must be 60 d; therefore, the required volume is $68.0\ \text{m}^3/\text{d} \times 60\ \text{d} = 4080\ \text{m}^3$.
   b. During the summer, the liquid temperature will be 25°C, and the degree-days will be 25 × 60 = 1500. From Fig. 13–35, the volatile solids reduction in the summer will be 50 percent.
2. Compute the winter and summer volatile solids reduction based on a total mass of volatile suspended solids.

   Total mass of VSS (VSS_M) = (0.65)(2100 kg/d) = 1365 kg/d

   a. Winter: $1365 \times 0.45 = 614$ kg VSS_M reduced/d
   b. Summer: $1365 \times 0.50 = 682$ kg VSS_M reduced/d
3. Determine oxygen requirements (see Table 13–45 for oxygen requirements).
   a. Winter: $614 \times 2.3 = 1412$ kg O_2/d
   b. Summer: $682 \times 2.3 = 1569$ kg O_2/d
4. Compute the volume of air required per d at standard conditions. For the density of air, see Appendix B–1. Note that air is approximately 23.2% oxygen by weight.

   a. Winter: $$V = \frac{1412\ \text{kg}}{(1.204\ \text{kg/m}^3)(0.232)} = 5055\ \text{m}^3/\text{d}$$

   b. Summer: $$V = \frac{1569\ \text{kg}}{(1.204\ \text{kg/m}^3)(0.232)} = 5617\ \text{m}^3/\text{d}$$

   Assuming an oxygen transfer efficiency of 10 percent, the air flowrates are

   Winter: $$q = \frac{(5055\ \text{m}^3/\text{d})}{(0.1)(1440\ \text{min/d})} = 35.1\ \text{m}^3/\text{min}$$

   Summer: $$q = \frac{(5617\ \text{m}^3/\text{d})}{(0.1)(1440\ \text{min/d})} = 39.0\ \text{m}^3/\text{min}$$
5. Compute the volume of sludge to be disposed of per day using Eq. (13–2).

   $$Q = \frac{2100\ \text{kg}}{(10^3\ \text{kg/m}^3)(1.03)(0.03)} = 68.0\ \text{m}^3/\text{d}$$
6. Compute the air requirement per m^3 of digester volume.

   $$q = \frac{(39.0\ \text{m}^3/\text{min})}{4080\ \text{m}^3} = 0.0096\ \text{m}^3/\text{min}\cdot\text{m}^3$$
7. Check the mixing requirements. Because the air requirement computed in Step 7 is below the range of values given in Table 13–44, mixing requirements will govern the design of the aeration system, unless separate mixing is provided.

Comment The above example is based on a single-stage aerobic digester. If a two-stage or more digester were used, a significant reduction in tank volume is possible. In a multistage arrangement, the air distribution between tanks would vary based on the expected demand as most of the volatile solids reduction will occur in the first stage where the biomass is most active.

Table 13–45

Advantages and disadvantages of ATAD

Advantages	Disadvantages
1. High reaction rate and low retention time requirements when compared to other digestion processes	1. High odor potential
2. Simple operation	2. Potential for poor dewatering characteristics^b
3. Greater reduction of bacteria and viruses are achieved as compared to mesophilic anaerobic digestion	3. Does not nitrify
4. Can meet Class A when the reactor is well mixed and maintained at or above 55°C	4. Requires upstream mechanical thickening
5. Fully enclosed reactors	5. Sidestreams contain high nutrient loads that may require further treatment
6. Lower energy requirements than conventional aerobic digestion	6. Foam control is necessary
	7. Many processes are proprietary
	8. Potential for erosion / corrosion

[a] Adapted, in part, from WEF (2012).

[b] Many newer designs now incorporate product cooling and an additional mesophilic digestion system which is reported to significantly improve dewatering characteristics.

Dual Digestion

Aerobic thermophilic digestion has also been used extensively in Europe as a first stage in the dual digestion process. The second stage is mesophilic anaerobic digestion. Dual digestion has also been tried in the United States using high-purity oxygen in the first stage. Residence times in the aerobic digester range typically from 18 to 24 h, and the reactor temperature ranges from 55 to 65°C. Typical residence time in the anaerobic digester is 10 d. The advantages of using aerobic thermophilic digestion in dual digestion are (1) increased levels of pathogen reduction, (2) improved overall volatile solids reduction, (3) increased methane gas generation in the anaerobic digester, (4) less organic material in and fewer odors produced by the stabilized sludge, and (5) equivalent volatile solids reductions can be achieved in one-third less tankage than a single-stage anaerobic digester. Prior hydrolysis in the aerobic reactor results in increased degradation during subsequent anaerobic digestion and gas production. Approximately 10 to 20 percent of the volatile solids is liquefied in the aerobic digester, while COD reduction is less than 5 percent. Provisions for foam suppression and odor control are required (Roediger and Vivona, 1998).

Autothermal Thermophilic Aerobic Digestion (ATAD)

Autothermal thermophilic aerobic digestion (ATAD), illustrated on Fig. 13–36, represents a variation of both conventional and high-purity oxygen aerobic digestion. In the ATAD process, the feed sludge is generally prethickened, and the reactors are insulated to conserve the heat produced from the oxidation of volatile solids during the digestion process. Thermophilic operating temperatures (generally in the range of 55 to 70°C) can be achieved without external heat input by using the heat released by the exothermic microbial oxidation process. The heat produced per kg of volatile solids destroyed is approximately 20,000 kJ. Because supplemental heat is not provided (other than the heat introduced by aeration and mixing), the process is termed autothermal.

Within the ATAD reactor, sufficient levels of oxygen, volatile solids, and mixing allow aerobic microorganisms to degrade organic matter to carbon dioxide, water, and nitrogen byproducts. The major advantages and disadvantages of ATAD are cited in Table 13–45.

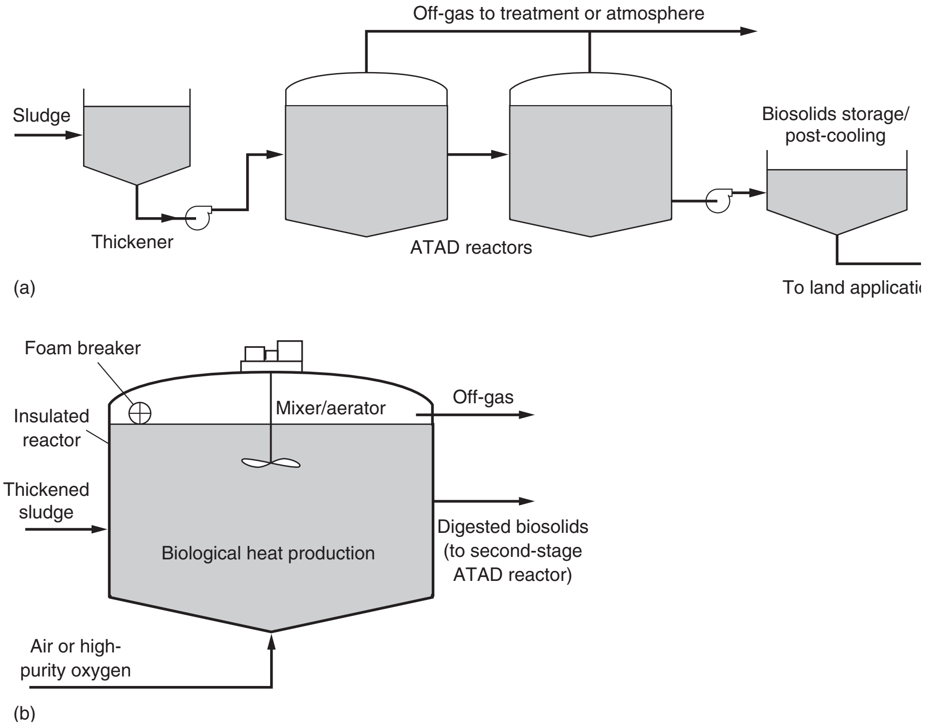


Figure 13–36

Autothermal thermophilic aerobic digester (ATAD) system: (a) process flow diagram and (b) detail of typical reactor.

Because the ATAD system is capable of producing Class A biosolids, it is growing in popularity. In a partial survey of ATAD systems in USA, ~25 ATAD systems were in operation (Meckes, 2011).

Process Theory. The biochemical conditions in thermophilic aerobic digesters differ significantly from conventional aerobic digesters. Because of the high operating temperatures, nitrification is inhibited, and aerobic destruction of volatile solids occurs as described by Eq. (13–19) without the subsequent reactions described by Eqs. (13–20) through (13–23). Additionally, most, if not all, ATAD systems may be operating under microaerobic conditions where oxygen demand exceeds oxygen supply (Stensel and Coleman, 2000). Under microaerobic conditions, proteinaceous cellular material will undergo fermentation where protein is represented as peptone as described by Eq. (13–26) (Chu and Mavinic, 1998):

$$4CH_2NH_2COOH + 4H_2O \rightarrow 3CH_3COOH + 2(NH_4)_2CO_3 \qquad (13\text{–}26)$$

Both Eqs. (13–19) and (13–26) result in the production of ammonia that reacts with water and carbon dioxide to form ammonium bicarbonate and ammonium carbonate to increase alkalinity. Because nitrification does not occur, the pH in the ATAD system will typically range from 8 to 9, higher than in conventional aerobic digesters. Ammonia-nitrogen produced will be present in the off-gas and in solution with concentrations of several hundred mg/L in each. Most of the ammonia nitrogen will be returned to the liquid process in sidestreams from the odor control and dewatering facilities. The acetic acid (or acetate) produced by the fermentation of proteins is one of the volatile fatty acids. Acetic acid will be oxidized subsequently in the presence of sufficient dissolved oxygen as described by Eq. (13–27):

$$CH_3COOH + 2O_2 \rightarrow 2CO_2 + 2H_2O \qquad (13\text{–}27)$$

Anaerobic conditions will occur at times in ATAD systems and will most likely take place in the pre-ATAD sludge holding facilities and in the first-stage ATAD reactors during and immediately after sludge transfers and batch feeding. Under anaerobic conditions, reduced sulfur compounds can be formed that can affect the design and performance of odor-control systems.

Process Design. ATAD systems are designed to have short hydraulic retention times within insulated reactors [see Fig. 13–36(b)]. As long as the ATAD system is well mixed and sufficient oxygen is provided, the temperature in the reactor will rise until a balance occurs; i.e., the heat lost equals the heat input from the exothermic reaction and mechanical energy input. The temperature will continue to rise until the process becomes oxygen mass-transfer-limited.

Factors that must be considered in designing an ATAD system include prethickening, number and type of reactors, postcooling/thickening, feed characteristics, detention time, feed cycle, aeration and mixing, temperature and pH, and foam and odor control. Nearly all of the ATAD systems currently installed in the United States utilize two or more reactors operated in series [see Fig. 13–36(a)]. Design considerations for ATAD systems are presented in Table 13–46; typical design criteria are summarized in Table 13–47.

ATAD systems must be designed to (1) transfer sufficient oxygen to meet the high demand of the reactors and (2) supply the required oxygen while minimizing the latent

Table 13–46

Typical design considerations for an ATAD system[a]

System component	Design consideration
Prethickening system	Thickening or blending facilities may be required to maintain an influent COD to the ATAD reactor greater than 40 g/L
Reactors	Number of reactors; a minimum of two enclosed insulated reactors in series equipped with mixing aeration, and foam control equipment
Screening	Fine screening 6 to 12 mm (0.25 to 0.5 in.) clear openings of raw wastewater or sludge feed stream should be provided for the removal of inert materials, plastics, and rags
Feed cycle	Continuous or batch processing is acceptable, except batch processing provides greater assurance in meeting Class A pathogen reduction requirements
Foam control	Foam suppression is required to ensure effective oxygen transfer and enhanced biological activity. Freeboard of 0.5 to 1.0 m (1.65 to 3.3 ft) is recommended
Post-ATAD storage/ dewatering	Postprocess cooling is necessary to achieve solids consolidation and to enhance dewaterability. A minimum of 20 d detention may be necessary unless heat exchangers are used for cooling the processed biosolids
Odor control	Because of high temperatures in the ATAD system, relatively high concentrations of ammonia are released. Reduced sulfur compounds also result, which can include hydrogen sulfide, carbonyl sulfide, methyl mercaptan, ethyl mercaptan, dimethyl sulfide, and dimethyl disulfide. Odor-control systems may include wet scrubbers, biofilters, or a combination of both (see Chap. 16)
Sidestreams	Liquid sidestreams from odor-control and dewatering systems, when returned to the liquid processing system, may contain constituents that could affect process performance unless accounted for or treated separately

[a] Adapted in part from WEF (2010a) and Stensel and Coleman (2000).

Table 13–47

Typical design parameters for autothermal aerobic digester (ATAD)[a]

	U.S. customary units			SI units		
Parameter	**Units**	**Range**	**Typical**	**Units**	**Range**	**Typical**
Reactor						
HRT	d	4–30	6–8	d	4–30	6–8
Volumetric loading						
TSS, 40 to 60 g/L	lb/10^3 ft^3·d	320–520		kg/m^3·d	5–8.3	
VSS, 25 g/L	lb/10^3 ft^3·d	200–260		kg/m^3·d	3.2–4.2	
Temperature						
Stage 1	°F	95–122	104	°C	35–50	40
Stage 2	°F	122–158	131	°C	50–70	55
Aeration and Mixing						
Mixer type			Aspirating			Aspirating
Oxygen transfer efficiency	lb O_2/kWh		4.4	kg O_2/kWh		2
Energy requirement	hp/10^3 ft^3	5–6.4		W/m^3	130–170	

[a] Adapted, in part, from Stensel and Coleman (2000).

heat loss in the exhaust air. It is difficult to define the oxygen transfer rate in an ATAD system while using typical design procedures used for selecting and sizing aeration equipment for wastewater treatment processes. The oxygen transfer coefficient α (alpha) and the oxygen saturation coefficient β (beta) have not been quantified under the environmental conditions present in an ATAD reactor (Stensel and Coleman, 2000). Factors affecting oxygen transfer are the high temperatures (that would reduce α values) and the foam layer and low dissolved oxygen levels (that might increase oxygen transfer). Nearly all ATAD systems utilize a type of aspirating aerator to introduce oxygen into the reactors. The types include hollow-shaft propeller or turbine aerators, pumped venturi aspirators, and jet aspirators. With all air aspirating systems, the equipment provides both mixing and oxygen transfer. Typical energy requirements for mixing and aeration are given in Table 13–47.

Substantial amounts of foam are generated in the ATAD process as cellular proteins, lipids, and oil and grease materials are broken down and released into solution. The foam layer contains high concentrations of biologically active solids that provide insulation of the reactor and improved oxygen utilization. It is important, therefore, that the foam layer be managed and controlled effectively. Mechanical foam cutters are used most commonly for foam control, but other methods such as spray systems have been employed. A freeboard of 0.5 to 1.0 m is generally recommended for controlling the foam layer (Stensel and Coleman, 2000).

Where ATAD systems are followed by mechanical dewatering, post-ATAD storage is recommended to allow for cooling of the biosolids to improve dewatering performance. Post-ATAD storage coupled with long detention times in the final-stage ATAD reactors may further increase the reduction of volatile solids.

Process Control. The provisions of the 40 CFR Part 503 regulations applicable for meeting the Class A biosolids requirements with the ATAD process are complex because several alternative pathogen-reduction requirements are given. The basic requirements that need to be demonstrated are (1) fecal coliform densities are less than 1000 MPN/g of total solids (dry weight basis), or (2) *Salmonella spp.* bacteria concentrations are below detection limits of 3 MPN/4 g of total solids (dry weight basis). For compliance with these pathogen regulations for Class A biosolids, the withdrawal and feeding of the sludge to the reactors is performed on a batch basis. (In flow-through systems, it is possible that some pathogens might pass through.) Two or more reactors in a series configuration are used typically to ensure that all particles in the reactor are subjected to the time and temperature requirements and that no insufficiently treated biosolids are released to the environment. The ATAD pumping system is designed to withdraw and feed the daily amount of sludge in 1 h or less. The reactor is then isolated for the remaining 23 h each day at a minimum temperature of 55°C.

Improved ATAD Systems

Because many of the first generation ATAD systems suffered from poor performance, odor problems, and high dewatering costs a second generation of ATAD systems has been developed. The differences between the second ATAD generation and previous ATAD installations includes: (1) a single thermophilic stage instead of 2 to 3 stages of shorter detention time; (2) use of pressurized (blower) air for aeration instead of aspirated air; (3) sufficient aeration pressure to maintain aerobic conditions and reduce odor; (4) aeration control based on ORP; (5) non-mechanical foam control; (6) a mesophilic aeration stage following the thermophilic stage. The volume of the first stage thermophilic reactor is approximately two thirds of the total treatment volume. In general, the second generation ATAD systems are reported to provide high volatile solids reduction, a class A biosolids product without offensive odor, and good dewatering with high solids concentration without high chemical demand (Smith et al., 2012). In addition, using a water scrubber and photo catalyter oxidizer for off gas odor control has been reported to be effective (Smith et al., 2012).

The first municipal installation of the second generation ATAD was in 2002 (Scisson, 2009). As of 2012, the largest second generation ATAD began operating at Middletown, OH in May 2009 with a processing capacity of 15 tonne/d (16.5 ton/d) (Pevec, 2010). Based on the performance data from this installation, the system has operated with minimal odor, produces an average volatile solids reduction of about 57 percent, and with centrifuge dewatering cake solids of about 31 percent have been achieved.

High-Purity Oxygen Digestion

High-purity oxygen aerobic digestion is a modification of the aerobic digestion process in which high-purity oxygen is used in lieu of air. The resultant biosolids are similar to biosolids from conventional aerobic digestion. Influent sludge concentrations vary from 2 to 4 percent. Recycle flows are similar to those achieved by conventional aerobic digestion. High-purity oxygen aerobic digestion is particularly applicable in cold weather climates because of its relative insensitivity to changes in ambient air temperatures due to the increased rate of biological activity and the exothermal nature of the process.

While one variation of the high-purity aerobic digestion process uses open tanks, aerobic digestion is usually done in closed tanks similar to those used in the high-purity oxygen activated sludge process. Using closed tanks for high-purity oxygen aerobic digestion will generally result in higher operating temperatures because of the exothermic nature of the digestion process. Maintenance of these higher temperatures in the digester results in a significant increase in the rate of volatile suspended solids destruction.

Where covered tanks are used, a high-purity oxygen atmosphere is maintained above the liquid surface, and oxygen is transferred into the sludge via mechanical aerators. Where an open aeration tank is used, oxygen is introduced to the liquid sludge by a special diffuser that produces minute oxygen bubbles. The bubbles dissolve before reaching the air-liquid interface.

The major disadvantage of high-purity oxygen aerobic digestion is the increased cost associated with oxygen generation. As a result, high-purity oxygen aerobic digestion is cost-effective generally only when used in conjunction with the high-purity oxygen activated sludge system. Also, neutralization may be required to offset the reduced buffering capacity of the system.

PROBLEMS AND DISCUSSION TOPICS

13–1 The water content of waste activated sludge is reduced from 98 to 95 percent. What is the percent reduction in volume by the approximate method and by the more exact method, assuming that the solids contain 70 percent organic matter of specific gravity 1.00 and 30 percent mineral matter of specific gravity 2.00? What is the specific gravity of the 98 and the 95 percent slurry?

13–2 Consider an activated sludge treatment plant with a flowrate of 40,000 m^3/d. The untreated wastewater contains 200 mg/L suspended solids. The plant provides 60 percent removal of the suspended solids in the primary settling tank. If the primary sludge alone is pumped, it will contain 5 percent solids. Assume that 400 m^3/d of waste activated sludge containing 0.5 percent solids is to be transferred to the digester. If the waste activated sludge is thickened in a gravity belt thickener to 6 percent TS, calculate the thickened waste activated sludge volume. Calculate the total reduction in daily volume of biosolids pumped to the digester that can be achieved by thickening the waste activated sludge in a gravity belt thickener as compared with discharging the primary and waste activated sludge directly to the digester. Assume complete capture of the waste activated sludge in the gravity belt thickener.

13–3 For Example 13–4 for gravity thickening, calculate the amount of dilution water required at average design flow using the data provided to maintain a hydraulic loading rate of 12 $m^3/m^2 \cdot d$ for the thickener size computed in the example.

13–4 Determine the required digester volume for the treatment of the sludge quantities specified in Example 13–5 using the (a) volatile solids loading factor, and (b) volumetric per capita allowance methods. Set up a comparison table to display the results obtained using the three different procedures for sizing digesters (two in this problem and one in Example 13–5). Assume the following data apply:

1. Volatile solids loading method
   a. Solids concentration = 5%
   b. Detention time = 15 d
   c. Loading factor = 2.4 kg $VSS/m^3 \cdot d$
   d. Volatile solids concentration = 75%
2. Volumetric loading method
   a. Sewer basin population = 70,000
   b. Per capita contribution = 0.72 g/capita·d
   c. Volume required 50 $m^3/10^3$ capita·d

13–5 A wastewater treatment plant is planning to provide for separate anaerobic sludge digestion for its primary sludge. The plant receives an influent wastewater with the following characteristics:

Average flowrate = 8000 m^3/d

Suspended solids removed by primary sedimentation 200 mg/L

Volatile matter in settled solids = 75%

Water in untreated sludge = 96%

Specific gravity of mineral solids = 2.60

Specific gravity of organic solids = 1.30

Using these data, determine (a) the required digester volume using an SRT of 20 d, and (b) the minimum digester capacity using the recommended loading parameters of kg $VM/m^3 \cdot d$ (kilograms of volatile matter per cubic meter per day).

13–6 A wastewater treatment plant currently dewaters on average 750 kg/d of primary and waste activated sludge to an average solid content of 22 percent TS. The plant currently uses post lime stabilization and on average mixes 300 kg/d of quicklime with their dewatered sludge in a pug mill. What is the theoretical temperature increase after adding quicklime? Discuss the advantages and disadvantage for the plant if they were to switch to anaerobic digestion for sludge stabilization in lieu of lime stabilization.

13–7 A digester is loaded at a rate of 300 kg COD/d. Using a waste-utilization efficiency of 75 percent, what is the volume of gas produced when SRT = 40 d? Assume $Y = 0.10$ and $b = 0.02\ d^{-1}$.

13–8 Volatile acid concentration, pH, or alkalinity should not be used alone to control a digester. How should they be correlated to predict most effectively how close to failure a digester is at any time?

13–9 A digester is to be heated by circulation of sludge through an external hot water heat exchanger. Using the following data, find the heat required to maintain the required digester temperature:

1. U_x = overall heat-transfer coefficient, $W/m^2 \cdot °C$
2. Wall above ground: $U_{air} = 0.85$, wall below ground: $U_{ground} = 1.2$, cover: $U_{cover} = 1.0$
3. Digester is a concrete tank with floating steel cover; diameter = 11 m and sidewall depth = 8 m, 4 m of which is above the ground surface. The tank walls and floor are 300 mm thick.
4. Sludge fed to digester = 15 m^3/d at 14°C
5. Outside temperature = –15°C
6. Average ground temperature = 5°C
7. Sludge in tank is to be maintained at 35°C
8. Assume a specific heat of the sludge = 4200 J/kg·°C
9. Sludge contains 4% solids
10. Assume a cone-shaped cover with center 0.6 m above digester top, and a cone-shaped bottom with center 1.2 m below bottom edge.

13–10 A wastewater treatment plant is considering options for expanding their existing anaerobic digestion system to handle increased sludge production. The plant currently sends 25,000 kg/d of 5 percent thickened sludge to anaerobic digestion for stabilization. The plant currently has three 6200 m^3 digesters but normally only operates two keeping the third as a redundant unit. The digesters are high rate complete mix digesters with no decanting.

a. Using one of the data sets as selected by the instructor determine how much additional digester volume would be required to maintain at least 15 d HRT at future build out with one digester out of service?

b. One option also being considered is adding a thermal hydrolysis process to increase the capacity of the existing anaerobic digesters to avoid building new digester volume. Assuming the solid content going to the digester is 9 percent (after thermal hydrolysis and dilution) would the existing digestion volume be sufficient to meet digestion requirements? What would be the theoretical steam requirement for the thermal hydrolysis system assuming the temperature out of the thermal hydrolysis process is 110°C?

Assume the specific gravity of the thickened sludge and diluted hydrolyzed sludge going to digestion is 1.03.

		Data set			
Item	Unit	1	2	3	4
Future sludge loading	kg/d	55,000	60,000	50,000	58,000
Raw sludge temperature	°C	10	15	20	12

13–11 A small wastewater plant currently utilizes aerobic digestion to stabilize their waste activated sludge prior to Class B liquid land application and they are looking at ways the system can be expanded to handle future loads. The plant currently sends waste activated sludge to a single stage complete mix aerobic digester with no decanting. Using one of the data sets to be selected by instructor, recommend possible options to consider in the upgrade if aerobic digestion is to be maintained. In the example use the following assumptions.

1. The winter and summertime liquid temperatures are 15 and 25 d, respectively
2. The system must be able to achieve $>40\%$ VSR in the winter and meet Class B requirements (SRT > 60 d at 15°C)
3. Specific Gravity of liquid waste activated sludge is 1.01
4. Air temperature in diffused air system is 20°C.
5. Assume a diffused air oxygen transfer efficiency of 10%

In the calculation be sure to comment on volume requirements, SRT, aeration requirements and mixing. Note any perceived advantage or disadvantage with options.

		Data set		
Item	Unit	1	2	3
Current sludge loading	kg/d	500	750	900
Waste activated sludge	% TS	1	0.8	1.3
Future sludge loading	kg/d	1500	2000	3000
Digester volume	m^3	3000	5600	2400
Current blower size	m^3/min	90	165	125

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14

Biosolids Processing, Resource Recovery and Beneficial Use

WORKING TERMINOLOGY

Term	Definition
Aerated static pile composting	A method of composting where the sludge or biosolids and bulking agent are mixed and distributed in a long pile over a grid of air piping which provides the air for reaction.
Belt-filter press	A device that uses a series of porous moving belts revolving over a series of pulleys to drain water from sludge or biosolids.
Carbon footprint	A measure of the impact human activities have on the environment in terms of the amount of greenhouse gases produced, measured in units of carbon dioxide equivalent.
Centrifuge	A dewatering device that relies on centrifugal force to separate particles of varying density (e.g., water and solids).
Composting	A stabilization process that relies on the aerobic decomposition of organic matter in sludge and biosolids by bacteria and fungi.

Term	Definition
Conditioning	A chemical, physical, or biological process designed to improve the thickening or dewatering characteristics of sludge or biosolids.
Decanting	Separating liquid from settled sludge or biosolids by drawing or pouring off the upper layer of liquid after the sludge or biosolids have settled.
Dewatering	A process, usually by belt-filter press or centrifuge that removes a portion of the water contained in solids. Dewatering is distinguished from thickening in that the resulting dewatered cake may be handled as a solid, not as a liquid.
Dissolved-air flotation	A clarification process in which minute bubbles become attached to flocculated material, float to the surface, and are removed by skimming. Heavier solids settle and are removed by mechanical scrapers.
Filter press	A dewatering device in which water is forced from semi-solid materials under high pressure.
Fluidized-bed incinerator	A furnace that uses a high-temperature gas to fluidize solid particles (usually sand and waste sludge and biosolids) to produce and sustain combustion.
Gravity-belt thickener	A thickening device that uses a porous filter belt to promote gravity water drainage.
Heat drying	The application of heat to evaporate water and reduce the moisture content in biosolids below that achievable by conventional dewatering methods.
Humus	Sludge removed from trickling filters.
Incineration	The reduction of the volume of a solid by the thermal destruction of organic matter.
In-vessel composting	A method of composting, mainly proprietary, that occurs inside an enclosed vessel or container.
Mass balance	A method for analyzing physical systems based on the law of conservation of mass.
Multiple-hearth incinerator	An incinerator consisting of numerous hearths that is used for the thermal destruction of organic sludge or biosolids.
Reed bed	A treatment system in which biosolids are used to grow reeds, which in turn utilize the water, nitrogen, and other nutrients to stabilize and dewater the biosolids.
Rheology	The flow properties of a liquid (generally biosolids and sludge) that include elasticity, viscosity, and plasticity.
Rotary drum thickener	A rotating cylindrical screen used to thicken liquid streams of sludge and biosolids.
Rotary press	A sludge or biosolids dewatering device in which the material to be dewatered flows through a channel that is bound between two rotating screens; filtrate passes through the screens and the dewatered material continues through the channel.
Sidestream	A portion of the wastewater flow that has been diverted from the main treatment process flow for specialized treatment (see Chap. 15).
Sludge drying beds	Devices used for the dewatering and drying of sludge and biosolids in which a semi-solid solution is spread over a porous (e.g., sand) or impervious medium and allowed separate and air dry or decant.
Solids	A term often used as a replacement for sludges that have not been stabilized by physical, chemical or biological treatment. The term solids is not used as a substitute for sludge in this chapter. The mass of dry material in sludge is referred to as the solids content.
Thickener	A tank, vessel, or device where residuals or a slurry are concentrated by removing a portion of the water.
Windrow composting	A method of composting where sludge or biosolids are mixed with a bulking agent and arranged in windrows (long piles) that are turned over periodically and remixed mechanically.

Processes used to reduce organic content and to render the processed sludge suitable for reuse or final disposal were considered in Chap. 13. The focus of Chap. 14 is on the many processes used for recovery and beneficial use of biosolids. Before biosolids can be processed or used beneficially, they are typically dewatered to reduce the volume that must be handled. However, to achieve effective dewatering, sludges and biosolids must be conditioned for enhanced water removal. The conditioning of sludge and biosolids is discussed in Sec. 14–1. Commercially available dewatering methods are identified and discussed in Sec. 14–2. Drying the dewatered sludge or biosolids is for the purpose of removing more water and further stabilization; producing granular material that is beneficially used as a fertilizer or energy source is presented in Sec. 14–3. Thermal oxidation of the sludges for destroying harmful constituents and producing ash-like material that can also be used beneficially is discussed in Sec. 14–4. Composting for stabilizing the biosolids and producing fertilizer like material is discussed in Sec. 14–5. Conveyance and storage of biosolids are discussed in Sec. 14–6. The preparation of solids balances for treatment facilities is described in Sec. 14–7. Biosolids resource recovery and energy recovery are covered in Sec. 14–8. The application of biosolids to land and conveyance and storage of biosolids after processing is discussed in Sec. 14–9.

14–1 CHEMICAL CONDITIONING

Sludges and biosolids are conditioned expressly to improve their dewatering characteristics. For proper mechanical dewatering systems such as centrifugation, belt filter press, rotary press, screw press, and pressure filter press, as discussed in Sec. 14–2 , sludges and biosolids must be chemically conditioned. Chemical conditioning results in the flocculation (aggregation) of the sludge and biosolids to achieve efficient solid-liquid separation. Other conditioning methods such as heat treatment and freeze-thaw, have also been used to a limited extent or experimentally and are discussed in previous editions of this textbook. Chemical conditioning uses inorganic chemicals and water soluble polymers, or both. Inorganic conditioners, which include lime, ferric chloride, ferrous sulfate, aluminum sulfate, and aluminum chloride, are used mainly for recessed chamber filter press discussed in Sec. 14–2. However, in certain applications when biosolids are hard to dewater and require high polymer dosages, iron salts are often used (Abu-Orf et al., 2001). Polymers do not increase the dry solids content, while iron salts and lime can increase the dry solids content by 20 to 30 percent. A general discussion of polymers, the most used conditioning agents for mechanical dewatering, is the primary focus of this section. Following a general description of polymers, polymer characteristics, factors affecting conditioning, dosage determination, mixing, and polymer makeup and feeding are also considered.

Polymers

Water-soluble polymers are used most commonly for sludges and biosolids conditioning prior to mechanical dewatering. Polymers are also termed *organic polyelectrolytes* as they dissociate upon addition to water into negatively and positively charged species. Polymers are chains of individual monomer units and linked together in linear, branched, or structured configuration with functional groups located along the chains that determine the charge of the polymer. Because sludges and biosolids are mainly negatively charged, cationic polymers are used most commonly for conditioning. Cationic polyacrylamide (PAM) is the backbone of most, if not all, commercially available polymers for biosolids and sludge conditioning (WEF, 2012). These polymers upon dissociation release anions such as chloride leaving behind long chains of high molecular weight polymeric molecules that are positively charged. These long chains flocculate the suspended solids and the colloidal

Table 14–1

Distribution of polymer charge density and molecular weight[a]

Relative charge density and molecular weight	Charge density, mole %	Relative molecular weight
Very high	> 70–100	> 6,000,000–18,000,000
High	> 40–70	> 1,000,000–6,000,000
Medium	> 10–40	> 200,000–1,000,000
Low	< 10	< 200,000

[a] Adapted from WEF (2012).

material in the sludge biofloc resulting in better solids liquid separation, and is the reason polymers are called flocculants. The form of the cationic PAM can be dry, liquid, or emulsion. The choice of form of polymer to use depends on (1) achieving desired dewatering performance; (2) cost effectiveness; (3) space for storage and handling of the neat product; (4) requirements for polymer make down, aging, and feeding equipment; and (5) safety considerations.

Polymers are proprietary chemicals and vary according to electrical charge, charge density, molecular weight and molecular structure (WEF, 2012). Polymers that are used in sludge or biosolids conditioning are usually cationic, high charge density, and high molecular weight. Degrees of polymer charge density and molecular weight are reported in Table 14–1 (WEF, 2012). The molecular structure can be straight, branched, or structured. In high shear dewatering devices like centrifuges, branched or structured polymers are used. Polymer characterization methods to determine charge density and molecular weight are not commonly practiced in wastewater treatment plants. However, for large treatment facilities that spend significant amounts of money on polymers, onsite characterization is recommended to ensure receiving consistent performing product (Abu-Orf et al., 2009).

Factors Affecting Polymer Conditioning

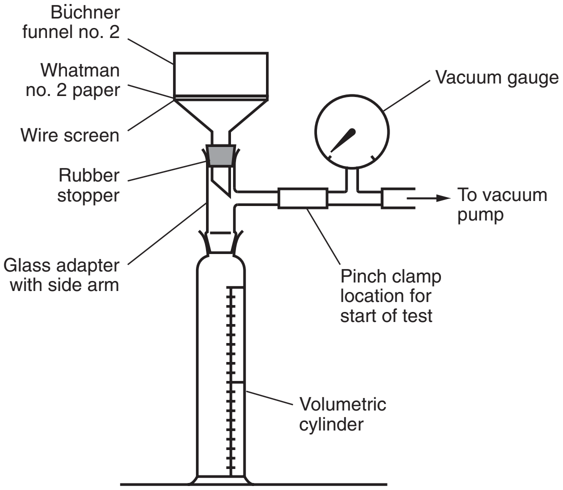
The selection of the type and dosage of polymers depends on the properties of the sludge and biosolids, mixing conditions between the chemicals and the sludge, and dewatering devices to be used. Important sludge and biosolids properties include source, solids concentration, electrical charge of sludge, biopolymer content, and rheology (WEF, 2012). The probable range of polymer doses required can be estimated based on the source of sludge or biosolids (e.g., primary sludge, waste activated sludge, and digested biosolids). Increasing the volatile solids destruction in aerobic and anaerobic digestion increases the polymer demand (Novak et al., 2004). Solids concentrations will affect the polymer dosage and its dispersion. The anionic electrical charge of the sludge or biosolids is an important factor in determining the cationic amount of polymer used. The biopolymer (protein and polysaccharides) content in the sludge and biosolids has a significant impact on polymer dose and dewatering. A linear relationship was found between the colloidal biopolymer content in the biosolids and the optimum polymer dose for dewatering and the dewatering potential (Novak et al., 2004). The method of dewatering affects the selection of the conditioning chemical because of the differences in mixing equipment used by various vendors and the characteristics of particular methods of dewatering.

Polymer Dosage Determination

The polymer dosage required for conditioning can be determined in the laboratory and need to be verified in full-scale trials. Laboratory tests used for selecting polymer dosage include the Buchner funnel test for the determination of specific resistance of filtration

Figure 14–1

Buchner funnel test apparatus used for the determination of the specific resistance of sludges and biosolids.



(see Fig. 14–1), capillary suction time test (CST) (Standard Methods 2012), and the standard jar test (ASTM, 2008). The Buchner funnel test is a method of testing sludge drainability or dewatering characteristics using various conditioning agents. The capillary suction test relies on gravity and the capillary suction of a piece of a standard thick filter paper to draw out water from a small sample of conditioned sludge or biosolids. The standard jar test, the easiest method to use, consists of testing standard volumes of sludge samples (usually 1 L) with different conditioner concentrations, followed by rapid mixing, flocculation, and settling using standard jar test apparatus. Other methods used successfully in laboratory and full scale for determination of optimum polymer dose for conditioning and dewatering include charge of the liquid stream (centrate or filtrate) as measured by streaming current detector (Abu-Orf and Dentel, 1997), viscosity of the liquid stream (Abu-Orf et al., 2003), and rheology of the conditioned sludge or biosolids (Abu-Orf and Ormeci, 2005). Laboratory or pilot-scale testing is recommended to determine the types of polymer and dose required, particularly for sludge and biosolids that may be difficult to dewater.

Mixing

The intimate and uniform mixing of sludge or biosolids and polymer is essential for proper conditioning. The intensity of the mixing must not break the flocculated material after it has formed, and the detention should be kept to a minimum so that sludge reaches the dewatering unit as soon after conditioning as possible. Mixing requirements vary depending on the dewatering method used. A separate mixing and flocculation tank is provided ahead of pressure filters; a separate flocculation tank may be provided for a belt filter press, or the polymer may be added directly to the sludge feed line of the belt filter press unit; and inline mixers are usually used with a centrifuge. It is generally desirable to provide at least two locations for the addition of conditioning chemicals.

In general, it has been observed that the type of sludge has the greatest impact on the quantity of chemical required. Difficult-to-dewater sludges that require larger doses of chemicals generally do not yield as dry a cake and have poorer quality of filtrate or centrate. Sludge types, listed in the approximate order of increasing conditioning chemical requirements, are as follows:

1. Untreated (raw) primary sludge
2. Untreated mixed primary and trickling filter sludge

3. Untreated mixed primary and waste-activated sludge
4. Anaerobically digested primary sludge
5. Anaerobically digested mixed primary and waste activated sludge
6. Aerobically digested mixed primary and waste activated sludge
7. ATAD biosolids
8. Aerobically digested waste activated sludge
9. Untreated waste activated sludge

Typical levels of polymer addition for various types of sludges using belt-filter press centrifuge, rotary press, and screw press dewatering can be found in Sec. 14–2. Actual dosages in any given case may vary considerably from the indicated values. Polymer dosages will also vary greatly depending on the molecular weight, ionic strength, and activity levels of the polymers used. Manufacturers should be consulted for applicability and dosage information.

Conditioning Makeup and Feed

Chemicals are most easily applied and metered in the liquid form. Dissolving tanks are needed if the chemicals are received as dry powder. In most plants, these tanks should be large enough for at least one day's supply of chemicals and should be furnished in duplicate. In large plants, tankage sufficient for one shift is usually adequate. The tanks must be fabricated or lined with corrosion-resistant material. Polyvinyl chloride, polyethylene, and rubber are suitable materials for tank and pipe linings for acid solutions. Metering pumps must be corrosion-resistant. These pumps are generally of the positive displacement type with variable speed or variable stroke drives to control the flowrate.

14–2 DEWATERING

Dewatering is a physical unit operation used to separate the solid matter and water in the sludge or biosolids resulting in a high solids content stream called "cake" and a liquid stream. The liquid stream contains fine, low-density solids and a high concentration of nutrients when anaerobically digested sludge is dewatered and is typically returned to the wastewater treatment system or treated separately (sidestream treatment, see Chap. 15) to reduce nutrient loading to the main treatment system. For effective solids liquid separation, chemical conditioning is required. Increasing the solids content of sludge and biosolids is mainly practiced for one or more of the following reasons:

1. The costs for trucking sludge and biosolids to the ultimate disposition site become substantially lower when the volume is reduced by dewatering.
2. Dewatered sludge and biosolids are generally easier to handle than thickened or liquid sludge. In most cases, dewatered sludge may be shoveled, moved about with tractors fitted with buckets and blades, and transported by belt conveyors.
3. Dewatering is required normally prior to the incineration of the sludge to increase the calorific value by removal of excess moisture.
4. Dewatering is required before composting to reduce the requirements for supplemental bulking agents or amendments.
5. Dewatering is required prior to thermal drying as it is cost effective to remove the water mechanically or by other means, compared to evaporating the water during drying.
6. In some cases, removal of the excess moisture may be required to render biosolids odorless and nonputrescible.
7. Dewatering is required prior to landfilling sludge and biosolids in monofills to reduce leachate production at the landfill site.

In mechanical dewatering devices, mechanically assisted physical means are used to dewater the sludge more quickly. Other dewatering devices rely on the application of electric and heat energy. Following a brief overview of dewatering, each of the major dewatering technologies is considered in this section.

Overview of Dewatering Technologies

To provide a perspective on the discussion of the individual dewatering technologies, it will be useful to first consider the fundamental principles of dewatering; some important factors in the selection of dewatering technologies, including the advantages and disadvantages of the various technologies; and the need for bench and pilot-sate testing.

Fundamental Principles of Dewatering. When considering the dewatering of sludge or biosolids, it is important to consider the various forms of water associated with the biosolids. In a relatively simplified overview, the four types of water associated with sludge, as proposed by Tsang and Vesiland (1990) and others and illustrated on Fig. 14–2(a), are (1) free water, (2) interstitial water, (3) surface water, and (4) bound water. Water not attached to particles that can be removed by gravitational forces, filtration, and centrifugation is known as *free water*. Water trapped within the sludge matrix is known as *interstitial water.* Water bound to the sludge particles by adsorption and adhesion is known as *surface water*. Intercellular and chemically bound water is known as *bound water*. The form of water that can be removed by the various dewatering technologies is illustrated on Fig. 14–2(b). Free and a portion of the interstitial water can be removed by physical means. Electro-dewatering can be used to remove interstitial and a portion of the bound water, depending on the adsorption forces. Thermal drying is required to remove the majority of bound water (Mahmoud et al., 2010).

Important Factors in Technology Selection. The selection of the dewatering device is determined by the type of sludge or biosolids to be dewatered, characteristics of the dewatered product, downstream processing, ultimate disposition, and the space available. The dewatering processes that are used commonly include centrifuges;

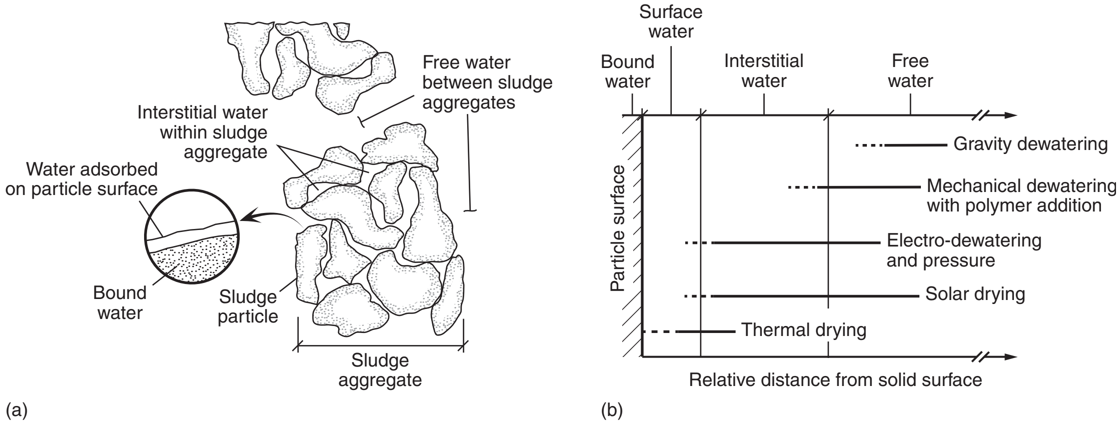


Figure 14–2

Overview of dewatering treated biosolids: (a) different forms of water associated with treated biosolids as proposed by Tsang and Vesiland (1990) and (b) potential operating regions for various technologies used for dewatering biosolids. The dashed portion represents the maximum potential achievable.

belt filter presses; rotary presses; screw presses; recessed-plate filter presses; electro-dewatering, a new innovative process; drying beds; and lagoons. Vacuum filtration, often used in the past for municipal sludge dewatering, has essentially been replaced by alternative mechanical dewatering equipment. The advantages and disadvantages of the various methods of sludge dewatering presented in this chapter are summarized in Table 14–2.

Table 14–2

Comparison of alternative methods for dewatering various types of sludge and biosolids[a]

Dewatering method	Advantages	Disadvantages
Solid bowl centrifuge	• Clean appearance, minimal odor problems, fast startup and shut down capabilities • Easy to install • Produces relatively dry sludge cake • Low capital cost-to-capacity ratio	• Scroll wear potentially a high maintenance problem • Requires grit removal and possibly sludge grinder in the feed stream • Skilled maintenance personnel required • Moderately high suspended solids content in centrate • Cannot observe dewatering zone to optimize/adjust performance
Belt filter press	• Low energy requirements • Relatively low capital and operating costs • Less complex mechanically and is easier to maintain • High pressure machines are capable of producing very dry cake • Minimal effort required for system shut down	• Hydraulically limited in throughput • Requires sludge grinder in feed stream • Very sensitive to incoming sludge feed characteristics • Short media life as compared to other devices using cloth media • Automatic operation generally not advised
Recessed plate filter press	• Highest cake solids concentration • Low suspended solids in filtrate • Simple operation • High solids capture rate	• Batch operation • High equipment cost • High labor cost • Special support structure requirements • Large floor area required for equipment • Skilled maintenance personnel required • Additional solids due to large chemical addition require disposal • Limitations on filter cloth life
Rotary Press	• Low speed 0.5 to 2.5 rev/min • Low noise $<$ 68 dBA • Enclosed design contains odors and aerosols • Relatively low energy use drive motor ranges from 0.56 to 15 kW (0.75 to 20 hp) depending on size of unit • Overdosing polymer does not clog screen and hinder dewatering • Washwater only used during shut down of system • Low shearing force reduces odors in dewatered cake stockpile	• Relatively large footprint per unit volume of dewatering capacity • Capacity limitations will require multiple units for wastewater facilities treatment facilities $> 19{,}000\ \text{m}^3/\text{d}$ (> 5 Mgal/d) • Cannot observe dewatering zone to optimize/adjust performance

(*continued*)

Table 14–2 (*Continued*)

Dewatering method	Advantages	Disadvantages
Screw Press	• Low speed 0.3 to 1.5 rev/min • Low noise < 68 dBA • Enclosed design with hinged access doors contains odors and aerosols • Low energy use drive motor ranges from 0.37 to 3.7 kW (0.5 to 5 hp) depending on size of unit • Overdosing polymer does not clog screen and hinder dewatering • Low shearing force reduces odors in dewatered cake stockpile	• Capacity limitations will require multiple units for wastewater facilities treatment facilities 19,000 m^3/d (> 5 Mgal/d) • Washwater required periodically throughout operating cycle • Cannot observe dewatering zone to optimize/adjust performance
Electro-dewatering	• Automatic operation • Good results for difficult sludge and biosolids • Mechanics are simple and easy to maintain • Odor improvement and pathogen kill on the sludge and biosolids • Some flexibility to incoming sludge characteristics • 3–5 times more energy efficient than dryers	• Batch operation • Moderate to high capital costs • Not particularly suited for larger plant 75,700 m^3/d (20 Mgal/d) and above • Limited final dryness achievable (max 45 to 50 percent DS) • Difficult to predict performance without bench scale testing • New technology • Requires odor treatment for the process off gases • Require predewatering, range of feed between 10 and 25 percent • Operational cost sensitive to local electricity tariff
Sludge drying beds	• Lowest capital cost method where land is readily available • Small amount of operator attention and skill required • Low energy consumption • Little to no chemical consumption • Less sensitive to sludge variability • Higher solids content than mechanical methods	• Requires large area of land • Requires stabilized sludge • Design requires consideration of climatic effects • Sludge removal is labor intensive
Sludge lagoons	• Low energy consumption • No chemical consumption • Organic matter is further stabilized • Low capital cost where land is available • Least amount of skill required for operation	• Potential for odor and vector problems • Potential for groundwater pollution • More land intensive than mechanical methods • Appearance may be unsightly • Design requires consideration of climatic effects

[a] Adapted in part from U.S. EPA (2000).

For smaller plants where land availability is not a problem, drying beds or lagoons are generally used. Conversely, for larger facilities or facilities situated on constricted sites, mechanical dewatering devices are often chosen. Odor control is an important design consideration as the level of odor release varies based on the type of sludge and the mechanical equipment selected. High shear dewatering and conveyance equipment can increase odor release, especially from anaerobically digested sludge (WERF, 2003) (see also Sec. 16–4 in Chap. 16).

Need for Bench and Pilot-Scale Testing. When particular types of sludge or biosolids must be dewatered mechanically, it is often difficult or impossible to select the optimum dewatering device and polymer dosage without conducting bench-scale or pilot studies. Bench-scale testing is usually conducted by manufacturers of dewatering equipment to narrow down the types of polymers and doses to be used in pilot testing. Trailer-mounted, full-size equipment is available from several manufacturers for field-testing purposes. In selecting the type of mechanical dewatering to be used, it is important not to rely on published industry standard performance information and data. Side-by-side pilot testing should be undertaken to select the most cost-effective dewatering device, suitable for the treatment plant sludge or biosolids. With side-by-side testing, dewatering devices can be compared on the same sludge or biosolids, as it is well established that sludge characteristics vary seasonally, if not daily, at some plants. Pilot testing should be designed carefully to determine solids throughput, optimum polymer dose, percent cake solids, and percent solids recovery, which are important factors in comparing capital and operation costs of the various dewatering devices.

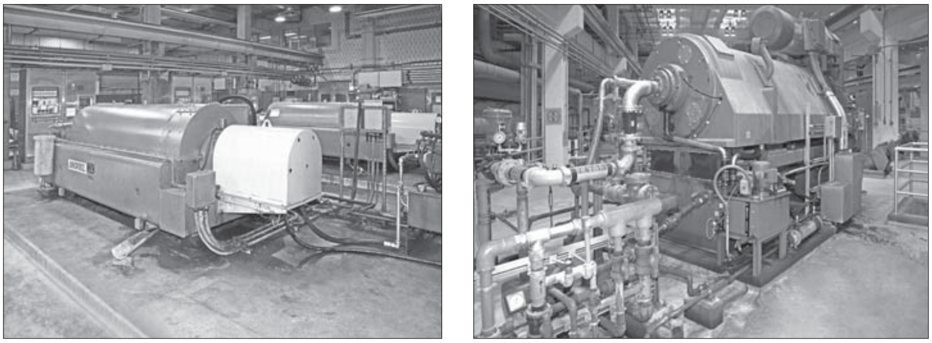
Centrifugation

The centrifugation process is used widely in industry for separating liquids of different density, thickening slurries, or removing solids. The process is applicable to the dewatering of wastewater sludges and has been used widely in both the United States and Europe. Solid-bowl centrifugal devices used for thickening sludge (see Sec. 13–6) may also be used for sludge and biosolids dewatering. In this section, standard solid-bowl and "high-solids" centrifuges are discussed. The high-solids centrifuge is a modification of the standard centrifuge.

Solid-Bowl Centrifuge. In the solid-bowl machine (see Figs. 14–3 and 14–4), biosolids or sludge is fed at a constant flowrate into the rotating bowl, where it separates into a dense cake containing the solids and a dilute liquid stream called "centrate." The centrate is returned to the wastewater treatment system or treated separately, if necessary.

Figure 14–3

View of typical solid bowl centrifuge dewatering installations.


(a)

(b)

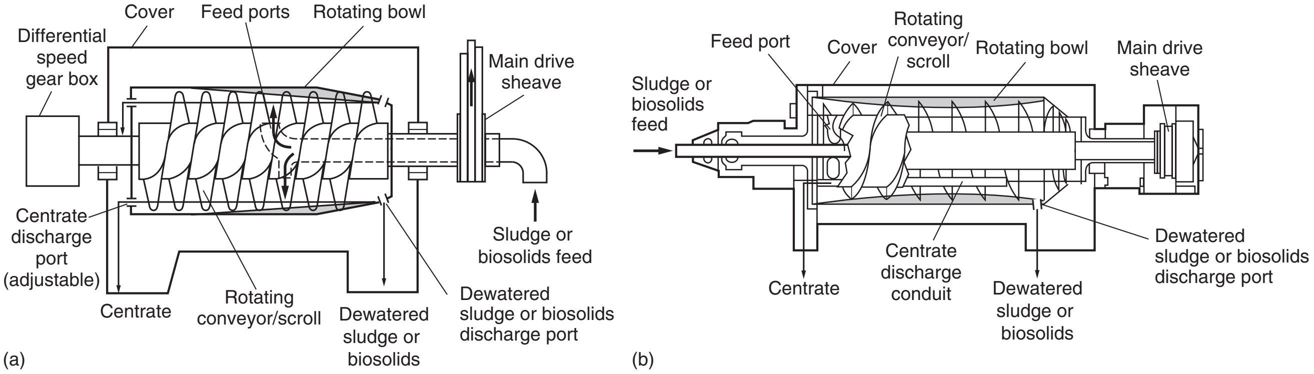


Figure 14–4

Schematic diagrams of two solid bowl centrifuge configurations for dewatering sludge: (a) countercurrent and (b) cocurrent.

The sludge cake is discharged from the bowl by a screw feeder into a hopper or onto a conveyor belt. Depending on the type of sludge or biosolids, solids concentration in the cake varies generally from 20 to 30 percent range. Cake concentrations above 25 percent are desirable for processing by incineration, drying, or by hauling to offsite processing, land application or disposition at a sanitary landfill.

Solid-bowl centrifuges are suitable generally for a variety of dewatering applications. Chemicals for conditioning are added to achieve the desired dewatering performance including cake solids and centrate quality and usually added to the feed line or within the bowl of the centrifuge. Dosage rates for conditioning with polymers vary from 1.0 to 25 g/ kg (2 to 50 lb/ton) of sludge (dry solids basis). Typical performance data for solid-bowl centrifuges are reported in Table 14–3.

Table 14–3

Typical dewatering performance for solid bowl centrifuges for various types of sludge and biosolids[a]

Type of feed	Feed solids, %	Cake solids, %	Polymer use, lb/ton dry TS	Polymer use, g/kg dry TS	Solids capture, %
Untreated sludge					
Primary	4–8	25–50	5–10	2.5–5	95+
Primary + WAS	3–5	25–35	5–16	2.5–8	95+
WAS	1–2	16–25	15–30	7.5–15	95+
Anaerobically digested biosolids					
Primary	2–5	25–40	8–12	4–6	95+
Primary + WAS	2–4	22–35	15–30	7.5–15	95+
Aerobically digested WAS	1–3	18–25	20–30	10–15	95+
ATAD biosolids	2–5	20–30	25–45	12.5–22.5	95+

[a] Adapted in part from U.S. EPA (2000) and feedback from centrifuge vendors.

High-Solids Centrifuge. High-solids (also called "high-torque") centrifuges are modified solid-bowl centrifuges that are designed to produce a dryer solids cake. These units have a slightly longer bowl length to accommodate a longer "beach" section, a lower differential bowl speed to increase residence time, and a modified scroll to provide a pressing action within the beach end of the unit. In some cases, the high-solids units are capable of achieving solids contents in excess of 30 percent in dewatering municipal wastewater sludges, although a higher polymer usage may be required.

Design Considerations. Centrifugation design is based on either helical or axial flow. Two basic designs of helical centrifuges are used: countercurrent flow and cocurrent flow [see Figs. 14–4 (a) and (b)]. The main difference in the designs is the location of the feed ports, removal of centrate, and internal flow patterns of the liquid and solid phases. In the countercurrent design, the feed slurry enters axially at the junction of the cylindrical conical section; solids travel to the conical end while the liquid phase moves in the opposite direction. In axial flow countercurrent centrifuges, flights are mounted on spokes in the path from the feed zone to the centrate dams. These flights reduce the flow velocity, resulting in enhanced separation. This design allows centrifuges to handle variations of the feed rate without needing to change the dams. As a result, axial flow is the most common type of centrifuge design (WEF, 2012). In the cocurrent design, the solid phase travels the full length of the bowl as does the liquid phase. Cocurrent centrifuge designs are seldom used because of maintenance problems (WEF, 2012).

Process Variables. Process variables affecting centrifuge performance, as measured by the sludge cake solids and TSS recovery, include feed flowrate, rotational speed, differential speed of the scroll, depth of the settling zone, conditioning dose, and the physicochemical properties of the suspended solids and suspending liquid. Important properties are particle size and shape, particle density, temperature, and liquid viscosity.

Selection of units for plant design is dependent on manufacturer's rating and performance data. Several manufacturers have portable pilot units, which can be used for field testing if sludges or biosolids are available. Unfortunately, biosolids or sludges from similar treatment processes but in different localities may differ markedly from each other. For this reason, pilot-plant tests as previously discussed should be conducted whenever possible before final design decisions are made.

Other Design Considerations. The area required for a centrifuge installation is less than that required for other dewatering devices of equal capacity, and the initial cost is lower. Higher power usage costs will partially offset the lower initial cost. Special consideration must also be given in providing sturdy foundations and soundproofing because of the vibration and noise that result from centrifuge operation. An adequate electric power source is required because large motors may be used.

Because centrifuges are enclosed, on-site odor generation may be better contained as compared to other types of dewatering systems. Ventilation of the centrifuge facility to control potential odors and moisture accumulation should be provided, however. On the other hand, cake solids produced from high solids centrifuges are more odorous compared to other dewatering devices, which could adversely affect beneficial use methods such as land application. Moreover, sudden increase in pathogen indicator organisms in cake solids after anaerobic digestion was observed after centrifugation (WERF, 2008a).

Combined Centrifuge Process. Technologies that combine centrifuge dewatering with flash air drying are available commercially. It is reported that these systems are able to process sludges or biosolids with total solids ranging from 2 to 7 percent and produce a

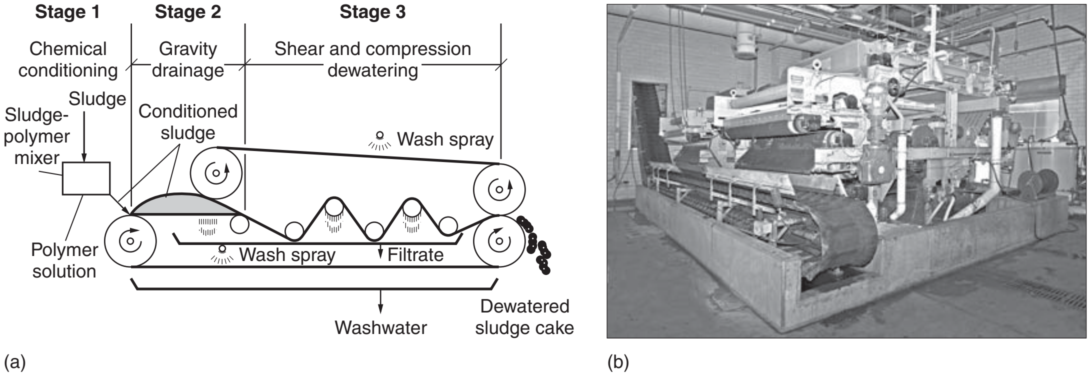


Figure 14–5

Belt press dewatering: (a) schematic of the three basic stages of belt press dewatering (b) view of a typical installation.

product with 60 to 90 percent dryness. Currently (2013), there is no installation in North America.

Belt-Filter Press

Belt-filter presses (BFPs) are continuous-feed dewatering devices that use the principles of gravity drainage and mechanically applied pressure to dewater chemically conditioned sludges or biosolids (see Fig. 14–5). The belt-filter press was introduced in the United States in the early 1970s and has become one of the predominant sludge dewatering devices. It has proven to be effective for almost all types of municipal wastewater sludge and biosolids.

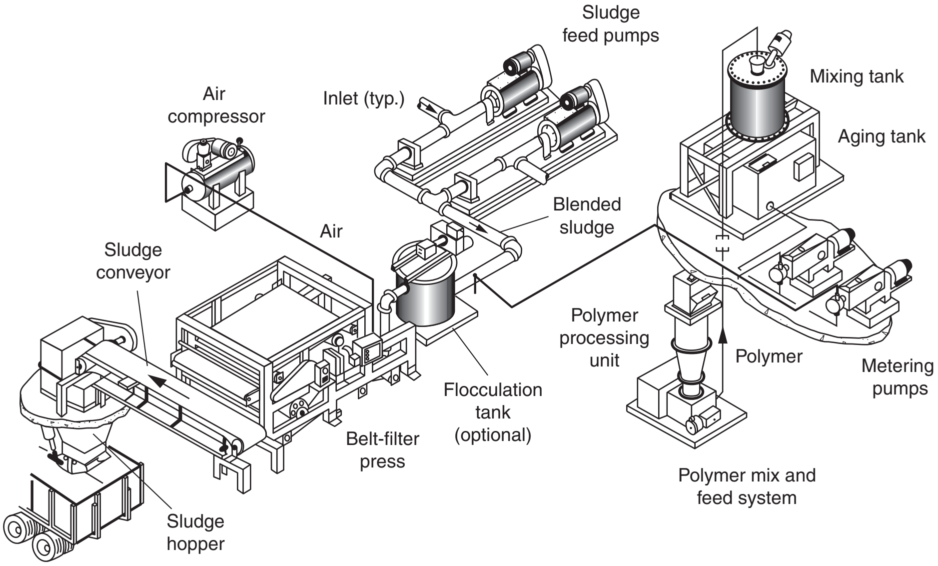
Description. In most types of BFPs, conditioned sludge or biosolids are first introduced on a gravity drainage section where it is allowed to thicken. In this section, a majority of the free water is removed by gravity. Following gravity drainage, pressure is applied in a low-pressure section, where the sludge is squeezed between opposing porous cloth belts. On some units, the low-pressure section is followed by a high-pressure section where the sludge is subjected to shearing forces as the belts pass through a series of rollers. The squeezing and shearing forces thus induce release of additional quantities of water from the sludge. Many vendors also offer systems with three belts which allow for independent control of the gravity section from the pressure section. The three belt system is ideal for dilute sludges. Improved dewatering can be achieved with this design as each zone can be optimized independently. The final dewatered sludge cake is removed from the belts by scraper blades.

System Operation and Performance. A typical BFP system consists of sludge feed pumps, polymer feed equipment, belt filter press, sludge cake conveyor, and support systems (sludge feed pumps, washwater pumps, and compressed air). Most units do not use a conditioning tank. A schematic diagram of a typical two belt-filter press installation is shown on Fig. 14–6.

Many variables affect the performance of the BFP: sludge or biosolids characteristics, method and type of chemical conditioning, pressures developed, machine configuration (including gravity drainage), belt porosity, belt speed, and belt width. The BFP is sensitive to wide variations in sludge characteristics, resulting in improper conditioning and reduced dewatering efficiency. Blending facilities should be included in the system design where

Figure 14–6

Schematic diagram of a complete belt press dewatering system.



the sludge or biosolids characteristics are likely to vary widely. Based on actual operating experience, it has been found that the solids throughput is greater and the cake dryness is improved with higher solids concentrations in the feed sludge or biosolids. Typical BFP performance data for various types of sludge and biosolids are reported in Table 14–4.

Table 14–4

Typical dewatering performance for belt filter presses[a]

Type of feed	Dry feed solids, %	Loading per meter of belt length		Dry polymer[b], g/kg dry solids	Cake solids, %	
		L/min	kg/h		Typical	Range
Untreated sludge						
Primary	4–8	230–640	1130–1590	1.5–2.5	30	26–35
WAS	1–2	190–380	180–340	5–10	16	12–20
Primary plus WAS	3–5	150–450	340–820	3–5.5	23	15–25
Primary plus trickling filter	3–6	150–450	360–910	3–7	27	16–30
SBR	1–2	190–380	250–360	5–7.5	16	12–19
MBR	1–2	260–420	230–320	5.5–10	15	11–18
Anaerobically digested:						
Primary	2–5	230–610	680–910	2–5	28	24–35
WAS	2–3	110–340	230–410	4–10	20	13–23
Primary plus WAS	2–4	150–450	320–540	4–8.5	24	15–28
Aerobically digested WAS	1–3	150–340	250–410	6–10	18	12–22
ATAD	2–5	110–490	360–590	5–12.5	19	12–22

[a] Based on feedback from belt filter press vendors.

[b] Polymer needs based on high molecular weight polymer (100 percent strength, dry basis).

Design Considerations. BFPs are available in metric sizes from 0.5 to 3.0 m in belt width. The most common size used for municipal sludge applications is 2.0 m. Solids loading rates vary from 180 to 1600 kg/m·h (400 to 3500 lb/m·h) depending on the sludge type and feed concentrations. Hydraulic throughput based on belt width ranges from 110 to 640 L/m·min (30 to 170 gal/m·min). Design of a BFP is illustrated in Example 14–1.

Safety considerations in design should include adequate ventilation to remove hydrogen sulfide or other gases mainly when dewatering undigested sludges, and equipment guards to prevent loose clothing from being caught between the rollers.

EXAMPLE 14–1 **Belt-Filter Press Design** A wastewater treatment plant produces 75,000 L/d of thickened biosolids containing 3 percent solids. A belt-filter press installation is to be designed based on a normal operation of 8 h/d and 5 d/wk, a belt-filter press loading rate of 275 kg/m·h, and the following data. Compute the number and size of belt-filter presses required and the expected solids capture, in percent. Determine the daily hours of operation required if a sustained 3-d peak solids load occurs.

1. Total solids in dewatered sludge = 22 percent.
2. Total suspended solids concentration in filtrate = 900 mg/L = 0.09 percent.
3. Washwater flowrate = 90 L/min per m of belt width.
4. Specific gravities of sludge feed, dewatered cake, and filtrate are 1.02, 1.07, and 1.01, respectively.

Solution

1. Compute average weekly sludge production rate.

$$\text{Wet biosolids} = (75{,}000\text{ L/d})(7\text{ d/wk})(10^3\text{ g/1 L})(1\text{ kg/}10^3\text{ g})(1.02) = 535{,}500\text{ kg/wk}$$

$$\text{Dry solids} = (535{,}500\text{ kg/wk})(0.03) = 16{,}065\text{ kg/wk}$$

2. Compute daily and hourly dry solids-processing requirements.

$$\text{Daily rate} = (16{,}065\text{ kg/wk})\left(\frac{1\text{ wk}}{5\text{ operating d}}\right) = 3213\text{ kg/d}$$

$$\text{Hourly rate} = \frac{(3213\text{ kg/d})}{(8\text{ h per operating d})} = 401.6\text{ kg/h (8 h operating d)}$$

3. Compute belt-filter press size 3

$$\text{Belt width} = \frac{(401.6\text{ kg/h})}{(275\text{ kg/m·h})} = 1.46\text{ m}$$

Use one 1.5-m belt-filter press and provide one identical size for standby.

4. Compute filtrate flowrate by developing solids balance and flow balance equations.
   a. Develop daily solids balance equation.

Solids in sludge feed = solids in sludge cake + solids in filtrate

$$3213\text{ kg/d} = (S\text{ kg/d})(0.22) + (F\text{ kg/d}) \times (0.0009)$$

$$3213\text{ kg/d} = (0.22)(\text{S}) + (0.0009)(F)$$

where S = sludge cake flowrate (wet), kg/d
F = filtrate flowrate (wet), kg/d

b. Develop mass flowrate equation.

Sludge flowrate + washwater flowrate = filtrate flowrate + cake flowrate
Daily sludge flowrate = (535,500 kg/wk)/(5 d/wk) = 107,100 kg/d
Washwater flowrate = (90 L/min·m)(1.5 m)(60 min/h)(8 h/d)(1 kg/L)(1.0)
= 64,800 kg/d
107,100 kg/d + 64,800 kg/d = 171,900 kg/d = $F + S$

c. Solve the mass balance and flowrate equations simultaneously.
First solve S in terms of F per the flowrate equation in 4.b.

$S = 171{,}900\ \text{kg/d} - F$

Next solve for F from the solids balance equation in 4.a.

$$\begin{aligned}3213\ \text{kg/d} &= 0.22\,(171{,}900\ \text{kg/d} - \text{F}) + 0.0009(\text{F})\\ &= 37{,}818\ \text{kg/d} - 0.2191(\text{F})\\ F &= 157{,}942\ \text{kg/d}\\ &= (157{,}942\ \text{kg/d})/(1\ \text{kg/L})/(1.01) = 159{,}521\ \text{L/d}\end{aligned}$$

5. Determine solids capture.

$$\text{Solid capture} = \frac{\text{solids in feed} - \text{solids in filtrate}}{\text{solids in feed}} \times 100$$

$$= \frac{[(3213\ \text{kg/d}) - (157{,}942\ \text{kg/d})(0.0009)]}{(3213\ \text{kg/d})} \times 100$$

$= 95.6\%$

6. Determine operating requirements for sustained peak biosolids load.

a. Determine peak 3-d load.
From Fig. 3–14 (b), the ratio of peak to average mass loading for 3 consecutive days is 2. The peak load is (75,000 L/d) (2) = 150,000 L/d.

b. Determine daily operating time requirements, neglecting sludge in storage.

$$\begin{aligned}\text{Dry solids per day} &= (150{,}000\ \text{L/d})(1\ \text{kg/L})(1.02)(0.03)\\ &= 4590\ \text{kg/d}\end{aligned}$$

$$\text{Operating time} = \frac{(4590\ \text{kg/d})}{(275\ \text{kg/m·h})(1.5\,\text{m})} = 11.1\ \text{h/d}$$

The operating time can be accomplished by running the standby belt-filter press in addition to the duty press, or operating the duty press for an extended shift.

Comment The value of sludge storage is important in dewatering applications because of the ability to schedule operations to suit labor availability most efficiently. Scheduling sludge dewatering operations during the day shift is also desirable if sludge has to be hauled off-site.

Rotary Press

The rotary press as a dewatering device for municipal sludges and biosolids has been adapted since 1994 with the first installation at the City of Montreal, Canada. Similar to centrifuges and BFPs, the rotary press feed solids are typically conditioned with cationic polymer that is injected into the feed upstream of the inlet to the dewatering unit. A typical process flow diagram for a rotary press system is shown on Fig. 14–7(a). Because rotary press systems are totally enclosed, they provide enhanced safety, odor containment, and relatively low noise levels compared with some other types of sludge or biosolids dewatering systems discussed in this textbook.

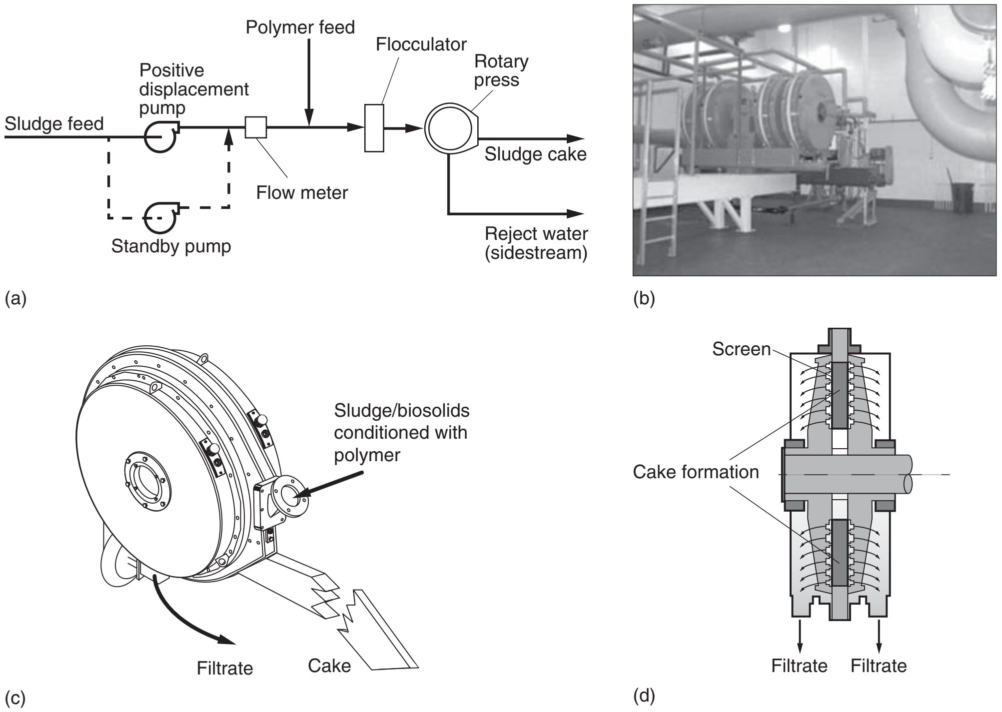


Figure 14–7

Rotary press for sludge dewatering: (a) schematic process flow diagram, (b) view of typical installation (courtesy of Scarborough Sanitart District, Scarborough ME), (c) view of a rotary press unit, and (d) cross-section through press (courtesy of Fournier Industries, Inc.).

Description. The rotary press is a slow-speed, enclosed, modular unit. The sludge or biosolids are fed into the unit at a relatively low pressure into the space between the two parallel filter screens. The sludge or biosolids are typically fed into a cylindrical coated carbon steel containment vessel that has a rectangular cross section. The solids within each module are retained between two parallel rotating stainless steel screens. The flocculated slurry material advances forward within the rotating channel that is formed between the parallel screens, and the filtrate passes out through sides of the screens. The sludge travelling around the channel (between the rotating screens) continues to be dewatered. The frictional force of the slow-moving screen assembly and the adjustable outlet restriction creates back pressure that forces additional filtrate out through the screen. The solids retained between the screens are dewatered, effectively forming an extrusion of relatively dry dewatered cake. Washwater is required intermittently to clean the screens and flush the solids from the unit during shutdown. A cutaway view of the typical rotary press is shown on Fig. 14–7(d).

Design Considerations. The design of the rotary press dewatering system is affected by a number of factors. These include quantity of sludge or biosolids, feed concentration, characteristics of sludge (digested or undigested, primary sludges or secondary sludges or both), safety, allowable noise levels, allowable odor levels, desired operating schedule, level of automation, space availability, reliability and redundancy requirements, washwater quality and availability, filtrate quality and treatment requirements, manufacturers' local service capabilities, and budgetary constraints.

The quantity of sludge or biosolids should be established at the beginning of the design process. The quantities should be developed for both current operation and projected future conditions. The characteristics of the sludges or biosolids that will be dewatered are an important consideration. The digested biosolids and raw secondary sludges are typically the most difficult to dewater effectively. Conversely, the presence of primary sludges tends to aid the dewatering process. The operating schedule at the facility, level of automation, reliability/redundancy requirements, and physical space availability need consideration when selecting the system and number of units. Washwater is used intermittently to flush the rotating screen assemblies during the shutdown sequence. The amount of washwater needed is relatively small, typically 190 L/min (50 gal/min) per channel. The time required for the flushing sequence is approximately 5 min per channel per d. All of the channels are normally flushed simultaneously, although sequential flushing can be used if the washwater supply has a capacity limitation. The units typically are offered with a single-channel, a dual-channel, a four-channel and a six-channel configuration. Special three- and five-channel units can also be provided, but the inlet configuration can create a flow imbalance due to the unequal number of channels on each side of the inlet connection. The screen diameter ranges from 460 to 1220 mm (18 to 48 in.) depending on the selected manufacturer. The effective dewatering area for various rotary press configurations is reported in Table 14–5.

The maximum hydraulic loading limit for a rotary press is approximately 8.5 m/h (3.5 gal/min·ft^2) at a total solids feed concentration of 3 percent. The maximum solids' loading is approximately 244 – 254 kg/h·m^2 (50 – 52 lb/h·ft^2). To optimize cake dryness and minimize polymer usage, rotary presses are normally operated below the maximum hydraulic and solids loading criteria. The typical average operating hydraulic loading rate for a rotary press system is approximately 2.4 m/h (1.0 gal/min·ft^2) at a total solids feed concentration of 4 percent and approximately 3.7 m/h (1.5 gal/min·ft^2) at a total solids feed concentration of 2 percent.

As discussed earlier, the types of sludges and biosolids to be dewatered have a significant impact on the actual allowable hydraulic loading, solids loading, polymer use, and dewatered cake moisture content. The characteristics and composition of sludge and biosolids vary from facility to facility so on-site pilot testing is recommended if strict performance and capacity criteria will be specified for the rotary press equipment.

System Operation and Performance. A typical rotary press system consists of a sludge feed pump, flow metering, polymer feed equipment, conditioning system (either inline or tank designs can be used), the rotary press, a dewatered cake conveyor (if direct discharge to a container or truck is not feasible), compressed air for valve control functions,

Table 14–5
Rotary press dewatering area[a]

	Effective dewatering area, m^2			
No. of Channels	460 mm diameter screens	610 mm diameter screens	915 mm diameter screens	1220 mm diameter screens
1	0.23	0.40	0.96–1.00	1.75
2	NA	0.79	1.91–1.20	3.49
4	NA	NA	3.84–4.00	7.00
6	NA	NA	6.00	NA

[a] Based on feedback from rotary press vendors.

Table 14–6
Typical dewatering performance of rotary press[a]

Type of feed	Process Parameter: Polymer use, lb/ton dry TS	Polymer use, g/kg dry TS	Cake solids, % TS	Solids capture, %
Untreated sludge				
Primary	4–12	2–6	28–45	95+
Primary plus WAS	15–20	7.5–10	20–32	92–98
WAS	20–35	12.5–17.5	13–18	90–95
Anaerobically digested biosolids				
Primary	15–20	7.5–10	22–32	90–95
Primary plus WAS	20–30	10–15	18–25	90–95
WAS	20–35	10– 17.5	12–17	85–90
Aerobically digested WAS	17–25	8.5–17.5	28–45	90–95

[a] Based on feedback from rotary press vendors.

electrical power for the drive components, and control/signal wiring for remote monitoring or operational control. A typical rotary press installation is shown on Fig. 14–7(b).

Rotary press process variables that can be adjusted to optimize performance or minimize operational costs include feed pump speed and flow, polymer concentration, polymer feed pump speed and flow, polymer mixing intensity, screen rotational speed, cake discharge outlet pressure, and washwater flushing frequency and duration. The performance of the unit is affected by each of these variables. In optimizing the performance of a rotary press, it is important to change one variable at a time so that the effect of each variable can be evaluated fully. The typical performance for a rotary press system varies from facility to facility and is affected by the type of solids feed, the process variables listed above, and the physical condition of the equipment and controls. Typical operating performance data for a rotary press system is reported in Table 14–6.

Screw Press

The screw press was adapted from industrial applications in 1990 for dewatering wastewater sludges or digested biosolids. The screw press feed solids are typically conditioned with cationic polymer that is injected into the feed upstream of the inlet to the dewatering unit. A typical process flow diagram for a screw press system is shown on Fig. 14–8(a).

Description. The screw press is a slow-speed, enclosed cylindrical unit. The sludges or biosolids are fed into the unit at a relatively low pressure into a stationary wedge wire screening basket with a rotating screw assembly that transfers the sludges or biosolids through the wedge wire screening basket. The flocculated material advances forward up along the rotating screw assembly. The filtrate passes out through the bottom and sides of the wedge wire screen. The sludges travelling along the screw continue to be dewatered. The frictional force of the slow moving screw assembly and the adjustable outlet restriction

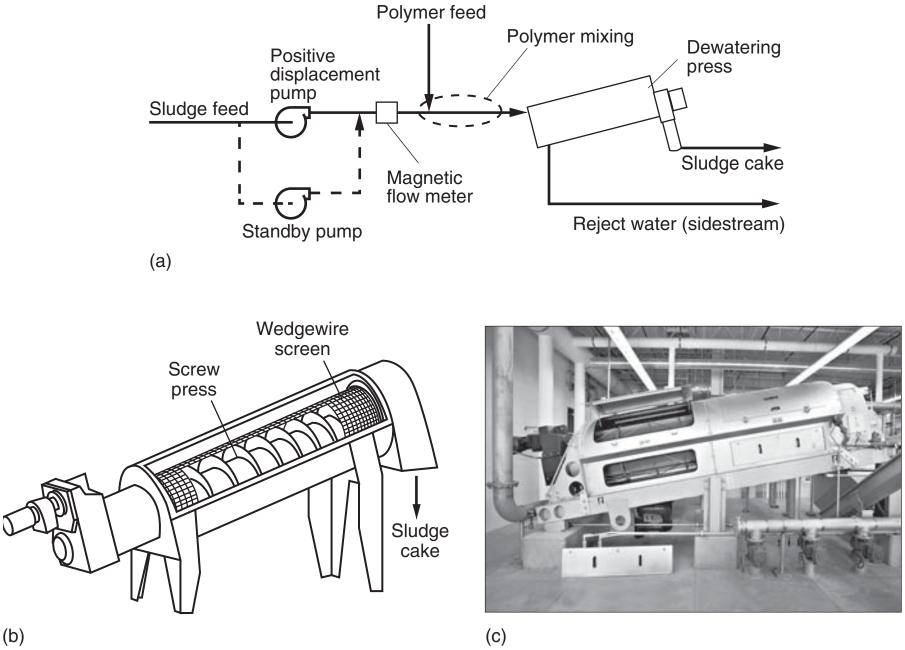


Figure 14–8

Screw press for sludge dewatering: (a) schematic process flow diagram (courtesy of PW Tech, Inc.), (b) cutaway view (adapted from Huber Technology Inc.) and (c) view of inclined screw press.

creates back pressure that forces additional filtrate out through the screen near the outlet end of the unit. The dewatered sludges form an extrusion of relatively dry dewatered cake. Washwater is required intermittently to clean the wedge wire screening assembly and flush the solids from the unit throughout the normal operating cycle. Units are available in either horizontal or inclined configuration depending on the manufacturers' preference. A cutaway view of a typical screw press is illustrated on Fig. 14–8(b). A view of a typical screen press is presented on Fig. 14–8(c).

Design Considerations. The design of the screw press dewatering system will be affected by a number of factors. These include quantity and feed concentration of sludges, volatile solids (VS) content, sludge characteristics (digested or undigested, primary sludges or secondary sludges or both), safety considerations, allowable noise and odor levels, facility operating schedule, extent of automation, building space availability, redundancy and reliability requirements, quality and availability of washwater, required filtrate quality and treatment, local service capabilities of the manufacturer, and financial constraints.

Similar to other types of dewatering systems the quantities of sludges or biosolids need to be established at the beginning of the design process. The quantities should be developed for both current operation, and projected future conditions. The characteristics of the material to be dewatered are an important consideration. Digested biosolids and raw secondary sludges are typically more difficult to effectively dewater than primary sludges. The screw press systems are low speed, totally enclosed units that provide enhanced safety, odor containment, and relatively low noise levels compared with some other types of dewatering systems discussed in this textbook. The schedule of operation, automation requirements, redundancy and reliability considerations, number of units, and building space availability need due consideration. The screw press intermittently utilizes washwater to flush the stationary wedge wire screen assembly regularly throughout the normal operating cycle. The amount of washwater needed is relatively small. Typically a flowrate of 7 to 45 L/min (2 to 12 gal/min) at a pressure between 2.8 to 5.5 bar-gauges (40 to 80 lb/in.^{2}-gauge) is required. The screen is flushed for approximately 15 s every 10 min during its normal operating cycle.

The units typically are offered with a single screw, dual screw, inclined, or horizontal configuration depending on the manufacturer that is selected. Units are available in various sizes and a single unit is capable of processing up to 500 kg/h (1100 lb/h) although the units may be hydraulically limited for thin waste activated sludges (< 1 percent TS). Dual screw units have a flow capacity that is approximately twice the capacity of the single screw configuration units for a given screw diameter and screen length.

As noted, the type of sludges or biosolids to be dewatered has a significant impact on the actual allowable hydraulic loading, solids loading, polymer use, and dewatered cake moisture content. The characteristics and composition of sludges or biosolids vary from facility to facility, so on-site pilot testing of the screw press system is recommended if strict performance and capacity criteria are to be specified for the screw press equipment.

System Operation and Performance. A typical screw press system consists of a sludge feed pump, flow metering, polymer feed equipment, inline solids conditioning system, a dewatered cake conveyor (if direct discharge to a container or truck is not feasible), electrical power for the drive components, and control/signal wiring for remote monitoring or operational control.

Screw press process variables that can be adjusted to optimize performance or minimize operational costs include feed pump speed and flow, polymer concentration, polymer feed pump speed and flow, polymer mixing intensity, screw rotational speed, cake discharge outlet pressure, and washwater flushing frequency and duration. The typical performance of for a screw press system varies from facility to facility and is impacted by the type of feed, the process variables listed above, and the physical condition of the equipment and controls. Typical operating performance data for a screw press system are presented in Table 14–7.

Table 14–7
Typical dewatering performance of screw press[a]

	Process parameter			
	Polymer use			
Type of feed	lb/ton dry TS	g/kg dry TS	Cake solids, % TS	Solids capture, %
Untreated sludge				
Primary	8–20	4–10	30–40	90+
Primary plus WAS	10–20	5–10	25–35	90+
WAS	17–22	8.5–11	15–22	88–95
Anaerobically digested biosolids				
Primary	20–35	10–17.5	22–28	90+
Primary plus WAS	20–35	10–17.5	17–25	90+
WAS	17–35	8.5–17.5	15–25	88–95
Aerobically digested WAS	17–25	8.5–17.5	15–20	88–95

[a] Based on feedback from screw press vendors.

Filter Presses

In a filter press, dewatering is achieved by forcing the water from the sludge or biosolids under high pressure. Advantages and disadvantages of filter presses are included in Table 14–2. When requiring cake solids contents greater than 35 percent on a routine basis, use of filter press is often dictated as other mechanical dewatering devices cannot achieve this high solids content consistently.

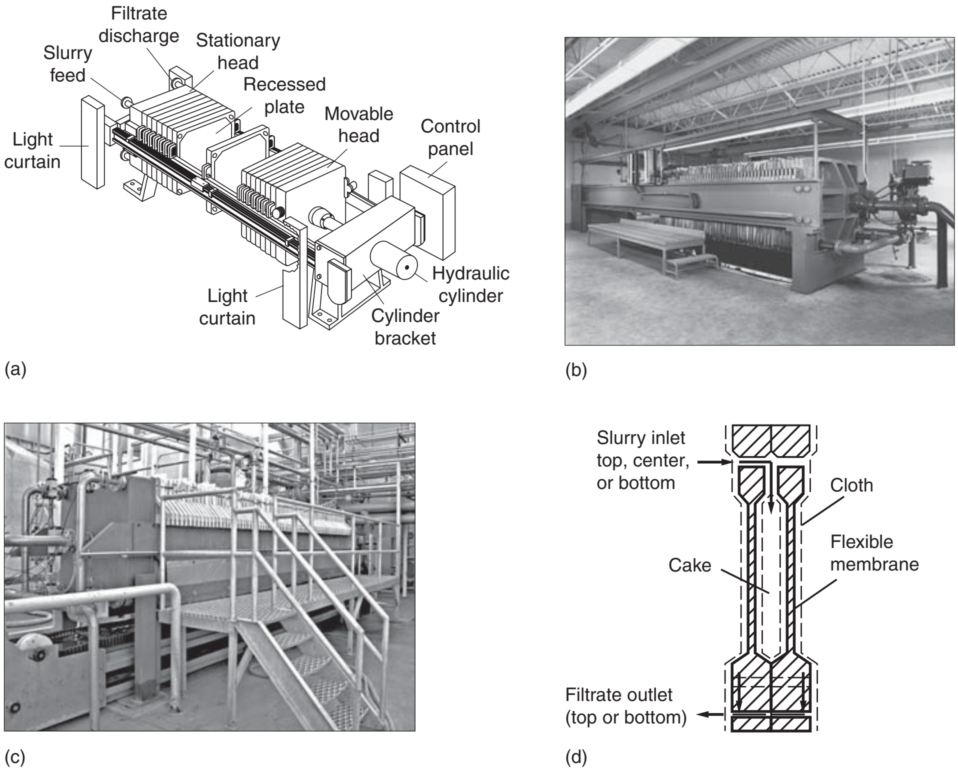
Various types of filter presses have been used to dewater sludges or biosolids. The two types used most commonly are the fixed-volume and variable-volume recessed-plate filter presses.

Fixed-Volume, Recessed-Plate Filter Press. The fixed-volume, recessed-plate filter press consists of a series of rectangular plates, recessed on both sides, that are supported face to face in a vertical position on a frame with a fixed and movable head [see Fig. 14–9(a)]. A filter cloth is hung or fitted over each plate. The plates are held together with sufficient force to seal them to withstand the pressure applied during the filtration process. Hydraulic rams or powered screws are used to hold the plates together.

In operation, chemically conditioned sludge or biosolids is pumped into the space between the plates, and pressure of 700 to 2100 kPa (100 to 300 $\text{lb}_f/\text{in.}^2$) is applied and maintained for 1 to 3 h, forcing the liquid through the filter cloth and plate outlet ports. The plates are then separated and the cake is removed. The filtrate normally is returned to the influent of the treatment plant. The cake thickness varies from about 25 to 38 mm (1 to 1.5 in.), and the moisture content varies from 45 to 70 percent. The filtration cycle time varies from 2 to 5 h and includes the time required to (1) fill the press, (2) maintain the press under pressure, (3) open the press, (4) wash and discharge the cake, and

Figure 14–9

Typical fixed-volume, recessed plate filter press used for dewatering sludge: (a) schematic of filter press, (b) and (c) views of a typical installations, and (d) cross section through a variable-volume recessed plate filter press.



(5) close the press. Depending on the degree of automation incorporated into the machine, operator attention must be devoted to the filter press during feed, discharge, and wash intervals.

Variable-Volume, Recessed-Plate Filter Press. Another type of filter press used for wastewater sludge dewatering is the variable-volume recessed-plate filter press, commonly called the "diaphragm press." This type of filter press is similar to the fixed-volume press except that a rubber diaphragm is placed behind the filter media, as shown on Fig. 14–9(d). The rubber diaphragm expands to achieve the final squeeze pressure, thus reducing the cake volume during the compression step. Generally about 10 to 20 min are required to fill the press and 15 to 30 min of constant pressure are required to dewater the cake to the desired solids content. Variable-volume presses are generally designed for 690 to 860 kN/m^2 (100 to 125 lb$_f$/in.^{2}) for the initial stage of dewatering followed by 1380 to 2070 kN/m^2 (200 to 300 lb$_f$/in.^{2}) for final compression. Variable-volume presses can handle a wide variety of sludges and biosolids with good performance results but require considerable maintenance (WEF, 2010).

Design Considerations. Several operating and maintenance problems have been identified for recessed-plate filter presses ranging from difficulties in the chemical feed and sludge-conditioning system to excessive downtime for equipment maintenance. Features that should be considered in the design of a filter press installation include (1) adequate ventilation in the dewatering room (6 to 12 air changes per hour are recommended depending on the ambient temperature), (2) high-pressure washing systems, (3) an acid wash circulation system to remove calcium scale when lime is used, (4) a grinder ahead of the conditioning tank, (5) cake breakers or shredders following the filter press (particularly if the dewatered sludge is incinerated), and (6) equipment to facilitate removal and maintenance of the plates. Other design criteria are can be found in WEF, (2010).

Combined Diaphragm Press with Vacuum Drying. Two components to the traditional diaphragm filter press are implemented with this combined process shown schematically on Fig. 14–10. First sludge is introduced into the press, and internal filter

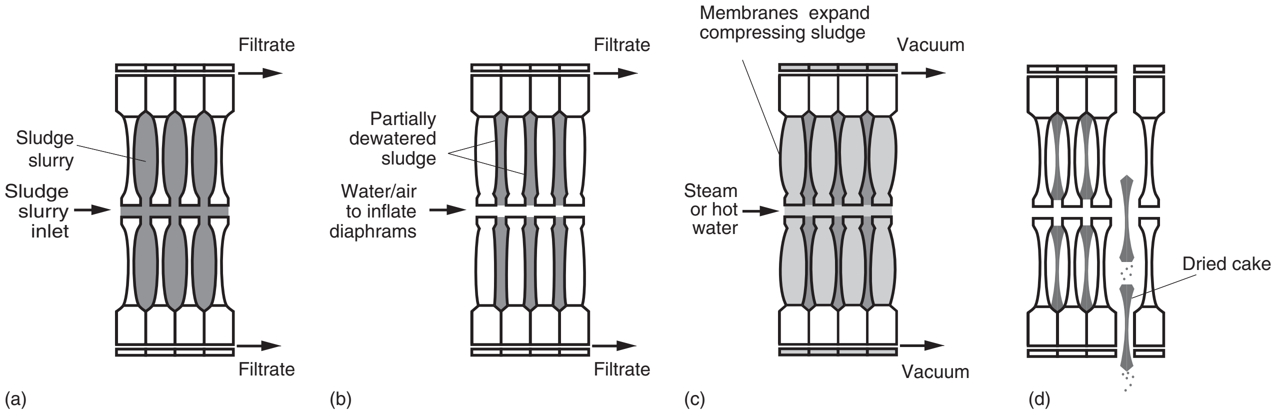


Figure 14–10

Schematic of the operation of variable-volume vacuum assisted recessed-plate filter press: (a) sludge slurry is introduced into the press (b) flexible membrane is expanded with air or water, (c) steam is added along with vacuum to reduce the boiling point of the water and, thus, further dewater the sludge, and (d) at the end of the cycle the filter press is open, and dried sludge is collected in the bin below the press.

plate diaphragms are inflated with air or water to squeeze out free water Next, hot water or steam is introduced to raise the temperature of the sludge while a vacuum is applied to lower the boiling point of water. Vaporized water in sludge is drawn out through the vacuum. At the end of the cycle, the press is opened to drop the dried sludge onto a receiving bin. There are several installations of this technology in North America and Europe (ca. 2013).

Electro-Dewatering

Electric field-assisted dewatering, also called electro-dewatering, is used to dewater municipal sludges and biosolids, industrial residuals, as well as less conventional biomass. The process can be used to dewater raw sludges or digested biosolids. The basic principle of operation, the commercial application of the principle, design considerations, and some performance data are presented and discussed below.

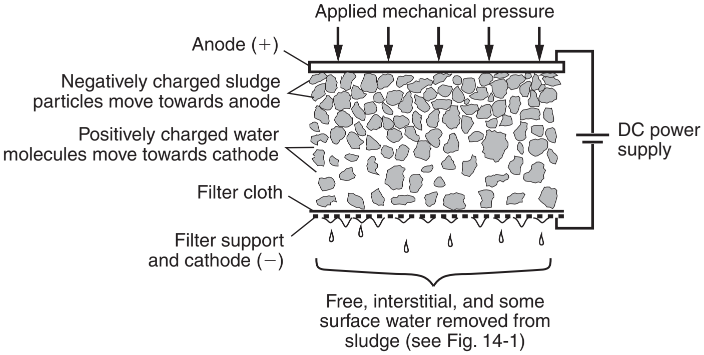
Description. In practice, because the bonding forces and the particle size of the sludge or biosolids, bound water cannot be removed readily by mechanical means. Electro-dewatering involves the application of a direct voltage to sludge or biosolids placed between two electrodes as illustrated on Fig. 14–11. Negatively charged sludge or biosolids move towards or gather at the positive electrode (anode). Positively charged water molecules move towards or gather at the negative electrode (cathode). At the negative electrode water moves through the filter cloth which covers the electrode. The filter cloth on the negative electrode does not clog, a common problem in mechanical filtration, as the sludge or biosolids particles are repelled by the negative electrode (Yoshida, 1993). Pressure is applied to the material to be dewatered to further accelerate the process and to allow the applied DC current to move more uniformly from the anode to the cathode [see Fig. 14–11(a)]. The advanced dewatering capability of electro-dewatering process as compared to other and other dewatering technologies was illustrated previously on Fig. 14–2(b).

Commercial Implementation of the Electro-Dewatering Process. In the most common commercial configurations encountered, a typical electro-dewatering system consists of a feeding module, the electro-dewatering unit, a high pressure wash system, a rectifier, a dewatered cake conveyor (if direct discharge to a container or truck is not feasible), compressed air for valve control functions, electrical power for the drive components and control/signal wiring for remote monitoring or operational control. A typical schematic of an electro-dewatering system is shown on Fig. 14–12(a).

Referring to Fig. 14–12(a), sludges or biosolids are fed in a hopper built as part of a feeding module. At the beginning of each treatment cycle, a conveyor system is actuated and

Figure 14–11

Definition sketch for the operation of electro-dewatering process. (Adapted from Mahmud et al., 2010.)



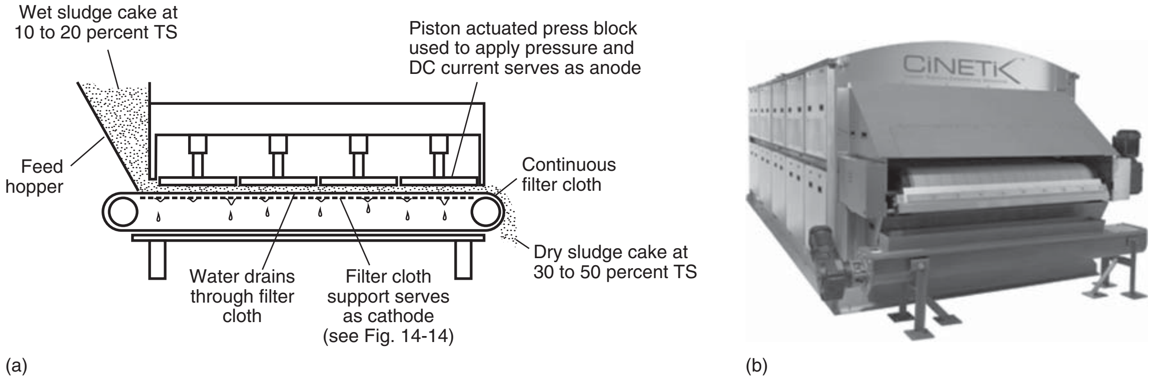


Figure 14–12

Electro-dewatering apparatus: (a) schematic diagram of operation and (b) view of discharge side of a linear electro-dewatering machine (courtesy of Ovivo, Inc.).

the feeding module extrudes a thin and uniform layer of the material to be dewatered on the filtering belt, to a predetermined thickness. Once the newly formed cake is moved into the treatment area as shown on Fig. 14–12 (a), power-blocks move down on the cake and apply a predetermined pressure. The applied pressure also allows DC current (generated by a rectifier) to flow through the cake and the filtering belt, between the anodes and the cathodes. The DC current is applied at predetermined levels for a controlled period of time until the required cake dryness is achieved, after which the power-blocks are lifted back to their upper position and the treated cake is discharged from the system. The filtering belt is cleaned using a high pressure wash system (filtrated process water or potable water) during its displacement. The high pressure wash system is also used daily to clean the cathodes and the equipment, during programmed automatic wash sequences and before a shutdown sequence. Machine controls and instrumentation ensure management of operating parameters such as pressure, voltage, current, treatment time, belt speed, and automatic wash cycles.

Design Considerations. Important design considerations for the electro-dewatering process are summarized in Table 14–8 and discussed briefly below. Specific design parameters are presented in Table 14–9. The quantity of sludge or biosolids should be established at the beginning of the design process. The quantities should be developed for both current operation and projected, future conditions. The characteristics of the sludge or biosolids that will be dewatered are the most important consideration. Electro-dewatering is particularly suited for processing digested biosolids and/or secondary sludge, typically the most difficult to dewater effectively with conventional mechanical dewatering processes. Electro-dewatering systems, such as shown on Fig. 14–12(b), are totally enclosed for enhanced safety and odor containment. Also the footprint is small and energy usage is relatively low as compared with thermal systems, discussed later in this chapter.

Integrating electro-dewatering into biosolids processing is worthy of consideration, particularly for small-to-mid-sized facilities typically less than 57,000 m^3/d (15 Mgal/d) range (Eschborn, 2011). For any size facility, factors favoring integrating electro-dewatering include high biosolids ultimate disposal costs and a desire to produce a Class A product. It is recommended to contact manufactures at the early stages of design process and obtain initial results from laboratory bench scale equipment. Piloting to evaluate the

Table 14–8

Design considerations for electro-dewatering[a]

Item	Comment
Quantity of sludge and biosolids	The quantity of solids should be established at the beginning of the design process. The quantities should be developed for both current operation and projected, future conditions.
Characteristics of the feed sludge or biosolids	Important feed characteristics include conductivity, pH, particles sizes, concentration, ionic composition and polymer used. Although unit capacity can sometimes be reduced with a wetter feed solids, better treatment response is commonly observed with feeds at 12 to 20 percent TS.
Type of sludge or biosolids	Electro-dewatering is particularly suited for processing digested biosolids and/or secondary sludge which are difficult to dewater effectively by conventional mechanical means.
Sludge or biosolids pretreatment	Conventional mechanical dewatering equipment such as belt filter presses, centrifuges, screw presses or rotary presses are used upstream of the electro-dewatering.
Final dryness desired	Depends on the ultimate use of the dewatered material. Cake solids are typically in the range between 25 and 50 percent TS.
Expected volume reduction	Typically, volume reduction will vary from 50 to 75 percent.
Filtrate characteristics and treatment requirements	The filtrate contains relatively high concentrations of organics (BOD_5 and COD), suspended solids (TSS), ammonia and organic nitrogen (TKN). Although the flow of this filtrate is small, these constituents can have an impact on downstream treatment processes, so they should be characterized and considered during the design of electro-dewatering.
Washwater quality and availability	Filtered process water or potable water is used to avoid clogging the filter. The amount of water needed is relatively small.
Electricity cost	Three to five fold reduction in energy usage compared to thermal drying.
Plant related issues in selection of system and number of units	Safety considerations, allowable odor levels, facility operating schedule, level of automation, space availability, reliability and redundancy requirements, manufacturers local service capabilities, and budgetary constraints.

[a] Courtesy of OVIVO.

Table 14–9

Design parameters for various linear electro-dewatering models[a]

		Effective dewatering area, m^2		
Item	Unit	4	8	16
Effective dewatering area	m^2	4	8	16
	ft^2	43	86	172
Footprint	m^2	11.6	19.5	27.9
	ft^2	125	210	300
Capacity at inlet	kg/h	270–600	545–1180	1090–2360
	lb/h	600–1320	1200–2600	2400–5200
Washwater	L/min	15.9	18.9	22.7
	gal/min	4.2	5	6

[a] Courtesy of OVIVO.

Table 14–10

Typical linear electro-dewatering performance for total solids and energy usage

	Cake solids, % TS		Energy usage[a]	
Type of feed	**Inlet**	**Outlet**	**kWh/ton**	**kWh/kg**
Untreated sludge				
Primary	22–24	29–49	110–260	0.12–0.29
WAS	13–17	28–43	150–270	0.17–0.30
	25	33–38	210–310	0.23–0.34
	16–20	32–43	230–310	0.25–0.34
Anaerobically digested biosolids	12–18	30–46	190–280	0.21–0.31
	20–23	32–48	165–260	0.18–0.29
Aerobically digested WAS	16–20	32–43	230–310	0.25–0.34

[a] Wet basis.

compatibility of the technology with the sludge or biosolids and to determine time required to achieve the desired cake dewatering along with electricity consumption is highly recommended, once satisfactory laboratory results are obtained.

Process Performance Data. Process variables that can be adjusted to optimize performance or minimize operational costs include cycle time (time of the batch treatment), applied voltage, applied current intensity, applied pressure, thickness of the formed cake (feed), sludge conductivity and ionic composition, polymer type and concentration (applied upstream, at the mechanical dewatering step). Typical operating performance data for linear electro-dewatering, which will vary facility to facility, is presented in Table 14–10. Thermal drying requires around 617–1200 kWh/m^3 of water removed (Gazbar et al., 1994; Mujumdar, 2007) whereas application of linear electro-dewatering often results in a 3 to 5 fold reduction in energy usage. In areas where electricity costs are high, economic evaluation to verify the cost benefit from obtaining the desired cake solids is required.

Sludge Drying Beds

Drying beds used to be the most widely used method of sludge dewatering in the United States. Sludge drying beds are typically used to dewater digested biosolids and settled sludge from plants using the extended aeration activated-sludge treatment process without prethickening. After drying, the dried material is removed and either disposed of in a landfill or used as a soil conditioner. The advantages and disadvantages of drying beds are summarized in Table 14–11. Conventional sand drying beds are the most commonly used sludge drying beds. Other types of drying beds include paved drying beds, wedge-wire drying beds, and vacuum-assisted drying beds. Because paved and wedge-wire drying beds are not used commonly, they are not covered in this textbook. Vacuum-assisted sludge

Table 14–11

Advantages and disadvantages of drying beds

Advantages	**Disadvantages**
1. Low capital and operating cost alternative	1. Requires a large footprint
2. Simple operation with minimal operator attention required	2. Performance highly susceptible to weather conditions
3. High-solids content in the dried product	3. Removing sludge can be labor-intensive high potential for odors
	4. Can attract insects

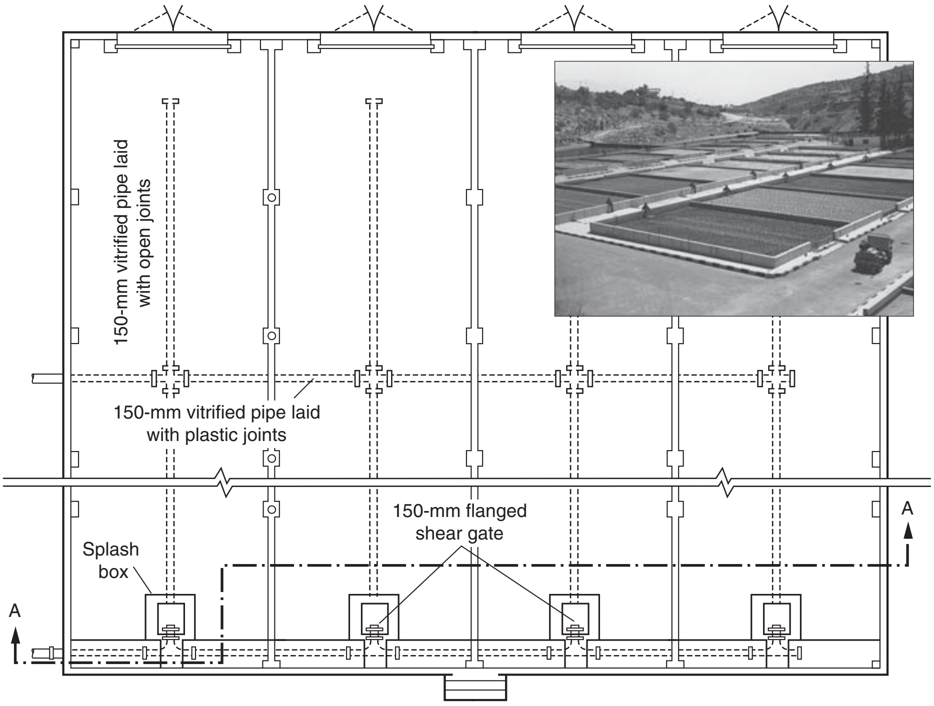
drying beds were used in the past but are no longer commercially available and are not included in this text (WEF, 2012). Because conventional sand drying beds are used most extensively, more detailed discussion is provided for these units.

Conventional Sand Drying Beds. Conventional sand drying beds are generally used for small- and medium-sized communities. For cities with populations over 20,000, consideration should be given to alternative means of sludge dewatering. In larger municipalities, the initial cost, the cost of removing the sludge and replacing sand, and the large area requirements generally preclude the use of sand drying beds.

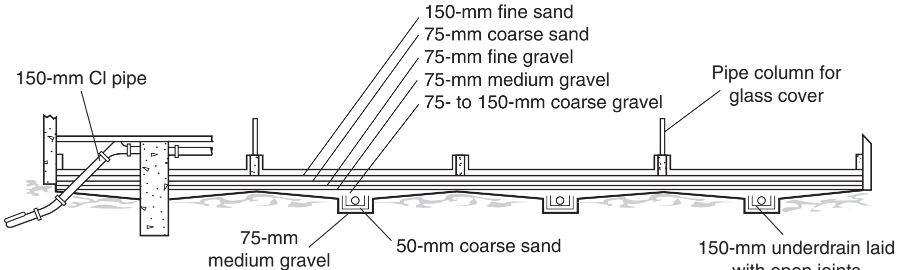
In a typical sand drying bed, sludge is placed on the bed in a 200 to 300 mm (8 to 12 in.) layer and allowed to dry. Sludge dewaters by gravity drainage through the sludge mass and supporting sand and by evaporation from the surface exposed to the air (see Fig. 14–13). Most of the water leaves the sludge by drainage; thus the provision of an adequate underdrainage system is essential. Drying beds are equipped with lateral drainage lines (perforated plastic pipe or vitrified clay pipe laid with open joints), sloped at a minimum of 1 percent and spaced 2.5 to 6 m (8 to 20 ft) apart. The drainage lines

Figure 14–13

Typical conventional sand drying bed: (a) plan and (b) section. Insert – view of sludge drying beds with sludge in various stages of dryness.



(a) Plan



(b) Section A–A

should be adequately supported and covered with coarse gravel or crushed stone. The sand layer should range from 200 to 460 mm (9 to 18 in.) deep with an allowance for some loss from cleaning operations. Deeper sand layers generally retard the draining process. Sand should have a uniformity coefficient of not over 4.0 and an effective size of 0.3 to 0.75 mm.

The drying area is typically partitioned into individual beds 7.5 m wide although the actual width for a particular site should be set to accommodate the sludge removal method. The bed length can vary and beds can be up to 30 to 60 m (100–200 ft) long. The partitions are typically constructed of earthen embankments, wooden planks, concrete planks or reinforced concrete blocks. The outer wall is normally constructed with a 500–900 mm (20–36 in.) freeboard above the sand area and the wall usually extends to the underdrain gravel. Concrete foundation walls are required if the beds are to be covered.

Sludges or biosolids are typically fed to drying beds via an open channel or closed piping. Distribution boxes or valves are required to divert the sludge or biosolids flow into the bed selected. Splash plates are placed in front of the sludge or biosolids outlets to spread the material over the bed and to prevent erosion of the sand. If the feed pipe is pressurized, a 90 elbow should be used to ensure the sludges or biosolids hit the splash plate at all feed flowrates.

Sludges or biosolids can be removed from the drying bed after it has drained and dried sufficiently. Dried material has a coarse, cracked surface and is black or dark brown. Sludge or biosolid removal is accomplished by manual shoveling into wheelbarrows or trucks, or by a scraper, front-end loader, or special mechanical sludge removal equipment. Provisions should be made for driving a truck onto or alongside of the bed to facilitate loading.

Open beds are used where adequate area is available and sufficiently isolated to avoid complaints caused by occasional odors. Open sludge beds should be located at least 100 m (about 300 ft) from dwellings to avoid odor nuisance. Covered beds with greenhouse types of enclosures are used where it is necessary to dewater sludge continuously throughout the year regardless of the weather, and where sufficient isolation does not exist for the installation of open beds.

Drying bed solids loadings are computed on dry solids loading per square foot per year (kilograms of dry solids per square meter per year). Designing drying beds based on a solids loading criteria is typically the preferred approach and loading requirements typically vary from 50 to 125 $kg/m^2 \cdot y$ (10–25 $lb/ft^2 \cdot y$) for open drying beds and 60–200 $kg/m^2 \cdot y$ (12–40 $lb/ft^2 \cdot y$) for closed drying beds (WEF 2010). Typical data for various types of biosolids are shown in Table 14–12. With covered drying beds, more biosolids can be

Table 14–12
Typical area requirements for open sludge drying beds

	Area[a]		Dry sludge loading rate	
Type of biosolids	**ft^2/person**	**m^2/person**	**$lb/ft^2 \cdot y$**	**$kg/m^2 \cdot y$**
Primary digested	1.0–1.5	0.1	25–30	120–150
Primary and trickling filter humus digested	1.25–1.75	0.12–0.16	18–25	90–120
Primary and waste activated digested	1.75–2.5	0.16–0.23	12–20	60–100
Primary and chemically precipitated digested	2.0–2.5	0.19–0.23	20–33	100–160

[a] Corresponding area requirements for covered beds vary from about 70 to 75 percent of those for the open beds.

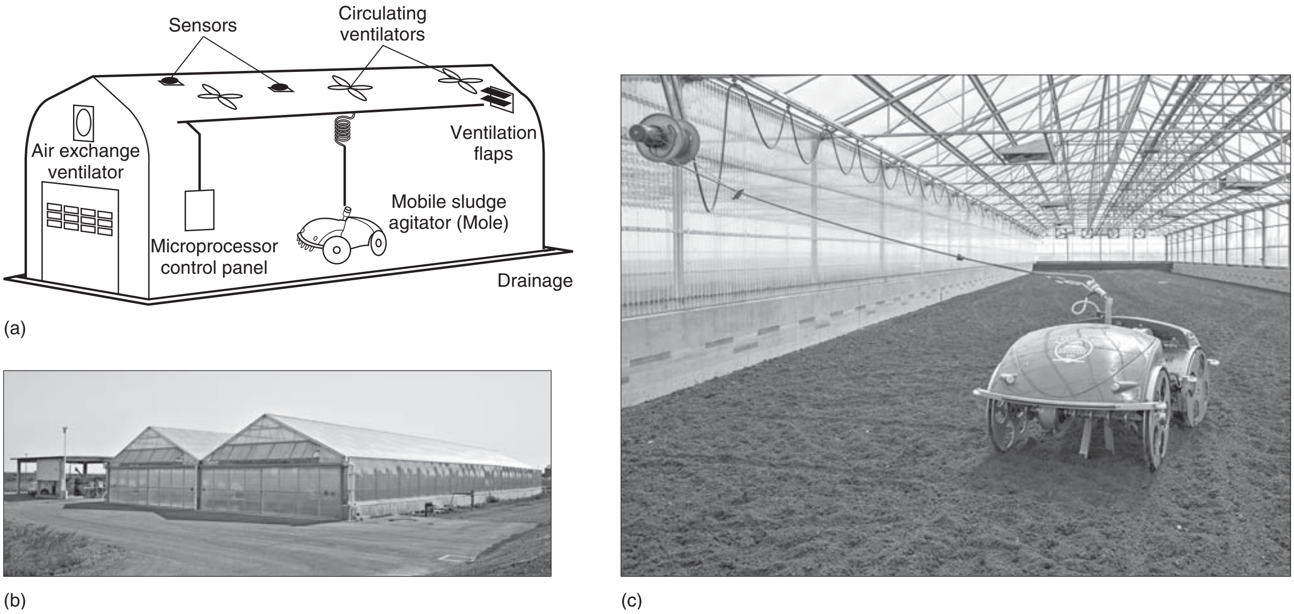


Figure 14–14

Solar sludge drying bed system employing mobile sludge agitator: (a) schematic (adapted from Parkson Corp.), (b) view of exterior of typical installation, and (c) view interior of a typical installation and mobile sludge mixer.

applied per year because of the protection from rain and snow. Polymer conditioning is also sometimes used to improve the performance of sludge drying beds.

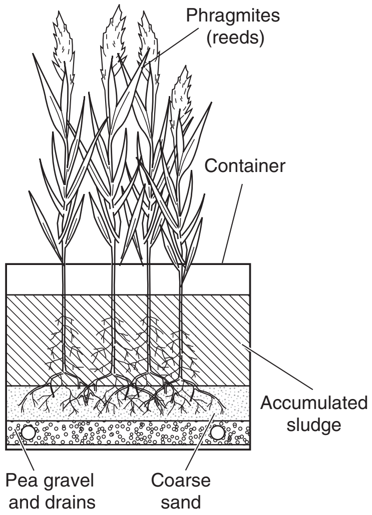
Solar Drying Beds. A method used to enhance the dewatering and drying of liquid, thickened, or dewatered biosolids is solar drying in covered drying beds (see Fig. 14–14). Solar drying systems are typically not used for undigested primary sludge due to concerns of odor and requirements to dry to greater than 90 percent to get Class A biosolids product instead of 75 percent drying for stabilized biosolids. The solar drying system, which is a sophisticated "greenhouse," consists of a rectangular base structure, translucent chamber, sensors to measure atmospheric drying conditions, air louvers, circulation fans, ventilation fans, a mobile electromechanical device that agitates and moves the drying biosolids, and a microprocessor that controls the drying environment [see Figs. 14–14(a) and (b)]. The system's main source of drying energy is solar radiation.

In most solar drying systems, mechanically dewatered biosolids are distributed in a greenhouse either manually or automatically. It is also possible to add liquid sludge directly to the greenhouse; however, the additional greenhouse area required typically outweighs the benefit from eliminating the mechanical dewatering step. Solar drying is best suited for tropical or arid environments however there are installations in both northern and mountainous climates.

During the drying cycle, the microprocessor evaluates the number of climatic variables, such as temperature, humidity and solar radiation, which initiates one or more operations that optimize the conditions inside the greenhouse. The greenhouse contains circulation fans and exhaust fans to provide convective drying and control of climatic conditions inside the greenhouse. The biosolids are periodically turned and aerated with varying devices depending on the manufacturer [see Fig. 14–14(c)]. In addition, low temperature waste heat can be used to

Figure 14–15

Cross section of a reed bed for dewatering and storage of biosolids.



enhance solar drying and reduce area requirements for the greenhouse. Ultimately, it is possible to obtain a dry pelletized material with a solids content as high as 90 percent.

Reed Beds

Reed beds can be used for biosolids dewatering at treatment plants with capacities up to 0.2 m^3/s (5 Mgal/d). A typical reed bed for biosolids dewatering, treatment, and storage is shown on Fig. 14–15. Reed beds are similar in appearance to subsurface flow constructed wetlands, which consist of channels or trenches filled with sand or rock to support emergent vegetation. The difference between reed beds used for biosolids application and subsurface flow wetlands is that the liquid biosolids are applied to the surface of the beds (as compared to subsurface application) and the filtrate flows through the gravel to underdrains.

Typically, reed beds are constructed of washed river-run gravel in the following layers: (1) a 250 mm (10 in.) deep drainage layer composed of 20 mm (0.8 in.) washed gravel, (2) a 250 mm (10 in.) deep layer composed of 4 to 6 mm (0.16 to 0.24 in.) washed gravel, and (3) a 100 to 150 mm (4 to 6 in.) layer of sand (0.4 to 0.6 mm). Sometimes an even coarser bottom layer is used. At least 1 m (3 ft) of freeboard above the sand layer is provided for a 10-y accumulation of sludge. Phragmites (reeds) are planted on 300 mm (12 in.) centers in the gravel layer just below the sand. Other wetland vegetation can be used, although reeds are the most popular. The first sludge application is made after the reeds are well established. Harvesting of the reeds is practiced typically in the winter by cutting the tops back to a level above the sludge blanket. Harvesting is necessary whenever the plant growth becomes too thick and restricts the even flow of biosolids. The harvested material can be composted, burned, or landfilled (Crites and Tchobanoglous, 1998).

The purpose of the plants is to provide a pathway for continuous drainage of water from the sludge layer. As the plants move back and forth due to wind currents, pathways are created for water to drain from the biosolids into the underdrains. The plants also absorb water from the sludge. Oxygen transfer to the plant roots assists in the biological stabilization and mineralization of the sludge. The reed bed system is a form of passive composting. The design loading rates for reed beds range from 30 to 60 $kg/m^2{\cdot}y$ (6 to 12 $lb/ft^2{\cdot}y$). Loading rates as high as 100 $kg/m^2{\cdot}y$ (20 $lb/ft^2{\cdot}y$) have been used, depending on the nature of the sludge and climatic conditions. The liquid sludge is applied intermittently, as in sand drying beds. The typical biosolids depth applied is 75 to 100 mm (3 to 4 in.) every week to 10 d (Crites and Tchobanoglous, 1998; Cooper et al., 1996).

Lagoons

Drying lagoons may be used as a substitute for drying beds for the dewatering of digested biosolids. Lagoons are not suitable for dewatering untreated sludges, limed sludges, or sludges with a high-strength supernatant because of their odor and nuisance potential. The performance of lagoons, like that of drying beds, is affected by climate; precipitation and low temperatures inhibit dewatering. Lagoons are most applicable in areas with high evaporation rates. Dewatering by subsurface drainage and percolation is limited by increasingly stringent environmental and groundwater regulations. If a groundwater aquifer used for a potable water supply underlies the lagoon site, it may be necessary to line the lagoon or otherwise restrict significant percolation.

Unconditioned digested biosolids are discharged to the lagoon in a manner suitable to accomplish an even distribution. Biosolids depths usually range from 0.75 to 1.25 m (2.5 to 4 ft). Evaporation is the principal mechanism for dewatering. Facilities for decanting of supernatant are usually provided, and the liquid is recycled to the treatment facility. Biosolids are removed mechanically, usually at a solids content of 25 to 30 percent. The cycle time for lagoons varies from several months to several years. Typically, biosolids are pumped to the lagoon for 18 months, and then the lagoon is rested for 6 mo. Solids loading criteria range from 36 to 39 $kg/m^3 \cdot y$ (2.2 to 2.4 $lb/ft^3 \cdot y$) of lagoon capacity (U.S. EPA, 1987a). A minimum of two cells is essential, even in very small plants, to ensure availability of storage space during cleaning, maintenance, or emergency conditions.

14–3 HEAT DRYING

Heat drying involves the application of heat to evaporate water and to reduce the moisture content of biosolids below that achievable by conventional dewatering methods. The advantages and disadvantages for heat drying are summarized in Table 14–13.

Heat-Transfer Methods

The classification of dryers is based on the predominant method of transferring heat to wet sludge or biosolids. Heat can be transferred by convection, conduction, radiation, or a combination of two or more methods. Dryers using infrared radiation have been used mostly for demonstration testing and are not covered in detail in this text.

All drying systems follow the three stages of drying which are illustrated in the drying curve shown on Fig. 14–16. The three stages of drying are warm up stage, constant

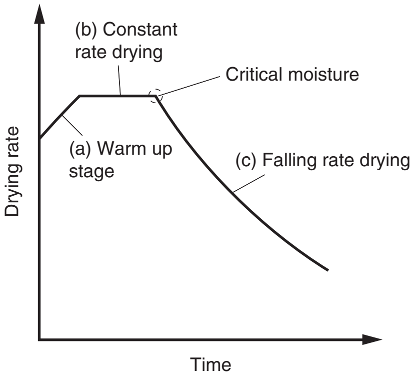
Table 14–13

Advantages and disadvantages of heat drying

Advantages	Disadvantages
1. Proven process	1. Relatively high capital cost
2. Product is readily marketable and suitable for a variety of uses that are consistent with the EPA's national beneficial reuse policy	2. Large fuel requirements
3. Small footprint	3. Payback from the sale of Class A production usually will not offset high operating costs
4. Reduced product transportation costs	4. Potential to create dust
5. Significant pathogen reduction (Class A)	5. Increased fire and explosion risk
6. Improved storage capability	6. Relatively complex system, requires highly trained operating staff
7. Does not require chemical additives	7. Odor potential and potential for odorous and dusty end product
8. Enhances heat value of biosolids	

Figure 14–16

Three stages of drying:
(a) warming up, (b) constant rate, and (c) falling rate.



rate drying stage, and the falling rate drying stage. During the warm up stage, the solids and moisture are heated up to the process temperature by transferring sensible heat. During the constant rate drying period, the moisture evaporated from the surface of the sludge or biosolids is replaced with internal moisture at an equal rate thus the surface remains saturated and the temperature of the sludge or biosolids surface being dried during this period is approximately equal to the wet bulb temperature of the gas present in the dryer. The constant rate period is typically the longest period and most drying occurs during this phase. The falling rate period occurs when most of the free moisture is removed and the rate of drying becomes controlled by diffusion of internal water to the surface of the sludge or biosolids being dried. During this period, the sludge or biosolids surface temperature begins to approach the gas or air temperature. The moisture content at the transition of the constant rate to the falling rate period is known as the critical moisture content.

Convection. In convection (direct drying) systems, wet sludge or biosolids are contacted directly with the heat transfer medium, usually hot gases. Under equilibrium conditions of constant rate drying, mass transfer is proportional to (1) the area of wetted surface exposed, (2) the difference between water content of the drying air and saturation humidity at the wet-bulb temperature of the solid-air interface, and (3) other factors, such as velocity and turbulence of drying air expressed as a mass transfer coefficient. The heat-transfer rate for evaporation is determined by the following equation (WEF, 2010):

$$q_{\text{conv}} = h_c A(T_g - T_s) \tag{14–1}$$

where q_{conv} = convective heat transfer rate, kJ/h (Btu/h)
h_c = convection heat transfer coefficient, kJ/m^2·h·°C (Btu/ft^2·h·°F)
A = area of the heated surface, m^2 (ft^2)
T_g = gas temperature, °C (°F)
T_s = temperature at sludge/gas interface, °C (°F)

The convection heat transfer coefficient can be obtained from dryer manufacturers or from pilot studies; however, many manufacturers consider this proprietary information.

Conduction. In conduction (indirect) drying systems, a solid retaining wall separates the wet sludge from the heat transfer medium, usually steam, thermal oil or another hot fluid. Heat transfer for conduction is determined by the following equation (WEF, 2010):

$$q_{\text{cond}} = h_{\text{cond}} A(T_m - T_s) \tag{14–2}$$

where q_{cond} = conductive heat-transfer rate, kJ/h (Btu/h)
h_{cond} = conductive heat-transfer coefficient, kJ/m^2·h·°C (Btu/ft^2·h·°F)
A = area of wetted surface exposed to gas, m^2 (ft^2)
T_m = temperature of heating medium, °C (°F)
T_s = temperature sludge at drying surface, °C (°F)

The conductive heat-transfer coefficient, a composite term, includes the effects of the heat-transfer surface films of the sludge and the medium. The conduction heat-transfer coefficient can be obtained from dryer manufacturers or from pilot studies.

Process Description

Heat dryers are classified as follows: direct, indirect, combined direct-indirect, and infrared. Direct and indirect dryers are described as they are the types most used commonly for municipal biosolids drying. Coal, oil, gas, infrared radiation, or dried sludge may be used as the means of supplying the energy for heat drying.

Direct Dryers. Direct (convection) dryers that have been used for drying municipal wastewater sludges and biosolids are the flash dryer, rotary dryer, and fluidized-bed dryer. Although as many as 50 municipal flash dryers have been installed in the United States since 1940, only Houston, TX, is still known to use this technology at two of its wastewater treatment plants. Because of safety concerns, high energy requirements, high O&M, and limited interest from vendors to work in the wastewater market, the popularity of this technology has declined, thus flash drying is not discussed in detail in this text (WEF, 2010). Rotary dryers are now very commonly used for wastewater sludges or biosolids. Fluidized-bed drying is a relatively new application in the United States.

Rotary Dryer. Rotary dryers have been used for the drying of raw primary sludge, waste-activated sludge, and digested biosolids from combined primary and WAS (see Fig. 14–17). In general, drying raw primary sludge is not recommended due to material handling concerns, odor and stability of the final product. A rotary dryer consists of a cylindrical steel shell that is rotated on bearings and usually is mounted with its axis at a slight slope from

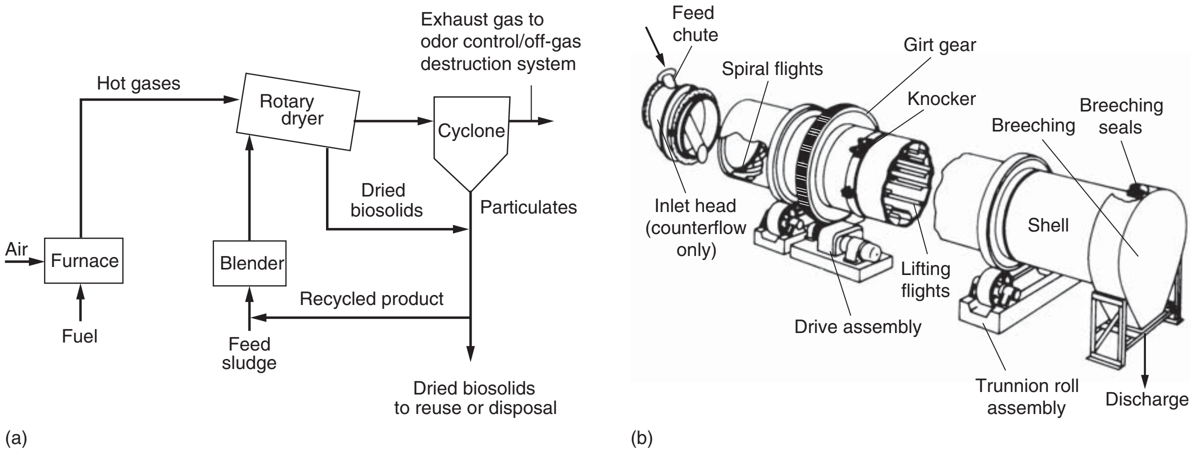
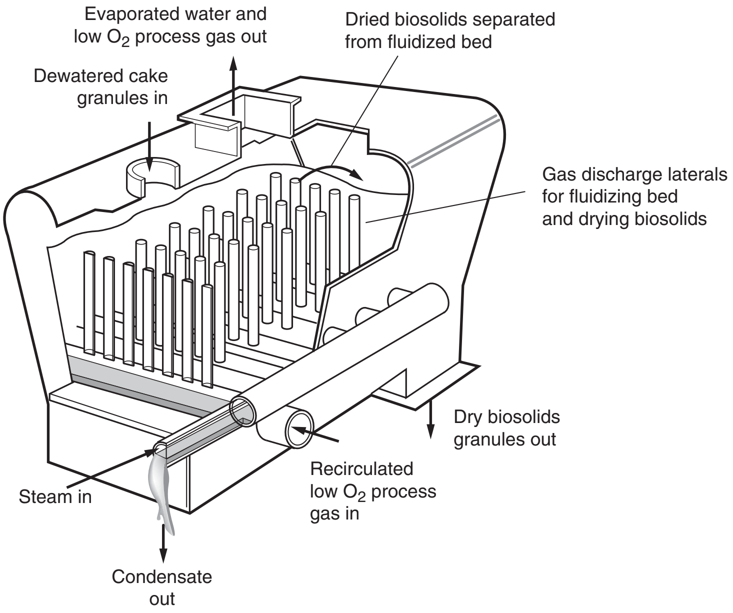


Figure 14–17

Rotary sludge dryer: (a) typical process flow diagram, and (b) isometric view of rotary dryer.

Figure 14–18

Cross-section through a fluidized-bed reactor. (From Andritz.)



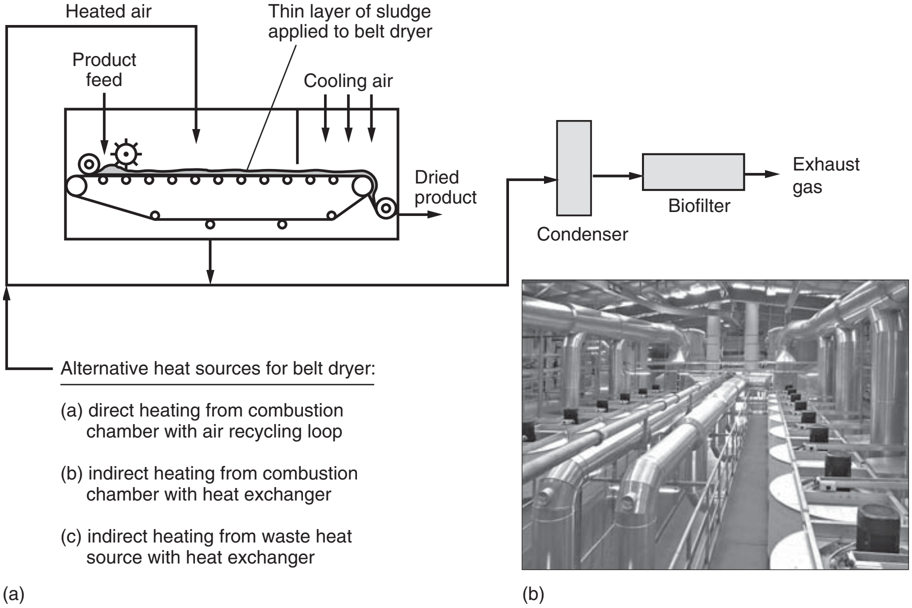
the horizontal [see Figs. 14–17(a) and (b)]. The feed sludge or biosolids is mixed with previously dried material in a blender located ahead of the dryer [see Fig. 14–17(a)]. The blended feed material has a moisture content of approximately 65 percent that improves its ability to move through the dryer without sticking. The mixture and hot gases are conveyed to the discharge end of the dryer. During conveyance, axial flights along the rotating interior wall pick up and cascade the sludge through the dryer. The product, which has a dry solids content of 90 to 95 percent, is screened and the oversize material passes through a crusher and then is transported to a recycle bin. The dried product is amenable to handling, storage, and marketing as a fertilizer or soil conditioner.

Fluidized-Bed Dryer. Fluidized-bed dryers, developed in Europe, were first adapted in the United States in the late 1990's (see Fig. 14–18). There are approximately 30 known installations worldwide with two in North America (WEF, 2012). The dryer has the capability of producing a pellet product, similar to that obtained from rotary drying systems (Holcomb et al., 2000). The method of feeding the dryer varies with different manufacturers. The fluidized-bed dryer consists of a stationary vertical chamber that is divided into three zones; the windbox or gas plenum, the heat exchanger, and the hood. The windbox is where the hot fluidizing gas is distributed through the bed. The heat exchanger is located within the fluidizing zone and heat is provided to the system either by steam or thermal oil. The hood is where the fluidizing gas exits the chamber and the dried biosolids are separated from the gas. The gas from the dryer typically passes through a cyclone before it is reheated and introduced back into the dryer. A portion of the gas, however, is removed from the closed loop and is scrubbed before being treated in an RTO or other odor control device.

The fluidizing gas, typically air, keeps the biosolids in suspension and provides agitation which assists with both heat transfer from the heat exchanger to the biosolids particles and also helps produce a relatively uniform dried biosolids product. A uniform temperature of 85 to 120°C (185 to 230°F) is maintained in the bed through intimate contact between the biosolids granules and the fluidizing air. Since fluid bed dryers operate at relatively low temperatures, they are suitable for low temperature waste heat recovery.

Figure 14–19

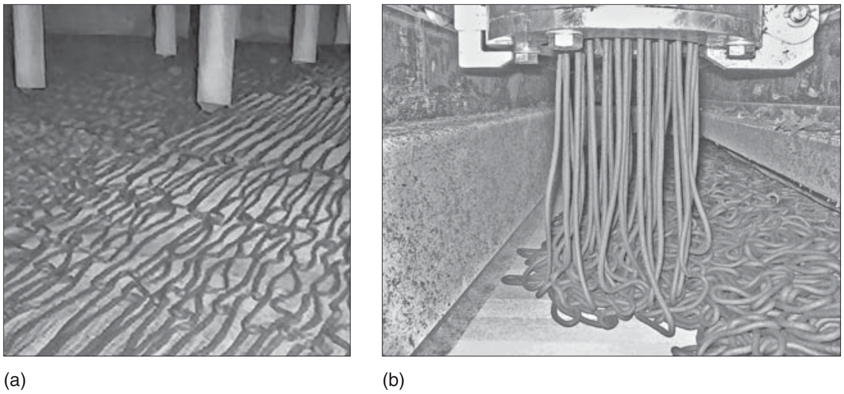
Example of belt dryer: (a) schematic of operation with three different heat sources, and (b) view of enclosed belt dryer (courtesy of SH+E Group U.S.).



Belt Dryers. At the time of writing this text, belt drying is still a relatively new type of sludge or biosolids dryer being implemented in the United States. However, there are up to seven belt dryers either in operation or under construction. Belt dryers are more widely used in Europe. Belt dryers are convection dryers that consist of a conveyor belt where biosolids are distributed in a thin layer (Fig. 14–19). The slowly moving belt transports the sludge or biosolids through the dryer while warm heated gases pass either through the belt and biosolids layer or pass across the biosolids layer to provide convective heat transfer. There are several manufactures currently available each with slightly different configurations and sludge or biosolids distribution methods. Some dryers contain metal mesh belts or perforated plates, while others contain fabric belts similar to what is used on belt filter presses. The dried product can be recycled and back mixed with dewatered cake, similar to rotary drum dryers, ensuring the biosolids being fed to the dryer are past the sticky phase and allowing for better distribution in the dryer. The sludge or biosolids fed to the dryer can also be extruded in ribbons (see Fig. 14–20) with a high drying surface area eliminating the need for back mixing systems. In the case of distributing the biosolids using extrusion method, belt drying cannot typically accept more than 30 percent cake solids.

Figure 14–20

Extruded biosolids for enhanced heat drying: (a) view of extruded biosolids strings (courtesy of Kruger) and (b) view of biosolids ropes being extruded (courtesy of SH+E Group U.S.).


(a) (b)

Belt dryers can be directly fired with flue gases from a hot gas furnace, or they can be indirectly heated with a heat exchanger using thermal oil, steam, or flue gas as the heating source to heat the drying air. Unlike many other types of dryers, belt dryers can also operate at very low temperatures, making this type of dryer very attractive for situations where low temperature waste heat is available. The low drying temperature and minimized agitation in the dryer makes this type of dryer inherently safer than other types of dryers. With the low temperature and high surface area requirements, belt dryers are typically better suited for small- to medium-sized plants as opposed to very large plants where they may become cost prohibitive. A typical covered belt drying installation is shown on Fig. 14–19(b).

Indirect Dryers. Indirect dryers are designed in either a horizontal or vertical configuration. A typical process flow diagram is shown on Fig. 14–21(a). Horizontal dryers employ paddles, hollow flights, or disks mounted on one or more rotating shafts to convey sludge or biosolids through the dryer [see Figs. 14–21(c) and (d)]. A heated medium, usually steam or oil, is circulated through the jacketed shell of the dryer and the hollow core of the rotating assembly. Dewatered sludge or biosolids are fed perpendicular to the dryer shaft and pass horizontally in a helical pattern through the dryer. The dryer performs the dual function of heat transfer and sludge or biosolids conveying. Drying occurs as the sludge or biosolids particles are broken up through agitation and come into contact with the heated metal surfaces in the dryer. The granular particles are both hot and abrasive.

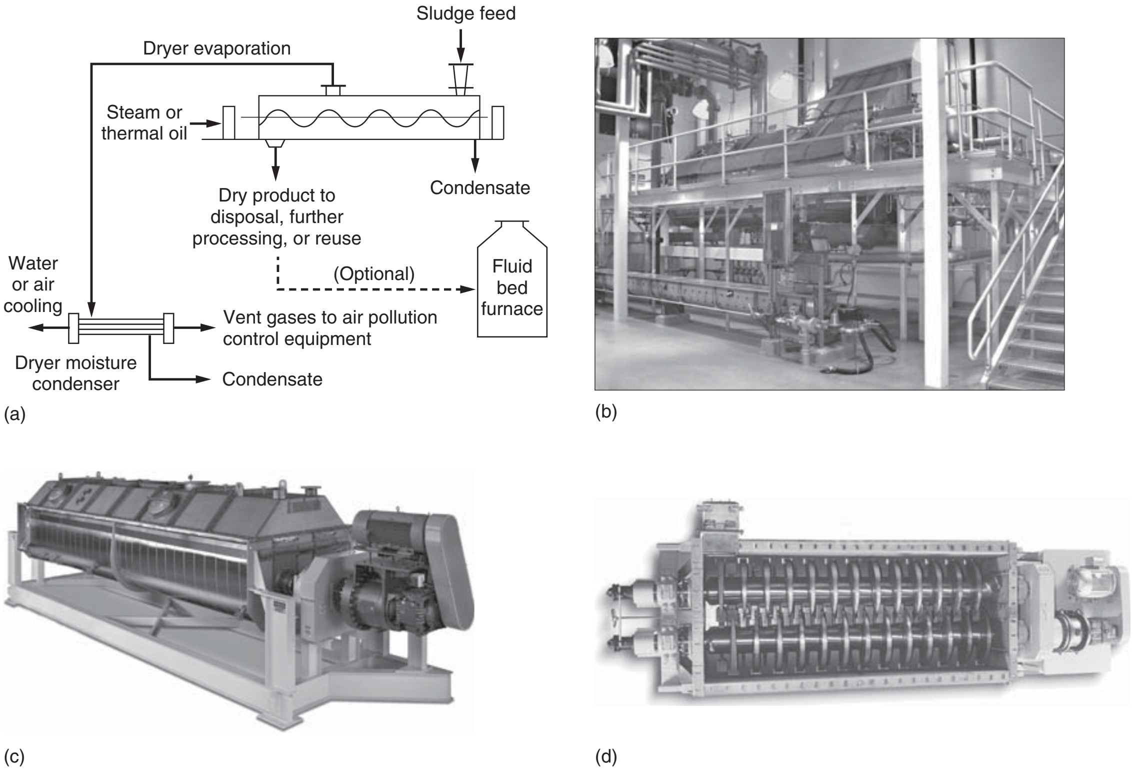


Figure 14–21

Indirect sludge dryer: (a) schematic process flow diagram, (b) view of typical dryer installation, (c) view of dryer without support facilities, and (d) view of interior paddle flights. [Figures (b), (c), and (d) courtesy of Komline-Sanderson.]

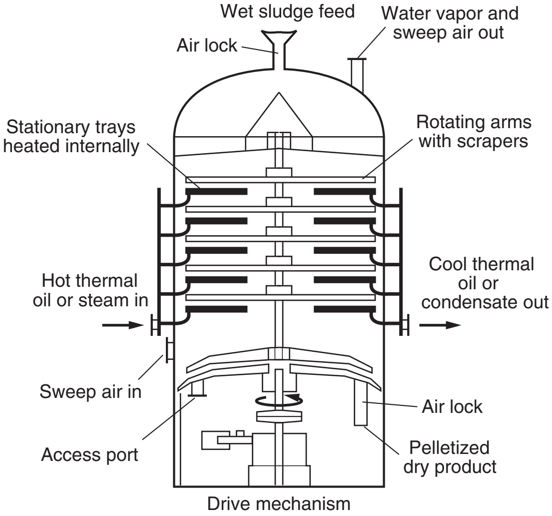


Figure 14–22

Cross section through a vertical indirect dryer (From Pelletech).

Moving parts, as in most dryers, will become abraded, and corrosion will accelerate the deterioration of the metal. The design of the agitator has to allow for efficient heat transfer, mixing of the sludge mass, and minimum fouling of the agitator. A weir at the discharge end ensures complete submergence of the heat-transfer surface in the biosolids being dried. The water vapor derived from the drying operation may be drawn off under a slightly negative pressure by an induced-draft fan located in the off-gas duct.

In vertical indirect dryers (see Fig. 14–22), sludge or biosolids contacts a metal surface heated by a medium, such as steam or oil, and the heat is conductively transferred to the sludge. The biosolids or sludge does not come in contact with the heating medium. Dewatered product (approximately 20 percent solids) mixed with recycled product is fed through the top inlet of a multistage dryer. Rotating arms move the sludge from the heated stationary tray to another in a rotating zigzag motion until it exits at the bottom as a dried, pelletized product. The rotating arms are equipped with adjustable scrapers that move and tumble the sludge in thin layers (20 to 30 mm) over the heated stationary trays. Dried product exits the dryer and is conveyed by a bucket elevator to a separation hopper. A portion of the properly sized pellets are cooled and conveyed to storage. Oversized particles are crushed and mixed with fines and a portion of the properly sized pellets and backmixed with the dewatered cake being fed to the dryer. Because of the backmixing process, this type of dryer produces a relatively uniform product similar to rotary drum dryers.

In indirect dryers, sludge or biosolids is dried to a specific level of dryness and discharged to a product conveyor for transfer to storage. Solids concentrations in the dryer product can range from 65 to over 95 percent depending on the ultimate use of the product.

Product Characteristics and Use

Dried biosolids are mainly used in a similar manner to commercial fertilizers, depending on the nutrient content of the material. Some of the dried material is well established as fertilizer and sold for profit to the producer. For example, OCEANGRO™, which is a dried pellet product from Ocean County Utilities Authority in New Jersey, is sold in bulk or in bags as a fertilizer to the agricultural community. The OCEANGRO™ guaranteed analysis is summarized in Table 14–14. The characteristics of the finished biosolids product depend on the type of sludge, the type of preprocessing, and the physical configuration of the drying surfaces. Raw primary sludge produces more of a fibrous, dusty, and odorous

Table 14–14
OCEANGROW™ guaranteed analysis

Ingredients	Unit	Value
Total Nitrogen (N)	%	5.0
Water Soluble Nitrogen	%	0.50
Water Insoluble Nitrogen[a]	%	4.50
Available Phosphate (P_2O_5)	%	5.00
Calcium (Ca)	%	2.50
Total Magnesium (Mg)	%	0.33
Combined Sulfur (S)	%	1.00
Total Copper (Cu)	%	0.04
Total Iron (Fe)	%	2.50
Total Zinc (Zn)	%	0.05
Standard Guide Number		150
Product Size Range	mm	1.5 to 2.5
Uniformity Index	%	60

[a] This product contains 4.50 percent slow release nitrogen.

material that is difficult to manage and pelletize. Digested sludge can be pelletized, depending on the type of dryer or downstream conditioning process, to produce an amorphous particle that can be easily handled and transported. Dryers with backmixing typically produce more uniform pellets but other downstream processes such as screening, conditioning agents or pelletizing can be added to dryers without backmixing if a more uniform product is required for marketing purposes. The best size range for marketing is approximately from 2 to 4 mm, but can vary depending on the specific market (WEF, 2012). To maximize marketing potential, screening for sizes smaller and larger than the selected size range will be required. The fines and oversized particles can be returned to blend with the incoming sludge, hence increasing the solids content entering the dryer but not changing the amount of moisture that must be evaporated.

Biosolids dried products can be used as an energy source depending on the characteristics of the sludge or biosolids that were dried. In recent years, dried biosolids have been used as a coal substitute for fuel in cement kilns. The use of dried product as an energy source is expected to increase, due to high energy costs and because the energy recovered from dried biosolids is considered renewable energy.

Product Transport and Storage

Although the granular product from heat drying is reasonably durable, long mechanical conveyors, such as screw conveyors, drag conveyors, and pneumatic conveyors that create an abrasive action could cause crumbling and dust formation. These types of conveyors, however, are commonly used at many sites. If product friability is a concern, open or folded belt conveyors may be preferred.

Upon exiting the dryer, the hot biosolids should be cooled to below 50°C (120°F) before placing in silos or storage vessels. The combination of initial heat plus heat that could be introduced by biological activity in the silo can cause smoldering or an open flame. This condition can occur where the drying operation is expected to pass through frequent start-and-stop cycles and drying may not be complete. In general, the product entering storage should be from 92 to 98 percent dry solids.

Fire and Explosion Hazards

With the fine particles and high levels of dryness in heat-dried sludge, hazards due to fire and explosion may exist in the dryer or when dried sludge is conveyed or stored. An organic dust suspended in air can rapidly combust if exposed to an ignition source. The heat of combustion can rapidly increase the volume and/or pressure of hot combustion products. If the pressure exceeds the rupture strength of the containing vessel, an explosion occurs. The phenomenon is called "deflageration," and deflageration explosions are the most serious concern when handling dried biosolids (Haug et al., 1993). Dried biosolids can also reheat if rewetted due to biological activity. In addition, the sludge being dried could contain a high level of fiber or grease content which could cause problems inside the dryer. Design considerations that are recommended for safety purposes and to prevent thermal events are given in Table 14–15.

Air Pollution and Odor Control

Two important control measures associated with heat drying of sludge are dust collection and odor control. All dryers produce some sort of offgas which results from the process gas and evaporated water that is continuously removed from the drying process. The offgas can contain

Table 14–15

Prevention measures to avoid dust hazards in heat drying[a]

Item	Prevention measure
Venting system (for processing, conveyance, and storage components)	Provide explosion relief vents. Size explosion vents for "worst case" explosion in an air atmosphere.
Temperature control	Use controls to prevent high temperatures which could cause unsafe conditions Temperature control can also be used to prevent the material from being over dried which could cause excess dust formation.
Water deluge	Include a water deluge or sprinkler system that reacts based on high temperature conditions. Water deluge can also be set to react to the presence of carbon monoxide or carbon dioxide which could indicate smoldering.
Nitrogen padding	Provide a nitrogen inerting atmosphere for all dried biosolids conveyance and processing facilities. Maintain oxygen levels below 5 percent by volume to reduce potential for self-heating and ignition of hot biosolids.
Electrical equipment	Design in accordance with appropriate National Fire Protection Association criteria. If dust is present, all equipment must be dust tight and electronic cabinets nitrogen purged. Motor control centers that contain sparking devices, such as starters and relays, must be located outside classified areas.
Ducts and vessels	Electrically bond and ground all conductive elements of the system that contact dried biosolids.
Maintenance	Keep areas clean to prevent accumulation of dust. Any vessel containing powder must have powder removed before opening or the powder must be cooled to ambient temperatures before safe entry clearance is given.
Product cooling	Ensure product is cooled to below 50°C (120°F) before being transported to product storage.
Miscellaneous	Eliminate or move outside all heat sources from classified areas. Equip electric motors located in a Class II, Division 2 area with Class F insulation to reduce "skin" temperatures.

[a] Adapted from Haug, et al. (1993) and WEF (2012).

high levels of odors due to volatilization and heat reactions. The method for dealing with dryer offgas varies depending on the dryer type, manufacturer and local site requirements.

Most dryers include wet scrubbers that have relatively high efficiencies and will condense the water evaporated from the sludge as well as remove some of the organic matter in the vent gas. Most wet scrubbers will also include mist eliminators to minimize carryover of water droplets. Some dryers also include cyclone separators, either on the process air circulation loop or upstream of the wet scrubber, to remove particles entrained in the air. The cyclones used typically have efficiencies of 75 to 80 percent and are suitable for vent gas temperatures up to 340 or 370°C (650 or 700°F). The final scrubbed off gas must then go through some sort of odor control treatment. The use of a thermal oxidizer is one common method used on many types of dryers to eliminate odors. Thermal oxidizers typically operate at temperatures greater than 815°C (1500°F) with a residence times of 0.75 to 1 s (WEF, 2012). The dryer offgas can also be sent to an incinerator (if one is onsite), or it can be mixed with combustion air for the dryer system's burner (if the dryer is indirectly fired). Other options that have been used for off-gas odor control include chemical scrubbers, biofilters, and directing the off-gas to diffusers in the aeration basin of the activated sludge process.

14–4 ADVANCED THERMAL OXIDATION

Incineration, referred to in this chapter as advanced thermal oxidation (ATO) of sludge and biosolids, involves the total conversion of organic solids to oxidized end products, primarily carbon dioxide, water, and ash. The major advantages and disadvantages of incineration are summarized in Table 14–16. ATO is used most commonly by medium to large sized plants with limited disposal or reuse options.

Sludges processed by ATO are usually first dewatered. It is normally unnecessary to stabilize sludge before incineration. In fact, such practice may be detrimental because stabilization, specifically aerobic and anaerobic digestion, decreases the volatile content of the sludge and consequently increases the requirement for an auxiliary fuel. When ATO is practiced for biosolids, it is desirable that the dewatered biosolids concentration be between 30 to 35 percent,

Table 14–16

Advantages and disadvantages of incineration

Advantages	Disadvantages
1. Maximum volume reduction and end product stability thereby lessening disposal requirements	1. High capital and operating cost
2. Maximum destruction of pathogens and toxic compounds	2. Complex system that requires highly trained operating staff
3. Relatively small process footprint	3. Possible adverse environmental effects from air emission and ash
4. Greatest control over final biosolids disposition	4. Disposal of residuals which may be classified as hazardous wastes if they exceed prescribed maximum pollutant concentrations
5. Well suited for larger plants with larger disposal requirements	5. Relatively lengthy implementation timeframe due to planning and public consultation requirements
6. Energy recovery potential	6. Air permitting can be tedious and difficult
7. Well proven process	7. May not be feasible for a nonattainment area

to reduce the use of an auxiliary fuel and achieve autogenous oxidation. In Europe, ATO has been used with dewatered biosolids from thermally hydrolyzed sludges as discussed in Chap. 13. Sludges may be thermally oxidized separately or in combination with municipal solid wastes. The ATO processes considered in the following discussion include multiple-hearth incineration, fluidized-bed incineration, and coincineration with municipal solid waste. Before discussing these processes, some fundamental aspects of complete combustion are introduced.

Fundamental Aspects of Complete Combustion

Combustion is the rapid exothermic oxidation of combustible elements in fuel. ATO is complete combustion. The predominant elements in the carbohydrates, fats, and proteins composing the volatile matter of sludge are carbon, oxygen, hydrogen, sulfur, and nitrogen (C-O-H-S-N). Other major components include moisture and ash. The approximate percentages of these may be determined in the laboratory by a technique known as ultimate analysis (ASTM, 2009) and proximate analysis (ASTM, 2009). Heating (or calorific) value may be obtained using ASTM (2011).

Oxygen requirements for complete combustion of a material may be determined by knowing its constituents, assuming that carbon and hydrogen are oxidized to the ultimate end products CO_2 and H_2O. The formula becomes

$$C_aO_bH_cN_d + (a + 0.25c - 0.5b)O_2 \rightarrow aCO_2 + 0.5cH_2O + 0.5dN_2 \tag{14–3}$$

The theoretical quantity of air required will be 4.35 times the calculated quantity of oxygen because air is composed of 23 percent oxygen on a mass basis. To ensure complete combustion, sufficient excess air is required. Too little excess air may lead to poor emissions performance and may also limit the throughput of the process. Too much excess air may lead to high auxiliary fuel usage and unnecessarily large equipment sizing. In some jurisdictions, air pollution regulations may require a minimum flue gas oxygen concentration. The design range of excess air required depends on the ATO process configuration, the system design, and the characteristics of the biosolids. For some units operating with very dry, high heating value biosolids, air is also used to quench, control, and spread the heat released in the process to maintain below an upper temperature limit. In this case, the air required for cooling may far exceed the theoretical quantity of air.

A materials balance must be made to include the above compounds and the inorganic substances in the sludge, such as the inert material (ash), moisture, and the other constituents of the air (primarily the approximately 77 percent N_2 in the air that is supplied along with O_2 reaction in Eq. (14–3). The specific heat of each of these substances and of the products of combustion must be taken into account in determining the heat required for the incineration process.

Heat requirements will include the sensible heat Q_s in the ash, plus the sensible heat required to raise the temperature of the flue gases to 760°C (1400°F) or whatever temperature of operation is selected for complete oxidation, elimination of odors, and assurance of suitable environmental performance less the heat recovered in preheaters or recuperators. Heat loses to the ambient space around the ATO process is also a heat requirement. Latent heat must also be furnished to evaporate all of the moisture in the sludge. The total heat required may be expressed as

$$Q = \sum Q_S + Q_E + Q_L = \sum C_P W_S (T_2 - T_1) + W_W \lambda + Q_L \tag{14–4}$$

where Q = total heat, kJ (Btu)
Q_S = sensible heat in the ash, kJ (Btu)
Q_E = latent heat, kJ (Btu)
Q_L = heat loss

C_p = specific heat for each category of substance in ash and flue gases, kJ/kg°C (Btu/lb°F)
W_s = mass of each substance, kg (lb)
W_w = mass of water, kg (lb)
T_1, T_2 = initial and final temperatures
λ = latent heat of evaporation, kJ/kg (Btu/lb)

Reduction of moisture content of the sludge is the principal way to lower heat requirements, and the moisture content may determine whether additional fuel will be needed to support combustion. It is desirable both from a cost perspective and from an environmental sustainability perspective not to have to add additional fuel to the process. Therefore, it is desirable that the heat required to drive the process should come from the exothermic oxidation of the biosolids' volatile components, which is called "autogenous combustion."

The heating value of a sludge may be estimated by using Eq. (2–66) presented previously in Chap. 2 and repeated here for convenience:

$$\text{HHV (MJ/kg)} = 34.91\ \text{C} + 117.83\ \text{H} - 10.34\ \text{O} - 1.51\ \text{N} + 10.05\ \text{S} - 2.11\text{A} \qquad (2\text{–}66)$$

where, HHV = higher heating value, MJ/kg (Btu/lb = MJ/kg × 0.00043)
C = carbon, percent by weight expressed as a decimal (dry basis)
H = hydrogen, percent by weight expressed as a decimal (dry basis)
O = oxygen, percent by weight expressed as a decimal (dry basis)
N = nitrogen, percent by weight expressed as a decimal (dry basis)
S = sulfur, percent by weight expressed as a decimal (dry basis)
A = ash, percent by weight expressed as a decimal (dry basis)

The comparable expression proposed by the U.S. EPA (1979) and WEF (2010) is

$$\text{HHV (MJ/kg)} = 33.83\ \text{C} + 144.70\ (\text{H} - \text{O}/8) + 9.42\ \text{S} \qquad (14\text{–}5)$$

The fuel value of sludge ranges widely depending on the type of sludge and the volatile solids content. The fuel value of untreated primary sludge is the highest, especially if it contains appreciable amounts of grease and skimmings. Where kitchen food grinders are used, the volatile and thermal content of the sludge will also be high. Digested biosolids have significantly lower heating values than raw sludge. Typical heating values for various types of sludge and biosolids are reported in Table 14–17. The heating value for sludge is equivalent to that of some of the lower grades of coal. Computation of the heating value of biosolids is illustrated in Example 14–2.

Table 14–17

Typical heating values for various types of sludge and biosolids[a]

	Btu/lb of total solids[b]		kJ/kg of total solid[b]	
Type of sludge or biosolids	**Range**	**Typical**	**Range**	**Typical**
Raw primary	10,000–12,500	11,000	23,000–29,000	25,000
Activated	8500–10,000	9000	20,000–23,000	21,000
Anaerobically digested primary	4000–6000	5000	9000–14,000	12,000
Raw chemically precipitated primary	6000–8000	7000	14,000–18,000	16,000
Biological filter	7000–10,000	8500	16,000–23,000	20,000

[a] Adapted, in part, from WEF (1988).

[b] Lower value applies to plants with long solids retention time.

EXAMPLE 14–2 **Energy Content of Biosolids** A wastewater treatment plant is considering an ATO for handling of their biosolids. The plant currently dewaters waste activated sludge on belt filter presses prior to lime addition for Class B land application. A sample of dewatered sludge, before lime addition, was sent out for an ultimate analysis and the results are shown below.

Parameter	As received basis	Dry basis
Carbon	6.84	41.33
Hydrogen	0.94	5.66
Oxygen	3.71	22.41
Nitrogen	0.92	5.57
Sulfur	0.14	0.86
Ash	4.00	24.17
Moisture	83.45	0.00
HHV (MJ/kg)	2.96	17.88

Compare the lab results of the HHV on a dry basis to the theoretical HHV's calculated from Eq (2–66) and Eq (14–5).

Solution

1. Determine the energy content of the biosolids using Eq. (2–66).

$$\text{HHV (MJ/kg)} = 34.91\text{ C} + 117.83\text{ H} - 10.34\text{ O} - 1.51\text{ N} + 10.05\text{ S} - 2.11\text{ A}$$

$$\text{HHV (MJ/kg)} = 34.91\,(41.33/100) + 117.83\,(5.66/100) - 10.34\,(22.41/100) - 1.51\,(5.57/100) + 10.05\,(0.86/100) - 2.11\,(24.17/100)$$

$$\text{HHV} = 18.27\text{ MJ/kg}$$

$$\text{Percent difference (measured vs. calculated)} = \left(\frac{17.88 - 18.27}{17.88}\right)100 = -2.18\%$$

2. Determine the energy content of the biosolids using Eq. (14–5).

$$\text{HHV (MJ/kg)} = 33.83\text{ C} + 144.70\,(\text{H} - \text{O}/8) + 9.42\text{ S}$$

$$\text{HHV (MJ/kg)} = 33.83\,(41.33/100) + 144.70\,(5.66/100 - (22.41/100)/8) + 9.42\,(0.86/100)$$

$$\text{HHV} = 18.20\text{ MJ/kg}$$

$$\text{Percent difference (measured vs. calculated)} = \left(\frac{17.88 - 18.20}{17.88}\right)100 = -1.79\ \%$$

Comment The computed results from both Eq. (2–66) and Eq. (14–5) are very close to each other. There is also good agreement, within 3 percent, between the predicted and measured values. For theoretical calculations Eq. (2–66) is favored because it was derived from an analysis of hundreds of organic feed stocks and includes nitrogen and ash, whereas Eq. (14–5) is a modification of the well-known Dulong formula for coal. To design an ATO system, a detailed heat balance must be prepared. Such a balance must include the energy required to evaporate water in the sludge and heat losses from the process equipment, ductwork, stack, and ash as well as any heat recovery (such as combustion air preheating). Heat is obtained from the combustion of volatile matter in the sludge and from the burning of auxiliary fuels. If the process is autogenous, the auxiliary fuel is needed only for warming up the ATO or adjusting from a process upset should the desired temperature be too low or operation become unstable. Regardless of the feed biosolids characteristics, the design should include provisions for auxiliary heat for startup and for assuring complete oxidation at the desired temperature under all conditions. Fuels such as oil, natural gas, or excess digester gas are typically used for supplemental heating purposes.

To design an ATO system, a detailed heat balance must be prepared. Such a balance must include heat losses through the walls and pertinent equipment of the incinerator, as well as losses in the stack gases and ash. Approximately 4.0 to 5 MJ (4000 to 5500 Btu) are required to evaporate each kg (2.2 lb) of water in the sludge. Heat is obtained from the combustion of the volatile matter in the sludge and from the burning of auxiliary fuels. If the process is autogenous, auxiliary fuel is needed only for warming up the incinerator or adjusting from a process upset should the desired temperature be too low or operation become unstable. Regardless of the feed biosolids characteristics, the design should include provisions for auxiliary heat for startup and for assuring complete oxidation at the desired temperature under all conditions. Fuels such as oil, natural gas, or excess digester gas are suitable.

Multiple-Hearth Incineration

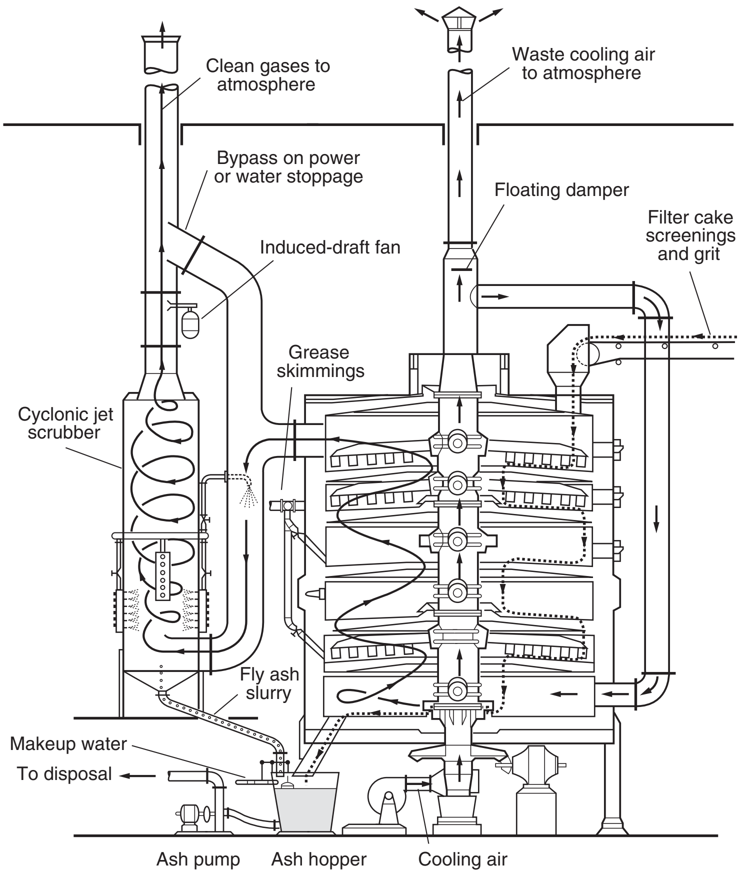
Multiple-hearth incineration is used to convert dewatered sludge cake to an inert ash. Because the process is complex and requires specially trained operators, multiple-hearth furnaces are normally used only in large plants. Multiple-hearth incinerators have been used at smaller facilities where land for the disposal of sludge is limited and at chemical treatment plants for the recalcining of lime sludges. As of 2011, about 70 percent of ATO installations in the United States utilized multiple-hearth units. These units are all fairly old, all having been installed in the mid and late twentieth century.

Process Description. As shown on Fig. 14–23 , a multiple-hearth incinerator is set up as a counter flow process with a series of hearths (typically 7–11) where the sludge introduced into the unit is successively, dried, combusted, cooled, and discharged. Air and flue gases are the heat exchange fluids in the vessel and generally travel counter-current to the biosolids. The sludge cake is fed onto the top hearth and is slowly raked to the center by a series of teeth mounted on a set of rake (or “rabble”) arms. The teeth direct the solids in a spiral pattern toward the center with each consecutive arm pushing a ridge of biosolids in toward the center and in the direction of rotation. The teeth also act to agitate the exposed surfaces of the biosolids and to mix the solids in an attempt to provide even exposure to the heat and air. From the center, sludge cake drops to the second hearth, where the rabble arms and teeth move it to the periphery. The sludge cake then drops to the third hearth and is again raked to the center. The hottest temperatures are typically on the middle hearths, where the sludge begins combustion after having evaporated enough water to raise the solids to a high enough temperature to sustain combustion. Preheated air is introduced to the lowest hearth and gains heat as it travels pass the sludge where the fixed carbon continues to burn off and the remaining ashes cool prior to discharge from the unit. The air is further heated by the sludge as the air rises past the middle hearths where combustion occurs and generates flue gases. The flue gases then cool as it gives up its heat to dry the incoming sludge on the top hearths.

Because air and flue gases are used as the primary heat exchange fluid in this counter flow system, proper air and flue gas flow, sufficient convection, and mixing is required to

Figure 14–23

Corss section through a typical multiple-hearth incinerator.



operate the system properly. The rabble tooth pattern is also important to provide even distribution, and frequent overturning of the biosolids on the hearths. Auxiliary fuel may be added at many different points on the unit. Typically, there are 3 to 4 burners around the circumference of a hearth and every second hearth is fitted with burners. By choosing the location and firing rate for these burners, the temperature and location of the combustion zone within the furnace can be changed to suit the material being combusted. The highest moisture content of the flue gas is found on the top hearths where sludge with the highest moisture content is heated and some water is vaporized. Cooling air is initially blown into the central column and hollow rabble arms to keep them from overheating. A large portion of this air, after passing out of the central column at the top, is recirculated to the lowest hearth as preheated combustion air.

Operational Controls. Operators may set feed rate, excess air, air injection location, burner location, burner firing rate, rake speed, etc. Given the large number of interrelated and codependent process variables for a multiple hearth furnace, consistent operation is sometimes challenging. These furnaces generally operate at relatively high levels of excess air (up to 100 percent) to avoid localized air/oxygen deficits and to provide sufficient buffer to ensure consistent, quality combustion.

Feed sludge must contain more than 15 percent solids because of limitations on the maximum evaporating capacity of the furnace. Auxiliary fuel is required usually when the feed sludge contains between 15 and 30 percent solids. Feed sludge containing more than

50 percent solids may create temperatures in excess of the refractory and metallurgical limits of standard furnaces. Average loading rates of wet cake are approximately 40 kg/m^2•h (8 lb/ft^2•h) of effective hearth area but may range from 25 to 75 kg/m^2•h (5 to 15 lb/ft^2•h).

Process Variables. Variations in the basic form exist with some furnaces having varying numbers of hearths, an array of systems to introduce auxiliary air for combustion, cooling, or mixing, upper hearths with little or no exposure to the biosolids (typically termed a "zero hearth" or an "on-hearth afterburner"), a wide degree of burner configurations, and flue gas recirculation to promote enhanced combustion controls.

In addition to dewatering, multiple hearth incinerators have a large amount of required ancillary processes. On the unit itself there are at least 2 or more fans for burner air and center shaft cooling air, additional fans are required for auxiliary combustion air and flue gas recirculation, a large number of burners (typically between 9 and 36 per unit) each with a gas and combustion air supply train, and the center shaft drive and gear reducer. In addition, there are ash handling systems and some type of wet or dry scrubber to meet air pollution requirements. Other ancillaries include an induced draft fan, bypass stack, any additional air pollution controls, and possibly energy recovery.

Air Pollution Control. In the most basic units, the air pollution controls are wet scrubbers. In these units, scrubber water comes in contact with and removes most of the particulate matter in the exhaust gases. The recycle BOD and COD is nil, and the total suspended solids content is a function of the particulates captured in the scrubber. Under proper operating conditions, particulate discharges to the air from wet scrubbers are less than 0.65 kg/10^3 kg (1.3 lb/ton) of dry sludge input. With recent (2011) upgrades to the regulations in the United States, many of the existing multiple hearth incinerators will require upgrades to the air pollution controls. Additional details on air pollution controls are discussed later in this chapter.

Ash Handling. Ash handling may be either wet or dry. In the wet system, the ash falls into an ash hopper located beneath the furnace, where it is slurried with water from the exhaust gas scrubber. After agitation, the ash slurry is pumped to a lagoon or is dewatered mechanically. The effluent water from the ash lagoon or ash dewatering process can either be sent back into the wastewater treatment process or discharged back into the plant effluent and sent to the outfall. In the dry system, the ash is conveyed mechanically to a storage hopper for discharge into a truck for eventual disposal. The ash is usually conditioned with water. Ash density is about 5.6 kg/m^3 (0.35 lb/ft^3) dry and 880 kg/m^3 (55 lb/ft^3) wet.

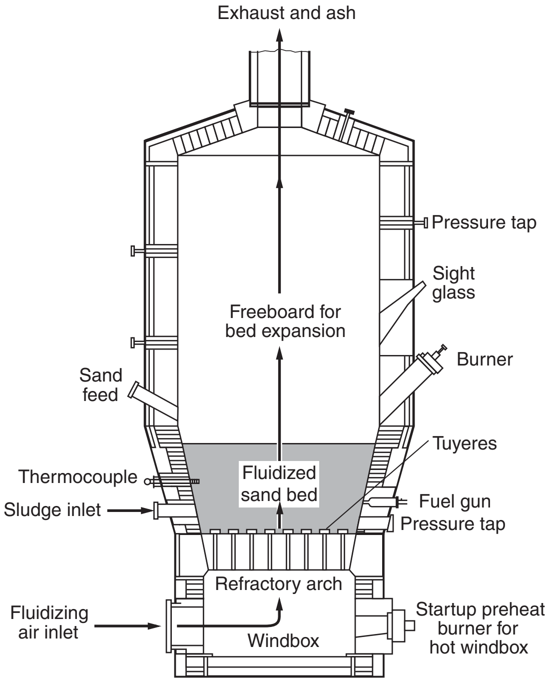
Fluidized-Bed Incineration

The fluidized-bed incinerator used commonly for sludge incineration is a vertical, cylindrically shaped, refractory-lined steel shell that contains a sand bed (media) and fluidizing air orifices to produce and sustain combustion (see Fig. 14–24). As of 2011, about 30 percent of ATO installations in the United States utilized fluidized-bed technology. These units tend to be much newer than the multiple-hearth installations that they are typically replacing. As of 2012, all of the new full-scale units currently being installed are fluidized-bed incinerators.

Process Description. The fluidized-bed incinerator ranges in size from 2.7 to 9.1 m (9 to 30 ft) in diameter. The incinerator is generally composed of three areas. These are (from bottom to top): (1) the windbox, (2) the sand bed, and (3) the freeboard. The windbox acts as an air plenum for distributing the fluidizing air, and may also have one or more burners for startup. Above the windbox is the sand bed. When quiescent, the

Figure 14–24

Cross section through a typical fluidized-bed incinerator.



sand bed is approximately 0.8 m (2.5 ft) thick and rests on a brick dome, refractory-lined grid, or a steel plate that keeps the sand bed separated from the windbox. The sand bed support area contains orifices, called "tuyeres," through which the fluidizing air is injected into the incinerator at a pressure of 20 to 35 kN/m^2 (3 to 5 lb$_f$/in.^{2}) by fluidizing air blowers. At low velocities, combustion gas "bubbles" appear within the fluidized bed. The main bed of suspended particles remains at a certain elevation in the combustion chamber and "boils" in place. Units that function in this manner are called "bubbling-bed" incinerators. The mass of suspended solids and gas, when active and at operating temperatures, expands to about double the at-rest volume. Sludge is mixed quickly within the fluidized-bed by the turbulent action of the bed. If required and auxiliary fuel injection ports are available, auxiliary fuel, such as oil, natural gas, or digester gas can be injected directly into the bed.

The minimum temperature needed in the sand bed prior to injection of sludge is approximately 700°C (1300°F). The temperature of the sand bed is controlled between 760 and 820°C (1400 and 1500°F). Evaporation of the water and combustion of the sludge solids takes place rapidly due to the high turbulence. The freeboard area, located above the bed, provides for residence time for the completion of combustion of the gaseous constituents. The flue gas (products of combustion) and ash leave the gas outlet through the top of the incinerator. No ash exits from the bed at the bottom of the incinerator as it is entrained with the flue gases.

Most fluid bed units that have been designed for autogenous operation should not require any auxiliary fuel after startup. The large thermal mass within the unit (i.e., the sand bed and refractory dome) helps to dampen fluctuations in heat input and also tends to retain the temperature in the unit if required to shut down for a limited duration. The large thermal mass tends to provide a stable, slow acting process that is robust, reliable, and controllable by most operators.

The fluidized-bed, though reliable, is complex and requires the use of trained personnel. Because fluidized-bed incinerators are complex, they are normally used in medium to large plants, but may be used in plants with lower flow ranges where land for the disposal of sludge is limited.

The main process vessel is very simple and contains no moving parts, which is a significant difference from the multiple-hearth incinerator design. The ancillary equipment required includes a fluidizing air blower, a burner combustion air blower, auxiliary fuel train, a sand system, an air pollution control system, and an ash handling system (which is typically combined with the air pollution control system).

Scrubber Water. Most of the ash (over 99 percent) is captured in the scrubber water, and the total suspended solids content is approximately 20 to 30 percent of the dry solids feed (inline with the ash content of feed sludge). Plant effluent is typically used as scrubber water at a rate of approximately 25 to 40 L/kg (3 to 5 gal/lb) of dry solids feed to the fluidized-bed. The ash slurry flows from the scrubber are normally directed to an ash lagoon or are mechanically dewatered to separate the ash from water. The effluent from the ash lagoon or ash dewatering process can either be sent back to the head of the wastewater treatment process or combined with the plant effluent at the outfall. The concentrations of BOD and COD are low, typically below 50 mg/L. Additional details on air pollution control are noted later in this chapter. Particulates, criteria air contaminants, and other air emissions are typically much less than a similarly sized multiple-hearth incinerator.

Process Modifications. Many new fluid bed units now incorporate at least one downstream air-to-air heat exchanger. The heat exchanger is used to transfer the heat from the exiting flue gases to the preheat the fluidizing air before it is introduced into the unit below the tuyeres. In this manner, heat is preserved within the fluid bed unit, despite the high volumetric flowrate of air through the bed. These units are called "hot windbox" fluid beds and are able to operate autogenously on biosolids with higher moisture contents. Some new units also incorporate energy recovery systems (separate from the fluidizing air preheat heat exchanger), which is discussed below.

A modification of the fluidized-bed incineration technology is the "circulating-bed" incinerator. In the circulating-bed unit, the reactor gas passes through the combustion chamber at much higher velocities, ranging from 3 to 8 m/s (10 to 25 ft/s). At these velocities, the bubbles in the fluidized bed disappear and streamers of solids and gas prevail. The entire mass of entrained particles flow up the reactor shaft to a particle separator, are deposited in storage momentarily, and are recirculated back to the primary combustion zone in the bottom of the reactor. Ash is removed continuously from the bottom of the bed. On turndown, the circulating bed becomes a bubbling bed.

Energy Recovery from Thermal Oxidation

Energy recovery from thermal oxidation processes is now an important component of the overall ATO process. The design of hot wind box fluidized-bed units incorporate at least one fluidizing air preheat heat exchanger as part of the main process to allow for highly efficient operation. Additional heat exchangers may be utilized to reheat the flue gases further downstream depending on the requirements of the air pollution control system. The reheating of the flue gases prior to stack tip discharge may also be desired for enhanced atmospheric dispersion or for aesthetic/plume suppression reasons. The heat recovery heat exchangers can also be used generate hot water, steam, or heated thermal oil for waste heat recovery or electricity production.

The quantity, quality, and form of the energy extracted from the ATO process depend on the heat and solids content of the dewatered sludge fed to the process. Undigested sludge with solids content greater than 25 percent offers good potential for energy recovery. One example

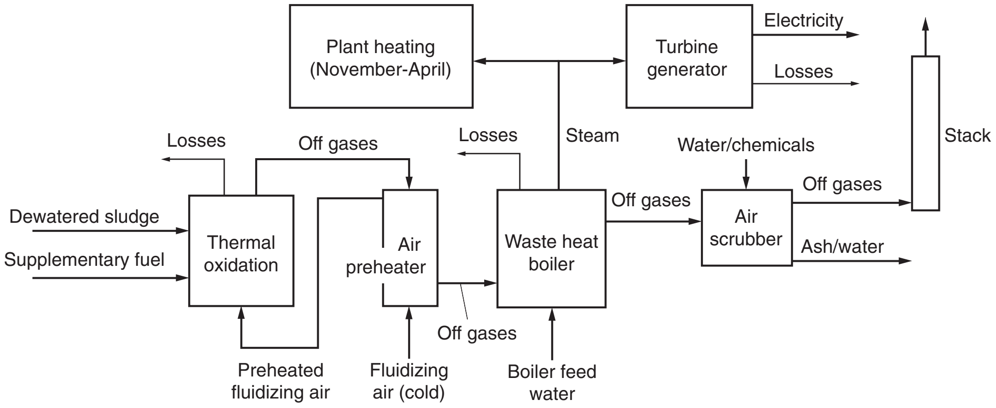


Figure 14–25

An example of an energy recovery process diagram.

of energy recovery is shown on Fig. 14–25. The hot off gases from thermal oxidation process is first introduced to an air preheater, then fed to a waste heat boiler that produces steam, which is fed to a steam turbine generator producing electricity. Depending on the electricity and fuel costs, the steam can be used to heat the buildings in the winter months instead of the turbine generator. Other uses for the steam extracted from the ATO process include process heating for digesters, heating FOG systems, or dewatering and drying processes.

Because generating and operating high pressure steam facilities require specialized boiler and pressure vessel operators, this application is not used widely. In this case, the hot off gases generated from the thermal oxidation process can be used to heat thermal oil or hot water that is used to generate electricity using the organic Rankin cycle. Other sources of waste heat (typically lower grade heat sources) which have been utilized at biosolids ATO facilities include capturing and recuperating heat lost from the exterior of the incinerator vessel via an air to glycol heat reclaim circuit, recovering heat from the scrubber water prior to the ash lagoon or ash dewatering with a water to water heat exchanger, and recovering heat from the residual temperature of the flue gas.

Coincineration with Municipal Solid Waste

Coincineration is the process of incinerating wastewater sludges with municipal solid wastes. Coincinerator types can vary but many consist of reciprocating grates that are designed for more of a solid fuel than the multiple heath and fluidized-bed incinerators discussed in the previous section. The major objective of coincineration is to reduce the combined costs of incinerating sludge and solid wastes. Coincineration is not practiced widely. The process has the advantages of producing the heat energy necessary to evaporate water from sludges, supporting combustion of solid wastes and sludge, and providing an excess of heat for steam generation, if desired, without the use of auxiliary fossil fuels. In properly designed systems, the hot gases from the process can be used to remove moisture from sludges to a content of 10 to 15 percent. Direct feeding of sludge cake containing 70 to 80 percent moisture over solid wastes on traveling or reciprocating grates has been found to be ineffective. For systems operating without heat recovery, a disposal ratio of 1 kg (2.2 lb) of dry wastewater sludge to 4.6 kg (11 lb) of solid wastes is fired in normal operation. In the case of the water-walled boiler with heat recovery, the ratio is approximately 1 kg of dry (industrial plant) sludge to 7 kg (17 lb) of solid wastes. Based on past experience in municipal solid-waste disposal, the application of coincineration will likely continue to proceed very slowly, despite the advantages to the community in combining the two waste-disposal functions.

Air-Pollution Control

Incineration methods for wastewater sludge have the potential to be significant contributors to air pollution. Air contaminants associated with incineration can be divided into two categories: (1) odors and (2) combustion emissions.

Odors from Incineration. Odors are particularly offensive to the human senses, and special attention is required to minimize nuisance odor emissions. Odors historically associated with multiple-hearth incinerators resulted from units with rudimentary air pollution controls and low incinerator exhaust exit temperatures. Today, many units are required to run at higher exhaust temperatures to meet stringent air emission requirements, and by doing so, nuisance odors from the incineration processes themselves are nonexistent. Odors may occur from the ancillary processes for both multiple-hearth and fluid bed incinerators. Dewatering, cake storage, and conveyance all emit odors that require proper capture, treatment, and discharge. Incinerators can generally use the foul air as combustion air in the ATO process resulting in cost effective odor elimination.

Emissions from Incineration. Combustion emissions vary depending upon the type of thermal reduction technology employed and the nature of the sludge and auxiliary fuel used in the combustion process. Combustion emissions of particular concern are the criteria air pollutants (i.e., particulate matter, carbon monoxide, oxides of nitrogen, and sulfur dioxide), acid gases (mainly hydrogen chloride), and hazardous air pollutants (mercury, cadmium, lead, dioxins and furans). Under Part 503 regulations in the United States, heavy metals (arsenic, beryllium, chromium, and nickel), and total hydrocarbons (or carbon monoxide) are also a concern for regulators. Regulations promulgated by the U.S. EPA in 2011 concerning incinerator emissions are discussed in Sec. 13–2 and are generally more strict than the rules in the European Union.

Generally, emission controls fall into three main categories. These are (1) source control, (2) combustion control, and (3) air pollution control equipment. For certain pollutants, namely the heavy metals, source control is one of the most cost effective methods of reducing air pollution from the incinerator stack.

Source Control. Source control programs and collection systems use bylaw enforcement, and directed control programs (such as dental amalgam separators for mercury reduction) are effective at reducing loadings into the downstream collection system, which eventually ends up in the biosolids.

Combustion Control. Nitrogen oxides, carbon monoxide, total hydrocarbons, certain volatile organic compounds, dioxin, and furans are all air pollutants that are formed during the ATO process and combustion controls are effective at reducing levels of these contaminants in the flue gases. Most of these pollutants will be decreased to acceptable levels with elevated temperatures in the incinerator exhaust (generally greater than 750°C) and sufficient residence time (greater than 1 s). However, increased levels of nitrogen oxides may form with highly elevated temperatures, so a careful balance must be maintained in the specific details of the incinerator design and operation.

Air Pollution Control. For both multiple-hearth and fluid bed incinerators, the typical (historically most common, as least in North America) type of air pollution control equipment has been discussed previously. However, advances in the level of air pollution control equipment are being driven by stricter legislation. In general because fluid bed ATO units produce cleaner emissions than multiple-hearths, the industry is moving towards fluid bed technology for new systems. Air pollution control equipment is used as the final step to reduce the remaining pollutant levels in the flue gases. Although the majority of systems in North America have historically been wet scrubbers, there are also "dry" systems which are more common in Europe and Asia and more closely resemble the equipment utilized on solid waste incinerators. For primarily fluid bed incinerators, ammonia or urea injection may be used for selective non-catalytic reduction of nitrogen oxides. This treatment is done early while the flue gases are hot as they exit from the incinerator.

Advanced wet systems are now being used which contain multiple throat venturi scrubbers for particulate and heavy metal removal, caustic wet scrubbing for residual acid gas and sulfur dioxide removal, wet electrostatic precipitation for enhanced fine particulate and metal removal and polishing. A packed bed of sulfur impregnated media is sometimes used for the removal of mercury. Unfortunately, the mercury removal process requires a significant quantity of water, which can result in large quantities of wastewater. The generation of wastewater is usually not a problem for biosolids incinerators, as they are most commonly co-located with a wastewater treatment plant. Wet ash handling is used similar to the traditional wet venturi scrubber system.

Dry systems typically employ some method of flue gas cooling (whether this is part of a waste heat recovery boiler or a quench tower) to temper the gases before proceeding to the downstream equipment. Powdered activated carbon is injecting into the flue gas stream for heavy metal removal, and dioxin and furan control. Lime may also be injected for sulfur dioxide and acid gas control. A bag house, employing filter bags, is used to filter the flue gases. For this type of system, the ash is dry and is sent to silos for truck loading and disposal.

Hybrid systems may be employed which incorporate components of wet and dry air pollution control trains depending on the requirements of the level of pollution control, the site constraints, and the availability of water.

14–5 COMPOSTING

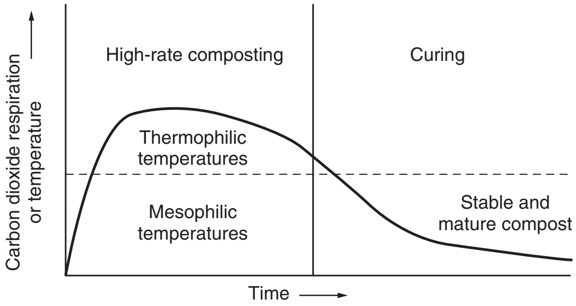
Composting is a process in which organic material undergoes biological degradation to a stable end product. Sludge that has been composted properly is a nuisance-free, humus-like material. Approximately 20 to 30 percent of the volatile solids are converted to carbon dioxide and water. As the organic material in the sludge decomposes, the compost heats to temperatures in the pasteurization range of 50 to 70°C (120 to 160°F), and enteric pathogenic organisms are destroyed. Properly composted biosolids may be used as soil conditioners in agricultural or horticultural applications, subject to any limitations based on the constituents in the composed biosolids (WEF, 2010). Composting is a cost-effective and environmentally sound alternate for the stabilization of wastewater sludges. Composting can be practiced for dewatered sludges or dewatered digested biosolids. However, composting biosolids is preferred, due to the odor issues during the composting process itself and the quality of the compost product.

Although composting may be accomplished under anaerobic or aerobic conditions, essentially all municipal wastewater biosolids composting applications are under mostly aerobic conditions (composting is never completely aerobic). Aerobic composting accelerates material decomposition and results in the higher rise in temperature necessary for pathogen destruction. Aerobic composting also minimizes the potential for nuisance odors.

The anticipated daily production of biosolids from a wastewater-treatment facility will have a pronounced effect on the alternate composting systems available for use, as will the availability of land for the construction of the composting facility. Other factors affecting

Figure 14–26

Stages during composting as related to carbon dioxide respiration and temperature (Epstein, 1997.).



the type of composting system are the nature of the biosolids produced; stabilization, if any, of the biosolids prior to composting; the type of dewatering equipment and chemicals used. Biosolids that are stabilized by aerobic or anaerobic digestion prior to composting may result in reducing the size of the composting facilities by up to 40 percent.

Process Microbiology

The composting process involves the complex destruction of organic material coupled with the production of humic acid to produce a stabilized end product. The microorganisms involved fall into three major categories: bacteria, actinomycetes, and fungi. Although the interrelationship of these microbial populations is not fully understood, bacterial activity appears to be responsible for the decomposition of proteins, lipids, and fats at thermophilic temperatures, as well as for much of the heat energy produced. Fungi and actinomycetes are also present at varying levels during the mesophilic and thermophilic stages of composting and appear to be responsible for the destruction of complex organics and the cellulose supplied in the form of amendments or bulking agents.

Composting Process Stages

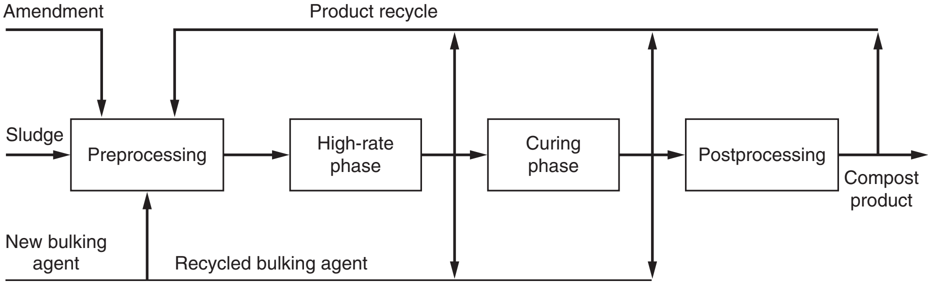
During the composting process, three separate stages of activity and associated temperatures are observed: mesophilic, thermophilic, and curing (cooling) (see Fig. 14–26). In the initial mesophilic stage, the temperature in the compost pile increases from ambient to approximately 40°C (104°F) with the appearance of fungi and acid-producing bacteria. As the temperature in the composting mass increases to the thermophilic range of 40 to 70°C (104 to 160°F), these microorganisms are replaced by thermophilic bacteria, actinomycetes, and thermophilic fungi. It is in the thermophilic temperature range that the maximum degradation and stabilization of organic material occur. The curing stage is characterized by a reduction in microbial activity, and replacement of the thermophilic organisms with mesophilic bacteria and fungi. During the curing stage, further evaporative release of water from the composted material will occur, as well as stabilization of pH and completion of humic acid formation.

Composting Process Steps

Most composting operations consist of the following basic steps (see Fig. 14–27): (1) preprocessing, the mixing of dewatered sludge with an amendment and/or a bulking agent; (2) high-rate decomposition, aerating the compost pile either by the addition of air, by mechanical turning, or by both; (3) recovery of the bulking agent (at the end of either the high-rate decomposition or curing phase, if practical); (4) further curing and storage, which allows further stabilization and cooling of the compost; (5) postprocessing,

Figure 14–27

Generalized process diagram for composting showing inputs of sludge (feed substrate), amendments, and bulking agents (Haug, 1993.).



screening for the removal of nonbiogradable material such as metals and plastics or grinding for size reduction, and (6) final disposition. A portion of the final product is sometimes recycled to the preprocessing step to aid in conditioning the compost mixture.

Feed Stock Amendments. An amendment is an organic material added to the feed substrate primarily to, reduce moisture content, and increase the air voids for proper aeration. Amendments can also be used to increase the quantity of degradable organics in the mixture. Commonly used amendments are sawdust, straw, recycled compost, and rice hulls. A bulking agent is an organic or inorganic material that is used to provide structural support and to increase the porosity of the mixture for effective aeration. Wood chips are the most commonly used bulking agents and can be recovered and reused. The characteristics of bulking agents used most commonly are reported in Table 14–18. An amendment

Table 14–18

Characteristics of compost bulking agents used in the aerobic composting of sludge from wastewater treatment[a]

Bulking agent	Comments
Wood chips	May have to be purchased High recovery rate by screening Provides supplemental carbon source
Chipped brush	Possibly available as a waste material Low recovery rate by screening Provides supplemental carbon source Longer curing time of compost
Leaves and yard waste	Must be shredded Wide range of moisture content Readily available source of carbon Relatively low porosity Nonrecoverable
Shredded tires	Often mixed with other bulking agents Supplemental carbon is not available Nearly 100 percent recoverable May contain metals
Ground waste lumber	Possibly available as a waste material Often a poor source of supplemental carbon

[a] Adapted in part from WEF (2010).

can also provide supplemental carbon to the mix and it is typically recommended to maintain a carbon to nitrogen ratio of 25:1 to 35:1. Insufficient carbon can lead to nitrogen loss as ammonia which can cause odor issues and reduce the nutrient content of the resulting compost (WEF, 2010).

High-Rate Decomposition. The high-rate decomposition stage of composting has been more engineered and controlled due to the need to reduce odors, supply high aeration rates, and maintain process control. The curing stage is often less engineered, less controlled, and given only small consideration in some designs, however, the curing stage is an integral part of the system design and operation, and both stages need to be designed and operated properly to produce a mature compost product.

Recovery of the Bulking Agent. If practicable, the bulking agent should be recovered at the end of either the high-rate decomposition or curing phase. The most common method for recovery of bulking agent is screening with mesh size based on the physical size of the bulking agent.

Postprocessing. Postprocessing is often used to prepare the finished compost for marketing. Preparation includes conveying the finished compost from the active composting area to the curing, screening, and preparation areas. Trommel screens and belt shredders are used frequently; shredding can precede or follow curing. In some cases, double screening is preferable, especially for the horticultural market to meet product quality requirements. Particle size of the finished product for general use ranges typically from 6 to 25 mm (1/4 to 1 in.).

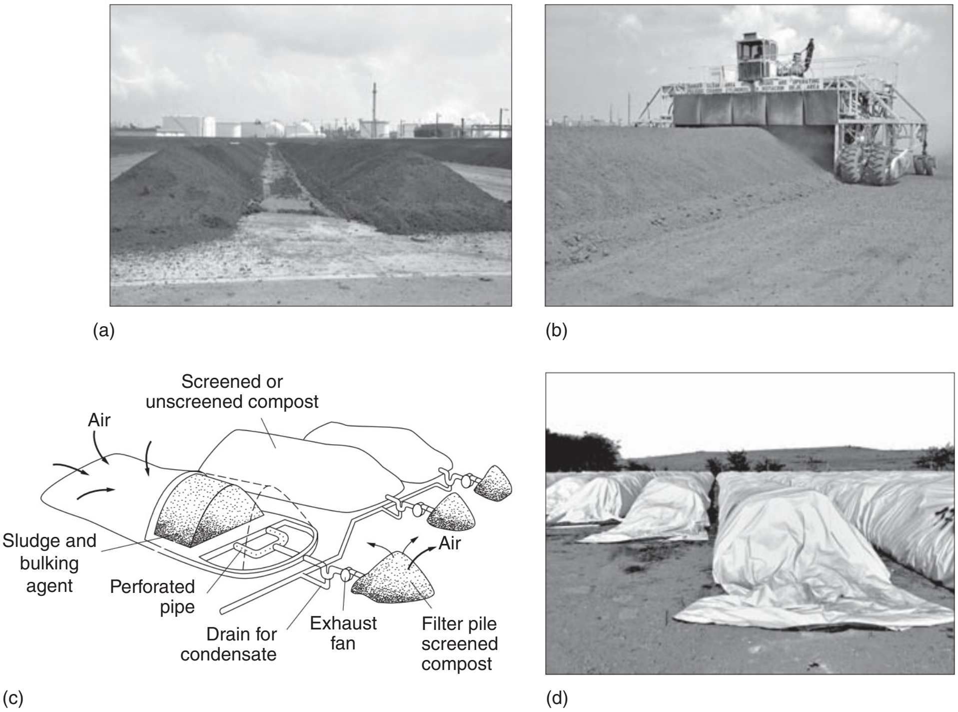
Composting Methods

The two principal methods of composting now in use in the United States may be classified as agitated or static. In the agitated method the material to be composted is agitated periodically to introduce oxygen, to control the temperature, and to mix the material to obtain a uniform product. In the static method, the material to be composted remains static and air is blown through the composting material. The most common agitated and static methods of composting are known as the windrow and static pile methods, respectively. Proprietary composting systems in which the composting operation is carried out in a reactor of some type are known as in-vessel composting systems.

Windrow. In a windrow system, a mixture of dewatered sludge and bulking agent is placed in windrows, which are typically from 1 to 2 m (3 to 6 ft) high and 2 to 4.5 m (6 to 14 ft) at the base [see Fig. 14–28(a)]. The windrows are turned and mixed periodically during the composting period using specialized equipment [see Fig. 14–28(b)]. Supplemental mechanical aeration is used in some applications. The composting period is about 21 to 28 d. Under typical operating conditions, the windrows are turned a minimum of five times while the temperature is maintained at or above 55°C. In windrow composting, aerobic conditions are difficult to maintain throughout the cross-sectional area of the windrow. Thus, the microbial activity within the pile may be aerobic, facultative, anaerobic, or various combinations thereof, depending on when and how often the pile is turned. Turning of the windrows is often accompanied by the release of

Figure 14–28

Composting systems: (a) view of compost windrows, (b) view of equipment for turning and grading windrows, (c) schematic of aerated static pile compost process, and (d) view of force-aerated static pile in compost bags for odor control.



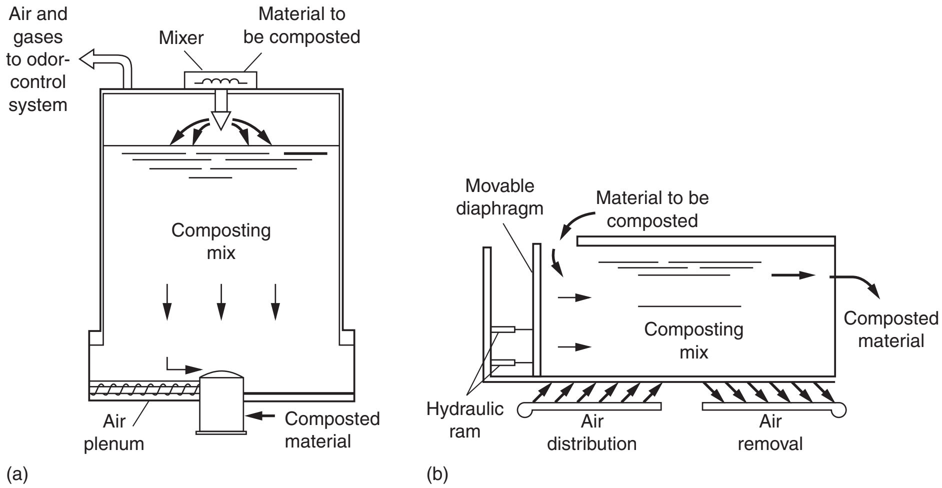
offensive odors. The release of odors occurs typically when anaerobic conditions develop within the windrow. Some windrow operations are covered or enclosed, similar to aerated static piles.

Aerated Static Pile. The aerated static pile system consists of a grid of aeration or exhaust piping over which a mixture of dewatered sludge and bulking agent is placed [see Fig. 14–28(c)]. An alternative version of the static pile composting method is carried out in an enclosed plastic bag with forced aeration [see Fig. 14–28(d)]. In a typical static pile system, the bulking agent consists of wood chips, which are mixed with the dewatered sludge by a pug-mill type or rotating-drum mixer or by movable equipment such as a front-end loader. Material is composted for 21 to 28 d and is typically followed by a curing period of 30 d or longer. Typical pile heights are generally about 2 to 2.5 m (6 to 8 ft). A layer of screened compost is often placed on top of the pile for insulation. Disposable corrugated plastic drainage pipe is commonly used for air supply and each individual pile is recommended to have an individual blower for more effective aeration control. Screening of the cured compost usually is done to reduce the quantity of the end product requiring ultimate disposal and to recover the bulking agent. For improved process and odor control, many facilities cover or enclose all or significant portions of the system.

In-Vessel Composting Systems. In-vessel composting is accomplished inside an enclosed container or vessel. Mechanical systems are designed to minimize odors and process time by controlling environmental conditions such as air flow, temperature, and oxygen concentration. The advantages of in-vessel composting systems are better process and odor control, faster throughput, lower labor costs, and smaller area requirements.

Figure 14–29

Plug flow in-vessel composting reactors: (a) unmixed vertical plug flow reactor and (b) unmixed tunnel (horizontal plug flow) reactor.



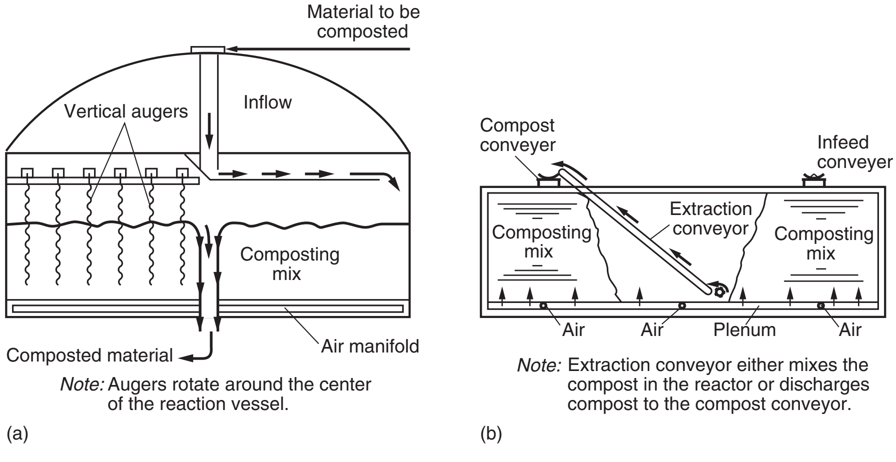
In-vessel composting systems can be divided into two major categories: plug flow and dynamic (agitated bed). In plug-flow systems, the relationship between particles in the composting mass stays the same throughout the process, and the system operates on the basis of a first-in, first-out principle. In a dynamic system, the composting material is mechanically mixed during the processing. In-vessel systems can be further categorized based on the geometric shape of the vessels or containers used. Examples of plug-flow reactors are shown on Fig. 14–29 and examples of dynamic-type systems are illustrated on Fig. 14–30.

Design Considerations

A number of factors, each of which must be evaluated to meet the specific requirements, must be considered in the design of a composting system (see Table 14–19). A design approach using a materials balance is particularly useful because the amount of each component (sludge or biosolids, bulking agent, and amendment) used during each phase of the process is determined. In a materials balance, the following parameters must be measured or calculated for each component: (1) total volume, (2) total wet weight, (3) total solids content (dry weight), (4) volatile solids content (dry weight), (5) water content

Figure 14–30

Dynamic (mixed) in-vessel composting units: (a) vertical reactor and (b) horizontal reactor.



Note: Augers rotate around the center of the reaction vessel.

Note: Extraction conveyor either mixes the compost in the reactor or discharges compost to the compost conveyor.

Table 14–19

Design considerations for aerobic sludge composting processes[a]

Item	Comment
Type of sludge	Both sludge and biosolids can be composted successfully. Sludge has a greater potential for odors, particularly for windrow systems. Sludge, as compared to biosolids, has more energy available, will degrade more readily, and has a higher oxygen demand.
Amendments and bulking agents	Amendment and bulking agent characteristics, such as moisture content, particle size, and available carbon, affect the process and quality of product. Bulking agents should be readily available.
Carbon-nitrogen ratio	The initial C:N ratio should be in the range of 25:1 to 35:1 by weight. At lower ratios, ammonia is given off. Carbon should be checked to ensure it is readily biodegradable.
Volatile solids	The volatile solids of the composting mix should be greater than 30 percent of the total solids content. Dewatered sludge will usually require an amendment or bulking agent to adjust the solids content.
Air requirements	Air with at least 50 percent of the oxygen remaining should reach all parts of the composting material for optimum results, especially in mechanical systems.
Moisture content	Moisture content of the composting mixture should be not greater than 40 percent for static pile, windrow, and in-vessel composting.
pH control	The pH of the composting mixture should generally be in the range of 6 to 9. To achieve optimum aerobic decomposition, pH should remain at 7 to 7.5 range.
Temperature	For best results, temperature should be maintained between 50 and 55°C for the first few days and between 55 and 60°C for the remainder of the active composting period. If the temperature is allowed to increase beyond 65°C for a significant period of time, biological activity will be reduced.
Control of pathogens	If properly conducted, it is possible to kill all pathogens, weeds, and seed during the composting process. To achieve this level of control, the temperature must be maintained at levels required by the EPA 503 regulations for Class A (see Table 13–11).
Mixing and turning	To prevent drying, caking, and air channeling, material in the process of being composted should be mixed or turned on a regular schedule or as required. Frequency of mixing or turning will depend on the type of composting operation.
Heavy metals and trace organics	Heavy metals and trace organics in the sludge and finished compost should be monitored to ensure that the concentrations do not exceed the applicable regulations for end use of the product.
Site constraints	Factors to be considered in selecting a site include available area, access, proximity to treatment plant and other land uses, climatic conditions, and availability of buffer zone.

[a] Adapted in part from Tchobanoglous, et al. (1993).

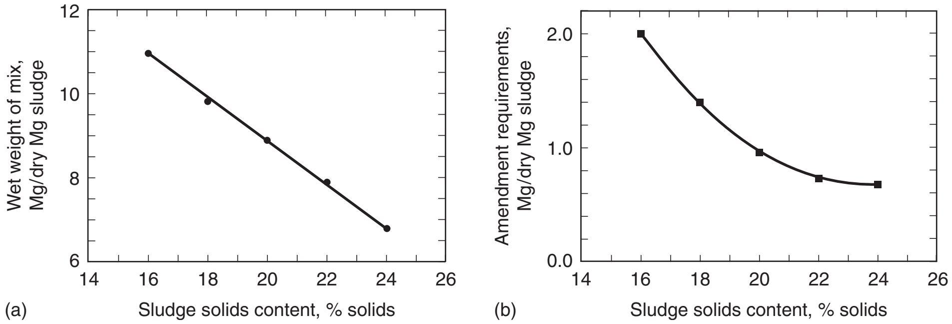
(weight), (6) bulk density (wet weight/unit volume), (7) percent water content, and (8) percent volatile solids of the compost mix.

An important output of the materials balance is to determine the composition of the compost mix. The compost mix should be about 40 percent dry solids to ensure adequate composting in windrow and static-pile composting. In-vessel systems require similar solids requirements, but slightly lower values may be used, depending on the aeration system. Starting the compost process with solids higher than 40 percent soon results in a “dry mix.” A dry mix is dusty, and sufficient biological activity and temperature levels are difficult to maintain. The addition of water will likely be required for the duration of the process; thus, provisions for a water supply should be made.

The effect of moisture content in the dewatered sludge on the compost mix is illustrated on Fig. 14–31. The moisture content of the sludge affects the wet weight of the mixture and the amount of amendment that has to be used. Using Fig. 14–31 (a), for

Figure 14–31

Effect of sludge solids content on compost mix and amendment quantities: (a) mix quantities versus sludge solids content, (b) amendment requirements versus sludge solids content. (U.S. EPA, 1989.)



example, if the sludge cake contains 24 percent solids, the wet weight of the mix is about 6.7 Mg (tonne) per dry Mg of sludge. If the sludge solids content decreases to 16 percent, the wet weight increases to about 11 Mg per dry Mg of sludge. The additional moisture content would require larger materials-handling systems and larger reactors. The amendment requirements, as indicated on Fig. 14–31(b), would triple over the same range of sludge solids. In compost-system design, the type of sludge-dewatering system and the consistency of resulting product have to be evaluated carefully.

Co-composting with Municipal Solid Wastes

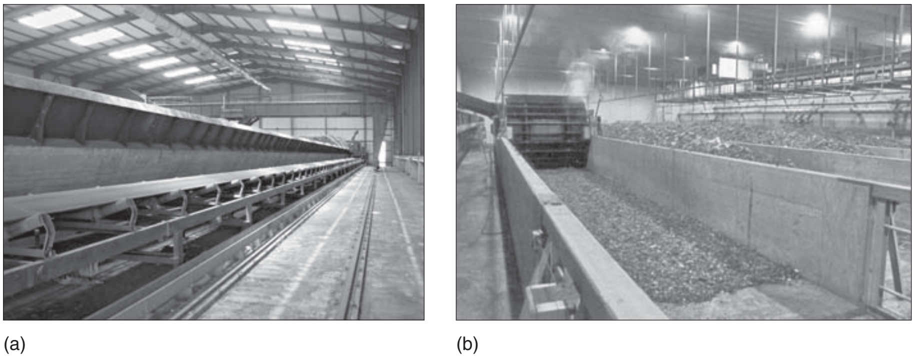
Co-composting of sludge and municipal solid wastes is a possible alternative where integrated waste-disposal facilities are considered. Mixing the sludge with the organic fraction of municipal solid waste or source separated yard wastes is beneficial because (1) sludge dewatering may not be required, and (2) the overall metals content of the composted material will be less than that of the composted sludge alone. Liquid treatment plant sludges typically have a solids content ranging from 3 to 8 percent. A 2-to-1 mixture of compostable municipal solid or yard wastes to sludge is recommended as a minimum. Both static and agitated compost systems have been tried (Tchobanoglous et al., 1993). Two examples of composting facilities in reactors with mechanical mixing, located within a building for odor control, are shown on Fig. 14–32.

Public Health and Environmental Issues

The principal public health and environmental issues concerning compost operations relate to exposure to pathogens and bioaerosols. Exposure to pathogens can occur during the composting process or through the use of the product if the composting process is not executed properly and the resulting product is not disinfected. The potential modes of infection for workers are (1) inhalation of aerosols containing airborne microorganisms,

Figure 14–32

Examples of compost reactors with mechanical mixing. The reactors are in enclosed buildings for odor control.


(a) (b)

(2) dermal contact, or (3) oral through inadvertent contact of dust or contaminated food or through hand-to-mouth contact such as cigarette smoking. Ingestion of contaminated product or contamination by the product to cigarettes or food is the greatest potential source of pathogen invasion by workers or users (Epstein, 1997).

Compost bioaerosols are organisms or biological agents that can be dispersed through the air and affect human health. Bioaerosols can contain living organisms including bacteria, fungi, actinomycetes, arthropods, protozoa and microbial products such as endotoxin and microbial enzymes. During composting, bioaerosols are not only present in waste materials but also can be generated during the process. The level and type of bioaerosols are a function of feedstock. The two bioaerosols of greatest interest to worker health and the environment surrounding composting facilities are *Aspergillus fumigatus* and endotoxin.

A. fumigatus, a common fungus, is of concern to both worker health and populations surrounding composting facilities, as it can cause lung disease. Endotoxin, part of the cell wall of gram-negative bacteria that is released to the environment during the composting process, is of primary concern to workers in composting, recycling, and other solid-waste processing facilities. There is little evidence that exposure to airborne endotoxin causes toxic conditions. Most of the data, however, concerning worker illness is associated with composting municipal solid waste, principally in Europe. Proper ventilation, dust control, and use of dust masks reduce worker exposure to bioaerosols (Epstein, 1997).

14–6 SLUDGE AND BIOSOLIDS CONVEYANCE AND STORAGE

The sludges from primary and biological treatment processes are concentrated and stabilized by mechanical, biological, and thermal means and are reduced in volume in preparation for final disposal. Because the methods of conveyance and final disposal often determine the type of stabilization required and the amount of volume reduction that is needed, they are considered briefly in the following discussion.

Conveyance Methods

Biosolids may be transported long distances by (1) pipeline, (2) truck, (3) barge, (4) rail, or any combination of these four modes. Truck transportation, however, is the most common method used currently [see Fig. 14–33(a)]. To minimize the danger of spills, odors, and dissemination of pathogens to the air, liquid biosolids should be transported in closed vessels, such as tank trucks, railroad tank cars, or covered tank barges. Stabilized, dewatered biosolids can be transferred in open vessels, such as dump trucks, or in railroad gondolas. If biosolids are hauled long distances, the vessels should be covered. The method of transportation chosen and its costs are dependent on a number of factors,

Figure 14–33

Views of sludge transport vehicle and storage facilities: (a) typical side dump sludge transport vehicle, and (b) large bins used for the temporary storage for lime-stabilized sludge. Overhead bucket is used to distribute sludge in bins and to transport to trucks.

(a) (b)

including (1) the nature, consistency, and quantity of biosolids to be transported; (2) the distance from origin to destination; (3) the availability and proximity of the transit modes to both origin and destination; (4) the degree of flexibility required in the transportation method chosen; and (5) the estimated useful life of the ultimate disposal facility.

Each transportation method contributes a minor air pollutant load, either directly or indirectly. A certain amount of air pollution is produced from the facility that generates electricity necessary for sludge pumping. The engines that move trucks, barges, and railroad cars also produce some air pollutants. On a mass (tonnage) basis, the transportation mode that contributes the lowest pollutant load is piping. Next, in sequence, are barging and unit train rail transportation. The highest pollutant load is from trucking. Other factors of environmental concern include traffic, noise, and construction disturbance.

Storage

It is often necessary to store biosolids that have been digested anaerobically before they are disposed of or used beneficially. Although sludge can be stored for short periods of time in clarifiers, biological systems, digesters or holding tanks, the available storage capacity may not be sufficient for long term storage. Storage of liquid biosolids can be accomplished in storage basins and lagoons, and storage of dewatered biosolids can be done on storage pads. An example of large temporary sludge storage bins for lime-stabilized biosolids is shown on Fig.14–33 (b). Drying beds, discussed previously, can also be used for storage.

Storage Basins and Lagoons. Biosolids stored in basins become more concentrated and are further stabilized by continued anaerobic biological activity. Long-term storage is effective in pathogen destruction. Depth of the biosolids storage basins may vary from 3 to 5 m (10 to 16 ft). Solids loading rates vary from about 0.1 to 0.25 kg VSS/m^2·d (20 to 50 lb VSS/10^3 ft^2·d) of surface area. If the basins are not loaded too heavily ($\leq$ 0.1 kg VSS/m^2·d), it is possible to maintain an aerobic surface layer through the growth of algae and by atmospheric reaeration. Alternatively, surface aerators can be used to maintain aerobic conditions in the upper layers. The number of basins to be used should be sufficient to allow each basin to be out of service for a period of about 6 months. Stabilized and thickened biosolids can be removed from the basins using a mud pump mounted on a floating platform or by mobile crane using a drag line. Biosolids concentrations as high as 35 percent solids have been achieved in the bottom layers of these basins.

Long-term storage of sludge and biosolids in lagoons is simple and economical if the treatment plant is in a remote location. A lagoon is an earthen basin into which untreated sludge, or digested biosolids are deposited. In lagoons with untreated sludge, the organic matter is stabilized by anaerobic and aerobic decomposition, which may give rise to objectionable odors. The stabilized biosolids settle to the bottom of the lagoon and accumulate, and excess liquid from the lagoon, if there is any, is returned to the plant for treatment. Lagoons should be located away from highways and dwellings to minimize possible nuisance conditions and should be fenced to keep out unauthorized persons. Lagoons should be relatively shallow, 1.25 to 1.5 m (4 to 5 ft), if they are to be cleaned by scraping. If the lagoon is used only for digested biosolids, the nuisances mentioned should not be a problem. If subsurface drainage and percolation are potential problems, the lagoon should be lined. Solids may be stored indefinitely in a lagoon and may be removed periodically after draining and drying.

Storage Pads. Where dewatered biosolids have to be stored prior to land application, sufficient storage area should be provided based on the number of consecutive days that biosolids hauling could occur without applying biosolids to land. Allowances also have to be made for paved access and for area to maneuver the biosolids hauling trucks, loaders, and application vehicles. The storage pads should be constructed of concrete and designed to withstand the truck loadings and biosolids piles. Provisions for leachate and stormwater collection and disposal also have to be included in the design of sludge storage pads.

14–7 SOLIDS MASS BALANCES

Sludge and biosolids processing facilities, such as thickening, digestion, and dewatering, produce waste streams that must be recycled to the treatment process or to treatment facilities designed specifically for the purpose. The recycled flows impose an incremental solids, hydraulic, organic, and nutrient load on the wastewater-treatment facilities that must be considered in the plant design. When the flows are recycled to the treatment process, they should be directed to the head of the plant and blended with the plant flow following preliminary treatment. Equalization facilities can be provided for the recycled flows so that their reintroduction into the plant flow will not cause a shock loading on the subsequent treatment processes. To predict the incremental loads imposed by the recycled flows, it is necessary to perform a materials mass balance for the treatment system.

Preparation of Solids Mass Balances

Typically, a materials mass balance is computed on the basis of average flow, average BOD and total suspended solids concentrations. To size certain facilities properly, such as sludge storage tanks and plant piping, it is also important to perform a materials mass balance for the maximum expected concentration of BOD and TSS in the untreated wastewater. However, the maximum concentrations will not usually result in a proportional increase in the recycled BOD and TSS. The principal reason is that the storage capacity in the wastewater and sludge-handling facilities tends to dampen peak solids loads. For example, for a maximum TSS load equal to twice the average value, the resulting peak solids loading to a dewatering unit may be only 1.5 times the average loading. Further, it has been shown that periods of maximum hydraulic loading typically do not correlate with periods of maximum BOD and TSS. Therefore, coincident maximum hydraulic loadings should not be used in the preparation of a materials mass balance for maximum organic loadings (see Chap. 5). The preparation of a mass balance is illustrated in Example 14–3.

Performance Data for Solids Processing Facilities

To prepare a materials mass balance, it is necessary to have information on the operational performance and efficiency of the various unit operations and processes that are used for the processing of waste sludge and biosolids. Representative data on the solids capture and expected solids concentrations for the most commonly used operations are reported in Tables 14–20 and 14–21. These data were derived from an analysis of the records from a number of installations throughout the United States. The wide variation that can occur in the reported values is apparent; thus, the values in Tables 14–20 and 14–21 should be used only if no other information is available. Wherever possible, local conditions and data should be used in performing the mass balance.

Impact of Return Flows and Loads

In addition to performance data for expected solids capture and constituent concentrations for the various process components, data for the expected concentrations of BOD and TSS in the return flows must also be included in preparing of mass balances. If the quantities

Table 14–20

Typical solids concentration and capture values for various sludge and biosolids processing methods

	Solids concentration, %		Solids capture %	
Operation	**Range**	**Typical**	**Range**	**Typical**
Gravity thickeners:				
Primary sludge only	3–10	5	85–92	90
Primary and waste activated	2–6	3.5	80–90	85
Flotation thickeners:				
With chemicals	4–6	5	90–98	95
Without chemicals	3–5	4	80–95	90
Centrifuge thickeners:				
With chemicals	4–8	5	90–98	95
Without chemicals	3–6	4	80–90	85
Belt-filter press:				
With chemicals	15–30	22	85–98	95
Filter press:				
With chemicals	20–50	36	90–98	95
Centrifuge dewatering:				
With chemicals	10–35	25	85–98	95

Table 14–21

Typical BOD and total suspended-solids (TSS) concentrations in the recycle flows from various sludge processes[a]

	BOD, mg/L		Suspended Solids, mg/L	
Operation	**Range**	**Typical**	**Range**	**Typical**
Gravity thickening supernatant:				
Primary sludge only	100–400	250	80–300	200
Primary + waste activated sludge	60–400	300	100–350	250
Flotation thickening subnatant	50–1200	250	100–2500	300
Centrifuge thickening centrate	170–3000	1000	500–3000	1000
Aerobic digestion supernatant	100–1700	500	100–10,000	3400
Anaerobic digestion (two-stage, high rate) supernatant	500–5000	1000	1000–11,500	4500
Centrifuge dewatering centrate	100–2000	1000	200–20,000	5000
Belt-filter press filtrate	50–500	300	100–2000	1000
Recessed-plate-filter press filtrate	50–250		50–1000	
Sludge lagoon supernatant	100–200		5–200	
Sludge drying bed underdrainage	20–500		20–500	

(*continued*)

Table 14–21 (*Continued*)

Operation	BOD, mg/L Range	BOD, mg/L Typical	Suspended Solids, mg/L Range	Suspended Solids, mg/L Typical
Composting leachate		2000		500
Incinerator scrubber water	20–60		600–8000	
Depth filter washwater	50–500		100–1000	
Microscreen washwater	100–500		240–1000	
Carbon adsorber washwater	50–400		100–1000	

[a] Adapted, in part, from U.S. EPA (1987c) and WEF (2010).

and characteristics of recycled flows and loads are not accounted for properly, the facilities that receive them may be underdesigned significantly. The major impacts of return flows and measures that can mitigate these impacts are summarized in Table 14–22. Impact of returned flows on the overall treatment process is discussed in detail in Chap. 15.

Table 14–22

Major impacts and potential mitigation measures for return flows from sludge and biosolids-processing facilities[a]

Source of return flow	Impact	Process impacted	Mitigation measure
Sludge thickening	Effluent degradation by colloidal SS	Sedimentation	Add flocculent aid ahead of sedimentation tank
			Separately thicken primary and biological sludges
			Optimize gravity thickener dilution water
	Floating sludge	Sedimentation	Minimize gravity thickener detention time
			Remove sludge continuously and uniformly
	Odor release and septicity	Recycle point	Reduce gravity thickener detention time
			Return flows ahead of aerated grit chamber
			Provide odor containment, ventilation, and treatment (scrubber or biofilter)
		Biological	Return odorous flows to aeration tank
			Remove sludge continuously and uniformly
			Provide separate return flow treatment (with other recycle streams)
	Solids buildup	Sedimentation	Increase dewatering unit operation time
		Biological	Remove sludge continuously and uniformly
			Include recycle loads in mass balance analysis
Sludge dewatering	Effluent degradation by colloidal suspended solids	Sedimentation	Optimize dewatering units solids capture by improved sludge conditioning
			Add flocculent aid ahead of sedimentation tank
			Return centrate/filtrate to thickener
			Provide separate return flow treatment (with other recycle streams)
	Solids buildup	Sedimentation	Increase dewatering unit operation time

(*continued*)

Table 14–22 (*Continued*)

Source of return flow	Impact	Process impacted	Mitigation measure
		Biological	Remove sludge continuously and uniformly Reduce trickling-filter recycle rate Include recycle loads in mass balance analysis
Sludge stabilization	Effluent degradation by excessive BOD load	Biological	Optimize supernatant/decant removal, i.e., remove smaller amounts over a longer period of time, or reschedule removal to off-peak periods Provide separate return flow treatment Increase RBC speed Increase MLVSS in activated-sludge system (decrease F:M ratio) Increase dissolved oxygen level in activated-sludge process
	Effluent degradation by nutrients	Biological	Regulate digester supernatant/decant removal Thicken sludge before stabilization Provide separate return flow treatment
Washwater from depth filters	Hydraulic surges	Sedimentation	Provide backwash storage for flow equalization Schedule filter backwashing for off-peak periods

[a] Adapted, in part, from U.S. EPA (1987b).

EXAMPLE 14–3 **Preparation of a Solids Mass Balance for a Secondary Treatment Facility** Prepare a solids balance for the treatment flow diagram shown in the following figure, using an iterative computational procedure.

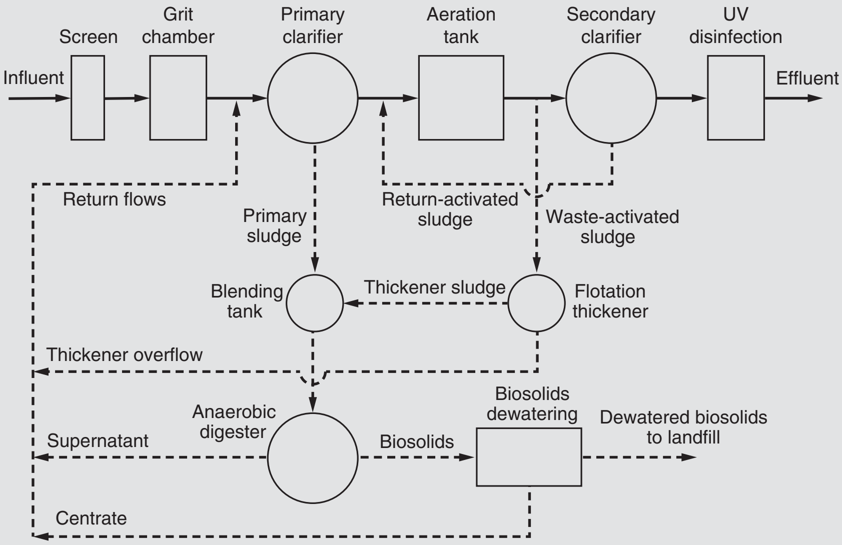
1. Definition of terms

   BOD_C = biochemical oxygen demand expressed as a concentration, g/m^3
   BOD_M = biochemical oxygen demand expressed as a mass, kg/d
   TSS_C = total suspended solids expressed as a concentration, g/m^3
   TSS_M = total suspended solids expressed as a mass, kg/d

   Assume for the purpose of this example that the following data apply:
2. Wastewater flowrates
   a. Average dry weather flowrate = 21,600 m^3/d
   b. Peak dry weather flowrate = 2.5(21,600 m^3/d) = 53,900 m^3/d
3. Influent characteristics
   a. $BOD_C = 375\ g/m^3$
   b. $TSS_C = 400\ g/m^3$ (assume VSS_c/TSS_c ratio in influent = 67%)
   c. TSS_C after grit removal = 360 g/m^3 (assume volatile fraction of the grit = 10%)
4. Sludge and biosolids characteristics
   a. Concentration of primary sludge = 6%
   b. Concentration of thickened waste-activated sludge = 4% (assume solids capture in the flotation thickeners = 90%)
   c. Total suspended solids in digested sludge = 5%
   d. For the purposes of this example, assume that the specific gravity of the solids from the primary sedimentation tank and the flotation thickener is equal to 1.0

   e. Fraction of the biological solids that are biodegradable = 65%
   f. The value of BOD$_C$ can be obtained by multiplying the value of UBOD by a factor of 0.68 (corresponds to a k value of 0.23 d^{-1} in the BOD equation, see Chap. 2)

5. Effluent characteristics
   a. BOD$_C$ = 20 g/m^3
   b. TSS$_C$ = 22 g/m^3
6. Primary clarifier
   a. Assumes performance of 33% BOD removal and 70% TSS removal
   b. Assume VSS$_C$/TSS$_C$ ratio in primary effluent going to the secondary treatment process = 85%
7. Secondary treatment
   a. Assume the mixed-liquor VSS$_C$/ TSS$_C$ ratio is 0.8
   b. Aeration tank volume, V_r = 4700 m^3
   c. Y = 0.5 kg/kg
   d. b = 0.06 d^{-1}
   e. SRT = 10 d
8. Sludge digestion
   a. Assume SRT = 20 d
   b. Assume VSR = 50%
   c. Assume digester gas production = 1.12 m^3/kg VSS destroyed
   d. Assume BOD$_C$ in digester supernatant = 1000 g/m^3
   e. Assume TSS$_C$ in digested sludge = 5%
9. Sludge dewatering
   a. Assume sludge cake = 22% solids
   b. Specific gravity of sludge = 1.06
   c. Solids capture = 93%
   d. Centrate BOD$_C$ = 2000 mg/L



Solution

1. Convert the given constituent quantities to daily mass values.
   a. BOD$_M$ in influent:

$$\begin{aligned} BOD_M &= (21{,}600\ m^3/d)(375\ g/m^3)/(10^3\ g/1\ kg)\\ &= 8100\ kg/d \end{aligned}$$

b. TSS_M in influent:

$$TSS_M = (21{,}600 \text{ m}^3/\text{d})(400 \text{ g/m}^3)/(10^3 \text{ g/1 kg}) = 8640 \text{ kg/d}$$

c. TSS_M after grit removal (influent to primary settling tanks):

$$TSS_M = (21{,}600 \text{ m}^3/\text{d})(360 \text{ g/m}^3)/(10^3 \text{ g/1 kg}) = 7776 \text{ kg/d}$$

2. Estimate the concentration of soluble BOD_C in the effluent using the following relationship:

   Effluent BOD_C = influent soluble BOD_C escaping treatment + BOD_C of effluent TSS_C

   a. Determine the BOD_C of the effluent TSS_C.
      - i. Biodegradable portion of effluent TSS_C is $0.65\ (22 \text{ g/m}^3) = 14.3 \text{ g/m}^3$
      - ii. UBOD of the biodegradable effluent TSS_C is $[0.65(22 \text{ g/m}^3)](1.42 \text{ g/g}) = 20.3 \text{ g/m}^3$
      - iii. BOD_C of effluent suspended solids $= 20.3 \text{ g/m}^3\ (0.68) = 13.8 \text{ g/m}^3$

   b. Solve for the influent soluble BOD_C escaping treatment.

      $20 \text{ g/m}^3 = S + 13.8 \text{ g/m}^3$

      $S = 6.2 \text{ g/m}^3$

3. Prepare the first iteration of the solids balance. (In the first iteration, the effluent wastewater total suspended solids and the biological solids generated in the process are distributed among the unit operations and processes that make up the treatment system.)

   a. Primary setting
      - i. Operating parameters:

        BOD_C removed = 33%

        TSS_C removed = 70% (see also Fig. 5–51)
      - ii. BOD_M removed $= 0.33(8100 \text{ kg/d}) = 2700 \text{ kg/d}$
      - iii. BOD_M to secondary $= (8100 - 2700) \text{ kg/d} = 5400 \text{ kg/d}$
      - iv. TSS_M removed $= 0.7(7776 \text{ kg/d}) = 5443 \text{ kg/d}$
      - v. TSS_M to secondary $= (7776 - 5443) \text{ kg/d} = 2333 \text{ kg/d}$

   b. Determine the volatile fraction of primary sludge.
      - i. Operating parameters:

        Volatile fraction of TSS_C in influent = 67%

        Volatile fraction of grit =10%

        Volatile fraction of incoming TSS_C discharged to the secondary process = 85%
      - ii. Volatile suspended solids (VSS_M) in influent prior to grit removal $= 0.67\ (8640 \text{ kg/d}) = 5789 \text{ kg/d}$
      - iii. VSS_M removed in grit chamber $= 0.10(8640 - 7776) \text{ kg/d} = 86 \text{ kg/d}$
      - iv. VSS_M in secondary influent, kg/d $= 0.85(2333 \text{ kg/d}) = 1983 \text{ kg/d}$
      - v. VSS_M in primary sludge, kg/d $= (5789 - 86 - 1983) \text{ kg/d} = 3710 \text{ kg/d}$
      - vi. Volatile fraction in primary sludge $= [(3710 \text{ kg/d})/(5443 \text{ kg/d})](100) = 68.2\%$

   c. Secondary process
      - i. Determine the secondary process operating parameters

$$\text{Mixed liquor } VSS_C = \frac{(Q)(Y)(S_o - S)\text{SRT}}{[1 + b(\text{SRT})]\ (V_r)}$$

$$= \frac{(21{,}600\ \text{m}^3/\text{d})(0.5)[(250 - 6.2)\ \text{g/m}^3](10\text{d})}{[1 + (0.06\,\text{d}^{-1})(10\,\text{d})](4700\ \text{m}^3)}$$

$$= 3500\ \text{g/m}^3$$

$$\text{Mixed Liquor TSS}_C = \frac{\text{VSS}_c}{0.8} = \frac{(3500\ \text{g/m}^3)}{0.8} = 4375\ \text{g/m}^3$$

$$Y_{\text{obs}} = \frac{Y}{1 + b(\text{SRT})} = \frac{0.5}{1 + 0.06 \times 10} = 0.3125$$

ii. Determine the effluent mass quantities.

$\text{BOD}_M = (21{,}600\ \text{m}^3/\text{d})(20\ \text{g/m}^3)/(10^3\ \text{g/1 kg}) = 432\ \text{kg/d}$
$\text{TSS}_M = (21{,}600\ \text{m}^3/\text{d})(22\ \text{g/m}^3)/(10^3\ \text{g/1 kg}) = 475\ \text{kg/d}$

iii. Estimate the mass of volatile solids produced in the activated-sludge process that must be wasted. [The required value is computed using Eq. (8–14)].

$$P_{x,VSS} = Y_{\text{obs}}Q(S_o - S)/(10^3\ \text{g/1 kg})$$

$$= \frac{0.3125(21{,}600\ \text{m}^3/\text{d})[(250 - 6.2)\ \text{g/m}^3]}{(10^3\ \text{g/1 kg})} = 1646\ \text{kg/d}$$

Note: The actual flowrate will be the primary influent less the flowrate of the primary underflow. However, the primary underflow is normally small and can be neglected. If the underflow is significant, the actual flowrate should be used to determine the volatile solids production.

iv. Estimate the TSS_M that must be wasted assuming the volatile fraction represents 0.80 of the total solids.

$\text{TSS}_M = 1646/0.80 = 2057\ \text{kg/d}$

Note: If it is assumed that the fixed solids portion of the influent suspended solids equals 0.15, the mass of fixed solids in the input from the primary settling facilities is equal to 0.15 × 2333 = 350 kg/d. This value can then be compared with the fixed solids determined in the above computations, which is equal to 2057 − 1646 = 411 kg/d. The ratio of these values is 1.18[(411 kg/d)/(350 kg/d)]. Values that have been observed for this ratio vary from about 1.0 to 1.3; a value of 1.15 is considered to be the most representative.

v. Estimate the waste quantities discharged to the thickener. (It is assumed in this example that wasting is from the biological reactor.)

$\text{TSS}_M = (2057 - 475)\ \text{kg/d} = 1582\ \text{kg/d}$

$$\text{Flowrate} = \frac{(1582\ \text{kg/d})(10^3\ \text{g/1 kg})}{(4375\ \text{g/m}^3)} = 362\ \text{m}^3/\text{d}$$

The assumed concentration value of MLSS of 4375 g/m^3 in the aeration tank will increase when the recycled BOD_C and TSS_C are taken into consideration in the second and subsequent iterations of the mass balance.

d. Flotation thickeners

i. Operating parameters:
Concentration of thickened sludge = 4%
Assumed solids recovery = 90%
Assumed specific gravity of feed and thickened sludge = 1.0

ii. Determine the flowrate of the thickened sludge.

$$\text{Flowrate} = \frac{(1582\ \text{kg/d})(0.9)}{(10^3\ \text{kg/m}^3)(0.04)} = 35.6\ \text{m}^3/\text{d}$$

iii. Determine the flowrate recycled to the plant influent.

$$\text{Recycled flowrate} = (362 - 35.6)\ \text{m}^3/\text{d} = 326.4\ \text{m}^3/\text{d}$$

iv. Determine the TSS_M to the digester.

$$TSS_M = (1582\ \text{kg/d})(0.9) = 1424\ \text{kg/d}$$

v. Determine the TSS_M recycled to the plant influent.

$$TSS_M = (1582 - 1424)\ \text{kg/d} = 158\ \text{kg/d}$$

vi. Determine the BOD_C of the TSS_C in the recycled flow.

$$TSS_C \text{ in recycled flow} = \frac{(158\ \text{kg/d})(10^3\ \text{g/1 kg})}{326\ \text{m}^3/\text{d}} = 485\ \text{g/m}^3$$

$$\begin{aligned} BOD_C \text{ of the } TSS_C &= (485\ \text{g/m}^3)(0.65)(1.42)(0.68) \\ &= 304.6\ \text{g/m}^3 \end{aligned}$$

$$BOD_M = (304.6\ \text{g/m}^3)(326\ \text{m}^3/\text{d})(1\ \text{kg}/10^3\ \text{g}) = 99\ \text{kg/d}$$

e. Sludge digestion

i. Operating parameters:

SRT = 20 d
VSS destruction during digestion = 50%
Gas production = 1.12 m^3/kg of VSS destroyed
BOD_C in digester supernatant = 1000 g/m^3 (0.1%)
TSS_C in digester supernatant = 5000 g/m^3 (0.5%)
TSS_C in digested sludge = 5%

ii. Determine the total solids fed to the digester and the corresponding flowrate.

TSS_M = solids from primary settling plus waste solids from thickener
TSS_M = 5443 kg/d + 1424 kg/d = 6867 kg/d

$$\begin{aligned} \text{Total flowrate} &= \frac{(5443\ \text{kg/d})}{0.06(10^3\ \text{kg/m}^3)} + \frac{(1424\ \text{kg/d})}{0.04(10^3\ \text{kg/m}^3)} \\ &= (90.7 + 35.6)\ \text{m}^3/\text{d} = 126.3\ \text{m}^3/\text{d} \end{aligned}$$

iii. Determine the VSS_M fed to the digester.

$$\begin{aligned} VSS_M &= 0.682(5443\ \text{kg/d}) + 0.80(1424\ \text{kg/d}) \\ &= (3712 + 1139)\ \text{kg/d} = 4851\ \text{kg/d} \end{aligned}$$

$$\begin{aligned} \text{Percent } VSS_M \text{ in mixture fed to digester} &= \frac{(4851\ \text{kg/d})}{(6867\ \text{kg/d})}(100) \\ &= 70.6\% \end{aligned}$$

iv. Determine the VSSM destroyed.

$$VSS_M = 0.5(4851\ \text{kg/d}) = 2426\ \text{kg/d}$$

v. Determine the mass flowrate to the digester.

Primary sludge at 6% solids:

$$\text{Mass flow} = \frac{(5443\ \text{kg/d})}{0.06} = 90{,}717\ \text{kg/d}$$

Thickened waste-activated sludge at 4% solids:

$$\text{Mass flowrate} = \frac{(1424\ \text{kg/d})}{0.04} = 35{,}600\ \text{kg/d}$$

$$\text{Total mass flowrate} = (90{,}717 + 35{,}600)\ \text{kg/d} = 126{,}317\ \text{kg/d}$$

Note: The total mass flow can also be computed by multiplying the total flowrate to the digester by the density of the combined primary sludge and the thickened biosolids, if known.

vi. Determine the mass quantities of gas and sludge after digestion. Assume that the total mass of fixed solids does not change during digestion and that 50% of the volatile solids is destroyed.

$$\text{Fixed solids} = \text{TSS}_M - \text{VSS}_M = (6867 - 4851)\ \text{kg/d} = 2016\ \text{kg/d}$$
$$\text{TSS}_M \text{ in digested sludge} = 2016\ \text{kg/d} + 0.5(4851\ \text{kg/d}) = 4441\ \text{kg/d}$$

Gas production assuming that the density of digester gas is equal to 0.86 times that of air (1.204 kg/m^3, see Appendix B):

$$\text{Gas} = (1.12\ \text{m}^3/\text{kg})(0.5)(4851\ \text{kg/d})(0.86)(1.204\ \text{kg/m}^3) = 2813\ \text{kg/d}$$

Mass balance of digester output:

Mass input = 126,317 kg/d
Less gas = − 2813 kg/d
Mass output = 123,504 kg/d (solids and liquid)

vii. Determine the flowrate distribution between the supernatant at 5000 mg/L and digested sludge at 5% solids. Let TSS_{SP} = kg/d of supernatant suspended solids.

$$\frac{\text{TSS}_{SP}}{0.005} = \frac{4441 - \text{TSS}_{SP}}{0.05} = 123{,}504\ \text{kg/d}$$

$$\text{TSS}_{SP} + 444.1 - (0.1)\text{TSS}_{SP} = 617.5\ \text{kg/d}$$

$$(0.9)\text{TSS}_{SP} = 173\ \text{kg/d}$$

$$\text{TSS}_{SP} = 192\ \text{kg/d}$$

$$\text{Digested solids} = (4441 - 192)\ \text{kg/d} = 4249\ \text{kg/d}$$

$$\text{Supernatant flowrate} = \frac{(192\ \text{kg/d})}{0.005(10^3\ \text{kg/m}^3)} = 38.4\ \text{m}^3/\text{d}$$

$$\text{Digested sludge flowrate} = \frac{(4929\ \text{kg/d})}{0.05(10^3\ \text{kg/m}^3)} = 85\ \text{m}^3/\text{d}$$

viii. Establish the characteristics of the recycled flow.

Flowrate = 38.4 m^3/d
$\text{BOD}_C = (38.4\ \text{m}^3/\text{d})(1000\ \text{g/m}^3)/(10^3\ \text{g/1 kg}) = 38\ \text{kg/d}$
$\text{TSS}_M = (38.4\ \text{m}^3/\text{d})(5000\ \text{g/m}^3)/(10^3\ \text{g/1 kg}) = 192\ \text{kg/d}$

f. Sludge dewatering. (Note: In the analysis that follows, the weight of the polymer or other sludge-conditioning chemicals that may be added was not considered. In some cases, their contribution can be significant and must be considered.)

i. Operating parameters for centrifuge:

Sludge cake = 22% solids
Specific gravity of sludge = 1.06
Solids capture = 93%
Centrate BOD_C = 2000 mg/L

ii. Determine the sludge-cake characteristics.

$$\text{Solids} = (4249 \text{ kg/d})(0.93) = 3952 \text{ kg/d}$$

$$\text{Volume} = \frac{(3952 \text{ kg/d})}{1.06\,(0.22)(0.22)(10^3 \text{ kg/m}^3)} = 16.9 \text{ m}^3/\text{d}$$

iii. Determine the centrate characteristics.

$$\text{Flowrate} = (85 - 16.9) \text{ m}^3/\text{d} = 68.1 \text{ m}^3/\text{d}$$

$$\begin{aligned}\text{BOD}_M \text{ (at 2000 g/m}^3) &= (2000 \text{ g/m}^3)(68.1 \text{ m}^3/\text{d})(10^3 \text{ g/1 kg}) \\ &= 136 \text{ kg/d}\end{aligned}$$

$$\text{TSS}_M = (4249 \text{ kg/d})(0.07) = 297 \text{ kg/d}$$

g. Prepare a summary table of the recycle flows and waste characteristics for the first iteration.

Operation	Flowrate, m^3/d	BOD_M, kg/d	TSS_M, kg/d
Flotation thickener	326.0	99	158
Digester supernatant	38.4	38	192
Centrate	68.1	136	297
Totals	432.5	273	647[a]

[a] The volatile fraction of the returned suspended solids will typically vary from 50 to 75 percent. A value of 60 percent will be used for the computation in the second iteration.

4. Prepare the second iteration of the solids balance.

a. Primary settling

i. Operating parameters = same as those in the first iteration

ii. TSS_M and BOD_M entering the primary tanks

$$\begin{aligned}\text{TSS}_M &= \text{influent TSS}_M + \text{recycled TSS}_M \\ &= 7776 \text{ kg/d} + 647 \text{ kg/d} = 8423 \text{ kg/d}\end{aligned}$$

$$\begin{aligned}\text{Total BOD}_M &= \text{influent BOD}_M + \text{recycled BOD}_M \\ &= 8100 \text{ kg/d} + 273 \text{ kg/d} = 8373 \text{ kg/d}\end{aligned}$$

iii. BOD_M removed = 0.33(8373 kg/d) = 2763 kg/d

iv. BOD_M to secondary = (8,373 − 2763) kg/d = 5610 kg/d

v. TSS_M removed = 0.7(8423 kg/d) = 5896 kg/d

vi. TSS_M to secondary = (8423 − 5896) kg/d = 2527 kg/d

b. Determine the volatile fraction of the primary sludge and effluent suspended solids.

i. Operating parameters:

Incoming wastewater = same as those for the first iteration
Volatile fraction of solids in recycle returned to headworks = 60%

ii. Although the computations are not shown, the computed change in the volatile fractions determined in the first iteration is slight and, therefore, the values determined previously are used for the second iteration. If the volatile fraction of the return is less than about 50%, the volatile fractions should be recomputed.

c. Secondary process

i. Operating parameters = same as those for the first iteration and as follows:

Aeration tank volume = 4700 m^3
SRT = 10 d
$Y = 0.50$ kg/kg
$b = 0.06\ d^{-1}$

ii. Determine the BOD_C in the influent to the aeration tank.

Flowrate to aeration tank = influent flowrate + recycled flowrate

$$= (21{,}600 + 432.5)\ \text{m}^3/\text{d} = 22{,}033\ \text{m}^3/\text{d}$$

$$\text{BOD}_C = \frac{(5610\ \text{kg/d})(10^3\ \text{g/1 kg})}{(22{,}032.5\ \text{m}^3/\text{d})} = 255\ \text{g/m}^3$$

iii. Determine the new concentration of mixed liquor VSS.

$$X_{\text{VSS}} = \frac{(Q)(Y)(S_o - S)\text{SRT}}{[1 + b(\text{SRT})](V_r)}$$

$$X_{\text{VSS}} = \frac{(22{,}035\ \text{m}^3/\text{d})(0.5)(255 - 6.2)(10\ \text{d})}{[1 + (0.06\ \text{d}^{-1})(10\ \text{d})](4700\ \text{m}^3/\text{d})} = 3648\ \text{g/m}^3$$

iv. Determine the mixed liquor suspended solids.

$$X_{\text{SS}} = \frac{X_{\text{VSS}}}{0.8}$$

$$X_{\text{SS}} = 3648/0.8 = 4560\ \text{g/m}^3$$

v. Determine the cell growth.

$$P_{x,\text{VSS}} = Y_{\text{OBS}}\, Q\, (S_o - S)/(10^3\ \text{g/1 kg})$$

$$= \frac{0.3125\ (22{,}032.5\ \text{m}^3/\text{d})[(255 - 6.2\ \text{g/m}^3)]}{(10^3\ \text{g/1 kg})} = 1714\ \text{kg/d}$$

$$P_{x,\text{TSS}} = 1714/0.8 = 2143\ \text{kg/d}$$

vi. Determine the waste quantities discharged to the thickener.

Effluent TSS_M = 432 kg/d (specified in the first iteration)

Total TSS_M to be wanted to the thicker = (2143 – 432) kg/d

= 1711 kg/d

$$\text{Flowrate} = \frac{(1711\ \text{kg/d})(10^3\ \text{g/1 kg})}{(4560\ \text{g/m}^3)} = 375\ \text{m}^3/\text{d}$$

d. Flotation thickeners

i. Operating parameters:

Concentration of thickened sludge = 4%
Assumed solids recovery = 90%
Assumed specific gravity of feed and thickened sludge = 1.0

ii. Determine the flowrate of the thickened sludge.

$$\text{Flowrate} = \frac{(1711\ \text{kg/d})(0.9)}{(10^3\ \text{kg/m}^3)(0.04)} = 38.5\ \text{m}^3/\text{d}$$

iii. Determine the flowrate recycled to the plant influent.

Recycled flowrate = $(375 - 38.5)\ \text{m}^3/\text{d} = 336.5\ \text{m}^3/\text{d}$

iv. Determine the TSS_M to the digester.

$TSS_M = (1711\ \text{kg/d})(0.9) = 1540\ \text{kg/d}$

v. Determine the TSS_M recycled to the plant influent.

$TSS_M = (1711 - 1540)\ \text{kg/d} = 171\ \text{kg/d}$

vi. Determine the BOD_C of the TSS_C in the recycled flow.

$$\text{TSS}_C \text{ in recycled flow} = \frac{(171 \text{ kg/d})(10^3 \text{ g/1 kg})}{(336.5 \text{ m}^3/\text{d})} = 508 \text{ g/m}^3$$

$$\text{BOD}_C \text{ of TSS}_C = (508 \text{ g/m}^3)(0.65)(1.42)(0.68) = 319 \text{ g/m}^3$$
$$\text{BOD}_M = (319 \text{ g/m}^3)(336.5 \text{ m}^3/\text{d})(10^3 \text{ g/1 kg})^{-1} = 107 \text{ kg/d}$$

e. Sludge digestion

i. Operating parameters = same as those in the first iteration

ii. Determine the total solids fed to the digester and the corresponding flowrate.
$\text{TSS}_M = \text{TSS}_\text{M}$ from primary settling plus waste TSSM from thickener
$\text{TSS}_M = 5443 \text{ kg/d} + 1540 \text{ kg/d} = 6983 \text{ kg/d}$

$$\text{Total flowrate} = \frac{(5443 \text{ kg/d})}{0.06(10^3 \text{ kg/m}^3)} + \frac{(1540 \text{ kg/d})}{0.04(10^3 \text{ kg/m}^3)}$$
$$= (90.7 + 38.5) \text{ m}^3/\text{d} = 129.2 \text{ m}^3/\text{d}$$

iii. Determine the total VSSM fed to the digester.

$$\text{VSS}_M = 0.682(5443 \text{ kg/d}) + 0.80(1540 \text{ kg/d})$$
$$= (3712 + 1232) \text{ kg/d} = 4944 \text{ kg/d}$$

$$\text{Percent VSS in mixture fed to digester} = \frac{(4944 \text{ kg/d})}{(6983 \text{ kg/d})}(100)$$
$$= 71.3\%$$

iv. Determine the VSS destroyed.

$$\text{VSS destroyed} = 0.5(4944 \text{ kg/d}) = 2472 \text{ kg/d}$$

v. Determine the mass flowrate to the digester.

Primary sludge at 6% solids:

$$\text{Mass flowrate} = \frac{(5443 \text{ kg/d})}{0.06} = 90{,}717 \text{ kg/d}$$

Thickened waste-activated sludge at 4% solids:

$$\text{Mass flowrate} = \frac{(1540 \text{ kg/d})}{0.04} = 38{,}500 \text{ kg/d}$$

$$\text{Total mass flowrate} = (90{,}717 + 38{,}500) \text{ kg/d} = 129{,}217 \text{ kg/d}$$

vi. Determine the mass quantities of gas and sludge after digestion. Assume that the total mass of fixed solids does not change during digestion and that 50% of the volatile solids is destroyed.

$$\text{Fixed solids} = \text{TSS}_M - \text{VSS}_M = (6983 - 4944) \text{ kg/d} = 2039 \text{ kg/d}$$
$$\text{TSS in digested sludge} = 2039 \text{ kg/d} + 0.5(4944) \text{ kg/d} = 4511 \text{ kg/d}$$

Gas production assuming that the density of digester gas is equal to 0.86 times that of air (1.204 kg/m^3):

$$\text{Gas} = (1.12 \text{ m}^3/\text{kg})(0.5)(4944 \text{ kg/d})(0.86)(1.204 \text{ kg/m}^3) = 2867 \text{ kg/d}$$

Mass balance of digester output:

Mass input = 129,217 kg/d
Less gas = −2867 kg/d
Mass output = 126,350 kg/d (solids and liquid)

vii. Determine the flowrate distribution between the supernatant at 5000 mg/L and digested sludge at 5 percent solids. Let TSS_{sp} = kg/d of supernatant suspended solids.

$$\frac{\mathrm{TSS}_{sp}}{0.005} = \frac{4441 - \mathrm{TSS}_{sp}}{0.05} = 126{,}350 \text{ kg/d}$$

$$\mathrm{TSS}_{sp} + 451.1 - (0.1)\mathrm{TSS}_{sp} = 631.8 \text{ kg/d}$$

$$(0.9)\mathrm{TSS}_{Sp} = 180.7 \text{ kg/d}$$
$$\mathrm{TSS}_{sp} = 201 \text{ kg/d}$$

$$\text{Digested } \mathrm{TSS}_M = (4511 - 201) \text{ kg/d} = 4310 \text{ kg/d}$$

$$\text{Supernatant flowrate} = \frac{(201 \text{ kg/d})}{0.005(10^3 \text{ kg/m}^3)} = 40.2 \text{ m}^3\text{/d}$$

$$\text{Digested sludge flowrate} = \frac{(4310 \text{ kg/d})}{0.05(10^3 \text{ kg/m}^3)} = 86.2 \text{ m}^3\text{/d}$$

viii. Establish the characteristics of the recycled flow.

Flowrate = 40.2 m^3/d
$\mathrm{BOD}_M = (40.2 \text{ m}^3\text{/d})(1000 \text{ g/m}^3)/(10^3 \text{ g/1 kg}) = 40 \text{ kg/d}$
$\mathrm{TBB}_M = (40.2 \text{ m}^3\text{/d})(5000 \text{ g/m}^3)/(10^3 \text{ g/1 kg}) = 201 \text{ kg/d}$

f. Sludge dewatering

i. Operating parameters for centrifuge = same as those in the first iteration
ii. Determine the sludge-cake characteristics.

$$\mathrm{TSS}_M = (4310 \text{ kg/d})(0.93) = 4008 \text{ kg/d}$$

$$\text{Volume} = \frac{(4008 \text{ kg/d})}{1.06(0.22)(10^3 \text{kg/m}^3)} = 17.2 \text{ m}^3\text{/d}$$

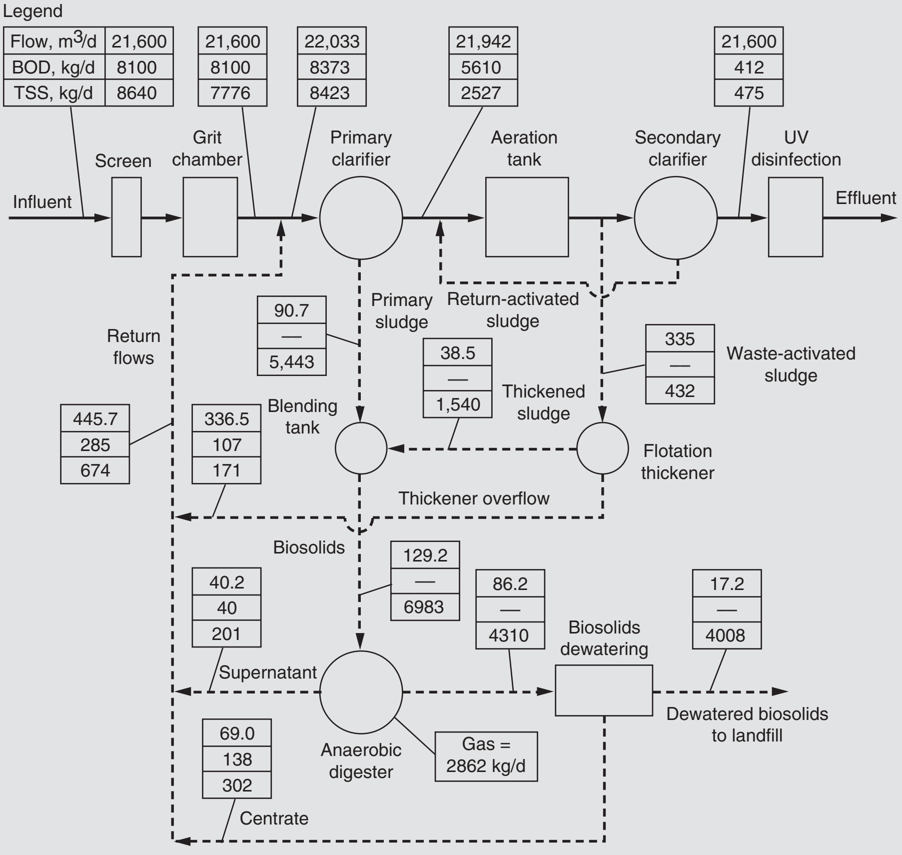
iii. Determine the centrate characteristics.

Flow = $(86.2 - 17.2) \text{ m}^3\text{/d} = 69 \text{ m}^3\text{/d}$
BOD_M (at 2000 g/m^3) = $(2000 \text{ g/m}^3)(69 \text{ g/m}^3)/(10^3 \text{ g/1 kg}) = 138 \text{ kg/d}$
$\mathrm{TSS}_M = (4310 \text{ kg/d})(0.07) = 302 \text{ kg/d}$

g. Prepare a summary table of the recycle flows and waste characteristics for the second iteration.

				Incremental change from previous iteration		
Operation/process	Flow, m^3/d	BOD_M, kg/d	TSS_M, kg/d	Flowrate, m^3/d	BOD_M, kg/d	TSS_M, kg/d
Flotation thickener	326.0	99	158	10.5	8	13
Digester supernatant	38.4	38	192	1.8	2	9
Centrate	68.1	136	297	0.9	2	5
Totals	432.5	273	647[a]	13.2	12	27

5. Because the incremental change in the return quantities is less than 5 percent, the values summarized in the above table are acceptable for design. Given that the above computations would be done on a spreadsheet program, additional iterations could be made to obtain an incremental change of less than 1 percent. The flow, TSSM, and BODM values for the various processes from the second iteration are presented in following figure.



Solids balance for second iteration

Comment In this example, an iterative approach was used to illustrate the computational steps in preparing a solids mass balance. Sludge and biosolids solids balances can be prepared using a specially designed spreadsheet or a proprietary solids balance software program. In general, if the iterative computational procedure is used, similar to the method used in this example, it should be carried out until the incremental change in all of the return quantities from the previous iteration is equal to or less than 5 percent.

14–8 RESOURCE RECOVERY FROM SLUDGES AND BIOSOLIDS

Sludges and biosolids can serve as a source of nutrients that can be used as a fertilizer, as a feedstock for the production of energy, and in the fabrication of value-added products. As discussed in Sec. 13–2 in Chap. 13, biosolids that have been stabilized properly according to EPA Part 503 Regulations can be used for a variety of beneficial uses.

Significant fractions of nutrients in biosolids become solubilized when biosolids are stabilized by anaerobic digestion and return flow from the dewatering process (often called centrate, filtrate or sidestream) contains high concentrations of nutrients. Nutrients in the sidestream can be returned to the main liquid treatment trains or recovered through sidestream treatment. Recovery and use of nutrients in solids is discussed in this section, and recovery of nutrients in the liquid stream is discussed in Chap. 15. The recovery of energy from biosolids and sludge is considered in the following section and in Chap. 17.

Recovery of Nutrients

Among the most valuable nutrients contained in sludges and biosolids are organic nitrogen and phosphorous. The recovery of phosphorous and ammonia is usually from the liquid stream of the dewatering process, as discussed in detail in Chap.15. Following dewatering, Class A or B biosolids can be used beneficially in agriculture land application or non-agriculture land application. The principal beneficial uses of biosolids include (1) agricultural land application, (2) non-agricultural land application, (3) energy recovery and generation, and (4) commercial use. Agricultural and non-agricultural land application are examined below. Energy recovery and generation are discussed in Sec. 14–9.

Agricultural Land Application

Biosolids contain a range of valuable nutrients such as nitrogen, phosphorus, iron, calcium, magnesium and various other macro and micro nutrients which are essential for plant growth. Many nutrients found in biosolids are also essential components in the healthy diet of animals. Biosolids are usually applied to agricultural land at rates designed to supply crops with adequate nitrogen (see Sec. 14–9 for details regarding biosolids land application). Recycling biosolids to agricultural land enables farmers to improve the economics of crop production in lieu of the use of expensive chemical fertilizers and reduces greenhouse gas emission generated from producing chemical fertilizer and carbon sequestration. Class B or A biosolids from processes such as anaerobic digestion and lime stabilization discussed in Chap. 13 are applied to land for crop production as cake; as compost material (see Sec. 14–5); or as a granular material in pellets form produced from drying processes (see Sec. 14–3).

Non-Agricultural Land Applications

Non-agricultural uses of biosolids include reclaiming mining sites, land reclamation, landscaping, and forest crops. Biosolids have been used widely for repairing land damaged by mining such as surface mined areas, abandoned mine lands, and coal refuse piles. Combining mine soils with biosolids, increases the organic matter, the cation exchange capacity, and soil nutrient levels. In mine damaged lands, biosolids application controls pH, metal content, and fertilization of the soils. Reclaiming and improving disturbed and marginal soils is another beneficial use of biosolids. Several biosolids characteristics make this use very successful. The organic matter in biosolids improves the soil physical properties through improving soil granulation and increasing soil water holding capacity. Biosolids increase soil cation exchange capacity, supply plant nutrients, and buffer soil. Land application to forest land is also practiced. However, this use has been difficult to achieve due to difficulties in spreading biosolids evenly through heavily forested areas. Finally, the use of biosolids for horticulture and landscaping, such as golf courses, is similar to agricultural land application but not used as a fertilizer. Biosolids compost is most popular product for landscaping uses as compost is primarily a soil conditioner and not a fertilizer.

14–9 ENERGY RECOVERY FROM SLUDGE AND BIOSOLIDS

Sludges and biosolids are considered renewable energy resources as they contain organic material that has a fuel value that can be harnessed. Under a properly engineered and controlled environment, energy recovery and generation from biosolids is considered at the top of the hierarchy of beneficial use due to the increased cost of energy and the need to obtain green energy credit by municipal utilities. Sludges from wastewater can be processed to generate energy by (WERF, 2008b):

1. Producing methane through anaerobic digester
2. Thermally oxidize sludge
3. Producing syngas through gasification process and/or pyrolysis
4. Producing oil and liquid fuel

Each of these options along with the appropriate method for energy recovery is considered briefly below. Additional details on energy recovery are presented in Chap. 17.

Energy Recovery through Anaerobic Digestion

Production of methane rich gas by anaerobic digestion is one pathway used to recover energy from biosolids. Gas production, collection and use methods are discussed in Sec. 13–9. Combined heat and power (CHP) recovery from the biogas generated from anaerobic digestion of sludge is illustrated on Fig. 14–34. As shown on Fig. 14–34 , the waste heat can be used as an energy source to further dry and stabilize the cake solids. Drying processes are discussed in Sec. 14–3. The dried material can also be used beneficially in agricultural or non-agricultural land application. Another beneficial use of the dried material can be hauling to cement kilns to supplement coal as a fuel. The value of the dried material is usually compared to the market value of coal on a heat content basis. Energy can be further recovered from the dried material on site to provide energy needed for drying as discussed below.

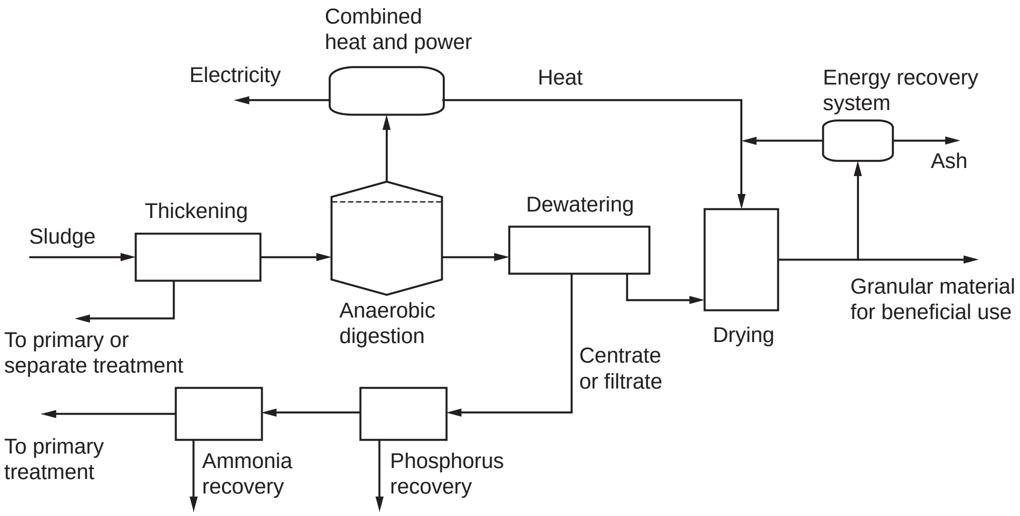


Figure 14–34

Schematic process flow diagram for the recovery of energy with anaerobic digestion. (The simultaneous recovery of nutrients is considered in Chap. 15.)

Table 14–23

Characteristics of different thermal processing technologies

Parameter	Combustion	Gasification	Pyrolysis
Temperature, °C (°F)	900–1100 (1650–2000)	590–980 (1100–1800)	200–590 (390–1100)
O_2 supplied	>Stoichiometric (excess air)	<Stoichiometric (limited air)	None
By-products	Flue gas (CO_2, H_2O) and ash	Syngas (CO, H_2) and ash	Pyrolysis gas, oils, tars and char

Energy Recovery by Thermal Oxidation

The three basic thermal processes are combustion or advanced thermal oxidation, gasification, and pyrolysis. All thermal processing of dried biosolids is followed by energy recovery methods. The difference between the three thermal processing technologies, in terms of operating temperature and oxygen requirements, and the main byproducts from each are summarized in Table 14–23. Thermal oxidation of sludge or biosolids is more common in the areas where ultimate disposal of sludge and biosolids is costly due to limited disposal sites. Sludges or biosolids to be incinerated, as described in Sec. 14–4, are typically dewatered to 15 to 30 percent dry solids. Further drying and combustion occurs simultaneously in the combustion reactor. The combustion process produces hot flue gas where the energy can be recovered for combustion air preheating and other energy needs or electrical production as shown previously on Fig. 14–25. The ash produced from the advanced thermal oxidation is an inert material that can be used in commercial applications such as cement making, asphalt, etc. Technologies to recover valuable material from the ash such as phosphorous are under development and once commercially developed may gain wide interest due to the potential market value of such resources. Excess heat from thermal oxidation of dewatered sludge has been used recently to generate steam for electricity generation.

Energy Recovery from Dried Material through Gasification and Pyrolysis

Gasification is an established process for converting organic materials to a fuel gas called synthetic gas or syngas, and has been practiced since the 1800s to generate fuel gas from coal and other biomass. Syngas is composed mainly of CO, CO_2, H_2, and CH_4 and has a low heating value of 4500–5500 kJ/m^3 (120–150 BTU/ft^3), which is approximately 25 percent of the heat value of biogas generated from anaerobic digestion. While gasification is still in the early phases of development for processing biosolids, there has been increased interest in applying this technology to sludges and biosolids. Sludges or biosolids have a higher ash content as compared to traditional organic materials used as a fuel source in gasification processes, which makes ash handling more difficult. Further details of gasification and energy recovery is considered in Chap. 17.

Pyrolysis is also an established technology used in the chemical industry to produce charcoal, activated carbon and methanol. Similar to gasification, pyrolysis at high temperatures generates a combustible gas, pyrolysis gas, with a low heating value but also can be used to generate char and oil. Pyrolysis is actually the first step that occurs in both gasification and combustion reactions. Pyrolysis of sludges and biosolids is still considered innovative technology and not widely practiced for municipal applications.

To effectively harness energy through a gasification or pyrolysis technology, most commercially available systems require dried biosolids with greater than 75 percent solids in granular form. Pelletization is not required; however, a certain degree of uniformity in the dried granular material along with low dust content is required. The required dryness depends on the technology. The energy required for drying is typically supplied by the thermal technology. Energy can also be recovered from waste heat, such as a combined heat and power (CHP) system, if available onsite.

Production of Oil and Liquid Fuel

Converting sludge to liquid and oil fuel is another use of sludge as energy source. However, the commercial application of such technologies is hindered by the high capital cost and requirement for large amount of sludge feed to lower the capital and operating costs.

14–10 APPLICATION OF BIOSOLIDS TO LAND

Land application of biosolids is defined as the spreading of biosolids on or just below the soil surface. Biosolids may be applied to (1) agricultural land, (2) forest land, (3) disturbed land, and (4) dedicated land disposal sites. In all four cases, the land application is designed with the objective of providing further biosolids treatment. Sunlight, soil microorganisms, and desiccation combine to further inactivate any residual pathogens and many toxic organic substances. Trace metals are trapped in the soil matrix and nutrients are taken up by plants and converted to useful biomass. In some cases, a geomembrane liner is installed below a dedicated land disposal area.

To qualify for application to agricultural and nonagricultural land, biosolids or material derived from biosolids must meet at least the pollutant ceiling concentrations, Class B requirements for pathogens, and vector attraction requirements. Bulk biosolids applied to lawns and home gardens and biosolids that are sold or given away in bags or containers must meet the Class A criteria and one of several available vector-attraction reduction processes.

Benefits of Land Application

The application of biosolids to land for agricultural purposes is beneficial because organic matter improves soil structure, tilth, water holding capacity, water infiltration, and soil aeration. Macronutrients (nitrogen, phosphorus, potassium) and micronutrients (iron, manganese, copper, chromium, selenium, and zinc) aid plant growth. Organic matter also contributes to the cation-exchange capacity (CEC) of the soil, which allows the soil to retain potassium, calcium, and magnesium. The presence of organic matter improves the biological diversity in soil and improves the availability of nutrients to the plants (Wegner, 1992). Nutrients in the biosolids also serve as a partial replacement for expensive chemical fertilizers.

Land application can also be of great value in silviculture and site reclamation. Forest utilization has been practiced extensively in the northwest, and biosolids application has been recognized as being beneficial to forest growth (WEF, 2010). Reclamation of disturbed land such as superfund sites has also been successful (Henry and Brown, 1997).

U.S. EPA Regulations for Beneficial Use and Disposal of Biosolids

As discussed in Sec. 13–2, the U.S. EPA published regulations for biosolids (sewage sludge is the term used in the regulations) use and disposal under the code of Federal Regulations (CFR), 40 CFR Part 503. For land application, the regulations provide numerical limits on 10 metals, management practice guidance, and requirements for monitoring, record keeping, and reporting. The regulations are summarized in Table 14–24 and discussed in the following paragraphs.

Table 14–24

U.S. EPA sludge regulations for land application

Classification	Class A: no restrictions[a] Class B: site restrictions
Management practices	See Table 14–25
Pathogen reduction alternatives	See Table 13–9
Vector attraction reduction	See Table 13–10
Site restrictions for Class B biosolids	See Table 14–26
Metal limits and loading rates	See Table 14–30

[a] Other than bag labeling (like a fertilizer).

Table 14–25

Land application management practices under U.S. EPA Part 503 rule[a]

For bulk biosolids[b]

- Bulk biosolids cannot be applied to flooded, frozen, or snow-covered agricultural land, forests, public contact sites, or reclamation sites in such a way that the biosolids enters a wetland or other waters of the United States (as defined in 40 CFR Part 122.2), except as provided in a permit issued pursuant to Section 402 (NPDES permit) or Section 404 (Dredge and Fill Permit) of the Clean Water Act, as amended.
- Bulk biosolids cannot be applied to agricultural land, forests, or reclamation sites that are 10 m or less from U.S. waters, unless otherwise specified by the permitting authority.
- If applied to agricultural lands, forests, or public contact sites, bulk biosolids must be applied at a rate that is equal to or less than the agronomic rate for the site. Biosolids applied to reclamation sites may exceed the agronomic rate if allowed by the permitting authority.
- Bulk biosolids must not harm or contribute to the harm of a threatened or endangered species or result in the destruction or adverse modification of the species' critical habitat when applied to the land. Threatened or endangered species and their critical habitats are listed in Section 4 of the Endangered Species Act. Critical habitat is defined as any place where a threatened or endangered species lives and grows during any stage of its life cycle. Any direct or indirect action (or the result of any direct or indirect action) in a critical habitat that diminishes the likelihood of survival and recovery of a listed species is considered destruction or adverse modification of a critical habitat.

For biosolids sold or given away in a bag or other container for application to the land[a]

- A label must be affixed to the bag or other container, or an information sheet must be provided to the person who receives this type of biosolids in another container. At a minimum, the label or information sheet must contain the following information:
  - The name and address of the person who prepared the biosolids for sale or give-away in a bag or other container.
  - A statement that prohibits application of the biosolids to the land except in accordance with the instructions on the label or information sheet.
  - An AWSAR (annual whole sludge application rate) for the biosolids that does not cause the annual pollutant loading rate limits to be exceeded.

[a] from U.S. EPA (1995).

[b] These management practices do not apply if the biosolids is of "exceptional quality."

Management Practices

Management practices that must be followed when biosolids are applied on land are specified in the Part 503 rule (see Table 14–25). The practices vary depending on whether the material that is applied is hauled in bulk or in individual bags.

Pathogen-Reduction Alternatives. As discussed in Chap. 13 and in Sec. 13–2, the Part 503 pathogen-reduction requirements for biosolids are divided into Class A and Class B categories (see Table 13–9). The goal of the Class A requirements is to reduce the pathogens in the biosolids (including Salmonella sp. bacteria, enteric viruses, and viable helminth ova) to below detectable levels. When this goal is achieved, Class A biosolids can be land applied without any pathogen-related restrictions on the site (U.S. EPA, 1995). The goal of the Class B requirements is to ensure that pathogens have been reduced to levels that are unlikely to pose a threat to public health and the environment under specific use conditions. Site restrictions on land application of Class B biosolids minimize the potential for human and animal contact with the biosolids until environmental factors have reduced pathogens to below detectable levels.

Vector Attraction Reduction. There are 10 potential vector attraction reduction measures that can be combined with pathogen-reduction alternatives for an acceptable land-application project using Class B biosolids (see Table 13–10). The list in Table 13–10 also includes some stabilization processes that reduce pathogens.

Site Restrictions for Class B Biosolids. Site restrictions, listed in Table 14–26, depend on the crops to be used and the contact control for animals and the public. Food crops and turf grass are given the longest time restrictions because of the potential for public exposure (U.S. EPA, 1995).

Table 14–26

Site restrictions for Class B biosolids[a]

Restrictions for the harvesting of crops and turf

- Food crops with harvested parts that touch the biosolids/soil mixture and are totally above ground shall not be harvested for 14 mo after application of biosolids.
- Food crops with harvested parts below the land surface where biosolids remains on the land surface for 4 mo or longer prior to incorporation into the soil shall not be harvested for 20 mo after biosolids application.
- Food crops with harvested parts below the land surface where biosolids remains on the land surface for less than 4 mo prior to incorporation shall not be harvested for 38 mo after biosolids application.
- Food crops, feed crops, and fiber crops, whose edible parts do not touch the surface of the soil, shall not be harvested for 30 d after biosolids application.
- Turf grown on land where biosolids is applied shall not be harvested for 1 y after application of the biosolids when the harvested turf is placed on either land with a high potential for public exposure or a lawn, unless otherwise specified by the permitting authority.

Restriction for the grazing of animals

- Animals shall not graze on land for 30 d after application of biosolids to the land.

Restrictions for public contact

- Access to land with a high potential for public exposure, such as a park or ball field, is restricted for 1 y after biosolids application. Examples of restricted access include posting with no trespassing signs or fencing.
- Access to land with a low potential for public exposure (e.g., private farmland) is restricted for 30 d after biosolids application. An example of restricted access is remoteness.

[a] From U.S. EPA (1995).

Exceptional Quality Biosolids. The category of "exceptional quality" biosolids has been defined as those biosolids that meet metal standards, Class A pathogen reduction standards, and vector reduction standards as defined in the Part 503 regulations.

Site Evaluation and Selection

A critical step in land application of biosolids is finding a suitable site. The characteristics of the site will determine the actual design and will influence the overall effectiveness of the land-application concept. The sites considered potentially suitable will depend on the land-application option or options being considered, such as application to agricultural lands, forest lands, etc. The site-selection process should include an initial screening on the basis of the factors and criteria described in the following discussion. For screening purposes, it is necessary to have at least a rough estimate of land-area requirements for each feasible option.

Ideal sites for land application of biosolids have deep silty loam to sandy loam soils, groundwater deeper than 3 m (10 ft), slopes at 0 to 3 percent; no wells, wetlands, or streams; and few neighbors. Site characteristics of importance are topography, soil characteristics, soil depth to groundwater, and accessibility and proximity to critical areas.

Topography. Topography is important as it affects the potential for erosion and runoff. Suitability of site topographies also depends on the type of biosolids and the method of application. As shown in Table 14–27, liquid biosolids can be spread, sprayed, or injected onto sites with rolling terrain up to 15 percent in slope. Dewatered sludge is usually spread on agricultural land that requires a tractor and spreader. Forested sites can accommodate slopes up to 30 percent if adequate setbacks from streams are provided.

Soil Characteristics. In general, desirable soil characteristics include (1) loamy soil, (2) slow to moderate permeability, (3) soil depth of 0.6 m (2 ft) or more, (4) alkaline or neutral soil pH (pH > 6.5), and (5) well drained to moderately well drained soil. Practically any soil can be adapted to a well-designed and well-operated system.

Soil Depth to Groundwater. A basic philosophy inherent in federal and state regulations is to design biosolids application systems that are based on sound agronomic principles, so that biosolids pose no greater threat to groundwater than current agricultural practices. Because the groundwater fluctuates on a seasonal basis in many soils, difficulties are encountered in establishing an acceptable minimum depth to groundwater. The quality of the underlying groundwater and the biosolids application option have to be considered carefully, especially where groundwater nondegradation restrictions apply.

Table 14–27

Typical slope limitations for land application of biosolids

Slope,%	Comment
0–3	Ideal; no concern for runoff or erosion of liquid or dewatered biosolids
3–6	Acceptable; slight risk of erosion; surface application of liquid or dewatered biosolids is acceptable
6–12	Injection of liquid biosolids required for general cases, except in closed drainage basin and/or when extensive runoff control is provided; surface application of dewatered biosolids is generally acceptable
12–15	No application of liquid biosolids should be made without extensive runoff control; surface application of dewatered biosolids is acceptable, but immediate incorporation into the soil is recommended
Over 15	Slopes greater than 15 percent are suitable only for sites with good permeability where the length of slope is short and where the area with steep slope is a minor part of the total application area

Generally, the greater the depth to the water table, the more desirable a site is for biosolids application. At least 1 m (3 ft) to groundwater is preferred for land-application sites. Seasonal water-table fluctuations to within 0.5 m (1.5 ft) of the surface can be tolerated. If the shallow groundwater is excluded as a drinking-water aquifer, the groundwater depth can be as shallow as 0.5 m before problems with trafficability of the soil arise. The presence of faults, solution channels, and other similar connections between soil and groundwater is undesirable unless the depth of overlying soil is adequate. When a specific site or sites are being considered for biosolids application, a detailed field investigation may be necessary to obtain the required groundwater information.

Accessibility and Proximity to Critical Areas. Buffer zones or setbacks are needed to separate the active application area from sensitive areas such as residences, wells, roads, surface waters, and property boundaries. Local and state regulations often include minimum distances for setbacks depending on the method of application; example minimum setback distances used in California are listed in Table 14–28.

Design Loading Rates

Design loading rates for land application of biosolids can be limited by pollutants (heavy metals) or by nitrogen. The long-term loadings of heavy metals are based on U.S. EPA 503 regulations. The annual loading rate is usually limited by the nitrogen loading rate.

Nitrogen Loading Rates. Nitrogen loading rates are set typically to match the available nitrogen provided by commercial fertilizers (Chang et al., 1995). Because municipal biosolids represent a slow-release organic fertilizer, a combination of ammonia and organic nitrogen must be made according to Eq. (14–6).

$$L_N = [(\mathrm{NO_3}) + K_V(\mathrm{NH_4}) + f_n(N_o)]F \qquad (14\text{–}6)$$

where L_N = plant available nitrogen in the application year, g N/kg (lb N/ton)

$\mathrm{NO_3}$ = percent nitrate nitrogen in biosolids, decimal

Table 14–28

Typical setback distances for land application sites[a]

Setback from:	Minimum distance ft	m
Property Boundaries	10	3
Domestic water supply wells	500	150
Nondomestic water supply wells	100	30
Public roads and onsite occupied residences	50	15
Surface waters (wetlands, creeks, ponds, lakes, underground aqueducts, and marshes)	100	30
Primary agricultural drainageways	33	10
Occupied nonagricultural buildings and offsite residences	500	150
Domestic water supply reservoir	400	120
Primary tributary to a domestic water supply	200	60
Domestic surface water supply intake	2500	750

[a] CSWRCB (2000).

K_v = volatilization factor for ammonia loss
= 0.5 for surface-applied liquid sludge
= 0.75 for surface-applied dewatered sludge
= 1.0 for injected liquid or dewatered sludge
NH_4 = percent ammonia nitrogen in sludge, decimal
f_n = mineralization factor for organic nitrogen
= 0.5 for warm climates and digested sludge
= 0.4 for cool climates and digested sludge
= 0.3 for cold climates or composted sludge
N_o = percent organic nitrogen in sludge, decimal
F = conversion factor, 1000 g/kg of dry solids (2000 lb/ton)

To use Eq. (14–6) requires knowledge of the method of application, the nitrogen content of the biosolids (nitrate, ammonia, and organic), the type of stabilization, and the type of climate. The use of the mineralization factors simplifies the previously used method of calculating the amount of organic nitrogen mineralized each year and adding up the total for an annual equivalent. The use of Eq. (14–6) is also appropriate if biosolids are applied to a single site once every 2 to 3 y.

The loading rate based on nitrogen loadings is then calculated from Eq. (14–7).

$$L_{SN} = \frac{U}{N_P F} \tag{14–7}$$

where L_{SN} = biosolids loading rate based on N, kg/ha·y (ton/ac·y)
U = crop uptake of nitrogen, kg/ha (lb/ac) (see Table 14–29)
N_p = plant available nitrogen in sludge, g/kg (lb/ton)
F = conversion factor, 10^{-3} kg/g (1 lb/lb)

Loading Rates Based on Pollutant Loading. The pollutants of concern are those listed in Table 14–30. To calculate the biosolids loading rate based on pollutant loading use Eq. (14–8).

$$L_S = \frac{L_C}{CF} \tag{14–8}$$

where L_S = maximum amount of biosolids that can be applied per year, kg/ha·y (tons/ac·y)
L_C = maximum amount of constituent that can be applied per year, kg/ha·y (lb/ac·y)
C = pollutant concentration in biosolids, decimal (mg/kg)
F = conversion factor, 10^{-6} kg/mg (2000 lb/ton)

Land Requirements. Once the maximum biosolids loading rate is determined [by comparing the values from Eqs. (14–7) and (14–8)], the field area can be calculated using Eq. (14–9).

$$A = \frac{B}{L_S} \tag{14–9}$$

where A = application area required, ha
B = biosolids production, kg of dry solids/y

Table 14–29
Typical nitrogen uptake values for selected crops[a]

	Nitrogen uptake			Nitrogen uptake	
Crop	**lb/ac·y**	**kg/ha·y**	**Crop**	**lb/ac·y**	**kg/ha·y**
Forage crops			Tree crops		
Alfalfa	200–600	220–670	Eastern forests		
Brome grass	115–200	130–220	Mixed hardwoods	200	225
Coastal Bermuda grass	350–600	390–670	Red pine	100	110
Kentucky bluegrass	175–240	195–270	White spruce	200	225
Quack grass	210–250	235–280	Pioneer succession	200	225
Orchard grass	220–310	250–350	Aspen sprouts	100	110
Reed canary grass	300–400	335–450	Southern forests		
Ryegrass	160–250	180–280	Mixed hardwoods	250	280
Sweet clover[b]	155	175	Loblolly pine	200–250	225–280
Tall fescue	130–290	145–325	Lake states forest		
Field crop			Mixed hardwoods	100	110
Barley	110	120	Hybrid poplar	140	155
Corn	155–180	175–200	Western forest		
Cotton	65–100	70–110	Hybrid poplar	270	300
Grain sorghum	120	135	Douglas fir	200	225
Potatoes	200	225			
Soybeans	220	245			
Wheat	140	155			

[a] Adapted from U.S. EPA (1981).

[b] Legume crops can fix nitrogen from the air but will take up most of their nitrogen from applied wastewater.

Table 14–30
Metals concentrations and loading rates for land application of biosolids[a]

	Ceiling concentration[b]		Cumulative pollutant loading rate[c]		Pollutant concentration for exceptional quality[d]		Annual pollutant loading rate[e]	
Pollutant	**lb/ton**	**mg/kg**	**lb/ac**	**kg/ha**	**lb/ton**	**mg/kg**	**lb/ac**	**kg/ha**
Arsenic	0.15	75	37	41	0.08	41	1.78	2.0
Cadmium	0.17	85	35	39	0.08	39	1.70	1.9
Chromium[f]	—	—	—	—	—	—	—	—
Copper	8.60	4300	1338	1500	3.00	1500	66.91	75
Lead	1.68	840	268	300	0.60	300	13.38	15

(*continued*)

Table 14–30 (*Continued*)

	Ceiling concentration[b]		Cumulative pollutant loading rate[c]		Pollutant concentration for exceptional quality[d]		Annual pollutant loading rate[e]	
Pollutant	**lb/ton**	**mg/kg**	**lb/ac**	**kg/ha**	**lb/ton**	**mg/kg**	**lb/ac**	**kg/ha**
Mercury	0.11	57	15	17	0.03	17	0.76	0.85
Molybdenum[f]	0.15	75	—	—	—	—	—	—
Nickel	0.84	420	374	420	0.84	420	18.74	21
Selenium	0.20	100	89	100	0.20	100	4.46	5.0
Zinc	15.00	7500	2498	2800	15.00	2800	124.91	140

[a] Adapted from Federal Register (1993).

[b] Dry weight basis, Table 1 from Part 503 regulations, instantaneous maximum.

[c] Dry weight basis, Table 2 from 503 regulations.

[d] Dry weight basis, Table 3 from 503 regulations, monthly average.

[e] Table 4 from 503 regulations.

[f] A February 25, 1994, Federal Register Notice deleted chromium; deleted the molybdenum values for Tables 2, 3, and 4; and raised the selenium value in Table 3 from 36 to 100.

EXAMPLE 14–4 **Metals Loadings in Land Application** A community has stockpiled biosolids in a storage lagoon. The lagoon needs to be cleaned and the biosolids disposed of to make room for a plant expansion. The metals concentrations (mg/kg) in the lagoon are as follows:

As = 45	Hg = 5
Cd = 30	Ni = 350
Cu = 1,200	Se = 15
Pb = 250	Zn = 3100

Determine if the biosolids are acceptable for land application.

Solution

1. Compare the concentrations for the above metals to the ceiling concentration (column 2) and the pollutant concentration for exceptional quality (column 4)
   a. All metals concentrations are under the ceiling limits in column 2. The biosolids are suitable for land application.
   b. Arsenic and zinc exceed the values for exceptional quality. Calculations of annual loadings are necessary.
2. Calculate the allowable annual biosolids loading rates, using Eq. (14–8), for the two metals using the annual pollutant loading rates in Table 14–30.
   a. Arsenic-based loading rate ($L_C = 2\text{ kg/ha·y}$)

$$L_S = \frac{L_C}{L_C(10^{-6})} = \frac{(2\text{ kg/ha·y})}{(45\text{ mg/kg})(1\text{ kg}/10^6\text{ mg})} = 44{,}444\text{ kg/ha·y}$$

b. Zinc based loading rate ($L_C = 140$ kg/ha·y)

$$L_S = \frac{(140\ \text{kg/ha·y})}{(3100\ \text{mg/kg})(1\ \text{kg}/10^6\ \text{mg})} = 45{,}161\ \text{kg/ha·y}$$

3. Compare the whole biosolids loading rates to determine the limiting rate. The 44,444 kg/ha·y biosolids loading based on arsenic is limiting.

Comment Nitrogen loadings typically are more limiting than metals loadings. If the nitrogen loading rate exceeds 20 Mg/ha·y (U.S. EPA, 1995), then the arsenic loading rate will determine the whole biosolids loading rate.

Application Methods

Application methods for biosolids range from direct injection of liquid biosolids to surface spreading of dewatered biosolids. The method of application selected will depend on the physical characteristics of the biosolids (liquid or dewatered), site topography, and the type of vegetation present (annual field crops, existing forage crops, trees, or preplanted land).

Liquid or Thickened Biosolids Application. Application of biosolids in the liquid or thickened state is attractive because of its simplicity. Dewatering processes are not required, and the liquid or thickened biosolids can be transferred by pumping. Typical solids concentrations of liquid or thickened biosolids applied to land range from 1 to 8 percent. Liquid or thickened biosolids may be applied to land by vehicular application or by irrigation methods similar to those used for wastewater distribution.

Vehicular application may be by surface distribution, subsurface injection or incorporation. Limitations to vehicular application include limited tractability on wet soil and potential reduction in crop yields due to soil compaction from truck traffic. Use of high-flotation tires can minimize these problems.

Surface distribution may be accomplished by tank truck or tank wagon equipped with rear-mounted spreading manifolds or by tank trucks mounted with high-capacity spray nozzles or guns. Specially designed, all-terrain biosolids application vehicles with spray guns are ideally suited for biosolids application on forest lands. Vehicular surface application is the most common method used for field and forage croplands. The procedure used commonly for annual crops is to (1) spread biosolids prior to planting, (2) allow the biosolids to dry partially, and (3) incorporate the biosolids by disking or plowing. The process is repeated then after harvest.

Liquid biosolids can be injected below the soil surface by using tank wagon or tank trucks with injection shanks or incorporated immediately after surface application by using plows or disks equipped with biosolids distribution manifolds and covering spoons (see Fig. 14–35). Advantages of injection or immediate incorporation methods include minimization of potential odors and vector attraction, minimization of ammonia loss due to volatilization, elimination of surface runoff, and minimum visibility leading to better public acceptance. Injection shanks and plows are very disruptive to perennial forage crops or pastures. To minimize such effects, special grassland biosolids injectors have been developed (Crites and Tchobanoglous, 1998).

Figure 14–35

Land application of liquid sludge: (a) self contained vehicle used to haul and to inject the liquid sludge into the ground. Self-contained vehicles of the type shown are used for relatively small amounts of liquid sludge and (b) tractor equipped with subsurface liquid sludge injection tines. The liquid sludge to be injected is supplied by a hose connected to the injection device. The tethered sludge supply hose is dragged along by the tractor. The injected liquid sludge is disked in the ground using the tractor and disk such as shown on Fig. 14–36(b).

(a) (b)

Irrigation methods include sprinkling and furrow irrigation. Typically, large-diameter, high-capacity sprinkler guns are used to avoid clogging problems. Sprinkling has been used mainly for application to forested lands and occasionally for application to dedicated disposal sites that are relatively isolated from public view and access. Sprinklers can operate satisfactorily on land too rough or wet for tank trucks or injection equipment and can be used throughout the growing season. Disadvantages to sprinkling include power costs of high-pressure pumps, contact of biosolids with all parts of the crop, possible foliage damage to sensitive crops, potential odors and vector attraction problems, and potentially high visibility to the public.

Furrow irrigation can be used to apply biosolids to row crops during the growing season. Disadvantages associated with furrow irrigation are localized settling of solids and the potential for ponding of biosolids in the furrows, both of which can result in odor problems.

Dewatered Biosolids Application. Application of dewatered biosolids to the land is similar to an application of semisolid animal manure. The use of conventional manure spreaders is an important advantage because farmers can apply biosolids on their lands with their own equipment. Typical solids concentrations of dewatered biosolids applied to land range from 20 to 30 percent. Dewatered biosolids are spread most commonly using tractor-mounted box spreaders or manure spreaders followed by plowing or disking into the soil (see Fig. 14–36). For high application rates bulldozers, loaders, or graders may be used. For forest application, a side-slinging vehicle has been tested that can apply dewatered biosolids up to 60 m (200 ft) (Leonard et al., 1992).

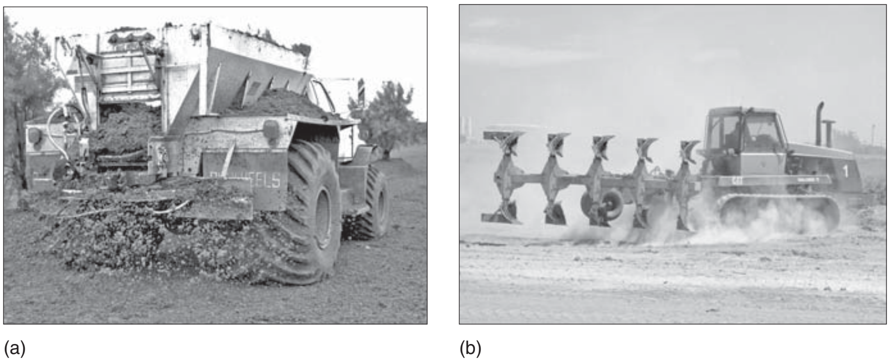


Figure 14–36

Land application of dewatered sludge: (a) typical example of vehicle used to apply dewatered sludge on the surface of the soil and (b) typical tractor and rotating two-way disk used to disk dewatered and/or liquid sludge into the ground.

(a) (b)

Table 14–31

Criteria for dedicated land disposal (DLD) sites for biosolids[a]

Parameter	Unacceptable condition	Ideal condition
Slope	Deep gullies, slope $>12\%$	$<3\%$
Soil permeability	$>1 \times 10^5$ cm/s[b]	$\leq 10^{-7}$ cm/s[c]
Soil depth	<0.6 m (2 ft)	>3 m (10 ft)
Distance to surface water	<90 m (300 ft) to any pond or lake used for recreational or livestock purposes, or any surface water body officially classified under state law	>300 m (1000 ft) from any surface water
Depth to groundwater	<3 m (10 ft) to groundwater table (wells tapping shallow aquifers)[d]	>15 m (50 ft)
Supply wells	Within 300 m (1000 ft) radius	No wells within 600 m (2000 ft)

[a] From U.S. EPA (1983).

[b] Permeable soil can be used for DLD if appropriate engineering design preventing DLD leachate from reaching the groundwater is feasible.

[c] When low-permeability soils are at or too close to the surface, liquid disposal operations can be hindered due to water ponding.

[d] If an exempted aquifer underlies the site, poor quality leachate may be permitted to enter groundwater.

Application to Dedicated Lands

Disturbed land reclamation and dedicated land disposal are two types of high-rate land application. Disturbed land reclamation consists of a one-time application of 110 to 220 Mg/ha (50 to 100 dry tons/ac) to correct adverse soil conditions. Lack of soil fertility and poor physical properties can be corrected by biosolids application to allow revegetation programs to proceed. For disturbed land reclamation to be the sole avenue for biosolids reuse, a large area of disturbed land must be available on an ongoing basis. Dedicated land disposal requires a site where high rates of biosolids application are acceptable environmentally on a continuing basis. Biosolids for a dedicated land disposal (DLD) operation should meet at least Class B requirements.

Site Selection. Siting criteria for a dedicated land disposal (DLD) site are presented in Table 14–31. Major issues in DLD siting are nitrogen control and the avoidance of groundwater contamination. Groundwater contamination can be avoided by (1) locating sites remote from useful aquifers, (2) intercepting of leachate, and (3) constructing an impervious geological barrier. Low percolation rates and deep aquifers will substantially reduce or eliminate potential contamination effects.

Where groundwater nondegradation restrictions apply, it has been found that for most DLD sites it is less costly to excavate the site entirely, install a geomembrane liner, and replace the excavated material, than to dispose of the sludge by some other means (e.g., dewatering and landfilling). The limited amount, if any, of leachate collected from the liner is returned to the treatment plant for processing.

Loading Rates. Annual biosolids loading rates have ranged from 12 to 2250 tonne/ha (5 to 1000 tons/ac). The higher rates have been associated with sites that

- Receive dewatered biosolids
- Mechanically incorporate the biosolids into the soil

- Have relatively low precipitation
- Have no leachate problems because of site conditions or project design

Design loading rates for DLD can be estimated using Eq. (14–10).

$$L_S = \frac{E(TS)F}{100} - TS \qquad (14\text{–}10)$$

where L_S = annual biosolids loading rate, Mg/ha (ton/ac)
E = net evaporation rate from soil, mm/y (in./y)
TS = total solids content, percent by weight
F = conversion factor, 10 Mg/mm (113.3 ton/in.)

The net soil evaporation can be estimated from Eq. (14–11).

$$E = (f)E_L - P \qquad (14\text{–}11)$$

where E = net evaporation rate from soil, mm/y (in./y)
$f = 0.7$
E_L = pan evaporation rate, mm/y (in./y)
P = annual precipitation, mm/y (in./y)

It should be noted that infiltration into the soil is not considered in Eq. (14–10). If infiltration is allowed, the term E should be increased by the annual infiltration rate in mm/y.

Once the annual loading rate is calculated, the field area can be determined using Eq. (14–9) (dividing the biosolids production by the loading rate). Other area requirements include buffer zones, surface runoff control, roads, and supporting facilities.

Landfilling

Landfilling of biosolids in a monofill is covered under 40 CFR Part 503. Landfilling of biosolids in a sanitary landfill with municipal solid waste is regulated by the U.S. EPA under 40 CFR 258. If an acceptable site is convenient, landfilling can be used for disposal of biosolids, grit, screenings, and other solids. Stabilization may be required depending on state or local regulations. Dewatering of biosolids is usually required to reduce the volume to be transported and to control the generation of leachate from the landfill. In many cases, solids concentration is an important factor in determining the acceptability of biosolids in landfills. The sanitary landfill method is most suitable if it is also used for disposal of the other types of solid wastes. In a true sanitary landfill, the wastes are deposited in a designated area, compacted in place with a tractor or roller, and covered with a 350 mm (14 in.) layer of clean soil. With daily coverage of the newly deposited wastes, nuisance conditions, such as odors and flies, are minimized.

PROBLEMS AND DISCUSSION TOPICS

14–1 A wastewater treatment plant is producing 55,000 L/d of thickened biosolids containing 2.8 percent solids. A belt-filter press installation is to be designed based on a normal operation of 8 h/d and 5 d/wk, a belt-filter press loading rate of 280 kg/m·h, and the following data. Compute the number and size of the belt filters, and the expected solids capture, in percent. Determine the daily hours of operation required if a sustained 5-d peak solids load occurs.

1. Total solids in dewatered sludge = 26 percent.
2. Total suspended solids concentration in filtrate = 800 mg/L.

3. Washwater flowrate = 90 L/min per m of belt width.
4. Specific gravities of sludge feed, dewatered cake, and filtrate are 1.02, 1.08, and 1.01, respectively.
5. Use Fig. 3–14 to estimate the peaking factor for a sustained 5-d peak solids load

14–2 The ultimate elemental analysis of a dried sludge yields the following data:

Element	Percent
Carbon	52.1
Hydrogen	2.7
Oxygen	38.3
Nitrogen	6.9
Total	100.0

How many kg of air will be required per kg of sludge for its complete oxidation?

14–3 Compute the fuel value of the sludge from a primary settling tank having a composition (by weight) of 64.5 percent carbon, 8.5 percent hydrogen, 21.0 percent oxygen, and 4 percent sulfur.

14–4 Calculate the theoretical heat requirement to dry 1000 m^3/d of sludge cake from solid contents of 15, 20 or 25 percent (to be selected by instructor) to 92 percent. Assume feed sludge temperature at 20°C and final temperature at 100°C. Latent heat of evaporation is 2260 kJ/kg H_2O. Assume a specific gravity of the dewatered cake is 1.05. Assume dry solids heat capacity of 1.5 kJ/kg·°C. Estimate the fuel requirement assuming 5 percent heat loss and 85 percent heater efficiency.

14–5 A community of 25,000 persons has asked you to serve as a consultant on their sludge disposal problems. Specifically, you have been asked to determine if it is feasible to compost waste activated sludge with the community's solid waste. If this plan is not feasible, you have been asked to recommend a feasible solution. Currently the waste sludge from the WWTP is dewatered on a belt filter press. Assume the following data are applicable:

Solid waste data:
  Waste production = 2 kg/person·d (wet basis)
  Compostable fraction = 55%
  Moisture content of compostable fraction = 22%

Sludge production:
  Net sludge production = 0.12 kg/person·d (dry basis)
  Concentration of sludge out of the belt filter press = 22%
  Specific gravity of dewatered cake = 1.05

Compost:
  Final moisture content of composted biosolids/solid waste mixture = 55%

14–6 A municipality with a population of 200,000 has hired you as a consultant to investigate alternative mechanical sludge dewatering options. The three alternatives to be investigated are belt press dewatering, centrifugation, and pressure filter press dewatering. The biosolids to be dewatered are stabilized by anaerobic digestion, are a mixture of primary sludge and waste activated sludge, and have a solids concentration of 5 percent. Ultimate disposal is by landfilling at a site located 50 kilometers from the treatment plant. Compare the various dewatering alternatives and recommend one. State the reasons for your recommendation. Should newer technologies such as rotary presses or screw presses be considered?

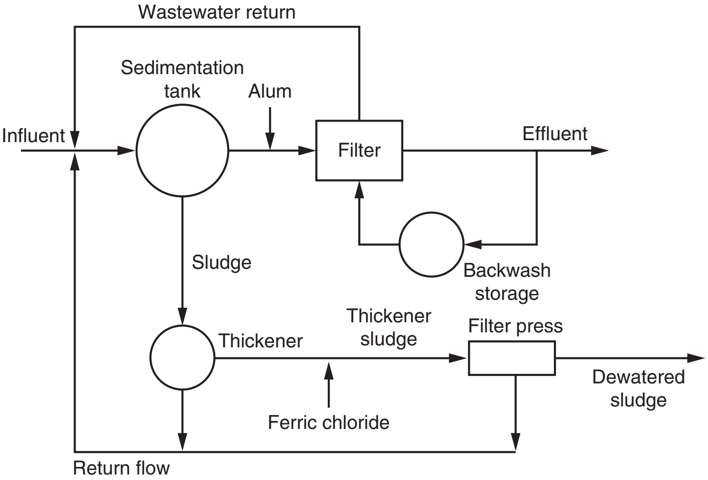
14–7 Prepare a solids balance for the peak loading condition for the treatment plant used in Example 14–3 for one of the data sets to be selected by instructor. Enter your final values on the solids balance figure in Example 14–3.

	Data set			
	1	2	3	4
Peak flowrate, m^3/d	54,000	60,000	50,000	54,000
Average BOD at peak flowrate, mg/L	340	300	350	300
Average TSS at peak flowrate, mg/L	350	320	330	320
TSS after grit removal, mg/L	325	300	310	300

Use data given in Example 14–3 for other parameters.

14–8 Prepare a solids balance, using the iterative technique delineated in Example 14–3, for the following treatment process flow diagram and one of the following data sets to be selected by instructor. Also determine the effluent flowrate and suspended solids concentration.

	Data set			
	1	2	3	4
Influent characteristics				
Flowrate, m^3/s	10,000	20,000	30,000	40,000
Suspended solids, mg/L	1000	350	400	300
Sedimentation tank				
TSS removal efficiency, %	75	60	65	60
Underflow TSS concentration, %	7	6.5	6	5.5
Specific gravity of sludge	1.1	1.1	1.1	1.1
Alum addition				
Dosage, mg/L	10	10	20	15
Chemical solution, kg alum/L of solution	0.5	0.5	0.5	0.5
Filters				
TSS removal efficiency, %	90	90	95	92
Washwater solids concentration, %	6	6	6.8	6.5
Specific gravity of backwash	1.08	1.08	1.089	1.085
Thickener				
Supernatant TSS, mg/L	400	300	200	250
Concentration of solids in underflow, %	12	8	9	8
Chemical addition				
Dosage, percent of underflow solids from thickener	0.8	1.0	1.0	1.0
Chemical solution, kg/L of solution	2.0	2.0	2.0	2.0
Filter press				
TSS concentration in filtrate, mg/L	200	300	250	200
Concentration of dewatered solids, %	40	38	42	42
Specific gravity of sludge cake	1.6	1.5	1.65	1.65



In preparing the solids balance assume that all of the unit operations respond linearly such that the removal efficiency for recycled solids is the same as that for the solids in the influent wastewater. Also assume that the distribution of the chemicals added to improve the performance of the filter and filter press is proportional to the total solids in the return flows and the effluent solids.

14–9 Assume the HHV value of biosolids as derived in Example 14–2, and compute the energy contents in biosolids with 25, 65 and 90 percent solid contents (to be selected by instructor). Compare the results with the heat requirements to vaporize water in the biosolids for each solids content level. Discuss supplemental energy requirements for the incineration process in relation to the solids contents of the feed biosolids.

14–10 Determine the dry sludge application rate for Reed canary grass on the basis of satisfying crop nitrogen uptake. Assume that biosolids containing 3 percent organic nitrogen by weight are applied to a soil that has an initial nitrogen content of zero. Use a mineralization rate of 30 percent for the first year, 15 percent for the second year, and 5 percent for the third and subsequent years.

14–11 Biosolids containing 50 ppm of cadmium on a dry basis are to be applied to land. If the limiting mass loading to the soil is set at 10 kg/ha, what would be the safe loading rate for 50 y of application?

14–12 Compare the advantages and disadvantages of land application of liquid biosolids to dewatered biosolids. Assume the land application site is located 15 km from the treatment plant in primarily an agricultural area and the biosolids are transported by truck. The biosolids are stabilized by anaerobic digestion and the liquid biosolids concentration is 6 percent and the dewatered solids concentration is 25 percent.

14–13 In problem 14–13, what would be the advantages and limitations of conveying the liquid biosolids by pipeline to the land application site? What types of facilities, i.e., structures, equipment, and vehicles, would be required and what are the operating and maintenance considerations?

The following two problems are best solved in groups of two or three.

14–14 A treatment plant is treating an average wastewater flowrate of 500,000 m^3·d with 430 mg/L TSS, 345 mg/L VSS, and 335 mg/L BOD. The volatile part of TSS is 80 percent. The peaking factors are 1.3, 1.4, and 1.5 for max month, max week, and max day, respectively. The main treatment processes include primary clarifiers and activated sludge system for secondary treatment. Primary sludge is thickened via gravity thickener while waste activated sludge is thickened by gravity belt thickening. Combined thickened sludges are stabilized via anaerobic digestion. The primary clarifier is expected to remove 50 percent of the TSS and 30 percent of the BOD. Students are encouraged to solve this problem through establishing a spreadsheet mass and energy balance.

1. Determine the average amount of primary sludge produced in kg/d at a TS concentration of 1.5 percent.
2. Determine the average amount of WAS produced in kg/d assuming the BOD effluent from the secondary clarifier to be 10 mg/L and a sludge yield of 0.6. Assume the mean cell residence time or SRT of the secondary treatment is 10 d; endogenous decay rate, k_d, is 0.06 d^{-1}; the VS content in the mixed liquor suspended solids is 70 percent.
3. Determine the average amount of primary thickened sludge in kg/d at a TS concentration of 5 percent and assuming a solids capture of 95 percent in the gravity thickener.
4. Determine the average amount thickened WAS in kg/d at a solids concentration of 6 percent assuming a 95 percent capture rate in the gravity belt thickener. Assume the WAS from the secondary clarifier is 1 percent TS.
5. The two thickened sludges are mixed in a combined primary and activated sludge (CPAS) tank with a volume of 40 m^3. Determine the combined sludges percent TS content out of the CPAS, and percent volatile solids content out of the CPAS. Determine the HRT of the CPAS tank in min.
6. The anaerobic digestion complex achieves 50 percent volatile solids destruction. Digesters are cylindrical with cone bottom. The dimensions of each digester are 37 m diameter, 11 m side water depth, 1.2 m freeboard, and cone bottom slope of 1:10. Assuming all sludge filled digester volume is usable:
   a. Determine the number of digesters if the HRT of all digesters is not reduced below 15 d HRT to meet max month per MOP 8 design criteria. Assume the digesters are high rate complete mix with no decant.
   b. Determine the number of digesters to process annual average sludge production.
   c. Determine the annual average and max month VS loading to the digester complex in $kg/m^3 \cdot d$ and compare it to typical design criteria
   d. Determine the amount of heat needed to maintain the mesophilic temperature within the digestion complex during the winter months assuming minimum income sludge temp to be 5°C and assuming no heat loss from the digesters.
   e. If you are the operations manager do you recommend processing the maximum day sludge in the digester complex? If not what do you recommend an operation strategy?
7. Determine the amount of biogas production in m^3/h assuming 0.95 m^3/kg VS destroyed.
8. The plant produces electricity from the biogas system through an internal combustion engine of 38 percent electrical efficiency and 40 percent thermal energy recovery.
   a. Determine the amount of electricity produced in KW if the generated biogas has 22,400 kJ/m^3.
   b. Is the recovered heat from the engine enough to supplement the energy used to keep the digesters heated during winter months?
9. If a portion of the biogas is used to supplement the energy needed for the digesters' heating through a boiler system with 80 percent energy efficiency prior to feeding the engine, determine the amount of electricity generated.
10. Provide your thoughts on which energy management system through comparing results from 8 and 9 above.

11. The biosolids produced from digestion process are dewatered via centrifuges. Determine the following:
    a. Dry solids loading to the centrifuge complex following digestion in kg/d
    b. Volatile solids content of the biosolids out of digestion in kg/d
    c. The percent solids total solids (%TS) content of the biosolids after digestion
    d. The volumetric flowrate of the biosolids in m^3/h going to the centrifuges.
    e. Determine the amount of wet tonnes produced daily from the centrifuge and the concentration (%TS) and the centrate flowrate (m^3/h) if the centrifuge complex is producing a solids cake of 23 percent with a 95 percent solids recovery. Assume the BOD concentration in the centrate is 1500 mg/L and is recycled to the primary clarifier.

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15

Plant Recycle Flow Treatment and Nutrient Recovery

WORKING TERMINOLOGY

Term	**Definition**
Acid absorption	The absorption of constituents from an air stream into an acid solution.
Air stripping	The use of air to remove (transfer) constituents from a liquid stream to an air stream.
Anaerobic ammonium oxidation (Anammox)	Biological oxidation of ammonia under an oxygen-free environment with nitrite as the electron acceptor.
AOB	Ammonia oxidizing bacteria.
Bioaugmentation	The introduction of a selected or engineered group of microorganisms to enhance treatment.
Crystallizers	A liquid-solid separation technique in which solid crystals are formed from constituents in solution.
Deammonification	A two-step biological process consisting of partial nitritation and the Anammox reaction.
Denitritation	Biological conversion of nitrite to nitrogen gas and other intermediate compounds containing nitrogen.
Denitratation	Biological conversion of nitrate to nitrite.
Fermentation	A biological process in which organic matter is converted under anaerobic conditions to volatile fatty acids.
Integrated sidestream-mainstream treatment	An integrated system is any process where the sidestream reactor waste solids are fed to the mainstream secondary process or mixed liquor solids are interchanged between the two processes.
Limit of technology	Represents the lowest achievable concentration of a specific constituent by the best available technology.
Nitratation	The conversion of nitrite to nitrate.
Nitritation	The conversion of ammonia to nitrite.
NOB	Nitrite oxidizing bacteria.
Sidestream	The collective term used to describe all recycle streams. Sidestreams derived from digestion are also known in some countries as liquor or reject water.

Term	Definition
Separate sidestream treatment process	Biological treatment process that is isolated from the mainstream treatment process and dedicated to treat sidestream flows.
Struvite	A precipitated compound form of magnesium ammonium phosphate hexahydrate, $MgNH_4PO_4{\cdot}6H_2O$.
Steam stripping	The use of steam to remove a constituent from a waste stream.

The separation of water from primary, secondary, combined, or digested sludges during solids processing generates a liquid stream, which has characteristics that prevent direct discharge of the stream with the wastewater treatment plant final effluent. At facilities that thicken primary and secondary waste sludges before aerobic or anaerobic digestion and dewater the digested solids, multiple recycle streams are generated, each with a different composition, flowrate, and impact on the treatment plant. Because anaerobic and aerobic digestion result in the release of soluble organic nitrogen-containing compounds, ammonium and orthophosphate into the bulk liquid, the post-digestion recycle stream generated by the dewatering of the digested solids will have elevated nutrient concentrations resulting in an increased nutrient loading to the primary and secondary treatment processes. In addition, the release of orthophosphate and ammonium during solids digestion often results in the formation of insoluble inorganic compounds such as magnesium ammonium phosphate, also known as struvite, which can cause operational and maintenance problems in mechanical dewatering equipment and pipes that convey the recycle stream.

Current practice at most wastewater treatment plants is to recycle these sidestreams to the head of the plant or directly to the secondary process for treatment. However, because these sidestreams can impact significantly the performance of the secondary treatment process, many treatment plants now treat these streams separately. Interest in reducing nutrient loadings from plant recycle streams rich in ammonium and phosphate through the implementation of dedicated or *sidestream* treatment processes has been increasing since the late 1980s due to more stringent effluent discharge limits on ammonium-N, total nitrogen (TN) and total phosphorus (TP) and a desire to reduce plant operating costs (energy, chemicals, and maintenance). In addition, the impact of suspended solids and colloidal material present in some recycle streams on treatment plant effluent quality must be considered.

The types of recycle streams commonly found in conventional wastewater treatment plants, their characteristics, and their potential impacts on the operation of wastewater treatment facilities are reviewed in this chapter. Following the review of types of recycle streams and their characteristics, the focus of the remainder of the chapter is on physiochemical and biological treatment technologies that have been developed and implemented to treat separately the nutrient-rich streams, which are typically the primary recycle streams of concern for nutrient removal facilities. Typical process design and performance information are also provided.

15–1 SIDESTREAM IDENTIFICATION AND CHARACTERIZATION

Recycle sidestreams from the thickening and dewatering of raw and digested solids are typically identified with the particular process or mechanical equipment from which they originate. The composition of each recycle sidestream also varies depending on the source of the solids being thickened or dewatered. For the purpose of simplifying terminology, all recycle streams discussed in this chapter will be called *sidestream*, but specific terminology that allows for the distinction of one type of recycle stream from another will be retained as needed and used interchangeably with the general name. Common sources of sidestreams, typical flowrates, and typical sidestream characteristics [total suspended solids (TSS), total Kjeldahl nitrogen (TKN), ammonium nitrogen (ammonium-N), total phosphorus (TP) and orthophosphate (ortho-P)] are summarized in Table 15–1. Additional characteristics that distinguish one type of sidestream from another are discussed below.

Sidestreams Derived from Primary and Secondary Sludges

The soluble nutrient concentrations in sidestreams generated from the thickening of primary and waste activated sludges (gravity thickening; dissolved air flotation; centrifugation) and the filtration of secondary effluent generally represent the soluble composition of the liquids from which they originate. The partial fermentation of sludges that often occurs in gravity thickeners, will generate volatile fatty acids, decrease the alkalinity, and increase the ammonium and phosphate concentrations. The wide ranges for the ammonium and orthophosphate concentrations shown in Table 15–1 for these sidestreams represent the typical range of weak to untreated domestic wastewaters (Table 3–18) and nitrifying and non-nitrifying secondary treatment systems. Concentrations of TKN, TP, and BOD are strongly dependent on the TSS concentration. In general, the daily mass nutrient loadings contributed by these sidestreams to the secondary process are relatively minor in comparison to the raw influent or primary tank effluent. The major impact of their return to the primary and/or secondary treatment process is on the solids mass balance for the treatment plant, as discussed in Sec. 14–7 in Chap. 14. The solids impact is dependent on the solids capture efficiency of the thickening or dewatering process. For most treatment facilities, sidestream flowrates from the thickening of primary and secondary sludges and backwash from final filtration are continuous or near-continuous.

Sidestreams Derived from Fermented Primary and Digested Primary and Secondary Sludges

Sidestream characteristics resulting from the fermentation of primary sludge and the digestion of primary and waste activated sludges are significantly different than sidestreams generated from the thickening and dewatering of primary and waste activated sludges. The principal reason is the release of organic nitrogen-containing compounds, ammonium and phosphate into the bulk liquid. The soluble nutrient concentrations in these higher strength sidestreams are, as indicated in Table 15–1, dependent on the process from which they originate. For example, in the fermentation of primary sludge to generate volatile fatty acids (VFAs) for the enhancement of biological phosphorus removal, the ammonium and orthophosphate concentrations in the fermentate are far lower than the concentrations in the sidestreams resulting from aerobic and anaerobic digestion. Further, the ammonium and orthophosphate concentrations resulting from digestion are dependent on the feed solids concentrations to the digester and the volatile solids destruction efficiency, as reflected in the broad concentration ranges shown in Table 15–1. For example, if thermal hydrolysis is applied in the pretreatment of sludge before anaerobic digestion, the total solids (TS) concentration in the feed to the digesters will be 8 to 11 percent by weight. With the higher TS concentration and enhanced volatile solids destruction efficiency in the digester associated with this advanced digestion process, the digester ammonium-N concentration will be two to three times the concentration observed in typical conventional anaerobic digesters.

Table 15–1

Characteristics of various sidestreams from the thickening, stabilization and dewatering of raw and digested solids

Operation	Flowrate, Percent of plant influent	Value, mg/L						
		TSS[a]	BOD[a]	TKN	NH4-N	NOx	TP	Ortho-P
Gravity thickening supernatant:								
Primary sludge	2–3	80–350	100–400	19–70	12–45	0	4–11	3–8[e]
Primary sludge + waste activated sludge	3–5	100–350	60–400	20–70	8–45	0–8	4–15[b]	2–7[b,e]
Primary sludge fermentate, including elutriation water	3–4	700–900	2000–2500	80–120	60–100	0	10–20	5–15
Flotation thickening subnatant (waste activated sludge)	0.7–1	100–2500	50–1200	8–250	0–45	0–30	2–50	0.05–8
Centrifuge thickening centrate (waste activated sludge)	0.7–1	500–3000	170–3000	40–280	0–45	0–30	8–60	0.05–8
Screw press–filtrate + pressate (alkaline and heat stabilization for Class A)[d]	0.3–0.5	400–500	600–1300	120–250	10–20	0–5	6–14	< 1
Aerobic digestion supernatant (mesophilic; continuous and intermittent aeration)	0.1–0.5	100–10,000	100–1700	100–1200	20–400	0–400	200–350[b,c]	200[b,c]
Anaerobic digestion supernatant (two-stage, high-rate)	0.1–0.5	1000–11,500	500–5000	850–1800	800–1300	0	110–470[b,c]	100–350[b,c]
Centrifuge dewatering centrate:								
Two-stage, high rate anaerobic digestion	0.5–1	200–20,000	100–2000	810–2100	800–1300	0	100–550[b,c]	100–350[b,c]
Thermal hydrolysis + single stage mesophilic anaerobic digestion	0.2–0.5	1500–10,000	1500–3000	2200–3700	2000–3000	0	220–800[b,c]	200–700[b,c]
Belt-filter press filtrate: two-stage, high rate anaerobic digestion, including belt washwater	1–2	100–2000	50–500	410–730	400–650	0	50–200[b,c]	50–180[b,c]
Recessed-plate-filter press filtrate	0.5–1	50–1000	50–250		800–1300	0		100–350[b,c]
Sludge lagoon supernatant		5–200	100–200					
Sludge drying bed underdrainage	0.3–0.5	20–500	100–200		0–400	0–400	2–210	2–200
Composting leachate		500	2000					

(continued)

Table 15–1 (*Continued*)

Operation	Flowrate, Percent of plant influent	Value, mg/L						
		TSS[a]	BOD[a]	TKN	NH4-N	NOx	TP	Ortho-P
Incinerator scrubber water		600–8000	30–80					
Depth filter washwater		100–1000	50–500					
Microscreen washwater		240–1000	100–500					
Carbon adsorber washwater		100–1000	50–400					
Dryer condensate								

[a] Adapted, in part, from U.S. EPA (1987b) and WEF (1998).

[b] Orthophosphate concentration does not include potential phosphorus release from waste activated sludges derived from plants operating with biological phosphorus removal.

[c] Orthophosphate concentration does not include reduction by chemical precipitation within the digester through the natural formation of salts such as hydroxyapatite and magnesium ammonium phosphate (struvite), through the addition of ferric or ferrous salts to the digester to control struvite formation in dewatering equipment and pipes conveying sidestream or for facilities that practice chemically enhanced primary or secondary treatment for phosphorus removal.

[d] Based on unpublished data collected from the stabilization and dewatering of waste activated sludge with the FKC Class A process at the Sequin, WA, wastewater treatment plant in July-August, 2011 (no primary sedimentation; oxidation ditch secondary process).

[e] Orthophosphate concentration does not reflect facilities that practice chemically enhanced primary treatment.

Nitrogen Content. In addition to a higher ammonium concentration, soluble organic nitrogen-containing compounds are released into the bulk liquid during digestion, accounting for roughly 10 percent of the soluble TKN. Of this soluble organic nitrogen fraction, approximately 50 percent is considered essentially non-biodegradable or recalcitrant dissolved organic nitrogen (rDON). Recalcitrant DON is typically less than 1 mg/L in plant effluents due to rDON in the raw influent and the production of soluble microbial products in the secondary treatment process. In general, the presence of rDON is not of concern for most nutrient removal plants. However, for limit of technology (LOT) plants with a low effluent TN limit (e.g., 3 mg/L), rDON becomes a larger fraction of the plant effluent TN. Typical digester sidestream will add approximately 0.2 mg N/L of rDON to the plant effluent, assuming no removal by adsorption onto the activated sludge in the secondary process or capture in primary tanks where chemically enhanced clarification is being performed. The concentration of rDON in digester sidestream will be higher if thermal hydrolysis is applied for sludge pretreatment before digestion; the concentration being dependent on the hydrolysis reactor temperature (Dwyer et al., 2008).

Phosphorus Content. The orthophosphate concentration range, as noted in Table 15–1, for the various digestion sidestreams does not include phosphate released by waste activated sludge from biological phosphorus removal processes, nor do the ranges account for reduction in phosphate due to precipitation within digester or in downstream mechanical equipment and pipes conveying sidestream. The values shown in Table 15–1 represent the stoichiometric amounts of phosphorus expected to be released into the bulk liquid for the various operating and performance conditions considered. For example, at facilities performing chemically enhanced primary treatment with ferric salts and anaerobically digesting the combined primary and waste activated sludges, the orthophosphate concentration in digester sidestream will typically be below 10 mg P/L due to precipitation or absorption onto iron floc in the digester. For facilities where ferric or ferrous salts are added directly to the anaerobic digester for control of hydrogen sulfide, partial removal of orthophosphate would also be expected. At facilities where no iron salts are added, partial precipitation of released phosphate as magnesium ammonium phosphate (struvite) at the digester outlet structure, in the biosolids dewatering process and in the pipes that convey digester sidestream is common. The amount that is formed is dictated by the ion molar concentrations of the struvite constituents and the elevated pH induced by CO_2 release from the bulk liquid at these locations.

Alkalinity Content. The alkalinity in sidestreams will vary depending on the source. The concentration of alkalinity in a sidestream is of importance in the operation of separate sidestream treatment processes. In sidestreams generated from fully aerated conventional low rate aerobic digesters, the alkalinity concentration is anticipated to be low due to the acidification of the digester sludge by nitrification. For aerobic digesters that operate with intermittent aeration to allow denitrification, the residual alkalinity in the digester will be slightly higher. In contrast, anaerobic digestion and autothermal thermophilic aerobic digestion (ATAD) sidestreams typically contain higher alkalinity concentrations, primarily in the form of bicarbonate. The cause of the high alkalinity concentrations is due to the retention of carbon dioxide in the digester bulk liquid to balance the positively charged ammonium ion at the typical pH range of the digesters (7.2 to 7.8). Therefore, the bicarbonate and ammonium-N concentrations will be equal on a molar basis. In terms of the standard measurement as $CaCO_3$, the alkalinity to ammonium-N mass ratio is 3.5 to 1 (kg $CaCO_3$/kg N).

Total Suspended Solids Content. Total suspended solids in digester sidestreams are composed largely of stabilized biologically inert solids with a relatively low volatile solids content (e.g., 65 percent). If the digester sidestream is sent to the primary sedimentation tanks, the TSS will settle readily and will be recycled to the solids processing train. If the sidestream is sent directly to the secondary treatment process, the solids will accumulate in the activated sludge. If the dewatering equipment is not working well, resulting in poor capture efficiency, the inert solids loading to the main plant can be substantial, potentially inducing foaming in the secondary process and effectively reducing the active fraction of the sludge in the secondary system.

Colloidal Material Content. Depending on the secondary treatment system configuration, digestion sidestreams, which can also contain colloidal material, can impact plant effluent quality. In addition to the potential for inorganic precipitate formation (e.g., struvite, hydroxyapatite), the presence of the colloidal material, can make the use of membrane-based technologies for separate treatment of digestion sidestream difficult due to membrane fouling, unless coagulation and filtration of these materials precedes the membrane process.

Temperature. Sidestream temperatures will vary depending on the source. The sidestream resulting from centrifugation of anaerobically-digested solids will typically have a temperature near the digester temperature (e.g., 30 to 35°C), depending on heat losses from a digested sludge holding tank and from the dewatering equipment. If a belt filter press is used for dewatering the same digested sludge, the temperature will typically be lower (e.g., 20 to 30°C) due to the inclusion of cooler belt washwater in the sidestream, unless warm filtrate is recycled and used as the washwater source. For aerobic digestion sidestreams, the temperature is also dependent on the type of process and the operating conditions. For example, a sidestream derived from ATAD process would have a higher temperature than a conventional non-insulated aerobic digester operating at a long hydraulic retention time. For the anaerobic sludge lagoons, the supernatant will be near ambient temperatures. The importance of temperature for the design of separate sidestream treatment processes is discussed in the individual sections dealing with physiochemical and biological treatment processes.

Flowrate. Sidestream flowrates from digestion processes, as reported in Table 15–1, are typically less than 1 percent of the daily average raw influent flow based on a continuous 7 d/wk operation. However, at facilities where digested sludge is not dewatered every day nor operate the dewatering equipment continuously (i.e., 24 h/d), the sidestream flow can result in a high instantaneous flow to the primary and secondary treatment processes if discharged directly without equalization. Estimation of peak sidestream flow is illustrated in Example 15–1.

EXAMPLE 15–1 **Estimate Peak Sidestream Flow** The sidestream from an anaerobic digester contributes 0.7 percent of the daily average influent flow of 0.5 m^3/s (11.4 Mgal/d), on a continuous basis, and is generated 5 d/wk, 8 h/d. What would be the instantaneous flow to the primary tanks? What percentage of the average influent flow would the instantaneous flow represent?

Solution

1. Determine the sidestream flowrate (SSF).

$$\text{SSF} = (0.5\ \text{m}^3/\text{s})(0.007) = 0.0035\ \text{m}^3/\text{s}$$

2. Determine the peak sidestream flowrate.

$$\begin{aligned}\text{Peak SSF} &= (0.0035\,\text{m}^3/\text{s})[(7\,\text{d/wk})/(5\,\text{d/wk})][(24\,\text{h/d})/(8\,\text{h/d})]\\ &= 0.0147\,\text{m}^3/\text{s}\end{aligned}$$

3. Determine the percentage of the flow represented by the instantaneous flow.

$$\text{Percent of total flow} = [(0.0147\ \text{m}^3/\text{s})/(0.50\ \text{m}^3/\text{s})](100) = 2.94$$

Comment Although the facility likely has sufficient hydraulic capacity to receive this sidestream flow over the 8-h sludge dewatering period, the primary concern is the nutrient loading associated with this sidestream flow, particularly if the sidestream load occurs during the peak diurnal nutrient loading to the facility. The impact of the 8-h sidestream peak nutrient load can be dampened through flow equalization or by treating the sidestream load in a separate process to minimize the nutrient loading to the secondary process.

15–2 MITIGATING RECYCLE FLOWS AND LOADS

Recycle streams from sludge thickening, digestion, dewatering and storage processes may have negative impacts on the performance of the mainstream process. These impacts vary in degree and are specific to each facility. Summary information on the impact of recycle streams on the mainstream plant and potential mitigation measures is provided in Table 15–2.

Sidestream Pretreatment

Sidestreams will increase the solids load to the mainstream process, affecting solids inventory and the mixed liquor solids concentration., Some sidestreams, as noted previously, will also contain colloidal material which may impact plant effluent quality. Depending on the plant operating condition and performance, sidestream pretreatment to reduce its suspended solids and colloidal material content may be beneficial. The reduction of suspended solids and colloidal material in sidestreams is discussed in Sec. 15–3.

For treatment plants with stringent nutrient removal requirements, pretreatment of the nutrient-rich sidestreams prior to return to the mainstream process may be cost effective. Physiochemical and biological treatment options for these sidestreams are discussed in Sec. 15–4 through 15–11. As demonstrated in Example 15–1, the hydraulic load contributed by sidestream generated by dewatering operations only performed 7–8 h/d will increase the instantaneous hydraulic load through the mainstream facility. However, this hydraulic load increase is typically not a concern.

Equalization of Sidestream Flows and Loads

Digested sludge dewatering operations are commonly carried out as a batch operation, over several hours during the daytime hours five or six days per week. Consequently, nutrient-rich sidestream constituent loads to the mainstream plant will increase near instantaneously and will coincide with the diurnal influent load peak period. Where nitrification, denitrification, chemical phosphorus removal or biological phosphorus removal are practiced, the contribution of the sidestream nutrient load will increase peak air and chemical demands (external organic carbon, iron salts or alum), and may deteriorate plant effluent quality if control systems are not adequate to respond to the sudden and significant increase in nutrient loads. Nitrification may also be destabilized if the alkalinity concentration in the plant influent is relatively low and no supplemental alkalinity is available.

Table 15–2

Major impacts and potential mitigation measures for return flows from sludge and biosolids processing facilities[a]

Source of return flow	Impact	Process impacted	Mitigation measure
Sludge thickening	Effluent degradation by colloidal SS	Sedimentation	Add flocculent aid ahead of sedimentation tank Separately thicken primary and biological sludges Optimize gravity thickener dilution water
	Floating sludge	Sedimentation	Minimize gravity thickener detention time Remove sludge continuously and uniformly
	Odor release and septicity	Recycle point	Reduce gravity thickener detention time Return flows ahead of aerated grit chamber Provide odor containment, ventilation, and treatment (scrubber or biofilter)
		Biological	Return odorous flows to aeration tank Remove sludge continuously and uniformly Provide separate return flow treatment (with other recycle streams)
	Solids buildup	Sedimentation	Increase thickening unit operation time or capacity to maintain desired solids inventory in sedimentation units
		Biological	Remove sludge continuously and uniformly Include recycle loads in mass balance analysis
Sludge dewatering	Effluent degradation by colloidal suspended solids	Sedimentation	Optimize dewatering unit solids capture by improved sludge conditioning Add flocculent aid ahead of sedimentation tank Return centrate/filtrate to thickener Provide separate return flow treatment (with other recycle streams)
	Solids buildup	Sedimentation	Increase dewatering unit operation time or capacity to maintain desired solids inventory in sedimentation units
		Biological	Remove sludge continuously and uniformly Reduce trickling-filter recycle rate Include recycle loads in mass balance analysis
Sludge stabilization	Effluent degradation by excessive BOD load	Biological	Optimize supernant/decant removal, i.e., remove smaller amounts over a longer period of time, or reschedule removal to off-peak periods Provide separate return flow treatment Increase RBC speed Increase MLVSS in activated sludge system (decrease F:M ratio) Increase dissolved oxygen level in activated sludge process
	Effluent degradation by nutrients	Biological	Regulate digester supernatant/decant removal Thicken sludge before stabilization Provide separate return flow treatment
Washwater from depth filters	Hydraulic surges	Sedimentation	Provide backwash storage for flow equalization Schedule filter backwashing for off-peak periods

[a] Adapted, in part, from U.S. EPA (1987*b*).

To minimize these potential impacts, sidestream flow equalization can be employed. In larger facilities, where sludge is dewatered continuously, the benefit of sidestream equalization depends on the variability of the sidestream nutrient loads.

Equalization of sidestream flow is also commonly used to reduce the peak air demand in a biological pretreatment process. As discussed in Sec. 15–11, flow equalization will also have a direct impact on the sidestream reactor volume requirement. For batch treatment processes, where interruption in sidestream flow to the treatment process will occur intermittently during the treatment cycle, flow equalization is required if no release of raw sidestream to the mainstream plant is desired.

Equalization Volume Requirement. The general principles of flow and load equalization discussed in Sec. 3–7 in Chap. 3 apply to nutrient-rich sidestreams, but the basis for calculating the equalization volume requirement is slightly different. The equalization volume may be based on full equalization where sidestream is returned continuously to the mainstream plant or sent to a pretreatment process at a relatively stable, but adjustable, flowrate. The required equalization volume can be estimated using Eq. (15–1).

$$V = (N)(Q_{\mathrm{dw}})(D_{\mathrm{dw}}/7) \tag{15–1}$$

Where V = equalization tank volume, m^3
N = maximum number of consecutive days without dewatering, d
Q_{dw} = average daily sidestream volume generated, m^3/d
D_{dw} = number of days per week sludge is dewatered, d

Depending on the sidestream constituent concentrations, the alkalinity of the mainstream plant influent, and the treatment objectives of the mainstream plant, capturing sidestream in an equalization tank and returning the non-pretreated sidestream load to the mainstream plant in the off-peak hours when the facility influent flow and loads are at their minimum may be sufficient. However, for facilities that perform nitrate and phosphate removal, continuous return of the sidestream load at a steady rate is recommended. Where space constraints and tank costs are of concern, a smaller equalization tank volume may be used if the sidestream load variability does not impact plant performance. Dynamic process modeling is commonly used to assess the feasibility of using a smaller tank volume. Depending on the results of such an assessment, it may also be found that increasing the number of days each week or the hours per day that sludge is dewatered is the most cost-effective option.

Design Considerations. The features of a sidestream equalization tank vary depending on the design objectives. For an ideally equalized stream, mechanical mixing is provided to return a uniform composition to the mainstream plant. A mixing power input of 8 to 13 $\mathrm{kW}/10^3\ \mathrm{m}^3$ (0.3 to 0.5 $\mathrm{hp}/10^3\ \mathrm{ft}^3$) may be required if high suspended solids concentrations of suspended solids frequently occur ($>$ 1000 mg/L). The mixing power input may be decreased by 50 percent if high suspended solids concentrations are infrequent. Aeration is not provided typically, unless the tank is performing a dual function of equalization and biological treatment, in which case the tank design is dictated by the desired biological treatment performance.

Depending on the sidestream characteristics, the tank may be covered and equipped with an odor control device to minimize odorous emissions. Liquid level sensors with high and low alarms, an overflow to the plant drain system, a sloped floor with drain line, and a variable speed return pump or control valve with flow metering are common features in equalization tank design. Struvite formation in the tank is a potential concern and will impact the design of the mixer and selection of pipe or pipe liner material. The flexibility of adding iron salts to the tank to limit struvite formation may be a consideration. Configuring sidestream piping systems to avoid introduction or contact with air will also help limit struvite formation.

Equalization and TSS Reduction. Equalization tanks for sidestream pretreatment processes often are used to reduce the TSS concentration if solids reduction is a requirement for the physiochemical or biological pretreatment process. In this design, mechanical mixing is not provided, the liquid level is greater and the operating liquid level range limited as to not disturb solids settling and thickening in the bottom section of the tank. Thickened solids are pumped intermittently to the solids processing facility. Because the tank is performing the dual function of sidestream storage and suspended solids reduction, the tank volume requirement may be greater than a tank only performing equalization, but a separate solids removal process is avoided.

Equalization and Biological Pretreatment. Sidestream equalization and biological pretreatment can be accomplished within a single tank. In a process called storage and treat (SAT), the tank is designed as a sequencing batch reactor but has the flexibility to operate in a continuous overflow mode during short periods of peak sludge dewatering. Tank volume and aeration capacity are dictated by the treatment objectives, rather than the equalization requirement. At full scale, SAT has been used to provide partial treatment, resulting in pretreated sidestream containing a reduced ammonium concentration and a mixture of nitrite and nitrate (Laurich, 2004). Biological treatment of sidestream is discussed in Sec. 15–7 through 15–11.

EXAMPLE 15–2 **Impact of Full Sidestream Equalization on Plant Influent Ammonium-N Concentration and Calculation of Equalization Tank Volume** Anaerobically digested sludge is normally dewatered from 8:00 a.m. to 4:00 p.m., 5 d/wk. Sidestream is produced at an average rate of 0.006 m^3/s during each 8-h period and contains an ammonium-N concentration of 1000 mg/L. The sidestream is sent to the facility's headworks. The longest number of consecutive days in which no sludge is dewatered is 3 d. For the average hourly facility influent flows and ammonium-N concentrations shown in the table below: (1) develop concentration-versus-time graphs for non-equalized and fully-equalized sidestream and (2) calculate the equalization tank volume required for full equalization.

Time period	Given data	
	Average flowrate during time period, m^3/s	Average NH_4^+-N concentration during time period, mg/L
M–1	0.275	20.0
1–2	0.220	18.8
2–3	0.165	17.9
3–4	0.130	20.2
4–5	0.105	17.3
5–6	0.100	15.6
6–7	0.120	15.3
7–8	0.205	13.4
8–9	0.355	19.6
9–10	0.410	27.0

(continued)

(*Continued*)

	Given data	
Time period	Average flowrate during time period, m^3/s	Average NH_4^+-N concentration during time period, mg/L
10–11	0.425	30.2
11–N	0.430	35.1
N–1	0.425	35.3
1–2	0.405	28.5
2–3	0.385	24.9
3–4	0.350	22.8
4–5	0.325	21.3
5–6	0.325	21.3
6–7	0.330	21.0
7–8	0.365	19.5
8–9	0.400	21.8
9–10	0.400	21.0
10–11	0.380	19.7
11–M	0.345	20.6
Average	0.307	

Note: $m^3/s \times 35.3147 = ft^3/s$.
$m^3 \times 35.3147 = ft^3$.
$mg/L = g/m^3$.

Solution

1. Calculate the fully equalized average sidestream flowrate.

$$\begin{aligned}\text{Equalized flowrate} &= (0.006\ m^3/s)(3600\ s/h)(8\ h/d)\ (5\ d/wk)/[(7\ d/wk)\\ &\quad (86{,}400\ s/d)]\\ &= 0.00143\ m^3/s\end{aligned}$$

2. Using the non-equalized and fully-equalized sidestream flows, calculate the hourly average ammonium-N in the facility headworks.

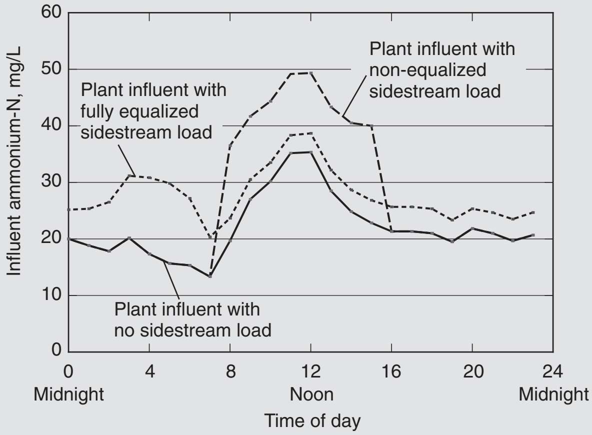
	Plant Influent without sidestream		No Sidestream Equalization		Sidestream Equalization	
Time period	Average flowrate during time period, m^3/s	Average NH_4^+-N concentration during time period, mg/L	Sidestream flowrate, m^3/s	Average ammonium-N concentration, mg/L	Sidestream flowrate, m^3/s	Average ammonium-N concentration, mg/L
M–1	0.275	20.0		20.0	0.00143	25.1
1–2	0.220	18.8		18.8	0.00143	25.3
2–3	0.165	17.9		17.9	0.00143	26.5
3–4	0.130	20.2		20.2	0.00143	31.2

(*continued*)

(*Continued*)

	Plant Influent without sidestream		No Sidestream Equalization		Sidestream Equalization	
Time period	Average flowrate during time period, m^3/s	Average NH_4^+-N concentration during time period, mg/L	Sidestream flowrate, m^3/s	Average ammonium-N concentration, mg/L	Sidestream flowrate, m^3/s	Average ammonium-N concentration, mg/L
4–5	0.105	17.3		17.3	0.00143	30.9
5–6	0.100	15.6		15.6	0.00143	29.9
6–7	0.120	15.3		15.3	0.00143	27.2
7–8	0.205	13.4		13.4	0.00143	20.3
8–9	0.355	19.6	0.006	36.5	0.00143	23.7
9–10	0.410	27.0	0.006	41.6	0.00143	30.5
10–11	0.425	30.2	0.006	44.3	0.00143	33.5
11–N	0.430	35.1	0.006	49.1	0.00143	38.4
N–1	0.425	35.3	0.006	49.4	0.00143	38.7
1–2	0.405	28.5	0.006	43.4	0.00143	32.1
2–3	0.385	24.9	0.006	40.5	0.00143	28.6
3–4	0.350	22.8	0.006	40.0	0.00143	26.9
4–5	0.325	21.3		21.3	0.00143	25.7
5–6	0.325	21.3		21.3	0.00143	25.7
6–7	0.330	21.0		21.0	0.00143	25.3
7–8	0.365	19.5		19.5	0.00143	23.4
8–9	0.400	21.8		21.8	0.00143	25.4
9–10	0.400	21.0		21.0	0.00143	24.6
10–11	0.380	19.7		19.7	0.00143	23.5
11–M	0.345	20.6		20.6	0.00143	24.7
Average	0.307					

3. Plot the resulting influent ammonium concentration.



4. Calculate the equalization tank volume required for full equalization with 3 consecutive days with no sludge dewatering.

$$V = (N)(Q_{\text{dw}})(D_{\text{dw}}/7)$$

$$V = (3)\,[(0.006\ \text{m}^3/\text{s})(3600\ \text{s/h})]\,(8\ \text{h/d})\,[(5\ \text{d/wk of sludge dewatering})/(7\ \text{d/wk})] = 370\ \text{m}^3\ (13{,}066\ \text{ft}^3)$$

15–3 REDUCTION OF SUSPENDED SOLIDS AND COLLOIDAL MATERIAL

The solids balance for a wastewater treatment plant, as illustrated in Sec. 14–7 in Chap. 14, is influenced by the TSS in the sidestreams derived from sludge thickening and biosolids dewatering processes. Sidestreams are commonly returned to the headworks where solids settle in primary sedimentation tanks. In plants that do not have primary sedimentation or return the sidestreams directly to the mainstream secondary process, the solids become integrated into the mixed liquor suspended solids. Because the majority of the post-digestion sidestream TSS are biologically inert, they will contribute directly to the secondary solids inventory and will exert only a minor oxygen demand on the process. The impact of these solids on the primary tank effluent or the secondary process operating conditions and performance varies with the sidestream solids loading rate, the operating conditions of the primary process, and the type of the secondary process employed.

Some sidestreams, particularly from post-digestion dewatering processes, contain colloidal matter that may impact the plant effluent quality, depending on the type of secondary process in the facility and its operating conditions. The particles will contribute to turbidity and may negatively impact effluent disinfection. Plant effluent colloidal particles are a particular concern for facilities that provide treated effluent for reuse.

Common practices used for the management of TSS and colloidal matter in thickening and dewatering sidestreams, at full scale, are discussed below.

Sidestreams Derived from Sludge Thickening

Thickeners that suffer from poor capture efficiencies will generate sidestreams with a high TSS concentration. The elevated sidestream solids may negatively impact the solids removal efficiency across the primary sedimentation tanks or contribute an unacceptable solids loading rate to a secondary process. Thickener operating conditions are commonly adjusted to improve solids capture to resolve this problem rather than installing a separate solids reduction process for the sidestream.

Sidestreams Derived from Biosolids Dewatering

Reducing the TSS concentration in post-digestion sidestream before the sidestream is sent to the mainstream plant is uncommon. If the dewatering sidestream solids concentration is frequently high and the solids returned to the mainstream plant prove detrimental, the dewatering process operating conditions must be adjusted to improve solids capture. A solids balance should be performed to assess its contribution to the mixed liquor solids inventory and concentration and determine if a sidestream solids reduction step would be beneficial. Life cycle cost analyses can be performed comparing operating costs of power, aeration, and chemicals in mainstream treatment with that of a sidestream pretreatment step.

Reduction of TSS is more common when sidestream is pretreated to reduce the nutrient load returned to the mainstream plant. The pretreatment requirement is specific to the physiochemical or biological process employed for nutrient reduction, as discussed in Sec. 15–4 through 15–11.

Process Options. Reduction of TSS before sidestream pretreatment has been accomplished in practice through gravity settling in sidestream equalization tanks. The suspended solids concentration in the clarified sidestream is typically sufficient for biological treatment processes that benefit from suspended solids less than 200 mg/L. If higher removal efficiency is required, separation processes such as Lamella inclined plate settlers or high rate clarification processes (see Sec. 5–7 in Chap. 5 for process descriptions) may be applied where chemically enhanced flocculation can be employed optimally to improve solids removal. Experience is largely limited to inclined plate settlers, without chemical addition, as the majority of the pretreatment processes applied at full scale do not require a low suspended solids concentration. Filtration is rarely used on sidestream due to the high fouling potential of the colloidal material and residual polymer and the formation of inorganic foulants such as struvite.

In belt filter presses, belt washwater contains the majority of the uncaptured solids and the filtrate has a TSS concentration less than 500 mg/L. Belt filter presses can be designed for separate collection of washwater and filtrate, potentially eliminating the need for a solids reduction step. Separate collection of washwater and filtrate results in higher nutrient concentrations in the filtrate, which will improve the nutrient removal efficiency in the pretreatment process. Some pretreatment processes may also benefit from the elevated temperature of the filtrate, which is no longer being cooled by washwater.

Removal of Colloidal Matter

Sidestreams derived from post-digestion dewatering processes commonly contain colloidal particles that may impact the facility effluent. Typically, the colloids are not removed by primary sedimentation. If chemically-enhanced primary treatment is employed, partial or complete capture of the colloidal particles may occur. In the absence of chemical addition, the colloidal particles will enter the secondary treatment process where some portion of the particles will be captured by the mixed liquor suspended solids. The degree of capture is dependent on the type of secondary process and its operating conditions. If a sidestream pretreatment process is employed for nutrient reduction, a portion of the colloidal particles may be removed before the stream is sent to the mainstream plant.

Removal of sidestream colloidal particles by chemically-enhanced flocculation in a separate pretreatment process is not a common practice. If colloidal particles derived from post-digestion dewatering sidestream are identified as a primary source of turbidity in a plant effluent, addressing the problem by sidestream pretreatment may be the most cost-effective option as the colloids are present at their highest concentration in the sidestream and the sidestream flow is less than 1 percent of the mainstream flowrate. A process configuration could consist of solids reduction by gravity settling in an unmixed sidestream equalization tank, followed by an advanced chemically-enhanced filtration process. Filtration options are presented in Chap. 11. Bench or pilot-scale testing is recommended to assess feasibility and chemical requirements.

15–4 PHYSIOCHEMICAL PROCESSES FOR PHOSPHORUS RECOVERY

Phosphate ore, which serves as the primary source of phosphorus for modern agricultural practices is a limited resource. Significant depletion of known global reserves had been projected to occur as early as the end of the 21st century (Cordell et al., 2009), but based on the most recent projections known reserves will be available for the next 300 to 400 years (Van Kauwenbergh, 2010). However, as demand for phosphorus increases and the exploration for new reserves, the mining of lower quality ores, and the adoption of more expensive processing equipment occur, the price of phosphorus will continue to escalate in the future. Consequently, the phosphorus in wastewater treatment facility sludges and return flows may be viewed increasingly as an asset that should be recovered and reused as fertilizer rather than a nutrient that must be treated and disposed.

The primary focus of this section is on phosphorus recovery processes that have been developed and demonstrated at a full-scale on nutrient-rich sidestreams and industrial wastewaters. Other technologies have been developed to recover phosphorus from sludge ash, but these processes are not presented here. The phosphorus recovery processes, based on crystallization, considered in this section include those for the recovery of magnesium ammonium phosphate (struvite) and calcium phosphate (hydroxyapatite). Descriptions of the following crystallization processes that have been demonstrated in full-scale facilities are presented in Table 15–3.

AirPrex® process
Cone-shaped fluidized bed crystallizer
Crystalactor®
NuReSys® process
Pearl® process
Phosnix® process
PHOSPAQ™ process

However, before discussing these processes in greater detail, it will be useful to consider the fundamental aspects of the crystallization process. The beneficial reuse of recovered nutrients as fertilizers is discussed in Sec. 15–6.

Description of the Crystallization Process

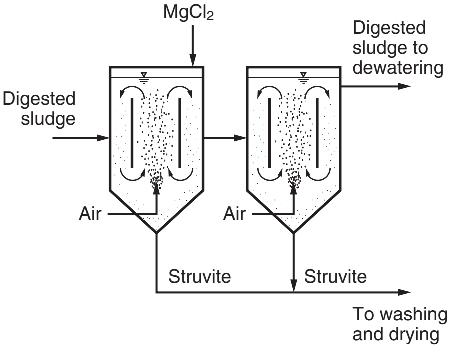
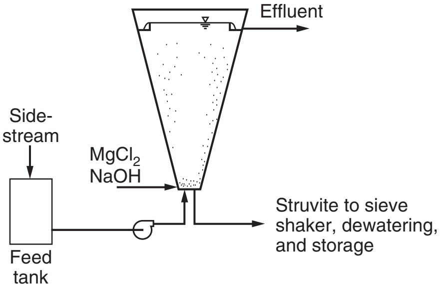
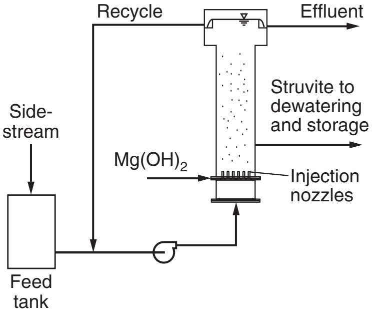
All of the technologies presented in Table 15–3 are based on the three fundamental stages that occur within the physical environment of the crystallizer (reactor): (1) supersaturated ion concentrations, (2) primary and secondary nucleation processes, and (3) crystal growth. The phosphate removal efficiency, crystal size distribution achieved within the reactor, and the purity of the final product are all influenced by the temperature, pH, ionic composition, and hydrodynamic conditions within the reactor. In the recovery of phosphate for reuse, a larger crystalline product is desired rather than an amorphous solid phase consisting of fine particles that are difficult to recover from the liquid and process to generate a reusable product.

Supersaturation. In supersaturated solutions, the product of the ion molar concentrations (expressed as activities) of the desired product exceeds the value of its solubility constant at given reaction conditions. As the degree of supersaturation increases, there is increased potential of forming fine particles through primary nucleation, not a desirable condition, as discussed below. Therefore, the supersaturated condition within a crystallizer is controlled to avoid fine particle formation and to provide sufficient driving force for mass transfer of ions to the surface of the growing crystals. Supersaturation is the driving force for subsequent processes of nucleation and crystal growth.

Nucleation. The process by which ions come together under supersaturated conditions and aggregate into a solid form, or “nuclei,” according to their solubility at a given temperature and pH is known as *nucleation.* The nucleation process cycles through

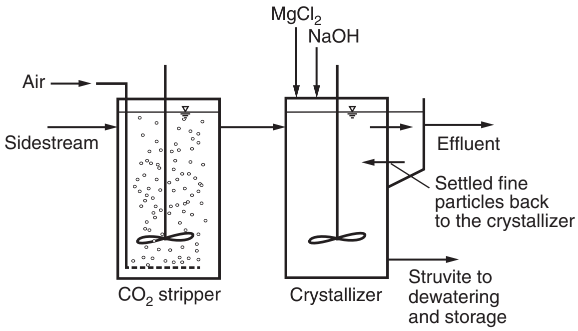
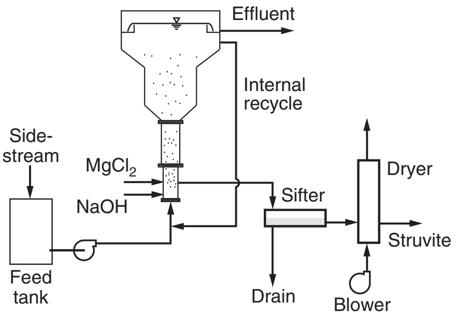
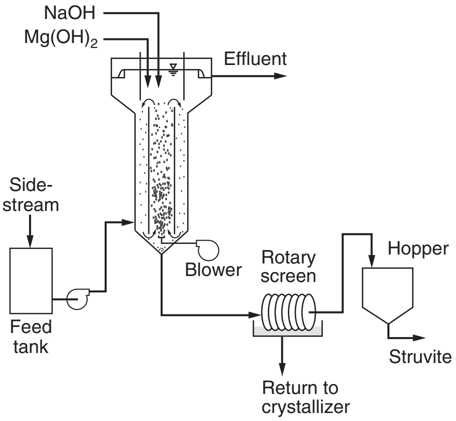
Table 15–3

Processes for recovery of sidestream phosphorus as magnesium ammonium phosphate (struvite)

	Description
(a) AirPrex® process  	The AirPrex® process was, developed by Berliner Wasserbetriebe (Germany) in collaboration with the Berlin Institute of Technology. In this process, struvite is crystallized directly from the sludge stream from an anaerobic digester, rather than from sidestream, to prevent struvite formation in the sludge dewatering process. AirPrex® consists of a dual-stage aerated tank configuration, either as separate tanks or as a single tank with a dividing wall, with a hydraulic retention time (HRT) of approximately 8 h. An air-lift aeration design is used in each stage to induce sludge mixing and strip CO_2 to increase pH. Magnesium chloride is used as the magnesium source and is added to the first, second or both stages. As struvite forms and develops into particle sizes of sufficient settling velocity, it settles into the bottom conical section of each stage. The product is withdrawn intermittently or continuously from each stage and transferred by a screw conveyer to a sand washer. Washed product is stored wet or is subsequently dried. Aerated sludge overflows the second stage and is sent to a sedimentation vessel where additional struvite may be recovered or to the dewatering process. Processing the exhaust air through an odor control system may be required.
(b) Cone-shaped fluidized bed crystallizer  	A cone-shaped fluidized bed crystallizer was developed by Multiform Harvest Inc. (USA). The crystallizer consists of a conical section and a solids-liquid separation zone located at the top. The dimensions of the conical section are selected to provide a desired range of superficial upflow velocities. The HRT is typically less than 1 h. As struvite crystals grow, they settle towards the bottom of the cone where they are removed intermittently, processed through a sieve shaker or drum screen, disinfected, and bagged for off-site processing. Magnesium chloride and sodium hydroxide are added through a proprietary injection system at the bottom of the cone to provide supersaturated conditions and to increase pH to the desired range.
(c) Crystalactor®  	The Crystalactor® is a fluidized bed crystallizer developed by DHV (The Netherlands). The crystallizer consists of a cylindrical reactor with a solids-liquid separation zone located at the top. Effluent is recirculated to the bottom of the reactor where it is blended with sidestream and injected into the crystallizer through nozzles to achieve optimum cross-section liquid distribution. The effluent recirculation rate is adjusted to maintain a superficial upflow velocity in the range of 40 to 75 m/h (130 to 250 ft/h) in the reaction section of the vessel. The HRT based on sidestream flow is typically less than 1 h. Quartz sand is added initially as seed material to accelerate startup, but further sand addition is not required once struvite crystals form. As the pellets grow, they settle to the bottom of the crystallizer, where a portion of the pellets is removed at regular intervals, dewatered, and stored for offsite transport.

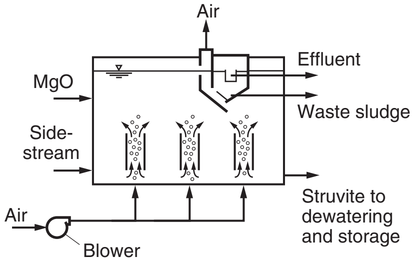
(*continued*)

Table 15–3 (*Continued*)

	Description
(d) NuReSys® process 	The NuReSys® process (NUtrient REcovery SYStem) was developed by Akwadok/NuReSys (Belgium). The process consists of a CO_2 stripping tank followed by a mechanically stirred crystallizer and a sedimentation zone. A HRT of 0.5 to 1 h is provided in the stripping tank where mechanical agitation and air are provided to strip CO_2 from the sidestream. Mixer speed and air flow are adjusted to control the pH to limit the formation of fine crystals in the stripping tank. In the crystallizer, mechanical stirring provides mixing and creates the hydrodynamic environment conducive to pelletized struvite formation. Stirrer speed and the product withdrawal rate are adjusted to provide the desired pellet size of the harvested product. Magnesium chloride is used as the magnesium source and NaOH is added to control the pH in the range of 8.1 to 8.3. A crystallizer HRT of 0.5 to 1 h is typical. Smaller crystals are settled in the sedimentation zone where they are returned to the crystallizer. Processing the exhaust air through an odor control system may be required.
(e) Pearl® process 	The Pearl® process was developed at the University of British Columbia for the crystallization of magnesium ammonium phosphate and was introduced at full-scale by Ostara Nutrients Recovery Technologies Inc. (USA). The Pearl® reactor is a fluidized bed crystallizer with a segmented construction where the segment or zone diameter increases from the bottom of the reactor to the top to reduce the upflow liquid velocity incrementally and retain struvite crystals of various sizes within each zone. A liquid/solids separation section is located at the top of the reactor. Effluent is recirculated to the bottom of the reactor to maintain the upflow velocity profile within the desired range. The HRT based on sidestream flow, is typically less than 1 h. As struvite pellet diameter increases, the pellets gradually sink from one zone to the next. The final product is removed from the bottom zone, separated from the liquid by screening, dried, and bagged. The effluent recirculation rate and struvite retention time are adjusted to control the pellet size in the final product. The magnesium source is typically magnesium chloride. Sodium hydroxide is used to maintain the pH within the desired range.
(f) Phosnix® process 	The Phosnix® crystallizer, developed by Unitika Ltd (Japan), consists of a cylindrical reaction zone with a conical bottom section and a larger diameter solids-liquid-gas separation section at the top. The crystallizer is aerated to provide mixing and to increase the pH by stripping CO_2 from the liquid. If required, the exhaust air is treated with an odor control system. The HRT in the reaction zone is less than 1 h. Magnesium hydroxide is typically used as the magnesium source and sodium hydroxide is added to control pH within the desired range. Larger struvite pellets that settle into the conical section of the reactor are pumped intermittently to a rotary drum screen. Liquid and associated smaller struvite particles that pass through the screens are sent to the mainstream plant or returned to the crystallizer to allow the fine particles to serve as seed material for struvite pellet growth.

(*continued*)

Table 15–3 (*Continued*)

	Description
(g) PHOSPAQ™ process 	PHOSPAQ™ was developed by Paques (The Netherlands). The process consists of an aerated reaction zone and a proprietary solids-liquid-air separation device in the upper section of the vessel. An air-lift aeration design is used to (1) provide mixing, (2) strip CO_2 from the liquid to increase pH, and (3) provide dissolved oxygen (DO) for biological treatment. The HRT is approximately 5 to 6 h. Magnesium oxide is typically used as the magnesium source. Struvite is harvested from the bottom of the reactor by pumping struvite-rich mixed liquor through a hydrocyclone. The recovered product is dewatered in a screw press and transferred to a container as a 70 percent dry material. The separation device above the reaction zone allows smaller struvite particles and biomass to settle and return to the reaction zone. An odor control system may be required for the exhaust air.

aggregate formation and dissolution until an aggregated cluster is of sufficient size to remain stable and provide a surface for crystal growth. Nucleation is subcategorized into primary and secondary processes. Primary homogeneous nucleation is a spontaneous formation of nuclei from supersaturated solution, while primary heterogeneous nucleation occurs on foreign solid surfaces (interior reactor surfaces, crystals, sand, colloids, dust). Of the two forms of primary nucleation, the latter is dominant in real applications.

Secondary nucleation also occurs in two fundamental forms: “fluid shear” nucleation and “contact” nucleation. In the fluid shear process, hydrodynamic forces sweep small nuclei from the crystal surface or cause breakage of branched protrusions from the crystal, resulting in the generation of a new “seed” surface for nucleation and growth of new crystals. Contact nucleation is a result of crystal attrition where physical forces fracture crystals, increasing the number of crystals and surface area for growth.

Crystal Growth. The growth of crystals comprises the third primary stage of the crystallization process. Under supersaturated conditions, ions diffuse through the boundary layer near the surface of each crystal and then integrate onto the crystal surface through a complex mechanism. As the dimensions of the crystal increase, it approaches a terminal size dictated by the hydrodynamic conditions in the reactor that cause fluid shearing and crystal attrition. Settling velocity increases with crystal size, which is advantageous as it allows the product to be separated more easily from the liquid and other solids within the reactor, a feature common to all of the processes shown in Table 15–3.

Recovery of Phosphorus as Magnesium Ammonium Phosphate (Struvite)

Magnesium ammonium phosphate, commonly known as struvite, has limited water solubility and forms encrustations in anaerobic digester outlet structures and downstream processes (dewatering equipment, pipes conveying sidestream) causing significant operational and maintenance problems. The objectives of struvite phosphorus recovery processes are to form struvite and produce a crystallized product with sufficient purity and physical characteristics that it qualifies for reuse as a fertilizer. The value of struvite in the fertilizer market is discussed in Sec. 15–6.

Reaction Stoichiometry. The chemistry of struvite formation is presented in detail in Sec. 6–5 in Chap. 6. The stoichiometry of the struvite precipitation reaction is shown in Eq. (6–25), repeated here for convenience.

$$Mg^{2+} + NH_4^+ + PO_4^{3-} + 6H_2O \rightarrow MgNH_4PO_4{\cdot}6H_2O \qquad (6\text{–}25)$$

Typically, the product of the ion molar concentrations of the struvite constituents in the anaerobic digester bulk liquid is not high enough to induce a substantial level of struvite formation, thus limiting the phosphate recovery efficiency. To overcome this limitation in the processes described in Table 15–3, a magnesium salt is added to the crystallizer. As discussed below, the addition of magnesium is not the only requirement for successful production of struvite. The pH and hydrodynamic conditions within the crystallizer must be controlled to optimize phosphate removal and generate a product of sufficient size that can be separated easily from the liquid phase and other suspended solids.

Calcium Inhibition. Calcium competes with magnesium for phosphate under conditions conducive to precipitation of calcium phosphates and magnesium ammonium phosphate. A Mg/Ca molar ratio of less than 2/1 results in a longer induction time and a lower struvite crystal growth rate. As the ratio decreases below 1/1, amorphous calcium phosphate formation is dominant (Le Corre et al., 2005).

Operational Considerations. The principal operational requirements that must be considered in the application of phosphorus recovery as struvite include (1) pretreatment requirements, (2) pH and temperature control, (3) chemical requirements, (4) seed requirements, and (5) mixing and hydraulics. Each of these factors is considered in the following discussion.

Pretreatment Requirements. Sidestream pretreatment to reduce TSS and colloidal material has not been standard practice. Upflow fluidized bed crystallizers can tolerate sidestream TSS concentrations ranging from 1500 to 5000 mg/L, depending on the technology. The upflow velocities are high enough to prevent digested biological solids from settling. Based on operating experience with the AirPrex® process, struvite crystals can be produced and separated from anaerobic digester effluent (biosolids), albeit with a larger reactor.

pH and Temperature Control. Struvite solubility decreases with increasing pH, reaching a minimum solubility near pH 10.3 (see Sec. 6–5 in Chap. 6). However, in practice, struvite crystallizers do not operate at pH greater than 9.0. Typically, pH is controlled in the range of 8.0 to 8.8 to minimize the addition of base chemicals, limit the degree of supersaturation, and limit the potential formation of other solids such as calcium carbonate and calcium phosphate (hydroxyapatite). These solids would lead to greater impurity of the final crystalline product. A reduction in the phosphorus recovery efficiency may also occur due to the formation of fine amorphous calcium phosphate precipitate that will exit the crystallizer with the treated effluent. In the pH range of 8.0 to 8.8, phosphorus recovery efficiencies greater than 80 percent and the production of a highly pure struvite product have been demonstrated.

The formation of struvite at a pH less than 8.0 has also been demonstrated. In the AirPrex® process, where struvite is formed directly from anaerobic digester effluent (sludge), operating in a pH range of 7.2 to 7.4 results in a high phosphorus recovery efficiency and generation of struvite crystals of sufficient size that allows separation from the sludge in the conical section of the reactor (Nieminen, 2010). The crystal growth rate is reduced as the operating pH decreases, which results in a longer residence time in the crystallizer to achieve the same level of phosphorus recovery and crystal sizes. In sidestreams derived from the digestion of WAS from biological phosphorus removal processes, the orthophosphate concentration and the resulting level of supersaturation are sufficiently high that a pH greater than 8.0 may not be required to achieve the desired level of phosphorus recovery.

The strategies used in attaining the desired operating pH range varies by process as summarized in Table 15–3 and the sidestream characteristics. Aeration of the crystallizer has been shown to be an effective method for increasing pH through stripping of CO_2. Anaerobic digester effluent and sidestream are supersaturated with CO_2 due to the high CO_2 content of the digester gas. Once exposed to air under atmospheric conditions, CO_2 will diffuse from the liquid and pH will increase above 8.0. In the AirPrex®, PHOSPAQ™, and Phosnix® technologies described in Table 15–3, aeration is applied directly to the crystallizer to strip CO_2 from the liquid. Aeration also provides mixing and induces hydrodynamic conditions conducive to developing crystals over a range of sizes that allows separation and processing. In the NuReSys® process, CO_2 is stripped from the sidestream through aeration and mechanical agitation in a vessel preceding the crystallizer. The application of high speed mechanical agitation and aeration reduces energy consumption and limits the amount of air that may require treatment in an odor control process. Mixer speed and air flow are adjusted to limit the formation of fine struvite crystals before the sidestream enters the crystallizer.

The dominant forms of orthophosphate in an anaerobic digester and sidestream are HPO_4^{2-} and $H_2PO_4^-$, with PO_4^{3-}, the form required for struvite formation, existing at a low concentration. As shown in Eq. (15–2), as struvite forms, HPO_4^{2-} and $H_2PO_4^-$ shift towards PO_4^{3-}, resulting in proton release or an increase in acidity.

$$H_2PO_4^- \rightleftarrows HPO_4^{2-} + H^+ \rightleftarrows PO_4^{3-} + 2H^+ \tag{15–2}$$

If magnesium chloride is used as the source of magnesium, additional acidity is generated through the chloride ions remaining in solution. The acidity generated by these chemical reactions will be buffered by neutralization with bicarbonate, but if sufficient acidity is generated, pH will decrease and a target pH above 8.0 may not be achieved by aeration alone. Magnesium oxide or hydroxide may be selected as the magnesium source as an alternative to magnesium chloride to provide alkalinity or sodium hydroxide can be added for pH control. Depending on the amount of acidity created as a result of struvite formation, magnesium oxide (or hydroxide) and sodium hydroxide may be required to control the process.

Struvite solubility is a function of temperature, but temperature control is not practiced in the processes presented in Table 15–3. Cooling or heating the reactor to operate at an ideal temperature to maximize the phosphate removal efficiency is not justified economically.

Chemical Requirements. The source of magnesium is typically magnesium chloride, magnesium hydroxide, or magnesium oxide, which forms magnesium hydroxide upon contact with water. The choice of chemical for each process varies and is based on vendor or end user preference and chemical cost.

Magnesium chloride has the advantage of disassociating faster than magnesium hydroxide, leading to higher reaction rates. Consequently, an optimized chemical feed dispersion system is required for the reactor to prevent localized excessively high supersaturated conditions near the chemical addition point.

Magnesium oxide and hydroxide provide both magnesium and alkalinity to the crystallizer. The primary role of both compounds is to provide magnesium. The alkalinity provided by the chemical is advantageous, but may not increase the pH and stabilize it within the desired operating range. In this case, sodium hydroxide addition will be required for pH control. Magnesium oxide and hydroxide have limited water solubility and are fed to the crystallizer as suspensions in water. Consequently, the solubilization rate of the compounds may control the level of supersaturation and the crystal growth rate. Nucleation may also occur on the surface of undissolved reagent leading to reduced purity of the harvested product.

Magnesium is added to the process in molar excess to maintain a supersaturated condition for crystal growth. A molar ratio of Mg^{2+} to PO_4^{3-} in range of 1.1 to 1.6 has been used in practice for municipal sidestreams, with a value of 1.3 being typical. If excess magnesium is added, resulting in a higher level of supersaturation, excessive primary nucleation will occur, leading to the formation of small crystals that may not be retained in the crystallizer, depending on its hydraulic design. At the Mg^{2+}/PO_4^{3-} molar ratio and operating pH range used in crystallizing struvite from municipal sidestreams, inhibition by calcium and the formation of calcium phosphate is minimal, leading to a high recovery efficiency and high purity of the harvested product.

Sodium hydroxide is used typically to adjust and control pH, depending on the process technology and desired operating pH. Sodium hydroxide is preferred due to ease of handling, and it can be stored at concentrations as high as 50 percent by weight.

Chemical requirements are specific to the sidestream as operating conditions in the mainstream process and solids processing facility dictate the phosphate concentration in the digester and its variability. Pilot or demonstration-scale testing of the selected crystallization technology is also commonly performed to assess the chemical requirements and other operating conditions for design of the full-scale process.

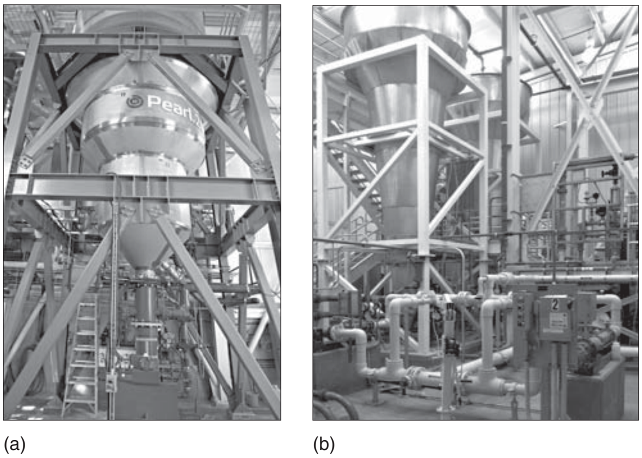
Seed Requirements. A seed material such as sand or struvite crystals is often added to a crystallizer to rapidly start the process. After a certain level of crystal inventory is achieved in the reactor, the process is self-sustaining at the target operating conditions (e.g., level of supersaturation, pH) and further seed addition is not required.

Mixing and Hydraulic Requirements. The mixing and hydraulic conditions within the reactor have an impact on the size of the crystals harvested from the process. In the AirPrex®, PHOSPAQ™ and Phosnix® processes, an air-lift aeration design is used to strip CO_2 and induce a mixing pattern within the reactor, resulting in the development of crystal sizes sufficient for separation and processing. In the AirPrex® and Phosnix® reactors, a conical bottom provides an environment where the larger struvite crystals can separate and thicken prior to their removal. Alternatively, solids from the bottom of the reactor can be intermittently pumped through a hydrocyclone to recover the product for further processing, as done in the PHOSPAQ™ process.

In the Pearl® and Crystalactor® processes, effluent from the crystallizer is recycled to maintain the upflow liquid velocity within the desired range to develop crystals and a pelletized product with a specific range of sizes. If the size of the crystals or pellets harvested from the reactor is of less importance, effluent recirculation may not be required. In the cone-shaped fluidized bed crystallizer (Table 15–3), where no internal recirculation is applied, crystals develop of sufficient size that allows separation from the liquid. The

Figure 15–1

Examples of full scale struvite crystallizers: (a) Ostara Pearl® reactor, Tigard, OR and (b) Multiform Harvest cone-shaped reactor, Yakima, WA (courtesy of Multiform Harvest Inc.).



(a) (b)

struvite is removed intermittently and processed offsite to generate product characteristics specific to a particular end use.

The specific geometry of the crystallizer may also be designed to create a range of upflow velocities that impact the characteristics of the harvested product and allow retention of smaller crystals, which continue to grow and sink into the lower sections of the crystallizer. The Pearl and cone-shaped crystallizers shown on Figs. 15–1(a) and (b) are examples of processes that create variable upflow velocities through an increasing cross-sectional diameter. In the Phosnix®, Crystalactor® and PHOSPAQ™ processes, the smaller crystals are retained in low upflow velocity zones or sedimentation zones in the upper section of the reactors and gradually return to the main reaction zone.

The hydrodynamic conditions required for crystallization and pellet formation can also be achieved through mechanical mixing. In the NuReSys® process, a mechanical mixer with a three-blade impeller has been used successfully to provide mixing and generate a pelletized final product.

Product Separation and Purification. Crystals are allowed to grow to sizes that allow separation from the liquid and other suspended solids and are easy to process after harvesting (e.g., screening, washing). Separation and recovery of product with a mean crystal size of 0.2 mm or greater from the liquid has been demonstrated. Depending on the crystallizer design, the hydrodynamic conditions will enhance crystal agglomeration and form spherical pellets with a mean diameter up to 2–4 mm. The mean diameter of the product is controlled by adjusting the product harvesting rate to increase or decrease the product residence time in the crystallizer.

Processing requirements for the struvite product vary. The product may be concentrated by screening, disinfected with chlorinated water or heat, rinsed, dewatered, and dried. Product purity typically exceeds regulatory requirements and no further purification steps are required. The degree to which the final product is processed is dependent on the end user requirements. Examples of processing equipment for the pelletized product harvested from a Pearl crystallizer are shown on Fig. 15–2.

Struvite Phosphorus Recovery Limitations. In practice, the orthophosphate is not recovered completely and a crystallizer effluent concentration below 5 mg P/L is not typical. For fluidized bed reactors with a relatively low hydraulic retention time (less than 1 hour), consistently achieving an effluent concentration below 10 mg P/L in these systems

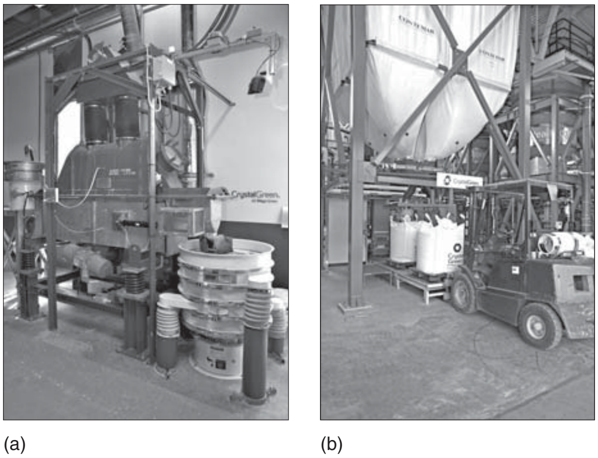

(a) (b)

Figure 15–2

Processing of pelletized struvite from a Pearl crystallizer: (a) drying; (b) product bagging.

requires greater chemical addition to maintain sufficient supersaturated conditions throughout the crystallizer as orthophosphate decreases. Based on economic considerations, i.e., chemical cost versus higher recovery efficiency, an effluent orthophosphate concentration in the range of 10 to 25 mg P/L is typical. In processes such as AirPrex® where the hydraulic retention time is much higher (e.g., 8 h), effluent concentrations below 10 mg P/L have been demonstrated, but at the expense of requiring a larger reactor.

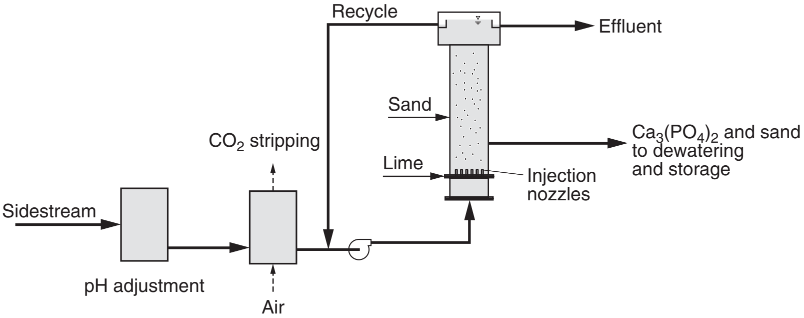
Recovery of Phosphorus as Calcium Phosphate

Phosphorus can be recovered from sidestream flows through precipitation as calcium phosphate. A typical process flow diagram for the recovery of phosphorus as calcium phosphate is presented on Fig. 15–3. The process was developed initially and demonstrated successfully in The Netherlands in the 1980s (Piekema and Giesen, 2001). However, the rise in chemical costs has made this process generally uneconomical, especially as compared to the recovery of phosphorus as struvite. Although the recovery of phosphorus as calcium phosphate is now largely limited to industrial, food, and dairy applications, there are still wastewater applications. Phosphate recovery from mainstream processes with lime precipitation continues to be used in several facilities and is discussed following the discussion of the recovery of calcium phosphate.

Reaction Stoichiometry. Calcium phosphate can exist in several forms depending on pH and the ionic composition of the sidestream. The basic chemistry of phosphate precipitation with lime is presented in Sec. 6–5. The primary product generated during

Figure 15–3

Process flow diagram for the production of calcium phosphate from nutrient-rich sidestream.



crystallization is hydroxyapatite [$Ca_5(PO_4)_3OH$], the most thermodynamically stable form of calcium phosphate.

Operational Considerations. The principal operational requirements that must be considered in the application of phosphorus recovery as calcium phosphate include (1) pretreatment requirements, (2) pH and temperature control, (3) chemical requirements, and (4) seed requirements. Each of these factors is considered in the following discussion.

Pretreatment Requirements. Most high-strength sidestreams contain a high bicarbonate concentration that will cause significant formation of calcium carbonate precipitate in the crystallizer. To limit the formation of calcium carbonate, the sidestream is pretreated through acidification to a pH less than 5 with a strong inorganic acid, and the acidified stream is subjected to air stripping to remove CO_2. The pretreated stream is then treated with calcium hydroxide slurry to precipitate phosphate.

pH and Temperature Requirements. Calcium phosphate crystallization is performed over a pH range of 8.0 to 9.0. The optimum pH for a given application is developed through pilot-scale tests. In practice, the crystallizer temperature is not controlled, due to economic considerations.

Chemical Requirements. An inorganic acid is required for sidestream pretreatment. The amount of acid can be estimated preliminarily by assuming complete conversion of bicarbonate to CO_2. Quick lime (CaO) is used typically as the calcium source for the crystallizer and fed as slurry (forming calcium hydroxide upon addition to water). Lime is added to increase the pretreated sidestream pH from 5.0 or less to the crystallizer pH setpoint and provide the desired level of supersaturation in the crystallizer. A lime overdose in the range of 0.5 to 5.0 mM has been used in practice (Piekema and Giesen, 2001). The inorganic acid and lime requirements are determined through pilot-scale tests.

Seed requirements. Calcium phosphate crystallization through primary nucleation creates a fine microcrystalline product that is difficult to separate from water. However, in the presence of seed material such as sand and under well-controlled supersaturated conditions, nucleation will occur on the surface of the sand particles and crystal growth on the newly formed pellet is sustained. By controlling the hydrodynamic conditions within the fluidized-bed crystallizer, a pelletized product with a mean diameter of approximately 1-mm is harvested from the bottom of the crystallizer. As pelletized product is removed intermittently, virgin seed material is added. The superficial upflow velocity through the crystallizer is in the range of 40 to 75 m/h (130 to 250 ft/h).

To eliminate the need for inorganic carbon removal via acidification and air stripping, alternative seed materials have been proposed. Calcite and calcium silica hydrates have been shown to be effective in the recovery of phosphorus by means of hydroxyapatite crystallization (Berg et al., 2006; Donnert and Salecker, 1999). The intermittent addition of seed material is a requirement, as it is with sand, as pelletized product is removed from the crystallizer. To date, the advantages of this approach, if any, for phosphorus recovery as compared to struvite crystallization have yet to be demonstrated at full scale.

Phosphorus Recovery from Mainstream Processes

Phosphate removal from the mainstream wastewater through chemical precipitation and adsorption with ferric and alum salts has been practiced widely. The chemical sludge produced by this reaction is amorphous and is formed in primary sedimentation tanks or in the activated sludge process where it is intimately blended with primary sludge or activated sludge. The form of the chemical sludge and its intimate mixing with non-chemical sludges makes phosphate recovery more challenging. Technologies for recovering phosphate from sludges and incinerator ash have been developed, but are not discussed in this chapter.

Phostrip Process. One of the first processes used for the recovery of phosphate from the mainstream process is the Phostrip process, developed specifically for enhanced biological phosphorus removal (EBPR). A description of the Phostrip process is presented in Table 8–27 in Sec. 8–8. In the Phostrip process, a portion of the phosphorus-rich return activated sludge (20 to 40 percent) from the EBPR process is subjected to anaerobic conditions at a sludge retention time in the range of 12 to 20 h. Under these conditions, endogenous production of readily biodegradable COD (rbCOD) causes release or *stripping* of orthophosphate from the sludge to the bulk water. The phosphate *stripper* tank typically consists of a gravity thickener with thickened sludge recirculation to elutriate the released phosphorus from the sludge to the thickener overflow. For nitrifying mainstream processes, a prestripper tank is provided prior to phosphate stripping to denitrify the RAS flow and optimize phosphate release. Thickened RAS is returned to the mainstream process or a portion is sent to the solids processing facility as waste activated sludge (WAS).

Traditionally, stripper overflow has been subjected to chemical precipitation with lime at a pH of 9.0–9.5 to remove phosphate. Due to the presence of bicarbonate in the stripper overflow, calcium carbonate formation will also occur, resulting in a mixed solids composition. Typically, the chemical solids are removed separately or settled in the primary sedimentation tanks. Alternatively, the overflow from the stripper tank can be blended with sidestream from post-digestion dewatering and fed to a crystallization process to produce magnesium ammonium phosphate.

Release of Phosphate from Waste Activated Sludge. The Phostrip concept can be adapted for waste activated sludge derived from an EBPR process where WAS is subjected to anaerobic conditions to allow solids hydrolysis and fermentation, which results in the release of phosphate. Releasing phosphate from WAS prior to anaerobic digestion significantly reduces struvite formation in the digester and digester overflow. The WAS fermenter or stripper may be configured as a WAS thickener with internal thickened sludge recycle, similar to the Phostrip process. The thickener overflow is blended with post-digestion sidestream for struvite recovery. Magnesium is also released from the WAS at a mass ratio of approximately 0.25 g Mg/g PO_4^{3-}-P released, reducing the magnesium chloride or magnesium oxide/hydroxide requirement in the struvite crystallizer.

Enhanced Release of Phosphate from Waste Activated Sludge. The addition of primary sludge fermentate or acetic acid to the WAS stripper will enhance the phosphate release rate and reduce the sludge residence time in the stripper from 12–20 h (endogenous process) to 2–5 h. A VFA to VSS ratio of 0.02 to 0.04 g/g is required (Schauer et al., 2011; Corrado, 2009). In general, the phosphate release rate is insensitive to the VFA concentration in the WAS stripper. The optimal configuration for the enhanced release WAS stripping process is a complete mixed tank followed by a solids thickening step, as shown in the process flow diagram presented on Fig. 15–4. Depending on the operating conditions, the Phostrip thickener configuration with thickened sludge recycle may also provide enhanced phosphate release.

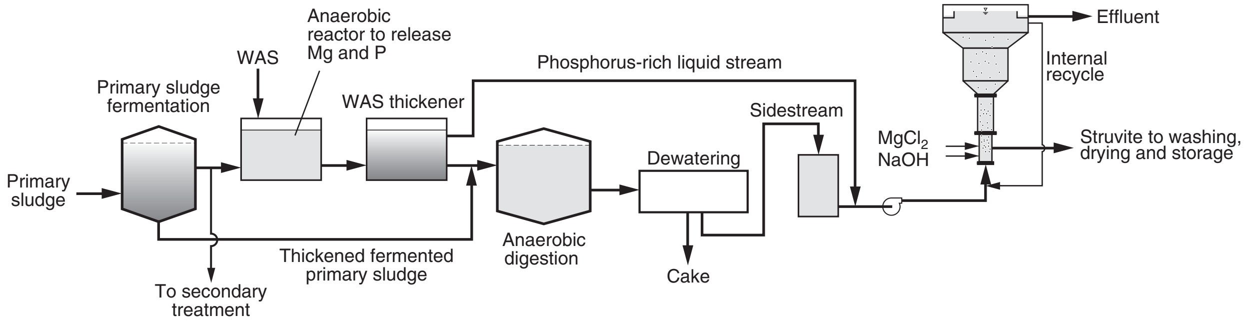


Figure 15–4

Process flow diagram for volatile fatty acid (VFA) enhanced stripping of phosphate from secondary waste sludge enriched with polyphosphate.

15–5 PHYSIOCHEMICAL PROCESSES FOR AMMONIA RECOVERY AND DESTRUCTION

Physiochemical processes for sidestream ammonium treatment are alternatives to biological treatment, which is the dominant method for ammonium removal in practice. Recovery of ammonia from wastewaters to produce aqueous ammonia or an ammonium salt (e.g., ammonium sulfate, ammonium nitrate) for use in industrial and agricultural applications has been of interest for decades, and a number of processes have been developed, demonstrated, and practiced at a full scale. Where ammonia reuse is not desired, a thermal catalytic destruction technology has also been developed and practiced at full scale where ammonia stripped from wastewater is catalytically converted to N_2 at high temperature.

Processes that have been demonstrated at a full scale and are currently used in recovering or destroying ammonia from wastewaters derived from industrial processes, municipal sludge digestion, landfill leachate treatment and animal manure digestion are described in this section. Technologies such as ion exchange and adsorption are rarely, if ever, practiced in the treatment of high strength wastewaters and are not considered in this section. Emerging technologies such as ammonia electrolysis to generate hydrogen for energy recovery (Vitse et al., 2005) and alternative ammonia stripping or volatilization technologies such as Vacuum Flash Distillation (Kemp et al., 2007) and Membrane Contactors (Membrana, 2007; du Preez et al., 2005) are also not considered as application at a full scale is limited or in development.

Recovery of Ammonia by Air Stripping and Acid Absorption

The recovery of ammonia from high strength wastewaters by air stripping–acid absorption technology has been used in both industrial and municipal applications. Most notable in the municipal sector is the process at the VEAS facility [3.5 m^3/s (80 Mgal/d)] in Oslo, Norway, where ammonium sulfate was produced from 1996 to 1998, and ammonium nitrate has been produced since 1998 (Sagberg et al., 2006). While this technology has not been applied widely in North America, several processes have been in operation in Europe since the late 1980s for the recovery of ammonia from municipal digester sidestreams, manure digestion sidestreams, landfill leachate, and industrial wastewaters.

Although sulfuric acid is the least expensive and most commonly used, other types of acids can also be used:

- Phosphoric acid—produces mono-ammonium phosphate (MAP) or di-ammonium phosphate (DAP)
- Hydrochloric acid—produces ammonium chloride (NH_4Cl)
- Acetic acid—produces ammonium acetate ($NH_4C_2H_3O_2$)
- Nitric acid—produces ammonium nitrate (NH_4NO_3)

The selection of a specific acid depends on local or regional demand for the resulting product and if revenue could be realized by the sale of the product. However, ammonium sulfate is the dominant product of choice from air stripping–acid absorption processes currently in practice, largely driven by chemical cost and market demand.

Process Description. The basic air stripping/acid absorption process flow diagram and an example of a full-scale process are shown on Fig. 15–5. The process is comprised of pH adjustment, TSS removal, a dual column air stripper–acid absorber system and chemical storage tanks with associated delivery systems and controls. As described in Sec. 11–10 in Chap. 11, stripping ammonia from wastewater with air requires ammonium to be converted to ammonia in the liquid phase through an increase in pH. The effects of temperature and pH on the percentage of ammonium in the form of ammonia for an aqueous ammonia solution are illustrated on Fig. 15–6. At the typical digestion sidestream temperature range of 25 to 35°C, a pH of 11 or higher is required to shift nearly 100 percent of the ammonium to ammonia. The ionic composition of the sidestream affects the chemical equilibrium illustrated on Fig. 15–6 due to non-ideal conditions created by

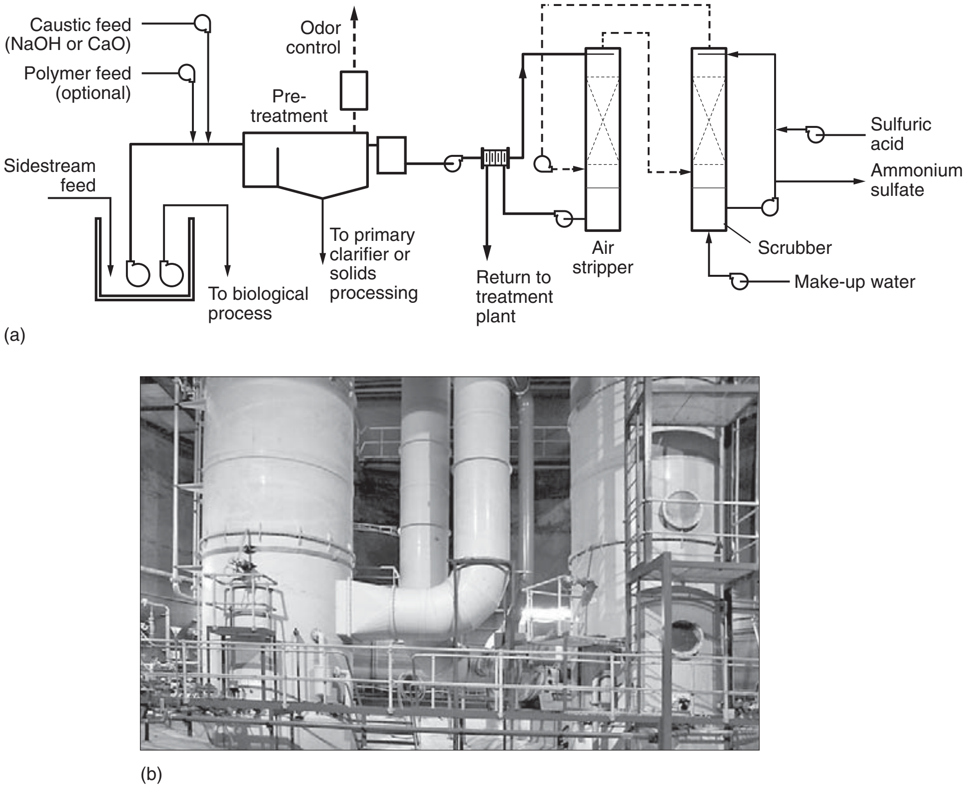
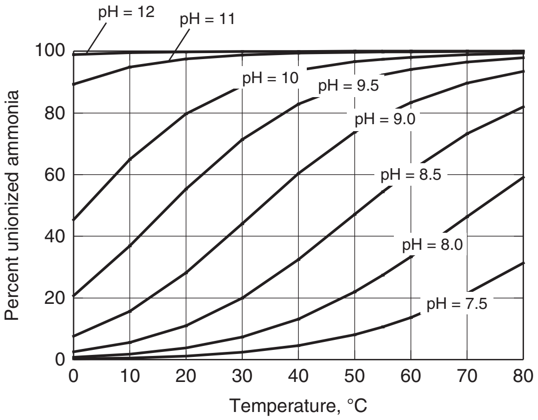


Figure 15–5

Air stripping–acid absorption process for recovery of ammonia and production of concentrated ammonium sulfate solution:
(a) process flow diagram and
(b) full-scale process at the VEAS wastewater treatment plant, Oslo, Norway (courtesy of Paul Sagberg and VEAS—Vestfjorden Avløpsselskap).

Figure 15–6

Ammonia-ammonium equilibrium as a function of pH and temperature.



ion-ion interactions. The impact of non-ideal conditions on ammonium-ammonia equilibrium is accounted for through the use of monovalent and divalent activity coefficients as multipliers on the ion concentrations. The calculation of activity coefficients is described in detail in Sec. 2–2 in Chap. 2.

Temperature Impacts. From the plots on Fig. 15–6 it would appear that as sidestream temperature increases, the required operating pH would be lower, resulting in a lower alkaline chemical requirement. However, as noted in Sec. 15–1, ammonium exists in digestion sidestream as ammonium bicarbonate in the pH range of 7.0 to 8.0. A shift from ammonium to ammonia due to an increase in temperature would be accompanied by the shift of bicarbonate to $CO_{2,\ aq}$ and subsequent release of CO_2 into the gas phase. Air stripping and acid absorption columns are typically designed as a closed air system, as illustrated on Fig. 15–5, to minimize the release of contaminated air to the atmosphere and eliminate evaporative cooling of the sidestream, which would impact the ammonia stripping efficiency. Because CO_2 would not be removed from the air with the ammonia in the absorption column, the CO_2 concentration in the air loop increases, resulting in the retention of CO_2 in the sidestream and pH suppression. Therefore, for an enclosed system, an elevated operating temperature cannot be used to reduce the operating pH and the alkaline chemical demand is unaffected.

Operationally, higher column temperatures are advantageous, as illustrated on Fig. 11–62 in Chap. 11. Higher temperatures increase the Henry's Law coefficient for ammonia, creating a higher driving force for mass transfer in the air stripping column. In addition, ammonia diffusivities in water and air increase with temperature, further increasing the mass transfer rate, mathematically expressed as the overall mass transfer coefficient, K_L, in the stripping column design calculations shown in Sec. 11–10 in Chap. 11. In total, higher operating temperatures have the net effect of lowering the air flowrate required to achieve the same ammonia removal efficiency (see Fig. 11–63). A reduction in the air requirement has the further effect of reducing the diameters of the stripping and absorption columns.

Alkaline Chemical Demand. Caustic soda (sodium hydroxide) and lime have been used in practice for adjusting the pH of digestion sidestream. The two primary reactions that govern the caustic soda requirement are

$$NH_4HCO_3 + NaOH \rightarrow NH_3 + H_2O + NaHCO_3 \qquad (15\text{–}3)$$

$$NaHCO_3 + NaOH \rightarrow Na_2CO_3 + H_2O \qquad (15\text{–}4)$$

Due to the solubility of sodium carbonate, the pH-adjusted sidestream contains a significant buffering capacity, allowing the air stripping column to operate at the desired pH throughout the column depth. For wastewater treatment facilities that require alkalinity addition to sustain nitrification, the sidestream processed by air stripping–acid absorption where caustic soda is used for pH adjustment can provide a portion or all of the alkalinity requirement for the secondary treatment system. Because the cost of caustic soda is a significant portion of the operating cost for an air stripping–acid absorption process, assuming partial or total credit for the alkalinity provided to the secondary treatment system is an important economic consideration.

Alternatively, lime (CaO) can be used for pH adjustment. The primary benefit of lime is the significantly lower cost compared to caustic soda. However, as described in Sec. 6–3 in Chap. 6, lime removes carbonate alkalinity through the precipitation of calcium carbonate. If a beneficial use of the calcium carbonate cannot be found (e.g., beneficially land applied as *Farmer's Lime*), disposal of the solids may result in an additional operating cost, offsetting the cost advantage lime has over caustic soda. The substantial loss in buffering capacity results in a decreasing pH gradient through the depth of the air stripping column as ammonia is stripped from the sidestream, reducing the effectiveness of the lower section of the column. The use of lime has the added disadvantage of increasing the air stripping column fouling rate, thereby increasing the frequency at which the air stripping column is removed from service for cleaning. Lime is also more difficult to handle in comparison to caustic soda.

Solids Removal for Enhanced Air Stripping. Digestion sidestreams contain TSS (see Table 15–1), which can cause significant fouling in the air stripping column. In addition, during pH adjustment, inorganic precipitates will form such as calcium carbonate, although the solids mass generated by pH adjustment with caustic soda will be far less than the amount of chemical sludge generated by lime. To reduce the negative impact of the TSS on the air stripping column, a solids removal step is recommended such as an inclined plate settler or a high rate clarification technology (see Sec. 5–7 in Chap. 5 for process descriptions). If lime is used for pH adjustment, the solids removal process is essentially conventional cold lime softening, which is commonly practiced in water treatment.

Air Stripping and Acid Absorption Column Operations. The basic design approach for determining the dimensions and packing depth of the air stripping column were introduced in Sec. 11–10 in Chap. 11. The design of the acid absorption column is more complicated because, in addition to mass transfer, a chemical reaction is occurring in the acidic ammonium salt solution recirculated through the column and heat is being generated, which affects the liquid and air temperatures in the column. An advanced chemical process model such as ASPEN+ or a similar modeling software package are typically employed to calculate the scrubbing efficiency, packing depth, column diameter, and temperature of the air returned to the air stripping column. Alternatively, an equipment provider may have a proprietary design method based on their experience that is specific to the type of mass transfer media being used in the column.

As illustrated on Fig. 15–5, the two columns contain randomly placed plastic medium and a closed air system where air blown through the stripper column, countercurrent to the sidestream flow, is sent to the absorber and air from the top of the absorber flows to the inlet of the fan. Sizing of the two columns, the fan and product recirculation pump is dependent on the sidestream flow and ammonia concentration, the desired sidestream ammonia removal efficiency and the operating conditions in the two columns (pH, temperature). Typically, the diameters of the stripper and absorption columns are equivalent, but the height of the absorption column and its associated packing depth is roughly 80 percent of stripping column. As the operating temperature increases beyond 40°C, the absorption column height and packing depth will increase and approach those of the stripping column. The air stripping–acid absorption process is typically controlled by pH (pretreated sidestream and recirculated product solution in absorption column) and the density of the recirculated product solution.

In the absorption column, acid is dosed into the product solution recirculation loop to maintain an acidic pH, which provides sufficient driving force for absorption. Real time measurement of the recirculated product solution density is used to fine tune pH control. The product solution is removed continuously from the bottom of the column. For the production of ammonium sulfate, the water content of the 93 percent sulfuric acid (by weight) typically used in larger applications is insufficient to meet the water requirement of a 40 percent (by weight) ammonium sulfate solution. Therefore, water is added continuously to the absorption column, primarily controlled by the density of the recirculated solution. The addition of concentrated sulfuric acid and the subsequent formation of ammonium sulfate results in significant chemical heat generation. A portion of this heat can be used to elevate the temperature of both columns by allowing the temperature of the absorption column to increase, which transfers heat to the air being recirculated back to the stripper column. Less chemical heat is generated with the use of concentrated nitric acid to produce ammonium nitrate.

The direct addition of low pressure waste steam to the stripper column can also be done as higher operating temperatures reduce the fan capacity and the column diameters up to a temperature limit of approximately 70°C. Beyond this temperature, the ammonia removal efficiency in the absorption column will begin to deteriorate and this is the temperature limit for the fiberglass reinforced plastic used for column construction.

Economic Considerations. As discussed in the overview of the air stripping–acid absorption process, concentrated inorganic acid is used in the absorption column and caustic soda or lime is used for pH adjustment of the digestion sidestream. Because these chemicals represent a major operating cost, the future pricing trends of these chemicals are critically important information to obtain to assess the economic viability of this process, in comparison to biological treatment. To offset the costs of the caustic soda or lime and the acid, selling the product to generate revenue will improve the economics of ammonia recovery. The use of ammonium sulfate or ammonium nitrate as a fertilizer is discussed further in Sec. 15–6.

Recovery of Ammonia by Steam Stripping

The use of steam to volatilize ammonia from water is practiced at several industrial installations, however, the implementation of steam stripping for municipal sidestreams has been limited. The only performance data reported are from pilot or demonstration-scale studies (Teichgräber and Stein, 1994; Gopalakrishnan et al., 2000). An ammonia concentration of 100 mg N/L in the steam-stripped sidestream is the practical limit for the process with energy consumption and associated operating cost being the limiting condition.

Process Description. As shown on Fig. 15–7, the steam stripping process consists of contacting sidestream with low pressure steam in a packed column with random-dumped media. At an operating temperature in the range of 95 to 100°C, ammonium bicarbonate is decomposed thermally into ammonia and carbon dioxide in the column and the dissolved gasses are subsequently stripped from the sidestream into the vapor phase.

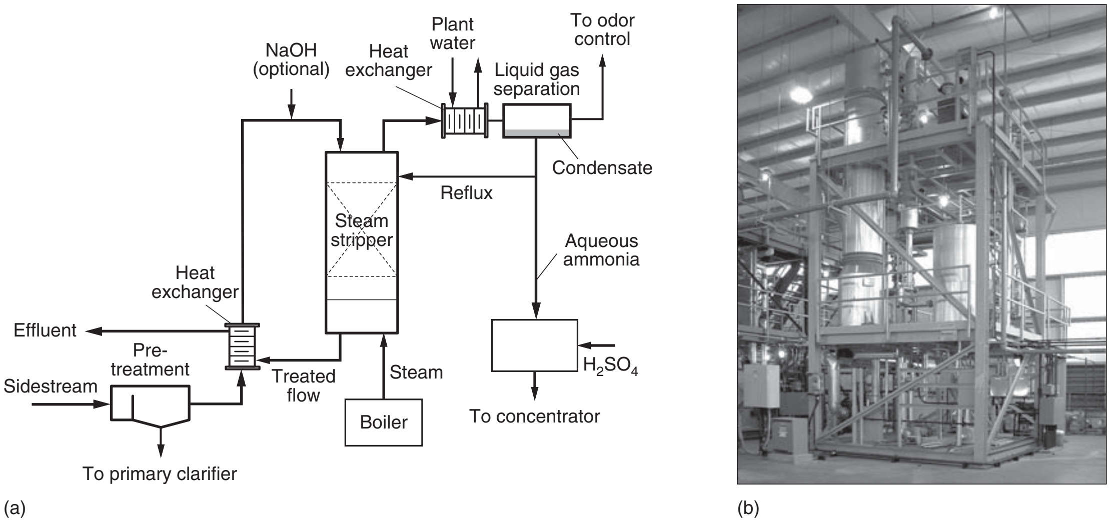


Figure 15–7

Steam stripping process for recovery of ammonia from sidestream for production of aqueous ammonia or ammonium sulfate: (a) process flow diagram and (b) view of pilot-plant facility (courtesy of The New York City Department of Environmental Protection).

The stripped sidestream is used to preheat the sidestream influent to reduce energy consumed by the process.

The vapor phase from the column is cooled in a condenser to a two-phase mixture using plant effluent as single-pass cooling water. The liquid and gas phases are separated and a portion of the liquid is returned to the stream stripping column ("reflux"). The remainder of the ammonia-rich liquid is processed further to produce concentrated aqueous ammonia or the liquid is neutralized with sulfuric acid to produce an ammonium sulfate or ammonium nitrate solution, which is further processed to generate a concentrated solution for reuse as a fertilizer. Due to the high volatility of carbon dioxide at the condenser temperature, the majority of the carbon dioxide remains in the gas phase, limiting the reformation of ammonium bicarbonate in the liquid. The gas is highly odorous and requires treatment.

Volatilizing CO_2 from the preheated sidestream in a "decarbonization" packed column prior to the steam stripping column may be beneficial. By removing the majority of the CO_2 before the stripping column, a pH of 9.5 to 9.9 can be sustained throughout the full packing depth resulting in a higher mass transfer rate and lower steam demand (Teichgräber and Stein, 1994). Steam is added to the decarbonization column to increase the sidestream temperature and enhance CO_2 volatilization, but the steam volume requirement does not induce significant ammonia stripping.

Energy Requirements. Approximately 0.15 to 0.18 kg of low pressure steam is required per kg of sidestream to achieve a stripped sidestream ammonia concentration of approximately 100 mg N/L, if energy is recovered from the stripped sidestream to preheat the sidestream influent. The steam demand will be lower if CO_2 is removed from the preheated sidestream in a decarbonization step. Additional energy is consumed to process the low strength aqueous ammonia to a higher concentration or to produce higher strength ammonium sulfate solution for reuse.

Chemical Requirements. Because ammonium bicarbonate is thermally decomposed into ammonia and carbon dioxide, sodium hydroxide is not a strict requirement in the process. However, addition of sodium hydroxide will increase pH throughout the packed column, which will increase the mass transfer rate and lower steam demand. In the process configuration where CO_2 is volatilized in a decarbonization column, sodium hydroxide addition to the steam stripping column may also lower steam demand.

Pretreatment Requirements. Reduction of suspended solids and colloidal material in the sidestream to less than 100 mg/L is advantageous to reduce fouling of the column media.

Off-gas Treatment. The CO_2-rich gas from the condenser (or decarbonization column) contains highly odorous reduced sulfur compounds and requires treatment. Volatile organic compounds will also be present in the off-gas, which also may require treatment, depending on the air quality regulations applicable to the facility.

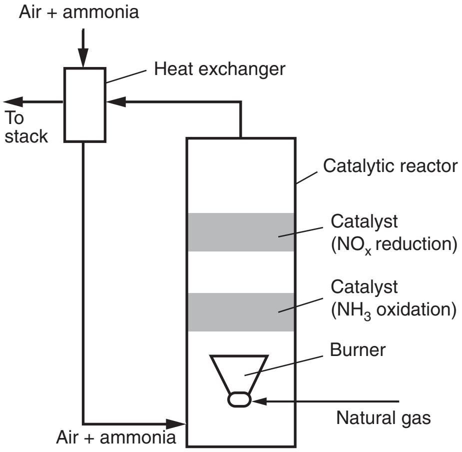
Operating Problems. Operating problems that have been encountered with steam stripping include: (1) extensive fouling (iron deposits, for example) within the heat exchanger and in the stripper due to the presence of waste constituents at elevated temperatures, (2) maintaining the required pH for effective stripping, (3) controlling the steam flow, and (4) maintaining the stripping tower temperature. Because of the importance of temperature, steam stripping should be carried out in enclosed facilities. Spiral wound heat exchangers have proven to be effective. In Europe, an acid wash is used to clean the piping, the heat exchanger, and the stripping column.

Air Stripping with Thermocatalytic Destruction of Ammonia

As an alternative to capturing the stripped ammonia to produce a fertilizer, the ammonia-laden air from the stripping column can, as shown on Fig. 15–8, be subjected to thermocatalytic oxidation (TCO). In TCO up to 98 percent of the ammonia is oxidized selectively to N_2 (~95 percent selectivity) with the balance of the oxidation products primarily being N_2O or NO, depending on the catalyst temperature. This process has not been applied to municipal sidestreams, but has been used for landfill leachate treatment applications where waste heat is available from the combustion of landfill gas to meet the thermal demand of the stripping column (Organics Limited, 2009).

Figure 15–8

Schematic of catalytic reactor used for the thermocatalytic destruction of ammonia in air.



The air stripping column is operated similarly to the column in the air stripping–acid absorption process, only the air is single-pass ambient air that must be heated via steam injection to compensate for evaporative cooling of the column to maintain the desired temperature. In full-scale applications on landfill leachate, sufficient waste heat is available to increase the stripping column temperature high enough to induce thermal decomposition of ammonium bicarbonate, similar to the steam stripping process (Organics Limited, 2009). Consequently, the caustic soda requirement is greatly reduced or eliminated and the air requirement for stripping is reduced.

The ammonia-laden air is then subjected to oxidation at temperatures in the range of 288 to 316°C (550 to 600°F) with a catalyst typically composed of mixtures of transition metals or transition metal oxides on silica or alumina supports and doped with a noble metal such as platinum. The reaction is exothermic (generates heat), reducing or eliminating the natural gas input to the oxidizer required to sustain the target temperature, depending on the ammonia concentration in the air. The primary reactions of interest in this process are:

$$NH_3 + 0.75\ O_2 \rightarrow 0.5\ N_2 + 1.5\ H_2O \tag{15–5}$$

$$NH_3 + O_2 \rightarrow 0.5\ N_2O + 1.5\ H_2O \tag{15–6}$$

$$NH_3 + 1.25\ O_2 \rightarrow NO + 1.5\ H_2O \tag{15–7}$$

Temperatures above 350°C (660°F) are avoided to limit the formation of NO_2. Even when operating within the desired temperature range, a certain portion of the ammonia will convert to NO, which must be reduced by selective catalytic reduction (SCR). Unoxidized ammonia will react with NO in the downstream SCR, but there is likely a need for supplemental addition of urea or anhydrous ammonia to maximize NO removal. Exhaust from the TCO/SCR process is used to preheat the ammonia-laden air to the TCO.

The application of air stripping/TCO is limited to cases where waste heat is available from gas engine exhaust or can be generated from otherwise unused digester gas. For leachate treatment applications, the energy demand has been estimated at 450 MJ per m^3 of leachate (Organics Limited, 2009). Depending on the ammonia concentration in the inlet to the TCO, the injection of natural gas to the TCO may also be required to maintain the temperature within the required range. An energy balance across the stripping/TCO system and a life cycle cost analysis is required to assess the economic viability of the process.

15–6 BENEFICIAL USE OF RECOVERED PHOSPHATE AND AMMONIUM PRODUCTS

Phosphate and ammonia recovered from sidestream can be reused as fertilizers or for other industrial applications. The principal products recovered from sidestreams include:

1. Magnesium ammonium phosphate hexahydrate (struvite)
2. Calcium phosphate (hydroxyapatite)
3. Ammonium sulfate
4. Ammonium nitrate

The beneficial use of these products is considered in this section.

Magnesium Ammonium Phosphate Hexahydrate (Struvite)

Struvite has been recognized as a fertilizer for over 150 years and is considered a slow-release fertilizer due to its low water solubility. The low dissolution rate into the soil limits high soluble nutrient concentrations around the root structure and the occurrence of "fertilizer burn," and minimizes the loss of nutrients due to surface runoff and groundwater percolation.

Struvite is only a minor contributor to the overall slow-release fertilizer market, dominated by slowly soluble urea-aldehyde reaction products and polymer and sulfur-coated controlled-release fertilizers. Due to high manufacturing costs, slow-release fertilizers have been predominately used in high valued-added applications in non-agricultural markets such as nurseries, greenhouses and golf courses. Struvite is currently available in a blend with potassium magnesium phosphate under the tradename of MagAmp or MagAmp®-K, typically defined as a 7 percent N, 40 percent P_2O_5, 6 percent K_2O fertilizer with 12 percent Mg.

Demand for slow-release fertilizers has been increasing world-wide to improve fertilizer utilization efficiency and limit the release of nutrients into water bodies. Consequently, struvite recovered from municipal sidestreams is being viewed increasingly as an asset rather than a nutrient that must be treated and disposed. As presented in Sec. 15–4, several struvite crystallization and recovery processes are in full-scale operation. Product purity reported from these facilities has been high, generally greater than 99 percent, with low heavy metal concentrations and coliform counts after the product is washed and decontaminated (Nawa, 2009; Baur et al., 2011; Moerman, 2011). Purity requirements and product accreditation vary by country, state or region, but the product purities have not limited their introduction into regional markets. The product is not considered an organic fertilizer, as it is derived from a wastewater treatment facility.

The recovered struvite may be sold directly by the municipality to a fertilizer blender who markets slow-release fertilizers, but it is a common business model by struvite crystallizer technology providers to assume this responsibility as part of their contract with the municipality. The fertilizer company will blend the struvite with other chemicals to create the desired nutrient blend for a specific application. The fertilizer application may also dictate the required physical characteristics of the product. For example, a pelletized product with a large mean diameter may be specified for a particular application. Consequently, the crystallizer operating conditions and processing requirements for the harvested product will be adjusted to generate a product that complies with this specification. Alternatively, the harvested product will be processed offsite to generate the desired physical characteristics.

Calcium Phosphate (Hydroxyapatite)

Although the recovery of phosphorus from municipal sidestreams as hydroxyapatite has fallen out of favor due to chemical pretreatment costs and the more favorable economics of struvite recovery, hydroxyapatite will continue to be favorable for certain industrial and dairy waste streams. If the use of calcite and calcium silica hydrate as seed materials for hydroxyapatite crystallization is proven to be successful at a full-scale, the production of hydroxyapatite may become a viable alternative to struvite production from municipal sidestreams.

As demonstrated in The Netherlands in the 1990s, hydroxyapatite can be crystallized from sidestream using a seed material such as sand to generate a granular product of high purity and low heavy metals content (Piekema and Giesen, 2001). The sand content of the harvested product is 5 percent by weight or less. The product can be used as feed stock by a phosphate rock processor to generate other phosphate compounds for the fertilizer market such as $Ca(HPO_4^{2-})_2$ (*Superphosphate*), it can also be blended with other nutrients to produce a formulated fertilizer or applied directly as a slow-release fertilizer. Similar to struvite, the pelletized product can be easily dewatered and stored.

Ammonium Sulfate

The primary use of ammonium sulfate is as a fertilizer, but there is no established market for ammonium sulfate produced from wastewater treatment plants. However, there is an established market for ammonium sulfate as a byproduct of various industries, primarily from the production of nylon. The product is typically marketed worldwide and produced in a crystallized form due to customer preference and to reduce shipping costs. Ammonium sulfate can be used in a direct application, it can be blended in custom fertilizer solutions, and it can be blended with biosolids as discussed below. For example, ammonium sulfate has been blended with urea ammonium nitrate (UAN) to produce a fertilizer solution that will increase the sulfur content of the solution fertilizer blend. While the use of sulfur-containing fertilizers are increasing due to the substantial reduction in atmospheric sulfur deposition in certain regions, the use of nitrate and anhydrous ammonia based fertilizers are decreasing due to security issues, ammonia volatilization, and a general shift to liquid fertilizers.

Use of Ammonium Sulfate (AS). Although used throughout the year, the highest use of AS (50–75 percent of annual usage) will typically occur in the late spring and early summer. Smaller amounts would be used in late summer and in the fall. For the remainder of the year (approximately six months), AS must be stored. Typically, a fertilizer blender/distributor will have storage capability, thus avoiding the need to install storage tanks at the wastewater treatment facility.

Blending Ammonium Sulfate with Biosolids. If biosolids are to be reused beneficially, the 40 percent by weight AS solution can be mixed with the dewatered solids to enhance N and S content. However, there are several potential concerns with this approach:

1. Blending the AS solution with the biosolids will the increase the mass of the biosolids and its associated hauling costs.
2. Where biosolids are land applied at an agronomic rate based on nitrogen content, the acreage needed will be larger as the nutrient content will be higher.
3. Increased potential for hydrogen sulfide odors: Because the solution has a high level of sulfate, there is a potential to generate hydrogen sulfide if the mixed biosolids becomes anaerobic during storage.

Marketing Ammonium Sulfate. The ammonium sulfate supply chain contains three distinct groups: manufacturer, blender/distributor, and applier/end-user. A treatment facility would sell their product, as a manufacturer, to a blender/distributor for them to either incorporate into a fertilizer blend or sell it as is.

The most critical issue with marketing AS for the fertilizer suppliers and farmers is its quality and consistency. The fertilizer/chemical suppliers generally will not accept the AS unless the quality and consistency are known and meet their specifications. However, product quality is anticipated to be higher than the standard specifications based on current experience with ammonium nitrate (Sagberg, 2006) and ammonium sulfate (ThermoEnergy, 2009), where the products were found to have low levels of heavy metals (<1 ppm and largely attributed to the quality of the commercial grade acids used in the process) and low total organic carbon (TOC < 50 ppm). Typically, TOC content is tightly specified for ammonium nitrate and monitored to limit explosion potential, but this would not be a factor with ammonium sulfate. Based on the data from operating processes, the TOC in the product is most likely methylamine, which is stripped from the digestion sidestream and absorbed into the product solution along with the ammonia.

Ammonium Nitrate

Ammonium nitrate is an important fertilizer used extensively throughout the world and is commercially available in both dry and liquid forms. Because ammonium nitrate can be used in explosives, the handling and purchase of dry ammonium nitrate is strictly regulated to prevent misuse. This regulation would not apply to "liquid" form, and liquid usage is increasing substantially. However, most nitric acid capacity is "captive," i.e., manufactured for internal use, and, therefore, the price and availability are uncertain. Based on experience in Norway, the fertilizer manufacturer would supply the nitric acid to the wastewater treatment plant and take the ammonium nitrate product; however, this arrangement results in a less cost competitive environment compared to ammonium sulfate production as the fertilizer manufacturer has more extensive control of the acid supply and purchase of the ammonium nitrate.

15–7 BIOLOGICAL REMOVAL OF NITROGEN FROM SIDESTREAMS

Sidestreams resulting from sludge and biosolids processing are most commonly treated in the mainstream plant. Typically, the sidestreams are returned to the headworks, the inlet of primary sedimentation tanks or directly to a location near or in the secondary treatment process (e.g., channels conveying primary effluent or RAS to the activated sludge reactors), depending on the nutrient, BOD and suspended solids loadings associated with the sidestream, and physical constraints such as the plant piping configuration and the location of the dewatering process relative to the process units associated with the main liquid treatment train.

Treatment of the nitrogen-rich sidestreams derived from the dewatering of digested solids, a principal focus of this chapter, can occur in separate treatment processes or in a treatment processes that are integrated with the mainstream treatment process. Both separate and integrated treatment processes are introduced and discussed in this section. The biological treatment processes used for the removal of nitrogen, introduced in this section, are described in greater detail in the subsequent three sections.

Nitrogen Removal Processes

Inorganic nitrogen can be removed biologically by three general processes:

1. Nitrification-denitrification
2. Nitritation-denitritation
3. Partial nitritation-anaerobic ammonium oxidation (deammonification)

Although these processes have been described previously in Chaps. 7 through 10 for the treatment of wastewater, the discussion in this chapter deals with the application of these processes for treatment of sidestreams containing high concentrations of ammonium. For the purpose of comparison, the pathways involved in each of these processes are illustrated on Fig. 15–9 and described below.

Nitrification-Denitrification. In the nitrification-denitrification process, as illustrated on Fig 15–9(a), ammonium is first oxidized to nitrite (nitritation) and subsequently to nitrate (nitratation). Ammonia oxidizing bacteria (AOB) and nitrite oxidizing bacteria (NOB) are responsible for the two steps in the nitrification process. In the denitrification process nitrate is first reduced to nitrite and subsequently to nitrogen gas. As shown on Fig. 15–9(a), oxygen must be added to complete the oxidation of ammonia to nitrate and a carbon source must be available to complete the reduction of nitrate to nitrogen gas.

Nitritation-Denitritation. In the nitritation-denitritation process, as illustrated on Fig 15–9(b), ammonium is first oxidized to nitrite (nitritation). In the next step, nitrite, under anoxic conditions, is reduced to nitrogen gas (denitritation). Short-circuiting the nitrification-denitrification pathway, as illustrated on Fig. 15–9(b), through the restriction

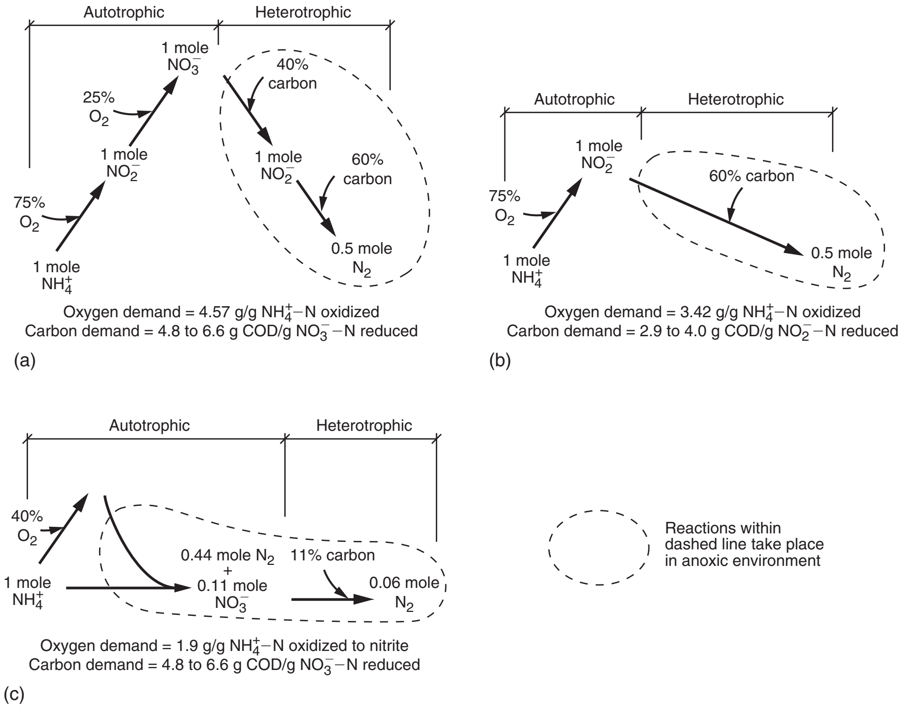


Figure 15–9

Biological reaction pathways for ammonium oxidation and inorganic nitrogen removal: (a) nitrification-denitrification, (b) nitritation-denitritation; and (c) deammonification.

or prevention of nitrite oxidation to nitrate, reduces the stoichiometric oxygen demand and associated aeration energy by 25 percent. In the subsequent anoxic step, the heterotrophic bacteria require 40 percent less degradable organic carbon to reduce nitrite to nitrogen gas. Reducing the aeration power and chemical requirements for sidestream treatment has been the primary driver in the development of alternative advanced biological treatment processes.

Partial Nitritation-Anaerobic Ammonium Oxidation (Deammonification). In the deammonification process, as illustrated on Fig 15–9(c), a portion of the ammonium is first oxidized to nitrite (partial nitritation). In the next step of the process, ammonia and nitrite are converted to nitrogen gas and nitrate under oxygen-free conditions (anaerobic ammonium oxidation) by a special group of autotrophic bacteria collectively known as anaerobic ammonium oxidizers (Anammox). As illustrated on Fig. 15–9(c), deammonification further reduces the organic carbon requirement.

Process Design Considerations. Although there are distinct differences in the three different nitrogen removal processes, there is commonality in the design approaches used to determine aeration, chemical, heat removal and tank volume requirements. A unified section on process design for the three major types of sidestream processes is presented in Sec. 15–11. Additional information on the specific design requirements for proprietary processes is provided in the process descriptions in Sec. 15–8 through 15–10.

Separate Treatment Processes for Nitrogen Removal

A sidestream treatment process is considered *separate* from the mainstream plant if the mixed liquor suspended solids in the sidestream reactor are isolated from the mainstream secondary treatment process [see Fig. 15–10(a)]. Waste solids from the sidestream treatment

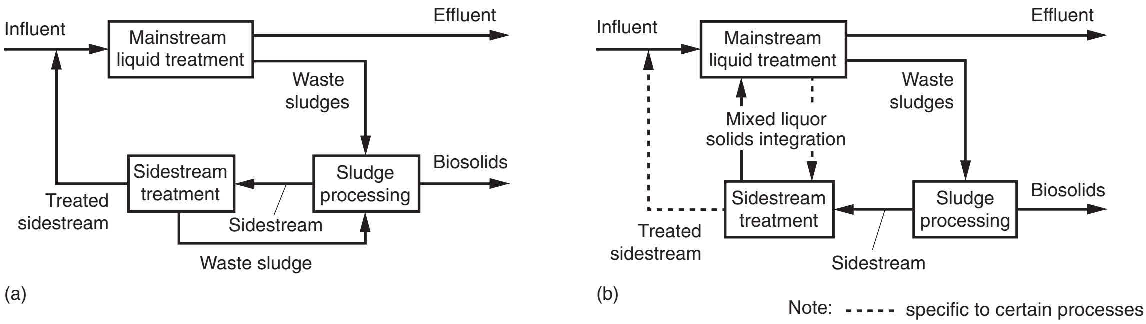


Figure 15–10

Biological process definitions: (a) separate sidestream treatment and (b) integrated sidestream-mainstream treatment.

process, as shown on Fig. 15–10(a), are sent directly to the solids processing train and no mixed liquor suspended solids from the mainstream secondary process are constantly or intermittently sent to the sidestream process.

Advantages and Disadvantages of Separate Sidestream Treatment. Biological treatment in a dedicated separate reactor presents two primary advantages. First, the nutrient and particulate loadings associated with the sidestream are greatly diminished, resulting in a stream that is less likely to impact the performance of the main treatment plant. Second, treatment in a dedicated reactor provides an opportunity to operate under conditions that can limit the final nitrification product to nitrite. This mode of operation results in lower oxygen and COD requirements for inorganic nitrogen removal (Sec. 15–9); it also provides an ideal environment for the growth of bacteria that possess unique characteristics and biochemistry that can be exploited to further reduce the cost of inorganic nitrogen removal (Sec. 15–10).

A separate sidestream treatment process may also be advantageous or cost-effective for facilities where a required marginal reduction in the plant effluent total nitrogen (TN) concentration or loading can be achieved by treating the sidestream alone and bioaugmentation via an integrated sidestream treatment process does not provide any distinct advantages in terms of overall plant performance (integrated systems are described below). The cost effectiveness of separate sidestream treatment, in comparison to alternative secondary process upgrades that achieve the same improvement in TN removal, is dependent on site specific factors unique to the specific facility, e.g., available footprint for plant expansion, constructability of the various plant upgrade options.

Historically, many separate sidestream treatment processes were designed to provide only nitrification. In these cases, the mainstream plant has limited nitrification capacity and full or partial nitrification of the sidestream ammonium load was the most cost effective option.

From an operations perspective, a separate sidestream treatment system is an additional process at the facility that requires operational oversight and maintenance, which may not be desirable. Also, some separate sidestream nitrification-denitrification processes require the addition of an alkalinity source, a supplemental COD source or both. Chemical consumption would only be a disadvantage if other sidestream nitrification-denitrification options do not require chemical addition for the facility to achieve the desired effluent discharge quality.

Impact of Other Treatment Processes. However, despite the isolation of the sidestream mixed liquor solids from the mainstream secondary process, the sidestream treatment process can be affected by the operating conditions of the other process units in the facility. For example, if the operation of the anaerobic digesters is disturbed or mechanical or performance problems are encountered with the post-digestion dewatering process, the quality and quantity of the sidestream may change, potentially resulting in a perturbation in the sidestream treatment process operating conditions and performance. However, as discussed in Sec. 15–2, the inclusion of sidestream equalization and solids removal will minimize the impact on the downstream process. Because the quality of the effluent from the sidestream treatment process is typically not sufficient to allow blending with the mainstream secondary effluent, the treated sidestream is sent to the mainstream secondary process for further treatment.

Integrated Sidestream-Mainstream Treatment and Bioaugmentation

A sidestream treatment process is defined as "integrated" when mixed liquor suspended solids in the mainstream secondary and sidestream processes are interchanged or nitrifier-enriched waste solids from the sidestream process are fed to the mainstream secondary process to induce bioaugmentation. An integrated configuration, as illustrated on Fig. 15–10(b), would be represented by a flow scheme where a portion of a mainstream return activated sludge (RAS) is fed to the sidestream reactor or, similarly, the sidestream is fed to a RAS reaeration tank. An integrated system would also be represented by any process where sidestream reactor waste solids are fed to the mainstream secondary process.

Advantages and Disadvantages of Integrated Sidestream Treatment. The main advantage of integrating the sidestream and mainstream processes is the enhancement and stabilization of mainstream nitrification with nitrifier-enriched mixed liquor solids from the sidestream reactor. The optimum configuration is dependent on the mainstream process configuration, operating conditions, and the desired plant effluent quality. Operating costs associated with chemical addition are also a consideration during process selection. A disadvantage of integrating the sidestream and mainstream processes is the introduction of additional inert solids into the secondary process mixed liquor; however, if the sidestream is equalized and the majority of the solids are removed and recycled to the solids processing train, the effect of the sidestream on the mainstream process solids balance is greatly diminished.

Augmentation of Mainstream Treatment Process. The augmentation of the mainstream process with nitrifier-enriched sidestream mixed liquor solids has been documented, yet a unified mechanistic model that can be used to predict the impact of bioaugmentation for all process configurations has not yet been developed. In general, the effectiveness of nitrifying biomass grown in a sidestream reactor in enhancing nitrification performance in the mainstream secondary process may be greatly impacted by the bulk liquid environment in which these organisms grow (e.g., osmotic pressure, temperature, pH, ionic composition and strength, substrate concentration). In integrated configurations where the sidestream is fed to a RAS reaeration tank, the nitrifying bacteria grow within the flocculated mixed liquor suspended solids in the RAS reaeration tank at the same temperature and nearly identical bulk liquid conditions as the mainstream secondary process; therefore, complete retention of nitrifier activity is anticipated as the nitrifier-enriched mixed liquor passes from the reaeration tank to the activated sludge tanks.

In process configurations where the difference in the bulk liquid sidestream and mainstream process environments are significant, the sidestream reactor operating conditions allow the growth and domination of specific types of nitrifying organisms over others, yet when placed in the mainstream environment, are at a competitive disadvantage to the nitrifying organisms that dominate under the mainstream operating conditions. Intuitively, the bioaugmentation effect would be lower with these configurations in comparison to the configurations where sidestream is substantially diluted into a RAS stream and only increase as the sidestream and mainstream processes are increasingly integrated through interchange of mixed liquor solids. However, an exact quantification of the effect has not been thoroughly demonstrated over a wide range of operating conditions.

Bioaugmentation Effect of Integrated Sidestream Treatment. The bioaugmentation effect of the sidestream mixed liquor solids is also impacted by the nitrifier mass discharge rate to the mainstream process. Therefore, operating conditions that lead to a reduced nitrifier mass in the sidestream process will reduce the mass in the waste sludge. For example, a higher operating temperature in the sidestream process will increase the decay rate of the nitrifying organisms. Enhanced decay also occurs as the sidestream process SRT is allowed to increase well above the minimum SRT required to sustain the desired sidestream treatment performance. In sidestream processes that operate at elevated temperatures and a high SRT, the combined effect of these two conditions will diminish the nitrifier mass available for bioaugmentation. Based on these observations operating the sidestream treatment reactor at (1) a temperature as close as practically possible to the mainstream process, (2) with a bulk liquid environment similar to the mainstream process and (3) at a low SRT that provides the desired performance will result in optimum retention of nitrifier activity and, hence, the greatest bioaugmentation effect.

15–8 NITRIFICATION AND DENITRIFICATION PROCESSES

Several biological treatment processes have been developed for sidestream treatment where the end product of ammonium oxidation is primarily nitrate and the nitrate is subsequently denitrified in part or in whole, depending on the treatment objective. The processes are subdivided into two categories: separate treatment and integrated sidestream-mainstream treatment, according to the definitions provided in Sec. 15–7. The implementation of one type of process over the other is dependent on the type of mainstream secondary process, its operating conditions, plant effluent quality objectives and economic considerations. The purpose of this section is to consider the application of nitrification and denitrification processes for the treatment of sidestreams.

Fundamental Process Considerations

To understand the nitrification-denitrification process, it is useful to consider (1) the process biology, kinetics, and stoichiometry, (2) the alkalinity requirements, (3) the importance of inorganic carbon, and (4) the need for degradable organic carbon.

Process Biology, Kinetics, and Stoichiometry. The microbiology, basic biochemical reaction stoichiometries and autotrophic growth kinetics associated with the oxidation of ammonium to nitrite and nitrite to nitrate are presented in Sec. 7–9 in Chap. 7. Although kinetic rates in sidestream biological treatment processes are influenced by the same environmental conditions as the mainstream nitrification processes, differences between ammonia oxidizing bacteria (AOB) and nitrite oxidizing bacteria (NOB) growth kinetics in the sidestream and mainstream environments must be considered for sidestream process design and operation. Specifically, the key differences are the impact of temperature on growth and decay, nitrous acid and free ammonia inhibition of AOB and NOB populations, and the effect of bicarbonate concentration on the autotrophic growth rates.

Biological denitrification reaction stoichiometry and kinetics are presented in Sec. 7–10 in Chap. 7. The pertinent stoichiometric reactions involved in nitrification-denitrification process, repeated here for convenience, are as follows.

Nitrification. The principal nitrification reactions are as follows.
Biological conversion of ammonium to nitrite

$$2NH_4^+ + 3O_2 \rightarrow 2NO_2^- + 4H^+ + 2H_2O \tag{7–88}$$

Biological conversion of nitrite to nitrate

$$2NO_2^- + O_2 \rightarrow 2NO_3^- \tag{7–89}$$

Total oxidation reaction:

$$NH_4^+ + 2O_2 \rightarrow NO_3^- + 2H^+ + H_2O \tag{7–90}$$

If cell tissue is neglected, the amount of alkalinity required to carry out the ammonium oxidation reaction is given by the following reaction, obtained by rewriting Eq. (7–90) as follows.

$$NH_4^+ + 2HCO_3^- + 2O_2 \rightarrow NO_3^- + 2CO_2 + 3H_2O \tag{7–91}$$

When cell mass synthesis is included in the overall oxidation of ammonium to nitrate, Eq. (7–91) becomes Eq. (7–93). The cell mass yield is based on yields of 0.12 g VSS/g NH_4-N and 0.04 g VSS/g NO_2-N for the nitritation and nitratation reactions, respectively.

$$\begin{aligned} &NH_4HCO_3 + 0.9852NaHCO_3 + 0.0991CO_2 + 1.8675O_2 \rightarrow \\ &\qquad 0.01982C_5H_7NO_2 + 0.9852NaNO_3 + 2.9232H_2O + 1.9852CO_2 \end{aligned} \tag{7–93}$$

Denitrification. The amount of biodegradable organic compound required to reduce nitrate to nitrogen gas is dependent on the carbon source as illustrated by the following equations where the organics in wastewater, methanol, or acetate are used for nitrate reduction.

Wastewater:

$$C_{10}H_{19}O_3N + 10NO_3^- \rightarrow 5N_2 + 10CO_2 + 3H_2O + NH_3 + 10OH^- \tag{7–110}$$

Methanol:

$$5CH_3OH + 6NO_3^- \rightarrow 3N_2 + 5CO_2 + 7H_2O + 6OH^- \tag{7–111}$$

Acetic Acid:

$$5CH_3COOH + 8NO_3^- \rightarrow 4N_2 + 10CO_2 + 6H_2O + 8OH^- \tag{7–112}$$

The above equations do not reflect the actual carbon requirement for denitrification as they do not include cell mass synthesis. When expressed as COD, the mass ratio of biodegradable COD to nitrate-N taking into account the biomass yield is given by Eq. (7–126).

$$\frac{\text{bsCOD}}{NO_3\text{-N}} = \frac{2.86}{1 + 1.42Y_n} \tag{7–126}$$

Where 2.86 = oxygen equivalent of nitrate-N (g O_2/g NO_3-N)
Y_n = net biomass yield, as defined by Eq. (7–121).

$$Y_n = \frac{Y}{1 + b(\text{SRT})} \tag{7–121}$$

Where b = heterotrophic anoxic decay rate

Based on these equations, the observed COD/N ratio is dependent on the SRT of the reactor and the value of b, which, in turn, is dependent on the reactor temperature. The reader should refer to Secs. 7–9 and 7–10 for more detailed discussions of the nitrification and denitrification reactions, respectively.

Alkalinity Requirements. The alkalinity requirements are given by Eqs. (7–91) and (7–93). Referring to Eq. (7–91), two moles of bicarbonate are required to neutralize the acidity generated per mole of ammonium oxidized during nitrification. When cell growth is included [see Eq. (7–93)], 1.98 moles of bicarbonate are required for acid neutralization and 0.099 moles of inorganic carbon are required for cell growth in the complete nitrification of one mole of ammonium-N. If nitrate is denitrified completely, 50 percent of the alkalinity destroyed during nitrification is recovered (see Sec. 7–10 in Chap. 7), resulting in a net alkalinity reduction of 1 mole of bicarbonate per mole of ammonium-N nitrified and denitrified (3.57 g $CaCO_3$/g NH_4-N).

For sidestreams derived from anaerobic digestion and ATAD processes, the alkalinity is primarily in the form of bicarbonate and is typically equal to the ammonium concentration on a molar basis, providing only one half of the alkalinity required for complete nitrification. The remaining alkalinity demand can be satisfied through the addition of an external alkalinity source (e.g., caustic soda; see Sec. 15–11), dilution of the sidestream into another stream with sufficient alkalinity (e.g., return activated sludge) or by generating alkalinity via denitrification. In the absence of sufficient alkalinity, a stoichiometric amount of the sidestream ammonium-N will be oxidized to a mixture of nitrite and nitrate in accordance with the available alkalinity.

The addition of a COD source, to enhance denitrification and generate alkalinity to support nitrification, may not eliminate the need for an external alkalinity source. For example, if ferric or ferrous chloride is added to the digesters to control struvite formation or to the sidestream to control struvite formation in the pipe or channel conveying sidestream, a reduction in sidestream alkalinity will occur due to the acidity associated with these chemicals. Therefore, in subsequent sidestream treatment, the addition of supplemental alkalinity may be required depending on the treatment objective.

Importance of Inorganic Carbon. The role of inorganic carbon in nitrifier growth is often ignored in mainstream nitrification processes where bicarbonate and CO_2 are readily available due to the degradation of abundant organic carbon compounds in the plant influent. However, the residual inorganic carbon concentration in separate sidestream reactors has an impact on the ammonium removal rate and removal efficiency due to the high autotrophic growth rates in the reactor. In high rate nitritation-denitritation sidestream treatment processes operated at high temperature (greater than 30°C) at a full-scale, the effect of inorganic carbon concentration on the autotrophic growth rate was not found to follow a conventional Monod kinetic form, but is represented by a logistic function ("S-curve") as defined by Eq. (15–8). Monod kinetic terms for dissolved oxygen, substrate concentration, free ammonia inhibition,and nitrous acid inhibition along with the decay rate and Arrhenius temperature function are not included in Eq. (15–8) for simplicity.

$$\mu_n = \mu_m \frac{e^{[(\mathrm{HCO}_3^- - k)/a]}}{e^{[(\mathrm{HCO}_3^- - k)/a]} + 1} \tag{15–8}$$

Where μ_n = growth rate of nitrifying bacteria, g new cells/g cells·d
μ_m = maximum specific growth rate of nitrifying bacteria, g new cells/g cells·d
HCO_3^- = bicarbonate concentration, mM
k = saturation constant, mM
a = constant (estimated value of 0.83 based on Wett and Rauch, 2003)

While the value of the saturation constant, *k*, for mainstream processes is 0.5 mM or less, the value has been estimated to be around 4 mM in the warm, high growth rate environment of a separate sidestream reactor (Wett and Rauch, 2003). Consequently, for sidestream biological treatment processes, the residual inorganic carbon concentrations in the reactor must be considered carefully when selecting the reactor operating conditions that provide optimum performance. Bicarbonate and CO_2 exist in equilibrium at concentrations dictated by the reactor bulk liquid conditions, such as temperature, pH, and calcium concentration (formation of calcium carbonate). As the reactor pH decreases below 7, more bicarbonate shifts to $CO_{2,aq}$, which is removed subsequently from the reactor via air stripping. Therefore, under increasingly acidic conditions in a sidestream biological reactor, a restricted nitrifier growth rate may occur due to an inorganic carbon limitation.

Need For Degradable Organic Matter. Depending on the COD source (e.g methanol, glycerol, volatile fatty acids, municipal wastewater), the observed sludge yield under anoxic conditions is typically in the range of 0.28 to 0.4 g VSS /g COD consumed, resulting in a degradable COD to nitrate-N ratio of 4.8 to 6.6 g/g [see Eq. (7–126)]. In a typical high strength sidestream, the ratio of degradable COD to TKN is less than 1. Thus, the available degradable COD for a sidestream treatment system is insufficient where a high denitrification efficiency is desired. Supplementation of the COD with commercially available organic carbon sources is an option; however, the use of COD sources within the facility (e.g primary sludge; endogenous decay of secondary mixed liquor solids) is desired as the purchase of a commercial COD source increases operating cost.

Treatment Processes

Over the past 20 years, a number of separate and integrated nitrification-denitrification treatment process configurations have been developed or implemented for sidestream treatment. The principal processes are

BAR/R-D-N process

InNitri® process

ScanDeNi® process

Sequencing batch reactor

Summary information on these processes, including process flow diagrams, is provided in Table 15–4.

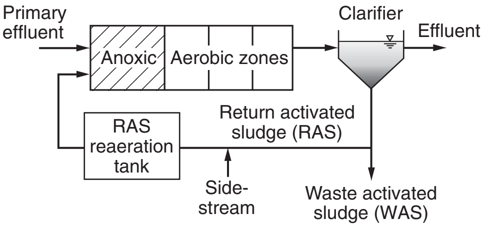
In addition to the process configurations described in Table 15–4, trickling filters have been used for sidestream treatment in a limited number of facilities, but typically their application has occurred where the mainstream process has been upgraded from trickling filters to a suspended growth activated sludge system, resulting in the availability of a

Table 15–4

Description of separate and integrated nitrification–denitrification processes for sidestream treatment

<table>
<tr><th>Process</th><th>Description</th></tr>
<tr><td>

(a) BAR and R-D-N process



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BioAugmentation Reaeration (BAR) or Regeneration-Denitrification-Nitrification (R-D-N) process consists of nitrification of the sidestream in a plug-flow return activated sludge (RAS) reaeration tank. In a typical design, the sidestream is mixed with the entire RAS flow at the head of the tank. An anoxic zone can be provided at the head of the tank with a HRT of 1 h to promote partial denitrification of the RAS and suppress odors associated with the sidestream. The aerobic zones that follow typically have a total hydraulic retention time (HRT) of 2 h to allow complete ammonium oxidation. A final anoxic zone may be provided at the end of the tank with a 1 h HRT to further promote endogenous denitrification if needed. Process modeling is used to refine the reaeration tank volume requirement. Addition of the entire RAS flow dilutes the sidestream ammonium-N and other constituents by 50 to 100-fold, creating a mixed liquor environment for nitrification that is similar to that in the mainstream activated sludge reactors. RAS alkalinity is typically sufficient to meet the alkalinity demand for full nitrification of the sidestream ammonium load. Ideally, mechanical mixers are used in the anoxic zone(s), but for large plants coarse bubble aeration has proven to be a cost-effective method of providing mixing with limited impact on denitrification.

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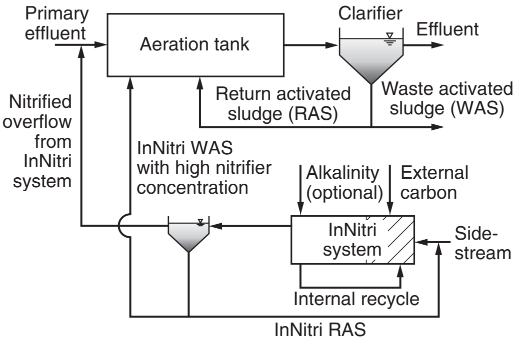
<tr><td>

(b) InNitri® process



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The InNitri® process was developed with the key objective of bioaugmentating the mainstream process with nitrifier-rich waste sludge. The InNitri process is typically designed in a Modified Ludzack-Ettinger (MLE) configuration with a separate gravity clarification step. An external carbon source is added to promote denitrification and generate alkalinity to enhance the nitrification efficiency. The internal recycle rate from the aerobic to the anoxic zone depends on the desired total inorganic nitrogen (TIN) removal efficiency. An external alkalinity source may be added if a high nitrification efficiency is required to meet the treatment objective. Alternatively, the InNitri reactor may be operated entirely in an aerobic mode, which would require external alkalinity addition to achieve a high nitrification efficiency. Primary or plant effluent is fed to the reactor to prevent the operating temperature from exceeding 38°C or to maintain the reactor within a specific temperature range. To maximize the amount of nitrifier mass sent to the mainstream secondary process, the InNitri sidestream reactor is operated at an aerobic SRT in the range of 3 to 5 d, which provides stable sidestream performance, but limits the loss of nitrifier mass through decay.

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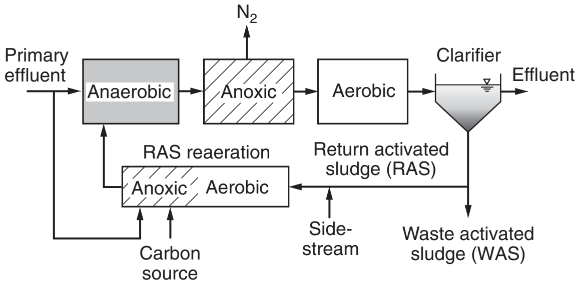
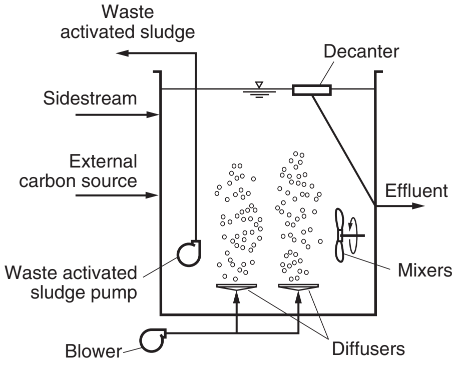
(*continued*)

Table 15–4 (*Continued*)

Process	Description
(c) ScanDeNi® process  	Similar to the BAR/R-D-N process, the ScanDeNi® process nitrifies the sidestream nitrogen load in a RAS reaeration tank. Unlike the other processes, a post anoxic zone is provided and an external carbon source is added to enhance denitrification. No internal recycle within the reaeration tank is provided since the RAS has sufficient alkalinity for complete nitrification. The design HRT of the anoxic zone will be impacted by the selected external carbon as the denitrification rates will vary with the source. The ScanDeNi process was developed to provide a means of nitrifying and denitrifying a sidestream ammonium load and provide a denitrified RAS flow to the mainstream process where biological phosphorus removal is performed.
(d) Sequencing Batch Reactor (SBR)  	Sequencing Batch Reactors (SBR) have been used most commonly for separate sidestream treatment. In the SBR configuration, sidestream is fed continuously during the *React* period or rapidly fed to the reactor at the beginning of the SBR cycle in a defined anoxic *Fill* period. Intermittent aeration is applied at defined time intervals to provide aerobic and anoxic periods for nitrification and denitrification. Typically, the total aerobic time is two-thirds of the *React* period. If a high ammonium removal efficiency is desired, an external alkalinity source is provided. If high ammonium and inorganic nitrogen removal efficiencies are desired, an external carbon source or a facility carbon source such as primary sludge/WAS is added for denitrification and to generate alkalinity for nitrification. Primary or plant effluent is fed to the SBR for temperature control as needed. At the end of the *React* period, the suspended solids are settled (*Settle* period), treated sidestream is decanted to the mainstream plant (*Decant* period) and waste sludge is pumped to the solids processing train. The total SRT is typically 10 d or higher.

trickling filter for sidestream treatment. Performance is largely limited to partial nitrification and a dilution water source (e.g., primary effluent) and a high internal recycle rate are required to maintain the desired hydraulic loading rate. Trickling filters are not considered further in this chapter.

Other attached growth process configurations such as moving bed biofilm reactors and rotating biological contactors, historically, have not been used for nitrification-denitrification of sidestreams, but have been proven effective for deammonification of sidestreams, as presented in Sec. 15–10. Submerged attached growth processes such as fluidized bed reactors and biological aerated filters have also not been applied full-scale for sidestream treatment. However, in pilot studies with submerged reactors with a sand medium nitritation-denitritation with methanol as the supplemental carbon source has been demonstrated successfully. The application of attached growth systems is considered further in Sec. 15–10.

Finally, activated sludge reactors have been applied successfully for the nitrification of digester sidestreams (Jeavons et al., 1998), but these systems have been decommissioned.

15–9 NITRITATION AND DENITRITATION PROCESSES

The development of the nitritation-denitritation processes in the 1990s was driven by the desire to reduce the energy and chemical requirements for high strength sidestream nitrogen removal. Nitritation-denitritation processes that have been implemented in full-scale facilities are described in this section along with general design information. Separate and integrated sidestream-mainstream processes and those that produce a mixture of nitrite and nitrate during the ammonium oxidation step are also included in this section.

Fundamental Process Considerations

To understand the nitritation-denitritation process it is useful to consider (1) the biological pathways, (2) process biology, kinetics, and stoichiometry**,** (3) the alkalinity requirements, (4) the need for biodegradable organic carbon, and (5) the operating modes that limit the oxidation of nitrite to nitrate.

Biological Pathways. As illustrated on Fig. 15–9(b), short-circuiting the nitrification-denitrification pathway through the restriction or prevention of nitrite oxidation to nitrate, reduces the stoichiometric oxygen demand and associated aeration energy by 25 percent. In the subsequent anoxic step, the heterotrophic bacteria require 40 percent less degradable organic carbon to reduce nitrite to nitrogen gas (*denitritation*), in comparison to the organic carbon required for nitrate reduction, as the oxidation state of nitrogen in nitrite (+3) is lower than nitrate (+5). The corresponding stoichiometry is given below.

Process Biology, Kinetics, and Stoichiometry. The microbiology, basic biochemical reaction and autotrophic growth kinetics associated with the oxidation of ammonium to nitrite are presented in this Section. Biological denitritation is discussed in Sec. 7–10 in Chap. 7. The pertinent nitritation-denitritation process stoichiometry is as follows.

Biological conversion of ammonium to nitrite (nitritation) is

$$\begin{aligned} &NH_4HCO_3 + 0.9852NaHCO_3 + 0.07425CO_2 + 1.4035O_2 \rightarrow \\ &\qquad 0.01485C_5H_7NO_2 + 0.9852NaNO_3 + 2.9406H_2O + 1.9852CO_2 \end{aligned} \tag{7–92}$$

The amount of biodegradable organic compound required to reduce nitrite to nitrogen gas is dependent on the carbon source as illustrated by the following reactions based on municipal wastewater, methanol, and acetate.

Wastewater:

$$C_{10}H_{19}O_3N + 16.66NO_2^- + 0.33H_2O \rightarrow 10CO_2 + NH_3 + 8.33N_2 + 16.66OH^- \tag{15–9}$$

Methanol:

$$5CH_3OH + 10NO_2^- \rightarrow 5N_2 + 5CO_2 + 5H_2O + 10OH^- \tag{15–10}$$

Acetic Acid:

$$5CH_3COOH + 13.33NO_2^- \rightarrow 10CO_2 + 3.33H_2O + 13.33OH^- \tag{15–11}$$

The equations given above can be compared to the nitrate reduction reactions in Eq. (7–110) through (7–112) given in Sec. 15–8. Comparing the corresponding equations, it can be seen that the mass of a specific organic compound can reduce a greater amount of inorganic nitrogen if nitrite is not allowed to oxidize to nitrate during the ammonium oxidation step in the process.

Similar to denitrification, the equations above do not reflect the actual carbon requirement for denitritation, as the reactions do not account for cell mass synthesis. When expressed on a COD basis, the actual biodegradable COD requirement for denitritation takes into account the effect of biomass yield and is calculated as follows.

$$\frac{\text{bsCOD}}{\text{NO}_2\text{-N}} = \frac{1.71}{1 - 1.42Y_n} \tag{15–12}$$

Where 1.71 = oxygen equivalent of nitrite (g O_2/g NO_2- N)
Y_n, = net biomass yield, calculated with Eq. (7–121), presented in Sec. 15–8.

Alkalinity Requirement. Alkalinity destruction during the complete oxidation of ammonium to nitrate occurs during the nitritation step. Therefore, the alkalinity demand for acid neutralization in a nitritation process is identical to a process performing complete nitrification. The alkalinity requirement for complete ammonium removal must be provided through an external alkalinity source or alkalinity generation through nitrite reduction. Providing the alkalinity requirement by diluting the sidestream into another plant stream (raw influent, primary tank effluent; plant effluent) is not a viable option, as dilution creates a growth environment where restriction of nitrite oxidation to nitrate is difficult to control. As discussed in Sec. 15-8, considering the impact of bicarbonate concentration on the autotrophic growth rate and the potential reduction in alkalinity through the use of chemicals (e.g. iron salts) upstream of the sidestream treatment process, the addition of supplemental alkalinity may be required depending on the treatment objective.

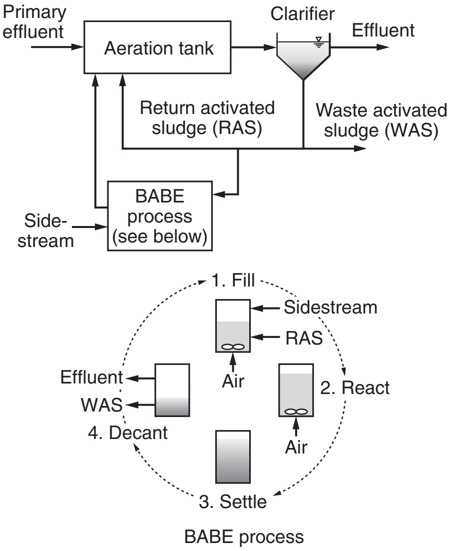
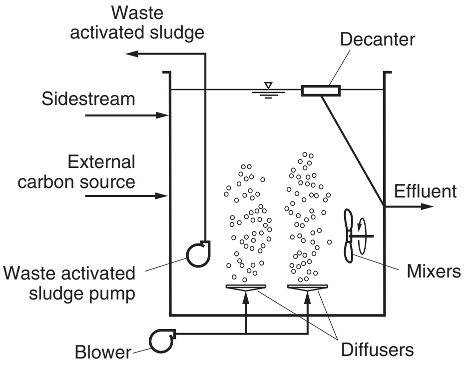
Need For Degradable Organic Matter. Despite a reduction in the organic carbon requirement, the amount of degradable organic carbon in a typical sidestream is insufficient to allow complete denitritation. An external organic carbon source or a carbon source from within the plant such as primary solids is required. In the absence of sufficient denitritation, an external alkalinity source is required for a high nitritation efficiency or incomplete ammonium oxidation to nitrite will occur.

Restriction of Nitrite Oxidation and the Impact of Nitrite Accumulation. Preventing or limiting NOB growth in a nitritation-denitritation process can be accomplished through four primary mechanisms: (1) low aerobic SRT at a reactor temperature greater than 20°C, (2) intermittent aeration at low dissolved oxygen (DO) concentration, (3) free ammonia inhibition, and (4) free nitrous acid inhibition, where the latter two inhibitory effects are dependent on reactor pH. The role of one or more of these mechanisms to restrict NOB growth is discussed in the process descriptions given in Table 15–5. For further discussion of ammonium and nitrite oxidizing bacteria growth kinetics and conditions that restrict NOB growth, refer to Sec. 7–9 in Chap. 7.

The accumulation of nitrite to high concentrations can result in a reduction in the ammonium oxidation rate through free nitrous acid inhibition. Therefore, inclusion of denitritation in the treatment process is preferred for very high strength sidestreams, or the sidestream can be diluted with primary or plant effluent, with the restriction that excessive dilution may result in sidestream reactor conditions that limit the control of nitrite oxidation to nitrate. Where a high residual nitrite concentration is present in the sidestream reactor effluent, the effluent should be discharged to an anoxic zone in the mainstream secondary process to prevent further oxidation of the nitrite to nitrate and to minimize the organic carbon demand for inorganic nitrogen removal.

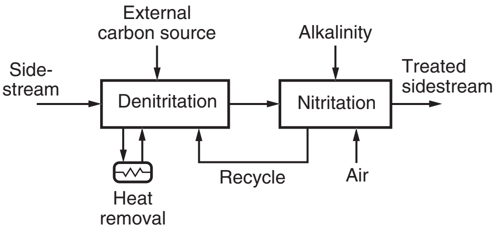
Table 15–5

Description of nitritation–denitritation processes for sidestream treatment

Process	Description
(a) BABE® process  	The BABE® (Biological Augmentation Batch Enhanced) process is an integrated sidestream-mainstream configuration designed to provide a source of nitrifier-enriched sludge for bioaugmentation of the mainstream process, but is operated under conditions that limit the oxidation of nitrite to nitrate (Berends et al., 2005). The BABE reactor can be a sequencing batch reactor (SBR) operated with intermittent aeration, a plug-flow reactor with external clarification where alternating anoxic and aerobic zones are provided or a plug-flow reactor in a MLE configuration. Regardless of the reactor configuration, the distinguishing feature of the BABE process is the addition of a portion of mainstream return activated sludge (RAS) ($<$ 10 percent of the total RAS flow) to the reactor, which serves to integrate the BABE reactor with the mainstream process and control the BABE reactor temperature at or below 25°C. Sidestream ammonium is oxidized to a mixture of nitrite and nitrate, which are subsequently reduced through endogenous denitrification with the RAS solids and the addition of an external carbon source. Free ammonia inhibition and transient anoxia are believed to be the primary mechanisms that limit nitrite oxidizing bacteria (NOB) growth in the BABE reactor. Waste sludge from the BABE reactor is sent to the mainstream process. The reactor volume requirement and the RAS flowrate (or daily volume) required for the BABE process are dependent on the overall facility inorganic nitrogen removal objective and typically developed through process modeling.
(b) Sequencing Batch Reactor (SBR)  	Nitritation-denitritation can be achieved in a SBR configuration through pH-controlled or time-based intermittent aeration with the DO concentration controlled at 1 mg/L or less during the aerated periods (Wett, 1998). An external carbon source or a facility carbon source such as primary sludge is added during the anoxic phases to promote denitritation. Continuously feeding sidestream to the SBR during the *React* phase is typically practiced in full-scale systems. However, for sidestreams with a COD/N ratio approaching 1, an intermittent feeding strategy may be beneficial to minimize the external carbon demand. The SBR typically operates at temperatures above 30°C due to biological heat generation and may require heat removal through an external mixed liquor cooling loop or through the addition of dilution water to maintain the reactor temperature below 38°C. A total solids retention time (SRT) in range of 5 to 10 d is typical.

(*continued*)

Table 15–5 (*Continued*)

Process	Description
(c) SHARON® process 	The SHARON® (Stable reactor system for High Ammonia Removal Over Nitrite) process consists of anoxic-aerobic continuous stirred tank reactors (CSTRs) in a series configuration, without suspended solids retention. The process is operated in the temperature range of 35 to 38°C. The design SRT of the aerobic zone is 1.5 d under this elevated operating temperature and controlled at this value through intermittent aeration, which allows growth of ammonia oxidizing bacteria, but provides a washout rate that prevents the growth of nitrite oxidizing bacteria (Hellinga, 1998). The design anoxic SRT is typically 0.75 d. The anoxic and aerobic volumes are based on the sidestream design flow and their respective design SRT values. If the sidestream is diluted with primary and plant effluent, the diluted flowrate serves as the basis for the anoxic and aerobic zone volumes. An internal recycle flow of 13 times the feed flow (undiluted or diluted sidestream) is provided to supply nitrite to the anoxic zone. An external carbon source is supplied to the anoxic zone for denitritation and to generate alkalinity to support nitritation. The addition of an external alkalinity source may also be required to sustain the target inorganic nitrogen removal efficiency. Heat generated by the biological reactions is removed through an external mixed liquor cooling loop, which can also serve as a method for heat addition if heat losses during the winter period are excessive. Highly concentrated sidestreams are typically diluted with primary or plant effluent to lower the ammonium concentration to 1500 mg N/L or less.

Treatment Processes

Nitritation-denitritation processes have also been developed and implemented at a full-scale for separate sidestream treatment. Process development has also led to integrated sidestream-mainstream configurations where the oxidation of nitrite to nitrate is not controlled completely, but nitrite remains the dominant product that accumulates during the ammonium oxidation step. The principal separate and integrated processes are

BABE® process

Sequencing batch reactor

SHARON® process

Summary information on these processes, including process flow diagrams, is provided in Table 15–5.

15–10 PARTIAL NITRITATION AND ANAEROBIC AMMONIUM OXIDATION (DEAMMONIFICATION) PROCESSES

Further reduction in aeration energy and chemical demand for sidestream inorganic nitrogen removal can be achieved by implementing processes that perform partial nitritation and support the growth and enrichment of Anammox bacteria. Anammox bacteria are autotrophic organisms capable of oxidizing ammonium under anoxic conditions, using nitrite as the electron acceptor. The development and implementation of processes designed to perform sequential partial nitritation and anaerobic ammonium oxidation ("Deammonification") has accelerated since the first full-scale systems were commissioned in Germany and The Netherlands in 2001–2002.

In the initial development of deammonification processes, researchers found that stable deammonification can be achieved within a single reactor where the two primary reactions occur in a suspended biomass or a biofilm (Kuai and Verstraete, 1998; Olav Sliekers et al., 2002; Siegrist et al., 1998; Seyfried et al., 2001). Two early process names (1) Completely Autotrophic Nitrogen removal Over Nitrite (CANON) and (2) Oxygen-Limited Autotrophic Nitrification Denitrification (OLAND) soon fell out of use, as specific trademarked process names became the norm. The processes described in this section are summarized in Table 15–6.

Fundamental Process Considerations

To understand deammonification processes and the advantages they present in reducing the cost of removing inorganic nitrogen from sidestreams, the biology, kinetics and reaction stoichiometry of Anammox bacteria, and their syntrophic relationship with aerobic ammonium oxidizing bacteria are summarized briefly below.

Process Biology, Kinetics, and Stoichiometry. Anammox organisms have been detected in marine and fresh water environments, soils, sediments, wetlands, and of particular interest for the discussion of sidestream treatment, in wastewater treatment plants (Kuenen, 2008; Van Hulle et al., 2010). Activated sludges from conventional nitrification/denitrification plants have been used for the startup of bench, pilot, and full-scale processes that incorporate the Anammox reaction (Fux et al., 2002; Third et al., 2005; van der Star et al., 2007).

The biochemistry and growth kinetics of the anammox bacteria are provided in Chaps. 7 and 8. As discussed in these chapters and summarized below, anammox bacteria have a maximum specific growth rate that is approximately one-tenth of the growth rate of aerobic nitrifying organisms. Therefore, a long SRT is a key feature in the design of all deammonification processes. Additional environmental factors that control their growth are also summarized below. The stoichiometry of the partial nitritation and Anammox reactions is given below.

Partial Nitritation. If ammonium is oxidized partially to nitrite (*partial nitritation*) according to the stoichiometric reaction given by Eq. (7–92), the stoichiometric ratio of ammonium and nitrite needed for the anammox reaction is obtained as shown in Eq. (15–13). The neutralization of the acidity produced by the nitritation reaction with bicarbonate alkalinity is included in the partial nitritation stoichiometry for discussion purposes.

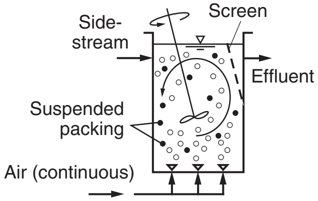
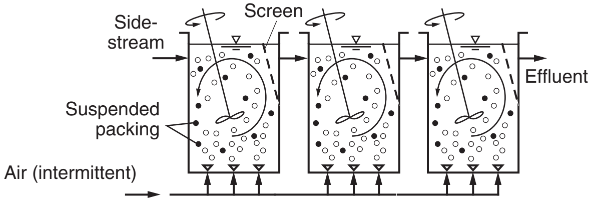
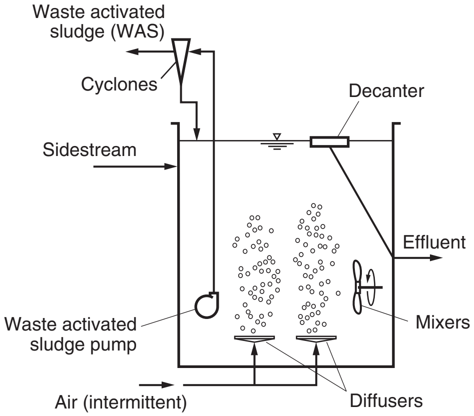
$$\begin{aligned}&2.34\mathrm{NH_4^+} + 1.87\mathrm{O_2} + 2.66\mathrm{HCO_3^-} \rightarrow \\ &\qquad 0.02\mathrm{C_5H_7NO_2} + \mathrm{NH_4^+} + 1.32\mathrm{NO_2^-} + 2.55\mathrm{CO_2} + 3.94\mathrm{H_2O}\end{aligned} \tag{15–13}$$

Anammox Reaction. Under oxygen-free conditions anammox bacteria will oxidize ammonium using nitrite as the electron acceptor. The principal stoichiometric reaction, including cell mass synthesis, is shown in Eq. (15–14) (Strous et al., 1998).

$$\begin{aligned}&\mathrm{NH_4^+} + 1.32\mathrm{NO_2^-} + 0.066\mathrm{HCO_3^-} + 0.13\mathrm{H^+} \rightarrow \\ &\qquad 1.02\mathrm{N_2} + 0.26\mathrm{NO_3^-} + 0.066\mathrm{CH_2O_{0.5}N_{0.15}} + 2.03\mathrm{H_2O}\end{aligned} \tag{15–14}$$

Table 15–6

Description of partial nitritiation—anaerobic ammonium oxidation (*deammonification*) processes

Process	Description
(a) ANITA™Mox–Single stage moving bed biofilm reactor process  	ANITA™Mox is a single-stage deammonification moving bed bioreactor (MBBR) system that is continuously aerated with a variable dissolved oxygen (DO) setpoint in the range of 0.5 to 1.5 mg/L, based on online reactor ammonium and nitrate measurements. DO setpoint adjustment is required to control nitrite oxidizing bacteria (NOB) growth and provide stable ammonium removal. Due to continuous aeration, mechanical mixing is only required during startup and periods of low ammonia loads. The fraction of the reactor volume occupied by the media does not exceed 50 percent to prevent insufficient mixing and media movement throughout the reactor volume. The AnoxKaldnes plastic media with an active specific area of 500 m^2/m^3 or higher have been used in this process.
(b) DeAmmon® moving bed biofilm reactor process  	DeAmmon® consists of a single or dual train reactor system with three stages per reactor (designed by Purac). The stages are operated in series, but piping flexibility is provided to allow parallel operation. Kaldnes (AnoxKaldnes/Veolia) K1 media has been used typically for biofilm support (active surface area of 500 m^2/m^3). Each stage is aerated intermittently to provide aerobic and anoxic periods for the partial nitritation and anammox reactions, respectively, and continuously mixed with mechanical mixers. The duration of the aerobic and anoxic periods are dependent on the ammonium loading to the system and the removal rate. Aeration and anoxic times of 20 to 50 min and 10 to 20 min, respectively, are typical. A DO concentration of 3 mg/L during the aeration periods is targeted, but higher concentrations are avoided to prevent the potential for NOB growth and to limit anammox inhibition. Typically, the fraction of the reactor volume occupied by the media does not exceed 50 percent to prevent insufficient mixing and media movement throughout the reactor volume.
(c) DEMON® Sequence Batch Reactor (SBR)  	The DEMON® SBR is a suspended growth reactor operated by pH-controlled or time-based intermittent aeration to provide aerobic periods for the partial conversion of ammonium to nitrite and anoxic periods for the anammox reaction (Wett, 2006). The peak DO concentration during each aerobic phase is controlled at approximately 0.3 mg/L to provide selective pressure against NOB growth and allow a rapid transition to an anoxic condition after the air is shut off to promote an optimal environment for the anammox bacteria. To minimize the impact of nitrite on anammox activity, the aeration period in each cycle is typically around 10 to 15 min (anoxic period is typically 5 to 10 min). If pH is used to control the aeration cycle, a pH interval of 0.01 or 0.02 units is applied. The mean operating pH is typically maintained above 6.8 to minimize inorganic carbon loss through air stripping of CO_2. The SBR is continuously fed during the *React* phase of each SBR cycle. Because the anammox bacteria grow in the SBR is in a dense granulated form, the waste sludge is pumped through hydrocyclones to separate the anammox granules from the remaining flocculated solids and return them to the reactor. Consequently, the solids retention time (SRT) of the anammox bacteria approaches 40 to 50 d and the SRT of the remaining solids [ammonia oxidizing bacteria (AOB), heterotrophs, inert solids] is maintained around 10 d, which provides further selective pressure against NOB growth and more stable performance over a range of loading conditions (Wett et al., 2010).

(*continued*)

Table 15–6 (*Continued*)

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<tr><th>Process</th><th>Description</th></tr>
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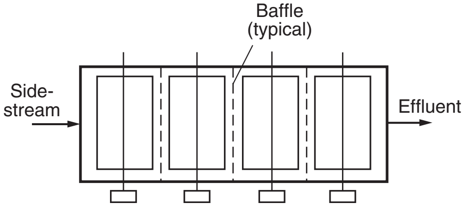
(d) Rotating Biological Contactors (RBCs)



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Rotating Biological Contactors (RBC) have been applied for leachate treatment and were found to promote the development of deammonification as the dominant nitrogen removal pathway (Seyfried et al., 2001). The deammonifying RBCs vary in configuration and media material, but demonstrated an average surface specific deammonification rate of approximately 2.5 g $N/m^2 \cdot d$ with peak rates up to 4.8 g $N/m^2 \cdot d$. The typical RBC bulk liquid DO concentration is 1 mg/L. For leachates rich in degradable organic carbon, biological removal of the carbon is required before the deammonification RBCs. Deammonification performance was stable at operating temperatures less than 20°C and a nitrogen removal efficiency of 70 percent was reported at temperatures as low as 10°C (Seyfried, 2002). Information is not readily available on submergence depth or rotational velocity required to induce or optimize deammonification performance.

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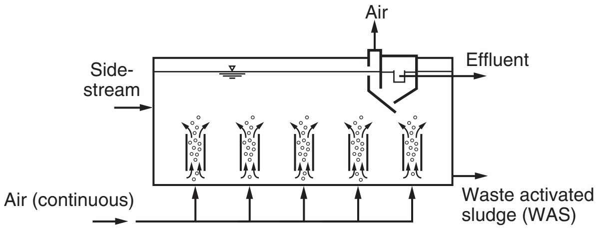
(e) Single-Stage ANAMMOX® process



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The single-stage ANAMMOX® reactor (designed by Paques BV, The Netherlands) was initially based on an air lift design where air is continuously applied to the bases of multiple riser tubes, resulting in upward liquid-solids movement through the risers and downward movement of liquid-solids outside the risers after gas disengagement. The hydrodynamic conditions favor the development of a thick well-granulated biomass where the AOB and anammox populations grow synergistically within the same granules. In a later development of this process, sludge granulation was shown to be stable without the need for riser tubes, simplifying the design. A proprietary gas-liquid-solids separator(s) located in the upper section of the reactor provides separation of the liquid-solids from the exhaust air and a settling zone for the granulated solids, which return to the main reaction zone. Pretreatment of the sidestream to remove denser inert solids is advisable to prevent their accumulation in the reactor. Online measurement of DO, nitrite and ammonium are used to control the process.

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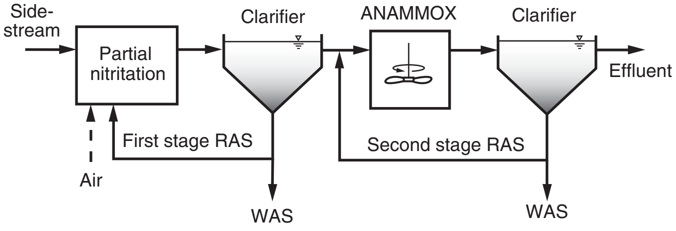
(f) Terra-N® process



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The Terra-N® MBBR process was developed by SÜD-Chemie/Clariant GmbH (Munich, Germany) where bentonite is used as the support media for biofilm growth (TERRANA® product; mean particle diameter in the range of 25 to 45 μm). Due to the rapid settling rate and compaction of the bentonite particles, gravity clarification is applied for separation of the media and biofilm from the bulk liquid. The process is designed in a two-stage configuration, with each stage consisting of a completely mixed reactor and a gravity clarifier (continuously aerated partial nitritation stage followed by a completely mixed anoxic stage for the anammox reaction), or as a single-stage SBR with intermittent aeration. In the SBR and the partial nitritation stage of the two-stage configuration, bentonite is added to a concentration of 10 to 12 g/L. With biomass attachment, the total suspended solids concentration is typically 15 to 20 g/L. In the anammox stage of the two-stage configuration, granulation of the anammox bacteria eliminates the need for a support media, although the addition of bentonite is not detrimental. Second-stage biomass concentrations of 5 to 7 g/L have been reported (Clariant/SÜD Chemie, 2012).

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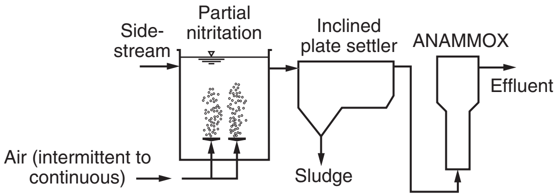
(g) Two-stage SHARON®–ANAMMOX® process



Deammonification can be accomplished in a two-stage system consisting of partial nitritation followed by the anammox reaction. In the first stage, a CSTR is operated at an aerobic SRT of approximately 1.5 d (based on the SHARON® concept) at a temperature above 30°C, without alkalinity or organic carbon addition, to convert approximately 50 percent of the sidestream ammonium to nitrite. The partially nitritated sidestream is fed to the second stage ANAMMOX® reactor (provided by Paques BV, The Netherlands) operated in an upflow configuration. Under a high upflow superficial liquid velocity, anammox bacteria form dense granules with settling velocities greater than 100 m/h, which allows the development of a sludge bed in the lower section of the reactor with a solids concentration as high as 5 to 7 percent. Flocculated particles with lower densities and settling velocities are washed out. Solids separation between the two reactors is recommended to prevent the accumulation of denser inert solids in the ANAMMOX® reactor sludge bed. Internal liquid mixing within the ANAMMOX reactor is provided by a gas-lift mechanism where nitrogen gas produced by the sludge bed is collected through a proprietary gas collection system located above the mid-depth point of the tank, which conveys the gas to a central riser pipe, inducing gas-lift of liquid to a gas-liquid separator located on top of the reactor. De-gassed liquid is returned to the bottom of the tank to increase the superficial liquid velocity in the sludge bed and dilute the partially nitritated sidestream entering the reactor.

Because the Anammox bacteria use inorganic carbon as their source of carbon for growth and nitrite is used as the electron acceptor, inorganic nitrogen is removed without the addition of an organic carbon substrate. As illustrated in Eq. (15–14), nitrate is a product of the reaction; hence, some organic carbon is required for complete inorganic nitrogen removal. For every kg of ammonium-N that is removed by the two-step deammonification pathway [see Fig. 15–9(c)], 0.11 kg of nitrate-N is produced. Consequently, deammonification reduces the organic carbon demand to about 11 percent of the carbon demand required by heterotrophic reduction of nitrate to nitrogen gas in a nitrification-denitrification process. The corresponding oxygen requirement for deammonification is also reduced to approximately 40 percent of the oxygen required for complete oxidation of ammonium to nitrate by the two-step aerobic autotrophic pathway.

Alkalinity Requirement. Bicarbonate in a typical high strength sidestream is equal to the ammonium concentration on a molar basis. As shown in Eqs. (15–13) and (15–14) above, the total amount of bicarbonate required to neutralize acidity produced by ammonium oxidation to nitrite and to support autotrophic cell growth will exceed the amount available in the sidestream. Nitrate reduction in the sidestream reactor, in part or entirely, depending on the degradable COD in the sidestream, will generate bicarbonate alkalinity, but the amount is insufficient to provide the balance required to satisfy the overall bicarbonate stoichiometric demand. In practice, deammonification processes typically achieve sufficiently high ammonium removal efficiencies that no external alkalinity source is added (greater than 80 percent removal, with 90 to 95 percent being typical). If ferrous or ferric chloride is added to the digesters or other locations to control struvite formation, the addition of alkalinity to the deammonification system to compensate for the alkalinity destroyed by iron salt addition may be justified or required if a high inorganic nitrogen removal efficiency is required.

Solids Retention Time. Anammox organisms have a maximum specific growth rate that is less than one-tenth the rate of aerobic nitrifying bacteria (see also Table 7–13 in Chap. 7). Hence, a SRT greater than 20 d is required to sustain these organisms in a biological reactor. Fortunately, anammox bacteria produce an excessive amount of exocellular polymeric substances (Cirpus et al., 2006), which results in tightly bound aggregation and granule formation as the organisms are concentrated in a suspended growth reactor. The formation of a granulated biomass results in high solids settling rates and a high solids specific gravity which can be exploited to separate the granules from flocculated solids. These features allow process configurations that can easily retain the anammox bacteria within the processes. Integration of the organisms within a biofilm also occurs readily once the environment conducive to their growth and sufficient surface area are provided.

Implementation of Deammonification Processes and Challenges. The first deammonification processes were developed for high strength sidestreams where the warm sidestream temperatures are advantageous in enhancing the growth rate of the anammox organisms. For suspended growth sequencing batch reactors, intermittent aeration combined with a low dissolved oxygen (DO) concentration (0.3 mg/L) during the aerobic period of each aeration cycle has proven to be an effective way to provide conditions for the growth of aerobic ammonia oxidizing and anammox bacteria while restricting the growth of aerobic nitrite oxidizing bacteria (Wett, 2007).

Nitrite accumulation in biofilms can also be induced by operating the attached growth reactor at a reduced bulk liquid DO concentration, providing an environment within the biofilm where the Anammox organisms can thrive. Sustaining the growth of aerobic ammonia oxidizing and anammox bacteria within a single reactor, while preventing growth of aerobic nitrite oxidizing bacteria, has also been demonstrated with continuously aerated gas-lift reactors where the hydrodynamic conditions induced by the aeration system creates a granulated sludge, which can be retained to a high concentration (Olav Sliekers et al., 2003; Abma et al., 2010). With a low DO concentration, aerobic and anoxic zones within the granulated sludge provide environments for aerobic ammonia oxidizer and anammox growth.

Although the anammox reaction is attractive from the perspective of reducing the plant operating cost associated with inorganic nitrogen removal, incorporating anammox bacteria within a biological reactor system has presented a number of technical challenges. First and foremost, due to their low growth rate, a long reactor startup period is required in comparison to a nitrification-denitrification or nitritation-denitritation system and a long recovery period is also needed in the event of a significant reduction in anammox activity. However, seeding a new installation with anammox-enriched biomass has been demonstrated to facilitate startup (Wett, 2006; Abma et al., 2007; Schneider et al., 2009; Christensson et al., 2011). Due to the increasing number of deammonification plants, seed availability is increasing. The degree to which startup can be hastened depends on ammonium load to the startup plant and mass of seed that can be obtained.

The anammox bacteria are also sensitive to nitrite, which can cause irreversible loss of activity. Various levels of nitrite exposure, both in terms of concentration and exposure time, and the corresponding loss of activity have been reported (Fux et al., 2004; Wett et al., 2007), which indicate that nitrite toxicity is a function of both concentration and exposure time. Exposure to oxygen also results in the inhibition of anammox activity, but the effect is reversible. Finally, anammox organisms compete with NOB for nitrite; therefore, the reactor operating conditions must be well-controlled to limit NOB growth or provide an environment where anammox bacteria have a competitive advantage.

Treatment Processes

Process configurations designed for deammonification include

ANITA™Mox moving-bed biofilm reactor process
DeAmmon® moving-bed biofilm reactor process
DEMON® suspended growth sequencing batch reactor
Rotating biological contactors
Single-stage ANAMMOX® process
Terra-N® moving-bed biofilm reactor process
Two-stage SHARON®-ANAMMOX® process

Summary information on these processes, including process flow diagrams, is provided in Table 15–6.

15–11 PROCESS DESIGN CONSIDERATIONS FOR BIOLOGICAL TREATMENT PROCESSES

Process design considerations for the biological systems described in Sec. 15–8 through 15–10 are presented in this section. The principal design considerations include (1) sidestream characteristics and treatment objectives (2) design loadings and load equalization,

(3) sidestream pretreatment, (4) reactor volume requirements, (5) aeration system design, (6) sludge retention time and mixed liquor suspended solids concentration, (7) chemical addition (organic carbon requirements for denitrification and denitritation; alkalinity), (8) operating temperature and pH, and (9) heat removal. An example problem is presented following the discussion of the design considerations, to illustrate their application.

Sidestream Characteristics and Treatment Objectives

The characteristics of high strength sidestream, as noted in Sec. 15–1, are dependent on the process from which the sidestream is derived and the process operating conditions. Therefore, when designing a process for sidestream treatment, the characteristics of the specific sidestream should be measured. The sidestream parameters that should be considered are summarized in Table 15–7. Because nitrogenous oxygen demand is an important process design parameter, both soluble and total TKN measurements are included in sidestream characterization, to estimate the biodegradable organic nitrogen concentration. However, a significant fraction of the organic TKN in sidestreams may be comprised of recalcitrant dissolved organic nitrogen (rDON) and nitrogen associated with biologically inert suspended solids. Typically, about 50 percent of the soluble organic TKN (soluble TKN minus ammonium-N) in sidestream derived from anaerobic digestion is biodegradable. Of the particulate TKN, typically 10 to 15 percent of TKN fraction is biodegradable under aerobic conditions.

If the sidestream is to be generated in the future as a result of the implementation of a new solids processing system, a bench or pilot-scale sludge digestion study is recommended so that the soluble constituents in the sidestream reflect the conditions in the

Table 15–7

Parameters to consider when designing a sidestream treatment system

Parameter	Unit	Remarks
Flowrate (average/minimum/maximum)	m^3/d	
Dewatering operation frequency	d/wk	
Dewatering operation hours	h/d	
Temperature (average/minimum/maximum)	°C	
Total suspended solids	mg/L	Dependent on pretreatment requirements
Volatile suspended solids	mg/L	Dependent on pretreatment requirements
Alkalinity	mg/L as $CaCO_3$	
Total $cBOD_5$[a]	mg/L	
Soluble $cBOD_5$[a]	mg/L	0.45μm membrane filtered
Ammonium-N	mg N/L	Typically 90 to 95 percent of soluble TKN
Soluble TKN[b]	mg/L	0.45μm membrane filtered
Total TKN[c]	mg/L	
Soluble ortho-P	mg/L	

[a] A degradable-COD to cBOD5 ratio of 1.5 g/g can be assumed to estimate the degradable particulate and soluble COD fractions.

[b] Difference between the soluble TKN and ammonium-N is the soluble organic N of which roughly 50 percent is typically considered biodegradable for sidestream derived from anaerobic digestion of primary and secondary sludges.

[c] Difference between the total TKN and soluble TKN is the particulate TKN of which roughly 10 to 15 percent is typically considered biodegradable for sidestream derived from anaerobic digestion of primary and secondary sludges.

digestion process and the characteristics of the sludges and biosolids. The TSS concentration in the sidestream is estimated by the capture efficiency range anticipated in the full-scale dewatering process. In the absence of such tests, the ammonium-N and other constituent concentrations can be estimated based on digestion process modeling.

Treatment objectives for the sidestream process will differ with each facility and are dependent on the facility effluent permit limits on nutrients and the costs associated with the energy and chemical requirements. Selection of a treatment option is also dependent on the potential benefits associated with bioaugmentation of the mainstream plant. Typically, the separate and integrated processes being considered are subjected to a life cycle cost assessment to identify the most cost effective option.

Design Loading and Load Equalization

The sidestream nitrogen and degradable carbon mass loading rates will affect the process volume and the aeration and chemical requirements. For separate treatment processes that employ solids separation for SRT control, the sidestream TSS loading rate may also impact the design as the majority of the solids are biologically inert and will accumulate in the reactor. Therefore, the maximum oxygen demand and solids loading rate are used commonly for reactor design.

The peak sidestream loading rate for reactor design is dependent on the operation of the biosolids dewatering process. Dewatering facilities at smaller treatment plants are commonly operated during daytime hours for 5 or 6 d/wk. At larger facilities, the dewatering operation typically is continuous. Where biosolids dewatering is intermittent, the implementation of flow and load equalization is a cost-effective option for reducing the volume of the treatment process and the aeration energy associated with the peak oxygen demand. Equalization of sidestream flows and loads were discussed previously in Sec. 15–2.

Sidestream Pretreatment

The majority of the separate sidestream treatment processes described in Sec. 15–8 to 15–10 will benefit from a reduction in the sidestream TSS concentration before treatment. Most of the sidestream TSS are biologically inert and will, therefore, accumulate in the sidestream reactor, potentially elevating the solids concentration to a level that can negatively impact the solids-liquid separation process and the ability to control the SRT within the desired range. Pretreatment requirements are dependent on historical sidestream TSS concentration data and the specific process being used or considered for sidestream treatment. Dynamic process modeling of many of the sidestream processes is an effective tool in assessing the need for pretreatment for existing sidestream where sufficient TSS data are available. Pretreatment options were considered in Sec. 15–3.

Reduction in the sidestream TSS to an average concentration of 200 mg/L or less is typically sufficient. For the DEMON deammonification SBR where hydrocyclones are used to separate anammox granules from other suspended solids, removal of dense particulate matter is particularly important because the hydrocyclones will also separate this material from the waste sludge and retain it in the reactor. Removal of large inert debris is critical for MBBR systems that use screens for media retention as the debris will accumulate in the reactor over time.

Sidestream solids removal is typically not a requirement for the SHARON nitritation-denitritation process as solids retention is not performed. However, if the sidestream solids concentration is high ($>$ 2000 mg/L) periodically, the removal of solids may prove beneficial in reducing oxygen demand in the aerated zone. For integrated sidestream-mainstream processes, a solids balance across the facility should be conducted to assess the impact of sidestream solids on the mainstream process.

Sidestream Reactor Volume

The conditions that define the minimum reactor volume requirement vary with the type of sidestream treatment process and are dictated by the treatment objective, the maximum achievable oxygen transfer rate, and the biological kinetic rates, as well as other considerations such as solids-liquid separation. The volume requirement for a separate sidestream reactor is commonly estimated through the use of a specific nitrogen loading rate on the system, expressed as the nitrogen loading per unit reactor volume per day (as kg $N/m^3 \cdot d$) or, in the case of fixed-film systems per unit area of active media surface area per day (g $N/m^2 \cdot d$). The maximum nitrogen loading rate (kg N/d) is selected commonly for design. The specific loading rate for a given type of process is often developed through pilot or demonstration-scale studies and full-scale operating experience. Typical values for the specific loading rate for the processes discussed in Sec. 15–8 through 15–10 are provided in Table 15–8.

A more detailed analysis is required to determine the volume requirement for a SBR or a suspended growth process with external clarification. For SBRs, the minimum volume is dictated largely by the aeration system and its ability to provide sufficient oxygen to satisfy the peak oxygen demand imposed by the sidestream (design condition) during the aerobic periods of the *react* phase. Separation of solids from the treated liquid is an additional design consideration. The reactor volume determined through the aeration system analysis must be sufficient to ensure that settled sludge inventory is not disturbed during the decanting of the treated sidestream, which further ensures that the desired SRT can be maintained. Where the anoxic reaction is occurring in a separate stage, the volume requirement depends on the reaction kinetics and is determined through process modeling or pilot studies.

For the integrated sidestream-mainstream processes, process modeling is commonly used to determine the reactor volume requirement. The volume will be influenced significantly by the desired level of inorganic nitrogen removal across the facility and the mainstream operating conditions, which affect the biological kinetic rates in the reactor or zones nitrifying and denitrifying the sidestream nitrogen load. Typical loading rates or hydraulic retention times for the various processes are also presented in Table 15–8.

Where a mainstream process of a facility does not have sufficient capacity to treat sidestream and unacceptable deterioration of the facility effluent quality would result, division of the sidestream reactor volume into multiple tanks may be a design requirement, giving the facility operator the flexibility to remove a reactor from service for maintenance and inspection. Process modeling is often used to evaluate these operating scenarios.

Aeration System

The application of oxygen demand in the design of the aeration system is specific to the type of sidestream process. The oxygen demand is comprised of the oxygen requirement for the carbonaceous oxygen demand and the oxidation of ammonium to nitrite or nitrate. Typically, complete conversion of ammonium to nitrite or nitrate is assumed in the calculation of the design oxygen demand for nitritation-denitritation and nitrification-denitrification processes, respectively. Where processes have been shown to produce a mixture of nitrite and nitrate, a conservative design approach is to provide sufficient aeration capacity for complete oxidation of ammonium to nitrate. For deammonification, the nitrogenous oxygen demand is based on the stoichiometric conversion of ammonium to nitrite that provides complete ammonium removal across the process.

Aeration Requirements for Sequencing Batch Reactors. For sequencing batch reactors where aerobic and anoxic reactions are occurring within the same reactor through alternating aerobic and anoxic phases, the temporal distribution of oxygen demand

Table 15–8

Typical design and control parameters for the biological sidestream treatment processes

Process	Key design parameters	HRT, h Ammonium-N loading rate (ALR), kg N/m^3·d Media loading rate, g N/m^2·d	SRT[a], d	Process control parameters
Nitrification and denitrification processes				
BAR and R-D-N process (RAS reaeration tank)	SRT (total), HRT (reaeration tank)	HRT = 2 (aerobic)[b] 1–2 (anoxic)	8–15 (SS+MS, total)	DO, SRT
InNitri or short SRT process (2-stage)	SRT, ALR	ALR = 0.4–0.5[c]	3–5 (SS, total)	DO, external Alk or COD add., temp., SRT
ScanDeNi process (RAS reaeration tank)	SRT (total), HRT (reaeration tank)	HRT = 1–2 (aerobic)[b] 1–2 (anoxic)	8–15 (SS+MS, total)	DO, SRT, external COD add.
Sequence Batch Reactor (SBR)	SRT, ALR	ALR = 0.3–0.4[d]	10–15 (SS, total)	DO, external Alk or COD add., temp, SRT
Nitritation and denitritation processes				
BABE process (SBR)	SRT (total), ALR, temp.	ALR = 0.4	4–8 (SS+MS, total)	RAS flow to BABE reactor, temp. (25°C, max), DO
Sequencing batch reactor	SRT, ALR	ALR = 0.4–0.6[d]	5–10 (total)	DO (<1 mg/L); external Alk/COD add.
SHARON Process	SRT (aerobic, anoxic), temp.	ALR < 0.7[e]	1.5/0.75[e] (aerobic/anoxic)	Temp (35–38 °C); aerobic SRT; external COD add.
Partial nitritation and anaerobic ammonium oxidation processes				
ANITA™Mox	Media surface area, ALR	ALR = 0.7–1.2[f]	> 20	DO (0.5–1.5 mg/L)
DeAmmon®	Media surface area, ALR	ALR = 0.6–0.8[g]	> 20	DO, aerobic-anoxic cycle times
DEMON® SBR	SRT, ALR	ALR = 0.7–1.2[h]	40–50 (Anammox granules) 10–15 (floc)	pH interval (0.01–0.02 s.u.); DO (0.3 mg/L); SRT
Rotating biological contactor	Media surface loading rate	loading rate = 2.3–2.8 g N/m^2·d[i]	> 20	DO (1 mg/L)
Single-stage ANAMMOX®	SRT, ALR, air-lift aeration design	ALR = 2.0	>20	DO, SRT, temp.
Terra-N® two-stage with intermediate clarification	Bentonite concentration, ALR, clarifier overflowrate	ALR = 1.2–2.1 (partial nitritation stage)[j] 1.2–2.1 (Anammox stage)[j]	> 20	DO

(*continued*)

Table 15–8 (*Continued*)

Process	Key design parameters	HRT, h Ammonium-N loading rate (ALR), kg N/m^3·d Media loading rate, g N/m^2·d	SRT[a], d	Process control parameters
Terra-N®-single-stage	Bentonite concentration, ALR	ALR = 0.25–0.7 (1.5)[k]	> 20	DO
Two-stage SHARON®-ANAMMOX®	SRT, ALR, upflow liquid velocity	ALR < 0.7 (SHARON); 3–10 (ANAMMOX)	1.5 (SHARON, aerobic) > 20 (ANAMMOX)	Aerobic SRT, temp (SHARON) Upflow liquid velocity, SRT (ANAMMOX)

[a] Notations: SS = sidestream reactor(s); MS = mainstream secondary reactor(s); total = sum of the aerobic and anoxic SRT values.

[b] Aerobic and anoxic hydraulic retention times and RAS reaeration tank volume based on the maximum RAS flow, typically a 100% return rate, at the minimum operating temperature. HRT values shown are typical. Tank geometry is plug flow and the volume is dependent on the desired nitrification performances in the RAS reaeration tank and the mainstream plant at the minimum operating temperature. Modeling is typically used in this assessment and the autotrophic growth rates are adjusted to approximately 0.5–0.6 d^{-1} (20°C) to account for the effect of low pH on the growth rate.

[c] Loading rate based on total system volume with an aerobic/anoxic volume ratio of 2:1 and a maximum OUR of 150 mg/L·h.

[d] Maximum ammonium removal rate based on three 8-h cycles per day with a *react* period of 6 h per cycle, an intermittent aeration cycle of 66 percent aerobic–34 percent anoxic, a maximum total OUR of 150 mg/L·h.

[e] Maximum specific rate based on total system volume with aerobic/anoxic ratio of 2:1 and a maximum OUR of 150 mg/L·h. Aerobic and anoxic SRTs based on influent flowrate, e.g. if dilution water is added, the volume requirements are based on the diluted sidestream flowrate.

[f] Loading rate dependent on the type of plastic media employed and its effective surface area for biofilm growth. A removal rate up to 1.2 kg N/m^3·d with AnoxKaldnes BiofilmChip™M (effective surface area of 1200 m^2/m^3 of media; 40 percent media fill volume) was reported by Christensson et al. (2011) and Lemaire et al. (2011).

[g] AnoxKaldnes K1 media, effective surface area for biofilm = 500 m^2/m^3 of media; deammonification rate = 1.5 to 2 g N/m^2 of effective surface area per day (Plaza et al., 2011).

[h] Demonstrated range of loading rates. Design loading rate is typically 0.7 kg N/m^3·d.

[i] Loading rates correspond to average nitrogen removal rates demonstrated by RBC systems treating landfill leachate. Peak removal rates up to 4 g/m^2·d have been demonstrated (Siegrist et al., 1998; Seyfried, 2002).

[i] Loading rates demonstrated for retrofitted tanks (Clariant/SÜD Chemie, 2012). Partial nitritation and anammox reactor volumes are typically equal. Proprietary clarifier design-clarifier surface overflowrate and typical solids return flowrate not disclosed by vendor.

[k] Range of values demonstrated for retrofitted tanks. Value in parenthesis is considered by the vendor to be the maximum design loading rate for a new tank (Clariant/SÜD Chemie, 2012).

has an effect on reactor volume and blower capacity requirements. For example, a SBR operating at three cycles per day with a combined *settle*, *decant* and *idle* time of 1.5 h/cycle, the total *react* time over the day is 19.5 h. If the reactor is operated with intermittent aeration during the *react* periods with an aerated:anoxic time ratio of 2/1, the total aeration time is 13 h/d. Therefore, the minimum reactor volume and the design of the aeration system are based on the daily oxygen demand being satisfied over the total aerated time period. Allocation of carbonaceous oxygen demand in the aerated and anoxic periods is dependent on if the reactor is fed continuously during the *react* period, fed rapidly at the beginning of each SBR cycle or fed intermittently during the anoxic phases of the aeration cycle. Process modeling is often used in assessing the total oxygen demand for design.

Type of Aeration System. Because the development of advanced biological treatment processes has been driven partially by the desire to reduce aeration energy, fine and ultra-fine bubble air diffusion systems are common in suspended growth reactors. To achieve a more compact reactor design, maximum floor coverage applicable to the type of diffuser is applied. A design oxygen uptake rate (OUR) of 150 mg/L·h is used commonly for fine and ultrafine bubble diffused air systems. Diffuser selection takes into account the material of construction as some manufacturers specify a maximum wastewater temperature limit for their diffusers (e.g., 30°C), which may be lower than the anticipated average operating temperature of the reactor.

Alpha and fouling factors (see Sec. 5–11 in Chap. 5) are developed through testing or provided by the process technology or diffuser vendor based on their experience with similar applications. The utilization of high purity oxygen is uncommon, but is applicable to separate and integrated sidestream-mainstream systems. High purity oxygen would satisfy a considerably higher OUR, which may allow a reduction in the reactor volume, depending on other operating conditions such as SRT, mixed liquor suspended solids concentration and kinetic rates.

In MBBR systems that employ plastic media, medium to coarse bubble stainless steel air diffusers are standard practice to reduce maintenance requirements and the risk of having to remove the media from the reactor to replace diffusers. The dissolved oxygen concentration requirement is specific to the process and typical values are provided in Table 15–8.

Sludge Retention Time and Mixed Liquor Suspended Solids Concentration

The design SRT is specific to each type of sidestream process. Typical SRT ranges employed in the processes described in Sec. 15–8 through 15–10 are presented In Table 15–8. For suspended growth separate reactors where solids-liquid separation is employed, the suspended solids concentration is typically kept below 4000 mg/L (at the maximum liquid level in a SBR), depending on the settling and compaction characteristics of the solids. Consequently, the impact of the sidestream TSS load on the reactor solids concentration should be examined to determine if the process would benefit from a sidestream pretreatment step where the suspended solids concentration is reduced.

Chemical Requirements

Alkalinity and/or organic carbon addition is often required for many of the nitrification-denitrification and nitritation-denitritation processes, especially if a high ammonium or inorganic nitrogen removal efficiency is desired. Deammonification processes generally do not require any chemical addition, but in certain cases alkalinity addition may prove beneficial. Some of the key considerations for each chemical category are summarized below.

Alkalinity. The importance of alkalinity for ammonium oxidation and the stoichiometric requirements are discussed in Sec. 15–8. As described in Sec. 15–1 and 15–8, the alkalinity in high strength sidestream typically provides up to 50 percent of the total alkalinity required for complete ammonium oxidation. Therefore, an external alkalinity source is added or alkalinity is generated through nitrate or nitrite reduction with an external organic carbon source. In the integrated sidestream-mainstream systems where a portion of the mainstream RAS is fed to the sidestream reactor, RAS will provide a fraction of the alkalinity requirement, reducing the external alkalinity demand. For the BAR/R-D-N/ScanDeNi processes where sidestream is diluted into the entire RAS stream in a RAS reaeration tank, the RAS alkalinity is typically sufficient to meet the alkalinity requirement.

Alkalinity Addition. Under the warm, high rate growth conditions in a separate sidestream reactor, inorganic carbon can impact the autotrophic growth rate at concentrations that would not impact their growth rate under mainstream process conditions (see discussion in Sec. 15–8). As shown in Eq. (15–8), as the bicarbonate concentration approaches the saturation coefficient value, which has been estimated at 4 mole/L (200 mg/L as $CaCO_3$). Concomitantly, the AOB growth rate is reduced, resulting in a reduction in the ammonium removal efficiency in systems operated at or near their design sidestream loading rate. External alkalinity addition has been shown to increase the nitrogen removal in deammonification pilot reactors, reducing the residual ammonium concentration (Yang et al., 2011). The improvement in reactor performance through alkalinity addition may be justified, depending on an economic analysis in which the cost of alkalinity addition is compared to the cost associated with treating the residual ammonium in the mainstream plant (e.g., aeration energy, organic carbon requirement).

Sources of Alkalinity. Caustic soda is the source of alkalinity used most commonly, due to ease of handling and availability. Magnesium hydroxide and sodium carbonate can also be used. Lime is generally avoided due to its limited water solubility and the potential formation of insoluble calcium carbonate. The selection of alkalinity source may be dependent on the regional market conditions for these chemicals.

Organic Carbon. The organic carbon requirement for design is based on complete nitrite or nitrate removal and the type of carbon source. A discussion of commercially available carbon sources is presented in Sec. 8–7 in Chap. 8. Although methanol and glycerol are the carbon sources used most commonly in sidestream treatment, any readily biodegradable carbon source can be used so long as nitrification reaction is not inhibited as a consequence of its addition. The COD demand for nitrate and nitrite reduction can be estimated using Eqs. (7–126) and (15–12), respectively. The observed sludge yields required in these two equations can also be estimated or the COD-to-N ratio can be developed through pilot testing.

Due to the cost of purchasing, storing, and handling an external carbon source, particularly if the carbon source is classified as a hazardous compound (e.g., methanol), a source of organic carbon from within the wastewater treatment facility may be preferred. Fermentation of degritted primary and secondary sludges are established methods for producing readily biodegradable COD in the form of volatile fatty acids to enhance nutrient removal in the mainstream process as described in Sec. 8–7 in Chap. 8. However, there are practical limitations to using primary sludge fermentate in separate sidestream reactors, as the rbCOD is dilute and typically separated from the residual primary solids by elutriation.

Primary sludge and combined primary and secondary sludges have been used as the carbon source for nitrite and nitrate reduction (Wett et al., 1998; Bowden et al., 2012).

In both cases, pretreatment of the sludges through hydrolysis and fermentation may enhance the utilization of the carbon for nitrite and nitrate removal. Screened and degritted sludge is also preferred to limit the introduction of inert materials and debris into the sidestream reactor. Diverting a portion of the plant sludge to the sidestream process results in an incremental decrease in anaerobic digester gas production, but such a diversion may be acceptable, if the sludge has greater economic value as a carbon source for nitrogen removal.

Operating Temperature and pH

The operating temperature of the separate sidestream reactor should not exceed 38°C to avoid inhibiting the ammonium oxidation rate. The lower temperature limit is specific to the process and the desired performance of the system. In general, separate sidestream reactors do not operate at temperatures below 20°C due to the sidestream temperature and biological heat generation. Integrated sidestream-mainstream processes where the sidestream is blended with a portion or the entire mainstream RAS will operate over a broad temperature range due to seasonal changes in the plant influent temperature.

Operating pH

The operating pH of separate sidestream reactors is typically above 6.8 to avoid a pH-limited autotrophic growth rate and limit free nitrous acid inhibition, if nitrite is accumulating to a high concentration. In the RAS reaeration tank of integrated sidestream-mainstream processes (BAR/R-D-N/ScanDeNi), the pH typically decreases to or below 6.5. The impact of lower pH on nitrification performance is accounted for typically in the design by reducing the autotrophic maximum specific growth rates by 40 to 50 percent when modeling the process to determine the reaeration tank volume requirement or to predict performance with an existing tank.

Energy Balance to Determine Reactor Cooling Requirements

The biological reactions associated with nitrification-denitrification, nitritation-denitritation and deammonification are exothermic. Due to the high ammonium concentration and temperature of sidestreams derived from anaerobic sludge digestion, biological heat generation may increase the separate sidestream reactor temperature beyond 38°C, impairing process stability and performance. Therefore, a heat balance is required using the following information to determine if heat removal or the addition of dilution water is required to maintain the sidestream reactor temperature within the desired range.

1. Sidestream flow and constituent concentrations
2. Sidestream enthalpy at reactor inlet
3. Biological heat generation rates
4. Heat losses due to evaporative cooling (aeration and induced by air movement across the open reactor surface), radiation and solar heat transfer and conductive heat transfer through reactor floor and walls
5. Treated sidestream enthalpy at reactor temperature
6. Mechanical energy inputs (blower compression energy, mechanical mixers)

Heat of reactions associated with the principal biological reactions are provided in Table 15–9. The values reported in Table 15–9 take into account cell mass growth. The influent and treated sidestream enthalpies are approximated by the enthalpies of pure water at their respective temperatures.

Table 15–9

Heat of reaction at standard conditions for biological reactions

Reaction	Heat of reaction at 25°C and 1 atm[a]
Nitritation[b]: $NH_4HCO_3 + 1.5O_2 + HCO_3^- \rightarrow NO_2^- + 2CO_2 + 3H_2O$	−14.3 MJ/kg-N
Nitrification[b]: $NH_4HCO_3 + 2O_2 + HCO_3^- \rightarrow NO_3^- + 2CO_2 + 3H_2O$	−21.8 MJ/kg-N
Deammonification[b]: $NH_4HCO_3 + 0.85O_2 + 0.11HCO_3^- \rightarrow 0.44N_2 + 0.11NO_3^- + 1.11CO_2 + 2.56H_2O$	−18.6 MJ/kg-N
Denitritation (external rbCOD): $COD + a\ NO_2^- + b\ CO_2 \rightarrow a\ HCO_3^- + 0.5b\ N_2 + c\ C_5H_7NO_2 + d\ H_2O$	$= [-17.0 + (25.5 \times Y_H{}^c)]$ MJ/kg-COD[d]
Denitritation (primary/secondary sludges): $COD_{VSS} + e\ NO_2^- + f\ CO_2 \rightarrow e\ HCO_3^- + 0.5f\ N_2 + g\ NH_4HCO_3 + h\ H_2O$	
Primary: $C_{4.66}H_{7.2}N_{0.21}O_{2.06}$	−23.9 MJ/kg-COD
Secondary: $C_5H_7NO_2$	−21.8 MJ/kg-COD
Denitrification (external rbCOD): $COD + i\ NO_3^- + j\ CO_2 \rightarrow i\ HCO_3^- + 0.5i\ N_2 + k\ C_5H_7NO_2 + k\ H_2O$	$= [-13.6 + (20.7 \times Y_H{}^c)]$ MJ/kg-COD[d]
Denitrification (primary/secondary sludges): $COD_{VSS} + l\ NO_3^- + m\ CO_2 \rightarrow l\ HCO_3^- + 0.5l\ N_2 + n\ NH_4HCO_3 + o\ H_2O$	
Primary: $C_{4.66}H_{7.2}N_{0.21}O_{2.06}$	−14.1 MJ/kg-COD
Secondary: $C_5H_7NO_2$	−14.3 MJ/kg-COD

[a] Heat of reaction values calculated by:

$$\Delta H^\circ = \left(\sum_i \nu_i \overline{H}_{f,i}{}^\circ\right)_{\text{products}} - \left(\sum_i \nu_i \overline{H}_{f,i}{}^\circ\right)_{\text{reactants}}$$

Where ν_i = the stoichiometric coefficient

$\overline{H}_{f,i}{}^\circ$ = standard heat of formation of species *i* at standard conditions of 25°C and 1 atm pressure.

Standard heats of formation for bicarbonate, nitrite and nitrate based on aqueous sodium salts (Green and Perry, 2007; Haynes, 2012). Heats of formation were not adjusted from standard conditions to typical sidestream reactor conditions. Heat of formation for cell mass estimated at −258 kJ/mole based on a calorimetric higher heat value of −24 kJ/g-VSS (−2712 kJ/mole). Heat of formation for primary sludge estimated at −290 kJ/mole based on a calorimetric higher heat value of −26 kJ/g-VSS (−2574 kJ/mole).

[b] Due to the low autotrophic cell mass yield under the typical sidestream reactor operating conditions, cell mass yield is ignored in the stoichiometric equation and calculation of the heat of reaction.

[c] Y_H defined as the net cell mass yield, mg VSS/mg COD.

[d] Equation based on heats of reaction for methanol, ethanol, glycerol and acetic acid.

Separate sidestream reactors are covered or uncovered depending on the climatic conditions and the desired reactor temperature range. If heat removal is a daily requirement, the reactors may be uncovered to promote evaporative cooling to minimize the design heat load or eliminate the need for heat removal. The heat load selected for design is typically based on summer climatic conditions and a low wind velocity, which corresponds to the lowest heat loss rate to the surrounding environment. The heat transfer calculations presented in Sec. 13–9 in Chap. 13 for anaerobic digesters are applicable for covered sidestream reactors constructed of concrete. For smaller reactors constructed of steel, the heat transfer coefficient estimation methods presented by Kumana and Kothari (1982) are recommended. The correlations presented by Al-Shammiri (2002) and

Bansal and Xie (1998) are recommended for estimating evaporation cooling induced by air movement across an open reactor surface.

EXAMPLE 15–3 **Estimating Reactor Volumes and Chemical Requirements for Nitritation-Denitritation and Deammonification Processes** For the facility described in Example 15–2, calculate the reactor volume and chemical requirements for (a) nitritation-denitritation SBR and (b) deammonification SBR processes for separate sidestream treatment. Assume the following design conditions apply:

1. Average equalized sidestream flow = 124 m^3/d
2. Peak equalized sidestream flow = 149 m^3/d
3. Dilution water requirement for reactor cooling during peak summer conditions
   a. Nitritation-denitritation = 30 m^3/d
   b. Deammonification = 15 m^3/d
4. Equalized sidestream ammonium concentration = 1000 g N/m^3 (valid for all flows)
5. Equalized sidestream TSS concentration < 200 g/m^3
6. Design maximum OUR = 150 $g/m^3 \cdot h$
7. Carbonaceous OUR = 3% of the nitrogenous OUR for nitritation-denitritation
8. Carbonaceous OUR = 6% of the nitrogenous OUR for deammonification
9. External organic carbon source: methanol
   a. Concentration = 100%
   b. Specific gravity = 790 kg/m^3
   c. COD/mass = 1.5 g/g
10. Biomass yield on methanol = 0.28 g VSS/g COD
11. Nitritation-denitritation and deammonification SBR basis: Three cycles per day with each cycle consisting of:
    a. 6 h react
    b. 1 h settle
    c. 1 h decant
12. Intermittent aeration during react period: 66% aerobic and 34% anoxic

Solution Part A—Nitritation-denitritation SBR

1. Determine nitritation-denitritation reactor volume.
   a. Determine design oxygen demand.

$$\text{Design oxygen demand} = \text{nitrogenous oxygen demand} + \text{carbonaceous oxygen demand at the maximum load}$$

$$\begin{aligned}\text{Nitrogenous oxygen demand} &= (149\ m^3/d)(1\ kg\ N/m^3)(3.43\ kg\ O_2/kg\ N)\\ &= 511\ kg\ O_2/d\end{aligned}$$

$$\text{Carbonaceous oxygen demand} = (0.03)(511\ kg\ O_2/d) = 15\ kg\ O_2/d$$

$$\text{Total oxygen demand} = 511\ kg\ O_2/d + 15\ kg\ O_2/d = 526\ kg\ O_2/d$$

   b. Determine the aerobic reaction time.

$$\begin{aligned}\text{Total aerobic time/d} &= (3\ \text{cycles/d})(6\text{-h react/cycle})(0.66\text{-h aerobic/h react})\\ &= 12\text{-h aerobic/d}\end{aligned}$$

   c. Determine AOR during aeration period.

$$\text{AOR during each aeration period} = (526\ kg\ O_2/d)/(12\ h/d) = 43.8\ kg\ O_2/h$$

d. Determine the required reactor volume.

Using the oxygen requirement computed above, reactor volume at minimum liquid level is:

$$\text{Reactor volume at minimum liquid level} = \frac{\text{AOR}}{\text{Design OUR}} = \frac{(43.8\text{ kg O}_2/\text{h})}{(150\text{ g O}_2/\text{m}^3\cdot\text{h})} \times \frac{10^3\text{ g}}{1\text{ kg}} = 292\text{ m}^3$$

Maximum reactor volume corresponds to the reactor volume at the minimum liquid level plus the additional volume needed for the maximum hydraulic load per cycle.

$$\text{Maximum hydraulic load/cycle} = \frac{(\text{Maximum hydraulic load/d})}{(\text{Number of cycles/d})}$$

$$= \frac{(\text{maximum sidestream flow}) + (\text{maximum dilution water})}{(\text{Number of cycles/d})}$$

$$= \frac{[(149\text{ m}^3/\text{d}) + (30\text{ m}^3/\text{d})]}{(3\text{ cycles/d})}$$

$$= 60\text{ m}^3/\text{cycle}$$

Reactor volume at maximum liquid level $= 292\text{ m}^3 + 60\text{ m}^3 = 352\text{ m}^3$

2. Determine methanol requirement. Assume 100% oxidation of ammonium to nitrite and 100% reduction of nitrite to N_2.
   a. Determine COD requirement for denitritation.
   Using Eq. (15–12) and assuming a biomass yield = 0.28 g VSS/g COD,

   $$\text{COD/N} = 1.71/(1 - 1.42 \times 0.28) = 2.85\text{ kg COD/kg NO}_2\text{ N}$$

   b. Calculate the annual ammonium loading.

   $$\begin{aligned}\text{Annual ammonium loading} &= (1000\text{ g N/m}^3)(124\text{ m}^3/\text{d})(365\text{ d/y})\\ &= (1\text{ kg N/m}^3)(124\text{ m}^3/\text{d})(365\text{ d/y})\\ &= 45{,}260\text{ kg N/y}\end{aligned}$$

   c. Determine methanol consumption.

   From the problem statement, COD of the methanol is 1.5 kg COD/kg-methanol.

   Average annual methanol consumption

   $$\begin{aligned}&= (45{,}260\text{ kg N/y})[(2.85\text{ kg COD/kg N})/(1.5\text{ kg COD/kg methanol})]\\ &= 85{,}994\text{ kg methanol/y}\end{aligned}$$

   Specific gravity of methanol is 790 kg/m^3, therefore annual methanol consumption is:

   $$\begin{aligned}\text{Annual methanol consumption, m}^3/\text{y} &= (85{,}994\text{ kg methanol/y})(1/790\text{ kg/m}^3)\\ &= 109\text{ m}^3/\text{y}\end{aligned}$$

Solution, Part B—Deammonification SBR

1. Determine deammonification reactor volume.
   a. Determine design oxygen demand.

   From Eq. (15–13), the oxygen requirement for complete deammonification = 1.87 kg O_2/kg NH_4 N.

$$\begin{aligned}\text{Maximum nitrogenous oxygen demand} &= (149\ \text{m}^3/\text{d})(1000\ \text{g N/m}^3)(1.87\ \text{kg O}_2/\text{kg N})\\ &= 279\ \text{kg O}_2/\text{d}\end{aligned}$$

Maximum carbonaceous oxygen demand = 6% of nitrogenous oxygen demand.

$$\text{Maximum carbonaceous oxygen demand} = (0.06)(279\ \text{kg O}_2/\text{d}) = 17\ \text{kg O}_2/\text{d}$$

$$\text{Total maximum oxygen demand} = 273\ \text{kg O}_2/\text{d} + 17\ \text{kg O}_2/\text{d} = 290\ \text{kg O}_2/\text{d}$$

b. Determine the time of aerobic reaction.

$$\begin{aligned}\text{Total aerobic time/d} &= (3\ \text{cycles/d})(6\text{-h react/cycle})(0.66\text{-h aerobic/h react})\\ &= 12\text{-h aerobic/d}\end{aligned}$$

c. Determine AOR during aeration period.

$$\text{AOR during each aeration period} = 290\ \text{kg O}_2/\text{d} \div 12\ \text{h/d} = 24\ \text{kg O}_2/\text{h}$$

d. Determine the reactor volume requirement.
Using the oxygen requirement obtained from Step (c), reactor volume at minimum liquid level is calculated as:

$$\begin{aligned}\text{Reactor volume at minimum liquid level} &= \frac{\text{AOR}}{\text{Design OUR}}\\ &= \frac{(24\ \text{kg O}_2/\text{h})(10^3\ \text{g/1 kg})}{(150\ \text{g O}_2/\text{m}^3{\cdot}\text{h})}\\ &= 160\ \text{m}^3\end{aligned}$$

Maximum reactor volume corresponds to the reactor volume at the minimum liquid level plus the additional volume needed for the maximum hydraulic load per cycle.

$$\begin{aligned}\text{Maximum hydraulic load/cycle} &= \frac{(\text{Maximum hydraulic load/d})}{(\text{Number of cycles/d})}\\ &= \frac{(\text{maximum sidestream flow}) + (\text{maximum dilution water})}{(\text{Number of cycles/d})}\\ &= \frac{[(149\ \text{m}^3/\text{d}) + (15\ \text{m}^3/\text{d})]}{(3\ \text{cycles/d})}\\ &= 55\ \text{m}^3/\text{cycle}\end{aligned}$$

Reactor volume at maximum liquid level = 160 m^3 + 55 m^3 = 215 m^3, or 61% of the volume requirement for nitritation-denitritation.

2. Determine methanol requirement.
The deammonification process, as shown in Eq. (15–14), does not require an external carbon source. Therefore methanol requirement = 0 m^3/y

Comment Although Deammonification does not require an external organic carbon source, the nitrate product from the anammox reaction would have to be reduced with a carbon source to yield an equivalent inorganic nitrogen removal efficiency. In the calculations presented above it was assumed that the maximum OUR of the aeration system controls the design reactor volume, which is typical of a conventional digester sidestream. In the case where sidestream TSS is consistently and considerably higher and no reduction is provided, the accumulation of inert suspended solids will increase the reactor MLSS concentration and may result in a need for a greater reactor volume. For this situation, reduction of the sidestream TSS may be cost effective.

PROBLEMS AND DISCUSSION TOPICS

15–1 Alkaline stabilization of combined primary and waste activated solids is currently practiced at a wastewater treatment facility, but there is a plan to replace this system with a single-stage mesophilic anaerobic digestion process. Centrifugation has been selected for biosolids dewatering. In the absence of pilot study data, a preliminary estimate of the daily sidestream volume and characteristics are required to assess the nutrient and solids loads returned to the mainstream nutrient removal process. Using the following data and assumptions, estimate the future daily sidestream volumes and concentrations of soluble TKN, total TKN, soluble orthophosphate, and total suspended solids at the average and peak thickened sludge loads to the proposed digestion and dewatering processes. In the calculation of the orthophosphate concentration, assume no precipitation of phosphate. The contribution of centrifuge wash-water to the sidestream flow can also be ignored.

Thickened combined raw sludge data

Parameter	Unit	Value
Daily volume, average—maximum two-week	m^3/d	530–700
Total solids concentration (applicable for all flowrates)	%	4.5
Volatile fraction	%	78
Volatile fraction nitrogen content	%	6.5
Volatile fraction phosphorus content	%	1.5
Specific gravity		1.02

Single-stage mesophilic digestion performance criteria

Parameter	Unit	Value
Temperature	°C	35
Volatile solids destruction efficiency at minimum digester SRT	%	45
Volatile solids destruction efficiency at average digester SRT	%	50
Digested volatile solids nitrogen content	%	6.5
Digested volatile solids phosphorus content	%	1.5

Biosolids dewatering performance criteria (applicable for all sludge loading rates)

Parameter	Unit	Value
Cake solids concentration	%	22
Solids capture efficiency	%	95

15–2 Digested primary and waste activated sludge is dewatered six day per week and eight hours per day. The average sidestream flow, ammonium-N concentration and soluble orthophosphate concentration are shown below:

Parameter	Unit	Value
Flowrate during biosolids dewatering,	m^3/h	83
Ammonium-N concentration	mg/L	1050
Orthophosphate-P concentration	mg/L	190
Maximum period without dewatering	d	2

Using the given information, (a) calculate the tank volume required for full equalization of the sidestream so that the equalized flow is returned continuously at constant flowrate to the mainstream plant and (b) the required volume if the equalized sidestream is to be returned to the mainstream plant seven days per week between the hours of 10 p.m. and 6 a.m.

15–3 The sidestream described in Problem 15-2 is discharged to the inlet of the primary clarifiers in the mainstream process. Ferric chloride is applied to the primary clarifiers at an average Fe/P mass ratio of 2 kg/kg to reduce the orthophosphate concentration in the primary tank effluent. Using the given data,

a. Estimate the volume of concentrated ferric chloride solution required to precipitate the orthophosphate-P contributed by sidestream to the primary tank influent. The physical property data for the ferric chloride solution is shown below:

Parameter	Unit	Value
$FeCl_3$ concentration, percent by weight	%	37
Specific gravity of concentrated solution		1.4

b. The implementation of a struvite crystallization process for sidestream pretreatment is being considered. Assuming negligible loss of phosphate during to precipitation in the equalization tank, a target soluble orthophosphate-P concentration of 15 mg/L in the crystallizer effluent and 100 percent recovery of the crystallized product, estimate the average daily mass of struvite that can be potentially harvested from this process. An air-dried product in the form of a hexahydrate salt should be assumed in the calculation.

15–4 For the sidestream described in Problem 15–2, a process consisting of struvite crystallization followed by deammonification in a SBR has been proposed for pretreatment before discharge to the mainstream plant. Using the struvite crystallization criteria provided in Problem 15–3(b),

a. Calculate the ammonium-N concentration in the struvite crystallizer effluent.
b. Estimate the deammonification SBR volume requirement. The following conditions apply:
    i. Sidestream flow will be equalized as calculated in Problem 15–2(a) and this equalized flow will serve as the design basis.
    ii. An energy balance has revealed that sufficient heat losses from the equalization tank, struvite crystallizer and SBR will occur to maintain the SBR temperature below 38°C during the peak summer conditions; thus, no dilution water is required.
    iii. Fine bubble diffusers will be used and can satisfy a maximum OUR of 150 mg/L·h.
    iv. Carbonaceous OUR is estimated to be 8 percent of the nitrogenous OUR.
    v. Intermittent aeration is applied with each aeration cycle time consisting of 66 percent aerobic and 34 percent anoxic.
    vi. Three SBR cycles per day with each cycle consisting of 6-h *react*, 1-h *settle* and 1-h *decant*.
    vii. Assume an ammonia removal efficiency of 100 percent as the design basis.

15–5 A sidestream derived from mesophilic anaerobic digestion of primary and waste activated sludge is produced at a equalized daily volume of 600 m^3/d and contains an ammonium concentration of 900 mg N/L. For the three biological treatment options, nitrification-denitrification, nitritation-denitritation and deammonification,

a. Calculate the daily biological heat generated by the three processes in megajoules/day using the following performance conditions:
    i. Ammonium-N removal efficiency = 95% in all processes
    ii. NOx-N removal efficiency = 95% for denitrification and denitritation.

iii. Denitrification of the nitrate produced by the anammox reaction with sidestream degradable carbon is not significant; thus, this reaction can be excluded from the heat calculation.

iv. Net sludge yield, $Y_H = 0.2$ g VSS/g COD and is applicable to denitrification and denitritation.

b. Determine if the biological heat generated in the three processes is sufficient to increase the reactor temperature beyond a maximum limit of 38°C. If the temperature will exceed 38°C, determine the fraction of the biological heat in each process that must be removed or absorbed via dilution water addition. The equalized sidestream has a peak temperature of 35°C.

15–6 For the sidestream and the process performance criteria described in Problem 15-5, calculate the daily mechanical mixing and aeration energy consumption (kWh/d) for nitrification-denitrification, nitritation-denitritation, deammonification or all three processes (instructor's preference). The following information should be used for the calculations:

Sidestream Characteristics

Parameter	Unit	Value
Equalized flowrate for design	m^3/d	600
Ammonium-N concentration	mg/L	900
Degradable COD concentration[a]	mg/L	200

[a] sum of the readily and complex biodegradable COD.

A sequencing batch reactor is selected for design. The SBR will be fed continuously during the *react* period of each SBR cycle. The following operating and design conditions apply to all three processes:

Parameter	Unit	Value
Number of SBR cycles per day	–	3
Settle period duration	h/cycle	1
Decant period duration	h/cycle	1
Aerobic fraction of react period	%	66
Average reactor temperature	°C	34
Idle period duration	h/cycle	0
Maximum OUR for design	mg/L·h	150
Maximum sidewater depth	m	7
Ammonia and degradable COD removal efficiencies for SBR design	%	100
Actual ammonia removal efficiency	%	90
Actual degradable COD removal efficiency	%	95
Net heterotrophic yield, Y_H	gVSS/gCOD	0.2

Based on a heat loss analysis for open top concrete reactors, the following average daily dilution water requirements were estimated for each process:

Process	Unit	Value
Nitrification-denitrification	m^3/d	200
Nitritation-denitritation	m^3/d	100
Deammonification	m^3/d	0

Fine bubble membrane disk diffusers and submersible mixers are selected for design.

Parameter	Unit	Value
Mech. mixing intensity (nitrification-denitrification)[a]	W/m^3	4
Mech. mixing intensity (nitritation-denitritation)[a]	W/m^3	4
Mech. mixing intensity (deammonification)[a]	W/m^3	6
Mixer total efficiency (electrical + mechanical)	%	84
Operating DO conc. (nitrification-denitrification)	mg/L	2.0
Operating DO conc. (nitritation-denitritation)	mg/L	0.5
Operating DO conc. (deammonification)	mg/L	0.3
Alpha factor, α	—	0.5
Fouling factor, F	—	0.85
Beta factor, β	—	0.95
Temperature correction factor, θ	—	1.024
Site barometric pressure	kPa	99.97
Distance from floor to membrane surface	m	0.25
Pressure drop (blower inlet)	kPa	1.7
Pressure drop (piping, valves, diffusers)	kPa	12
Standard oxygen transfer efficiency, SOTE[b]	%/m	6
Average ambient air temperature	°C	20
Blower mechanical efficiency	%	75
Blower motor electrical efficiency	%	90

[a] Mixing intensity at maximum liquid level. The submersible mixers only operate during the anoxic periods and when they operate, the mixers run at a constant speed regardless of the liquid depth.

[b] SOTE per meter of diffuser submergence. For simplification, assume a linear relationship with depth.

15–7 A wastewater treatment facility requires a capacity expansion to accommodate the projected growth in the service population. The following processes will be added to the facility:

i. Primary sedimentation tanks
ii. Gravity thickening of primary sludge
iii. Dissolved air flotation for thickening of waste activated sludge
iv. Mesophilic anaerobic digestion of combined primary and waste activated sludges
v. Screw press for dewatering digested sludge

Using the plant information given below:

a. Calculate the flow and suspended solids concentrations for the gravity thickener overflow and the dissolved air flotation subnatant.
b. Calculate the pressate flow and concentrations of soluble TKN, total suspended solids and soluble phosphorus.
c. Calculate the percent contribution of the pressate soluble TKN and soluble phosphorus to the primary effluent TKN and TP loads to the secondary process.

Future plant influent flowrate and characteristics:

Parameter	Unit	Value
Average daily flowrate	m^3/d	26,500
COD	mg/L	580
Carbonaceous BOD	mg/L	275
Total suspended solids	mg/L	290
Volatile suspended solids	mg/L	226
TKN	mg/L	40
Ammonium-N	mg/L	23
Total phosphorus	mg/L	7
Ortho-phosphorus	mg/L	3.6
Particulate TKN/VSS ratio	—	0.04
Particulate P/VSS ratio	—	0.015
Particulate COD/VSS ratio	—	1.6

Future solids removal, thickening and digester performance criteria:

Parameter	Unit	Value
Gravity thickener TSS capture efficiency	%	93
Thickened primary sludge percent solids	%	6
DAF TSS capture efficiency	%	95
Thickened waste activated sludge percent solids	%	5
Primary tank TSS removal efficiency	%	60
Primary tank underflow percent solids	%	1
Primary tank cBOD removal efficiency	%	30
Screw press TSS capture efficiency	%	95
Digested sludge cake percent solids	%	25
Digester volatile solids destruction	%	50
Nitrogen content of digested VSS	%	6
Phosphorus content of digested VSS	%	1.8

Future secondary process performance and solids production:

Parameter	Unit	Value
SRT	d	12
cBOD removal efficiency	%	98
Observed yield	gVSS/gBOD	0.55
Volatile content of WAS TSS	%	80
WAS MLSS concentration	mg/L	7500
Nitrogen content of WAS VSS	%	9.5
Phosphorus content of WAS VSS	%	2
Filtered plant effluent TSS concentration	mg/L	< 2

15–8 Using Eqs. (7-96) in Sec. 15-8 and (15-11) in Sec. 15-9, estimate the amount of methanol that can be saved with the conversion of a sidestream treatment process from nitrification-denitrification to nitritation-denitritation. Express the results as kilograms of methanol saved per kilogram of NOx-N-removed.

15–9 For the sidestream described in Problem 15-2, and assuming a sidestream alkalinity of 3750 mg/L as $CaCO_3$, estimate the soda ash (Na_2CO_3) dosing rate required for complete nitrification of the sidestream ammonium-N in a separate reactor performing only nitrification. Describe how the soda ash dosing requirement would change if the sidestream treatment process is modified from nitrification-denitrification to nitritation-denitritation.

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16

Air Emissions from Wastewater Treatment Facilities and Their Control

WORKING TERMINOLOGY

Term	Definition
Absorption	The process by which atoms, ions, molecules, and other constituents are transferred from one phase and are distributed uniformly in another phase (see also adsorption).
Adsorption	The process by which atoms, ions, molecules, and other constituents are transferred from one phase and accumulate on the surface of another phase (see also absorption).
Air stripping	The removal of volatile and semi-volatile contaminants from a liquid by passing air and liquid counter currently through a packed tower.
Buffer zone	An area around a facility which serves to diminish the impact of any odors emitted from the facility. Trees are sometimes planted at the periphery of buffer zones to further reduce the impact of odors.
Catalytic incineration	Controlled process used to oxidize VOCs with the help of a catalyst such as platinum and palladium.
Chemical scrubber	A reactor used to provide contact between air, water, and chemicals, if used, to provide oxidation or entrainment of odorous compounds.
Biofilter	Open or closed packed-bed filters used for the removal of odors biologically. In open biofilters, gases to be treated move upward through the filter bed. In closed biofilters, the gases to be treated are either blown or drawn through the packing lateral.
Biotrickling filter	Similar to biofilters, with the exception that moisture is provided continuously or intermittently over the packing. Liquid is recirculated and nutrients are often added.
Digester gas	Gas produced from anaerobic digestion of sludges. Also often referred to as biogas. Digester gas typically contains 60 percent or higher methane gas by volume and can be used as a fuel source.
Gas stripping	The purposeful introduction of air or other gases to transfer volatile constituents such as VOCs and odors from a liquid phase to a gaseous phase.
Global warming potential (GWP)	A measure of relative effects of a gas to trap heat in the atmosphere in reference to that of carbon dioxide.
Greenhouse gases	Gases that have been identified as contributing to global warming.
Mass transfer	The transfer of material from one homogeneous phase to another; aeration, gas stripping, and adsorption are examples of mass transfer.
Mechanical aerators	Devices used to agitate water to promote mixing with atmospheric air.
Mixing	The agitation of a liquid-solids suspension for the purpose of blending the mixture and keeping solids in suspension, entraining gases, or for accelerating a chemical reaction.
Odor threshold	The concentration at which an odor is detectable by human sense of smell.
Off-gas	The gaseous emission from a process; off-gas may be odorous and/or contain greenhouse gases and VOCs.
Stripping tower	A closed vertical reactor used to bring about the transfer of VOCs from a liquid phase to a gaseous phase.
Thermal oxidation	Controlled process used to oxidize VOCs at high temperatures.
Vapor phase adsorption	Process whereby hydrocarbons and other compounds arc are adsorbed selectively on the surface or such materials as activated carbon, silica gel, or alumina.
Volatile organic compounds (VOCs)	A term often used generically to mean total organic carbon; in the context of air quality the term means total nonmethane hydrocarbons.
Volatilization	The release of VOCs from water surface to the atmosphere.

The treatment of wastewater results in the release of a variety of air emissions, many of which are odorous and/or may contain air pollutants. Activities that result in emissions of air pollutants to the outdoor atmosphere may require approval of the state or regional environmental agency or the U.S. Environmental Protection Agency (U.S. EPA). The extent of the air permitting requirements is based on a number of factors such as the nature of the air pollutant, the quantity of emissions, the air quality in the vicinity of the facility, the existing air emissions and emission sources, the state air permitting requirements, and emissions control regulations mandated by the governing agency and federal regulations. The governing agency may be at the state, region, city, or tribal level. Air permitting is typically at the federal and state level, but authority may vary. As used herein, "state" refers to agencies other than the U.S. EPA that have jurisdiction to implement air permitting under the Clean Air Act (CAA) at the state, regional, city, or tribal level. Topics discussed in this chapter include (1) types of emissions, (2) regulatory requirements, (3) odor management, (4) volatile organic carbon emissions and their control, (5) emissions from combustion of gases and solids, and (6) emission of greenhouse gases.

16–1 TYPES OF EMISSIONS

Typical air emissions from wastewater treatment plants that are subject to regulation under the CAA are provided in Table 16–1. Some states require that additional classes of compounds (e.g., odors and greenhouse gases) be calculated at these facilities. For example, a recently promulgated regulation by the U.S. EPA requires that the total emissions from all sources of greenhouse gases be calculated and compared to regulatory thresholds. Certain processes within wastewater treatment plants have qualified for exemptions to the rule due to the "biogenic" nature of their emissions (i.e., non-fossil fuel origin). Emission of greenhouse gases is discussed further in Sec 16–6. Sources of emissions within odor management is particularly important for the operation of wastewater treatment facilities, and it is discussed in detail in Sec 16–3.

16–2 REGULATORY REQUIREMENTS

The CAA is the basis for federal and state regulations, which are codified in Subchapter C, Air Programs, of Title 40 of the Code of Federal Regulations. The CAA and amendments establish air permitting programs at the federal level and provide a regulatory framework for state and regional regulations. Permit authority may be at the federal, state, or regional level. Emissions of air pollutants fall into three general categories: criteria pollutants, non-criteria pollutants, and hazardous air pollutants (HAPs). Criteria pollutants have an associated air quality standard. The key air quality regulations cover broad categories, as described herein.

Ambient Air Quality and Attainment Status

The U.S. EPA has established National Ambient Air Quality Standards (NAAQS) to protect public health and public welfare. Primary standards are based on observable human health responses and are set at levels that provide an adequate margin of safety for sensitive segments of the population. Secondary standards are intended to protect public welfare interests such as structures, vegetation, and livestock. States may also establish ambient air quality standards that are more stringent than the federal standards and may retain a revoked federal standard.

Table 16–1

Typical air pollutants associated with wastewater treatment plants

Air pollutant	Source(s)
Criteria pollutants	
Carbon monoxide (CO)	Incomplete combustion, partial oxidation of organic material
Nitrogen dioxide (NO_2)	Combustion processes
Sulfur dioxide (SO_2)	Combustion processes
Total suspended particulate matter (TSP)	Combustion processes, material handling, aggregate handling, other process sources
Respirable particulate matter with a diameter of up to 10 microns (PM_{10})	Combustion processes, material handling, aggregate handling, other process sources
Fine particulate matter with a diameter of up to 2.5 microns ($PM_{2.5}$)	Combustion processes, material handling, aggregate handling, other process sources
Ozone (O_3):	Generated at ground level through photochemical oxidation of precursors NOx and VOC
Oxides of nitrogen (NOx)	Combustion processes
Volatile organic compounds (VOC)	Combustion processes, organic material storage and use, other process sources
Lead (Pb)	Combustion processes
Non-criteria pollutants	
Hydrogen sulfide (H_2S)	Process sources, anaerobic reduction of sulfur compounds
Methane (CH_4)	Anaerobic digestion, combustion processes
Carbon dioxide (CO_2)	Combustion processes, anaerobic digestion
Ammonia (NH_3)	Solids processing/polymer breakdown
Nitrous oxide (N_2O)	Biological nitrogen removal systems
Hazardous air pollutants (HAPs)	
Toluene, benzene, xylene, etc	Influent constituents from industrial sources
Methanol	Nutrient removal systems
Trimethylammine and dimehtylammine	Solids processing/polymer breakdown
Carbon disulfide and carbonyl sulfide	Total reduced sulfur compounds from sulfate oxidation
Formaldehyde and hexane	Combustion by-products from combined heat and power/boilers/process heaters/etc.
Chlorine (Cl_2)	Used in chlorination processes
Mercury, other heavy metals and polycyclic organic compounds (POM)	Wastewater sludge incinerators

The CAA requires the U.S. EPA and states to identify by category the ambient air quality compliance status for specific geographic regions (air quality control regions or portions thereof). Areas may be designated “attainment” or “nonattainment” of the NAAQS based on monitoring data, or “unclassifiable” if insufficient ambient monitoring data exists; “unclassifiable” areas are generally considered as “attainment” areas. The attainment status of the existing/new source location determines the applicability of pre-construction permitting programs.

Table 16–2

Permitting programs and control technologies

Requirements	Regulations, rules and technologies
Preconstruction and operating permitting	Nonattainment new source review (NNSR or NANSR)
	Prevention of significant deterioration (PSD)
	Title V operating permits
	Minor source permitting programs
Stationary source control permitting	New Source Performance Standards (NSPS per 40 CFR Part 60)
	National Emission Standards for Hazardous Air Pollutants (NESHAPs per 40 CFR Parts 61 and 63)
	Reasonably Available Control Technology (RACT)

Preconstruction and Operating Permitting Programs

New sources and modifications/reconstructions of existing sources that result in air pollution emissions (above specified thresholds) will be subject to air permitting requirements. Air permitting programs are listed in Table 16–2. The applicability of nonattainment new source review (NNSR) and prevention of significant deterioration (PSD) depends upon the nature of the project, the air quality designation in the vicinity of the project, and the quantity of annual emissions. Both are preconstruction approval programs applicable to major projects, and both prohibit commencement of construction until the preconstruction approval is issued. If major source or major modification criteria are met, NNSR may be applicable to a nonattainment pollutant (and precursor pollutants) while PSD may be applicable to an attainment pollutant. If NNSR is applicable, then the lowest achievable emission rate (LAER) control technology that is technically feasible must be incorporated in the project design, regardless of cost.

If PSD is applicable, then the best available control technology (BACT) must be incorporated into the project design. Factors such as cost, energy requirements, and other environmental consequences/benefits are part of a BACT analysis. The U.S. EPA may delegate authority to approve NNSR and PSD projects to state agencies; alternatively, it may approve state regulations that meet U.S. EPA requirements (e.g., approval of State Implementation Plan). Within the PSD program, more stringent protection of air quality is afforded to "Class I" areas, which include national parks, wildlife areas, and other designated areas. Impacts to visibility and ecology must also be minimized for the project to obtain an air permit to construct.

The CAA Amendments of 1990 added the Title V operating permit program, which established state and federal permitting procedures for major facilities that allow for the consolidation of facility specific requirements into a single, federally-enforceable permit. These requirements include emission limitations, work practice standards, monitoring requirements, recordkeeping requirements, and submittal/notification requirements. Applicable requirements may be based on regulations or preconstruction approval conditions. Title V permitting programs are usually administered at the state level. Minor source permitting programs apply to sources that are not subject to Title V operating permits. These requirements are established by state agencies.

Stationary Source Control Technology Requirements

Stationary source emissions control technology requirements are listed in Table 16–2. Unlike the control technology requirements of the NNSR and PSD preconstruction programs, regulations such as new source performance standards (NSPS) and national emission standards for hazardous air pollutants (NESHAPs) apply to equipment based on the function and category of the equipment. The NSPS apply to new, modified, or reconstructed sources that meet the applicability criteria specific to the source type. The NESHAP requirements apply primarily to major sources of HAPs, but some requirements are applicable to "area sources" (minor sources) of HAPs. Reasonably available control technology (RACT) broadly refers to the control technology requirements specified by state regulations that establish the minimum level of emissions control. RACT requirements can be applied retroactively to a source and are generally more stringent in nonattainment areas than in attainment areas.

16–3 ODOR MANAGEMENT

The potential release of odors is a major concern of the public relative to modifying existing wastewater treatment facilities and constructing new facilities. Thus, the control of odors has become a major consideration in the design and operation of wastewater collection, treatment, and disposal facilities, especially with respect to the public acceptance of these facilities. In many instances, projects have been rejected because of the fear of potential odors. In several states, wastewater management agencies are now subject to fines and other legal action over odor violations. In view of the importance of odors in the field of wastewater management, the following topics are considered in this section: (1) the types of odors encountered, (2) the sources of odors, (3) measurement of odors, (4) the movement of odorous gases, (5) strategies for odor control, (6) odor control methods, and (7) the design of odor control facilities.

Types of Odors

For humans, the importance of odors at low concentrations is related primarily to the psychological stress the odors cause, rather than to the harm they do to the body. The principal types of odors encountered in wastewater management facilities are reported in Table 16–3. With few exceptions, odorous compounds typically contain either sulfur or nitrogen. The characteristic odor of organic compounds containing sulfur is that of decayed organic material. Of the odorous compounds reported in Table 16–3, the rotten egg smell of hydrogen sulfide is the odor encountered most commonly in wastewater management facilities. As noted in Chap. 2, gas chromatography has been used successfully for the identification of specific compounds responsible for odors. Unfortunately, this technique has not proved as successful in the detection and quantification of odors derived from wastewater collection, treatment, and disposal facilities, because of the many compounds that may be involved. It should be noted that at higher concentrations, many of the odorous gases (e.g., hydrogen sulfide) can, depending on exposure, be lethal.

Sources of Odors

The principal sources of odors in wastewater management facilities and the relative potential for release of odor are presented in Table 16–4. Minimization of odors from these sources is the concern of odor management.

Wastewater Collection Systems. The principal sources of odorous compounds in collection systems are from (1) the biological conversion, under anaerobic conditions, of organic matter containing nitrogen and sulfur, and (2) the discharge of industrial wastewater that may contain odorous compounds or compounds that may react with compounds in the wastewater to produce odorous compounds. Odorous gases released to the sewer atmosphere can accumulate and be released at air release valves, cleanouts, access ports (i.e., manholes), and house vents.

Table 16–3

Odor thresholds of odorous compounds and their characteristics associated with wastewater management

Odorous compound	Chemical formula	Molecular weight	Odor threshold, ppm_v[a]	Characteristic odor
Ammonia	NH_3	17.0	46.8	Pungent, irritating
Chlorine	Cl_2	71.0	0.314	Pungent, suffocating
Chlorophenol	ClC_6H_4OH	128.51	0.00018	Medicinal odor
Crotyl mercaptan	$CH_3\text{-}CH{=}CH\text{-}CH_2\text{-}SH$	90.19	0.000029	Skunk like
Dimethyl sulfide	$CH_3\text{-}S\text{-}CH_3$	62	0.0001	Decayed cabbage
Diphenyl sulfide	$(C_6H_5)_2S$	186	0.0047	Unpleasant
Ethyl mercaptan	$CH_3CH_2\text{-}SH$	62	0.00019	Decayed cabbage
Ethyl sulfide	$(C_2H_5)_2SH$	91.9	0.000025	Nauseating odor
Hydrogen sulfide	H_2S	34	0.00047	Rotten eggs
Indole	C_8H_6NH	117	0.0001	Fecal, nauseating
Methyl amine	CH_3NH_2	31	21.0	Putrid, fishy
Methyl mercaptan	CH_3SH	48	0.0021	Decayed cabbage
Skatole	C_9H_9NH	132	0.019	Fecal odor, nauseating
Sulfur dioxide	SO_2	64.07	0.009	Pungent, irritating
Thiocresol	$CH_3\text{-}C_6H_4\text{-}SH$	124	0.000062	Skunk like, irritating
Trimethyl amine	$(CH_3)_3N$	59	0.0004	Pungent, fishy

[a] Parts per million by volume.

Table 16–4

Sources of odor in wastewater management systems[a]

Location	Source/cause	Odor potential
Wastewater collection system		
Air release valves	Accumulation of odorous gases released from wastewater	High
Cleanouts	Accumulation of odorous gases released from wastewater	High
Access ports (manholes)	Accumulation of odorous gases released from wastewater	High
Industrial wastewater discharges	Odorous compounds may be discharged to wastewater collection system	
Raw wastewater pumping station	Wetwell/septic raw wastewater, solids and scum deposits	High
Wastewater treatment facilities		
Headworks	Release odorous gases generated in the wastewater collection system due to turbulence in hydraulic channels and transfer points	High
Screening facilities	Putrescible matter removed by screening	High
Preaeration	Release of odorous compounds generated in wastewater collection system	High

(*continued*)

Table 16–4 (*Continued*)

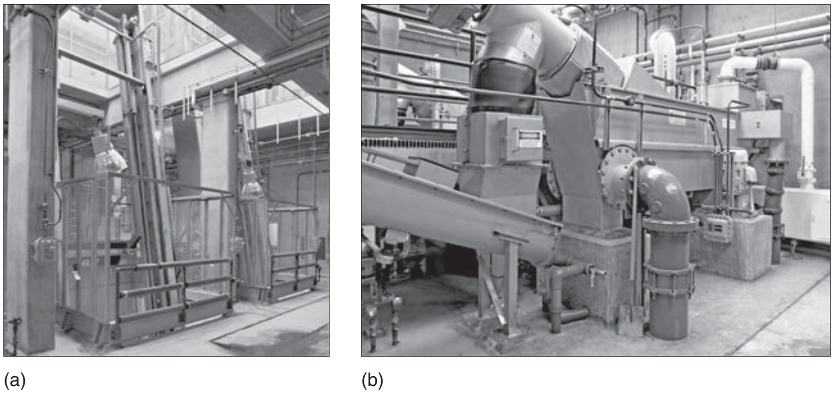
Location	Source/cause	Odor potential
Grit removal	Organic matter removed with grit	High
Flow equalization basins	Basin surfaces/septic conditions due to accumulation of scum and solids deposits	High
Septage receiving and handling facilities	Odorous compounds can be released at septage receiving stations, especially when septage is being transferred	High
Sidestream returns[b]	Return flows from biosolids processing facilities	High
Primary clarifiers	Effluent weirs and troughs/turbulence that releases odorous gases. Scum-either floating or accumulated on weirs and baffles/putrescible matter. Floating sludge/septic conditions	High/moderate
Fixed film processes (trickling filters or RBCs)	Biological film/septicity due to insufficient oxygen, high organic loading, or plugging of trickling filter medium; turbulence causing release of odorous material	Moderate/high
Aeration basins	Mixed liquor/septic return sludge, odorous sidestream flows, high organic loading, poor mixing, inadequate DO, solids deposits	Low/moderate
Secondary clarifiers	Floating solids/excessive solids retention	Low/moderate
Sludge and biosolids facilities		
Thickeners, solids holding tanks	Floating solids; weirs and troughs/scum and solids septicity due to long holding periods, solids deposits, and temperature increases; odor release by turbulence	High/moderate
Aerobic digestion	Incomplete mixing in reactor	Low/moderate
Anaerobic digestion	Leaking hydrogen sulfide gas/upset conditions, high sulfate content in solids	Moderate/high
Sludge storage basins	Lack of mixing, formation of scum layer	Moderate/high
Mechanical dewatering by belt filter press, recessed plate filter press, or centrifuge	Cake solids/putrescible matter; chemical addition, ammonia release	Moderate/high
Sludge loadout facilities	Release of odors during the transfer of biosolids from storage to transfer facilities	High
Composting facilities	Composting solids/insufficient aeration, inadequate ventilation	High
Alkaline stabilization	Stabilized solids/ammonia generation resulting from reaction with lime	Moderate
Incineration	Air emissions/combustion temperature is not high enough to destroy all organic substances	Low
Sludge drying beds	Drying solids/excess putrescible matter due to insufficient stabilization	Moderate/high

[a] Adapted in part from WEF (1996a).

[b] Sidestreams could include digester decant, dewatering return flows, or backwash water (see Chap. 15).

Figure 16–1

Typical examples of odor management for preliminary treatment processes: (a) bar screen in enclosed building with odor control facilities and (b) enclosed grit processing facilities.



Wastewater Treatment Facilities. In considering the potential for the generation and release of odors from treatment plants, it is common practice to consider the liquid and solids processing facilities separately. The headworks and preliminary treatment operations have the highest potential for release of odor, especially for treatment plants that have long collection systems where anaerobic conditions can develop (see Fig. 16–1). Sidestream discharges including return flows from filter backwashing and from sludge and biosolids processing facilities are often a major source of odors, especially where these flows are allowed to discharge freely into a control structure or mixing chamber.

Sludge and Biosolids Handling Facilities. Typically, the most significant sources of odors at wastewater treatment plants are sludge thickening facilities, anaerobic digesters, and sludge loadout facilities. The highest potential for odor release occurs when unstabilized sludge is handled (e.g., turned, spread, or stored).

One of the major contributors to odor in solids processing and an important item to consider in treatment plant design is shear. Shear is the cutting or tearing of solids by shear stress. When solids undergo mixing by either high shear dewatering or conveyance equipment, particle size reduction occurs, and odor production increases. Solids that exit a dewatering facility can be sheared enough to release odors. The major mechanism appears to be the release of proteinaceous biopolymer. Once released, these proteins are degraded, liberating a number of odorous compounds, but mostly mercaptans. The increase in solution protein also makes dewatering more difficult. The solution proteins can be "coagulated" by addition of polymer, but the synthetic polymers are degraded and the protein becomes degradable. The synthetic polymer can also generate methylamines when degraded (Novak, 2001; Murthy, 2001).

Trimethylamine (TMA) is present in the liquid phase in many anaerobically digested sludges. Trimethylamine, like ammonia, is soluble below pH 9, but above this pH level is a gas, which can be released into the air. Adding lime to digested sludge for odor control may, in fact, enhance the release of odors by converting TMA to a gas (Novak, 2001; Murthy, 2001). Some plants may be unable to land apply dewatered sludge because of the increased odor production. Thus, in evaluating processing and disposal options, the ramifications of odor generation and control have to be evaluated carefully.

Measurement of Odors

The sensory (organoleptic) measurement of odors by the human olfactory system is used most often to detect odors emanating from wastewater treatment facilities. Detection of odors by the human olfactory system and instrumental methods is described in Chap. 2.

Although there are several different ways of assessing the impact of odors, the primary method at wastewater treatment facilities is dilution-to-threshold ratio (D/T). The D/T ratio is a measure of the number of dilutions of fresh air needed to render the odorous ambient air nondetectable. The higher the D/T value, the greater amount of fresh air needed to render the odor nondetectable. The D/T values, determined by an odor panel, are used as a regulatory standard for assessing off-site odor impacts. Typical D/T values vary from 1 to 50, with a median value of 5. The location where D/T impacts are measured varies with the governing jurisdiction, but typically is at the location of the receptor of odor, e.g., property fence line, residence, community, park, etc. Modeling, discussed subsequently, is used to predict the D/T values at the receptor location using source-specific D/T values.

There are two challenges with the measurement of odors; one challenge is that the detection threshold varies between individuals, and the second challenge is that odorous compounds tend to be mixed together, such as hydrogen sulfide and methyl mercaptan. The first challenge can be overcome through the use of standard methods for measuring odor, such as ASTM E679–04 where a specialized device presents different D/T mixtures to subjects for testing. The second challenge can only be overcome through the use of an odor panel, as the impact of individual odors cannot be summed. Typical probability distributions of threshold odor levels for different odorous compounds, as reported in the literature, are illustrated on Fig. 16–2. As shown on Fig. 16–2, the range of threshold odor values can vary by as much as seven orders of magnitude.

Odor Dispersion Modeling

Odor dispersion modeling is used to assess the impact from an odor source at the receptor location and the type and the extent of the odor management facilities must be implemented. Typically, the odor dispersion model is run over a set time period, and a probability of exceeding a given D/T value at the receptor location is estimated. Input to the model must be obtained from the emission source, and usually these samples must be sent for analysis by an odor panel. The model run time corresponds to the amount of time a receptor can be exposed to an odor. Average expose times range between 3 and 60 min. Depending on the governing jurisdiction or agency, the odor D/T value at the receptor location must be lower than the established regulatory D/T value (e.g., 5), 98 to 100 percent of the time.

Movement of Odors from Wastewater Treatment Facilities

Under quiescent meteorological conditions (i.e., calm winds and low atmospheric mixing), odorous gases that develop at treatment facilities tend to hover over the point of generation (e.g., sludge thickening facilities, sludge storage lagoons), because the odorous gases are more dense than air. Depending on the local meteorological conditions, it has been observed that odors may be measured at undiluted concentrations at great distances from the point of generation. The following events appear to happen: (1) in the evening or early morning hours, under quiescent meteorological conditions, a cloud of odors will develop over the wastewater treatment unit prone to the release of odors; and (2) the concentrated cloud of odors can then be transported (i.e., pushed along), without breaking up, over great distances by the weak evening or early morning breezes, as they develop. In some cases, odors have been detected at distances of up to 25 km from their source. This transport phenomenon has been termed the puff movement of odors (Tchobanoglous and Schroeder, 1985). The puff movement of odors was first described by Wilson (1975). Air dispersion modeling can be used to predict whether quiescent conditions will tend to persist over prolonged periods of time. The most common method used to mitigate the effects of the

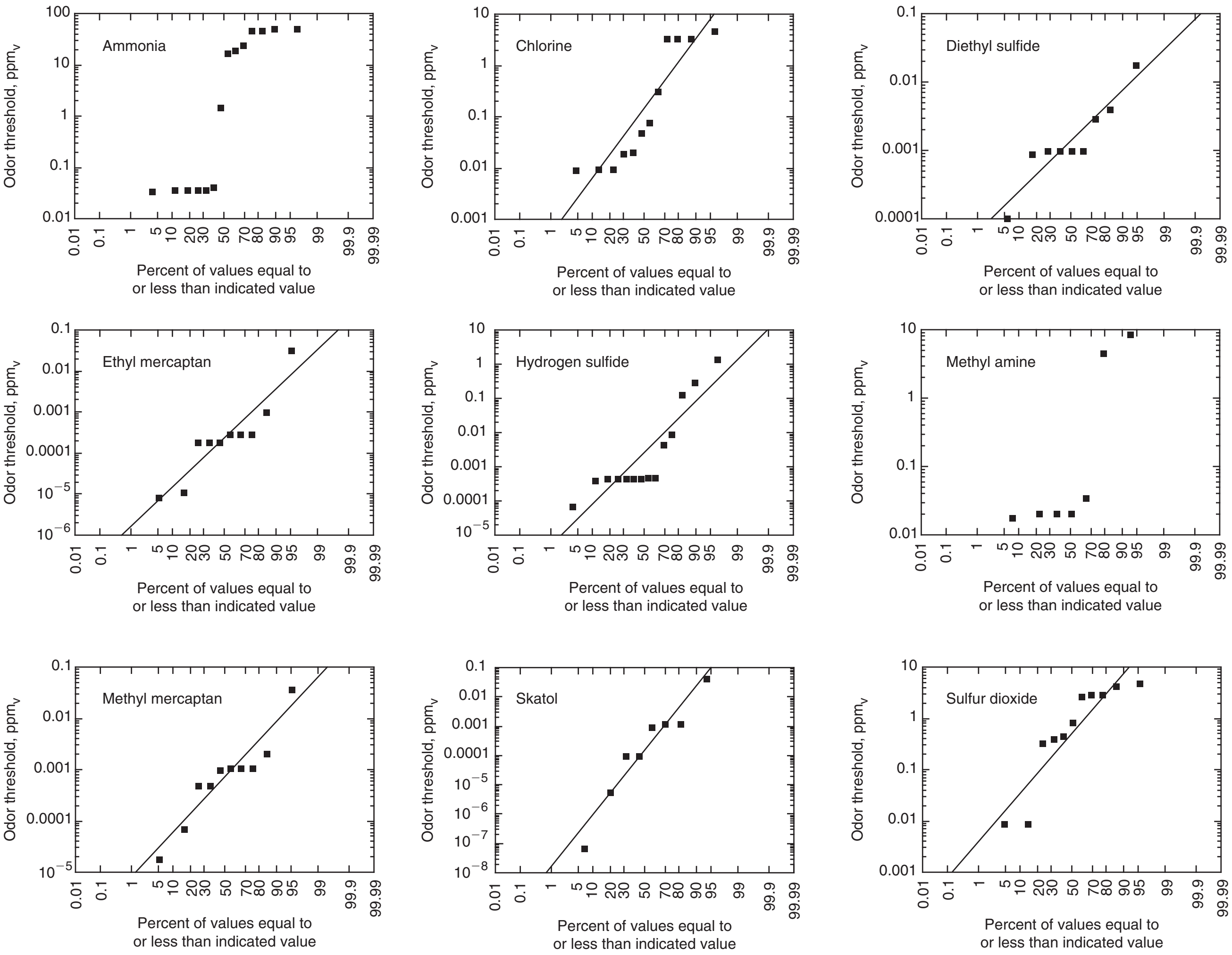


Figure 16–2

Log probability plots of odor threshhold odor values reported in the literature for a variety of odorous compounds.

odor puff is to install barriers to induce turbulence, thus breaking up and dispersing the cloud of concentrated odors, and/or to use wind generators to maintain a minimum velocity across the source.

Strategies for Odor Management

Strategies for the management and control of odors are presented and discussed below. An overview of some of the methods used to control and treat odorous gases is presented in the following section. Where chronic odor problems occur at treatment facilities, approaches to solving these problems may include (1) control of odor-causing wastewaters discharged to the collection system and treatment plant that creates odor problems, (2) control of odors generated in the wastewater collection system, (3) control of odors generated in wastewater treatment facilities, (4) installation of odor containment and treatment facilities, (5) application of chemicals to the liquid (wastewater) phase, (6) use of odor masking and neutralizing agents, (7) use of gas-phase turbulence-inducing structures and facilities, and (8) establishment of buffer zones.

Control of Discharges to Wastewater Collection System. The elimination and/or control of wastewater discharges containing odorous compounds to the collection system can be accomplished by (1) adopting more stringent waste discharge ordinances and enforcement of their requirements, (2) requiring pretreatment of industrial wastewater, and (3) providing flow equalization at the source to eliminate slug discharges of wastewater.

Odor Control in Wastewater Collection Systems. The release of odors from the liquid phase in wastewater collection systems can be limited by (1) maintaining aerobic conditions through the addition of hydrogen peroxide, pure oxygen, or air at critical locations in the collection system and to long force mains (see Fig. 16–3), (2) controlling anaerobic microbial growth by disinfection or pH control, (3) oxidizing or precipitating odorous compounds by chemical addition, (4) design of the wastewater collection system to minimize the release of odors due to turbulence, and (5) off-gas treatment at selected locations.

Odor Control in Wastewater Treatment Facilities. With the proper attention to design details, such as the use of submerged inlets and weirs, the elimination of hydraulic jumps in influent piping and channels, the elimination of physical conditions leading to the formation of turbulence, proper process loadings, containment of odor sources, off-gas treatment, and good housekeeping, the routine release of odors at treatment plants can be minimized. It must also be recognized, however, that odors will develop occasionally. When they do, it is important that immediate steps be taken to control them. Some steps that can be taken include operational changes or the addition of chemicals, such as chlorine, sodium hypochlorite, hydrogen peroxide, lime, or ozone.

Design and operational changes that can be instituted can include (1) minimizing free-fall turbulence by controlling water levels, (2) reducing of overloading of plant processes, (3) increasing the aeration rate in biological treatment processes, (4) increasing the plant treatment capacity by operating standby process units, (5) reducing solids inventory and sludge backlog, (6) increasing the frequency of pumping of sludge and scum, (7) adding chlorinated dilution water to sludge thickeners, (8) controlling the release of aerosols, (9) increasing the frequency of disposal of grit and screenings, (10) cleaning odorous accumulations more frequently, and (11) containment, ventilation, and treatment of odorous gases.

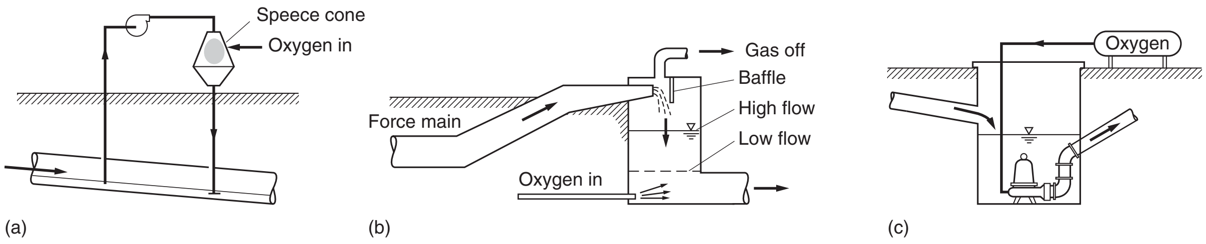
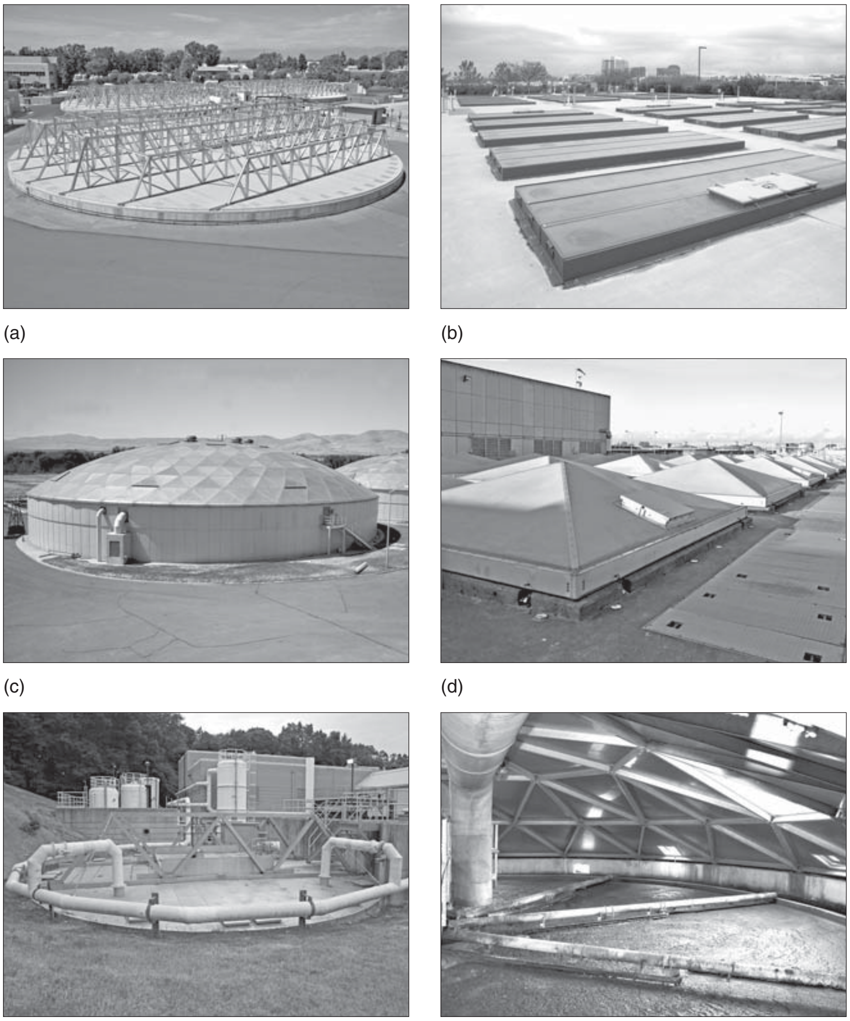


Figure 16–3

Typical uses of commercial oxygen in wastewater collection systems for odor control: (a) sidestream oxygenation and reinjection of wastewater into a gravity sewer, (b) Injection of oxygen into a hydraulic fall and (c) Injection of oxygen into two-phase flow in force main. (From Speece et al., 1990.)



(a) (b) (c) (d) (e) (f)

Figure 16–4

Typical odor containment facilities at wastewater treatment plants: (a) and (b) covered primary sedimentation tanks, (c) covered trickling filter, (d) cover over the solids contact portion of a trickling filter solids contact biological treatment process, (e) covered primary sludge fermentor, and (f) view inside of an enclosed sludge thickener.

Odor Containment. Odor containment includes the installation of covers, collection hoods, and air handling equipment for containing and directing odorous gases to disposal or treatment systems. In cases where the treatment facilities are close to developed areas, it has become common practice to enclose or cover treatment units such as the bar screens and grit processing units [see Fig. 16–1(a) and (b)], primary clarifiers [see Figs. 16–4(a) and (b)], trickling filters [see Fig. 16–4(c)], (d) biological treatment processes [see Fig. 16–4(d)], (e) sludge fermentors [see Fig. 16–4(e)], sludge thickeners [see Fig. 16–4(f)], sludge-processing facilities, and sludge-loadout facilities. Where covers are used, the trapped gases must be collected and treated. The specific method of treatment will depend on the characteristics of the odorous compounds. Typical containment alternatives for the control of odor emissions from wastewater management facilities are reported in Table 16–5.

Chemical Additions to Wastewater for Odor Control. Odors can be eliminated in the liquid phase through the addition of a variety of chemicals to achieve (1) chemical oxidation, (2) chemical precipitation, and (3) pH control. The most common oxidizing chemicals that can be added to wastewater include oxygen, air, chlorine, sodium

Table 16–5

Odor containment and process alternatives for the control of odors the emission from wastewater management facilities

Source	Suggested control strategies
Wastewater sewers	Seal existing access ports (i.e., manholes). Eliminate the use of structures that create turbulence and enhance volatilization.
Sewer appurtenances	Isolate and cover existing appurtenances.
Pump stations	Vent odorous gases from wet-well to treatment unit. Use variable-speed pumps to reduce the size of the wet-well.
Bar racks	Cover existing units. Reduce headloss through bar racks.
Comminutors	Cover existing units. Use inline enclosed comminutors.
Parshall flume	Cover existing units. Use alternative measuring device.
Grit chamber	Cover existing aerated grit chambers. Reduce turbulence in conventional horizontal-flow grit chambers; cover if necessary. Avoid the use of aerated grit chambers.
Equalization basins	Cover existing units. Use submerged mixers, and reduce air flow.
Primary and secondary sedimentation tanks	Cover existing units [see Figs. 16–4(a) and (b)]. Replace conventional overflow weirs with submerged weirs.
Biological treatment	Cover existing units. Use submerged mixers and reduce aeration rate.
Transfer channels	Use enclosed transfer channels.

hypochlorite, potassium permanganate, hydrogen peroxide, and ozone. While all of these compounds will oxidize hydrogen sulfide (H_2S) and other odorous compounds, their use is complicated by the chemical matrix in which the odorous gases exist. The only way to establish the required chemical dosages for the removal of chemical compounds is through bench- or pilot-scale testing.

Odorous compounds can also be reduced by precipitation. For example, ferrous chloride and ferrous sulfate can be used for the control of H_2S odors by precipitation of the sulfide ion as ferrous sulfide. As with the oxidation reactions, the required chemical dosage can be determined only through bench- or pilot-scale testing. The release of H_2S can also be controlled by increasing the pH value of the wastewater. Increasing the pH of the wastewater results in reduced bacterial activity and also shifts the equilibrium so that the sulfide ion is present as HS^-. With most of the odor control methods involving the addition of chemicals to wastewater, some residual product is formed that must ultimately be dealt with. Shock treatment involving the addition of sodium hydroxide (NaOH) can be used to reduce microbial slimes in sewers. The high pH also reduces sulfide (S^{2-}) formation. Additional details on chemical addition may be found in Rafson (1998).

Use of Odor Masking and Neutralization. On occasion, chemicals have been added to wastewater or offgases to mask an offensive odor with a less offensive odor. Masking chemicals are based on essential oils with the most common aromas being vanilla, citrus, pine, or floral (Williams, 1996). Typically, enough masking chemical is added to wastewater to overpower the offensive odor. Masking chemicals, however, do not modify or neutralize the offensive odors. Neutralization involves finding chemical compounds that can be combined with the odorous gases in the vapor state so that the

(a) (b)

Figure 16–5

High barrier fence placed around sludge holding lagoons to induce air turbulence and mixing, and thus limit the release of odors off the treatment plant site (a) aerial view of sludge storage basins (coordinates N 38.439, W121,480) and (b) view of high barrier fence.

combined gases cancel each other's odor, produce an odor of lower intensity, or eliminate the odorous compounds. Although odor masking and neutralization are viable options for short-term management of odor problems, the key to long-term odor management is to identify the source of the odors and implement corrective measures.

Use of Turbulence-Inducing Structures and Facilities for Odor Dispersion. In a number of wastewater treatment plants, physical facilities used to induce atmospheric turbulence have been constructed specifically for the purpose of gas-phase odor reduction. The high barrier fence [3.7 m (12 ft)] shown on Fig. 16–5 surrounds sludge-storage lagoons. Operationally, any odorous gases that develop under quiescent conditions over the lagoons are diluted as they move away from the storage lagoons, due to the local turbulence induced by the barrier. Trees are also used commonly to dilute odorous gases by inducing turbulence (i.e., the formation of eddies) and mixing. Trees are also known to help purify the air as a result of respirometric activity.

Use of Buffer Zones. The use of buffer zones can also help in reducing the impact of odors on developed areas. Typical buffer zone distances used by regulatory agencies are presented in Table 16–6. If buffer zones are used, odor studies should be conducted that identify the type and magnitude of the odor source, meteorological conditions, dispersion characteristics, and type of adjacent development. Trees that grow rapidly are often planted at the periphery of the buffer zones to further reduce the impact of odors.

Odor Treatment Methods

The general classification of odor treatment methods is presented in Table 16–7, along with typical applications in wastewater management. Odor treatment methods are designed either to treat the odor producing compounds in the wastewater stream or to treat the foul air. Most of the methods in Table 16–7 are meant to be used to treat the foul air (i.e., the gas phase). As noted above, to control the release of odorous gases from treatment facilities, it has become more common to cover wastewater treatment processes (see Fig. 16–4).

The principal methods used to treat odorous gases include the use of (1) chemical scrubbers, (2) activated-carbon absorbers, (3) vapor-phase biological treatment processes (i.e., compost filters), (4) treatment in conventional biological treatment processes, and (5) thermal processes. Each of these methods is discussed below. The specific method of odor control and treatment that should be applied will vary with local conditions. However, because odor control measures are expensive, the cost of making process changes or

Table 16–6

Methods to control and treat odorous gases found in wastewater management systems[a]

Method	Description and/or application
Physical methods	
Adsorption on activated carbon	Odorous gases can be passed through beds of activated carbon to remove odors. Carbon regeneration can be used to reduce costs. Additional details may be found in Chap. 11.
Adsorption on sand, soil, or compost beds	Odorous gases can be passed through sand, soil, or compost beds. Odorous gases from pumping stations may be vented to the surrounding soils or to specially designed beds containing sand or soils. Odorous gases collected from treatment units may be passed through compost beds.
Combustion	Gaseous odors can be eliminated by combustion at temperatures varying from 650 to 815°C (1200 to 1500°F). Gases can be combusted in conjunction with treatment plant solids or separately in a fume incinerator.
Containment	Installation of covers, collection hoods, and air handling equipment for containing and directing odorous gases to disposal or treatment systems.
Dilution with odor-free air	Gases can be mixed with fresh air sources to reduce the odor unit values. Alternatively, gases can be discharged through tall stacks to achieve atmospheric dilution and dispersion.
Masking agents	Perfume scents can be sprayed in fine mists near offending process units to overpower or mask objectionable odors. In some cases, the odor of the masking agent is worse than the original odor. Effectiveness of masking agents is limited.
Oxygen injection	The injection of oxygen (either air or pure oxygen) into the wastewater to control the development of anaerobic conditions has proven to be effective.
Scrubbing towers	Odorous gases can be passed through specially designed scrubbing towers to remove odors. Some type of chemical or biological agent is usually used in conjunction with the tower.
Thermal oxidation	Combustion of off-gases at temperatures from 800 to 1400°C will eliminate odors. Lower temperatures (400 to 800°C) are used with catalytic incineration.
Turbulence inducing facilities	Use of wind breaks, such as high fences and trees, and propeller fans.
Chemical oxidation	Oxidizing the odor compounds in wastewater is one of the most common methods used to achieve odor control. Chlorine, ozone, hydrogen peroxide, and potassium permanganate are among the oxidants that have been used. Chlorine also limits the development of a slime layer.
Chemical precipitation	Chemical precipitation refers to the precipitation of sulfide with metallic salts, especially iron.
Scrubbing with various alkalies	Odorous gases can be passed through specially designed scrubbing towers to remove odors. If the level of carbon dioxide is high, costs may be prohibitive.
Biological methods	
Activated sludge aeration tanks	Odorous gases can be combined with the process air for activated sludge aeration tanks to remove odorous compounds.
Biological conversion	Biological processes in the wastewater can reduce odors by converting malodorous constituents through oxidation.
Biological stripping towers	Specially designed towers can be used to strip odorous compounds. Typically, the towers are filled with plastic packing of various types on which biological growths can be maintained.
Compost filters	Gases can be passed through biologically active beds of compost to remove odors.
Sand and soil filters	Gases can be passed through biologically active beds of compost to remove odors.
Trickling filters	Odorous gases can be passed through existing trickling filters to remove odorous compounds.

[a] Adapted in part from U.S. EPA (1985).

Table 16–7

Suggested minimum buffer distances from treatment units for odor containment[a,b]

Treatment process unit	Buffer distance	
	ft	m
Sedimentation tank	400	125
Trickling filter	400	125
Aeration tank	500	150
Aerated lagoon	1000	300
Sludge digester (aerobic or anaerobic)	500	150
Sludge handling units	1000	300
Open drying beds	500	150
Covered drying beds	400	125
Sludge holding tank	1000	300
Sludge thickening tank	1000	300
Vacuum filter	500	150
Wet air oxidation	1500	450
Effluent recharge bed	800	250
Secondary effluent filters		
Open	500	150
Enclosed	200	75
Advanced wastewater treatment		
Tertiary effluent filters		
Open	300	100
Enclosed	200	75
Denitrification	300	100
Polishing lagoon	500	150
Land disposal	500	150

[a] Source: New York State Department of Environmental Conservation.

[b] Actual buffer distance requirements will depend on local conditions.

modifications to the facilities to eliminate odor development should always be evaluated and compared to the cost of various alternative odor control measures before their adoption is suggested.

Chemical Scrubbers. The basic design objective of a chemical scrubber is to provide contact between air, water, and chemicals (if used) to provide oxidation or entrainment of the odorous compounds. The principal wet scrubber types, as shown on Fig. 16–6, include single-stage countercurrent packed towers, countercurrent spray chamber absorbers, and cross-flow scrubbers. In most single-stage scrubbers, such as shown on Fig. 16–7, the scrubbing fluid (usually sodium hypochlorite) is recirculated. The commonly used oxidizing scrubbing liquids are sodium hypochlorite, potassium permanganate, and hydrogen peroxide solutions. Because of safety and handling issues, chlorine gas is not used commonly in scrubbing applications at wastewater treatment facilities. Sodium hydroxide is also used in scrubbers where H_2S concentrations in the gas phase are high.

Figure 16–6

Typical wet scrubbers systems for odor control: (a) countercurrent packed tower, (b) spray chamber absorber, and (c) cross-flow scrubber.

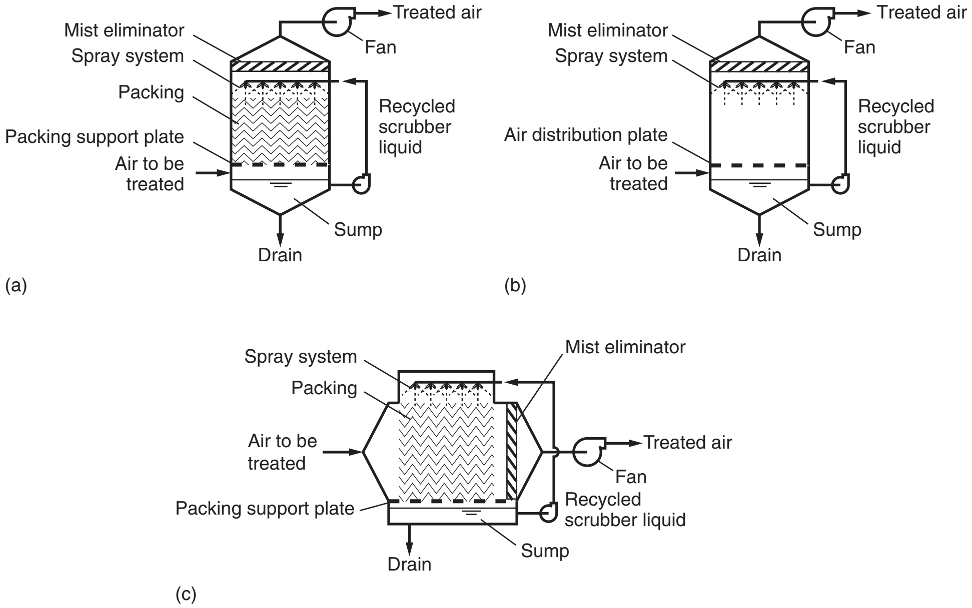


Figure 16–7

Typical sodium hypochlorite scrubber used to treat the odors from the trickling filters shown on Fig. 16–4(c).

Chemical Scrubbing Reactions for Hydrogen Sulfide. Typical simplified scrubbing reactions for H_2S with chlorine, sodium hypochlorite, potassium permanganate, and hydrogen peroxide are as follows:

With chlorine

$$\underset{(34.06)}{\mathrm{H_2S}} + \underset{(4\times 70.91)}{\mathrm{4Cl_2}} + \mathrm{4H_2O} \rightarrow \mathrm{H_2SO_4} + \mathrm{8HCl} \tag{16–1}$$

$$\underset{(34.06)}{\mathrm{H_2S}} + \underset{(70.91)}{\mathrm{Cl_2}} \rightarrow \mathrm{S^\circ}\downarrow + \mathrm{2HCl} \tag{16–2}$$

With sodium hypochlorite

$$\underset{(34.06)}{\mathrm{H_2S}} + \underset{(4\times 74.45)}{\mathrm{4NaOCl}} + \mathrm{2NaOH} \rightarrow \mathrm{Na_2SO_4} + \mathrm{2H_2O} + \mathrm{4NaCl} \tag{16–3}$$

$$\underset{(34.06)}{\mathrm{H_2S}} + \underset{(74.45)}{\mathrm{NaOCl}} \rightarrow \mathrm{S^\circ}\downarrow + \mathrm{NaCl} + \mathrm{H_2O} \tag{16–4}$$

With potassium permanganate

$$\underset{(3\times 34.06)}{\mathrm{3H_2S}} + \underset{(2\times 142.04)}{\mathrm{2KMnO_4}} \rightarrow \mathrm{3S} + \mathrm{2KOH} + \mathrm{2MnO_2} + \mathrm{2H_2O}\ \text{(acidic pH)} \tag{16–5}$$

$$\underset{(3\times 34.06)}{\mathrm{3H_2S}} + \underset{(8\times 142.04)}{\mathrm{8KMnO_4}} \rightarrow \mathrm{3K_2SO_4} + \mathrm{2KOH} + \mathrm{8MnO_2} + \mathrm{2H_2O}\ \text{(basic pH)} \tag{16–6}$$

With hydrogen peroxide

$$\underset{(34.06)}{\mathrm{H_2S}} + \underset{(34.0)}{\mathrm{H_2O_2}} \rightarrow \mathrm{S^\circ}\downarrow + \mathrm{2H_2O}\ (\mathrm{pH} < 8.5) \tag{16–7}$$

In the reaction given by Eq. (16–3), 8.74 mg/L of sodium hypochlorite is required per mg/L of hydrogen sulfide, or 9.29 mg/L if the hydrogen sulfide is expressed as sulfide. In addition, in the reaction given by Eq. (16–3), 2.35 mg/L of sodium hydroxide (caustic) will be required per mg/L of hydrogen sulfide to make up for alkalinity consumed in the reaction. In practice, the required sodium hypochlorite dosage for the reaction given by Eq. (16–3) will vary from 8 to 10 mg/L per mg/L of H_2S. For the reaction given by Eq. (16–4), 2.19 mg/L of sodium hypochlorite is required per mg/L of hydrogen sulfide.

When potassium permanganate is used, the reactions that occur are typically various combinations of the reactions given by Eqs. (16–5) and (16–6). Reaction products that can occur, depending on the local wastewater chemistry, include elemental sulfur, sulfate, thionates, dithionates, and manganese sulfide. Stoichiometrically, about 2.8 and 11.1 mg/L of $KMnO_4$ are required for each mg/L of H_2S oxidized as given by Eqs. (16–5) and (16–6), respectively. However, based on operating data from actual field installations, about 6 to 7 mg/L $KMnO_4$ are required for each mg/L of H_2S oxidized. Potassium permanganate is generally used in smaller installations because of the cost (U.S. EPA, 1985; WEF, 1995).

In the reaction given in Eq. (16–7), 1.0 mg/L of hydrogen peroxide is required for each mg/L of sulfide expressed as hydrogen sulfide. In practice, the required dosage can vary from 1 to 4 mg/L per mg/L of H_2S. Because the systems used to carry out the reactions defined by Eqs. (16–1) through (16–7) are complex, especially where competing reactions may occur, the proper dosage should be established by site-specific testing.

Hypochlorite scrubbers can be expected to remove oxidizable odorous gases when other gas concentrations are minimal. Typical removal efficiencies for single-stage scrubbers are reported in Table 16–8. In cases where the concentrations of odorous components in the exhaust gas from the scrubbers are above desirable levels, multistage scrubbers (see Fig. 16–8) are often used. In the three-stage scrubber shown on Fig. 16–8, the first stage is a pretreatment stage used to raise the pH so that a portion of the odorous gases (e.g., hydrogen sulfide) is reduced before treatment with chlorine in the second and third stages. The reaction that occurs in the first stage of a three-stage unit can be represented as follows.

$$H_2S + 2NaOH \rightarrow Na_2S + 2H_2O \qquad (16\text{–}8)$$

To reduce maintenance problems due to precipitation, it is recommended that a low hardness (less than 50 mg/L as $CaCO_3$) be used for the makeup water.

Chemical Scrubbing of Ammonia and/or Ammine Compounds. In situations where there is a potential for the release of high concentrations of ammonia or ammine compounds, such as from solids handling and lime stabilization processes, an additional tower may be needed. The removal of ammonia or ammine compounds is accomplished with an

Table 16–8

Effectiveness of hypochlorite wet scrubbers for removal of several odorous gases[a]

Gas	Expected removal efficiency, %	
	Range	Typical
Hydrogen sulfide	90–99	98
Ammonia	90–99	98
Sulfur dioxide	90–96	95
Mercaptans	85–92	90
Other oxidizable compounds	70–90	85

[a] Adapted in part from U.S. EPA (1985).

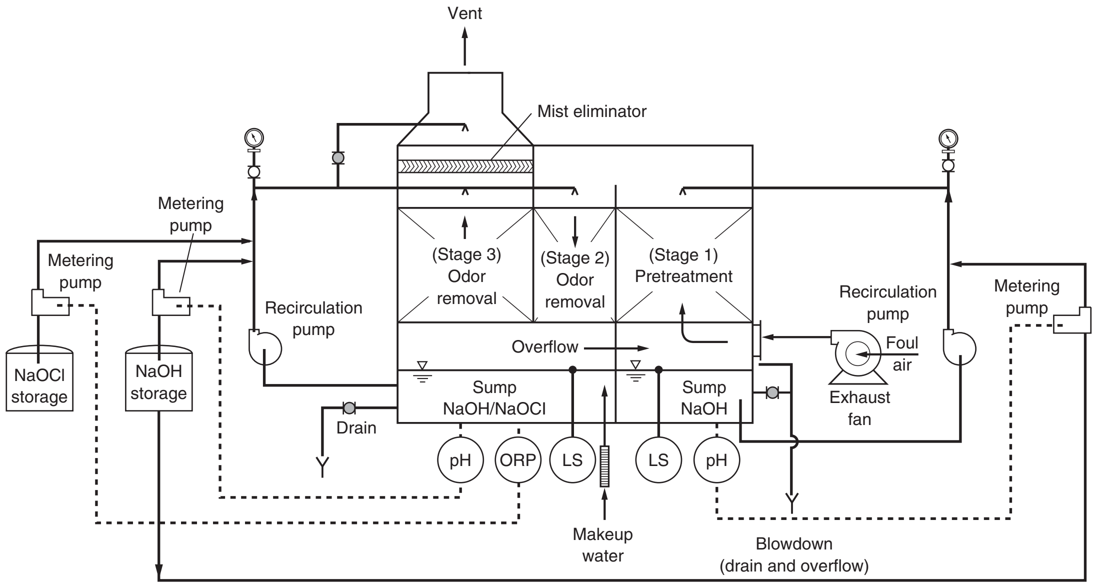


Figure 16–8

Three-stage odor control process flow diagram. (From Lo/Pro Systems, Inc.)

acid/base reaction with sulfuric acid. Typically, the pH of recirculating sulfuric acid solutions is around 4.0 to 6.0 and the following reaction occurs:

$$2NH_3 + H_2SO_4 \rightarrow (NH_4)_2SO_4 \tag{16–9}$$

Some of the spent scrubbing liquid that contains the solid ammonium sulfate is discharged from the scrubber. The waste spent scrubbing fluid is typically returned to the treatment plant headworks. The remaining scrubbing liquid is recirculated back into the tower with additional sulfuric acid added to maintain the proper pH of the scrubbing liquid.

Activated Carbon Adsorbers. Activated carbon adsorbers are used commonly for odor control (see Fig. 16–9). The rate of adsorption for different constituents or compounds will depend on the nature of the constituents or compounds being adsorbed (nonpolar versus polar). It has also been found that the removal of odors depends on the

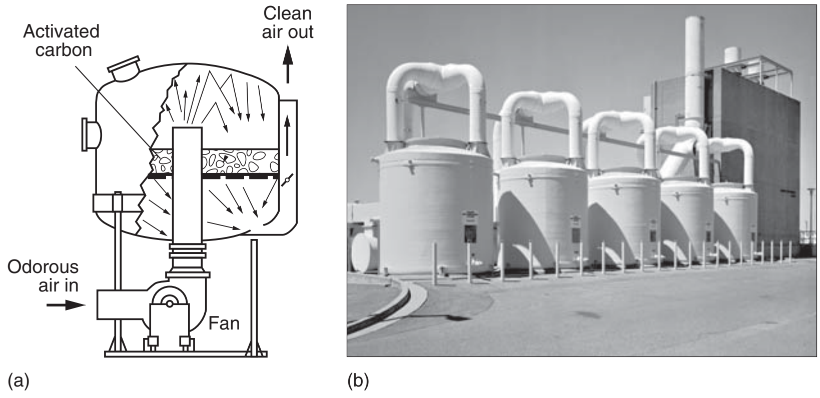


Figure 16–9

Use of activated carbon for odor control: (a) schematic of typical downflow activated carbon reactor and (b) view of multiple activated carbon odor control reactors.

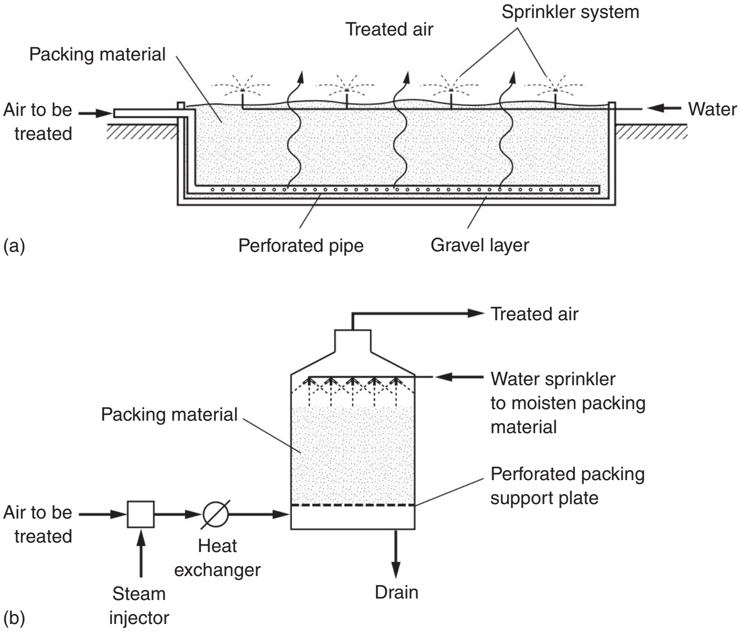
concentration of the hydrocarbons in the odorous gas. Typically, hydrocarbons are adsorbed preferentially before polar compounds such as H_2S are removed (note activated carbon is nonpolar). Thus, the composition of the odorous gases to be treated must be known if activated carbon is to be used effectively. Because the life of a carbon bed is limited, carbon must be regenerated (see Sec. 11–7 in Chap. 11) or replaced regularly for continued odor removal. To prolong the life of the carbon, two-stage systems have been used, with the first stage being a wet scrubber followed by activated-carbon adsorption.

Vapor-Phase Biological Treatment Processes. The two principal biological processes used for the treatment of odorous gases present in the vapor phase are (1) biofilters and (2) biotrickling filters (Eweis et al., 1998). The use of microbial growths for the treatment of odors was the subject of an early patent by Pomeroy (1957), one of the important early researchers in the area of odor management in wastewater collection and treatment facilities.

Biofilters. Biofilters are packed-bed filters. In open biofilters [see Fig. 16–10(a)], the gases to be treated move upward through the filter bed. In closed biofilters [see Fig. 16–10(b)], the gases to be treated are either blown or drawn through the packing material. As the odorous gases move through the packing in the biofilter, two processes occur simultaneously: sorption (i.e., absorption/adsorption) and bioconversion. Odorous gases are absorbed into the moist surface biofilm layer and the surfaces of the biofilter packing material. Microorganisms, principally bacteria, actinomycetes, and fungi, attached to the packing material, oxidize the absorbed/adsorbed gases and renew the treatment capacity of the packing material. Moisture content and temperature are important environmental conditions that must be maintained to optimize microorganism activity (Williams and Miller, 1992a, 1992b; Yang and Allen, 1994; Eweis et al., 1998). Although compost biofilters are used commonly, one drawback is the large surface area (footprint) required for these units.

Figure 16–10

Typical packed bed biofilters: (a) open bed type and (b) enclosed reactor type.



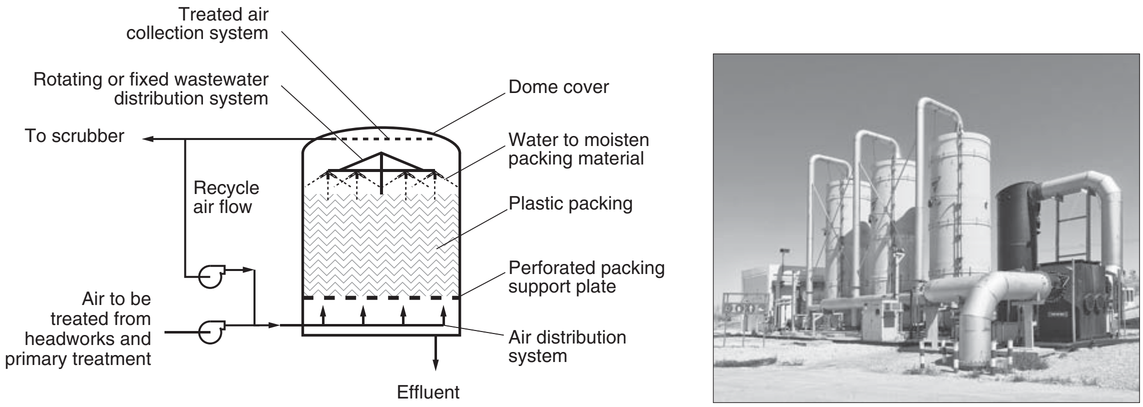


Figure 16–11

Typical covered biological stripping tower used for odor control (a) schematic and (b) view of multiple biological stripping towers.

Biotrickling Filters. Biotrickling filters are essentially the same as biofilters with the exception that moisture is provided continuously or intermittently by applying (typically spraying) a liquid (e.g., treated effluent) over the packing (see Fig. 16–11). The liquid is recirculated and nutrients are often added. Because water is lost in the gas leaving the filter, makeup water must be provided. Similarly, because of the accumulation of salts in the recycled water, a blowdown stream is required. Compost is not a suitable packing material for biotrickling filters because water will accumulate within the compost, thereby limiting the free movement of air within the filter. Typical packing materials include Pall rings, Raschig rings, lava rock, and granular activated carbon (Eweis et al., 1998; see also Sec. 11–8, Gas Stripping, in Chap. 11).

Conventional Biological Treatment Processes. The ability of microorganisms to oxidize hydrogen sulfide and other similar odorous compounds dissolved in the liquid under aerobic conditions is the basic concept used for the treatment of odors in liquid-based systems. The two principal types of conventional liquid-based systems used in wastewater treatment plants are the activated sludge process and the trickling filter process. In the activated sludge process, the odorous compounds are introduced into the aeration basin either with the existing air supply or injected separately through a manifold system. A major concern with this method of odor management is the high rate of corrosion in the air piping and blowers that occurs due to the presence of moist air containing hydrogen sulfide. The ability to transfer the odorous gaseous compounds to the liquid phase is also of concern.

With conventional uncovered trickling filters the major issues are how to transfer the air containing the odorous compounds to the trickling filter and how to avoid the release of untreated odorous compounds to the atmosphere. To control the release of odorous compounds, existing trickling filters that are to be used for odor control are almost always covered [see Fig. 16–4(c)].

Thermal Processing. Three thermal processing techniques have been used: (1) thermal oxidation, (2) catalytic oxidation, and (3) recuperative and regenerative thermal oxidation.

The Oxidation Process. The oxidation of CH_4 (methane) and H_2S can be used to illustrate the basic principle of all three thermal processes.

$$CH_4(\text{gas}) + 2O_2(\text{gas}) \rightarrow CO_2(\text{gas}) + 2H_2O(\text{vapor}) + \text{heat} \tag{16–10}$$

$$H_2S(\text{gas}) + 2O_2(\text{gas}) \rightarrow H_2SO_4(\text{vapor}) \tag{16–11}$$

If the gas to be combusted does not liberate enough heat to sustain the combustion process, it is usually necessary to use an external fuel source such as fuel oil, natural gas, or propane. Unfortunately, because of the low concentrations of odorous combustible gases in most waste streams, sustainable thermal oxidation is seldom possible, and large amounts of fuel are typically required to maintain the combustion temperatures needed to eliminate odors.

Thermal Oxidation. Thermal oxidation is used, more commonly, for concentrated waste streams. The flaring of odorous gases is a relatively crude form of thermal combustion [see Fig. 16–12(a)]. Depending on the design of the combustion facility, incomplete combustion can occur due to variations in gas flow. For this method of odor control to be sustainable, the waste gas must typically contain 50 percent of the fuel value of the gas stream to be combusted.

Catalytic oxidation. A flameless oxidation that occurs in the range from 310 to 425°C (600 to 800°F) in the presence of a catalyst is defined as catalytic oxidation [see Fig. 16–12(b)]. Common catalysts include platinum, palladium, and rubidium. The decrease in temperature as compared to complete thermal oxidation reduces the energy requirements significantly. However, because the catalysts can become fouled, the gas to be

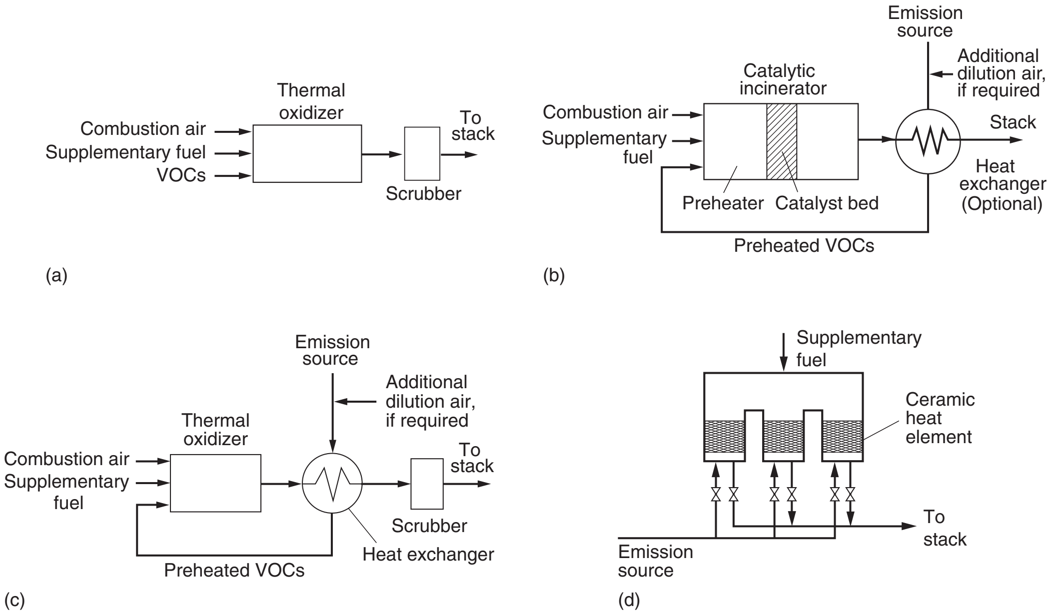


Figure 16–12

Schematic diagrams of thermal processes for treatment of VOCs: (a) thermal oxidation, (b) catalytical oxidizer, (c) recuperative thermal oxidizer, and (d) regenerating thermal oxidizer.

oxidized must not contain particulate material or constituents that will result in a residue. Additional information on the physical facilities used for thermal processing of VOCs may be found in Sec. 16–4.

Recuperative and Regenerative Thermal Oxidation. This process involves preheating the odorous gases before passing them into the combustion chamber so that complete oxidation can be achieved. Combustion occurs at temperatures in the range from 425 to 760°C (800 to 1400°F). Recuperative and regenerative thermal oxidation processes are used to reduce fuel consumption by preheating the incoming air, especially in large installations. In recuperative oxidizers, thin wall tubes are used to transfer heat recovered from exhaust air to the incoming air [see Fig. 16–12(c)]. In regenerative oxidizers, ceramic packing material is used to capture the heat from the hot exhaust gas and subsequently to release it to the incoming air. To maintain optimal heat recovery, the exhaust and incoming air are cycled through the packing material so that the incoming air is always passed through the hottest packing material. Typically, three stages of packing material are used in regenerative oxidizers [see Fig. 16–12(d)].

Selection and Design of Odor Control Facilities

The following steps are involved in the selection and design of odor control and treatment facilities:

1. Determine the characteristics and volumes of the gas to be treated.
2. Define the exhaust requirements for the treated gas.
3. Evaluate climatic and atmospheric conditions.
4. Select one or more odor control and treatment technologies to be evaluated.
5. Conduct pilot tests to determine design criteria and performance.
6. Perform life cycle economic analysis.

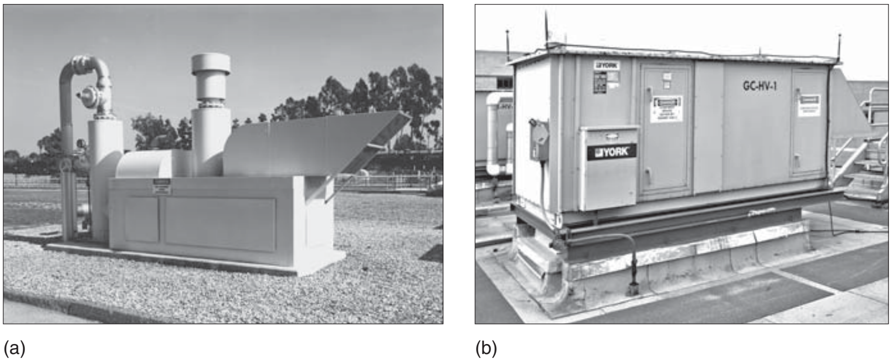
Many of the chemical odor control technologies are supplied as complete packages, designed to meet a given performance specification. The analysis of chemical scrubbers and the design of biofilters is considered in the following discussion.

Design Considerations for Chemical Scrubbers

Most chemical scrubbers are supplied as a complete unit (see Fig. 16–13). Typical design factors for chemical scrubbers are presented in Table 16–9. Determination of the chemical requirements for odor scrubbing is illustrated in Example 16–1.

Figure 16–13

Typical self contained chemical odor stripping unit.



(a) (b)

Table 16–9

Typical design factors for chemical scrubbers[a]

Item	Unit	Value
Packing depth	m	1.8–3
Gas residence time in packing	s	1.3–2.0
Scrubbant flowrate	kg H_2O/kg air flow	1.5–2.5
	L/s per m^3/s air flow	2–3
Makeup water flow	L/s per kg sulfide at pH 11	0.075
	L/s per kg sulfide at pH 12.5	0.004
pH	unitless	11–12.5
Temperature	°C	15–40
Caustic usage	kg NaOH/kg sulfide	2–3

[a] Adapted in part from from WEF (1995), Devinny et al. (1999).

EXAMPLE 16–1 **Chemical Requirements for Odor Scrubbing** Hydrogen sulfide is to be scrubbed from a waste airstream using sodium hypochlorite. Determine the chemical (i.e., sodium hypochlorite and caustic) and water requirements for the following conditions.

1. Waste airstream flowrate = 1000 m^3/min
2. H_2S concentration in waste stream = 20 ppm_v at 20°C
3. Specific weight of air = 0.0118 kN/m^3 at 20°C
4. Density of air = 1.204 kg/m^3 at 20°C (see Sec. B-3 in Appendix B)
5. Assume liquid to gas ratio for scrubber = 1.75
6. Density of 50 percent NaOH solution = 1.52 kg/L

Solution

1. Determine the volume occupied by one mole of a gas at a temperature of 20°C and a pressure of 1.0 atm using Eq. (2–44).

$$V = \frac{nRT}{P}$$

$$V = \frac{(1\text{ mole})(0.082057\text{ atm}\cdot\text{L/mole}\cdot\text{K})[(273.15 + 20)\text{K}]}{1.0\text{ atm}}$$

$$= 24.055\text{L} \qquad \text{use } 24.1\text{L}$$

2. Estimate the sodium hypochlorite requirement.
   a. Using Eq. (2–45), convert the H_2S concentration from ppm_v to g/m^3.

$$20\text{ ppm}_v = \left(\frac{20\text{ m}^3}{10^6\text{ m}^3}\right)\left[\frac{(34.8\text{ g/mole H}_2\text{S})}{(24.1 \times 10^{-3}\text{ m}^3\text{/mole of H}_2\text{S})}\right]$$

$$\text{H}_2\text{S concentration} = 28.3 \times 10^{-3}\text{ g/m}^3$$

   b. Determine the amount of H_2S that must be treated per day.

$$(1000\text{ m}^3\text{/min}) \times (28.3 \times 10^{-3}\text{ g/m}^3)(1440\text{ min/d})(1\text{ kg}/10^3\text{ g}) = 40.8\text{ kg/d}$$

c. Estimate the sodium hypochlorite dose. From Eq. (16–3), 8.74 mg/L of sodium hypochlorite are required per mg/L of sulfide, expressed as hydrogen sulfide.

$NaOCl_2$ required per day = (40.8 kg/d)(8.74) = 356.6 kg/d

3. Estimate the water requirement for the scrubbing tower.
   a. Determine the mass air flowrate.

   $(1000 \text{ m}^3/\text{min})(1.204 \text{ kg/m}^3) = 1204 \text{ kg/min}$

   b. Determine the water flowrate.

   $(1204 \text{ kg/min})(1.75) = 2107 \text{ kg/min} = 2.1 \text{ m}^3/\text{min}$

4. Determine the amount of sodium hydroxide (caustic) that must be added to replace the alkalinity consumed in the reaction.
   a. From the reaction given by Eq. (16–3), 2.35 mg/L of NaOH is required for each mg/L of H_2S removed.
   b. Determine the amount of NaOH required.

   NaOH = (40.8 kg/d)(2.35) = 95.9 kg/d

   c. Determine the volume of NaOH required. The amount of caustic per liter at 50% NaOH is

   NaOH = (1.52 kg/L)(0.50) = 0.76 kg/L

$$\text{Volume of NaOH} = \frac{(95.9 \text{ kg/d})}{(0.76 \text{ kg/L})} = 126.2 \text{ L/d}$$

Comment The water requirement for the scrubbing tower will be specified initially by the scrubber supplier and field adjusted based on the results of pilot-plant studies and past operating experience. If the wastewater has sufficient alkalinity, it may not be necessary to add sodium hydroxide.

Design Considerations for Odor Control Biofilters

Important design considerations for biofilters include (1) the type and composition of the packing material, (2) facilities for gas distribution, (3) maintenance of moisture within the biofilter, and (4) temperature control. Each of these topics is considered below. Design and operational parameters are presented and discussed following the discussion of the above topics. Additional details on biofilters may be found in van Lith (1989), Allen and Yang (1991, 1992), WEF (1995), Eweis et al., (1998), Devinny et al. (1999), and WEF (2004).

Packing Material. The requirements for the packing used in biofilters are (1) sufficient porosity and near-uniform particle size, (2) particles with large surface areas and significant pH-buffering capacities, and (3) the ability to support a large population of microflora (WEF, 1995). Packing materials used in biofilters include compost, peat, and a variety of synthetic mediums. Although soil and sand have been used in the past, they are less used today because of excessive headloss and clogging problems (Bohn and Bohn, 1988). Bulking materials used to maintain the porosity of compost and peat biofilters include perlite, Styrofoam™ pellets, wood chips, bark, and a variety of ceramic and plastic materials. A typical recipe for a compost biofilter is as follows (Schroeder, 2001):

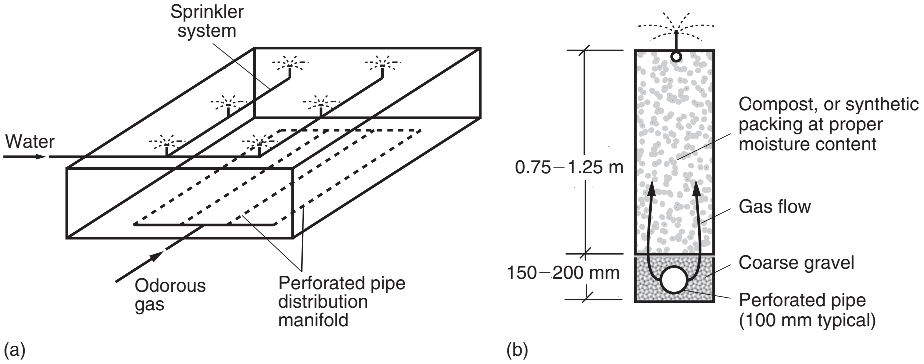
Compost = 50 percent by volume

Bulking agent = 50 percent by volume

1 meq $CaCO_3$/g of packing material by weight

Figure 16–14

Definition sketch for packed bed biofilters: (a) open bed and (b) trench type.



Optimal physical characteristics of a packing material include a pH between 7 and 8, air-filled pore space between 40 and 80 percent, and organic matter content of 35 to 55 percent (Williams and Miller, 1992a, 1992b). When compost is used, additional compost must be added periodically to account for the loss due to biological conversion. Bed depths of up to 1.8 m (6.0 ft) have been used. However, because most of the removal takes place in the first 20 percent of the bed, the use of deeper beds is not recommended.

Gas Distribution. An important design feature of a biofilter is the method used to introduce the gas to be treated. The most commonly used gas distribution systems include (1) perforated pipes, (2) prefabricated underdrain systems, and (3) plenums. Perforated pipes are usually placed in a gravel layer below the compost (see Fig. 16–14). Where perforated pipes are used, it is important to size the pipe so that it performs as a reservoir and not a manifold to assure uniform distribution (Crites and Tchobanoglous, 1998). A variety of prefabricated underdrain systems are available that allow for the movement of gas upward through the compost bed and allow for the collection of drainage. Air plenums are used to equalize the air pressure to achieve uniform flow upward through the compost bed. The height of air plenums will typically vary from 200 to 500 mm.

Moisture Control. Perhaps the most critical item in the successful operation of a biofilter is to maintain the proper moisture within the filter bed. If the moisture content is too low, biological activity will be reduced. If the moisture content is too high, the flow of air will be restricted and anaerobic conditions may develop within the bed. Also, biofilters tend to dry out unless moisture or humidity is added. The optimal moisture content is between about 50 and 65 percent, defined as follows:

$$\text{Moisture content, \%} = \left(\frac{\text{mass of water}}{\text{mass of water} + \text{mass of dry packing}}\right) \times 100 \qquad (16\text{–}12)$$

Moisture can be supplied by adding water to the top of the bed (usually by spraying) or by humidifying the incoming gas in a humidification chamber. The relative humidity of the gas entering the biofilter should be 100 percent at the operating temperature of the biofilter (Eweis et al., 1998). The liquid application rate for biotrickling filters is typically about 0.75 to 1.25 $m^3/m^2 \cdot d$.

Temperature Control. The operating temperature range for biofilters is between 15 and 45°C, with the optimal range being between 25 and 35°C. In cold climates, biofilters must be insulated, and the incoming gas must be heated. Where the incoming gas is warmer, it may have to be cooled before being introduced to the biofilter. Operation at higher temperatures (e.g., 45 to 60°C) is often possible, as long as the temperature remains relatively constant.

Design and Operating Parameters for Biofilters. The sizing of biofilters is typically based on a consideration of the gas residence time in the bed, the unit air loading rate, and the constituent elimination capacity. Terms that will be encountered in the literature and the relationships commonly used to describe the performance of bulk media filters are summarized in Table 16–10. The empty bed residence time (EBRT) [see Eq. (16–13)], used to define the relationship between the volume of the contactor and the volumetric gas flowrate, is similar to Eq. (11–62) used for the analysis of activated-carbon systems. The true residence time is determined by incorporating the porosity, a [see Eq. (16–14)]. Surface and volumetric mass loading rates are often used to define the operation of bulk media filters. The elimination capacity, as given by Eq. (16–20), is used to compare the performance of different odor control systems.

The residence time for foul air from wastewater treatment facilities is typically between 15 and 60 s, and surface loading rates have ranged up to 120 $m^3/m^2 \cdot min$ for H_2S concentrations up to 20 mg/L. Constituent elimination rates are determined experimentally and are usually reported as a function of the constituent loading rate (e.g., mg $H_2S/m^3 \cdot h$). An essentially linear, 1 to 1, constituent elimination rate up to a critical loading rate has been observed for hydrogen sulfide and other odorous compounds

Table 16–10

Parameters used for the design and analysis of bulk media filters[a]

Parameter	Definition
Empty bed residence time $\text{EBRT} = \frac{V_f}{Q}$ (16–13) Actual residence time in filter $\text{RT} = \frac{V_f \times \alpha}{Q}$ (16–14) Surface loading rate $\text{SLR} = \frac{Q}{A_f}$ (16–15) Surface mass loading rate $\text{SLR}_m = \frac{Q \times C_o}{A_f}$ (16–16) Volume loading rate $\text{VLR} = \frac{Q}{V_f}$ (16–17) Volume mass loading rate $\text{VLR}_m = \frac{Q \times C_o}{V_f}$ (16–18) Removal efficiency $\text{RE} = \frac{C_o - C_e}{C_o} \times 100$ (16–19) Elimination capacity $\text{EC} = \frac{Q(C_o - C_e)}{V_f}$ (16–20)	EBRT = empty bed residence time, h V_f = total volume of filter bed contactor, m^3 Q = volumetric flowrate, m^3/h RT = residence time, h, min, s a = porosity of filter bed contactor SLR = surface loading rate, $m^3/m^2 \cdot h$ A_f = surface area of filter bed contactor, m^2 VLR = volumetric loading rate, $m^3/m^3 \cdot h$ RE = removal efficiency, % C_o = influent gas concentration, mg/L EC = elimination capacity, $g/m^3 \cdot h$ C_e = effluent gas concentration, mg/L

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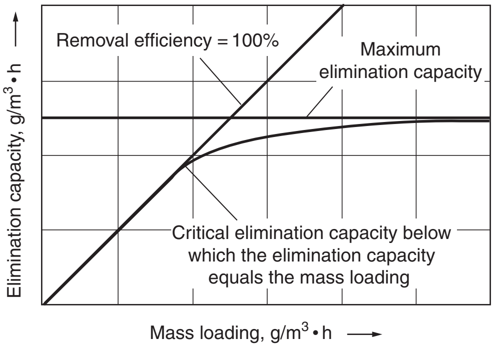
[a] Adapted in part from from Eweis et al. (1998); Devinny et al. (1999).

Figure 16–15

Typical odor elimination capacity versus applied load.



(see Fig. 16–15). Yang and Allen (1994) have reported a linear 1 to 1 elimination rate for H_2S, with loading rates for compost filters up to a maximum value of about 130 g $S/m^3 \cdot h$, beyond which the elimination rate becomes essentially constant at a rate of 130 g $S/m^3 \cdot h$ with increased loading. It should be noted that H_2S is eliminated easily as it passes through a biofilter.

Typical design criteria for biofilters are presented in Table 16–11. Typical biofilters are shown on Fig. 16–16. Some states regulate the design of compost biofilters including

Table 16–11

Typical design factors for biofilters[a]

		Type of biofilter	
Item	**Units**	**Biofilter**	**Biotrickling filter**
Oxygen concentration	Parts oxygen/ parts oxidizable gas	100	100
Moisture			
Compost filter	%	50–65	50–65
Synthetic media	%	55–65	55–65
Temperature, optimum	°C	15–35	15–35
pH	unitless	6–8	6–8
Porosity	%	35–50	35–50
Gas residence time	s	30–60	30–60
Depth of medium	m	1–1.25	1–1.25
Inlet odorous gas concentration	g/m^3	0.01–0.5	0.01–0.5
Surface loading rate	$m^3/m^2 \cdot h$	10–100[b]	10–100[b]
Volume loading rate	$m^3/m^3 \cdot h$	10–100	10–100
Liquid application rate	$m^3/m^2 \cdot d$		0.75 to 1.25
Elimination capacity			
H_2S (in compost filter)	$mg/m^3 \cdot h$	80–130	80–130
Other odorous gases	$mg/m^3 \cdot h$	20–100	20–100
Back pressure, maximum	mm of water	50–100	50–100

[a] Adapted in part from van Lith (1989), Yang and Allen (1994), WEF (1995), and Devinny et al. (1999).

[b] Loading rates as high as 500 $m^3/m^2 \cdot h$ have been reported, depending on the compound and its concentration.

Figure 16–16

Typical bulk biofilters for odor control: (a) compost biofilter (b) gravel-type biofilter.

(a) (b)

loading rates, biofilter emission rates, odor-sampling procedures, and setbacks from property lines. A typical odor-emission limit at the surface of the biofilter is 50 dilutions to threshold [see Eq. (2–52) in Chap. 2)] (Finn and Spencer, 1997). The design of a compost biofilter for the elimination of hydrogen sulfide is illustrated in Example 16–2.

EXAMPLE 16–2 **Design of a Compost Biofilter for Odor Control** Determine the size of compost biofilter needed to scrub the air from a 100 m^3 enclosed volume using the design criteria given in Table 16–11. Also estimate the mass of the buffer compound needed to neutralize the acid formed as a result of treatment within the filter. Assume 12 air changes per hour are needed. Assume a bed porosity of 40 percent. Will the volume selected be adequate if the air contains 40 ppm_v of H_2S in addition to other odorous constituents? Assume an elimination rate of 65 g $S/m^3{\cdot}h$, which incorporates a factor of safety of 2 as compared to the maximum rate given in Table 16–11. The temperature of the air is 20°C.

Solution

1. Estimate the airflow to be scrubbed.
   Flow = volume/time
   Flow = $(100\ m^3)(12\ \text{changes per h}) = 1200\ m^3/h$
2. Select a loading rate from Table 16–11; use 90 $m^3/m^2{\cdot}h$.
3. Select a filter-bed depth from Table 16–11; use 1.0 m.
4. Calculate the area needed for the filter bed.
   Area = gas flow/loading rate
   Area = $(11200\ m^3/h)/(90\ m^3/m^2{\cdot}h)$
   Area = $13.3\ m^2$
5. Check the empty bed residence time using Eq. (16–13).

$$\text{EBRT} = \frac{V_f}{Q} = \frac{(13.3\ m^2)(1\ m)}{(1200\ m^3/h)}$$
$$= 0.011\ h = 39.9\ s\ (\text{OK } 39.9\ s > 30\ s)$$

6. Determine if the volume of the biofilter determined in Step 5 is adequate to treat the H_2S.

a. Determine the concentration of H_2S in g/m^3 using Eq. (2–45). From Example 16–1 the volume of gas occupied by one mole of a gas at a temperature of 20°C and a pressure of 1.0 atm is 24.1 L. Thus, the concentration of H_2S is

$$\text{g/m}^3 = \left(\frac{40\ \text{L}^3}{10^6\ \text{L}^3}\right)\left[\frac{(34.08\ \text{g/mole H}_2\text{S})}{(24.1 \times 10^{-3}\ \text{m}^3\text{/mole of H}_2\text{S})}\right]$$

$$= 0.057\ \text{g/m}^3$$

b. Determine the mass loading rate of S^{2-} in g S/h.

$$M_s = \left(\frac{1200\ \text{m}^3}{\text{h}}\right)\left(\frac{0.057\ \text{g H}_2\text{S}}{\text{m}^3}\right)\left(\frac{32\ \text{g S}^{2-}}{34.08\ \text{g H}_2\text{S}}\right)$$

$$= 64.2\ \text{g S}^{2-}\text{/h}$$

c. Determine the required volume assuming an elimination rate of 65 g S/m^3·h.

$$V = \frac{(64.2\ \text{g S/h})}{(65\ \text{g S/m}^3\cdot\text{h})} = 0.99\ \text{m}^3$$

Because the volume of the bed (13.3 m^3) is significantly greater than the required volume, the removal of H_2S will not be an issue.

7. Determine the mass of the buffer compound needed to neutralize the acid formed as a result of treatment within the filter.

a. Determine the mass of H_2S in kg applied per year.

$$\text{H}_2\text{S, kg/year} = \frac{(1200\ \text{m}^3\text{/h})(0.057\ \text{g/m}^3)(24\ \text{h/d})(365\ \text{d/y})}{(10^3\ \text{g/1 kg})}$$

$$= 599.2\ \text{kg/y}$$

b. Determine the mass of buffer compound required. Assume the following equation applies:

$$\underset{34.06}{\text{H}_2\text{S}} + \underset{74.08}{\text{Ca(OH)}_2} + 2\text{O}_2 \rightarrow \text{Ca}_2\text{SO}_4 + 2\text{H}_2\text{O}$$

Thus, about 2.18 kg of $Ca(OH)_2$ (74.08/34.06) will be required per kg of H_2S. If the compost biofilter has a useful life of 2 y, then a total of 2457 kg of $Ca(OH)_2$ equivalent will be required to be added to the bed. Typically, 1.25 to 1.5 times as much are added. The buffer compound is mixed in with the compost and the bulking agent.

Comment Based on the results of the computation carried out in Step 6, it is clear why compost and soil filters are so effective in the elimination of H_2S.

16–4 CONTROL OF VOLATILE ORGANIC CARBON EMISSIONS

Many of the organic priority pollutants of concern in wastewater treatment are, as noted in Chap. 2, also classified as volatile organic compounds (VOCs). At some wastewater-treatment facilities, volatile organic compounds (VOCs) such as trichloroethylene (TCE) and 1,2-dibromo-3-chloropropane (DBCP) have been detected in wastewater.

The uncontrolled release of such compounds that now occurs in wastewater collection systems and wastewater-treatment plants is of concern because (1) once such compounds are in the vapor state they are much more mobile and, therefore, more likely to be released to the environment, (2) the presence of some of these compounds in the atmosphere may pose a significant public health risk, and (3) they contribute to a general increase in reactive hydrocarbons in the atmosphere, which can lead to the formation of ground-level ozone. The physical properties of selected VOCs, the mechanisms governing the release of these compounds, the locations where the release of these compounds is most prevalent, and the methods of controlling the discharge of these compounds to the atmosphere are discussed in this section.

Physical Properties of Selected VOCs

The physical properties of selected VOCs are presented in Table 16–12. Organic compounds that have a boiling point less than or equal to 100°C and/or a vapor pressure greater than 1 mm Hg at 25°C are generally considered to be volatile organic compounds (VOCs). For example, chloroethene (vinyl chloride), which has a boiling point of −13.9°C and a vapor pressure of 2548 mm Hg at 20°C, is an example of an extremely volatile organic compound.

Emission of VOCs

The release of VOCs in collection systems and at treatment plants, especially at the headworks, is of particular concern with respect to the health of the collection system and treatment plant workers. The principal mechanisms governing the release of VOCs in wastewater collection and treatment facilities are (1) volatilization and (2) gas stripping. These mechanisms and the principal locations where VOCs are released are considered in the following discussion.

Volatilization. The release of VOCs from wastewater surfaces to the atmosphere is termed volatilization. Volatile organic compounds are released because they partition between the gas and water phase until equilibrium concentrations are reached (Roberts et al., 1984). The mass transfer (movement) of a constituent between these two phases is a function of the constituent concentration in each phase relative to the equilibrium concentration. Thus, the transfer of a constituent between phases is greatest when the concentration in one of the phases is far from equilibrium. Because the concentration of VOCs in the atmosphere is extremely low, the transfer of VOCs usually occurs from wastewater to the atmosphere. However, because of the dynamic nature of the flows found within a wastewater treatment plant equilibrium conditions rarely exist, VOCs are either volatilized and/or degraded biologically.

Gas Stripping. Gas stripping of VOCs occurs when a gas (usually air) is entrained temporarily in wastewater or is introduced purposefully to achieve a treatment objective. When gas is introduced into a wastewater, VOCs are transferred from the wastewater to the gas. The forces governing the transfer between phases are the same as described above. For this reason, gas (air) stripping is most effective when contaminated wastewater is exposed to contaminant-free air. In wastewater treatment, air stripping occurs most commonly in aerated grit chambers, aerated biological treatment processes, and aerated transfer channels. Specially designed facilities (e.g., stripping towers) for gas stripping are considered in Sec. 11–10 in Chap 11 and Sec 15–5 in Chap. 15.

Table 16–12

Physical properties of selected volatile and semi-volatile organic compounds[a,b]

Compounds	mw	mp, °C	bp, °C	vp, mm Hg	vd	sg	Sol., mg/L	Cs, g/m³	H, m³·atm/mol	log K_{ow}
Benzene	78.11	5.5	80.1	76	2.77	.8786	1780	319	5.49×10^{-3}	2.1206
Chlorobenzene	112.56	−45	132	8.8	3.88	1.1066	500	54	3.70×10^{-3}	2.18–3.79
o-Dichlorobenzene	147.01	18	180.5	1.60	5.07	1.036	150	N/A	1.7×10^{-3}	3.3997
Ethylbenzene	106.17	−94.97	136.2	7	3.66	0.867	152	40	8.43×10^{-3}	3.13
1,2-Dibromoethane	187.87	9.8	131.3	10.25	0.105	2.18	2699	93.61	6.29×10^{-4}	N/A
1,1-Dichloroethane	98.96	−97.4	57.3	297	3.42	1.176	7840	160.93	5.1×10^{-3}	N/A
1,2-Dichloroethane	98.96	−35.4	83.5	61	3.4	1.25	8690	350	1.14×10^{-3}	1.4502
1,1,2,2-Tetrachloroethane	167.85	−36	146.2	14.74	5.79	1.595	2800	13.10	4.2×10^{-4}	2.389
1,1,1-Trichlorethane	133.41	−32	74	100	4.63	1.35	4400	715.9	3.6×10^{-3}	2.17
1,1,2-Trichloroethane	133.4	−36.5	133.8	19	N/A	N/A	4400	13.89	7.69×10^{-4}	N/A
Chloroethene	62.5	−153	−13.9	2548	2.15	0.912	6000	8521	6.4×10^{-2}	N/A
1,1-Dichloroethene	96.94	−122.1	31.9	500	3.3	1.21	5000	2640	1.51×10^{-2}	N/A
c-1,2-Dichloroethene	96.95	−80.5	60.3	200	3.34	1.284	800	104.39	4.08×10^{-3}	N/A
t-1,2-Dichloroethene	96.95	−50	48	269	3.34	1.26	6300	1428	4.05×10^{-3}	N/A
Tetrachloroethene	165.83	−22.5	121	15.6	N/A	1.63	160	126	2.85×10^{-2}	2.5289
Trichloroethene	131.5	−87	86.7	60	4.54	1.46	1100	415	1.17×10^{-2}	2.4200
Bromodichloromethane	163.8	−57.1	90	N/A	N/A	1.971	N/A	N/A	2.12×10^{-3}	N/A
Chlorodibromomethane	208.29	<−20	120	50	N/A	2.451	N/A	N/A	8.4×10^{-4}	N/A
Dichloromethane	84.93	−97	39.8	349	2.93	1.327	20000	1702	3.04×10^{-3}	N/A
Tetrachloromethane	153.82	−23	76.7	90	5.3	1.59	800	754	2.86×10^{-2}	2.7300
Tribromomethane	252.77	8.3	149	5.6	8.7	2.89	3130	7.62	5.84×10^{-4}	N/A
Trichloromethane	119.38	−64	62	160	4.12	1.49	7840	1027	3.10×10^{-3}	1.8998
1,2-Dichloropropane	112.99	−100.5	96.4	41.2	3.5	1.156	2600	25.49	2.75×10^{-3}	N/A
2,3-Dichloropropene	110.98	−81.7	94	135	3.8	1.211	insol.	110	N/A	N/A
t-1,3-Dichloropropene	110.97	N/A	112	99.6	N/A	1.224	515	110	N/A	N/A
Toluene	92.1	−95.1	110.8	22	3.14	0.867	515	110	6.44×10^{-3}	2.2095

[a] Data were adapted from Lang (1987).

[b] All values are reported at 20°C.

Note: mw = molecular weight, mp = melting point, bp = boiling point, vd = vapor density relative to air, s_g = specific gravity relative to water, Sol = solubility, C_s = saturation concentration in air, H = Henry's Law Constant, log K_{ow} = logarithm of the octanol-water partition coefficient.

Locations Where VOCs Are Emitted. The principal locations where VOCs are emitted from wastewater collection and treatment facilities are summarized in Table 16–13. The degree of VOC removal at a given location will depend on local conditions. Mass transfer is considered in the following section.

Table 16–13

Sources, methods of release, and control of VOCs from wastewater facilities

Source	Method of release	Suggested control strategies
Domestic, commercial, and industrial discharges	Discharge of small amounts of VOCs in liquid wastes	Institute active source control program to limit the discharge of VOCs to municipal sewers
Wastewater sewers	Volatilization from the surface enhanced by flow induced turbulence	Seal existing manholes. Eliminate the use of structures that create turbulence and enhance volatilization
Sewer appurtenances	Volatilization due to turbulence at junctions, etc., volatilization and air stripping at drop manholes and junction chambers	Isolate and cover existing appurtenances
Pump stations	Volatilization and air stripping at influent wet-well inlets	Vent gases from wet-well to VOC treatment unit Use variable-speed pumps to reduce size of wet-well
Bar racks	Volatilization due to turbulence	Cover units, reduce headloss through bar racks
Comminutors	Volatilization due to turbulence	Cover units, use inline enclosed comminutors
Parshall flume	Volatilization due to turbulence	Cover units, use alternative measuring device
Grit chamber	Volatilization due to turbulence in horizontal-flow grit chambers Volatilization and air stripping in aerated grit chambers Volatilization in vortex-type grit chambers	Cover aerated and vortex-type grit chambers Reduce turbulence in horizontal-flow grit chambers; cover if necessary
Equalization basins	Volatilization from surface enhanced by local turbulence Air stripping where diffused air is used	Cover units, use submerged mixers. Reduce air flow
Primary and secondary sedimentation tanks	Volatilization from surface. Volatilization and air stripping at overflow weirs, in effluent channel, and at other discharge points	Cover tanks, replace overflow weirs with drops with submerged launders
Biological treatment	Air stripping in diffused-air activated sludge. Volatilization in activated sludge processes with surface aerators. Volatilization from surface enhanced by local turbulence. Volatilization from trickling filters	Cover units, in activated sludge systems, use submerged mixers and reduce aeration rate
Transfer channels	Volatilization from surface enhanced by local turbulence. Volatilization and air stripping in aerated transfer channels	Use enclosed transfer channels
Digester gas	Uncontrolled release of digester gas. Discharge of incompletely combusted or incinerated digester gas	Controlled thermal incineration, combustion, or flaring of digester gas

Mass Transfer Rates for VOCs

The mass transfer of VOCs can, for practical purposes, be modeled using the following equation (Roberts et al., 1984, and Thibodeaux, 1979):

$$r_{\mathrm{VOC}} = -(K_La)_{\mathrm{VOC}}(C - C_s) \tag{16–21}$$

where r_{VOC} = rate of VOC mass transfer, mg/m^3·h
$(K_La)_{\mathrm{VOC}}$ = overall VOC mass transfer coefficient, 1/h
C = concentration of VOC in liquid, mg/m^3
C_s = saturation concentration of VOC in liquid, mg/m^3

Due to chemical handling and analytical requirements, measuring $K_L\mathrm{a}_{\mathrm{VOC}}$ is much more difficult than measuring $K_La_{O_2}$. Therefore, a practical approach is to relate the K_La_{VOC} to the $K_La_{\mathrm{O_2}}$. The following equation is used to relate the mass transfer coefficients as a function of the VOC and O_2 diffusion coefficients in water:

$$(K_La)_{\mathrm{VOC}} = (K_La)_{\mathrm{O_2}}\left(\frac{D_{\mathrm{VOC}}}{D_{\mathrm{O_2}}}\right)^n \tag{16–22}$$

where $(K_La)_{\mathrm{VOC}}$ = system mass transfer coefficient, T^{-1} (1/h)
$(K_La)_{\mathrm{O_2}}$ = system oxygen mass transfer coefficient, T^{-1} (1/h)
D_{VOC} = diffusion coefficient of VOC in water, $\mathrm{L^2T^{-1}}$ (cm^2/s)
$D_{\mathrm{O_2}}$ = diffusion coefficient of oxygen in water, $\mathrm{L^2T^{-1}}$ (cm^2/s)
n = coefficient

Diffusion coefficient values for different compounds can be obtained from Schwarzenbach et al. (1993) or in other handbooks. It should be noted that there is often considerable variation in the values reported in the literature. However, based on the results a variety of experimental investigations of the relationship between K_La_{VOC} and $K_La_{\mathrm{O_2}}$ it has been found that Eq. (16–22) is generally applicable, and that the value for n varies depending on whether the gas/liquid transfer is accomplished by surface aeration, diffused aeration, or a packed column air stripper, and the power intensity of the gas transfer device (Roberts and Dandliker, 1983; Matter-Muller et al., 1981; Hsieh et al., 1993; Libra, 1993; and Bielefeldt and Stensel, 1999). For practical power intensities of less than 100 W/m^3 a reasonable value of n is 0.50 for packed columns and mechanical aeration and 1.0 for diffused aeration. For higher power intensities the work of Hsieh et al. (1993) should be consulted. The K_La_{VOC} value has also been found to be essentially the same in wastewater as in tap water (Bielefeldt and Stensel, 1999).

Mass Transfer of VOCs from Surface and Diffused-Air Aeration Processes

The amount of VOCs released from a complete-mix reactor used for the activated-sludge process will depend on the method of aeration (e.g., surface aeration or diffused aeration).

Complete-Mix Reactor with Surface Aeration. A materials balance for the stripping of a VOC written around a complete-mix reactor is as follows, assuming no other removal mechanisms for the VOC compound such as biodegradation or solids sorption are applicable.

1. General word statement:

$$\begin{array}{c}\text{Rate of accumulation}\\\text{of VOC within}\\\text{the system boundary}\end{array} = \begin{array}{c}\text{rate of flow of}\\\text{VOC into the}\\\text{system boundary}\end{array} - \begin{array}{c}\text{rate of flow of}\\\text{VOC out of the}\\\text{system boundary}\end{array} + \begin{array}{c}\text{amount of VOC}\\\text{removed through}\\\text{system boundary}\\\text{by stripping}\end{array} \tag{16–23}$$

2. Simplified word statement:

Accumulation = inflow-outflow + decrease due to stripping (16–24)

3. Symbolic representation:

$$\frac{dC}{dt}V = QC_i - QC_e + r_{\text{VOC}}V \tag{16–25}$$

where dC/dt = rate of change in VOC concentration in reactor
V = volume of complete mix reactor, L^3 (m^3)
Q = liquid flowrate, L^3T^{-1} (m^3/s)
C_i = concentration of VOC in influent to reactor, ML^{-3} (mg/m^3)
C_e = concentration of VOC in effluent from reactor, ML^{-3} (mg/m^3)
rVOC = rate of VOC mass transfer, $ML^{-3}T^{-1}$ ($mg/m^3 \cdot h$)

Substituting for rVOC from Eq. (16–21) and τ for V/Q yields

$$\frac{dC}{dt} = \frac{C_i - C_e}{\tau} + [-(K_La)_{\text{VOC}}(C_e - C_s)] \tag{16–26}$$

If steady-state conditions are assumed and it is further assumed that C_s is equal to zero, then the amount of VOC that can be removed by surface aeration is given by the following expression:

$$1 - \frac{C_e}{C_i} = 1 - [1 + (K_La)\tau]^{-1} \tag{16–27}$$

If a significant amount of the VOC is adsorbed or biodegraded, the results obtained with the above equation will be overestimated. The above analysis can also be used to estimate the release of VOCs at weirs and drops by assuming the time period is about 30s.

Complete Mix Reactor with Diffused-Air Aeration. The corresponding expression to Eq. (16–24) for a complete-mix reactor with diffused-air aeration is developed by a mass balance on the VOC compound. At steady state the VOC in equals the VOC out, and the corresponding mass balance is

$$\begin{matrix}\text{Inflow in} \\ \text{liquid stream}\end{matrix} = \begin{matrix}\text{outflow} \\ \text{liquid stream}\end{matrix} + \begin{matrix}\text{outflow} \\ \text{in exit gas}\end{matrix}$$

$$QC_i = QC_e + Q_gC_{g,e} \tag{16–28}$$

where Q = liquid flowrate, L^3T^{-1} (m^3/s)
C_i = VOC concentration in influent, ML^{-3} (mg/m^3)
C_e = VOC concentration in effluent, ML^{-3} (mg/m^3)
Q_g = gas flowrate, L^3T^{-1} (m^3/s)
$C_{g,e}$ = VOC concentration in exit gas, ML^{-3} (mg/m^3)

The general expression for the removal of VOC by gas sparging through a liquid is (Bielefeldt and Stensel, 1999)

$$Q_gC_{g,e} = Q_gH_uC_e(1 - e^{-\phi}) \tag{16–29}$$

where H_u = Henry's law constant, dimensionless
ϕ = VOC saturation parameter defined as

$$\phi = \frac{(K_La)_{\text{VOC}}V}{H_uQ_g} \tag{16–30}$$

Eq. (16–29) can be rearranged as follows:

$$Q(C_i - C_e) = H_u C_e (1 - e^{-\phi}) \tag{16–31}$$

Solving Eq. (16–31) for C_e/C_i yields

$$\frac{C_e}{C_i} = \left[1 + \frac{Q_g}{Q} H_u (1 - e^{-\phi})\right]^{-1} \tag{16–32}$$

and the fraction removed is given by

$$1 - \frac{C_e}{C_i} = 1 - \left[1 + \frac{Q_g}{Q} H_u (1 - e^{-\phi})\right]^{-1} \tag{16–33}$$

The application of the above equations is illustrated in Example 16–3.

EXAMPLE 16–3 **Stripping of Trichloroethene in the Activated Sludge Process** Determine the amount of trichloroethene (TCE) that can be stripped in a complete-mix activated sludge reactor equipped with a diffused-air aeration system. Assume the following conditions apply:

1. Wastewater flowrate = 4000 m^3/d
2. Aeration tank volume = 1000 m^3
3. Depth of aeration tank = 6 m
4. Air flowrate = 50 m^3/min at standard conditions
5. Oxygen mass transfer rate, $(K_L a)_{O_2} = 6.2/h$
6. $H_{TCE} = 1.17 \times 10^{-2}\ m^3 \cdot atm/mol$ (see Table 16–10)
7. $n = 1.0$
8. Temperature = 20°C
9. Oxygen diffusivity = $2.11 \times 10^{-5}\ cm^2/s$
10. Trichloroethene diffusivity ~ $1.0 \times 10^{-5}\ cm^2/s$

Solution

1. Determine the quantity of air referenced to the mid-depth of the aeration tank, which represents the depth for an average bubble size. Using the universal gas law, the air flowrate at mid-depth (3 m) is:

$$Q_g = (50\ m^3/min)\frac{(10.33\ m)}{(10.33\ m + 3\ m)} = 38.7\ m^3/min$$

Note: 10.33 = standard atmospheric pressure expressed in m of H_2O.

2. Determine the air/liquid ratio.

$$Q = \frac{(4000\ m^3/d)}{(1440\ min/d)} = 2.78\ m^3/min$$

$$\frac{Q_g}{Q} = \frac{(38.7\ m^3/min)}{(2.78\ m^3/min)} = 13.9$$

3. Estimate the mass transfer coefficient for TCE using Eq. (16–22).

$$(K_L a)_{VOC} = (K_L a)_{O_2}\left(\frac{D_{VOC}}{D_{O_2}}\right)^n = (6.2/h)\left[\frac{(1.0 \times 10^{-5}\ cm^2/s)}{(2.11 \times 10^{-5}\ cm^2/s)}\right]^{1.0}$$

$$(K_L a)_{VOC} = 2.94\ /h = 0.049/min$$

4. Determine the dimensionless value of the Henry's law constant for TCE using Eq. (2–51).

$$H_u = \frac{H}{RT}$$

$$H_u = \frac{0.0117}{0.000082057 \times (273 + 20)} = 0.487$$

5. Determine the saturation parameter ϕ using Eq (16–30).

$$\phi = \frac{(K_La)_{\text{VOC}}V}{H_uQ_g}$$

$$\phi = \frac{(0.049/\text{min})(1000\ \text{m}^3)}{(0.228 \times 38.7\ \text{m}^3/\text{min})} = 5.55$$

6. Determine the fraction of TCE removed from the liquid phase using Eq. (16–33).

$$1 - \frac{C_e}{C_i} = 1 - \left[1 + \frac{Q_g}{Q}(H_u)(1 - e^{-\phi})\right]^{-1}$$

$$1 - \frac{C_e}{C_i} = 1 - [1 + 13.9(0.487)(1 - e^{-5.55})]^{-1}$$

$$1 - \frac{C_e}{C_i} = 1 - 0.13 = 0.87 \text{ or } 87\%$$

Comment The computations presented in this example are based on the assumption that the concentration of TCE in the influent is not being reduced by diffusion and turbulence in the headworks, primary sedimentation tank or by adsorption or biological degradation in the aeration tank.

Control Strategies for VOCs

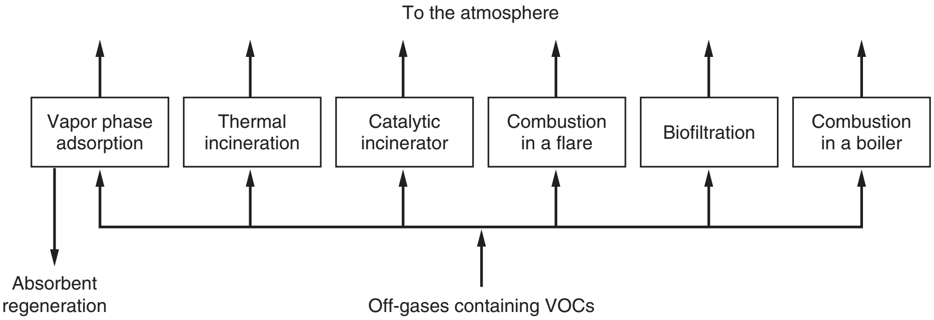
Volatilization and gas stripping are, as noted previously, the principal means by which VOCs are released from wastewater-treatment facilities. In general, it can be shown that the release of VOCs from open surfaces is quite low compared to the release of VOCs at points of liquid turbulence and by gas stripping. Thus, the principal strategies for controlling the release of VOCs, as reported in Table 16–13, are (1) source control, (2) elimination of points of turbulence, and (3) the covering of various treatment facilities. Three serious problems associated with the covering of treatment facilities are (1) treatment of the off-gases containing VOCs, (2) corrosion of mechanical parts, and (3) provision for confined space entry for personnel for equipment maintenance.

Treatment of Off-Gases

The off-gases containing VOCs from covered treatment facilities will have to be treated before they can be discharged to the atmosphere. Options for the off-gas treatment include (see Fig. 16–17) (1) vapor-phase adsorption on granular activated carbon or other VOC selective resins, (2) thermal incineration, (3) catalytic oxidation, (4) combustion in a flare, (5) biofiltration, and (6) combustion in a boiler or process heater (U.S. EPA, 1986; WEF, 1997). The application of these processes will depend primarily on the volume of air to be treated and the types and concentrations of the VOCs contained in the airstream. The first four of these off-gas treatment processes are considered in greater detail in the

Figure 16–17

Options for treating off-gases containing VOCs. (Adapted from Eckenfelder, 2000.)



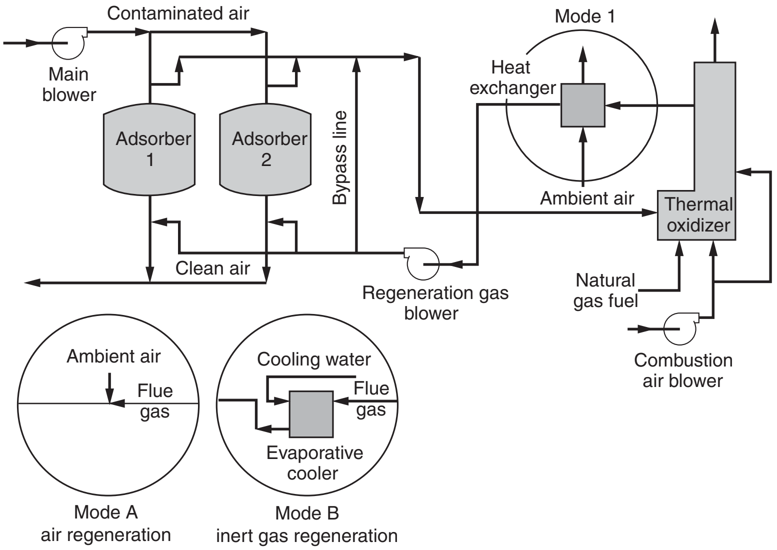
following discussion. Biofiltration was discussed previously in Sec. 16–3. A boiler or process heater is used only where a combustion process is included as part of the plant facilities.

Vapor-Phase Adsorption. Adsorption is the process whereby hydrocarbons and other compounds are adsorbed selectively on the surface of materials such as activated carbon, silica gel, or alumina. Of the available adsorbents, activated carbon is used most widely. The adsorption capacity of an adsorbent for a given VOC is often represented by adsorption isotherms that relate the amount of VOC adsorbed (adsorbate) to the equilibrium pressure (or concentration) at constant temperature. Typically, the adsorption capacity increases with the molecular weight of the VOC being adsorbed. In addition, unsaturated compounds are generally adsorbed more completely than saturated compounds, and cyclical compounds are adsorbed more easily than linearly structured materials. It should be noted that careful evaluation of the adsorption media should be undertaken to be assured that the compound being adsorbed does not react with the adsorbent. Also, the adsorption capacity is enhanced by lower operating temperatures and higher concentrations. VOCs characterized by low vapor pressures are more easily adsorbed than those with high vapor pressures (U.S. EPA, 1986).

Steps in VOC Adsorption Process. The two main steps involved in the VOC adsorption process are (1) the continuous adsorption in multiple beds (see Fig. 16–18) and (2) batch regeneration of the adsorbent. For control of continuous emission streams, at least one bed

Figure 16–18

Gas phase carbon adsorption and regeneration system for the treatment of VOCs in off-gas.



remains online in the adsorption mode while the other is being regenerated. In a typical batch operation, the off-gas containing VOCs is passed through the carbon bed where the VOCs are adsorbed on the bed surface. As the adsorption capacity of the bed is approached, traces of VOCs appear in the exit stream, up to the level where the breakthrough point of the bed has been attained. The off-gas is then directed to a parallel bed containing regenerated adsorbent, and the process continues. Concurrently, the saturated bed is regenerated by the passage of hot air (see Fig. 16–18, Mode A), hot inert gases (see Fig. 16–18, Mode B), low-pressure steam, or a combination of vacuum and hot gas. Because adsorption is a reversible process, the VOCs adsorbed on the bed can be desorbed by supplying heat (equivalent to the amount of heat released during adsorption). Small residual amounts of VOCs are always left on the carbon bed, because complete desorption is technically difficult to achieve and economically impractical. Regeneration with hot air and a hot inert gas is considered in the following discussion.

Hot Air Regeneration. When the VOCs are either nonflammable or have a high ignition temperature and thus do not pose a risk of carbon fires, hot air regeneration is used. A portion of the hot flue gas in the oxidizer is mixed with ambient air to cool the gas to below 180°C (350°F). The regeneration gas is driven upflow (or countercurrent to adsorption flow) through the GAC adsorber. As the temperature of the carbon bed rises, the desorbed organics are transferred to the regeneration gas stream. The regeneration gas containing the desorbed VOCs is sent directly to the thermal oxidizer where the VOCs are oxidized. After the bed has been maintained at the desired regeneration temperature for a sufficient period of time, regeneration is ended. The bed is then cooled to approximately ambient temperature by shutting off the hot regeneration gas and continuing to pass ambient air through the carbon bed. The regeneration and cooling times are predetermined based on the amount of carbon in the adsorber and the expected loading on the carbon (U.S. EPA, 1986).

Inlet Gas Regeneration. Where the VOCs contained in the off-gas include compounds such as ketones and aldehydes that may pose fire risks at elevated temperatures in the presence of oxygen, inert gas regeneration is used. A relatively inert gas can be obtained by passing a portion of the hot flue gas from the thermal oxidizer through an evaporative cooler. It is possible, therefore, to keep the oxygen concentration in the regeneration gas at 2 to 5 percent by volume. The desorbed VOCs are transferred along with the regeneration gas to the thermal oxidizer. A controlled amount of secondary air is added to the oxidizer. The addition of air ensures complete combustion of the VOCs but limits the excess oxygen level in the oxidizer to an acceptable range (e.g., 2 to 5 percent by volume). Regeneration is complete when the carbon bed has reached the necessary temperature for a given period of time, and VOCs are no longer being desorbed from the bed. Cooling of the bed is accomplished by increasing the water flowrate to the evaporative cooler and reducing the regeneration gas temperature to between 105 and 120°C (220 and 250°F).

Thermal Incineration. Thermal incineration [see Fig. 16–12(a)] is used to oxidize VOCs at high temperatures. The most important variables to consider in thermal incinerator design are the combustion temperature and residence time because these design variables determine the VOC destruction efficiency of the incinerator. Further, at a given combustion temperature and residence time, destruction efficiency is also affected by the degree of turbulence, or mixing of the emission stream and hot combustion gases, in the incinerator. In addition, halogenated organics are more difficult to oxidize than unsubstituted organics; hence, the presence of halogenated compounds in the emission stream requires higher temperature and longer residence times for complete oxidation. When emission streams treated by thermal incineration are dilute (i.e., low heat content), supplementary fuel is required to maintain the desired combustion temperatures. Supplementary fuel requirements may be reduced by recovering the energy contained in the hot flue gases from the incinerator. Also, depending on the byproducts of incineration (e.g., HCl, H_2SO_4, HF), it may be necessary to provide an acid gas scrubber.

Catalytic Oxidation. In catalytic oxidation [see Fig. 16–12(b)], VOCs in an emission stream are oxidized with the help of a catalyst. A catalyst is a substance that accelerates the rate of a reaction at a given temperature without being appreciably changed during the reaction. Catalysts typically used for VOC oxidation include platinum and palladium; other formulations are also used, including metal oxides for emission streams containing chlorinated compounds. The catalyst bed (or matrix) in the oxidizer is generally a metal mesh-mat, ceramic honeycomb, or other ceramic matrix structure designed to maximize catalyst surface area. The catalysts may also be in the form of spheres or pellets. Before passing through the catalyst bed, the emission stream is preheated, if necessary, in a natural gas-fired preheater (U.S. EPA, 1986).

The performance of a catalytic oxidizer is affected by several factors including (1) operating temperature, (2) space velocity (reciprocal of residence time), (3) VOC composition and concentration, (4) catalyst properties, and (5) presence of catalyst poisons or inhibitors in the emission stream. In catalytic incinerator design, the important variables are the operating temperature at the catalyst bed inlet and the space velocity. The operating temperature for a particular destruction efficiency is dependent on the concentration and composition of the VOC in the emission stream and the type of catalyst used (U.S. EPA, 1986). As with incineration, it may be necessary to provide an acid gas scrubber.

Combustion in a Flare. Flares, another type of thermal incineration commonly used for disposal of waste digester gas, can be used to destroy most VOCs found in off-gas streams. Flares can be designed and operated to handle fluctuations in emission VOC content, inerts content, and flowrate. Several different types of flares are available, including steam-assisted, air-assisted, and pressure-head flares. Steam-assisted flares are employed in cases where large volumes of waste gases are released. Air-assisted flares are generally used for moderate off-gas gas flows. Pressure-head flares are used for small gas flows.

16–5 EMISSIONS FROM THE COMBUSTION OF GASES AND SOLIDS

Various types of fuels are used at wastewater treatment facilities to generate heat and electricity onsite to supplement energy supplied from outside sources and to operate equipment. As discussed in Chaps. 13 and 14, part or all of gases and solids generated at wastewater treatment facilities can also be used as fuels within the wastewater treatment facility. Fuel sources, combustion systems used at wastewater treatment plants and their emissions, and flaring of excess digester gas are discussed in this section.

Sources of Fuels

Fuels used for the operation of a wastewater treatment plant include various grades of fuel oils, natural gas, and digester gas. The types of fuels used will depend on the types of combustion systems used at a treatment facility. Emissions of certain pollutants will vary with the quality of the fuel and the type and condition of the combustion system. Typical fuels used at a wastewater treatment plant are listed in Table 16–14.

Table 16–14

Fuels used commonly at wastewater treatment plants

Fuel type	Characteristics
No. 2 fuel oil[a]	A distillate fuel oil that has a distillation temperature of 640 degrees Fahrenheit at the 90-percent recovery point. It is used in atomizing type burners for domestic heating or for moderate capacity commercial/industrial burner units.
No. 4 fuel oil[a]	A distillate fuel oil made by blending distillate fuel oil and residual fuel oil stocks. It is used extensively in industrial plants and in commercial burner installations that are not equipped with preheating facilities. It also includes No. 4 diesel fuel used for low- and medium-speed diesel engines.
Natural gas[b]	Typical composition of natural gas before refining is 70 to 90 percent methane, up to 20 percent of ethane, propane, and butane, and up to 8 percent carbon dioxide. Sulfur content is typically low but unprocessed natural gas may have higher sulfur content.
Digester gas	Typically contains 50 to 65 percent methane. Siloxanes in the digester gas can be detrimental to the combustion system. The level of cleaning required for use depends on the types of the combustion system. Emission of sulfur oxides depends on the sulfur contents in the digester gas.

[a] From U.S. Energy Information Administration.

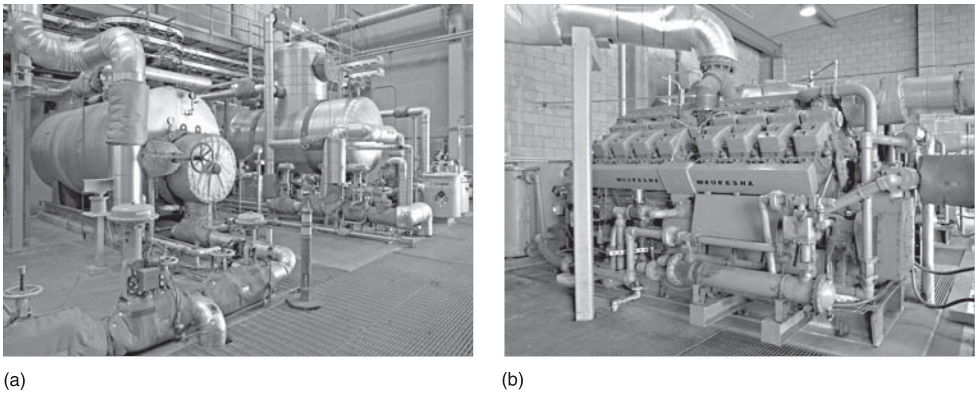
[b] Typical composition from Natural Gas Supply Association (naturalgas.org).

Combustion Systems Used at Wastewater Treatment Plants

Major combustion systems used at wastewater treatment plants include boilers and power generators. In some cases, solids are incinerated within the treatment facility (see Sec. 14–4 in Chap. 14). Flares are used to combust excess digester gas.

Boiler. Boilers used at wastewater treatment facilities use fuel oil, natural gas, digester gas, or a combination of these fuels to produce hot water or steam [see Fig. 16–19(a)]. The main purposes of using boilers are to provide heat for maintaining temperature of anaerobic digesters and to provide heat and hot water in the buildings. In addition, boilers are used to produce process steam for use in sludge drying and to treat biosolids thermally in various processes. The use of steam boilers is less common, especially at smaller wastewater treatment facilities, as additional treatment of the boiler feed water may be required. Combustion of a fuel is controlled to heat water to a desired temperature, and hot water or steam is circulated for various purposes.

Reciprocating Engine. Internal combustion reciprocating engines are used most commonly for the generation of electricity [see Fig. 16–19(b)]. Reciprocating engines can also be used to operate pumps, gas compression, or chillers. Reciprocating engines are categorized as spark ignition or compression ignition. Diesel fuel is used for compression ignition engines, or the engines can be set up to use natural gas with a small amount of diesel to ignite with compression. Emission of nitrogen oxides (NOx) is correlated with the combustion temperature as well as nitrogen in the fuel (typically a concern with liquid fuels), and NOx removal may be necessary if the emission exceeds the permit level. Emission of particulate matters and carbon monoxide (CO) are generally correlated with incomplete combustion of fuels.



(a) (b)

Figure 16–19

Combustion systems used commonly at wastewater treatment facilities: (a) view of heat recovery boiler and (b) view of internal combustion engine.

Gas Turbine. In gas turbines, compressed fuel and air is combusted in a combustion chamber, and the expanded hot exhaust gas is used to spin the power turbine. Because gas turbines generally operate at a significantly higher temperature than reciprocating engines, the emission of NOx has to be controlled. The use of gas turbines is increasing with the use of combined heat and power (CHP) system at wastewater treatment facilities [see Fig. 17–8(d) in Chap. 17]. Steam is generated using a heat exchanger that removes excess energy from turbine exhaust gases. The steam produced in this manner can be used to replace boiler-produced steam for heating or used in a steam turbine to produce additional electricity.

Emissions of Concern from Combustion Sources

The major emissions of concern from combustion sources are NOx, CO, particulate matter (PM) and sulfur dioxide (SO_2). VOCs are generally emitted in small quantities from the incomplete combustion of fuel and are only a concern when a significant combustion equipment is being used. In addition, combustion emissions may contain hazardous air pollutants (HAPs) in small quantities such as formaldehyde, and any metals contained in the fuel.

Fuel gas such as natural gas and digester gas is oxidized completely to carbon dioxide (CO_2) when a combustion temperature is at or above 850°C (1560°F) and the combustion time is 0.3 s or longer. Lower combustion temperatures, shorter combustion times, and incomplete mixing of fuel and air can cause the incomplete oxidation of the fuel resulting in the production of CO. Carbon monoxide emission is controlled by adjusting the operating conditions, or by the use of a post-combustion catalytic oxidation system. Emission of VOC is essentially the emission of unburned hydrocarbons in the fuel as a result of incomplete combustion.

At combustion temperatures above 1200°C, oxidation of the nitrogen gas contained in the combustion air will occur, and NOx will be formed. Particulate matter emissions from fuel combustion include both organic and inorganic fractions. The organic fractions of particulate matter include by-products of incomplete combustion, such as polycyclic organic matter. Inorganic fractions include metals as well as acid mists. Emissions of sulfur dioxide occur as a result of the oxidation of sulfur contained in the fuel. Both liquid and gaseous fuels contain sulfur both in elemental as well as in organic fractions.

Emission Factor. An emission factor is a representative value that is used to relate the quantity of a pollutant released to the atmosphere by a specific activity associated with the release of that pollutant (U.S. EPA, 1995). Pollutant emissions are estimated using Eq. (16–34):

$$E = (A)(EF)(1 - ER/100) \tag{16–34}$$

where, E = emissions
A = activity rate, e.g., rate of the fuel consumption
EF = emission factor for given activity
ER = overall emission reduction efficiency for given activity, %

Activity rate is a general term for the rate of the fuel consumption when emissions from a combustion system are being considered. For example, if there is no device to destroy or capture the pollutant, the overall emission reduction efficiency (*ER*) is zero. Thus, without emission reduction Eq. (16–34) is simply the product of the rate of fuel consumption and the emission factor. As noted above, emission factors are available for specific emission generating activities. The U.S. EPA has been compiling the emission factors in AP 42, Compilation of Air Pollutant Emission Factors (U.S. EPA, 1995). Emission factors for criteria pollutants from fuel oil and natural gas combustion are reported in Table 16–15. The overall emission reduction is a product of destruction or removal efficiency and the capture efficiency of the control system such as cyclones, scrubbers, and catalytic reduction systems. It should be noted that the estimate based on the emission factor from AP42 is usually conservative. For more accurate estimate, emission data from the equipment supplier should be referenced.

Control of Nitrogen Oxides. Methods to control the emission of NOx include control of combustion temperature and residence time, and the use of catalytic reduction systems. Combustion temperature and residence time may be controlled by adjusting air to fuel ratio or by injecting water or steam. The latter is used mainly for the gas turbine systems. It should be noted that the use of dilution air could result in the emission of CO, and the operating conditions must be controlled to minimize the emission of both NOx and CO. Catalytic reduction systems include selective catalytic reduction with the use of ammonia or urea, and non-selective catalytic reduction. Non-selective catalytic reduction is applicable only for the combustion system with exhaust oxygen levels of 4 percent or less. In selective catalytic reduction, nitrogen oxides, ammonia (or urea) and oxygen react in presence of the catalyst to form nitrogen gas (N_2) and water (H_2O). Selective catalytic reduction is used for both reciprocating engines and gas turbines.

Flaring of Digester Gas

Digester gas, a gas generated from anaerobic digestion of sludges, typically contains 55 to 65 percent methane and can be used as a fuel source to generate heat and electricity. When a wastewater treatment facility does not utilize all digester gas, the excess gas must be fared (see Fig. 16–20).

Combustion of methane gas can be expressed as

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O \tag{16–35}$$

Air contains approximately 21 percent oxygen and requiring 2 moles of oxygen to completely oxidize 1 mole of methane. Therefore, a theoretical ratio of methane to air to achieve complete combustion of methane is approximately 9.5. If digester gas contains 60 percent methane, the theoretical digester gas to air ratio is 5.7. Practically, excess air with a well controlled flaring system is necessary to achieve effective combustion of digester gas. Flaring of digester gas will result in emissions described previously in Table 16–1.

Table 16–15

Emission factors for criteria pollutants from fuels used commonly at wastewater treatment facilities[a]

Category	Unit	CO	NOx	SO$_2^b$	PM (filterable)	PM (condensable)	Lead
Boilers, larger than 106 GJ/h (100 MBtu/h)							
Natural gas							
Uncontrolled[c]	kg/m^3 (lb/10^6 ft^3)	1344 (84)	3040 (190)	9.6 (0.6)	30.4 (1.9)	91.2 (5.7)	0.08 (0.0005)
Controlled, low NOx burner	kg/ m^3 (lb/10^6 ft^3)	1344 (84)	2240 (140)	9.6 (0.6)	30.4 (1.9)	91.2 (5.7)	0.08 (0.0005)
Controlled, flue gas recirculation	kg/ m^3 (lb/10^6 ft^3)	1344 (84)	1600 (100)	9.6 (0.6)	30.4 (1.9)	91.2 (5.7)	0.08 (0.0005)
Fuel oil No. 2	kg/ m^3 (lb/10^3 gal)	0.60 (5)	2.88 (24)	17.0S (142S)	0.24 (2)	0.16 (1.3)	
Fuel oil No. 2, low NOx burner, flue gas recirculation	kg/ m^3 (lb/10^3 gal)	0.60 (5)	1.20 (10)	17.0S (142S)	0.24 (2)	0.16 (1.3)	
Fuel oil No. 4, normal firing	kg/ m^3 (lb/10^3 gal)	0.60 (5)	5.63 (47)	18.0S (150S)	0.24 (2)	0.16 (1.3)	
Fuel oil No. 4, tangential firing	kg/ m^3 (lb/10^3 gal)	0.60 (5)	3.83 (32)	18.0S (150S)	0.24 (2)	0.16 (1.3)	
Boilers, smaller than 106 GJ/h (100 MBtu/h)							
Natural gas							
Uncontrolled	kg/1 m^3 (lb/10^6 ft^3)	1344 (84)	1600 (100)	9.6 (0.6)	30.4 (1.9)	91.2 (5.7)	0.08 (0.0005)
Controlled, low NOx burner	kg/ m^3 (lb/10^6 ft^3)	1344 (84)	800 (50)	9.6 (0.6)	30.4 (1.9)	91.2 (5.7)	0.08 (0.0005)
Controlled, flue gas recirculation	kg/ m^3 (lb/10^6 ft^3)	1344 (84)	512 (32)	9.6 (0.6)	30.4 (1.9)	91.2 (5.7)	0.08 (0.0005)
Fuel oil No. 4, oil fired	kg/ m^3 (lb/10^3 gal)	0.60 (5)	2.40 (20)	18.0S (150S)	0.84 (7)	0.16 (1.3)	

(*continued*)

Table 16–15 (*Continued*)

Category	Unit	CO	NOx	SO_2^b	PM (filterable)	PM (condensable)	Lead
Gas turbine							
Natural gas							
Uncontrolled	kg/GJ	0.19	0.74	2.19S	0.0042	0.011	
	(lb/MMBtu)	(0.082)	(0.32)	(0.94S)	(0.0019)	(0.0047)	
Water steam injection	kg/GJ	0.070	0.30	2.19S	0.0042	0.011	
	(lb/MMBtu)	(0.030)	(0.13)	(0.94S)	(0.0019)	(0.0047)	
Lean premix	kg/GJ	0.035	0.23	2.19S	0.0042	0.011	
	(lb/MMBtu)	(0.015)	(0.099)	(0.94S)	(0.0019)	(0.0047)	
Digester gas, uncontrolled	kg/GJ	0.040	0.37	2.19S	0.0042	0.011	
	(lb/MMBtu)	(0.017)	(0.16)	(0.94S)	(0.0019)	(0.0047)	
Natural gas fired reciprocating engines							
4-stroke, lean burn, 90–105% load	kg/GJ	0.737	9.49	(5.58×10^{-4})	(7.71×10^{-5})	(9.91×10^{-3})	
	(lb/MMBtu)	(0.317)	(4.08)				
4-stroke, lean burn, <90% load	kg/GJ	1.30	1.97	(5.58×10^{-4})	(7.71×10^{-5})	(9.91×10^{-3})	
	(lb/MMBtu)	(0.557)	(0.847)				
Large diesel (>600hp) and dual fuel reciprocating engines							
Large diesel engines, uncontrolled	kg/GJ	2.0	7.4	$2.35S_1$	0.14	0.018	
	(lb/MMBtu)	(0.85)	(3.2)	$(1.01S_1)$	(0.062)	(0.0077)	
Large diesel engines, controlled	kg/GJ	2.0	4.4	$2.35S_1$			
	(lb/MMBtu)	(0.85)	(1.9)	$(1.01S_1)$			
Dual fuel engines[d]	kg/GJ	2.70	6.3	$0.12S_1 + 2.08S_2$			
	(lb/MMBtu)	(1.16)	(2.7)	$(0.05SS_1 + 0.895S_2)$			

[a] From U.S. EPA (1998, 1999).

[b] S = weight percent of sulfur in the fuel oil.

[c] Values for post New Source Performance Standard (NSPS).

[d] S_1 = weight percent of sulfur in the fuel oil, S_2 = weight percent of sulfur in natural gas.

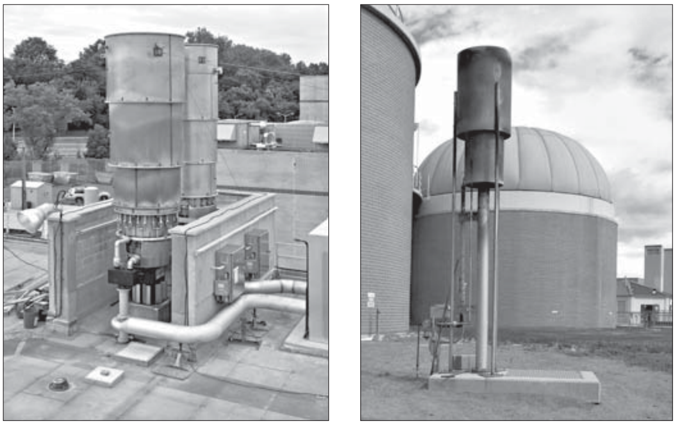


Figure 16–20

Flares for the combustion of excess digester gas: (a) ground effect flare and (b) open air flare.

Similar to the internal combustion systems, the emissions from flaring of digester gas will depend on the completeness of the combustion and the combustion temperature. Digester gas to air ratio should be adjusted according to the quality of the digester gas, but typically 150 to 200 percent of excess air in addition to the stoichiometric air requirement is necessary to achieve sufficient oxidation of methane while minimizing the generation of NOx (IEA Bioenergy, 2000). To minimize emission of CO or NOx, the flare temperature should be controlled between 850 to 1200°C. Technical specification of flaring systems must also meet the Code of Federal Regulations, CFR 40, 60.18, and 60.31.

EXAMPLE 16–4 **Calculation of NO_x Emissions from a Natural Gas-Fired Boiler** Using U.S. EPA's AP42 compilation of air emission factors, estimate NOx emissions from a natural-gas fired boiler. Use the following data to calculate the emissions.

Item	Unit	Basis
Boiler heat input rating	52,753 MJ/h (50.0 MMBtu/h)	Vendor or Manufacturer
Boiler fuel	Natural Gas	
Natural gas heat content	39.1 MJ/m^3 (1050 MMBtu/MMscf)	U.S. EPA AP42[a], Appendix A
Boiler operation	8760 h/y	
Emission factor	1600 kg/m^3 (100 lb/MMscf)	U.S. EPA AP42[a], 1.4–5, Small Boilers, Uncontrolled

[a] Latest version of AP42 sections can be found in: http://www.epa.gov/ttnchie1/ap42/.

Solution

1. Determine the amount of fuel that is combusted over a a period of a year. The amount of fuel is given by

$$\text{Amount of fuel} = (\text{Boiler heat input rating / fuel heat content}) \times \text{h of operation/y}$$

$$(52{,}753 \text{ MJ/h}) / (39.1 \text{ MJ/m}^3) \times 8760 \text{ h/y}$$

$$= 11.82 \times 10^6 \text{ m}^3\text{/y of natural gas consumed}$$

2. Using the emission factor found in U.S. EPA AP42 (see Table 16–15), estimate the anticipated emissions:

Emissions = fuel consumed × emission factor

$$\text{Emissions} = (11.82 \times 10^6 \text{ m}^3/\text{y}) \times (1600 \text{ kg/m}^3) \times (1 \text{ tonne}/10^3 \text{ kg})$$

$$= 18.9 \text{ tonne of NOx emissions/y}$$

Comment As reported in Table 16–15, the emission factor for low NOx burner is 800 kg/10^3 m^3 (50 lb/10^6 ft^3), which is half of uncontrolled small natural gas boilers, and the emission factor is 512 kg/10^3 m^3 (32 lb/10^6 ft^3), less than one-third, with a flue gas recirculation.

16–6 EMISSION OF GREENHOUSE GASES

Recognizing the evidence and projected impacts of the emission of greenhouse gases (GHG), their reduction and/or elimination has become an important element of wastewater management. Assessment of GHG emissions is important in establishing priorities for future capital investments. In planning wastewater treatment facilities, reduction of GHG emissions is often considered in the environmental assessment of the triple bottom line (TBL) approach (see Chap. 18). Because there was no standard protocol for the measurement of GHG emissions until recently and the methodology can affect the outcome of the assessment, it is useful to first review the GHG measurement framework and protocols. After the review of the framework and protocols, opportunities for the reduction of GHG emissions from wastewater treatment plant are discussed.

Framework for Greenhouse Gases Reduction

The United Nations Framework Convention on Climate Change (UNFCCC), first signed in Rio de Janeiro in June 1992, was the first major world-wide recognition of the challenges posed by climate change. The Kyoto Protocol, adopted in 1997, is the first international agreement that set binding targets for industrialized countries towards actions to reduce greenhouse gas emissions. Following the Kyoto Protocol, Intergovernmental Panel on Climate Change (IPCC) issued National Greenhouse Gas Inventories Guidelines (updated most recently in 2006). Protocols developed thereafter to measure the GHG emissions in smaller scales generally follow the greenhouse gases specified in the Kyoto Protocol. In the United States, US EPA issued the Mandatory Reporting of Greenhouse Gases Rule (74 FR 56260) in 2008.

Assessment Protocols

The protocol entitled “GHG Protocol Corporate Accounting and Reporting Standard,” often referred simply as the *GHG Protocol*, developed by the World Resources Institute (WRI) and the World Business Council for Sustainable Development (WBCSD) is consistent with the IPCC Guidelines and accepted widely as the standard protocol to quantify GHG emissions from a specific business(WRI and WBCSD, 2004). In the United States, the Local Government Operations (LGO) protocol was developed based on the *GHG Protocol* in partnership with the California Air Resources Board (ARB), California Climate Action Registry (CCAR), ICLEI-Local Government for Sustainability, in collaboration with the Climate Registry and dozens of stakeholders (ARB et al., 2010). In the LGO protocol, procedures to calculate GHG emissions from government-owned facilities including water and wastewater treatment facilities are identified. The procedure to calculate GHG emissions from a wastewater treatment plant is illustrated in Example 16–4. It should be noted that these protocols are being updated to reflect the latest scientific findings, and latest publications of relevant protocols should be consulted.

Greenhouse Gases. Based on the Kyoto Protocol, the following six greenhouse gases are usually considered in the assessment: carbon dioxide (CO_2), methane (CH_4), nitrous oxide (N_2O), hydrofluorocarbons (HFCs), perfluorocarbons (PFCs), and sulfur hexafluoride (SF_6). Because each greenhouse gas has a different level of impact on retaining heat in the atmosphere, the amount of GHG is usually reported as carbon dioxide equivalent, using global warming potentials (GWPs) determined for each greenhouse gas (see Table 16–16). For wastewater treatment facilities, only CO_2, CH_4 and N_2O are usually assessed because emissions of HFCs, PFCs, and SF_6 are generally negligible.

GHG Emissions Categories. Generally, GHG emissions are categorized in three categories:

Scope 1: All direct GHG emissions
Scope 2: Indirect emissions associated with the consumption of purchased or acquired electricity, steam, heating, or cooling
Scope 3: All other indirect emissions not covered in Scope 2

In the LGO protocol, Scopes 1 and 2 are considered mandatory for reporting, while Scope 3 is stated as optional. In the LGO protocol it is also recommended that CO_2 emissions from the combustion of biomass (or biomass-based fuels including digester gas) be quantified, but not included in the Scope 1 emissions, and reported separately (LGOP, 2008) as “biogenic” emissions.

The protocol, PAS 2050, prepared by the British Standards Institution (BSI), is another approach that can be used for the assessment of greenhouse gas emissions (BSI, 2011). The PAS 2050 protocol is based on the life cycle assessment, and encompasses the boundary to the Scope 3 in the WRI/WBSCD protocol.

Nitrous Oxide Emission. Because of its global warming potential (about 310 times the effect of carbon dioxide), emission of nitrous oxide from wastewater treatment facilities could potentially be significant relative to the emission of other greenhouse gases.

Table 16–16

Fuels used commonly at wastewater treatment plants[a]

Greenhouse gas	Global warming potential (GWP)
Carbon dioxide (CO_2)	1
Methane (CH_4)	21
Nitrous oxide (N_2O)	310
Hydrofluorocarbons (HFCs)	12–11,700
Perfluorocarbons (PFCs)	6500–9200
Sulfur hexafluoride (SF_6)	23,900

[a] From LGO Protocol (2010). GWP values vary for HFCs and PFCs depending on the specific chemical compounds.

Currently, the protocols described above use an approximate estimate of nitrous oxide emission from wastewater treatment facilities based on the nitrogen loading to the plant or population served, types of treatment, and types of receiving waters. The emission is considered as a direct emission from the facility and included in the Scope 1 item. As discussed in Chap. 7, the mechanism of nitrous oxide emission from wastewater treatment facility is highly dependent on the operating conditions, making it difficult to make accurate estimates. Based on the protocol, nitrous oxide emission could make up almost one third, or more, of the total GHG emissions from a wastewater treatment facility.

EXAMPLE 16–5 **Calculation of GHG Emissions from a Wastewater Treatment Plant** Calculate the GHG emissions from the wastewater treatment plant. The treatment plant is designed for BOD removal and nitrification/denitrification, and effluent is discharged to an estuary. The treatment plant serves a combination of residential and industrial/commercial customers. The data required to complete the calculations for the treatment plant were collected and summarized as below. Report Scope 1 and Scope 2 emissions, and report biogenic emissions separately.

Item	Unit	Value
Energy use		
Electricity	kWh/y	14,100,000
Natural gas	m^3/y	17,300
Fuel oil #2	m^3/y	390
Digester gas production	m^3/y	1,047,900
Digester gas used	m^3/y	755,000
Digester gas flared	m^3/y	290,500
Digester gas vented	m^3/y	2400
Plant performance		
Annual average flowrate	m^3/d	100,000
Average influent ammonium concentration	mg/L	18
Average influent total nitrogen concentration	mg/L	32
Average effluent ammonium concentration	mg/L	1.5
Average effluent total nitrogen concentration	mg/L	8.3
Other required information		
Population served	persons	430,000
Methane content in digester gas	%	60
Carbon dioxide in digester gas	%	35
Energy content in natural gas	GJ/m^3	0.0383
Energy content in fuel oil #2	GJ/m^3	38.47
Energy content in digester gas	GJ/m^3	0.0224

(*continued*)

(*Continued*)

Item	Unit	Value
Emission factors		
Electricity	g CO_2e/kWh	720
Natural gas		
CO_2	g/GJ	50,253
CH_4	g/GJ	0.948
N_2O	g/GJ	0.0948
Fuel oil #2		
CO_2	g/GJ	70,100
CH_4	g/GJ	2.844
N_2O	g/GJ	0.569
Digester gas		
CO_2	g/GJ	49,353
CH_4	g/GJ	3.033
N_2O	g/GJ	0.597

Solution

1. Calculate Scope 1 emissions which Include emissions from stationary combustion of natural gas, fuel oil, fugitive emission of digester gas, and emissions associated with wastewater treatment process:
   a. Emissions from natural gas

$$\begin{aligned}\text{CO}_2\text{ emission} &= (\text{Natural gas use, m}^3/\text{y})(\text{energy content, GJ/m}^3)(\text{emission factor})\\ &= (17{,}300\text{ m}^3/\text{y})(0.0383\text{ GJ/m}^3)(50{,}253\text{ g/GJ})(1\text{ tonne}/10^6\text{ g})\\ &= (33{,}297{,}135\text{ g/y})(1\text{ tonne}/10^6\text{ g})\\ &= 33.3\text{ tonne/y}\end{aligned}$$

$$\begin{aligned}\text{CH}_4\text{ emission} &= (\text{Natural gas use, m}^3/\text{y})(\text{energy content, GJ/m}^3)(\text{emission factor})\\ &= (17{,}300\text{ m}^3/\text{y})(0.0383\text{ GJ/m}^3)(0.948\text{ g/CH}_4\text{/GJ})(1\text{ tonne}/10^6\text{ g})\\ &= (628.1\text{g CH}_4/\text{y})(1\text{ tonne}/10^6\text{ g})\\ &= 0.000628\text{ tonne CH}_4/\text{y}\end{aligned}$$

$$\begin{aligned}\text{N}_2\text{O emission} &= (\text{Natural gas use, m}^3/\text{y})(\text{energy content, GJ/m}^3)(\text{emission factor})\\ &= (17{,}300\text{ m}^3/\text{y})(0.0383\text{ GJ/m}^3)(0.095\text{ g N}_2\text{ O/y})(1\text{ tonne}/10^6\text{ g})\\ &= (628.1\text{g N}_2\text{ O/y})(1\text{ tonne}/10^6\text{ g})\\ &= 0.0000628\text{ tonne N}_2\text{O/y}\end{aligned}$$

$$\begin{aligned}\text{Total emissions} &= (\text{CO}_2\text{ emission})\ \text{GWP}_{\text{CO}_2}\\ &\quad + (\text{CH}_4\text{ emission})\ \text{GWP}_{\text{CH}_4}\\ &\quad + (\text{N}_2\text{O emission})\ \text{GWP}_{\text{N}_2\text{O}}\\ &= 33.3 \times 1.0 + 0.000628 \times 21 + 0.0000628 \times 310\\ &= 33.3\text{ tonne CO}_2\text{ e/y}\end{aligned}$$

b. Emissions from fuel oil #2

$$\begin{aligned}\text{CO}_2\text{ emission} &= (\text{Fuel oil use, m}^3\text{/y})(\text{energy content, GJ/m}^3)(\text{emission factor})\\ &= (390\text{ m}^3\text{/y})(38.47\text{ GJ/m}^3)(70{,}100\text{ g CO}_2\text{/GJ})(1\text{ tonne}/10^6\text{ g})\\ &= (1{,}051{,}731{,}330\text{ g/y})(1\text{ tonne}/10^6\text{ g})\\ &= 1051.7\text{ tonne/y}\end{aligned}$$

$$\begin{aligned}\text{CH}_4\text{ emission} &= (\text{Fuel oil use, m}^3\text{/y})(\text{energy content, GJ/m}^3)(\text{emission factor})\\ &= (390\text{ m}^3\text{/y})(38.47\text{ GJ/m}^3)(2.844\text{ kg CH}_4\text{/GJ})(1\text{ tonne}/10^6\text{ g})\\ &= (42{,}669.4\text{ g CH}_4\text{/y})(1\text{ tonne}/10^6\text{ g})\\ &= 0.0427\text{ tonne CH}_4\text{/y}\end{aligned}$$

$$\begin{aligned}\text{N}_2\text{O emission} &= (\text{Fuel oil use, m}^3\text{/y})(\text{energy content, GJ/m}^3)(\text{emission factor,}\\ &\qquad \text{kg-N}_2\text{ O/GJ})\\ &= (390\text{ m}^3\text{/y})(38.47\text{ GJ/m}^3)(0.569\text{ kg N}_2\text{ O/GJ})(1\text{ tonne}/10^6\text{ g})\\ &= (8536.9\text{ g N}_2\text{O/y})(1\text{ tonne}/10^6\text{ g})\\ &= 0.00854\text{ tonne N}_2\text{O/y}\end{aligned}$$

$$\begin{aligned}\text{Total emissions} &= (\text{CO}_2\text{ emission})\,\text{GWP}_{\text{CO}_2}\\ &\quad + (\text{CH}_4\text{ emission})\,\text{GWP}_{\text{CH}_4}\\ &\quad + (\text{N}_2\text{O emission})\,\text{GWP}_{\text{N}_2\text{O}}\\ &= 1050.7 \times 1.0 + 0.0427 \times 21 + 0.00854 \times 310\\ &= 1054.2\text{ tonne CO}_2\text{ e/y}\end{aligned}$$

c. Emissions from digester gas used. Note CO_2 emission from digester gas combustion is considered "biogenic." Biogenic emissions are often not required to be reported as part of the GHG emission but to be reported separately.

$$\begin{aligned}\text{CO}_2\text{ emission} &= (\text{Digester gas use, m}^3\text{/y})(\text{energy content, GJ/m}^3)(\text{emission factor})\\ &= (755{,}000\text{ m}^3\text{/y})(0.0224\text{ GJ/m}^3)(49{,}353\text{ kg-CO}_2\text{/GJ})(1\text{ tonne}/10^6\text{ g})\\ &= (834{,}657{,}936\text{ g/y})(1\text{ tonne}/10^6\text{ g})\\ &= 834.7\text{ tonne/y (biogenic)}\end{aligned}$$

$$\begin{aligned}\text{CH}_4\text{ emission} &= (\text{Digester gas use, m}^3\text{/y})(\text{energy content, GJ/m}^3)(\text{emission factor})\\ &= (755{,}000\text{ m}^3\text{/y})(0.0224\text{ GJ/m}^3)(3.033\text{ kg CH}_4\text{/GJ})(1\text{ tonne}/10^6\text{ g})\\ &= (51{,}294\text{ g CH}_4\text{/y})(1\text{ tonne}/10^6\text{ g})\\ &= 0.0513\text{ tonne-CH}_4\text{/y}\end{aligned}$$

$$\begin{aligned}\text{N}_2\text{O emission} &= (\text{Digester gas use, m}^3\text{/y})(\text{energy content, GJ/m}^3)(\text{emission factor})\\ &= (755{,}000\text{ m}^3\text{/y})(0.0224\text{ GJ/m}^3)(0.597\text{ kg N}_2\text{ O/GJ})(1\text{ tonne}/10^6\text{ g})\\ &= (10{,}096\text{ g N}_2\text{ O/y})(1\text{ tonne}/10^6\text{ g})\\ &= 0.0101\text{ tonne N}_2\text{O/y}\end{aligned}$$

$$\begin{aligned}\text{Total emissions} &= (\text{CO}_4\text{ emission})\ \text{GWP}_{\text{CH}_4}\\ &\quad + (\text{N}_2\text{O emission})\ \text{GWP}_{\text{N}_2\text{O}}\\ &= 0.0513 \times 21 + 0.0101 \times 310\\ &= 4.21\text{ tonne CO}_2\text{ e/y}\end{aligned}$$

$$\begin{aligned}\text{Biogenic emissions} &= (\text{CO}_2\text{ emission})\ \text{GWP}_{\text{CO}_2}\\ &= 834.7 \times 1.0\\ &= 834.7\text{ tonne CO}_2\text{ e/y}\end{aligned}$$

d. Emissions from digester gas flared. In the LOG Protocol, digester gas flaring is assumed to leave 1 percent of the methane gas within the digester gas due to incomplete combustion, and no nitrous oxide emission is assumed from the digester gas flaring. The approach taken by the LOG protocol is followed in this example. Similarly to combusted digester gas, CO_2 emission is considered "biogenic."

$$\begin{aligned}\text{CO}_2\text{ emission} &= (\text{Digester gas flared, m}^3\text{/y})(\text{energy content, GJ/m}^3)(\text{emission factor})\\ &= (290{,}500\text{ m}^3\text{/y})(0.0224\text{ GJ/m}^3)(49{,}353\text{ kg CO}_2\text{/GJ})(1\text{ tonne}/10^6\text{ g})\\ &= (834{,}657{,}936\text{ g/y})(1\text{ tonne}/10^6\text{ g})\\ &= 321.15\text{ tonne/y (biogenic)}\end{aligned}$$

$$\begin{aligned}\text{CH}_4\text{ emission} &= (\text{Digester gas flared, m}^3\text{/y})(\text{methane content})\\ &\quad \times (\text{incomplete combustion})(\text{mass of methane, g/m}^3)\\ &= (290{,}500\text{ m}^3\text{/y})(0.60)(0.01)(656\text{ g/m}^3)(1\text{ tonne}/10^6\text{ g})\\ &= (1{,}143{,}408\text{ g CH}_4\text{/y})(1\text{ tonne}/10^6\text{ g})\\ &= 1.143\text{ tonne CH}_4\text{/y}\end{aligned}$$

$$\begin{aligned}\text{Total emissions} &= (\text{CH}_4\text{ emission})\ \text{GWP}_{\text{CH}_4}\\ &= 1.143 \times 21\\ &= 24.0\text{ tonne CO}_2\text{ e/y}\end{aligned}$$

$$\begin{aligned}\text{Biogenic emissions} &= (\text{CO}_2\text{ emission})\ \text{GWP}_{\text{CO}_2}\\ &= 321.15 \times 1.0\\ &= 321.2\text{ tonne CO}_2\text{ e/y}\end{aligned}$$

e. Emissions from digester gas vented. Of the digester gas vented to the atmosphere, 60 percent is methane, to be reported as Scope 1 emission. Carbon dioxide (35 percent) is not counted in the GHG inventory.

$$\begin{aligned}\text{CH}_4\text{ emission} &= (\text{Digester gas vented, m}^3\text{/y})(\text{methane content})\\ &\quad (\text{mass of methane, g/m}^3)\\ &= (2400\text{ m}^3\text{/y})(0.60)(656\text{ g/m}^3)(1\text{ tonne}/10^6\text{ g})\\ &= (944{,}640\text{ g-CH}_4\text{/y})(1\text{ tonne}/10^6\text{ g})\\ &= 0.945\text{ tonne-CH}_4\text{/y}\end{aligned}$$

$$\begin{aligned}\text{Total emissions} &= (\text{CH}_4\text{ emission})\,\text{GWP}_{\text{CH}_4}\\ &= 0.945 \times 21\\ &= 19.8\text{ tonne CO}_2\text{ e/y}\end{aligned}$$

f. Determine process N_2O emissions from WWTP with nitrification/denitrification. In the LGO Protocol, process N_2O emission from WWTP with nitrification/denitrification is estimated using an emission factor of 7 g-N_2O/person/y. Using the given data, the emission is estimated as

$$\begin{aligned}\text{Process N}_2\text{O emission} &= [(P_{\text{total}} \times F_{\text{ind-com}})EF_{\text{nit/denit}}(1\text{ tonne}/10^6\text{ g})]\text{GWP}_{\text{N}_2\text{O}}\\ &= [(430{,}000 \times 1.25) \times 7 \times 10^{-6}] \times 310\\ &= 1166.4\text{ tonne CO}_2\text{ e/y}\end{aligned}$$

where, P_{total} = total population served by the treatment plant
$F_{\text{ind-com}}$ = factor for industrial and commercial co-discharge waste into the sewer system = 1.25 for WWTPs
$EF_{\text{nit/denit}}$ = emission factor = 7 g-N_2O/person·y
GWP_{N_2O} = global warming potential for N_2O = 310

g. Determine process N_2O emissions from effluent discharge. In the LGO Protocol, process N_2O emission from wastewater discharge is estimated based on the measured average total nitrogen discharge. The emission factor for N_2O from effluent discharge is 0.005 kg N_2O-N/kg-N discharged in the effluent. Using the given data, the N_2O emission is estimated as follows.

$$\begin{aligned}\text{Process N}_2\text{O emission} &= (\text{N load})(EF_{\text{effluent}})(365.25)(44/28)(10^{-6})\text{CWP}_{\text{N}_2\text{O}}\\ &= [(8.3\text{ g-N/m}^3)(100{,}000\text{ m}^3\text{/d})](0.005\text{ kg N}_2\text{O-N/kg-N})\\ &\quad (365.25\text{ d/y})(44\text{ g-CO}_2/28\text{ g-N})(1\text{ tonne}/10^6\text{ g})(310)\\ &= (2{,}381{,}952\text{ g/y})(1\text{ tonne}/10^6\text{ g})(310)\\ &= 738.4\text{ tonne CO}_2\text{e/y}\end{aligned}$$

h. Prepare a summary of Scope 1 emissions.

$$\begin{aligned}\text{Total Scope 1 emissions} &= 33.3 + 1054.2 + 4.21 + 24.0 + 19.8 + 1166.4\\ &\quad + 738.4 = 3040\text{ tonne CO}_2\text{e/y}\end{aligned}$$

$$\text{Biogenic emissions} = 834.7 + 321.2 = 1156\text{ tonne CO}_2\text{e/y}$$

2. Calculate Scope 2 emissions.
Emissions associated with electricity:

$$\begin{aligned}\text{CO}_2\text{ emission} &= (\text{Electricity use, kWh/y})(\text{emission factor, g CO}_2\text{e/kWh})\\ &= 14{,}100{,}000 \times 720\text{ g/y}\\ &= (10{,}152{,}000{,}000\text{ g/y})(1\text{ tonne}/10^6\text{ g})\\ &= 10{,}152\text{ tonne/y}\end{aligned}$$

3. Prepare a summary of the total and biogenic emissions.

$$\text{Total emissions} = \text{Scope 1 emissions} + \text{Scope 2 emissions}$$

$$= 3040 + 10{,}152 = 13192 \text{ tonne } CO_2e/y$$

$$\text{Total biogenic emissions} = 1156 \text{ tonne } CO_2e/y$$

Comment The calculations shown in this example are based on the LGO Protocol, presented in SI units. It should be noted that nearly two-thirds of Scope 1 emissions is attributed to nitrous oxide emissions, and it constitutes about 14 percent of the total emissions.

Opportunities for GHG Reduction at Wastewater Treatment Facilities

Opportunities typically investigated for the reduction of GHG emissions from wastewater treatment facilities are summarized in Table 16–17. Many of the GHG reduction opportunities are related directly to reduction of energy use, discussed further in Chap. 17. Other opportunities are related to shifting the energy sources to those with lower GHG emissions. Examples of opportunities for GHG reductions are discussed briefly in the following subsections. It should be noted that there are numerous options, most of them site-specific, to achieve energy conservation and GHG emission reduction. It is also important to note that a thorough evaluation of existing conditions must be made before options for GHG reduction are evaluated. In conducting the evaluation of existing conditions and opportunities for GHG reduction, latest publications from U.S. EPA, WEF and other sources should be consulted.

Control of Dissolved Oxygen. Aeration of activated sludge reactors is the most significant process element when evaluating energy use as it takes nearly half of the total process-related energy at a typical conventional secondary treatment facility (see Chaps. 4 and 17). Control of dissolved oxygen not only helps reduce energy use, it also helps improve the treatment performance. The use of automated DO control with online DO analyzers has become a common practice. Other process upgrades to improve DO control and associated energy efficiencies include the improvements in blowers and diffusers (see Chap. 5).

Treatment Process Modification. In selecting a treatment process, or planning a process upgrade and modifications, considerations should be given to the treatment processes that require less energy to achieve the treatment objective. For example, by controlling the nitrogen removal process to nitritation and denitritation (see Chap. 7, 8 and 15), oxygen requirements for nitrogen removal could be reduced by 25 percent. Generation and recovery of energy from wastewater and waste energy within the treatment facility is also a significant consideration in GHG reduction as energy recovered from wastewater constituents (e.g., in the form of digester gas) is considered as biogenic and not counted in the Scope 1 and Scope 2 emissions. Greenhouse gas emissions associated with the use of chemicals of fossil fuel sources are usually counted as a Scope 3 emission. However, CO_2 emissions from the use of fossil fuel-origin methanol for denitrification may be counted as process-oriented Scope 1 emission (ICLEI, 2012).

Management of Nitrous Oxide Emission. Emission of nitrous oxide is estimated with assumptions that may not reflect actual emissions from the treatment processes and receiving waters, and it constitutes a significant fraction of total GHG emissions from wastewater treatment facilities. Depending on the mode of operation and

Table 16–17

Considerations in greenhouse gas reduction options for wastewater treatment facilities

Unit process	Description
General	• Selection of lower GHG emission energy sources (e.g., digester gas or natural gas vs. fuel oil) • Waste energy recovery within wastewater treatment plant (see Chap. 17)
Inlet/preliminary treatment	• Pump efficiency improvement • Flow equalization
Primary treatment	• Improvements in solids removal
Secondary treatment and sidestream treatment	• Use of diffusers with high oxygen transfer efficiencies • Use of blowers with high energy efficiencies • Selection of blowers or combination of blowers sized to allow operation in high efficiency range over full range of possible air requirements • Use of energy efficient mixing system for activated sludge anoxic/anaerobic zones • Control of aeration with DO monitoring • Control of aeration with NH_4-N monitoring • Selection of biological process requiring less oxygen (e.g., nitritation/denitritation, partial nitritation/deammonification) (see Chaps. 7, 8, 9 and 15) • Process configuration and control to minimize generation of N_2O (see Chap. 7)
Sludge processing and biosolids utilization/disposal	• Optimization of sludge thickening and dewatering • Elimination of unaccounted and un-combusted methane emissions • Complete utilization of digester gas • Enhancement in digester gas production (see Chap. 13) • Use of waste heat for pre-heating of sludge drying (see Chap. 14) • Heat recovery from sludge incineration systems (see Chap. 14) • Onsite energy and heat production from digester gas and biosolids (see Chaps. 14 and 17)
Disinfection	• Use of high-output lamps for UV disinfection
Advanced treatment	• Use of membrane systems with lower energy requirements • Energy recovery from residual pressure in the membrane treatment system

control, emission of nitrous oxide from biological treatment process could increase significantly. Generation of nitrous oxide during biological treatment is discussed in Sec. 7–12 in Chap. 7.

Use of Digester Gas. Even though the use of digester gas for boilers and power generation has been a common practice at many wastewater treatment facilities, not all facilities with anaerobic sludge digestion utilize all of the digester gas. Because methane has 21 times higher global warming potential than carbon dioxide, one of the major GHG emission sources can be the release of uncombusted digester gas into the atmosphere. By simply flaring the digester gas, the GHG emission is reduced by 21 fold, and because the CO_2 released from the flare is originated from biological sources, it is counted as a "biogenic" emission, which is reported separately, but not included in the GHG emissions inventory. By maximizing the use of digester gas as a fuel source, part of the GHG emissions associated with the use of electricity and fuels from other sources could be avoided. The use of digester gas and management of energy use are discussed further in Chap. 17.

PROBLEMS AND DISCUSSION TOPICS

16–1 Verify that for each mg/L of H_2S removed with chlorine 10.87 mg/L of alkalinity as $CaCO_3$ will be required.

16–2 Determine the amount of hydrogen peroxide (H_2O_2) required for the oxidation of hydrogen sulfide H_2S.

16–3 Using the following half reaction for permanganate (MnO_4^-), estimate the amount of permanganate that would be required per day to oxidize 100 ppm_v of H_2S from an foul air stream with a flowrate of 1500, 2000, 1800 or 2200 m^3/min (flowrate to be selected by instructor).

$$MnO_4^- + 4H^+ + 3e^- \rightarrow MnO_2(s) + 2H_2O$$

16–4 Determine the amount of ferrous sulfate ($FeSO_4$) that would be required to remove 150 mg/L of H_2S from digester supernatant. Assume the sulfide ion in H_2S will be converted to ferrous sulfide in an exchange reaction.

16–5 Four different waste air streams have been sampled and the results are summarized below. For one of these waste air streams (to be selected by instructor), determine the chemical requirements. Sodium hypochlorite and sodium hydroxide are to be used in the chemical scrubber.

		Plant			
Item	Unit	1	2	3	4
Air waste stream flowrate	m^3/min	1000	2500	3200	1800
H_2S concentration	ppm_v	75	45	65	35
Liquid to gas ratio	kg/kg	1.85	2.0	2.1	1.9
Temperature	°C	28	33	30	25
Density of 50 percent NaOH solution	kg/L	1.52	1.52	1.52	1.52

16–6 Using the design criteria given in Table 16–11, determine the size of compost filter needed to scrub 65 ppm_v H_2S from foul air at a flowrate of 1500, 1880, 2100 or 2300 m^3/min (value to be selected by instructor). Also estimate the mass of the buffer compound needed to neutralize the acid formed as a result of treatment within the filter. Assume a packed bed porosity of 43 percent. The temperature of the foul air is 20°C.

16–7 Using U.S. EPA's AP42 compilation of air emission factors, estimate (CO, NOx or SO_2 emissions, to be selected by instructor) from a dual-fuel reciprocating engine. The data used to calculate the emissions are summarized below and emission factors are reported in Table 16–15.

Item	Unit
Engine power rating	2386 kW (3200 bhp)
Fuel	Fuel oil #2, natural gas
Fuel oil #2 heat content	38.47 GJ/m^3
Natural gas heat content	0.0383 GJ/m^3
Average load by fuel oil	35 percent
Engine operation	8640 h/y

16–8 In Example 16–4, part of the digester gas was flared and vented. Consider the situation where all unutilized digester gas in Example 16–4 is utilized to reduce natural gas consumption, assuming the equipment at the treatment plant is capable to switch the use between natural gas and digester gas. Calculate: (a) the amount of natural gas consumption saved per year, and (b) reduction in the total GHG emissions.

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17

Energy Considerations in Wastewater Management

WORKING TERMINOLOGY

Term	**Definition**
Biogas	A generic term used for the gas produced during anaerobic treatment of organic compounds. Typical composition of biogas generated by anaerobic digestion of sludges is 60 to 65 percent methane, and approximately 30 percent carbon dioxide. The term is used interexchangeably with digester gas.
Bomb calorimeter	A device in which a known mass of sample is combusted and the released energy is measured.
Carnot cycle	A reversible power cycle consisting of two isothermal processes and two isentropic processes without internal heat transfer.
Coefficient of performance (COP)	A ratio of the energy extracted from an energy recovery process to the energy input to the energy recovery process such as heat pump.
Convection	Heat transferred by the movement of gas or liquid.
Combined heat and power (CHP)	A system used to generate both electricity and heat. Also called cogeneration.
Electrical efficiency	The percentage of energy that is converted into electricity relative to the total energy input.
Enthalpy	A thermodynamic property of a system expressed as the sum of the internal energy and the product of pressure and volume.
Exothermic reaction	A chemical reaction resulting in a release of energy from the molecules.
Fuel cell	A device used to generate electricity from the reaction of hydrogen and oxygen gas. In wastewater treatment facilities, methane in biogas can be used to generate hydrogen. Recoverable thermal energy can also be generated with a fuel cell.
Heat exchanger	A device that allows the extraction of heat through a refrigerant with external power input.
Heat pump	A device used to transfer heat from one body to another using a refrigerant and energy from an external power source.
Higher heating value (HHV)	The total amount of energy released by complete combustion of a unit mass of a substance. Latent heat of vaporization is counted in HHV.
Integrated resource recovery (IRR)	A management approach in which the recovery of energy, water, and other resources is achieved by integrating the management of various waste streams such as wastewater, solid waste, and others.
Latent heat	Heat released or absorbed during the change in phases of a chemical substance (e.g., from liquid to gas).
Lower heating value (LHV)	The total amount of energy released by complete combustion of a unit mass of a substance, less the latent heat of vaporization of the water vapor formed by the combustion.

Term	Definition
Organic Rankine process	A process to generate electricity using low grade heat such as waste heat from a stationary combustion system and with an organic fluid that changes phases at a lower temperature than water.
Sensible heat	Energy transferred to a substance that results in a change in temperature of the substance.
Specific heat	The amount of heat necessary for the temperature of a unit mass of a material to increase by one unit.
Syngas	Syngas, or synthesis gas, is a mix of hydrogen, carbon monoxide, and other gaseous compounds generated as a result of the incomplete oxidation of organic materials under restricted oxygen environment. Syngas can be used as a fuel gas or to generate liquid fuel through the Fischer-Tropsch process.
Thermal efficiency	The ratio of net usable thermal output to the energy input, expressed in percentage. Also known as total system efficiency.

In the past, due to the relatively low and stable cost of fossil fuels and electricity, the use, recovery, and management of energy was not typically emphasized in the design and operation of wastewater management facilities. With increasing energy costs, uncertainties about future fossil fuel supplies, and increasing awareness of the impacts of greenhouse gas emissions, the efficient management of energy is now of greater concern with both private and public entities. Recognizing the importance of energy in the implementation of wastewater treatment facilities, and opportunities to recover and utilize energy from various sources within the treatment facilities, the focus of this chapter is on the subject of energy recovery and utilization. Topics considered in this chapter include: (1) need for energy management, (2) energy in wastewater, (3) fundamentals of heat balance, (4) energy usage in treatment plants, (5) recovery and utilization of chemical energy (6) recovery and utilization of heat energy, (7) recovery and utilization of hydraulic potential energy, (8) energy management, and (9) future opportunities for alternative wastewater treatment processes. The recovery of nutrients is considered in Chaps. 14 and 15.

17–1 FACTORS DRIVING ENERGY MANAGEMENT

The principal driving forces for achieving more efficient management of energy in wastewater treatment are

1. Potential for energy cost savings, including opportunity to become a net energy supplier
2. Potential for improved energy supply reliability
3. Considerations for sustainability, including the greenhouse gas reduction goals put forth by local, state, and federal governmental agencies

Potential for Energy Cost Savings

The operation of wastewater treatment facilities, as discussed in Chap. 4, depends largely on the use of energy resources to bring about various reactions. Therefore it is important to appraise energy requirements to better manage the energy usage, which constitutes the second largest expenditure in the operation of wastewater treatment facilities after labor costs. Examples of energy cost saving opportunities include the use of energy efficient equipment, process control for optimized energy use, and selection of energy sources including pricing negotiation.

Energy Supply Reliability

Reliability of energy supply for wastewater treatment facilities is an important consideration because of the potential for unforeseeable events such as an area-wide blackout of electrical power supply and disruption in energy supply after natural disasters. Generally, wastewater treatment facilities are equipped with emergency generators to operate critical elements of the treatment facility during disruption of energy supply, but few treatment plants are able to operate the entire treatment process with the emergency power supply. In recent years, it has been recognized that wastewater theoretically contains more energy than that required for treatment. It is also recognized that wastewater treatment plants could become net exporters of energy if the energy contained in incoming wastewater could be recovered effectively. Becoming self sufficient in the production and utilization of energy would significantly improve treatment plant reliability, even during power outages. As discussed in this chapter, however, inherent inefficiencies with energy recovery process are significant challenges. The use of anaerobic digestion and combustion of biogas, for example, usually achieves one third or less of typical energy requirements for the conventional wastewater treatment facility. Some of the technologies for energy recovery and utilization considered in this chapter are relatively new to the wastewater field, but many of them have been used in various industrial and commercial applications.

Considerations for Sustainability

In addition to the need to reduce costs for purchasing energy from outside sources, the reduction of greenhouse gas (GHG) emissions has become one of key factors affecting decisions on wastewater treatment processes and equipment selection. As discussed in Sec. 4–1, many existing treatment plants built during the 1970s and 1980s will need to be upgraded to meet increasingly stringent discharge limits. In planning treatment plant upgrades, reduction of GHG emissions (see Chap. 16) is often highlighted in the objectives of the project. Even though justifications for funding tend to focus on the immediate cost savings, when similar alternatives are compared, the options which address GHG reduction goals and other environmental impacts are considered favorably. As wastewater treatment facilities are considered more frequently to be an integral part of a regional energy management scheme, various alternative organic wastes such as food waste and grease, could be brought to the treatment plant to enhance energy production (Chap. 14).

17–2 ENERGY IN WASTEWATER

The energy contained in wastewater is comprised of (1) chemical energy, (2) thermal energy, and (3) hydraulic energy. Chemical energy is the energy contained in organic molecules which can be released by chemical reactions. Thermal energy is the heat retained in wastewater. Hydraulic energy of wastewater fluid is the sum of the gravitational potential energy due to elevation head, energy associated with pressure head, and kinetic energy embodied in the wastewater flows as velocity head. Each of the three forms of energy in wastewater is considered in the following discussion.

Chemical Energy

Wastewater contains organic and inorganic molecules, and exothermic reactions of these constituents will result in a release of chemical energy retained in the molecules. A majority of the chemical energy in wastewater is contained in organic compounds measured as COD, even though some inorganic constituents including ammonia also contain chemical energy that could be extracted. In wastewater treatment, part of the chemical energy is removed from the liquid stream in the form of sludge during preliminary and primary treatment. During the biological treatment process, some of the chemical energy is transformed into biomass and reaction products such as carbon dioxide and methane, or released as heat through metabolism of microorganisms. In sludge processing, part of the chemical energy may be recovered in the form of methane gas and utilized as an energy source.

Energy Contents of Wastewater Constituents. The energy content of wastewater, as described in Chaps. 2 and 14, can be estimated from an elemental analysis of the organic constituents in wastewater using the following empirical expression, which is a modified form of the DuLong formula developed by Channiwala (1992), repeated here from Chap. 2 for convenience.

$$\text{HHV (MJ/kg)} = 34.91\,\text{C} + 117.83\,\text{H} - 10.34\,\text{O} - 1.51\,\text{N} + 10.05\,\text{S} - 2.11\text{A} \tag{2–66}$$

Where C, H, O, N, S, and A are the weight fraction of carbon, hydrogen, oxygen, nitrogen, sulfur, and ash, respectively. Note that higher heating value (HHV) is the total amount of energy released by complete combustion of a unit mass of a substance when water remains as liquid. In actual combustion systems, the combustion temperature is well above 100ºC and water will be vaporized, absorbing the latent heat of vaporization. The total amount of energy released from a complete combustion of unit mass of a substance, considering the heat required for vaporization of water is the lower heating value (LHV). For example, LHV of natural gas is typically 10 percent lower than the HHV. While Eq. (2–66) can be used to estimate the energy content, experimental data on chemical energy in wastewater samples are collected using a bomb calorimeter.

For simple molecules, chemical energy can be calculated based on the enthalpy of reaction. The enthalpy of reaction is defined by Hess' Law as the difference between the sum of enthalpy of formation of reactants and the sum of enthalpy of formation of all products:

$$\text{H}_{\text{reaction}} = \sum \text{H}^{o}_{f\,\text{products}} - \sum \text{H}^{o}_{f\,\text{reactants}} \tag{17–1}$$

where $\text{H}_{\text{reaction}}$ = enthalpy of reaction
H^{o}_{f} = enthalpy of formation

The reaction is an exothermic reaction when the enthalpy of reaction is negative. Enthalpy of formation of chemical compounds found commonly in wastewater is shown in Table 17–1, and a more complete list can be found in references in chemical engineering. Calculation of the heat of reaction is illustrated in Example 17–1.

EXAMPLE 17–1 **Calculation of Enthalpy of Reaction** Calculate the enthalpy of reaction for the oxidation of methane.

The enthalpy of formation for each reactant and product at 25ºC is

$CH_{4(g)}$	−74.6 kJ/mole
$O_{2(g)}$	0 kJ/mole
$CO_{2(g)}$	−393.5 kJ/mole
$H_2O_{(l)}$	−285.8 kJ/mole

Solution

1. The oxidation of methane can be expressed as follows.

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$$

2. Determine the enthalpy for the oxidation of methane using Eq. (17–1).

$$H_{\text{reaction}} = \sum H^o_{f\,\text{products}} - \sum H^o_{f\,\text{reactants}}$$

$$H_{\text{reaction}} = [(H^o_{fCO_2}) + 2(H^o_{fH_2O})] - [(H^o_{fCH_4}) + 2(H^o_{fO_2})]$$

$$= [(-393.5) + 2(-285.8)] - [(-74.6) + 2(0)]$$

$$= -890.5 \text{ kJ/mole}$$

Comment Because the enthalpy of reaction was calculated for 25°C, water is in the liquid phase and the calculated value is the higher heating value at 25°C. The volume of one mole of methane at 25°C at atmospheric pressure is approximately 24L. Assuming 65 percent methane content by volume in digester gas, 1 m^3 of digester gas contains approximately 27 moles of methane. Thus, the HHV in digester gas can be estimated as 890 (kJ/mole) × 27 ($mole/m^3$) = 24,112 kJ/m^3 (~647 Btu/ft^3). Stationary combustion systems are operated at a much higher temperature and the latent heat of water vaporization must be accounted for by estimating the lower heating value.

Table 17–1

Enthalpy of formation for selected chemical compounds at 25°C[a]

Substance	State[b]	ΔH^0_f, kJ/mole	Substance	State[b]	ΔH^0_f, kJ/mole
Ca^{2+}	aq	−542.8	H_2O	g	−241.8
$CaCO_3$	s	−1206.87	HS	aq	−17.6
$Ca(OH)_2$	s	−986.6	H_2S	g	−20.6
$CaSO_4$	s	−1434.5	H_2S	aq	−39.3
CH_4	g	−74.6	H_2SO_4	l	−814
CH_3CH_3	g	−84.67	Mg^{2+}	aq	−466.9
CH_3COOH	aq	−488.4	$Mg(OH)_2$	s	−924.5
CH_3COO^-	aq	−486.0	Na^+	aq	−240.1
$C_6H_{12}O_6$	s	−1275	NH_3	g	−45.9
Cl_2	g	0	NH_3	aq	−80.83
Cl_2	aq	−23.4	NH_4^+	aq	−132.5
Cl^-	aq	−167.20	NO_2^-	aq	−104.6
CO_2	g	−393.51	NO_3^-	aq	−207.4
CO_2	aq	−412.92	O_2	g	0
CO_3^{2-}	aq	−677.10	O_2	aq	−11.71
HCO_3^-	aq	−692.0	OH^-	aq	−230.0
H_2CO_3	aq	−699.0	S^{2-}	aq	30.1
H_2O	l	−285.8	SO_4^{2-}	aq	−909.3

[a] From Sawyer et al. (2003).

[b] g = gas, aq = aqueous solution, l = liquid.

For larger and more complex molecules in wastewater, theoretical values cannot be calculated with Eq. (17–1) because the enthalpy of formation values for many of the reactants are not available. There have been attempts to find correlations between COD and chemical energy in wastewater, and the values ranging between 14.7 and 17.8 kJ/gCOD have been reported for untreated domestic wastewater (Shiraz and Bagley, 2004; Heidrich et al., 2011). However, as described by Heidrich et al. (2011), the amount of energy per gram COD can vary widely. Typical energy content in primary sludge has been reported to be between 23,000 and 29,000 kJ/kg dry solids (see Table 14–17 in Chap. 14). Determination of energy in primary sludge is illustrated in Example 17–2.

EXAMPLE 17–2 **Chemical Energy in Sludge Removed by Primary Clarification** The flowrate to a wastewater treatment plant is 1000 m^3/d. The average total suspended solids concentration in the wastewater is 720 g/m^3. Using the typical energy contents in primary sludge reported in Table 14–17, estimate the chemical energy removed by primary clarification assuming 50 percent solids removal.

Solution

1. Calculate TSS loading.

   $$\text{TSS loading} = 1000\ \text{m}^3/\text{d} \times 720\ \text{g/m}^3 = 720{,}000\ \text{g/d} = 720\ \text{kg/d}$$

2. Estimate the amount of chemical energy removed by primary clarification.

   $$\text{TSS removed} = (720\ \text{kg/d}) \times 0.5 = 360\ \text{kg/d}.$$

   Chemical energy in the primary sludge is 23,000 to 29,000 kJ/kg. Assuming 26,000 kJ/kg,

   $$\text{Chemical energy removed} = (360\ \text{kg/d}) \times (26{,}000\ \text{kJ/kg}) = 9{,}360{,}000\ \text{kJ/d}$$
   $$= 9.36\ \text{GJ/d}$$

Comment The amount of energy that can be recovered and utilized will depend on the solids processing, energy recovery and energy use efficiencies, as discussed later in this Chapter. It should be noted that the loss of dissolved and volatile organic compounds during sample preparation have been recognized and high variability in values between samples have been reported (Shiraz and Bagley, 2004; Heidrich et al., 2011).

Energy Content of Ammonia. As described in Sec. 15–5, thermal oxidation of ammonia is an exothermic reaction, and combustion of ammonia will release energy. The heat balance based on the heat of formation at 25°C at 1 atm is expressed as:

$$NH_3 + 0.75\ O_2 \rightarrow N_2 + 1.5\ H_2O \qquad \Delta H^\circ = -317\ \text{kJ/mole} \qquad (17\text{–}2)$$

$$NH_3 + O_2 \rightarrow 0.5\ N_2O + 1.5\ H_2O \qquad \Delta H^\circ = -276\ \text{kJ/mole} \qquad (17\text{–}3)$$

$$NH_3 + 1.25\ O_2 \rightarrow NO + 1.5\ H_2O \qquad \Delta H^\circ = -227\ \text{kJ/mole} \qquad (17\text{–}4)$$

Anhydrous ammonia has been used as a fuel in the past and is currently the subject of extensive research as a potential alternative fuel source. Even though the heat content is significantly lower than that of methane (802.6 kJ/mole or 50,163 kJ/kg), there is a potential to recover ammonia from nitrogen-rich waste streams and use the energy to supplement other fuel sources.

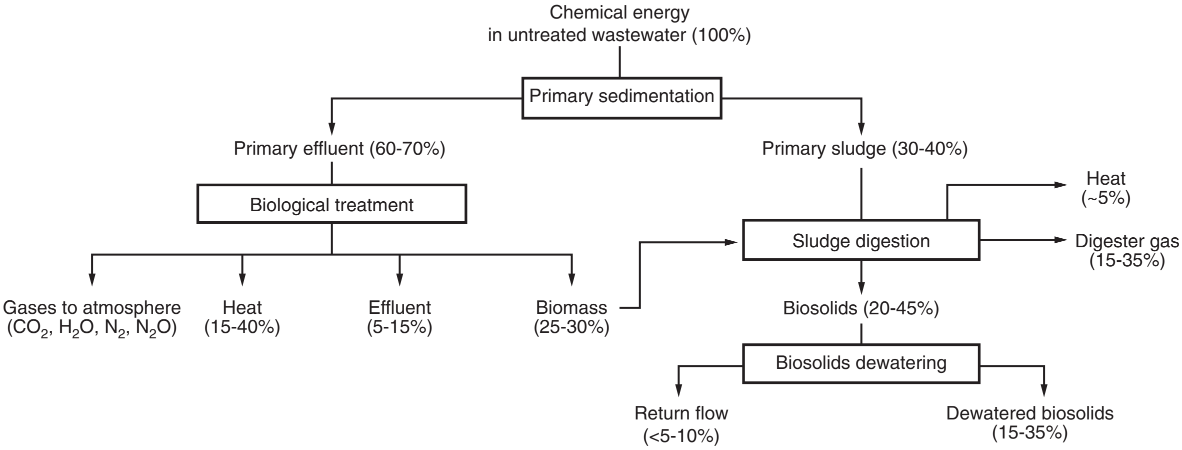


Figure 17–1

Fate of chemical energy in wastewater treatment with activated sludge and anaerobic sludge digestion. A range is shown for the percentages as they vary with the treatment technologies and wastewater characteristics.

Distribution of Chemical Energy during Wastewater Treatment. The distribution and fate of chemical energy in conventional wastewater treatment is illustrated conceptually on Fig. 17–1. Transformation of chemical energy occurs primarily during two major treatment processes: biological treatment of liquid stream, and treatment and processing of sludge. Even though each wastewater has different characteristics and distribution of energy varies with treatment processes, it is worth noting that a considerable amount of chemical energy is sent to the biological treatment process where additional energy is added to the process to convert chemical energy into CO_2, H_2O, N_2, N_2O, heat, and other byproducts that cannot be utilized as an energy source.

Heat generated from exothermic reactions can be significant in processes such as autothermal aerobic digestion (ATAD) and sidestream treatment involving oxidation and reduction of highly concentrated organic and nitrogenous compounds. In these processes, a heat balance must be prepared to determine if cooling of the treatment unit process or dilution of the treated stream is necessary (see Chap. 15) to maintain a desired range of operating temperature. However, in normal domestic wastewater, the heat generated from biochemical reactions is not significant enough to require heat balance evaluation.

Part of the chemical energy is retained in biomass, which can be transformed into an energy source such as biogas and syngas through sludge processing (see Chaps. 13 and 14). With a conventional treatment system, only a fraction of chemical energy that reaches solids processing can be recovered. Historically, wastewater treatment systems were not designed with the intent to maximize the flow of chemical energy to the processes capable of energy recovery.

Thermal Energy

The thermal energy in wastewater is in the form of temperature. When the temperature of a liquid or gas changes from T_1 to T_2, the change in thermal energy is expressed as:

$$Q = mc\Delta T \qquad (17–5)$$

where, Q = change in heat content in wastewater flowrate, kJ/h

m = mass flowrate of water, kg/h

c = specific heat of a substance, kJ/kg·°C

ΔT = temperature change, °C

Thermal energy can be recovered from water (e.g., wastewater effluent) or heated air, such as exhaust from unit processes involving combustion of fuel (e.g., engine generators, boilers, incinerators). It is common to utilize the excess heat in air and/or water for various uses within the treatment facility including digester heating, solids drying, hot water supply, and space heating. Excess heat can also be provided to users outside of the facility, if the amount of heat produced at the plant is large enough.

When the temperature of the wastewater is significantly different from the ambient temperature or the temperature of other streams at the treatment facility, there is a potential that heat can be transferred in a form that could be used for other purposes. Heat added to (or lost from) wastewater can be approximated by assuming that the specific heat of wastewater is equal to that of water at the given temperature (see Chap 2). The specific heat of water, c_w, is 4.1816 kJ/kg·°C at 20°C, and even though it varies with temperature the difference is within 1 percent over the range between 0 to 100°C. For example, if the temperature of wastewater is raised from 21 to 23°C and the flowrate is 160,000 L/h, the heat added to wastewater is

$$\begin{aligned} Q &= mc_w\Delta T \\ &= 160{,}000 \times 4.18 \times 2 \\ &= 1{,}337{,}600\,\text{kJ/h} \end{aligned}$$

Hydraulic Energy

Exclusive of chemical and heat energy, wastewater fluid can also contain energy in the form of elevation head, h_e, (the relative position of the influent to effluent free water surface), pressure head, h_p, (as in pressurized processes such as reverse osmosis) and, velocity head h_v, (associated with the kinetic energy of the moving fluid). These forms of energy are usually quantified in terms of the Bernoulli equation. It should be noted that each of the three terms represents a linear dimension. The energy involved, expressed in kJ or other appropriate units, is obtained by taking into account the corresponding mass of the fluid. The power, expressed in W, kJ/h, or other appropriate units, is the energy per unit time.

Elevation Head, h_e. Most conventional wastewater treatment plants are designed to allow wastewater to flow by gravity from the headworks to the receiving waters. To minimize the power requirements, the hydraulic profile is set to minimize the excess head at the end of the treatment process. Because the majority of wastewater treatment plants are located adjacent to the receiving water body, the plant is often designed to have minimal head at the discharge point.

Pressure Head, h_p. Some wastewater treatment processes such as reverse osmosis operate under pressurized conditions. The pressure head is expressed as p/γ, where γ is a specific weight of the fluid. The recovery of energy from reverse osmosis is considered in Sec. 11–7 in Chap. 11.

Velocity Head, h_v. Because the wastewater is moving through the plant, it also contains kinetic energy, expressed as $v^2/2g$. In general, the velocity head contribution to the total energy is relatively small.

Determination of Total Fluid Head. If H_t, represents the total head transferred to (+) or from (−) the fluid (e.g., in a pump, fan, or turbine) then application of the law of conservation of energy between any two points can be expressed as follows:

$$(h_e + h_p + h_v)_1 \pm H_t = (h_e + h_p + h_v)_2 + \text{losses} \tag{17–6}$$

The losses in Eq. (17–6) represent the head that has been transformed into nonrecoverable forms of energy (e.g., heat or noise). The general expression for an incompressible liquid may be rewritten as

$$\pm H_t = (p_2/\gamma - p_1/\gamma) + (v_2^2/2g - v_1^2/2g) + (z_2 - z_1) + h_L \tag{17–7}$$

where H_t = total head transferred to/received from fluid, m (ft)
p_1, p_2 = pressure, kN/m^2 (lb$_f$/in.^{2})
γ = specific weight of water, kN/m^3 (lb/ft^3)
v = velocity of water, m/s (ft/s)
g = acceleration due to gravity, 9.81 m/s^2 (32.2 ft/s^2)
z_1, z_2 = height above any assumed datum plane, m (ft)
h_L = headloss, m (ft)

Conversion of Fluid Potential Energy to Other Forms of Energy. The fluid energy as given by Eq. (17–7) can be converted to electrical power for a given flowrate using the following expression:

$$P_e = \rho Q g H_t \eta_t \eta_e \tag{17–8}$$

where, P_e = electrical power obtained, W
ρ = density of wastewater, kg/m^3
Q = flowrate, m^3/s
g = acceleration due to gravity, 9.81 m/s^2
η_t = efficiency of mechanical device (e.g., Pelton Wheel, reverse pump, pump, etc.) expressed as a fraction, dimensionless
η_e = efficiency of electrical conversion device expressed as a fraction, dimensionless

Efficiency factors for power generation systems are given in Sec. 17–7. The application of Eqs. (17–7) and (17–8) is demonstrated in Example 17–3.

EXAMPLE 17–3 **Determine Hydraulic Energy of Wastewater Discharge** Consider a wastewater treatment facility located near a coastline, with an effluent discharge point 3 m above mean sea level and an average effluent flowrate of 4 ML/d. Treated effluent flowing out of the chlorine contact tank has a velocity of 0.5 m/s. Calculate the potential energy contained in the effluent. Estimate the actual electrical energy output that can be obtained if a hydraulic turbine generator is located at sea level. Assume a turbine generator combination is to be used with an overall efficiency on 40 percent, The wastewater temperature is 20°C.

Solution

1. Determine the potential and velocity energy.
   a. The potential energy is equal to

$$(z_2 - z_1) = [0 - 3\text{ m})] = -3\text{ m}$$

   b. The velocity energy is equal to

$$(v_2^2/2g - v_1^2/2g) = 0 - (0.5\text{ m/s})^2/2g = -0.0127\text{ m}$$

2. Calculate the potential mechanical energy that can be transferred using Eq. (17–7) neglecting any losses.

$$H_t = (z_2 - z_1) + (v_2^2/2g - v_1^2/2g) = -3\text{ m} + (-0.0127)\text{ m} = -3.0127\text{ m}$$

Because the sign is negative, energy can be produced.

3. Calculate the electrical energy that can be produced with a turbine generator combination using Eq. (17–8) with a combined efficiency factor η and neglecting any other losses.

$$\begin{aligned} P_e &= \rho Q g H_t \eta \\ &= (1000\ \text{kg/m}^3)(4000\ \text{m}^3/\text{d})(9.81\ \text{m/s}^2)(3.0127\ \text{m})(0.40) \\ &= 47.3 \times 10^6\ \text{kg·m}^2/\text{s}^2\text{·d} \\ &= 47.3\ \text{MJ/d} \end{aligned}$$

Comment Depending on the receiving water, the hydraulic potential that could be made available for power generation may vary with tides, seasons or other factors altering the surface level. Recovery of hydraulic potential is further discussed in Sec. 17–7.

17–3 FUNDAMENTALS OF A HEAT BALANCE

In the following sections, the various forms of energy utilized in wastewater treatment and the management of that energy usage in wastewater treatment facilities are introduced and discussed. To understand the issues involved, it will be useful to review the basic concept involved in the preparation of heat balances.

Concept of a Heat Balance

The concept of heat balance is based on the first law of thermodynamics, i.e., enthalpy is conserved. The mathematical approach is similar to the mass balance calculation described in Chap. 1. For a given system boundary, the general heat balance analysis is given by:

1. General word statement:

$$\underset{(1)}{\begin{array}{c}\text{Rate of accumulation}\\ \text{of heat within the}\\ \text{system boundary}\end{array}} = \underset{(2)}{\begin{array}{c}\text{rate of flow of}\\ \text{heat into the}\\ \text{system boundary}\end{array}} - \underset{(3)}{\begin{array}{c}\text{rate of flow of}\\ \text{heat out of the}\\ \text{system boundary}\end{array}} + \underset{(4)}{\begin{array}{c}\text{rate of generation}\\ \text{of heat within the}\\ \text{system boundary}\end{array}} \tag{17–9}$$

2. The corresponding simplified word statement is

$$\text{Accumulation} = \text{inflow} - \text{outflow} + \text{generation} \tag{17–10}$$

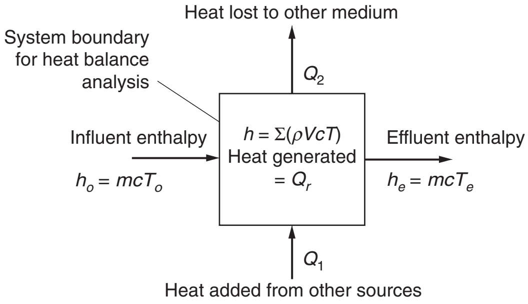
3. Assuming there is no change in mass flowrate and specific heat of the mass, symbolic representation of heat balance is (see Fig. 17–2):

$$\Delta H = mcT_o + Q_1 - mcT_e - Q_2 + Q_r \tag{17–11}$$

where ΔH = change of enthalpy within the system boundary
m = influent mass flowrate
c = specific heat of the mass
T = temperature
Q_1, Q_2 = heat added to or lost from the system boundary, such as added energy through mechanical mixing, or heat loss through the reactor wall
Q_r = heat generated/absorbed within the system boundary, such as heat generated /absorbed during due to chemical reaction and latent heat of vaporization

Figure 17–2

Conceptual diagram of heat balance.



Preparation of a Heat Balance

The preparation of a heat balance is similar to the preparation of a mass balance in that all sources of heat that enter, exit, or are released or taken up within the system boundary must be considered. In most cases, however, internal energy within chemical compounds is not considered except for the energy released into the system in the form of thermal energy. The steps given below should be followed in the preparation of heat balances, as the techniques involved are being mastered.

1. Prepare a simplified schematic or flow diagram of the system or process.
2. Draw a system boundary to define the limits over which the heat balance is to be applied. Proper selection of the system boundary is important because, in many situations, it may be possible to simplify the heat balance computations.
3. List all of the pertinent data and assumptions that will be used in the preparation of the heat balance on the schematic or flow diagram.
4. List all of the rate expressions for the biological or chemical reactions that occur in the process.
5. Select a convenient basis on which the numerical calculations will be based.
6. Solve the balance equations.

The application of the heat balance calculation is illustrated in Example 17–4.

EXAMPLE 17–4 **Heat Balance Calculation** Evaluate the heat balance for an anaerobic digester, using the following process information. How much additional heat will be required to maintain the digester temperature? Assume that the sludge in the reactor is well mixed and has a uniform temperature and specific heat. Ignore heat in the digester gas.

Process information	Unit	Value
Incoming sludge flowrate	m^3/d	100
Incoming sludge temperature	°C	10
Volume of sludge in the digester	m^3	2000
Liquid temperature in the digester	°C	32
Specific heat of sludge in the digester	kJ/kg·°C	4.2
Heat loss by conduction[a]	kJ/d	1.9×10^6
Heat added by sludge mixing	kJ/d	negligible

[a] Conduction losses occur through the digester walls above and below the ground level, the bottom, and the top, depending on the design of the digester.

Solution

1. Describe the heat balance.
   Accumulation/loss of heat = (heat in incoming sludge) + (heat added by sludge mixing) − (heat in effluent sludge) − (heat loss by conduction through the wall)
2. Develop a symbolic representation from Eq. (17–3).

$$\Delta H = mcT_o + Q_1 - mcT_e - Q_2 + Q_r$$

3. Using the given data and assuming specific gravity of water is 1.0, solve for ΔH.

$$\begin{aligned}\Delta H &= (100\,\text{m}^3/\text{d})(10^3\,\text{kg/m}^3)(4.2\,\text{kJ/kg})(10^\circ\text{C}) + 0 - (100\,\text{m}^3/\text{d})(10^3\,\text{kg/m}^3)\\ &\quad (4.2\,\text{kJ/kg})(32^\circ\text{C}) - (1.9 \times 10^6\,\text{kJ/d}) + 0\\ &= 4.2 \times 10^6 - 13.44 \times 10^6 - 1.9 \times 10^6\,\text{kJ/d}\\ &= -11.14 \times 10^6\,\text{kJ/d}\end{aligned}$$

To maintain the heat balance, an additional 11.14×10^6 kJ/d, or 11.14 GJ/d of heat must be added to the digester.

Comments For simplicity, heat in the digester gas was not included in the calculation. Even though a single number was assumed in this example to account for conduction losses, evaluation of the heat loss from the digester tank is a significant part of the heat balance calculation around the digester (see Chap. 13).

In normal municipal wastewater treatment, the heat balance around the liquid treatment train is important primarily to maintain temperature for biological treatment. However, it is sometimes observed that the temperature of mixed liquor is slightly higher than the incoming wastewater. The elevated temperature is generally attributed to biological oxidation of organics and nitrogen which is an exothermic reaction as discussed in Sec. 17–5 and Sec. 15–6 in Chap. 15. The aeration system will also contribute to the heat balance in bioreactors as compressed air, at an elevated temperature, is introduced into the bioreactor. Mixing in open activated sludge tanks can add heat (if done with the aeration system), but mixing also enhances the removal of heat liquid by evaporative cooling as moist air exits from the reactor. In sidestream treatment, heat generated from oxidation and reduction of nitrogen is significant and the heat balance must be evaluated to maintain the operating temperature with dilution water or a heat removal mechanism (see Sec. 15–11). Excess heat in the incoming wastewater and treated effluent can be recovered for beneficial purposes as discussed in Sec. 17–6.

17–4 ENERGY USAGE IN WASTEWATER TREATMENT PLANTS

The energy contained in wastewater and the fundamentals of the heat balance used to assess the potential for the recovery and utilization of energy were introduced in the previous sections. To achieve efficient use of energy and utilization of energy extracted from the treatment facility, it is useful to review the usage of energy in wastewater treatment plants. The review of energy usage in wastewater treatment plants is also important because the cost for energy ranges between 15 and 40 percent of the total operation and maintenance costs for wastewater treatment (WEF, 2009), the second highest after labor costs. The topics considered in this section include (1) types of energy sources, (2) energy use for wastewater treatment, (3) energy usage by various treatment processes, and (4) advanced and new wastewater treatment technologies.

Types of Energy Sources Used at Wastewater Treatment Facilities

The energy required to operate a wastewater treatment facilities is supplied by a number of different sources, but primarily by electricity. Electricity is used to operate motors for blowers, pumps and other facilities with moving parts. Electricity is also used for instrumentation and control equipment, lighting, and cooling and heating of buildings with air conditioning equipment. Fuel oils and natural gas are used primarily to operate boilers for heating, and to operate a stationary combustion system to produce electricity. Emergency generators are typically operated by fuel oils. The heat generated from the combustion system can also be utilized (combined heat and power, CHP). In some cases, reciprocating dual fuel engines are used to drive pumps or blowers. Digester gas is used commonly for both boilers and stationary combustion systems for electricity generation. Wind and solar electric power generation systems are used in some facilities, but typically for a minor power demand. The use of electricity from onsite power generation and the use of renewable energy such as solar, wind, tidal, and other energy has been considered and implemented in a number of wastewater treatment facilities (U.S. EPA, 2006a). However, the contribution of these energy sources to the total energy consumption has been minor due primarily to relatively low and stable cost of electricity.

Energy Use for Wastewater Treatment

At the present time, as discussed in Chap. 1, nearly all of the publicly owned wastewater treatment plants in the United States provide secondary or higher levels of treatment. Treatment plants with more stringent treatment limits generally require greater amounts of electric energy per volume of water treated. Plants that have biological treatment for nutrient removal and filtration use on the order of 30 to 50 percent more electricity for aeration, pumping, and solids processing than conventional activated sludge treatment (EPRI, 1994). Wastewater reclamation plants with advanced treatment processes described in Chap. 11 also require significantly more energy to operate.

Energy Use by Individual Treatment Processes

The energy requirements of wastewater treatment systems depend on the flowrate, the characteristics of the incoming raw wastewater, and the treatment process employed. Treatment processes and equipment requiring electric energy in a municipal wastewater treatment plant are presented in Table 17–2. Various types of electric motor-driven equipment are involved in these operations and processes including pumps, blowers, mixers, sludge collectors, and centrifuges. In conventional secondary treatment, most of the electricity is used for (1) biological treatment by either the activated sludge process that requires energy for aeration blowers or trickling filters that require energy for influent pumping and effluent recirculation; (2) pumping systems for the transfer of wastewater, liquid sludge, biosolids, and process water; and (3) equipment for the processing, dewatering, and drying of residuals and biosolids.

Typical energy requirements for individual treatment processes are reported in Table 17–3. The electrical energy required for wastewater treatment vary widely but typically between 950 MJ/10^3 m^3 and 2900 MJ/10^3 m^3 (between 1000 and 3000 kWh/Mgal) for most treatment facilities (AWWARF, 2007). The amount of electricity consumed varies with plant size, influent and effluent characteristics, pumping requirements for influent

Table 17–2

Commonly used electric motor driven equipment used in wastewater treatment

Process or operation	Commonly used electric motor driven equipment
Pumping and preliminary treatment	Chemical feeders for prechlorination, influent pumps, screens, screenings press, grinders and macerators, blowers for preaeration and aerated grit chambers, grit collectors, grit pumps, air lift pumps
Primary treatment	Flocculators, clarifier drives, sludge and scum pumps, blowers for channel aeration
Secondary (biological) treatment	Blowers for channel and activated sludge aeration, mechanical aerators, trickling filter pumps, trickling filter distributors, clarifier drives, return and waste activated sludge pumps
Disinfection	Chemical feeders, evaporators, exhaust fans, neutralization facilities, mixers, injector water pumps, UV lamps
Advanced wastewater treatment	Blowers for nitrification aeration, mechanical aerators, mixers, trickling filter pumps, pumps for depth filters, blowers for air backwash, pumps for membrane filtration
Solids processing	Pumps, grinders, thickener drives, chemical feeders, mixers for anaerobic digesters and blending tanks, aerators for aerobic digesters, centrifuges, belt presses, heat dryer drives, incinerator drives, conveyors
Ancillary systems	
Odor control	Odor control fans, chemical feeders
Process water	Pumps
Plant air	Compressors

and effluent, energy requirements for odor control, and type of treatment system employed. However, generally energy use per volume tends to be lower at larger treatment plants, and activated sludge processes tend to require more energy than trickling filters (see Fig. 17–3). A typical percentage distribution of energy use in a conventional activated sludge treatment plant is illustrated on Fig. 17–4.

The operational requirements for wastewater collection and treatment systems are correlated with the wastewater load (see Fig. 3–6 in Chap. 3). If a diurnal electricity demand curve were developed for the treatment facilities, it would be of a similar shape to the flowrate and loading curves shown on Fig. 3–6. Therefore the peak energy demand at a wastewater treatment facility would likely occur from midday to the early evening hours when peak electricity consumption occurs in the community.

Advanced and New Wastewater Treatment Technologies

With the introduction of new technologies for wastewater treatment, energy usage requirements will change, particularly where effluent from a conventional treatment process is treated further for reuse applications. As reported in Table 17–3, higher levels of treatment, or new technologies whose operation is based on electric energy, i.e., membrane treatment, UV disinfection, and advanced oxidation, tend to require more energy to operate.

Table 17–3

Typical energy consumption of various treatment processes on wastewater treatment[a]

Technology	Energy consumption[b]	
	$kWh/10^3$ gal	kWh/m^3
Conventional secondary treatment WWTP[c]	**0.38 to 0.67**	**0.10–0.18**
Wastewater influent pumping	0.12–0.17	0.032–0.045
Screens	0.001–0.002	0.0003–0.0005
Grit removal (aerated grit removal)	0.01–0.05	0.003–0.013
Trickling filters	0.23–0.35	0.061–0.093
Trickling filter-solids contact	0.35	0.093
Activated sludge for BOD removal	0.53–4.1	0.14
Activated sludge with nitrification/denitrification	0.87–0.88	0.23
Membrane bioreactor	1.9–3.8	0.5–1.0[d]
Return sludge pumping	0.03–0.05	0.008–0.013
Secondary settling	0.013–0.015	0.003–0.004
Dissolved air flotation	0.12–0.15	0.03–0.04
Tertiary filtration (depth filtration)	0.1–0.3	0.03–0.08
Tertiary filtration (surface filtration)		
Chlorination (sodium hypochlorite)	0.001–0.003	0.0003–0.0008
UV (ultraviolet) disinfection	0.05–0.2	0.01–0.05
Microfiltration/ultrafiltration	0.75–1.1	0.2–0.3
Reverse osmosis (without energy recovery)	1.9–2.5	0.5–0.65
Reverse osmosis (with energy recovery)	1.7–2.3	0.46–0.6
Electrodialysis (TDS range 800–1200 mg/L)	4.2–8.4	1.1–2.2
UV photolysis with O_3 or H_2O_2 (advanced oxidation)[e]	0.2–0.4	0.05–0.1
Sludge pumping	0.003	0.0008
Gravity thickening	0.001–0.006	0.0003–0.0016
Aerobic digestion	0.48–1.2	0.13–0.32
Mesophilic anaerobic digestion (primary plus waste activated sludge)[f]	0.35–0.6	0.093–0.16
Mesophilic anaerobic digestion with thermal hydrolysis pretreatment (primary plus waste activated sludge)[f]	0.58–0.6	0.015–0.02
Sludge dewatering (centrifuge)	0.02–0.05	0.005–0.013
Sludge dewatering (belt filter press)	0.002–0.005	0.0005–0.0013

[a] Adapted in part from Burton (1996).

[b] Energy requirement per unit volume of wastewater treated.

[c] For treatment, not including conveyance. From Global Water Research Coalition (2008).

[d] From Krzeminski et al. (2012).

[e] For RO permeate.

[f] Energy recovery is not counted. Including electrical power and heating requirements.

Figure 17–3

Comparison of electrical energy used for different types of treatment processes as a function of flowrate. (From Burton, 1996.)

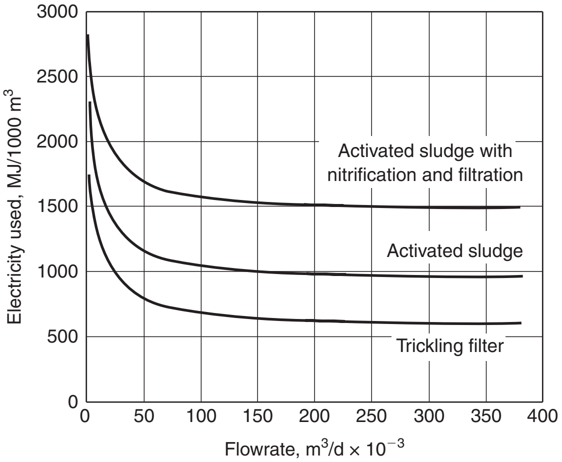
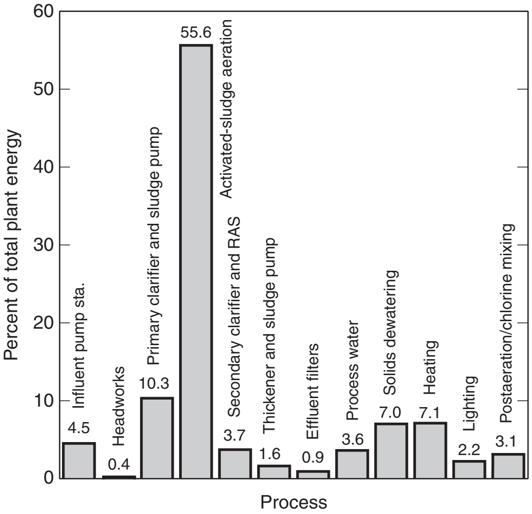


Figure 17–4

Distribution of energy usage in a typical wastewater treatment plant employing the activated-sludge process.
(From EPRI, 1994.)



17–5 ENERGY AUDITS AND BENCHMARKING

An energy audit is a procedure and methodology used to identify energy conservation opportunities. The performance of an energy audit is an important first step to identify inefficiencies with existing plant operation and opportunities to improve energy efficiency. Different levels of energy audits can be performed varying from a preliminary "walk-through" to obtain an overview of principal equipment to a detailed process audit in which the energy used by each unit processes is evaluated (EPRI, 1994). In general, one of the initial tasks in an energy audit is the evaluation of the unit processes known to consume a significant amount of energy, such as influent pumping, aeration for activated sludge plants, pumping systems at trickling filter plants, and sludge processing units. A typical procedure used to conduct an energy audit is presented in Table 17–4. Common energy audit recommendations are reported in Table 17–5.

Table 17–4

Typical procedure for energy audit

1. Establish energy and/or GHG reduction goals
2. Develop data collection questionnaire that includes
    - Energy demands
    - Process description
    - Operating parameters
    - Discharge goals
    - Equipment list with power rating, size, number in operation, etc.
    - Other important energy related information
3. Conduct an energy audit tour (visual inspection)
4. Evaluate energy consumption of:
    - Unit processes
    - Systems (i.e. secondary treatment, digestion, etc.)
    - Structures
5. Develop a matrix of energy conservation measures (ECMs)
    - Unit based
    - System/Process based
    - Structure based
    - Institutional (including changes in energy regulations including required design specification language)
6. Conduct economic analysis
    - Capital, operating expenditures (including labor)
    - Life cycle costs [(Capex + Opex)/useful life expectancy]
    - Simple payback (Capex/annual savings)
    - Compound payback (Capex + Opex)/annual energy savings
    - Net present value
    - Internal rate of return
    - Annual energy savings, [e.g., Capex/kWh, and Capex/MT CO_2(e) avoided]
    - Triple bottom line assessment
7. Produce report detailing the energy roadmap with relations to operational modifications, process modifications, and capital improvement plan modifications

Benchmarking Energy Usage

Benchmarking of energy use at a wastewater treatment plant is a fundamental and essential tool in assessing energy usage and conservation opportunities. Energy benchmarking is used to estimate the energy consumption reduction potential and to provide a basis for identifying increases or decreases in energy efficiencies as new processes, technologies, and energy conservation measures are employed. Benchmarking the energy usage of a wastewater treatment plant involves the comparison of energy consumption to a so called "standard" wastewater treatment plant. Because the design flowrates, influent loadings, process steps, operational modes, and treatment technologies can all impact energy consumption and vary significantly from plant to plant, the energy consumption must first be normalized to allow a comparison between treatment plants with different characteristics. For example, two physically identical plants with the same flowrate, but different BOD loadings will have different benchmark results when based on flow but should have more similar results when based on BOD loading.

Table 17–5

Summary of common audit recommendations for energy savings at several U.S. wastewater treatment plants[a]

1. Install adjustable speed drives on pumps and blowers for variable flowrate operations
2. Install DO monitoring and control in aeration tanks
3. Conduct periodic pumps tests and repair or replace inefficient pumps
4. Operate emergency generators to reduce peak hour power demand
5. Install CHP when replacing emergency generators
6. Install electric load monitoring devices
7. Install capacitors to improve power factor
8. Operate less reactors during prolonged under-loading conditions
9. Change or reduce pumping operations
10. Reduce odor control/ventilation areas where possible
11. Install motion sensors for lighting in areas not occupied frequently
12. Control heating and cooling in areas that are not occupied all day
13. Replace oversized motors
14. Change selected operations to off-peak periods

[a] Adapted in part from Burton (1996).

Benchmarking Protocol

In the protocol published by AWWARF (2007) the parameters that appear to have the greatest impact on energy consumption were identified based on the analysis of data collected from 266 wastewater treatment plants throughout the United States. Using multi-linear log-regression, an empirical model was developed for benchmarking energy in wastewater treatment plants. This method is based on the multi-parameter benchmark score method used by U.S. EPA Energy Star rating for buildings (U.S. EPA, 2007) with minor modifications.

In the AWWARF protocol, six parameters were identified as key variables affecting the energy use, including (1) daily average flowrate, (2) design flowrate, (3) influent and (4) effluent BOD concentrations, (5) fixed versus suspended media, and (6) conventional treatment versus biological nutrient removal. Using multi-parameter log-regression analysis, the following wastewater treatment plant energy use model was developed:

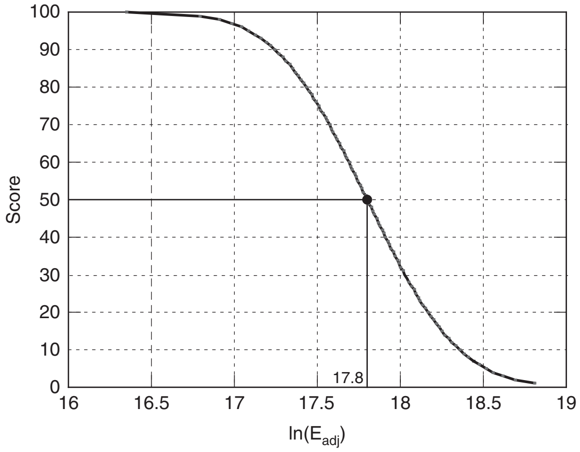
$$\begin{aligned}\ln(E_s) = {} & 15.8741 \\ & + 0.8944 \times \ln(\text{influent average Mgal/d}) \\ & + 0.4510 \times \ln(\text{influent BOD mg/L}) \\ & - 0.1943 \times \ln(\text{effluent BOD mg/L}) \\ & - 0.4280 \times \ln(\text{influent average flowrate / influent design flowrate} \times 100) \\ & - 0.3256 \times (\text{trickling filter? Yes-1, No-0}) \\ & + 0.1774 \times (\text{nutrient removal? Yes-1, No-0})\end{aligned} \tag{17–12}$$

where E_s = source energy use estimated from the model, kBtu/y (defined below)

Because the model was developed for treatment plants in the United States, the units used for the development of the model are U.S. customary units. An SI version is not available at the present time.

Figure 17–5

Benchmarking score plot for wastewater treatment plant energy use. (Adapted from AWWARF, 2007.)



Performance Rating. The value obtained from Eq. (17–12) is an estimate of the energy usage at a specific facility. To compare the estimated performance of the specific facility to other facilities, the ratio of the predicted performance to the average performance derived from the complete data set is used. The range of $\ln(E_s)$ values in the complete data set used to derive Eq. (17–12) ranges from approximately 16 to 19.6 with a mean value (50th percentile) of 17.8 (see Fig. 17–5). Thus the adjustment factor, F_{adj}, used to normalize the individual energy use factors is given as the following expression.

$$F_{adj} = \ln(E_s) / 17.8 \tag{17–13}$$

In effect the adjustment factor allows a utility to assess its benchmark score by its location on the distribution curve.

Conversion of Energy Usage Data to Source Energy Use. The source energy use is defined as the amount of raw fuel energy that is required to operate the facility (U.S. EPA, 2011). Whereas the energy usage data obtained at the wastewater treatment facility is the amount of energy measured at the point of use. The difference between the two values is due to the energy losses during production, transmission, and delivery. To allow for a consistent comparison, conversion factors (source energy factor, F_s) between the actual energy use and the source energy use were developed based at a national average, as presented in Table 17–6. It should be noted that the energy generated onsite from renewable energy sources such as the use of digester gas or solar power is not included in the source energy use calculation. The value

Table 17–6

Source energy factor used for the benchmarking of wastewater treatment plant energy usage[a]

	US customary units		SI units	
Energy Source	**Unit**	**Value**	**Unit**	**Value**
Electricity	kBtu/kWh	11.1	kBtu/kWh	11.1
Natural Gas	kBtu/therm	102.5	kBtu/MJ	0.97
Fuel Oil	kBtu/gal	141	kBtu/L	37.25
Propane	kBtu/gal	91	kBtu/L	24.04
Digester Gas	kBtu/ft^3	0.6	kBtu/m^3	21.2

[a] Adapted from AWWARF (2007).

for digester gas is included in Table 17–6 as a reference, and the use of the value is discussed in Example 17–5. Actual source energy use, E_{as}, is calculated as

$$E_{as} = \sum E_u \cdot F_s \tag{17–14}$$

where, E_{as} = Actual source energy use

E_u = Energy use measured at the point of use (= data from the treatment plant)

F_s = Source energy factor

The source energy factor will convert all energy use into the source energy use, and also convert the unit into kBtu/y, by which energy usage from various energy sources could be added to determine the total source energy use.

Calculation of Adjusted Energy Use. The adjustment factor is used to calculate the adjusted energy use factor, a normalized log-value of the energy use:

$$\ln(E_{adj}) = \ln(E_{as}) / F_{adj} \tag{17–15}$$

The adjusted energy use factors allow comparison between treatment facilities with different treatment levels and wastewater characteristics. The benchmark score can be obtained from Fig. 17–5, or the value for each score can be looked up from the AWWARF report (AWWARF, 2007). The score corresponds to the percentile in terms of normalized energy as compared to other studied plants. Benchmarking is illustrated in Example 17–5.

EXAMPLE 17–5 **Benchmarking of Energy Use at a Wastewater Treatment Plant** A BNR activated sludge process is used to treat the wastewater from a community. Based on the plant data given below, calculate the source energy usage and compare with the actual energy use to determine the benchmark score. How does the score changes if the use of energy generated onsite from digester gas was counted in the source energy use?

Item	Unit	Value
Energy use		
Electricity	kWh/y	14,100,000
Natural gas	m^3/y	17,300
Fuel oil #2	m^3/y	390
Digester gas production	m^3/y	1,047,900
Digester gas used	m^3/y	755,000
Digester gas flared	m^3/y	290,500
Digester gas vented	m^3/y	2400
Plant performance		
Annual average flowrate	m^3/d	100,000
Average influent BOD concentration	mg/L	180

(*continued*)

(*Continued*)

Item	Unit	Value
Average influent ammonium concentration	mg/L	18
Average influent total nitrogen concentration	mg/L	32
Average effluent BOD concentration	mg/L	4
Average effluent ammonium concentration	mg/L	1.5
Average effluent total nitrogen concentration	mg/L	8.3
Other required information		
Design flowrate	m^3/d	180,000
Population served	persons	430,000

Solution

1. Calculate the natural log of the source energy use value using Eq. (17–12) and the given data:

$$\begin{aligned}\ln(E_s\ \text{kBtu/y}) = {} & 15.8741 \\ & + 0.8944 \times \ln\{[100{,}000\ (\text{m}^3/\text{d})]/[3785\ (\text{m}^3/\text{Mgal})]\} \\ & + 0.4510 \times \ln(180) \\ & - 0.1943 \times \ln(4) \\ & - 0.4280 \times \ln(100{,}000/180{,}000 \times 100) \\ & - 0.3256 \times (0) \\ & + 0.1774 \times (1) \\ = {} & 15.8741 + 2.9284 + 2.3420 - 0.2694 - 1.7194 - 0 + 0.1774 \\ = {} & 19.33\end{aligned}$$

2. Calculate the adjustment factor from the value obtained in Step 1 using Eq. (17–13):

$$\text{Adjustment factor} = 19.33 / 17.8 = 1.086$$

3. Calculate the natural log of the source energy use value using the energy usage data from the wastewater treatment plant and source energy factor in Table 17–6.
   a. Calculate actual source energy use for the energy from outside sources (not including energy generated from digester gas) using Eq. (17–14).

$$\begin{aligned}E_{as}\ (\text{no digester gas}) = {} & (14{,}100{,}000\ \text{kWH/y})(11.1\ \text{kBtu/kWH}) \\ & + (17{,}300\ \text{m}^3/\text{y})(35.31\ \text{ft}^3/1\ \text{m}^3)(1.025\ \text{kBtu/ft}^3) \\ & + (390\ \text{m}^3/\text{y})(264.2\ \text{gal}/1\ \text{m}^3)(141\ \text{kBtu/gal}) \\ = {} & 156{,}510{,}000 + 626{,}135 + 14{,}528{,}358 \\ = {} & 171{,}664{,}493\ \text{kBtu/y}\end{aligned}$$

   b. Calculate source energy use for the energy including the energy generated onsite from digester gas.

$$\begin{aligned}\text{Digester gas energy} &= (755{,}000\ \text{m}^3/\text{y})(35.31\ \text{ft}^3/1\ \text{m}^3)(0.6\ \text{kBtu/ft}^3) \\ &= 15{,}995{,}430\ \text{kBtu/y}\end{aligned}$$

$$\begin{aligned}E_{as}\ (\text{with energy from digester gas}) &= 171{,}664{,}493\ \text{kBtu/y} + 15{,}995{,}430\ \text{kBtu/y} \\ &= 187{,}659{,}923\ \text{kBtu/y}\end{aligned}$$

4. Convert the source energy usage calculated in Step 3 to the natural log of the adjusted energy use using the adjustment factor from Step 2 and Eq. (17–15).
    a. Calculate $\ln(E_{adj})$ not including energy generated from digester gas.

$$\ln(E_{adj}) = \ln(171{,}664{,}493 \text{ kBtu/y}) / 1.086 = 18.96 / 1.086$$
$$= 17.46$$

    b. Calculate $\ln(E_{adj})$ including energy generated from digester gas.

$$\ln(E_{adj}) = \ln(187{,}659{,}923 \text{ kBtu/y}) / 1.086 = 19.05 / 1.086$$
$$= 17.54$$

5. Using Fig. 17–5, find the benchmark score with and without digester gas usage.

Score without counting the use of energy generated from digester gas = 78

Score counting the use of energy generated from digester gas = 72

Comment In this example, the treatment plant's normalized percentile score for energy usage, without including the use of energy generated from digester gas as specified in the protocol, is 78, or slightly better than the average of the treatment facilities used for the development of the model. Including the use of energy generated onsite from digester gas will lower the benchmark score to 72. In other words, if this treatment plant was generating digester gas but not utilizing it (i.e., flaring it) and depending solely on energy from outside sources, this plant would have scored 72 on benchmarking. Therefore the use of energy generated onsite from digester gas will significantly improve the benchmark score.

17–6 RECOVERY AND UTILIZATION OF CHEMICAL ENERGY

Recovery and utilization of chemical energy involves transformation of wastewater constituents containing chemical energy into fuel, and the use of the fuel for beneficial purposes. In some cases, pretreatment of the fuel is necessary before it is used. Recovery of chemical energy has been practiced at wastewater treatment facilities primarily by producing digester gas (biogas) from sludges with anaerobic sludge digestion, and digester gas has been used widely for boilers and other combustion systems to supplement other energy sources. Dried biosolids have also been used as an energy source where sludge incineration is practiced.

Fuels Derived from Wastewater

The types of fuels derived from wastewater constituents can be categorized as (1) gaseous fuels, (2) solids, and (3) liquids/oils. The gaseous fuels include biogas from anaerobic digestion and syngas from gasification. The solids include primary sludge, waste secondary sludge, and stabilized biosolids. Liquid fuels and oils could be produced from the solid contents of wastewater, but the generation and utilization of liquid fuels and oils derived from wastewater constituents is not common.

Generally, chemical energy in wastewater is extracted from the solid contents, and by means of a physical/chemical process or a biological treatment process. In anaerobic treatment processes (see Chap. 10), some dissolved organics are converted biologically to methane, but the use of anaerobic treatment processes for liquid treatment in municipal wastewater has been uncommon. In conventional municipal wastewater treatment facilities with aerobic biological treatment, only a fraction of the dissolved chemical energy is assimilated into biomass which is subsequently collected for the solids processing and converted into biogas through anaerobic digestion.

Biogas. The method used most commonly to recover energy from the solid contents of wastewater is to produce methane through anaerobic digestion (see Chap. 13). Typical production of digester gas through an anaerobic biological process varies between 0.75 and 1.12 m^3/kg volatile solids destroyed (12 to 18 ft^3/lb VSS). Typically, digester gas contains 55 to 70 percent methane, 30 to 40 percent CO_2, and small amounts of N_2, H_2, H_2S, water vapor, and other gases. The energy content of digester gas is typically in the range of 22 to 24 MJ/m^3 (600 to 650 Btu/ft^3) in HHV. The methane gas content depends primarily on the pH of the digester as it affects the amount of CO_2 that is released to the gas phase. The theoretical biogas production reaction, as described in Chap. 10, is

$$\begin{aligned} &C_vH_wO_xN_yS_z + \left(v - \frac{w}{4} - \frac{x}{2} + \frac{3y}{4} + \frac{z}{2}\right)H_2O \rightarrow \\ &\quad \left(\frac{v}{2} + \frac{w}{8} - \frac{x}{4} - \frac{3y}{8} - \frac{z}{4}\right)CH_4 + \left(\frac{v}{2} - \frac{w}{8} + \frac{x}{4} + \frac{3y}{8} + \frac{z}{4}\right)CO_2 + yNH_3 + zH_2S \end{aligned} \tag{10–4}$$

As shown in Eq. (10–4), ammonia and hydrogen sulfide are produced, and other volatile compounds are also produced during anaerobic digestion. Because some of the compounds could be detrimental to the combustion system, gas cleaning may be required before collected biogas is used for the combustion system. The cleaned biogas can also be used to generate electricity using a reciprocating engine, gas turbine, microturbine, or fuel cell. The use of biogas for boilers typically does not require gas cleaning. The biogas conveyance and storage system must be kept in a positive pressure to prevent accidental mixing of biogas with air, which could result in explosion.

Syngas. Syngas is a mixture of gases comprised mainly of CO, H_2, CO_2, and CH_4 generated through a gasification process (see Chap. 14) and has a lower heating value (LHV) ranging between 4 and 15 MJ/m^3 with a typical range in the wastewater application between 4.5 and 5.5 MJ/m^3 (120–150 BTU/ft^3). Energy content of syngas varies widely and depends on the gasification process technology and its operating conditions such as amount of air and moisture injected to the gasification process. The unit processes to produce syngas are described in Sec. 14–9 in Chap. 14.

Syngas can be oxidized immediately, or cleaned and used for internal combustion systems in a two stage system. Cleaned syngas could be processed further using a catalytic Fisher-Tropsch (FT) process to a liquid fuel, which can be used in an internal combustion engine-generator, boiler, or fuel cell. Liquid fuel can also be used for the production of various chemicals (Valkenburg et al., 2008). Limited information is available regarding the production rate of syngas, as it varies widely with the gasification process, characteristics of the feed solids, and the operating conditions used. In a full-scale installation of fluidized bed gasification process in Balingen, Germany, 1000 kg of dewatered sludge with 32 percent dry solid content was dried to produce 400 kg of dried sludge with 80 percent solids content, which generated to 510 m^3 of syngas and 160 kg of mineral granule (WERF, 2008).

Solid Fuels. Solid contents in wastewater are largely organic compounds. Depending on the water content of the sludges or biosolids separated from wastewater they can be incinerated without additional fuels. Typical heating values of sludges and biosolids are reported in Table 14–17 in Chap. 14, and theoretical values could be derived from Eq. (2–66) if the chemical composition is known (see Sec. 17–2).

Liquid Fuel and Oils. Technologies are available to produce liquid fuel and oils from the solid constituents present in wastewater. For example, syngas could be converted into a liquid fuel through the Fisher-Tropsch (FT) process, and char and oil are generated from pyrolysis of solid contents. However the production of oil and liquid fuel from wastewater solid contents has not been implemented in full-scale.

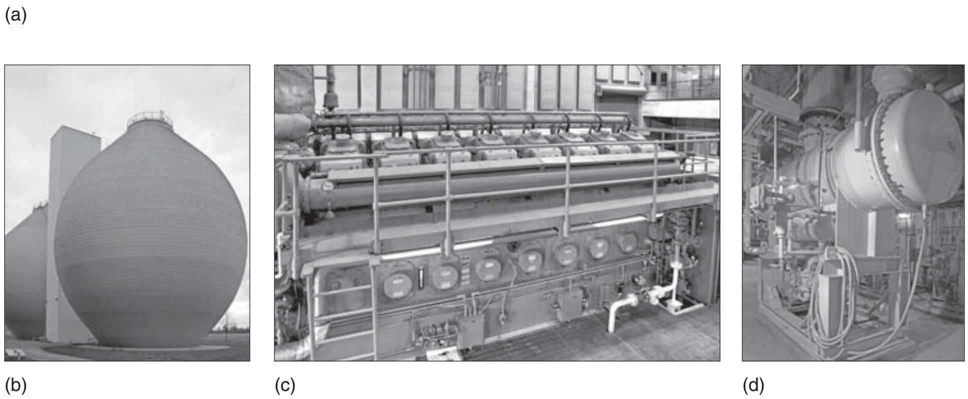
Energy Recovery from Gaseous Fuels with Engines and Turbines

Reciprocating engines, gas turbines, and microturbines are the principal technologies used to generate electricity from the combustion of gaseous fuels derived from wastewater. Even though it is practiced less commonly, pumps and blowers can be operated with a direct drive from the engines fueled by gaseous fuels. Typical energy recovery systems with engines and turbines include, as illustrated on Fig. 17–6, a gas generation process, gas holding vessel, compressors, gas cleaning, an engine/turbine, emission control (see Chap. 16), and a waste heat recovery system. Fuel cells are also used to generate electricity from biogas but by a different mechanism. Thus fuel cells are considered separately in this section.

Exhaust heat from these combustion processes can be used to heat water for building heating, heat anaerobic digesters, or to provide hot water supply, as discussed further in Sec. 17–6. It should be noted that the use of biogas in boilers and engines will result in the emission of flue gas, which may be subject to strict regulations to minimize air pollution. Emissions from boilers and engines are considered in Sec. 16–5 in Chap. 16.

Figure 17–6

Typical process for the recovery and utilization of biogas generated by anaerobic digestion of sludges: (a) process flow diagram, (b) egg shape anaerobic digester, (c) view of dual fuel type reciprocating engine, and (d) view of heat recovery boiler.



(a)

(b) (c) (d)

Gas Pretreatment Requirements. Constituents in biogas and syngas that affect the operation of equipment to generate power and heat include hydrogen sulfide (H_2S), siloxanes, carbon dioxide (CO_2), and moisture (H_2O). Syngas also contains particulate matters that may need to be removed. Hydrogen sulfide (H_2S) is an odorous gas and highly corrosive. At relatively low concentrations H_2S is hazardous to human health. The immediately dangerous to life or health (IDLH) value used by the National Institute for Occupational Safety and Health (NIOSH) for H_2S is 100 ppm (CDC, 1994). An H_2S value of 100 ppm is often exceeded in biogas generated from anaerobic digestion. Siloxanes are known to cause detrimental impacts to the combustion systems. Water condensation occurs in the biogas transfer pipelines and a compressor, and enhances corrosion by hydrogen sulfide. To minimize these adverse impacts, digester gas is often cleaned before combustion. Carbon dioxide may be removed only when biogas is to be sold as natural gas (sometimes referred to as bio-methane), thus CO_2 removal is rarely practiced at a wastewater treatment facility unless there is an incentive to sell bio-methane instead of using the biogas within the facility. When biogas is mixed with natural gas at a treatment facility, it is common to mix natural gas with air to reduce the heat content to the level of biogas, instead of removing CO_2 from the biogas to raise the biogas heat content to that of natural gas. Because the air added to the natural gas is not sufficient to cause explosion, the removal of oxygen is not required for the natural gas/air mixing process.

Typical treatment systems for biogas and syngas are shown on Fig. 17–7. The types of gas cleaning technologies used for digester gas and syngas are reported in Table 17–7. Description of each gaseous compound and its treatment is described in detail in Sec. 13–9 in Chap. 13. In many cases, not all of these gas cleaning processes are necessary, and the cleaning requirements are determined according to the gas quality requirements for the end use.

Total System Efficiency. The total system efficiency, also known as thermal efficiency, is defined as the ratio of the total energy output, either work, heat, or electricity, to the total energy input. For an engine generator without heat recovery, total system efficiency is the electrical power output per fuel energy input. In a CHP system, the total system efficiency is a sum of the power output and useful net thermal output per total fuel input through a downstream heat recovery process per energy input by a fuel. Typical thermal efficiencies of internal combustion engines and fuel cell are presented in Table 17–8. As reported in Table 17–8, the overall thermal efficiency is improved significantly by employing CHP. Recovery of waste heat is further discussed in Sec. 17–6. Equipment suppliers should be consulted to obtain expected thermal efficiency for specific equipment, but thermal efficiency data in Table 17–8 may be useful for making preliminary estimates on the energy output based on the fuel consumption when accurate data on energy output

Figure 17–7

Typical gas cleaning process flow diagrams: (a) for biogas from anaerobic digestion and (b) syngas from gasification.

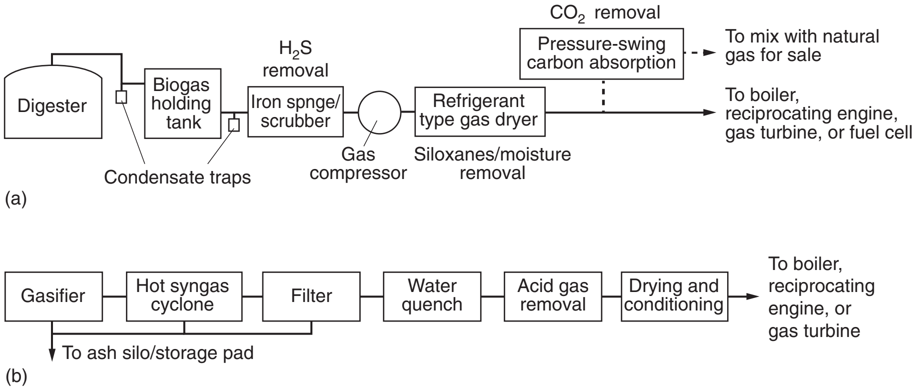


Table 17–7

Major contaminants in digester gas, and treatment method for their removal[a]

Gaseous compound	Typical concentration in digester gas, mg/L[b]	Treatment methods
Hydrogen sulfide (H_2S)	1000–2000	Adsorption Scrubber Chemical addition to digester
Siloxanes (in silica)	0.3–11	Refrigeration/drying Scrubber Adsorption (activated carbon, proprietary media)
Water vapor (H_2O)	Saturation	Solid desicants Liquid absorption Refrigeration/drying
Carbon dioxide (CO_2)	30 to 35%	Scrubber. CO_2 is removed only for the application to produce natural gas quality gaseous fuel and the application has been limited

[a] Compiled from various sources.

[b] Unless otherwise noted.

are not available. When digester gas is used for engines and boilers, the actual heat content of digester gas should be measured to accurately project the expected energy output.

Reciprocating Engine Generators. Reciprocating engines and gas turbines are used widely to produce electricity on site at wastewater treatment facilities using recovered chemical energy in the form of digester gas [see Figs. 17–8(a) and (c)]. Both spark ignition engines (Otto-cycle engines) and compression ignition engines (diesel-cycle engines) have been used with biogas. When compression ignition engines are used, a supplemental fuel oil will be added to induce the ignition. The typical size of the reciprocating engines ranges between 20 kW and 6 MW (see Table 17–8).

Gas Turbines. Gas turbines used at wastewater treatment facilities have a wind range of capacity [see Fig. 17–8(d)], in the range of 1 to 250 MW. Their electrical efficiency is slightly lower than reciprocating engines, varying from 30 and 40 percent. As shown in Table 17–8, overall thermal efficiency can be in the same range with reciprocating engines when it is used for CHP because exhaust gas from gas turbines has a higher temperature and can be used for boilers and a wide range of heat recovery processes. However, reciprocating engines have been used more commonly at wastewater treatment facilities because the reciprocating engines generally have a wider operational range than the gas turbines and the energy output could be adjusted more easily in response to the diurnal energy demand variations.

Microturbines. A microturbine is a smaller gas turbine, often packaged by the burbine suppliers. The typical size of microturbines used at wastewater treatment facilities range from 30 to 250 kW. In the typical microturbines, a recuperation cycle is used to utilize the heat in the exhaust gas to preheat the combustion gas. Remaining heat is recovered to produce hot water or connected to other thermal energy recovery devices. Microturbines are suited for distributed generation applications to supplement a specific electrical load, or to be used in parallel to serve large loads.

Table 17–8

Devices used to recover energy from digester gas and syngas

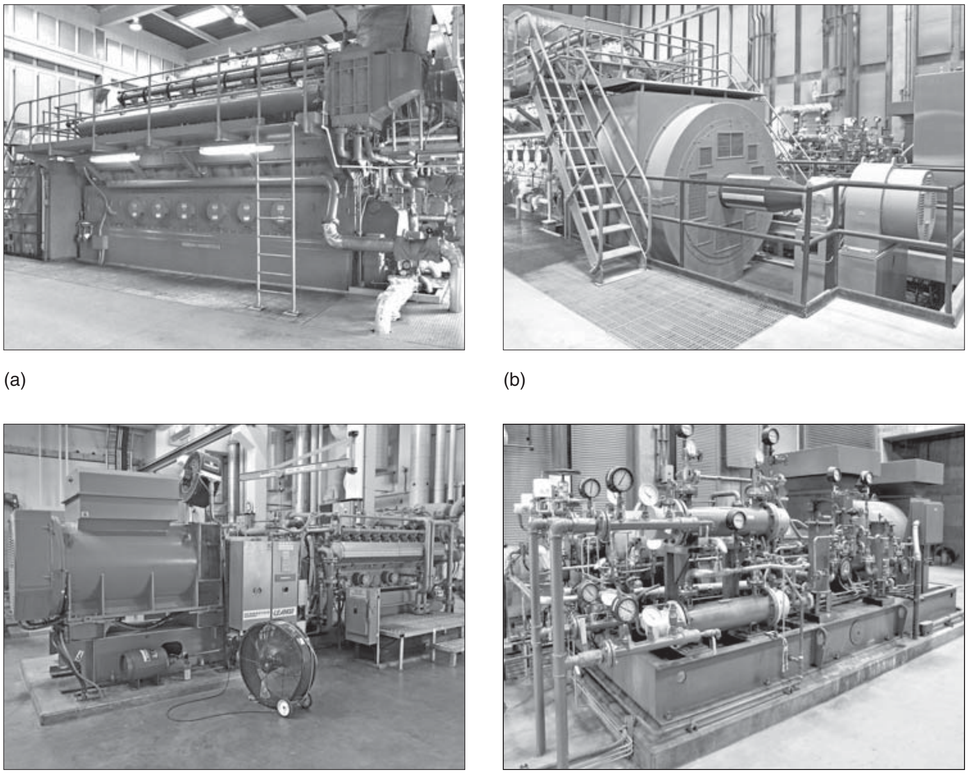
Device	Typical efficiency[a], %	Typical efficiency with CHP, %	Gas cleaning requirements	Typical size, kW	Conversion of energy
Reciprocating engine	25–50	70–80	siloxiane	20–6000	electricity, mechanical power, waste heat
Gas turbine (simple cycle)	25–40	70–80	siloxiane, H_2S	1000–250,000	electricity, mechanical power, waste heat
Gas turbine (combined cycle)	40–60	70–80	siloxiane, H_2S	1000–250,000	electricity, mechanical power, residual heat to produce steam for more electricity generation
Micro-turbine	25–35	70–85	siloxiane, H_2S	30–250	electricity, mechanical power, waste heat
Sterling engine	~30	~80	no cleaning required		electricity, mechanical power, waste heat
Fuel cell	40–60	70–85	siloxiane, H_2S, H_2O	200–3000	electricity, waste heat
Boiler	80–90+	-	typically no cleaning required		steam, hot water

[a] Efficiencies based on the manufacturers' rating for a new system, not including heat recovery.

Sterling Engine. Sterling engine is an external combustion engine and a wide variety of fuels can be used without the level of gas cleaning required for internal combustion engines. Electrical efficiency is approximately 30 percent, and a total system efficiency can be up to 80 percent when heat recovery is included.

Energy Recovery from Gaseous Fuels with Boilers

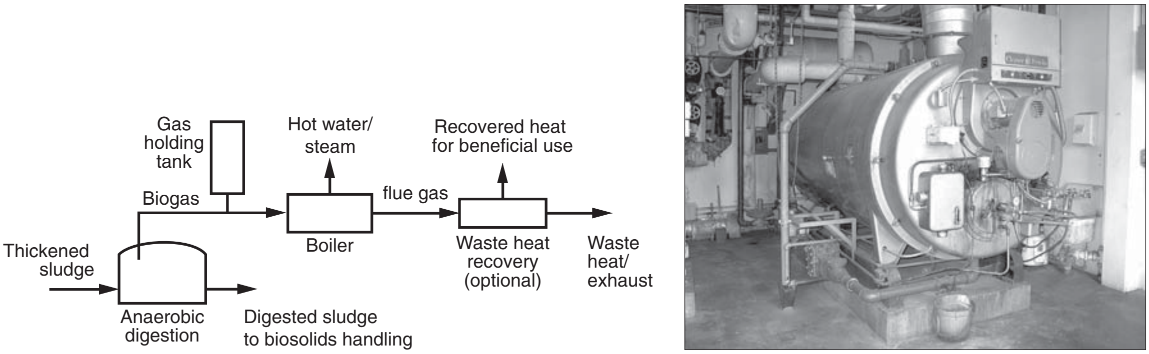
Boilers are used at wastewater treatment facilities to generate hot water or steam for steam turbines, space heating, and hot water supply. Heating requirements for wastewater treatment processes include heating of anaerobic digester and building heating and various solids handling processes such as sludge pretreatment and sludge drying (see Chaps. 13 and 14). Boilers are used also for heating of administration and control building, as well as hot water supply. Typically gas pretreatment is not included when biogas or syngas is used only for the boilers, and it is designed to maximize the thermal efficiency of the boiler, even though further recovery of heat from the waste heat is possible (see Fig. 17–9).



(a) (b) (c) (d)

Figure 17–8

Typical devices utilizing biogas generated at wastewater treatment facilities: (a) view of large reciprocating dual fuel engine, (b) generator connected to engine shown on (a), (c) view of alternative dual fuel engine, and (d) view of gas turbine generator.



(a) (b)

Figure 17–9

Energy recovery system utilizing biogas in a dual-fuel boiler: (a) process flow diagram and (b) view of boiler.

Types of Boilers. The typical boilers used commonly at wastewater treatment facilities are hot water boilers to provide building heating, heating for treatment processes, and hot water supply. The boilers producing steam are also used especially when a high temperature heat source is necessary for sludges and biosolids processing, such as thermal hydrolysis. Boilers can generally be categorized as fire tube type and water tube type. In fire tube boilers, hot gases flow through the tubes to heat water that surrounds the tubes to generate hot water and steam. Fire tube boilers are suited for smaller, low pressure (less than 17 bar or 250 $lb_f/in.^2$) steam generation. In water tube boilers, water runs through the tubes which are placed in a combustion chamber, filled with hot combustion gases. Water in the tube is heated and produces steam in the steam drum. Water tube boilers are suited for producing higher (higher than 17 bar or 250 $lb_f/in.^2$) pressure steam.

Thermal Efficiency of Boilers. Selection of the type and size of boilers depends on the amount and quality of heat to be delivered and the types of fuel to be used. Either dual-fuel or gas boilers could be used with biogas. It should be noted that sometimes the term "thermal efficiency" is used specifically for the efficiency of the heat exchanger but does not count radiation and convection losses. The term "boiler efficiency" may be used instead for the thermal efficiency to account for the radiation and convection losses. Typically, the thermal efficiency of boilers is higher than 80 percent based on the manufacturers' specifications, and it is affected by the flue gas temperature, fuel type, amount of excess air, ambient temperature, and heat losses. Condensing boilers are the type of boilers equipped with preheating of the cold feed water. By preheating the feed water, overall efficiency can be improved to greater than 90 percent (see Table 17–8). Build up of soot or scaling in boilers will result in reduced thermal efficiency. Periodic cleaning and maintenance are necessary to maintain the thermal efficiency of boilers.

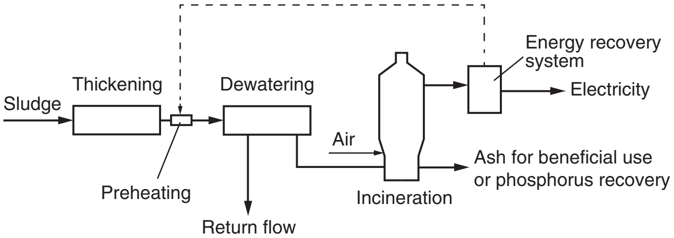
Energy Recovery from Solid Fuels

Thermal oxidation (i.e., incineration) of solid contents is a process typically used by medium- to large-size plants with limited disposal or reuse options (Chap. 14). To maximize the energy recovery from thermal oxidation, the solid contents are used directly without a solid stabilization process even though incineration of stabilized biosolids is also practiced. The energy recovery system from solid fuels typically includes sludge/biosolids thickening, dewatering, drying, thermal oxidation, and energy recovery system. Part of the heat from the incineration is used to preheat the incoming sludge and air (see Fig. 17–10). The flue gas contains particulate matters and other pollutants, and appropriate emission control must be placed to meet the air emission requirements.

The two most important parameters to determine energy balance are dewatered cake solids content and volatile solid content. Typically the dewatered solids content must be greater than 22–28 percent solids to balance energy release and evaporative cooling. Solids content is also an important operating parameter as it controls the combustion temperature. Higher temperature will result in the emission of nitrogen oxides and formation of amalgam from the ash, whereas lower temperature will result in an incomplete combustion and increased emission of particulate matters (see Sec. 16–5 in Chap. 16). The overall heat balance can vary based on the solids calorific value and overall efficiencies of the drying, gasification, and energy recovery systems. Undigested sludges typically have a higher volatile content so supplemental fuel use can be eliminated at a lower solids content. Anaerobically digested biosolids contain lower calorific values per unit dry material as some of the volatile material is consumed and converted to biogas. Digested sludge with chemical addition within the treatment system generally contains lower calorific value

Figure 17–10

Typical process flow diagram for the recovery of chemical energy by combustion of sludges and biosolids.



than digested sludge without chemical precipitation (Barber, 2007). Because the calorific value of digested biosolids is lower, a higher dewatered solid content is required to achieve an energy neutral drying and gasification process.

Combustion of Sludge and Biosolids. Combustion of solid and semi-solid sludge and biosolids has been implemented in some parts of world where land application and/or landfill disposal is not economical. The main types of combustors used for biosolids combustion include multiple hearth, fluidized bed, and electric infrared. Devices used to recover energy from wastewater solids are summarized in Table 17–9.

Combustion Stoichiometry. The stoichiometric expression for the complete combustion of sludges and biosolids, discussed in Sec. 14–4 in Chap. 14 and repeated here for convenience, is:

$$C_aO_bH_cN_d + (a + 0.25c - 0.5b)O_2 \rightarrow aCO_2 + 0.5cH_2O + 0.5dN_2 \tag{14–3}$$

If it is assumed that dry air contains 23.15 percent of oxygen by weight, the amount of air required for complete oxidation is approximately 4.3 times the calculated amount of required oxygen. In addition, excess air is required to induce near sufficient turbulence and mixing in the combustion chamber. For the incineration of dried sludges and biosolids, typically about 50 percent of excess air is required. Excess air will also affect the combustion temperature, and the air to fuel ratio must be controlled to maximize the efficiency of combustion and minimize the emission of pollutants such as nitrogen oxides, carbon monoxide, volatile organic compounds and other potentially hazardous compounds (see Chaps. 14 and 16).

Table 17–9

Devices used to recover energy from wastewater solid contents[a]

Device	Output[a]	Remarks
Fluidized bed incinerator	Waste heat	• Used widely for dewatered sludge incineration • Typical feed solids content 20 to 35 percent
Multiple heath incinerator	Waste heat	• Used widely for dewatered sludge incineration • Typical feed solids content 20 to 35 percent
Fluidized bed gasifier	Syngas, waste heat	• Uniform product • Require high solids content (>85 percent) • Operating temperature 700 to 900°C
Fixed bed updraft gasifier	Syngas, waste heat	• Relatively low solids content is acceptable (70 to 80 percent) • Relatively high tar content in syngas • Operating temperature ~1000°C
Fixed bed downdraft gasifier	Syngas, waste heat	• Require high solids content (>80 percent) • Relatively low tar content in syngas • Operating temperature 900 to 1000°C • Unconverted carbon 4 to 7 percent
Entrained flow gasifier	Syngas, waste heat	• Relatively low solids content is acceptable • Low tar and CO_2 • Low CH_4

[a] Adapted in part from DOE (2002) and McKendry (2002).

[b] Depending on the heat balance, waste heat could be used to operate turbine to generate electricity and steam or hot water.

Heat Released from Combustion. Part of the heat released from the combustion of solid materials is stored in the combustion product to raise the temperature of the material, and partly transferred by convection, conduction, and radiation to the wall of the combustion system. Depending on the dewatering and drying process, sludges and biosolids may contain a significant amount of water which will be vaporized in the combustion chamber, taking up latent heat of evaporation. As discussed in Chap. 14, the heat required to maintain the combustion system at the final temperature is expressed as:

$$Q = \sum(Q_S) + Q_e + Q_L = \sum C_P W_S(T_2 - T_1) + W_W\lambda + Q_L \qquad (14\text{–}4)$$

where Q = total heat, kJ (Btu)
Q_s = sensible heat in the ash, kJ (Btu)
Q_e = latent heat, kJ (Btu)
Q_L = heat loss
C_p = specific heat for each category of substance in ash and flue gases, kJ/kg·°C (Btu/lb·°F)
W_s = mass of each substance, kg (lb)
T_1, T_2 = initial and final temperatures
W_w = mass of water, kg (lb)
λ = latent heat of evaporation, kJ/kg (Btu/lb)

By comparing the required heat and heat content in the combusted material as calculated by Eq. (2–66), additional energy requirement, or the excess heat that could be recovered and utilized, can be calculated. In Example 17–6, a heat balance for combustion of biosolids is presented.

EXAMPLE 17–6 **Computation of Heat Balance for Combustion of Biosolids** Dewatered biosolids with the following characteristics is to be combusted in an incinerator. Using the information, given below, on the operating conditions for the incinerator and related process data, determine the air requirement and develop a heat balance. Determine heat content of the flue gas after it was used to preheat the inlet air. For simplicity, assume all combustible materials listed below are oxidized to CO_2, H_2O, N_2 and SO_2, and ignore SO_2 in heat balance calculations.

Biosolids composition and elemental analysis

Element	Percent of total weight
Combustible	
Carbon	14.2
Hydrogen	1.0
Oxygen	5.2
Nitrogen	0.3
Sulfur	0.4
Inerts	8.9
Water	70
Total	100

Operating conditions

Item	Unit	Value
Ambient air temperature	°C	20
Relative humidity at 20°C	%	50
Target combustion temperature	°C	850
Heat loss	% of gross heat input	0.5

Thermodynamic properties

Component	Unit	Value
Specific heat of water	kJ/kg·°C	4.19
Specific heat of biosolids	kJ/kg·°C	1.26
Specific heat of dry air	kJ/kg·°C	1.01
Specific heat of water vapor	kJ/kg·°C	1.88
Specific heat of the ash	kJ/kg·°C	1.05
Latent heat of water evaporation	kJ/kg	2257

Air composition at 20°C and 50 percent humidity

Gas	Percent
N_2	77.1
O_2	20.7
CO_2	0.04
H_2O	1.17
Other gas (ignore in calculations)	1.00

Enthalpy of ideal gas, kJ/mole

	20°C	850°C
N_2	8.53	34.2
O_2	8.54	35.7
CO_2	9.18	49.6
H_2O	9.74	41.1

Solution

1. Estimate the air requirement.
   a. Estimate the stoichiometric requirement to completely oxidize the sludge, assuming CO_2, H_2O, N_2, and SO_2 are the only combustion products. Set up a computation table to determine the moles of oxygen and kg of air required per kg of biosolids.

Component	Weight fraction	Atomic weight, kg/mole	Atomic weight units[a], mole/kg	O_2 required[b], mole	Combustion reaction and products	Product gas formed, mole/kg
Carbon	0.142	0.012	11.83	11.83	$C + O_2 \rightarrow CO_2$	11.83
Hydrogen	0.010	0.001	10.00	2.50	$4H + O_2 \rightarrow 2H_2O$	5.0
Oxygen	0.052	0.016	3.25	−1.63	$2O \rightarrow O_2$	0
Nitrogen	0.003	0.014	0.214	–	$2N \rightarrow N_2$	0.11
Sulfur	0.004	0.0321	0.125	0.062	$S + O_2 \rightarrow SO_2$	0.12
Water	0.70	0.018			H_2O (vapor)	38.9
Inerts	0.086					
Total	1.00			12.83		56.0

[a] Atomic weight unit = weight fraction/atomic weight.

[b] Moles required = atomic weight unit × mole of O_2 required per atom being oxidized based on the combustion reaction. For oxygen, O_2 saved due to oxygen in the biosolids is recorded, indicated with a negative sign.

From the calculations above, 1 kg of biosolids will require 12.8 moles of O_2. Oxygen content in the air = 20.7 percent by volume at 20ºC and 50 percent humidity, and assuming mole fraction = volume fraction, moles of air required is:

Air required = 12.8/0.207 = 62.0 mole air/kg biosolids

b. Determine the amount of gas generated from combustion of 1 kg biosolids.
For CO_2, 1 mole C is converted to 1 mole CO_2. From the summary above, CO_2 formed is 11.8 mole/kg biosolids (assuming complete combustion). Similarly, H_2O from hydrogen, N_2 from nitrogen, and SO_2 from sulfur are calculated and summarized in the table. Note that the water content in biosolids (70 percent) also becomes water vapor.

2. Develop a heat balance for various quantities of excess air for a unit mass of biosolids. Consider 0, 50, and 100 percent excess air and prepare a computation table to summarize the results. Ignore SO_2 for the rest of the calculations.

a. Determine air flows.
From Step 1, stoichiometric air requirement is 62.0 mole/kg biosolids
For 50 percent excess air: 62.0 × 1.5 = 93.0 mole/kg biosolids

b. Determine heat content of added air.
From air composition and enthalpy data given in the problem statement, heat content of added air at 20ºC without excess air is

$$H = [(8.53 \times 0.771 + 8.54 \times 0.207 + 9.18 \times 0.0004 + 9.74 \times 0.0117 \text{ (kJ/mole)}] \times 62.0$$
$$\text{mole air/kg biosolids} = 524 \text{ kJ/kg biosolids} = 0.524 \text{ MJ/kg biosolids}$$

Heat content with 50 and 100 percent excess air is calculated as summarized in the computation table.

c. Determine the heat content in biosolids at 20ºC.
Solid content of biosolids is 30 percent and specific heat of dry biosolids is given in the problem statement. Thus heat content is 0.30 × 1.26 × 20 = 7.56 kJ/kg biosolids. Water content of biosolids is 70 percent. Thus, the heat content is 0.70 × 4.19 × 20 = 58.7 kJ/kg biosolids. Total heat content is 7.6 + 58.7 = 66.3 kJ/kg biosolids = 0.066 MJ/kg biosolids.

d. Determine the flue gas composition for 0, 50, and 100 percent excess air flows and calculate heat content at 850ºC. Calculations for 0 percent excess air flow is shown as an example.

   i. Determine gas composition.

      N_2: (N_2 from N in biosolids) + (N_2 in the air) = 0.11 (mole/kg biosolids) + 62.0 (mole air/kg biosolids) × 0.771 = 47.9 mole/kg biosolids

      O_2: With no excess air, all oxygen in the air is used. For 1 kg of biosolids, 12.8 moles of oxygen is used

      CO_2: 11.8 (mole/kg biosolids) + 62.0 (mole air/kg biosolids) × 0.0004 = 11.9 mole/kg-biosolids

      H_2O: [5.00 + 38.9 (mole/kg biosolids)] + 62.0 (mole air/kg biosolids) × 0.011 = 44.6 (mole/kg biosolids)

      Similarly, the flue gas composition with excess air is calculated and the results are summarized in the table.

   ii. Calculate the heat content using the data given in the problem statement.

      N_2: 47.9 × 34.2 = 1638 kJ/kg biosolids

      O_2: 0 × 35.7 = 0 kJ/kg biosolids

      CO_2: 11.9 × 49.6 = 587 kJ/kg biosolids

      H_2O: 44.6 × 41.1 = 1832 kJ/kg biosolids

      Total = 1638 + 0 + 587 + 1832 = 4057 kJ/kg biosolids = 4.057 MJ/kg biosolids

e. Calculate the heat content remaining in the ash.

   Assuming complete combustion, ash = inert. Heat content is
   0.089 × 1.05 × 850 = 79.4 kJ/kg biosolids = 0.079 MJ/kg biosolids

f. Estimate the heat released from combustion of biosolids, assuming complete combustion.

   Using Eq. (2–66) and elemental analysis data given in the problem statement, estimate the heat contents of the solids:

   $$\text{HHV (MJ/kg)} = 34.91\text{ C} + 117.83\text{ H} - 10.34\text{ O} - 1.51\text{ N} + 10.05\text{ S} - 2.11\text{A}$$

   $$\begin{aligned}\text{HHV (MJ/kg)} &= 34.91 \times 0.142 + 117.83 \times 0.010 - 10.34 \times 0.052\\ &\quad -1.51 \times 0.003 + 10.05 \times 0.004 - 2.11 \times 0.089\\ &= 5.446\text{ MJ/kg-biosolids)}\end{aligned}$$

g. Estimate evaporative cooling from the vaporization of water in biosolids and water formed by combustion of biosolids.

   From 1 kg of biosolids, 43.9 moles of water are formed. Latent heat of vaporization is 2257 kJ/kg. Therefore, latent heat of vaporization associated with every kg of biosolids is

   $$[(43.9 \times 18)/1000] \times 2257 = 1783\text{ kJ/kg-biosolids} = 1.783\text{ MJ/kg-biosolids.}$$

h. Estimate the heat loss. Assume 0.5 percent of gross heat input.

   Gross heat input = (heat of biosolids) + (heat of inlet air) + (heat of combustion) − (heat loss from evaporation). Note that latent heat of vaporization should be subtracted from the gross heat input as the heat of combustion calculated in Step f above is HHV.

   Gross heat loss = (0.524 + 0.066 + 5.446 − 1.783) × 0.005 = 0.021 MJ/kg-biosolids

3. Evaluate the heat balance to determine the air flow to maintain the operating temperature at 850ºC.

From the summary table below, energy balance is barely positive with stoichiometric air flow, and the balance is negative with excess air flow. If the heat balance is linear to the air flow, heat balance is exactly zero at 4.3 percent excess air.

	Unit[a]	Stoichiometric air flow	With 50% excess air	With 100% excess air
Air added	mole/kg biosolids	60.3	90.4	120.5
Flue gas composition				
N_2	mole/kg biosolids	46.6	69.8	93.0
O_2	mole/kg biosolids	0	6.2	12.5
CO_2	mole/kg biosolids	11.4	11.5	11.5
H_2O	mole/kg biosolids	45.1	45.5	45.9
Heat content of added air at 20°C	MJ/kg biosolids	0.524	0.786	1.05
Heat content of biosolids at 20°C	MJ/kg biosolids	0.066	0.066	0.066
Flue gas heat content at 850°C	MJ/kg biosolids	4.057	5.119	6.180
Ash heat content at 850°C	MJ/kg biosolids	0.079	0.079	0.079
Energy released from combustion	MJ/kg biosolids	5.446	5.446	5.446
Heat loss by evaporation of water	MJ/kg biosolids	1.783	1.783	1.783
System heat loss	MJ/kg biosolids	0.021	0.023	0.024
Net energy balance		0.095	−0.7	−1.5

[a] Units are per kg biosolids on a wet basis.

Note: Because values were calculated on a spreadsheet and rounded, some values may not match exactly with manual calculations.

4. Determine the water content that will allow self-sustained combustion. From Step 3, the heat generated by combustion of chemical contents in the biosolids is barely sufficient to maintain the heat balance. Therefore in this example, 30 percent solids, 70 percent water content was the limit to sustain combustion. At water content of 71 percent, the heat balance is −0.12 MJ/kg biosolids. It is important to note that at stoichiometric air flow, the heat content in the flue gas is approximately 4.1 MJ/kg biosolids. The heat in the flue gas could be used to preheat biosolids and inlet air or to generate electricity and/or hot water.

Comment The dewatered sludge solid content assumed in this example is on the higher end of the typical water content with conventional centrifuge. Even though the calculations used in this example may not be completely representative of the actual combustion system as ideal conditions were assumed, they are helpful in illustrating how to conduct a preliminary assessment of the dewatering/drying requirements and operational conditions such as excess air requirement and operating temperature. In biosolids incineration facilities, inlet air is often preheated with flue gas to save the use of supplemental fuel. In a typical incineration facility, a solid content of 26 to 28 percent is considered the threshold needed to sustain combustion without supplemental fuel, but no significant excess energy will be available for other purposes. During the plant start up, supplemental fuel must be used to raise the temperature of the reactor.

Energy Recovery from Syngas

There are two approaches to utilize energy from syngas. One is to clean the syngas and use it for conventional boilers and engine generators, often referred to as two-stage gasification. The other approach is to use the syngas directly in the thermal oxidation chamber, referred to as close-coupled gasification.

Two Stage Gasification. In two stage gasification systems, as depicted on Fig. 17–11(a), the syngas produced from gasifying the dried biosolids is cleaned and the cleaned syngas can be used as a fuel source for an internal combustion engine. Cleaning the syngas is required to remove sulfur, siloxanes, and other contaminants such as tar that could damage the engine. The syngas cleaning process is not fully developed for the application of biosolids and currently considered in the innovative phase. Syngas cleaning, however, is commercially practiced in the coal industry, but on a much larger scale.

Close-Coupled Gasification. Close-coupled gasification, illustrated on Fig. 17–11(b) does not require syngas cleaning and instead the syngas is thermally oxidized. Syngas oxidation generates high temperature, approximately 980°C (1800°F), flue gas which can be used for thermal heat recovery. The energy recovered from the flue gas can be used as the energy source to dry to the biosolids to the desired dryness and thus minimize or eliminate the need for fossil fuels (e.g., natural gas or fuel oil). The hot flue gas can also be used as an energy source for generating electricity through the use of boiler and steam turbine or an organic Rankine cycle (ORC) engine. Electricity generation with close-coupled gasification is practiced commonly on other types of biomass, however, this system is not common for biosolids because it is generally more economical to use the energy to offset the drying energy requirement.

Energy Recovery with Fuel Cell

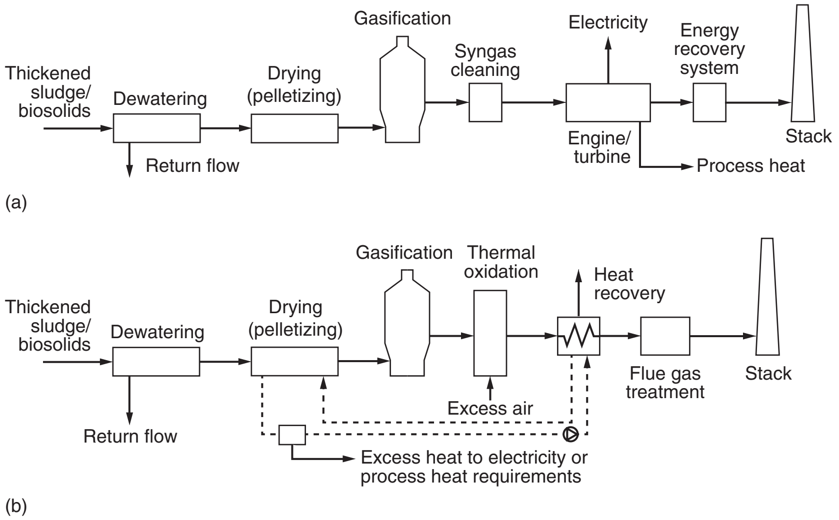
Fuel cell systems used at wastewater treatment facilities utilize methane gas generated from anaerobic digestion. In the fuel cell system, methane gas is used to generate hydrogen as expressed by Eqs. (17–16) and (17–17):

$$CH_4 + 2H_2O \rightarrow CO + 3H_2 \tag{17–16}$$

$$2H_2O + CO \rightarrow H_2 + CO_2 \tag{17–17}$$

Figure 17–11

Recovery and utilization of chemical energy by gasification: (a) two-stage gasification and (b) closed-couple gasification.



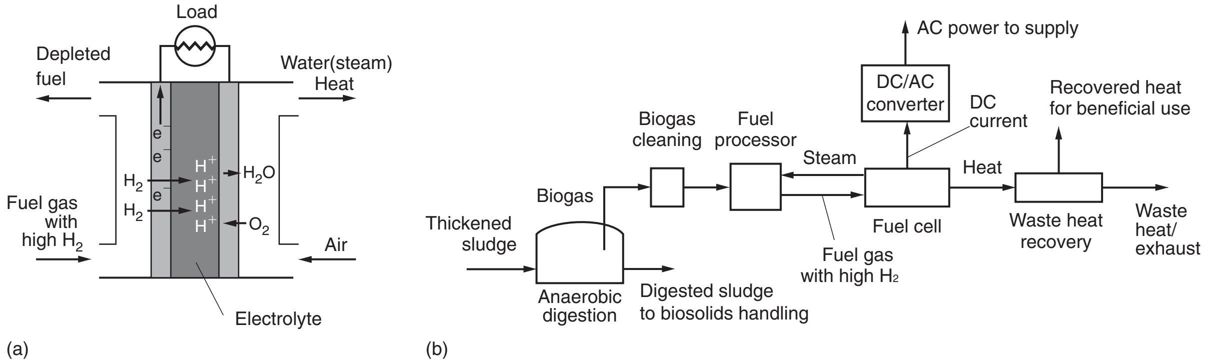


Figure 17–12

Fuel cell energy recovery system: (a) schematic illustration of fuel cell (adapted from U.S. EPA, 2006b) and (b) typical process flow diagram of a fuel cell system used at a wastewater treatment facility.

Hydrogen gas is then introduced to the anode side of the fuel cell system, where electrons are released and protons move within the fuel cell to the cathode. On the cathode side of the fuel cell, protons react with oxygen to produce water [see Fig. 17–12(a)].

Fuel cells are in general sensitive to impurity in the source fuel, and pretreatment is necessary to remove at least hydrogen sulfide (H_2S) and halides before digester gas can be sent to fuel cells. Fuel-cell systems also require essentially non-detect for siloxanes. Cleaned gas is mixed with steam to induce the reactions shown in Eqs. (17–16) and (17–17), and introduced into the fuel cell stack. Steam is recycled from the fuel cell where protons (H^+) and oxygen react on the cathode side of the fuel cell to produce heat and water. Electricity generated from a fuel cell is in direct current (DC), and is converted to alternating current (AC) for use in AC circuits. A typical process flow diagram is shown on Fig. 17–12(b).

Among a number of fuel cell types being used or studied, three fuel cell types: phosphoric acid fuel cells (PAFC), molten carbonate fuel cells (MCFC), and solid-oxide fuel cells (SOFC) are considered promising for the application at wastewater treatment facilities (U.S. EPA, 2006b). Of the three, PAFC has an advantage that the system is relatively tolerant to impurities in the source fuel. For example, phosphoric acid fuel cells have been installed at four wastewater treatment facilities operated by the New York City Department of Environmental Protection (Carrio, 2011). Waste heat produced from the fuel cell system can be recovered for other purposes (cogeneration). A comparison of the three fuel cell types is reported in Table 17–10. The total system efficiency of a fuel cell system is typically 40 to 60 percent without additional heat recovery. With heat recovery, total system efficiency can vary 70 to 85 percent.

17–7 RECOVERY AND UTILIZATION OF THERMAL ENERGY

Recovery and utilization of thermal energy involves transferring heat energy from a heat source to a heat demand. In heat recovery systems, thermal energy is exchanged from one medium to the other, sometimes with additional energy input, and conveyed to the point of use where the recovered energy is used for an intended purpose, such as space heating, heating of an anaerobic digester, or electrical power generation.

Table 17–10

Fuel cell systems used at wastewater treatment facilities[a]

System	Typical power output, kW	Efficiency, %	Advantage	Disadvantage
Alkaline (AFC)	10–100	60	• High performance due to fast cathode reaction in alkaline electrolyte • Low cost components	• Sensitive to CO_2 in fuel and air • High electrolyte management needs
Phosphoric acid (PAFC)	100–400	40	• Suitable for combined heat and power due to high operating temperature • Tolerant to fuel impurities	• Expensive catalyst (Pt) • Long start-up time • Low current and power
Molten carbonate (MCFC)	300–3000	45–50	• High efficiency compared to PAFC • Fuel flexibility • A variety of catalysts could be used • Suitable for combined heat and power due to high operating temperature	• Susceptible to high temperature corrosion and breakdown of cell components • Long start-up time • Low power density

[a] Adapted from DOE (2011).

Sources of Heat

Major sources of thermal energy in wastewater treatment facilities include heat from the combustion processes, as discussed in Sec. 17–5, and sensible heat in wastewater. Excess heat from engine generators and boilers has been used widely to heat water for building heating, anaerobic digesters, or hot water supply. In some wastewater treatment plants, lower quality heat such as the residual heat from the exhaust air stream in heat recovery ventilation systems and the heat in treated wastewater effluent has been recovered and used for various purposes.

Combined Heat and Power (CHP) System. In a CHP system, or cogeneration system, both electricity and usable heat are generated. Typically thermal energy is transformed into hot water or steam and used for space heating, digester heating, drying, and other purposes. The typical range of total system efficiency from internal combustion engines without CHP is between 25 and 50 percent, as reported in Table 17–8, depending on the type of engine and operational conditions, whereas the total system efficiency with CHP can be in the range of 70 to 85 percent.

Low Grade Waste Heat. Low grade waste heat is the heat either coming directly out of combustion processes or after the heat has been recovered from the heat source by other heat recovery systems, and considered not sufficient to generate steam or high temperature hot water (80 to 90°C) directly. The low grade waste heat may have the temperature from 30°C or lower (e.g., wastewater effluent) to as high as 230°C (e.g., exhaust air from existing heat recovery devices), depending on the heat recovery system considered

(DOE, 2008). Devices such as heat pumps and organic Rankine cycle engines have been used to recover energy from low grade waste heat.

Other Sources of Heat. Digested sludge, process air exiting bioreactors, and air in the building exhaust also contain heat that can be used for preheating incoming sludge or process air. Sidestream treatment processes can be a potential source of excess heat as temperature of sidestream tends to be higher than normal wastewater and due to high concentrations of nitrogen, the heat generated from oxidation and reduction of nitrogen is typically significant enough to require dilution of the sidestream or heat removal through a heat exchanger (see Chap. 15). However, the feasibility of heat recovery and its use at a specific treatment facility has to be evaluated as the cost savings from the use of recovered heat may not be justifiable based on the installed cost of the heat recovery equipment and conveyance of recovered heat to the point of heat demand or use.

Demands for Heat

Heating requirements in wastewater treatment facilities are primarily for sludge and biosolids processing, building heating, and generation of hot water for various uses. Heating requirements in solids processing include anaerobic digester heating to maintain the operating temperature and for drying sludges and biosolids. Effluent disinfection by pasteurization, a relatively new development for wastewater applications, also requires significant amount of heat (see Sec. 12–10 in Chap. 12).

Heating and Cooling of Buildings. Typically, heating, ventilation, and air conditioning (HVAC) requirements are assessed during the design of building structures and a dedicated HVAC system is installed to maintain the climatic condition in the buildings. For high level assessment, the estimates on heating demand may be developed from published energy data. More detailed heating demand estimates are typically developed using microclimate-specific energy models. Major heating requirements must be identified and heat recovery considered during the design phase when heat supply for major facilities is considered. Smaller heat recovery systems may be considered separately for specific heating needs and available heat. In rare cases, excess heat at a wastewater treatment plant may be exported to outside of the facility, either to a specific building or to be integrated with a district heat system. Typically, however, the amount and quality of heat is not significant enough for energy input to district heating systems and the use is limited to within the treatment facility as the piping distances are much shorter.

Digester Heating. In conventional mesophilic anaerobic digestion, the heating of the digester is the most significant heating demand. The heating demand in sludge processing will increase significantly with thermophilic anaerobic digestion and also when thermal hydrolysis is applied for the pretreatment of the sludge. Heating of the feed sludge up to the operating temperature is the largest heating demand, and additional energy may be required to maintain the digester temperature, depending on the heat loss from the digester walls, floor, and the roof (see Example 13–7 in Chap. 13).

Drying. Sludge and biosolids drying is an energy intensive operation, and it can raise the treatment facility's energy demand significantly. Heating demand is determined based on the temperature of incoming sludges/biosolids and initial and final water contents. The heat transfer rate is estimated from the heat-transfer coefficient, contact area of heat source, and the temperature difference between the sludge and heating medium as expressed in Eqs. (14–6) and (14–7) (See Sec. 14–3 in Chap. 14). Depending on the heat recovery system employed, the heat required for drying can be supplied by combusting the dried sludge. In Example 17–6, the heat balance for a sludge drying process is presented.

EXAMPLE 17–7 **Computation of Heat Balance for Sludge Drying Process** Two different sludges and anaerobically digested biosolids are to be dried and subsequently combusted to generate heat for a sludge drying process. Heat content of the sludges and biosolids are presented below. Assume the sludge drying process requires 3.5 MJ to remove 1 kg of water from sludge, and dried cake has 90 percent solids content. Determine the initial water content above which heat content of the solids is sufficient for the drying process. Assume the heat loss through the transfer of heat from the combustion process to sludge drying process is inclusive of the heating requirement (3.5 MJ/kg).
Latent heat of water vaporization is 2.257 MJ/kg.

Type of feed	Heat content, MJ/kg
Primary sludge	11.6
Waste activated sludge	14.0
Anaerobically digested biosolids	16.3

Solution

1. Calculate the mass of wet cake for 1000 kg dry solids and solids contents between 15 and 35 percent.
   For 15 percent solids content, the mass of wet cake is

   Wet cake = (1000 kg)/0.15 = 6666.7 kg

   Set up a spreadsheet to calculate the mass of wet cake between 15 and 35 percent solids content.
2. Calculate the amount of water to be evaporated and the heat required to dry wet cake from the initial solids content to 90 percent.

   Mass of 90 percent solids content cake = (1000 kg)/0.90 = 1111.1 kg

   The amount of water to be evaporated for 15 percent cake is

   Water to be evaporated = 6666.7 − 1111.1 = 5555.6 kg

   Heat required to evaporate = 5555.6 × 3.5 = 19,444 MJ

   Similarly, calculate the amount of water to evaporate for the range of solids content and compute the heat required to evaporate water.
3. Calculate heat generated from combustion of the 90 percent solids.
   For primary sludge, 1000 kg of dry solids will generate 1000 × 11.6 = 11,600 MJ
   The remaining water (10 percent, or 111.1 kg) will be evaporated.

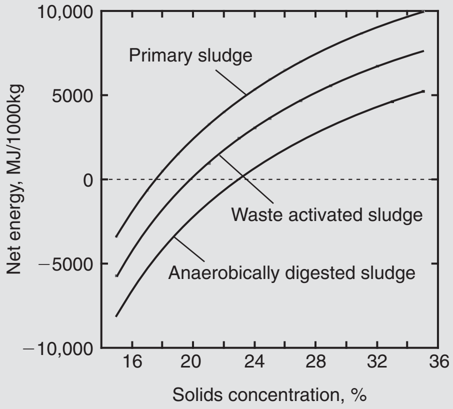
   The latent heat of vaporization is 111.1 × 2.257 = 250.8 MJ

   Heat generated from the combustion of 90 percent solids cake = 11,600 − 250.8 = 11,349 MJ.

4. Calculate the net heat balance and plot the results for the three types of sludge for the range of solids content.
   For primary sludge with 15 percent solids content, the heat balance is

   Net heat generated = 11,349 MJ − 19,444 MJ = −8905 MJ.

   Similarly, calculate the heat balance and plot the results for primary sludge, waste activated sludge, and digested sludge. The result is shown on the following plot.



Using the solver function on the spreadsheet program, the solids contents to achieve zero net energy are 20.0, 19.8, and 17.6 percent for primary sludge, waste activated sludge, and digested sludge, respectively.

Comment Depending on the initial solids concentration and efficiency of the solids combustion and drying process, heat generated from combustion may be sufficient to operate the drying process without supplemental fuel. Detailed evaluation of the thermal efficiency was not included in this example, but a simplified evaluation is often used for a preliminary assessment.

Devices for Waste Heat Recovery and Utilization

Suitable heat recovery applications are determined based on the thermal energy contents and the type of output (e.g., electricity, hot water, hot air, cold water, cold air). Examples of devices used at wastewater treatment facilities to recover and utilize heat include heat exchangers, heat pumps, and heat absorption chillers. Organic Rankine cycle engines have also been used to recover energy from waste heat to produce electricity.

Heat Exchanger. A heat exchanger is a device that transfers heat from one source (hotter fluid) to another (colder fluid) through a conductive material that separates fluids with different temperatures. Heat exchangers are used widely at wastewater treatment plants, especially where anaerobic digesters are used to stabilize sludges and produce digester gas. Exhaust heat from the water jacket of engine generators is often used for digester heating and heating of buildings. Various types of heat exchangers have been installed at wastewater treatment plants. Common heat exchanger types are (1) coil heat exchangers, (2) plate heat exchangers, (3) shell-and-tube heat exchangers [see Fig. 17–13(a)], and (4) spiral heat exchangers [see Fig. 17–13(b)].

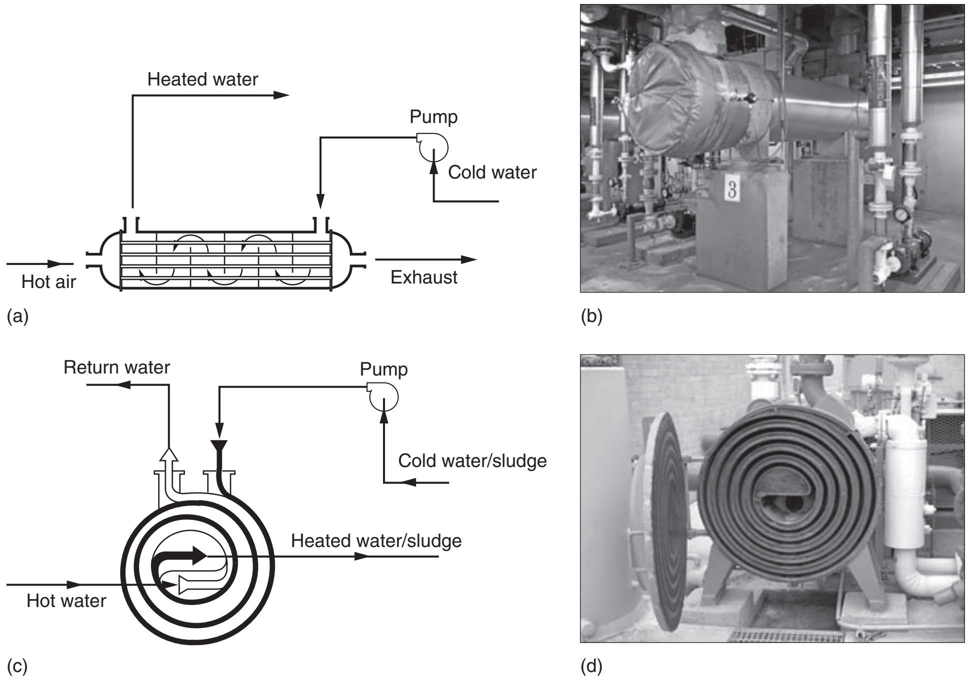
A simplified schematic of a shell-and-tube heat exchanger is shown on Fig. 17–14. In simpler forms, heat exchanger configurations can be categorized as cocurrent [see Fig. 17–14(a)] and countercurrent [see Fig. 17–14(b)]. Overall heat transfer with a shell-and-tube heat exchanger is expressed as

$$Q = UAF\Delta T \tag{17–18}$$

where Q = total heat load transferred, MJ/h (Btu/h)
U = heat transfer coefficient, MJ/m^2·h·°C (Btu/ft·h·°F)
A = surface area, m^2 (ft^2)

Figure 17–13

Typical heat exchangers used at wastewater treatment facilities: (a) schematic of shell-and-tube heat exchanger for the recovery of heat from engine exhaust, (b) view of a shell-and-tube heat exchanger, (c) schematic of spiral heat exchanger for sludge heating, and (d) view of inside of a spiral heat exchanger.



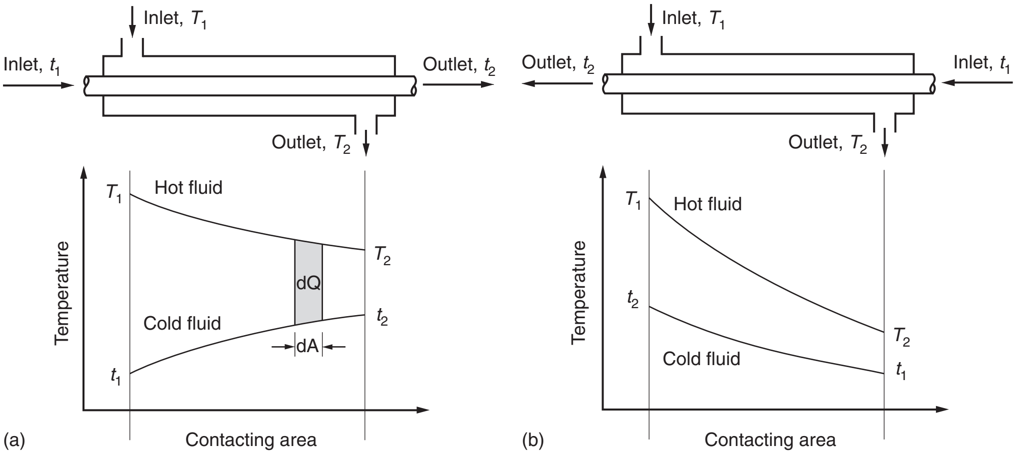


Figure 17–14

Simplified diagram of heat exchangers and corresponding temperature gradients: (a) cocurrent type and (b) counter current type.

F = correction factor for specific flow arrangements within the heat exchanger, unitless
ΔT = mean temperature difference, °C (°F)

Heat Transfer Coefficient. The heat transfer coefficient depends on the conduction heat transfer coefficients for the materials used, and to a lesser degree the heat convection resistance (Holman, 2009). A detailed analysis of heat exchangers can be found in references on chemical engineering (for example, Green and Perry, 2007). The typical range of heat transfer coefficients for various materials are reported in Table 17–11. The correction factor, F, for various heat exchanger configurations could be found in references on heat exchangers, such as Holman (2009), Kuppan (2000), and others.

Table 17–11

Typical heat transfer coefficients for various materials[a]

Heat exchanger type and media	Heat transfer coefficient, U W/m²·K
Steam condenser	1100–5600
Feed water heater	1100–8500
Water to water heat exchanger	850–1700
Finned-tube heat exchanger, water in tubes, air across tubes	25–55
Water-to-oil heat exchanger	110–350
Finned-tube heat exchanger, steam in tubes, air over tubes	28–280
Ammonia condenser, water in tubes	850–1400
Alcohol condenser, water in tubes	255–680
Gas-to-gas heat exchanger	10–40

Adapted from Hewitt (1992).

Temperature Difference. The temperature difference between hot fluid and cold fluid changes along the heat exchanger. Temperature profiles of two fluids in a cocurrent flow double-pipe heat exchanger are illustrated on Fig. 17–14(a). Through an incremental area, dA, heat dQ is transferred, and the total heat transferred across the heat exchanger is an integral of the incremental heat transfer. Solving for the total energy transfer, the mean temperature difference, ΔT_{LM}, in Eq. (17–18) for a double-pipe heat exchanger can be written as

$$\Delta T_{LM} = \frac{(T_2 - t_2) - (T_1 - t_1)}{\ln[(T_2 - t_2)/(T_1 - t_1)]} \tag{17–19}$$

where T_1, T_2 = temperature of hot fluid, °C
t_1, t_2 = temperature of cold fluid, °C

The temperature difference expressed by Eq. (17–19) is termed *log mean temperature difference* (LMTD). The equation is also valid for counter current flow [see Fig. 17–14(b)]. When the LMTD is used to calculate the heat transfer for other configurations a correction factor, F, is used to estimate the mean temperature difference. Thus, the overall heat transfer equation is described as shown previously in Eq. (17–18).

Fouling Considerations. When selecting heat exchangers, the designer must take into account the fouling factors which will impede the heat transfer. Consideration of the fouling factor, is a critical design factor, especially when the fluid is not within a closed loop and is not a clean fluid. Generally, selection of a heat exchanger would be determined by the cost and specifications by manufacturers providing packaged heat exchangers. The characteristics of the fluid from which heat is recovered, including the type of fluid, temperature, any constituents in the fluid, and desired heat recovery are put together and manufacturers are consulted. Critical information that would determine the feasibility of heat recovery and the type of heat exchanger includes the heat transfer requirements, capital cost, physical size, pressure drop across the heat exchanger, and ease of cleaning. Design of a heat exchanger that meets the heat transfer requirements will depend on the relative weights of cost versus pressure drop and physical size limitations (Holman, 2009). The use of Eq. (17–18) and Table 17–11 would be useful in conducting a preliminary assessment before detailed information is collected for design.

Heat Pump. The heat pump is a device that uses a refrigerant to take heat from one source typically at a lower temperature and transfer it to another medium which usually has a higher temperature (DOE, 2003). A simplified schematic is shown on Fig. 17–15. Briefly, the low temperature-low pressure refrigerant is vaporized using a heat source. The vaporized refrigerant is compressed to a high pressure-high temperature vapor by a compressor which requires energy (electricity). It then goes through another heat exchanger to transfer the latent heat of vaporization to a heating medium as the refrigerant is condensed to liquid. The refrigerant that is condensed to liquid still retains relatively high temperature. In the expansion valve, the temperature and pressure of the refrigerant is lowered. The low pressure, low temperature refrigerant goes back to the first step to be vaporized by taking up heat from wastewater effluent. The heat transferred to a medium could be used for heating of buildings, water, digesters, and other purposes.

When wastewater is used as a heat source, usually an intermediate medium such as propylene glycol is used to take heat from wastewater by a heat exchanger suitable for wastewater, and glycol is used to operate the heat pump. The intermediate step is recommended to avoid fouling of the heat exchanger within the heat pump. The use of a heat pump for heating can also provide a side benefit by lowering the temperature of the effluent to be discharged to receiving water that is sensitive to temperature difference. The most common use of heat pump at wastewater treatment facilities is for space-heating and ventilation preheating.

The same principle for a heat pump could also be used on an opposite direction to cool a medium. A refrigerator is an example of a cooling cycle heat pump. In a warm climatic condition, wastewater can be used as a heat sink (i.e., heat taken out for cooling is disposed) in a heat pump operation to provide space cooling. In a large scale system, both heating and cooling cycles could be used so that the heat pump system could be utilized throughout a year. A district energy system installed in Tokyo is such an example (Funamizu et al., 2001). It is a key design consideration to determine if the heat pump system would be used only for heating, or used for both heating and cooling.

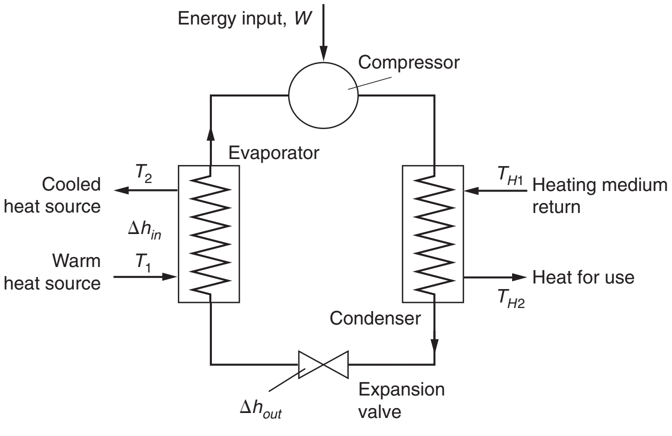
Heat Balance. Referring to Fig. 17–15, the heat transferred from the effluent, Δh_{in}, J/s, is

$$\Delta h_{in} = m_o C_o (T_1 - T_2) \tag{17–20}$$

where m_o = mass flowrate of effluent, kg/s
C_o = specific heat of heat source, J/kg·°C
T_1 = heat source temperature entering the heat pump, °C
T_2 = heat source temperature exiting the heat pump, °C

Figure 17–15

Schematic illustration of a heat pump system.



The heat transferred from the heat pump to the heating water, Δh_{out}, is

$$\Delta h_{out} = m_{\mathrm{H}} C_{\mathrm{H}} (T_{\mathrm{H2}} - T_{\mathrm{H1}}) \tag{17–21}$$

where m_{H} = mass flowrate of heating medium, kg/s
T_{H1} = heating medium temperature entering the heat pump, °C
C_{H} = specific heat of heating medium, J/kg·°C
T_{H2} = heating medium temperature exiting the heat pump, °C

Energy loss through the piping, h_c, is

$$h_c = h_f + h_a = g m_e H(2cL) + UA\Delta T \tag{17–22}$$

where, h_f = energy loss due to friction headloss, J/s
h_a = heat loss through the pipe, J/s
g = acceleration of gravity, m/s^2
H = headloss per unit pipe length, m/m
c = friction loss per unit pipe length, m/m
L = distance between heat pump and heat delivery point, m
Other terms were defined previously [see Eq. (17–18)].

If the electricity used to compress the refrigerant is W_1 (J/s) and the compressor efficiency is E_c, the energy balance is as follows:

$$m_o C_o (T_1 - T_2) + E_c W = m_{\mathrm{H}} C_{\mathrm{H}} (T_{\mathrm{H2}} - T_{\mathrm{H1}}) + h_c \tag{17–23}$$

where E_c = compressor efficiency, unitless
W = energy input, J/s

Coefficient of Performance. Coefficient of performance (COP) is an expression used to quantify heat pump performance in terms of the heat delivered from a unit of energy input.

$$\mathrm{COP} = h_h/h_w \tag{17–24}$$

where, COP = coefficient of performance, unitless
h_h = heat output from the heat pump, Joule (or J/s)
h_w = energy input to the heat pump, Joule (or J/s)

Using the terms shown previously, COP can also be written as

$$\mathrm{COP} = \frac{m_{\mathrm{H}} C_{\mathrm{H}} (T_{\mathrm{H2}} - T_{\mathrm{H1}})}{W} = E_c W + C_o(T_2 - T_1) - h_c/W \text{ (use the same format as the term before)} \tag{17–25}$$

The level of COP achievable for the specific application will depend on the climatic condition and the type of application. Therefore a direct comparison of COP values between different applications does not provide meaningful information, but the estimate of COP for alternatives may be useful for a decision making process. In relatively cold climate, such as central Canada, heat recovery from wastewater effluent could provide sufficient energy to heat a small building within a wastewater treatment facility, and typical COP would be in the range of 3 to 4, depending on the pumping power and fluid temperatures. In more temperate regions, the use of heat pump for both heating and cooling can provide a significant contribution to energy savings. In some cases, higher COP values are achievable with cooling (i.e., use wastewater effluent as a heat sink).

Waste Heat Utilization for Cooling and Refrigeration. Low grade waste heat, such as exhaust heat of engine generators that has gone through a series of heat exchangers, could be utilized to operate heat absorption chillers or other technologies to provide cooling and refrigeration (see Fig. 17–16). In absorption chillers, lithium bromide is used

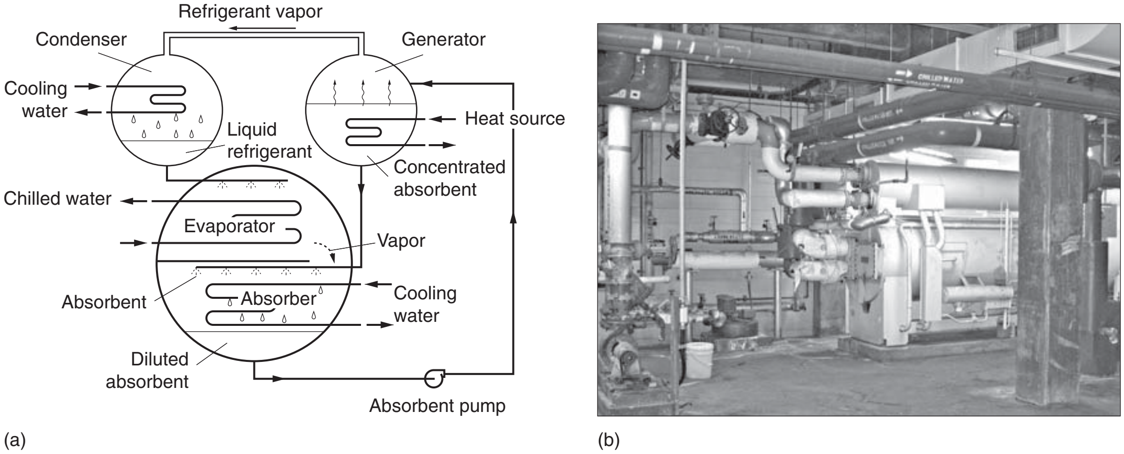


Figure 17–16

Chiller using low grade heat: (a) schematic of chiller composed of generator, condenser evaporator and absorber and (b) view of typical chiller (courtesy of Philadelphia Water Department).

typically to absorb the vaporized water (refrigerant) to create a vacuum environment. Water absorbed in the lithium bromide solution is vaporized with the low grade heat, and concentrated lithium bromide solution is recirculated. Water vapor is sent to a condenser, and sent back to the vacuum environment, where it boils at a low temperature and evaporative cooling is used to produce chilled water. The quality of energy source suitable for the absorption chillers varies with the chiller types, but suitable heat source is typically between 70 and 130°C.

The performance of absorption chiller or other cooling and refrigerating systems can be quantified in a similar manner as the heat pump, in terms of COP. The range of COP depends on the energy source and the cooling requirement. When hot water is used as a heat source, the COP value can be 4.0 or higher. Because the design of chillers will involve evaluation of the heat source, chilled water temperature, cooling water temperature and other site specific conditions, manufacturers of chillers should be consulted and a computer modeling is used to determine the sizing and operating conditions.

Organic Rankine Cycle Engine. The organic Rankine cycle (ORC) engine is an electrical power generator utilizing an organic refrigerant. The mechanism of ORC engines is similar to that of the heat pump, and a generator is used in place of the expansion valve. The ORC is capable of generating electricity from waste heat as low as 67°C (150°F), and the efficiency is generally 10 to 20 percent (DOE, 2008).

Design Considerations for Thermal Energy Recovery Systems

Evaluation of the characterization of waste heat is an important step to determine the feasibility of thermal energy recovery and the selection of the recovery systems. Important parameters that must be determined include (DOE, 2008):

1. Heat quantity
2. Heat temperature/quality
3. Composition
4. Minimum allowed temperature
5. Operating schedules, availability, and other logistics
6. Diurnal and seasonal variations in the parameters identified above

Heat Recovery from Combustion Systems. Waste heat from combustion systems such as exhaust heat from engine generators has sufficiently high temperature to generate hot water or steam. At treatment facilities utilizing digester gas for power generation, it is common to design the system to utilize waste heat for other purposes through a CHP system (see Chap. 13).

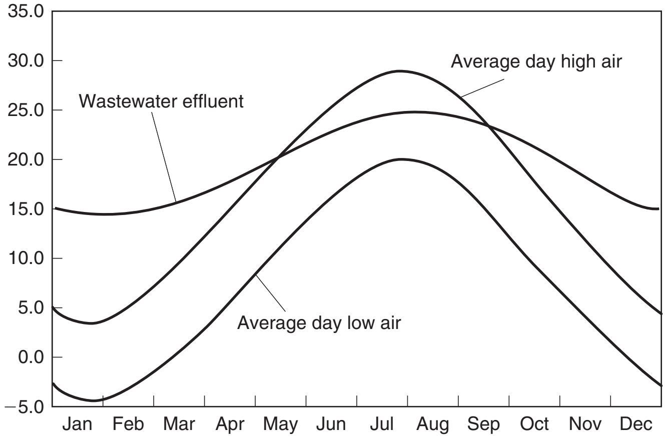
For a biosolids incineration system, the applicability and selection of a CHP technology is based on the amount of biosolids to be processed and the amount of excess energy in the hot flue gas generated. Another factor which can affect the selection of an appropriate type of CHP is the size of the facility and the goals for energy recovery. Small to medium size facilities, with a goal of producing electricity may use a technology such as the ORC with ~10–20 percent electrical efficiency to recover energy from excess flue gas generated. Larger facilities with similar goals may select high pressure steam turbines with approximately 15–38 percent electrical efficiency. Economic and operational issues must be evaluated to determine if an add-on CHP system is practical and what the appropriate size and type of system should be.

Heat Recovery from Wastewater. The temperature of wastewater is typically higher than that of potable water, and the variation in wastewater temperature is smaller than that of ambient temperature (see Fig. 17–17) throughout a year. The heat from wastewater is a reliable source of thermal energy for beneficial uses during the colder season, and it can be used as a heat sink during the warmer season. Depending on the size of the treatment facility, thermal energy can be used to supplement heating requirements within the wastewater treatment facility, or integrated into a district heating/cooling system. Because the available heat in wastewater is low, a heat pump is used to extract heat from wastewater (Pallio, 1977). Depending on the availability of heat, heating of a specific building within a wastewater treatment facility may be considered, or it may be integrated into the centralized heating system for the entire treatment plant. The estimation of heat to be recovered from wastewater is illustrated in Example 17–7. Heat recovery from raw wastewater in the collection system or at the beginning of the treatment facility can also be considered. When heat is recovered from raw wastewater, the impact of lowered wastewater temperature on the treatment performance must be evaluated (Wanner et al., 2005).

In North America, heating or cooling capacity of air conditioning equipment is often measured in terms of the tons of refrigeration. One ton of refrigeration is approximately 12,000 Btu/h or 3.517 kW. Similarly, the power requirement for the heat extraction is often expressed in brake horse power (bhp) per ton of refrigeration. One horse power (hp) is approximately 0.745 kW. Thus, if a power requirement for a heat pump is rated as 1.0 bhp/ton, then 0.745 kW is required to provide 3.517 kW of heat extraction.

Figure 17–17

Seasonal variation of ambient temperature and wastewater temperature in north eastern United States.



EXAMPLE 17–8 **Heat Pump for Effluent Heat Recovery** A water to water heat pump is to be used to provide heating in a building at a wastewater treatment plant. The heat source is treated effluent. The lowest effluent temperature is 12°C, and temperature drop of the effluent in the shell and tube heat exchanger is assumed to be 4°C. A 40 percent propylene glycol mixture is used as an intermediate heat transfer medium to extract heat from the effluent, and the heat pump will have an 8°C entering temperature and a 5°C leaving fluid temperature. The peak heating requirement for the building was estimated to be 360 kW.

A heat pump supplier was contacted and a heat pump was selected requiring 0.222 kW electrical input per kW heat output to extract 300 kW from the effluent.

1. Calculate the COP for the heat pump, and the heat output and the electrical power input in kW of the heat pump. For simplicity, assume compressor efficiency = 1
2. Three circulation pumps, 14, 11, and 15kW, are used to transfer effluent to the heat exchanger and to pump the glycol on each side of the heat pump system. Calculate the overall system COP including pumping power.
3. Given the energy to be extracted from wastewater and assumed temperature drop, determine the wastewater flowrate that needs to be transferred to the heat pump system.

Solution

1. Calculate the COP for the heat pump using Eq. (17–24).

   Power input = 0.222 kW per 1 kW of output

   $$\text{COP} = \frac{1.0}{0.222} = 4.5$$

   Using Eq. (7–25) and ignoring the heat loss from the transfer of the heating medium,

   $$4.5 = \frac{W + m_o C_o (T_1 - T_2)}{W} = \frac{W + 300}{W}$$

   Power input = W = 86 kW

   Using Eq. (17–23),

   Total power output = 300 kW + 86 kW = 386 kW

2. Calculate the COP for the entire system.

   $$\text{COP} = \frac{386\,\text{kW}}{(86\,\text{kW} + 14\,\text{kW} + 11\,\text{kW} + 15\,\text{kW})}$$
   $$= 3.06$$

3. Calculate the wastewater flowrate to be transferred to the heat pump system.

   300 kW = 300 kJ/s

   Specific heat of water = 4.2 kJ/kg·°C

   Assume density of water = 1.0

   $$\text{Wastewater flowrate required, m}^3\text{/s} = \frac{(300\,\text{kJ/s})}{(4.2\,\text{kJ/kg·°C})(1000\,\text{kg/m}^3)(4°\text{C})}$$
   $$= \frac{300}{4.2 \cdot 1000 \cdot 4} = 0.0179\ \text{m}^3\text{/s} = 1.07\ \text{m}^3\text{/min}$$

Comment Additional energy required to convey (i.e., pump) wastewater and heated water or glycol can be significant if the heat extracted by a heat pump is used for a large area or outside of the wastewater treatment facility (e.g., for district heating system).

17–8 RECOVERY AND UTILIZATION OF HYDRAULIC POTENTIAL ENERGY

Utilization of hydraulic potential energy involves conversion of hydraulic potential energy into electricity, and transferring of hydraulic pressure in a stream to different parts of the treatment process. The purpose of this section is to introduce (1) the type of devices used to recover energy from hydraulic potential, (2) the hydraulic energy recovery applications adopted by wastewater treatment facilities, and (3) considerations for the selection and design of hydraulic energy recovery systems.

Type of Hydraulic Potential Energy Recovery Devices

Hydraulic turbines for electrical power generation are the primary machines used for the recovery of hydraulic potential energy from the main liquid process flow. Devices to utilize the pressure head or kinetic energy of the stream are usually integrated into a specific energy-intensive unit process to reduce the energy requirement of the process. Electrical power generated from hydraulic potential energy can be sent to the electrical distribution system within the treatment facility to supplement the electricity drawn from the electrical grid or directly connected to the grid and sold to the electric utility.

Hydraulic Turbines. Turbines are turned by water flow, and the movement of the turbine is directed to the attached generator to produce electricity. Hydraulic turbines to generate electricity are essentially pumps in reverse. An electrical generator connected to the turbine produces electricity rather than an electric motor which consumes electricity to move water. The type of turbine follows the flow and head characteristics of the related pumps. The head and flow ranges for most wastewater treatment facilities would place the hydroturbines generators in the small ($<$ 5 MW) or micro ($<$ 250kW) range.

The two types of hydraulic turbines are: (1) reaction turbines, and (2) impulse turbines. In the reaction turbines, water pressure is applied to the face of the runner blades to turn the turbine. Radial (Francis) turbines, and axial type (Kaplan and propeller) turbines are the common reaction turbines. In the impulse turbines, water entering the turbine has sufficient velocity (i.e., kinetic energy), and turn the turbine as a high velocity water hits the buckets on the periphery of the runner. Common impulse turbines include Pelton turbines [see Figs. 17–18(a) and (b)], Turgo turbine, and cross-flow turbines (ESHA, 2004). Impulse turbines exhaust to atmosphere and are not suitable for energy recovery from force mains or where the continuity of the water column must be maintained through the turbine to capture the available head. The application ranges for various types of hydraulic turbines are graphically presented on Fig. 17–19. The Pelton and Turgo are the best fit for low flow and high head conditions. The axial flow Kaplan-type turbines are suitable for low head, high flow conditions that would be typical of wastewater treatment plant effluent streams. Crossflow turbines are used most commonly for small scale applications, and suitable for low head applications.

Devices for the Transferring of Pressure Head. Hydraulic turbochargers and pressure exchangers used for the nanofiltration and reverse osmosis systems are the examples of devices used to reduce energy requirements by transferring hydraulic potential energy from a stream to the other part of the process. Hydraulic turbochargers utilize the hydraulic head in the membrane reject stream to boost the inlet pressure for the membrane modules. The use of energy recovery devices in membrane treatment systems are discussed in detail in Sec. 11–6 in Chap. 11.

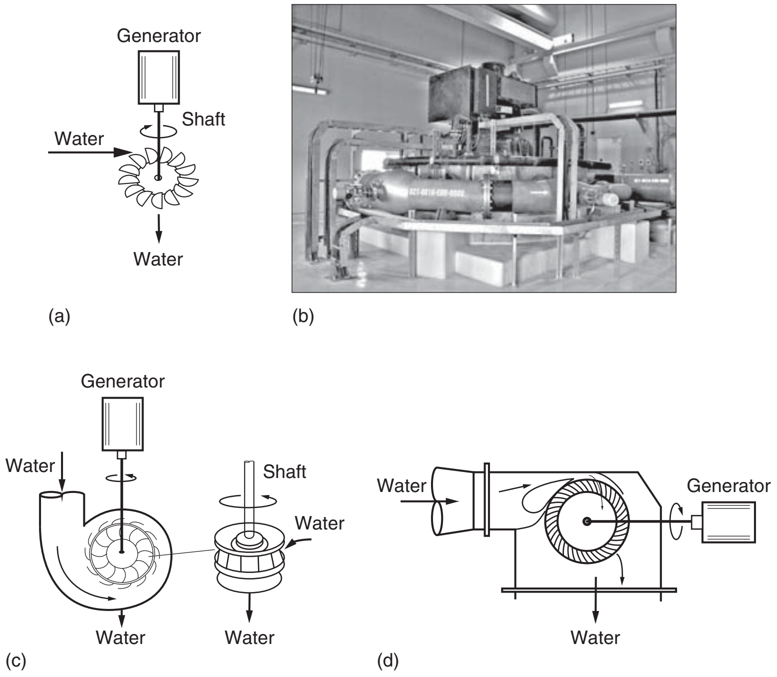


Figure 17–18

Examples of hydraulic turbines: (a) schematic of a pelton wheel turbine, (b) view of pelton turbine used with raw wastewater for power generation in Jordan, (c) schematic of Francis turbine, and (d) schematic of crossflow turbine. (Adapted from BHA, 2005.)

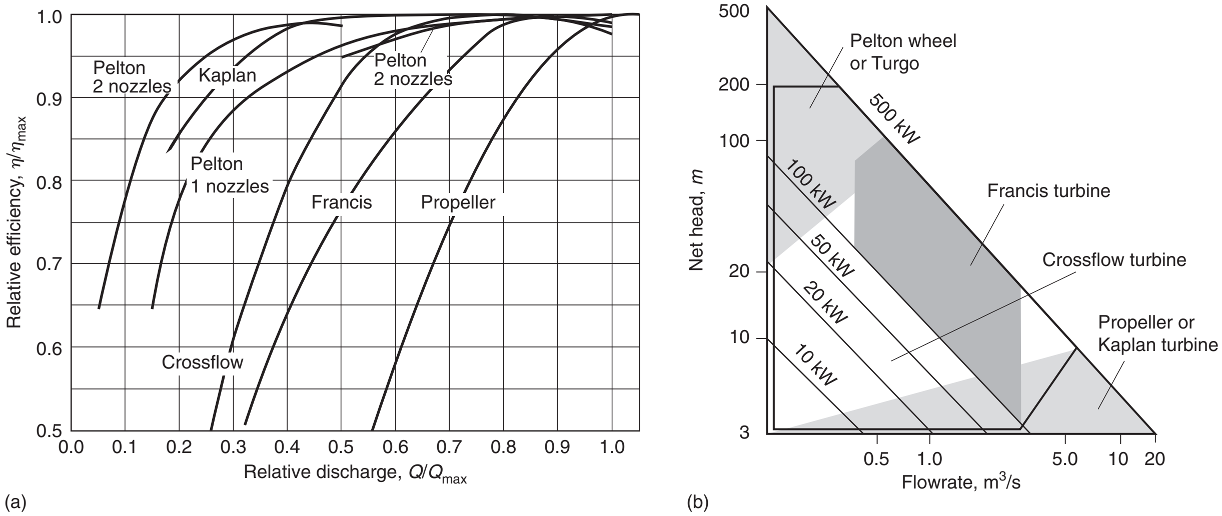


Figure 17–19

Guidelines for the selection of hydraulic turbine: (a) typical turbine efficiency for various turbine types, and (b) range of operation for hydraulic turbine based on the net head and flowrate (adapted from the BHA, 2005, and ESHA, 2009).

Application of Hydraulic Energy Recovery Devices

The two most common applications to recover hydraulic potential energy from wastewater treatment facilities are (1) recovery of hydraulic potential at the influent or effluent of the wastewater treatment plant and (2) recovery of residual head from high-pressure membrane treatment processes.

Recovery of Hydraulic Potential and Net Head. The hydraulic energy that can be converted to electrical energy from a turbine-generator system is defined previously as.

$$P_e = \rho Q g \Delta H \eta_t \eta_e \qquad (17\text{–}8)$$

Typical turbine efficiency is reported on Fig. 17–19(a). Typically, generator efficiency for new generators is between 90 and 95 percent. In practical applications, additional efficiency factors may need to be considered. For example, where power is to be generated using a turbine and generator combination, other factors such as penstock efficiency and transformer efficiency must be considered to account for the efficiency for the entire electrical power generation system.

Selection of Hydraulic Turbines. The key criteria in selecting the type, geometry, and dimensions of the turbine include the following (ESHA, 2004):

1. Available head
2. Flowrate range
3. Variation in tailwater elevation
4. Cost
5. Available space and access

Generally, impulse turbines are suitable for high head, low flowrate streams, and reaction turbines are suitable for low head, high flowrate streams.

Design Considerations for Hydraulic Turbines for Wastewater. When a hydraulic turbine is used with raw wastewater, the turbine is located downstream of preliminary treatment to remove large debris, and wastewater flows through a penstock to the turbine, just before the main treatment facility. Pelton turbines may have an advantage for raw wastewater as they are less prone to the damage by small debris. However, Pelton turbines require significant head, in form of kinetic head, and are not suitable for typical wastewater treatment plants.

For the recovery of hydraulic energy from the treated effluent before discharge to the receiving water, the use of low head reaction type turbines such as Francis and Kaplan turbines, or crossflow turbines for small scale applications, are more applicable than the impulse turbines. Because wastewater effluent at the end of the treatment processes will have minimal kinetic energy, the turbine may have to be located close to the discharge point to maximize the head. General guidance for the selection of turbine types is provided on Fig. 17–19. It is important to understand the range of flowrate and the available head in determining the appropriate turbine. The ranges of net head, H_n, for major types of hydraulic turbines are presented in Table 17–12.

Table 17–12

Typical range of net head for major hydraulic turbine types[a]

Turbine type	Range of net head, H_n, m
Kaplan and propeller	2–40
Francis	25–350
Pelton	50–1300
Crossflow	2–250

[a] Adapted from ESHA (2004).

Recovery of hydraulic head from wastewater plant effluent flow will not fit the typical range of turbines used for hydroelectric generating facilities. The recovery of energy from a large WWTP (432 ML/d) with 7 meters of available head is illustrated in Example 17–9.

EXAMPLE 17–9 **Power Generation from Wastewater Discharge Location** Estimate the power that can be generated by a hydro-power generation unit located at the effluent discharge for the following conditions. Also, select the type of turbine that would be most suitable the given conditions.

Parameter	Unit	Value
Available head	m	7
Flowrate	m^3/s	5
Penstock efficiency	%	0.92
Turbine efficiency	%	0.8
Generator efficiency	%	0.92
Transformer efficiency	%	0.95

Solution

1. Estimate the the power that can be generated using Eq. (17–8).

   $$P_e = (1000)(5)(9.81)(7)(0.8)(0.92) = 2.53 \times 10^5 \text{ W} = 253 \text{ kW}$$

2. Selection of turbine type.
   From Fig. 17–19(b), for the flow and available head, a Kaplan or propeller turbine could be used.

Comment Consideration should be given to looking at the hydroturbine as a pump in reverse and coupling the appropriate pump to a generator. The reverse pump may not be as efficient as a turbine designed specifically for hydropower, but it may be difficult to find a hydroturbine in the flow and head ranges seen at wastewater treatment facilities. Note that flywheels are often used to prevent freewheeling of the turbine if the connection to the electrical system is interrupted.

Use of Residual Pressure Head in Treatment Processes

For industrial and indirect/direct potable reuse applications, reverse osmosis has become a common treatment unit process used to remove dissolved constituents that are otherwise difficult to remove. In water reuse applications, typical feed water pressure for the RO unit is in the range of 12 to 18 bar, and the concentrate (brine) will retain the pressure ranging 10 to 16 bar. Because transfer of the brine usually does not require as much head to the discharge point or to the brine storage, the residual pressure is often used to recover energy. Devices used for the recovery of energy from the residual pressure sources were discussed previously in Sec. 11–6. Typically, the energy recovery from RO units are considered economically viable for larger treatment plants as the energy savings is close to proportional to the flowrate whereas the additional costs for the energy recovery devices are not proportional to the size of the units.

17–9 ENERGY MANAGEMENT

Important energy management considerations include maximizing the use of available energy as discussed in previous sections, reducing the use of energy at wastewater treatment facilities, increasing the energy production onsite by utilizing waste from other sources, and utilizing renewable energy such as solar or wind power. In addition, a wide range of management goals must be evaluated, including (NYSERDA, 2010)

1. Improving energy efficiency and managing total energy consumption
2. Controlling peak demand for energy
3. Managing energy cost volatility
4. Improving energy reliability

In addition to these management goals, greenhouse gas reduction has become a management goal for wastewater treatment facilities. Assessment of greenhouse gas emissions is described in Sec. 16–6 in Chap. 16. Topics considered in this section include (1) process optimization and modification for energy saving, (2) process modification for increased energy production, (3) peak flowrate management, and (4) selection of energy sources. It should be noted that the topics covered in this section are illustrative of what can be done to improve energy management, and are not meant to be inclusive of all energy management opportunities. Additional information on energy management can be found in a number of publications including WEF (2009), and U.S. EPA (2008, 2010, 2012).

Process Optimization and Modification for Energy Saving

Opportunities to improve energy efficiency in existing wastewater treatment facilities are identified through the energy audit described in Sec. 17–4. Typically, the evaluation following the energy audit involves process optimization and modification. Examples of process optimization and modifications for energy saving are summarized in Table 17–13. Energy saving may be achieved through operational changes to existing processes or a modification to processes and equipment. Even though the opportunities for energy saving from operational changes are limited, they can be made with little or no capital cost. Process modifications and the use of energy efficient equipment could make improvements in energy management, but may require significant capital expenditures. A life cycle cost analysis should be conducted for the shortlisted options to determine if the energy savings generated by making the change justify the capital cost of the change. To illustrate the energy reduction through process modifications, energy saving opportunities with large pumps and activated sludge aeration systems are considered below.

Major Pumps at Wastewater Treatment Facilities. Many wastewater treatment facilities require raw wastewater be pumped at the headworks to provide sufficient hydraulic head for wastewater to flow through the rest of the liquid treatment processes by gravity. These pumps, often identified as *main sewage pumps (MSPs)*, are typically the largest pumps at a wastewater treatment plant. The MSPs are required to handle a wide range of flowrates through diurnal variations of the flowrate as well as the wet weather flows. Treatment facilities designed for nutrient removal typically have significant mixed liquor or return activated sludge (WAS) flows. In MBR systems, the recycle flow is often 6 times the influent wastewater flowrate to maintain high MLSS concentrations (see Chap. 8). Energy efficiency in these major pumps can affect the overall energy efficiency of the treatment facility significantly.

Table 17–13

Example of process optimization and modifications to manage energy consumption

Operational modification	Expected outcome	Potential issues
Main sewer pump control	Reduced energy requirement	Modification to the pumps may be necessary
Wastewater loading distribution by flow equalization	Reduced peak-hour energy use	
DO control in activated sludge	Reduced power consumption Increased process performance reliability	Usually require online monitoring of DO and other parameters Increased monitoring requirements Potential process upset if online analyzers are not working properly
Conversion of nitrification to nitritation or simultaneous nitrification/denitrification	Reduced oxygen requirement	precise process control is required for nitritation
Use of energy efficient diffusers	Reduced air flow requirement	Air flow distribution should be checked
Use of energy efficient blowers	Reduced energy requirement	Some energy efficient blowers may not be suitable for deep reactors due to pressure head limitations
Use of energy efficient UV lamps	Reduced energy requirement	
Energy recovery system for RO	Reduced energy requirement	
Use of off-peak hours for biosolids dewatering	Reduced peak-hour energy use	

The overall efficiency of a pumping system is defined as

$$E = E_p \times E_m \times E_c \tag{17–26}$$

where E = overall efficiency
E_p = pump efficiency
E_m = motor efficiency
E_c = control efficiency

The relative efficiency of each of the three elements identified in Eq. (17–26) is reported in Table 17–14. Pump and motor efficiencies depend on the selection of specific models, whereas the control efficiency can vary depending on the type of the control device used. The same principle applies to the control of blowers as described in the following discussion.

Variable Frequency Drive. The flow control efficiency depends on the type of control used, and the use of variable frequency drives (VFDs) is the most energy efficient, corresponding to the higher end of the range of control efficiency reported in Table 17–14. Operationally, VFDs manipulate the frequency of the alternating current to vary the speed

Table 17–14

Pump systems used for wastewater treatment and typical efficiency[a]

Pump system component	Range, %	Typical value, %
Pump	30–75	60
Flowrate control	20–98	60
Motor	85–95	90
Overall efficiency	5–80	

[a] Adapted from U.S. EPA (2010).

of a motor. With a VFD, the pump can continue to operate along the same pump system curve while shifting the performance curve [see Fig. 17–20(a)]. Throttling and bypass controls artificially shift the system curve while maintaining the power input, thereby reducing the actual flow delivered from the pump [see Fig. 17–20(b-d)]. Stop/start control does not use part of the output that is available (U.S. EPA, 2010).

Aeration Control for the Activated Sludge Process. Energy demand associated with aeration of the activated sludge process could be managed by (1) controlling the aeration rate to match actual oxygen requirements, (2) using energy efficient and appropriately sized blowers, (3) using diffusers with higher oxygen transfer efficiencies for

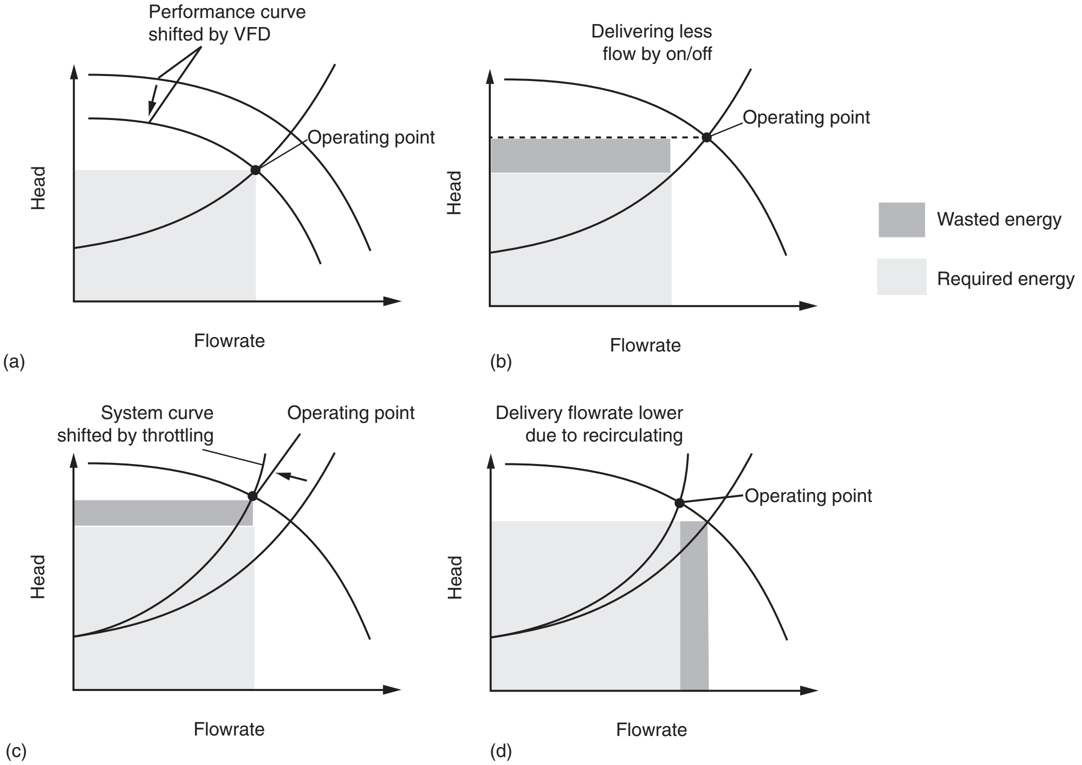
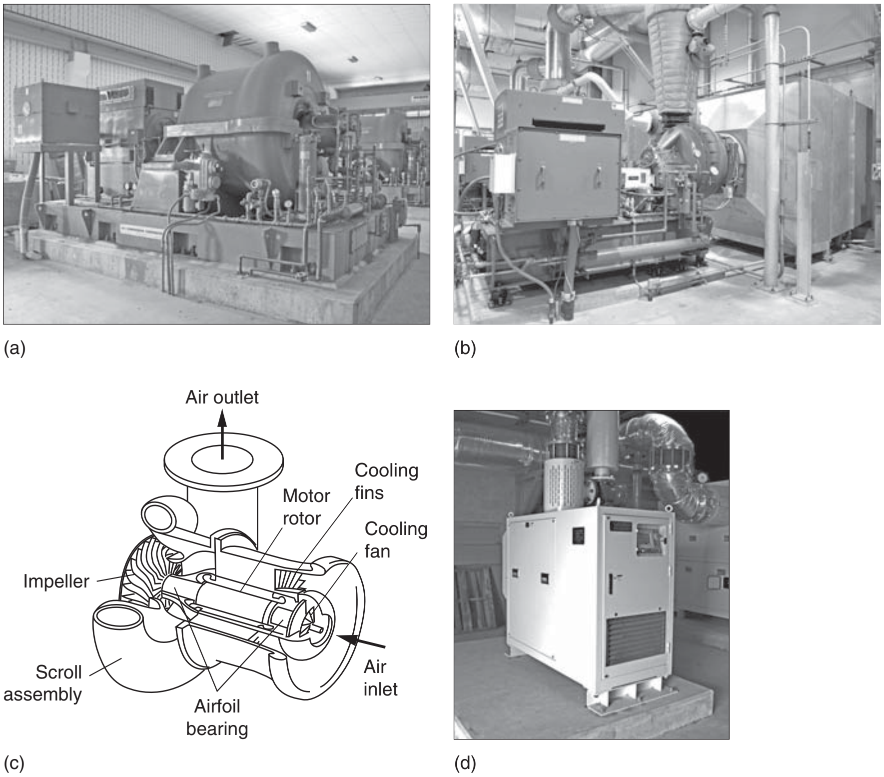


Figure 17–20

Pump efficiency with various control methods: (a) with VFD control, (b) energy loss due to intermittent pumping, (c) energy loss due to throttling, and (d) energy loss due to recirculation. (Adapted from U.S. DOE, 2006.)

Figure 17–21

Examples of aeration blowers for activated sludge process: (a) view of large centrifugal blower driven by 1120 kW (1500 hp) electric motor, and (b) view of single stage centrifugal blower with adjustable inlet guide vanes, (c) schematic of high speed turbo blower (adapted from APG Neuros), and (d) view of high speed turbo blower installation in self contained sound-dampening enclosure.



the given locations and other conditions, and (4) controlling organic loading entering the activated sludge process. Minimizing energy loss from inefficient piping of air supply system can also contribute to the overall energy demand, but the contribution is generally minor compared to other factors (WEF/ASCE 2010).

Excessive DO concentration in the aerobic zone of BNR processes is not only wasteful of energy but can result in the carryover of DO into anoxic or anaerobic zones. Excessive DO can also result in biological oxidation of organic carbon meant for denitrification. A deoxigenation zone is often allocated to avoid the DO intrusion into anoxic and anaerobic zones.

Blowers. The types of blowers used commonly in North America include single stage centrifugal, multi stage centrifugal, rotary lobe positive displacement, and high speed single stage turbo blowers (see Fig. 17–21 and Fig. 5–72 in Sec. 5–11 in Chap. 5). High speed turbo blowers are relatively new technology that utilizes air bearings or magnetic bearings to achieve high speed. Comparison of the blower types and blower efficiencies are summarized in Table 17–15. It should be noted the technologies for blowers are advancing quickly and most recent information should be consulted to determine the applicability of the turbo blowers. Airflow control devices include inlet dampers, inlet guide vanes at the inlet to the blower, and outlet dampers. Inlet guide vanes are used most commonly with centrifugal blowers and are more energy efficient than dampers or throttling. Inlet guide vanes create swirls to the airstream entering the blower, lowering the load on the blower, and are an effective airflow control between 80 and 100 percent of full flow. Another method of airflow control is fan speed adjustment, typically by VFD (DOE, 1989). Similar to the pumps, the use of VFD is an energy-efficient control of the air flowrate. The turn-down ratio, the ratio between the lowest operable flowrate to the maximum flowrate, for the specific blower must be evaluated so that the air flowrate for both

Table 17–15

Types of blowers and typical blower efficiency[a]

Blower type	Range of discharge pressure Bar	Nominal blower efficiency %	Nominal turndown % of rated flow[b]
Positive displacement, VFD	0.55–1.03	45–65	50
Positive displacement hybrid, VFD	0.20–1.5	70–80	25
Centrifugal, multi-stage, throttled	0.55	60–70	60
Centrifugal, multi-stage, inlet guide vanes	0.55	60–70	60
Centrifugal, multi-stage, VFD	0.55	60–70	50
Centrifugal, single stage, internally geared[c]	0.83	75–80	45
Centrifugal, single stage, gearless (high-speed turbo)	1.03	75–80	45–50

[a] Adapted, in part, from U.S. EPA (2010).

[b] The value depends on the blower pressure.

[c] Constant Speed with dual vane system, Outlet diffuser vanes and inlet guide vanes.

low loading and high loading conditions can be provided by the selected blower and the VFD (U.S. EPA, 2010). For example, a high efficiency constant speed blower may be used to cover the base flow without a VFD, and additional blowers, either the same size or a smaller size, could be equipped with a VFD to provide the required range of air flow.

Aeration Diffusers. Oxygen transfer efficiencies of various diffuser types were discussed in Sec. 5–11 in Chap. 5. Most activated sludge facilities designed and constructed in the 1970s and 1980s and earlier employed coarse bubble diffusers or surface aerators. As these facilities approach their design life and design capacity, their replacement with higher

EXAMPLE 7–10 **Energy Saving by DO Control** Using the complete-mix activated sludge process to treat 22,700 m^3/d of primary effluent for COD removal without nitrification as shown in Example 8–3 in Chap. 8, estimate the reduction in power demand by changing the DO setpoint from 3.5 mg/L to 2.0 mg/L.

Use the following design information extracted from Example 8–3 and supplemental information necessary to determine the energy demand.

Parameter	Unit	Value
Required oxygen transfer rate, OTR_f	kg/h	111.3
Alpha factor, α	–	0.50
Fouling factor, F	–	0.90
Beta factor, β	–	0.95
Dissolved oxygen surface saturation concentration at 12°C, C_{st}	mg/L	10.78
Dissolved oxygen surface saturation concentration at standard temperature (20°C), C^*_{s20}	mg/L	9.09

(*continued*)

(*Continued*)

Parameter	Unit	Value
Pressure correction factor, P_b/P_s		0.94
Steady-state DO saturation concentration, $C^*_{\infty 20}$		10.64
Empirical temperature correction factor, θ		1.024
Temperature of mixed liquor	°C	12
Ambient temperature	°C	15
Oxygen in the air	kg/kg air	0.232
Oxygen transfer efficiency, OTE	%	25
Universal gas constant	kJ/kmole	8.314
Blower inlet absolute pressure	kPa	101.3
Blower discharge absolute pressure	kPa	121.5

For simplicity, assume the same oxygen transfer efficiency of 25 percent for both DO concentrations.

Solution

1. Determine the SOTR for DO in aeration basin = 3.5 mg/L using Eq. (5–70).

$$\text{SOTR} = \left[\frac{\text{OTR}_f}{(\alpha)(F)}\right]\left[\frac{\text{C}^*_{\infty 20}}{(\beta)(\text{C}_{\text{st}}/\text{C}^*_{\text{s}20})(\text{P}_\text{b}/\text{P}_\text{s})(\text{C}^*_{\infty 20}) - \text{C}}\right][(1.024)^{20-t}]$$

Using the data given above, SOTR is calculated as

$$\text{SOTR} = \left[\frac{(111.3 \text{ kg/h})}{(0.50)(0.90)}\right]\left\{\frac{10.64}{\left[0.95\left(\frac{10.78}{9.09}\right)(0.94)(10.64) - 3.5\right]}\right\}(1.024^{20-12}) = 409.6 \text{ kg/h}$$

2. Determine air flowrate in terms of kg/min.

$$\text{Air flowrate, kg/min} = \frac{(\text{SOTR kg/h})}{[(E)(60 \text{ min/h})(0.232 \text{ kg O}_2\text{/kg air})]}$$

$$= \frac{409.6}{(0.25)(60)(0.232)} = 117.7 \text{ kg/min}$$

3. Calculate an estimated power requirement using Eq. (5–77a) in Chap. 5.

$$P_w = \frac{wRT_1}{28.97\, n\, e}\left[\left(\frac{p_2}{p_1}\right)^n - 1\right]$$

Using the data give above, P_w is calculated as

$$P_w, \text{kW} = \frac{(117.7/60)(8.314)(273.15 + 15)}{28.97(0.283)(0.85)}\left[\left(\frac{121.5}{101.3}\right)^{0.283} - 1\right]$$

$$= 35.6 \text{ kW}$$

4. Repeat Steps 1 through 3 for DO = 2.0mg/L.

$$\text{SOTR} = \left[\frac{(111.3\ \text{kg/h})}{(0.50)(0.90)}\right]\left\{\frac{10.64}{\left[0.95\left(\dfrac{10.78}{9.09}\right)(0.94)(10.64) - 2.0\right]}\right\}(1.024^{20-12}) = 343.3\ \text{kg/h}$$

$$\text{Air flowrate, kg/min} = \frac{(\text{SOTR kg/h})}{[(\text{OTE})(60\ \text{min/h})(0.232\ \text{kg O}_2/\text{kg air})]}$$

$$= \frac{343.3}{(0.25)(60)(0.232)} = 98.6\ \text{kg/min}$$

$$P_w, \text{kW} = \frac{(98.6/60)8.314\cdot(273.15 + 15)}{28.97\cdot 0.283\cdot 0.85}\left[\left(\frac{121.5}{101.3}\right)^{0.283} - 1\right]$$

$$= 29.8\ \text{kW}$$

5. Compare the energy demand at DO setpoint at 3.5 and 2.0 mg/L.
From Step 3 and Step 4, the energy requirement for aeration is reduced from 35.6 to 29.8 kW by reducing the DO setpoint from 3.5 to 2.0 mg/L. The reduction in energy requirement is 16.2 percent.

Comment The energy demand estimate with Eq. (5–77) is for a single-stage centrifugal blower and assuming the air flowrate is controlled by VFD. The estimated energy saving cannot be recognized if the air flow is controlled with throttling and other methods which do not reduce energy consumption with reduced air flowrates.

oxygen transfer efficiency aerators will, in most cases, save energy. When such a retrofit/upgrade is considered, it is also important to review and make necessary improvements in the air piping, air flow meters, and the DO monitoring and control system.

Process Modification for Increased Energy Production

Increased energy production can be achieved by improving the removal of organic material from wastewater prior to biological secondary treatment for anaerobic digestion, and by improving the volatile solids destruction in the anaerobic digesters. In recent years, the use of waste products from other sources has been studied for the enhanced energy production at wastewater treatment facilities. Enhanced energy production with waste products from other sources such as food waste, fats, oil and grease, and waste from industrialized livestock operations can be implemented at a wastewater treatment facility. Depending on the scale of the energy production, it may be possible to produce more energy than the treatment facility consumes (net energy positive). The use of renewable energy generated onsite, such as solar and wind power, has been implemented on a smaller scale, however, the contribution of these energy sources is relatively minor in comparison to the total energy demand at a wastewater treatment facility.

Additional Removal of Organic Matter from Wastewater. One effective way to remove additional organic matter from wastewater is through primary effluent filtration. In recent studies, it has been possible to remove up to 70 percent of the TSS remaining in primary effluent. Primary effluent filtration is considered in Sec. 5–9 in Chap. 5. Another technology that has been tested and found to remove a significant fraction of COD and TSS from raw wastewater is charged bubble flotation, also considered in Sec. 5–9 in Chap. 5, in which air bubbles are coated with a polymer. In the coming years it is anticipated that a variety of new technologies will be developed to capture the organic matter in wastewater before aerobic oxidation. Removal of organic matter at the primary treatment can also be improved by chemically enhanced primary treatment (CEPT) but care must be taken in evaluating the effectiveness of CEPT for the enhanced energy recovery because the increase in inorganic precipitates in the sludge could potentially lower the efficiency of biogas production with anaerobic digestion. Other design considerations with enhanced removal of organic matter include the impacts of iron on UV disinfection (when iron salts are used for precipitation), and carbon requirements for denitrification in BNR processes.

Process Modifications to Enhance Digester Gas Production. It is often recognized that older anaerobic digesters are not operated and maintained to maximize the production of biogas. Two common issues are insufficient digester heating and insufficient digester mixing. Another common loss of energy is due to leakage in biogas piping and storage facilities. Improvements in the operating conditions of the existing anaerobic digesters and biogas capturing could improve the amount of energy available for beneficial use. In addition to these operational and maintenance issues, methods to enhance solids loading and digester performance are discussed in detail in Sec. 13–9 in Chap. 13.

Use of Waste Products from Other Sources. Anaerobic digesters are often designed with multiple tanks but not all tanks are built in the beginning, and additional space is allocated for the later stage of the design horizon. Even with the phasing of the digester expansion, anaerobic digesters often operate at below the design capacity. The excess capacity available in most anaerobic digesters can be utilized to increase the digester gas production. The use of food waste, fat, and grease for enhanced production of biogas has been studied extensively in recent years, based on the fact that enhanced biogas production could be implemented with minimal modifications to the existing anaerobic digestion processes. The use of food waste, fat, and grease for the enhanced production of biogas is considered in Chaps. 10 and 13.

Use of Renewable Energy Sources. Considerations for the use of renewable energy sources such as wind and solar energy have become a common practice in many buildings but implementation at wastewater treatment facilities has been relatively minor, typically due to space limitations. It is considered more commonly when a building or the facility is pursuing a sustainability certificate, such as a LEED (leaders in energy and environmental design) program, but often these energy sources could fill only a small fraction of the total energy demand at a treatment plant. Common usage of renewable energy at a wastewater treatment facility includes the use of solar panels for lighting, and the use of solar energy for the production of hot water.

Peak Flowrate Management (Peak Energy Usage)

The pricing structure for the electricity for large power users such as wastewater treatment facilities typically includes a demand charge based on the peak equipment power demand, and energy charge based on the amount of energy consumed. The pricing for the energy, in most cases, is different during peak demand hours. In some regions, a spot market is adopted for the electricity pricing. When considering energy management in terms of energy cost savings, one of the effective measures is to shift the use of electricity during the hours with lower energy charge. The shift could be made, for example, by operating sludge dewatering only during off-peak hours.

Flow equalization can be an effective measure to divert part of the wastewater entering the facility during the peak power demand (highest power cost) hours and treat it during off-peak hours. Peak flowrate management not only reduces the use of power during peak demand hours, but also allows for more uniform flow and loading to the treatment processes which will also increase the energy efficiency for most electrically driven equipment. Flow equalization of return flows for treatment in off hours will also reduce peak demand. Flow equalization of return flows is considered in Chap. 15.

Opportunities exist for managing energy use in wastewater treatment by employing the concept of demand-side management. The electric power industry has long recognized the importance of integrating traditional supply-side planning with demand-side management to reduce peak demand. The goal of demand-side management is to change the electrical load characteristics (the amount of energy used at different times of the day) by improving energy efficiency and managing equipment operation. In demand-side management, it is also recognized that continued load growth will occur as the systems expand to meet new domestic and industrial wastewater collection and treatment requirements.

Selection of Energy Sources

Wastewater treatment agencies typically sign contracts with an electrical power supplier and suppliers of fuels (e.g., natural gas, fuel oils and other fuels) to set the pricing. The pricing for electrical power can be a based on a fixed plus usage fee, which can be a fixed rate or dynamic rate. When the electrical power charge is based on the dynamic pricing, a significant cost saving can be expected by implementing peak flowrate management with flow equalization, by which aeration power demand could be shifted to off-peak hours, or during the time electrical power cost is lower.

The mechanism of fuel cost contracts may affect the selection and design of mechanical equipment as different types of contracts become available for the wastewater treatment facilities. For example, a 20-y contract for natural gas at a fixed price can lead to the selection of direct drive motors fueled by natural gas over electrical motors. The fixed price contract can be compared to the purchase of electrical power which can be unpredictable.

17–10 FUTURE OPPORTUNITIES FOR ALTERNATIVE WASTEWATER TREATMENT PROCESSES

The principal sources of energy in wastewater have been identified and discussed previously in Sec. 17–2. The means used currently to recover and utilize energy from the available sources has been discussed in Secs. 17–5 through 17–7. The purpose of this section is to highlight briefly some innovations that are being developed that will make it possible to (1) recover additional energy from the various sources, (2) reduce energy usage through the implementation of different biological treatment technologies, and (3) reduce energy usage though the implementation of alternative treatment process options. Energy reduction through the use of improved process equipment is discussed in the previous section dealing with energy management.

Enhanced Energy Recovery of Particulate Organic Matter

Currently, conventional primary clarification achieves about 50 percent removal of the applied TSS and 30 to 40 percent of incoming COD. The TSS and COD not removed by sedimentation must be treated in the downstream biological treatment process. If all of the particulate COD could be removed before biological treatment, additional energy could be recovered from the material removed and the amount of energy need for carbonaceous oxidation could be reduced.

The TSS and associated COD can be removed by depth, surface, or membrane filtration. Primary effluent filtration (PEF) has been studied since the late 1970s (Matsumoto et al.,1980, 1982). Using PEF in combination with primary sedimentation, removal rates of TSS and COD around 90 and 60 percent, respectively, can be achieved. The PEF process was not adopted widely because the cost of energy at the time the process was first studied was less than $0.03/kWh and the recovery of energy was not an issue. With the current interest in the recovery of energy, the desire to reduce energy usage, and the development of a number of new filtration technologies, there is a resurgence of interest in PEF. A major advantage of PEF is that a separate waste stream is not created, because the solids removed by filtration, either continuously or intermittently, can be mixed with the primary sludge. Primary sludge from PEF or CEPT processes is converted readily to biogas using anaerobic digestion. Alternatively, the solids can be diverted to a fermenter for the production of volatile fatty acids for use in the enhanced removal of phosphorus (see Sec. 8–8 in Chap. 8).

Reduced Energy Usage in Biological Treatment

As discussed in Chap. 7 and more extensively in Chap. 15, new biological processes are being developed that offer potential savings in energy and chemical consumption, especially with respect to the removal of nitrogen. For example, the use of nitritation/denitritation and partial nitritation and deammonification processes can be used to reduce both oxygen and carbon demand. Nitrogen removal through nitritation/denitritation would require 25 percent less oxygen and 40 percent less external carbon source than conventional nitrification/denitrification processes. With partial nitritation and deammonification, the requirement for oxygen is 60 percent less and demand for an external carbon source is nearly 90 percent less than the conventional nitrification and denitrification (see Chap. 15).

When partial nitritation and deammonification is used for sidestream treatment and the main treatment process is also designed for nutrient removal, the use of external carbon in the sidestream treatment may be eliminated and the remaining nitrate treated in the main stream process, which will simplify the chemical dosing system for the sidestream treatment. While the application of this process is predominantly for the treatment of recycle flow with high ammonium concentration, studies are ongoing to apply the process for the treatment of main stream wastewater at ambient temperatures (Al-Omari et al., 2012).

Reduced Energy Usage through the Use of Alternative Treatment Processes

Future treatment process flow diagrams will incorporate alternative biological processes as well treatment without biological processes to achieve more effective reduction and utilization of the energy in wastewater. Examples of alternative approaches are illustrated on Fig. 17–22. With advances in treatment technologies, a number of alternative approaches are expected to emerge.

Anaerobic Treatment at Ambient Temperature. Anaerobic processes which were discussed in detail in Chap. 10 are in general less energy intensive than aerobic processes as they do not require aeration for the bulk removal of bCOD. Anaerobic processes also generate biogas containing methane which could be recovered and used for heat and power generation. In the future, it is anticipated that treatment process flow diagrams will be developed for the anaerobic treatment of wastewater at ambient temperature (McCarty, 2011).

Use of Trickling Filters for Treatment of Filtered Wastewater. Another example of an alternative treatment process flow diagram involves the use of a cloth screen for primary treatment followed by a filtration step as discussed above. The filtered effluent could be applied directly to a tricking filter in a single stage, or two stage process where nutrient removal is an issue. A sedimentation tank would not be needed following the trickling filter as the small amount of residual solids could be removed, if needed, with a high-rate filtration process (Koltz, 1985).

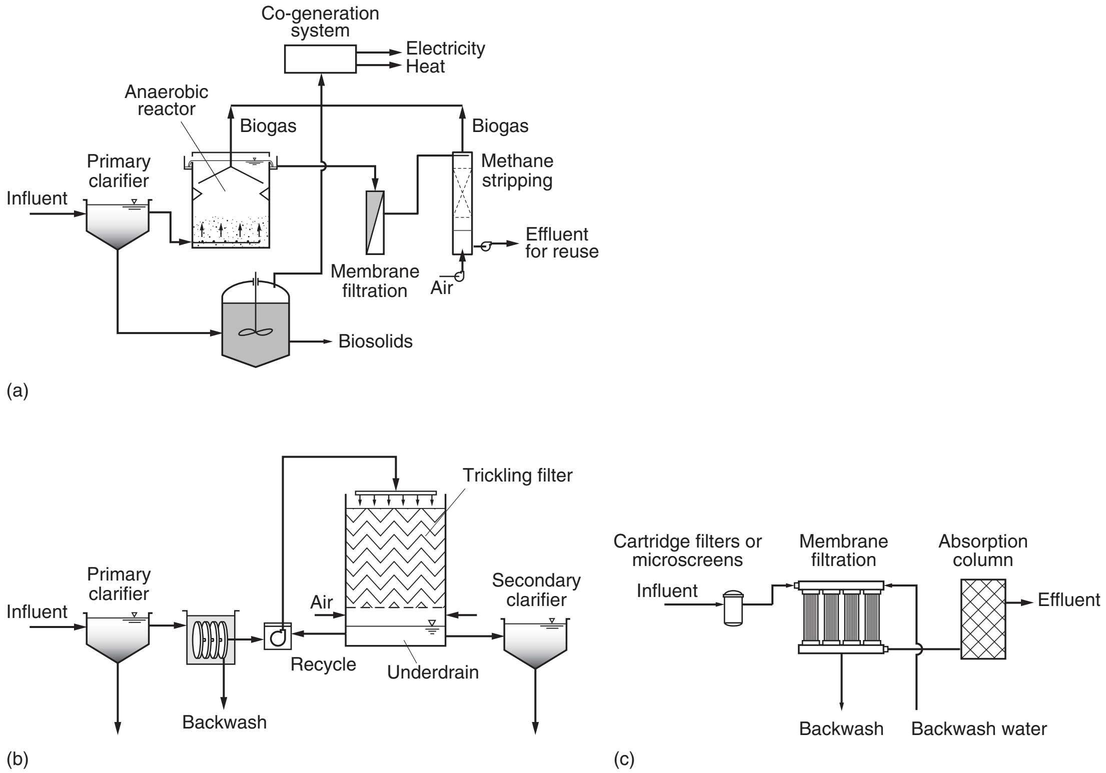


Figure 17–22

Process flow diagrams of alternative wastewater treatment processes: (a) anaerobic treatment at ambient temperature (adapted from McCarty, 2011), (b) trickling filter for treatment of filtered wastewater, and (c) membrane absorption process.

Membrane-Absorption Process. It is also anticipated that processes will be developed to treat wastewater that will not employ biological treatment. Filtered wastewater, as discussed above, could be processed further with a variety of different membranes. Any residual organic matter passing the membrane could be removed by adsorption. In turn, the adsorbed organic matter could be digested or processed thermally (Adams et al., 2011).

Prospects for the Future

The key to thinking about alternative treatment processes is that the technologies available currently, or under development, will make it possible to alter the characteristics of wastewater to optimize any given treatment process while maximizing the recovery and utilization the energy in wastewater.

PROBLEMS AND DISCUSSION TOPICS

17–1 Review three current (since 2005) peer-reviewed articles on co-digestion of wastewater sludge and food waste. Summarize the key findings with respect to increased gas production. Discuss the obstacles/challenges with implementation of co-digestion based on your review.

17–2 Given the following summary information for a wastewater treatment plant, determine the benchmark score for this treatment plant using the AWWARF methodology described in Sec. 17–4. Calculate the benchmark score for 30, 60, and 100 percent use (value to be selected by instructor) of biogas to offset natural gas use. Assume LHV of 35 MJ/m^3 for natural gas, 22 MJ/m^3 for digester gas.

Parameter	Unit	Value
Flowrate (annual average/design capacity)	10^3 m^3/d	24/30
Influent BOD	mg/L	120
Influent ammonium	mg N/L	24
Influent total nitrogen	mg N/L	36
Effluent BOD	mg/L	6.2
Effluent ammonium	mg N/L	2.1
Effluent total nitrogen	mg N/L	5.5
Energy use		
Electricity	kWh/y	3,600,000
Biogas production	m^3/y	605,000
Biogas used	m^3/y	0
Natural gas	m^3/y	372,700

17–3 Using the data presented in Example 17–5 for a wastewater treatment facility, calculate the benchmarking score when all biogas is utilized to generate electricity and heat is recovered to offset natural gas use. Assume typical energy efficiencies for CHP as reported in Table 17–8.

17–4 A wastewater treatment facility receives raw wastewater at an average flowrate of 7500 m^3/d, with a velocity of 1.5 m/s, and a pressure head of 2.5 bar. Identify the type of hydroturbine that could be used to generate electricity using the incoming raw wastewater, and estimate the potential power generation. Use Fig. 3–14 to estimate the peaking factor and Fig. 17–19 to estimate the average efficiency at the average flowrate when the turbine is sized for the peak flowrate. Discuss potential operational issues in using raw wastewater for hydraulic turbine.

17–5 Calculate the heat balance for sludge combustion assuming the following solids composition (to be selected by instructor). Use the procedure presented in Example 17–6.

	Sample constituent composition, % by wt		
Element	1	2	3
Combustible			
Carbon	13.3	18.9	15.0
Hydrogen	0.9	1.3	1.2
Oxygen	4.9	6.9	5.3
Nitrogen	0.3	0.4	0.3
Sulfur	0.4	0.5	0.4
Inerts	8.3	11.9	7.8
Water	72	60	70
Total	100	100	100

17–6 Determine the coefficient of performance for the heat pump as well as the entire system and the wastewater flowrate necessary to provide the heating requirement based on the information summarized in the table below (to be selected by instructor).

Parameter	Unit	Value
Heating requirement (value to be selected by instructor)	kW	(1) 200 (2) 240 (3) 280
Heat pump electrical input	kW per kW output	0.24
Wastewater temperature	°C	15
Temperature drop through heat exchanger	°C	5
Total power requirements for pumps (value to be selected by instructor)	kW	(1) 40 (2) 45 (3) 50

17–7 For Problem 17–6 above, discuss benefits and potential issues to extract heat from raw wastewater in the collection system, instead of the wastewater effluent, for district heat supply in a small commercial area.

17–8 For Example 17–10 and Example 8–3 in Chap. 8, the treatment process operated at the design capacity is upgraded with a primary effluent filtration device between primary clarifiers and activated sludge process, and BOD loading to the activated sludge process was reduced by 20, 30, or 40 percent (value to be selected by instructor). Calculate the air requirement and aeration power demand for the same COD removal without nitrification process. Assume the DO set point is 2.0 mg/L.

17–9 The wastewater treatment plant in Example 8–3, designed for BOD removal, is currently operating at 60 percent of its capacity. The municipality is experiencing lower population growth in the service area as compared to the projected growth when the treatment plant was designed and built. As part of an integrated resource recovery scheme being studied by the municipality, you are asked to determine if implementing the use of kitchen food waste grinders in residential homes will result in saving energy. It is projected that installation of food waste grinders will result in 600, 1000, or 1200 kg/d (value to be selected by instructor) increase in biodegradable COD loading, of which 75 percent will be captured as primary sludge, and 20 percent will be treated in the activated sludge process. Primary sludge is transferred to anaerobic digestion to produce biogas.

a. Determine the aeration power demand at 60 percent loading.

b. Estimate the increased aeration power demand if the use of food waste grinders was implemented.

c. Determine the increase in digester gas production if food waste grinders were implemented (assume anaerobic digester has sufficient capacity)

d. Assuming 45 percent engine efficiency and 93 percent generator efficiency, determine whether the implementation of food waste grinders will result in overall energy saving.

For simplicity, ignore the increased biomass and biogas production from WAS and the corresponding energy savings from heat recovery in the calculations.

17–10 A wastewater pump station is used to transfer 20,000 m^3/d screened raw wastewater from a wet well to an aerated grit chamber. The pump station contains three pumps. Two pumps are used continuously and one pump serves as a standby, with an on/off control. Discuss potential changes to the pump system that can be made for energy conservation and explain how the savings are achieved.

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18

Wastewater Management: Future Challenges and Opportunities

WORKING TERMINOLOGY

Term	**Definition**
Climate change	Significant change in measures of climate such as temperature, precipitation or wind, lasting for an extended period of time (decades or longer).
Decentralized (satellite) treatment system	System used for the treatment of wastewater located close to the point of reuse. Satellite treatment plants generally do not have solids processing facilities; solids are returned to the collection system for processing in a central treatment plant located downstream. Three types of satellite systems are identified: (1) interception type, (2) extraction type, and (3) upstream type.
Direct potable reuse	The introduction of purified water either directly into the potable water supply distribution system downstream of a water treatment plant, or into the raw water supply immediately upstream of a water treatment plant.
Histogram	A graphical representation of the frequency with which an event occurs over a range of different conditions.
Indirect potable reuse	The planned incorporation of purified water into a raw water supply, such as in potable water storage reservoirs or a groundwater aquifer, resulting in mixing, dilution, and assimilation, thus providing an environmental buffer.
Linear correlation	The linear relationship between two variables.
Low impact design (LID)	A design approach that protects surface and ground water quality, maintains the integrity of aquatic living resources and ecosystems, and preserves the physical integrity of receiving streams with designs that are in harmony with nature.
Natural disasters	Extreme sudden natural events such as floods, tornadoes, hurricanes, heat waves, or droughts that injure people and damage property.
NPDES	The National Pollution Elimination Discharge System (NPDES) was established based on uniform technological minimums with which each point source discharger has to comply.
Pareto analysis	A statistical decision making technique to select a limited number of tasks or variables that will result in a significant overall impact.
Pilot plant studies	Studies conducted at test beds at a scale larger than bench-scale, to establish the suitability of a process in the treatment of a specific wastewater under specific environmental conditions and to obtain data that can be used for full-scale design and operation.
Population demographics	The study of populations based on statistics such as economics; migration patterns; and birth, deaths, and disease.
Potable reuse, direct	See Direct potable reuse.
Potable reuse, indirect	See Indirect potable reuse.
Privatization	Private sector ownership and operation of facilities and services used by government entities in performing their public function.
Satellite treatment system	See decentralized treatment.
Sustainability	The principle of optimizing the benefits of a present system without diminishing the capacity for similar benefits in the future.
Triple bottom line	Project analysis in which profit (economics), people (social) and planet (environment) are considered equally.
Uncontrollable events	Events in nature beyond the control of humans, such as tornadoes and hurricanes.
Unintended consequences	Consequences (outcomes) that are not anticipated or intended by a particular action.
Variability, inherent	Based on the laws of chance, all physical, chemical, and biological treatment processes exhibit some measure of variability with respect to the performance that can be achieved. Variability is inherent in biological treatment processes.
Wet-weather flow	The runoff that results when it rains or snows.

The concepts and elements involved in the general design of wastewater treatment plants have been presented and discussed in the previous chapters. The purpose of this chapter is to consider the various concepts and design elements in light of the challenges and opportunities for wastewater management in the future. The topics considered are (1) a general discussion of some important future challenges and sustainability issues, (2) the impact of global and uncontrollable events, (3) upgrading existing treatment plant performance through process optimization, (4) upgrading existing treatment plant performance through process modification, and (5) the management of wet-weather flows. Important future challenges and sustainability issues and the impact of global and uncontrolled events are discussed first because improvements to existing plants as well as the design of new plants must be undertaken in light of these challenges and issues. Wherever possible, reference will be made to the pertinent section and/or chapter where additional information can be found on the individual topics discussed in this chapter.

18–1 FUTURE CHALLENGES AND OPPORTUNITIES

In the twentieth century the primary focus of wastewater treatment was on the removal and treatment of settleable and floatable solids, organic matter expressed as biochemical oxygen demand (BOD), total suspended solids (TSS), and pathogenic microorganisms (see Chap. 1). Late in the twentieth century, nutrient removal and odors also became issues, and controlled, non-potable use of reclaimed water became a common practice in many parts of the world (Asano et al., 2007). In the twenty-first century, as a result of numerous environmental issues and events, a paradigm shift has occurred in how wastewater is viewed. In the twenty-first century, wastewater is no longer viewed as a waste requiring disposal, but as a "renewable recoverable source of energy, resources, and potable water" (Tchobanoglous, 2010; Tchobanoglous et al., 2011). In light of this view of wastewater, it is appropriate to consider briefly some challenges as well as the opportunities that will become increasingly important in the future, including: (1) asset management, (2) the need to design wastewater treatment plants (WWTPs) for the recovery of energy and resources, (3) the need to produce effluent suitable for the production of potable water, (4) the implementation of decentralized (satellite) treatment systems, (5) integrated wastewater management, (6) the use of low impact development, and (7) the use of the triple bottom line for project evaluation. It should be noted that the challenges discussed below are beyond the need to meet more stringent discharge standards for trace organics and residuals processing as discussed in Chap. 4. The impact of global and uncontrollable events on the future of wastewater management is considered in the following section.

Asset Management

With aging infrastructure and reduced funding for repair and replacement, wastewater agencies are exploring a number of different techniques that can be used to prioritize future capital expenditures that will yield the greatest value added for the consumer. Asset management, in its many forms, is one of the techniques now being investigated and applied. The U.S. EPA defines asset management (AM) as "Managing infrastructure assets to minimize the total cost of owning and operating them, while delivering the service levels customers desire" (U.S. EPA, 2012a). The U.S. EPA further reports (U.S. EPA, 2012b) the implementation of AM based decisions have the potential to save 20 to 30 percent of future lifecycle costs within U.S. wastewater utilities and defined strategies and procedures have reportedly shown operational cost savings and more efficient working practices.

Objectives of Asset Management. The principal objective of AM is to develop sufficient information in an organized fashion to support strategic decision making. All agencies conduct some form of AM, but since the late 1990's there has been an increased emphasis on the adoption of more focused approaches that encourage the understanding and reporting of all aspects of the lifecycle of an asset. From the initial conceptual need through to complete implementation, owners can draw from a wide range of proven techniques and available technologies to assist in the design, operation, and maintenance of new and replacement assets. The application of these techniques and technologies rely on the collection and analysis of data that enable the owner to analyze a range of performance metrics to support strategic decision making.

Practice of Asset Management. To date, the most widespread development and implementation of AM has been in Australia, New Zealand, and the UK where the performance of all aspects of the water and wastewater industries are more closely regulated. From the experience gained in these countries, it is clear that although it takes a number of years to see true financial returns, there are very real benefits by adopting more advanced AM techniques. While regulatory requirements in the United States do not require the implementation of AM, many agencies including WEF, AWWA, and the Association of Metropolitan Water Agencies (AMWA) have taken a lead role in promoting AM. The U.S. EPA continues to blend education of the principles and tools, with a formal need for implementation as a component of compliance regulations for individual agencies and owners.

Asset Management Methodologies. There are many AM techniques and technologies and it is beyond the scope of this text to discuss each of them adequately. Numerous publications are available to provide guidance but the fundamental core is in understanding the required performance from an asset and being able to operate and maintain the asset cost-effectively. In general, AM techniques include (1) developing an asset inventory, (2) assessment of the condition of asset, (3) determining the level of service to be provided, (4) identifying the critical asset to sustain the performance, (5) determining the cost for the entire life cycle of the asset, and (6) determining the best long-term strategy. Agencies that have a mature understanding of AM will likely have a better understanding of the condition and remaining life of their infrastructure and of the best maintenance methodologies. This level of knowledge can prove to be beneficial as agencies, with such an approach, are considered to be well managed businesses. Information on AM is also needed to justify funding either to customers, financial institutions, or within their own organization.

Ramifications of Asset Management. While much of the current focus of AM is related to existing infrastructure, it is clear that the lifecycle of every asset is significantly influenced by the decisions that are made at the concept and design stages. The designer has the opportunity to determine the optimum performance of the asset while considering criteria relevant to operations and maintenance. The correct combination of these interrelated factors will enable the end user to derive the most benefit with assets that provide the anticipated levels of service with reduced operations, maintenance, and energy costs.

While the principles of AM are considered by many to be the most effective way to manage infrastructure, there are many organizations in the United States that are still in the early stages of developing suitable approaches. Organizations that have adopted more advanced AM methodologies are however seeing benefits both in terms of performance levels and financial management. It is clear, therefore, that a more defined implementation of AM should be considered for all aspects of infrastructure management from the development and justification of need, to the ultimate disposal.

Design for Energy and Resource Recovery

The chemical and heat energy content of wastewater has been delineated and discussed previously in Chaps. 2, 14, and 17. As noted in the earlier discussions, wastewater treatment plants could potentially become net exporters of energy, and especially so if external sources of energy contained in food waste and fats, oils, and grease are included. The challenge in the future is how to extract the energy in wastewater most effectively. For example, food waste could be ground up in kitchen food waste grinders and transported to the wastewater treatment facilities in the collection system, or it could be intercepted at various upstream locations and extracted from the wastewater using a micro- or clothscreen such as described in Sec. 5–9 in Chap. 5. The solids removed from the wastewater could be placed directly in an anaerobic digester. Alternatively, conventional aerobic treatment processes could be replaced with ambient-temperature low hydraulic retention time anaerobic treatment processes (McCarty, 2011). Heat recovered from wastewater could be used for drying screenings as well as in other applications, especially in the processing of biosolids. The key concept here is to think about how the characteristics of wastewater could be altered to enhance the recovery of energy from wastewater.

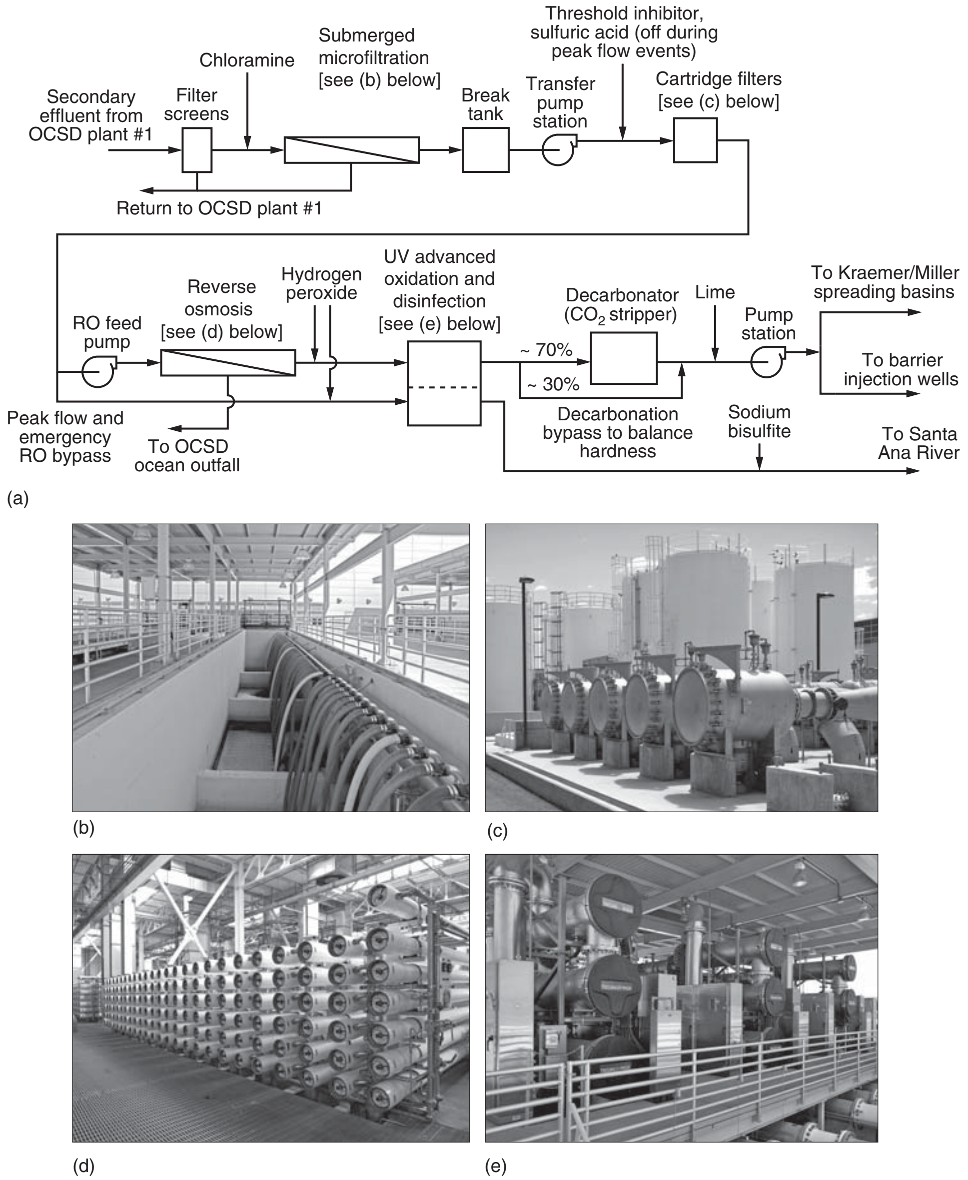
In the future, the recovery of resources from wastewater will occur simultaneously with the recovery of energy. To date, the removal of nitrogen and phosphorus has received the greatest attention as nitrogen and phosphorus discharge standards have become more stringent. The option of recovering, rather than simply removing, these constituents has became economically feasible, especially from return flows. Biological phosphorus removal was considered in Sec. 8–8 in Chap. 8. The recovery of nitrogen and phosphorus in the form of struvite is considered in Sec. 6–5 in Chap. 6 and Sec. 15–4 in Chap. 15. The recovery of nitrogen as ammonium sulfate is considered in Sec. 15–5 in Chap. 15. The recovery of resources from fly ash following combustion is considered in Sec. 14–4 in Chap. 14. The recovery of nutrients including nitrogen, phosphorus, and potassium from urine (see Table 3–15 in Chap. 3) is another resource recovery opportunity that has received considerable attention, especially in Europe and Australia. What role urine separation will play in the United States remains to be seen. Clearly, finding the optimum cost and energy-effective approach for the recovery of resources, coupled with the recovery of energy and potable water from wastewater will be a major challenge in the future.

Design of Wastewater Treatment Plants for Potable Reuse

As a result of population growth, urbanization, and climate change, public water supplies are becoming stressed, and the chances of tapping new water supplies for metropolitan areas are getting more difficult, if not impossible. As a consequence, existing and new water supplies must go further. One way to achieve this objective is by increased water reuse, particularly in supplementing municipal water supplies by means of indirect or direct potable reuse (IPR or DPR) (Leverenz et al., 2011; Tchobanoglous et al., 2011). As a result of the development and demonstration of full-scale advanced treatment processes (see Fig. 18–1), the use of purified water that has been recovered from municipal wastewater directly for potable purposes is now considered to be a viable alternative (NRC, 2012). It is also recognized that there is a continuum of possibilities. The challenge for the future is how to design or upgrade treatment plants so that the effluent produced will be

Figure 18–1

Advanced water treatment facility at the Orange County Water District, Fountain Valley, CA: (a) schematic flow diagram for 2.65×10^5 m^3/d (70 Mgal/d) advanced water treatment facility, (b) microfilters, (c) cartridge filters, (d) reverse osmosis module, and (e) advanced UV oxidation reactors.



suitable for the production of purified water. Technical issues involved in the implementation of potable reuse are identified in Table 18–1.

When considering wastewater treatment plants for potable reuse, a key question is, what should the treatment plant of the future look like? In examining existing treatment process flow diagrams for IPR and DPR, it can be concluded that the production of purified water for DPR was an afterthought. Basically, additional unit processes were tacked on to the end of the existing secondary treatment process flow diagrams to remove specific compounds. However, at some point in the future there will need to be a complete rethinking of urban infrastructure to obtain the highest levels of performance and reliability. For water and wastewater systems, the advanced infrastructure model will likely include decentralization (also known as distributed), remote management, resource recovery, source separated waste streams, and application of specific optimization of water quality.

Table 18–1

Technical issues in the implementation of potable reuse

Consideration	Comments / questions
Source control	• Identification of constituents that may be difficult to remove (depends on technologies used) • Development of baseline sources and concentrations of selected constituents • Define the improvements that need to be made to existing source control programs where DPR is to be implemented
Influent monitoring	• Development of influent monitoring systems, including constituents, parameters, and monitoring recommendations • Investigate potential benefits of various influent monitoring schemes that may be used for early detection of constituents • Consideration of how influent monitoring data could be used to adapt treatment operations depending on variable influent characteristics
Flow equalization	• Determination of the optimum location and type (inline or offline) in secondary treatment process with respect to enhanced reliability and removal of trace constituents • Determination of optimum size of flow equalization before advanced treatment • Quantify the benefits of flow equalization on the performance and reliability of biological and other pretreatment processes
Wastewater treatment	• Quantify benefits of optimizing conventional (primary, secondary, and tertiary) processes to improve overall reliability of entire system • Quantify the benefits of complete nitrification or nitrification and denitrification on the performance of membrane systems used for DPR applications
Performance monitoring	• Determine monitoring schemes to document reliability of treatment performance for each unit process and validate end-of-process water quality
Analytical/monitoring requirements	• Selection of constituents and parameters that will require monitoring, including analytical methods, detection limits, quality assurance/quality control methods, and frequency • Determination of how monitoring systems should be designed in relation to process design • Development of appropriate monitoring systems for use with alternative buffer designs
Advanced wastewater treatment (water purification)	• Develop baseline data for treatment processes employing reverse osmosis. OCWD can be used as a benchmark • Development of alternative treatment schemes with and without demineralization that can be used for water purification • Quantify benefit of second stage (redundant) reverse osmosis
Engineered storage buffer	• Development of sizing guidelines based principally on existing analytical, detection, and monitoring capabilities to assess technical and economic feasibility of utilizing engineered storage buffer • Characterize the impact of existing monitoring response times on the safety and economic feasibility of implementing an engineered storage buffer
Balancing mineral content	• Development of recommendations for balancing water supply mineral content in consideration of site-specific factors, such as magnesium and calcium • Determination of potential impacts of various water chemistries on infrastructure and public acceptance • Development of specifications for chemicals used for balancing water quality

(*continued*)

Table 18–1 (*Continued*)

Consideration	Comments / questions
Blending	• Development of guidance on what level of blending, if any, is required based on the quality of the purified water and alternative water sources • Investigation of the significance of and rationale for blend ratios in terms of engineered buffer, protection of public health, public acceptance, and regulatory acceptance • Investigation of potential impacts of purified water on drinking water distribution system, e.g., corrosion issues, water quality impacts, etc.
Emergency facilities	• Stand-by power systems in the event of power loss or other emergency • Availability of all replacement parts and components that would be required in the event of a process breakdown • Process redundancy so that treatment trains can be taken offline for maintenance • Facilities for the by-pass or discharge of off-spec water in the event that the water does not meet the established quality requirements
Pilot testing	• Utilization of a review panel for advice and recommendations on the design, operation, monitoring plan for a project's pilot system to ensure that it will be representative of the proposed full-scale system • Development of monitoring protocol for collection of baseline data for "raw" water input to AWT pilot plant; how much testing and for what duration (e.g., 6 mo to 1 y). • Development of pilot study design so that results can be used to assess reliability with proposed source water

From Tchobanoglous et al., (2011).

What is needed is the development of integrated water management systems in which new wastewater treatment plants are planned and designed from the ground up to optimize treatment performance with respect to the production of purified water, along with the recovery of energy and resources.

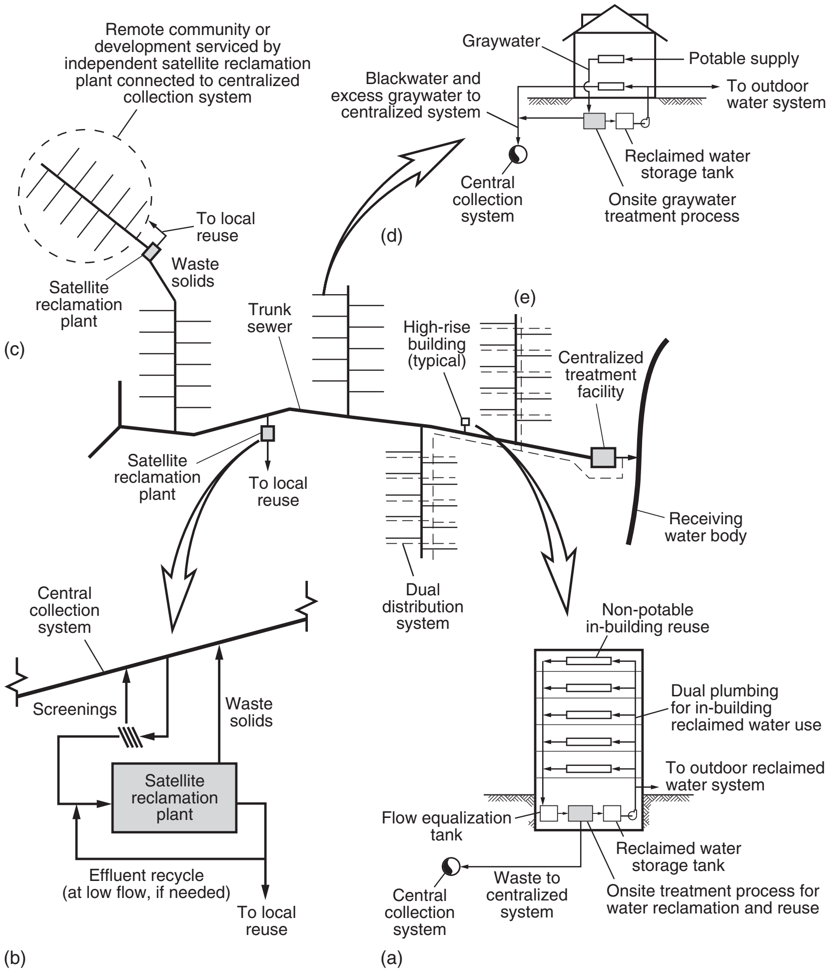
Decentralized (Satellite) Wastewater Treatment

In most collection and treatment systems, wastewater is transported through the collection system to a centralized treatment plant located at the downstream end of the collection system near the point of dispersal (disposal) to the environment. Because centralized wastewater collection systems are generally arranged to route wastewater to these remote locations for treatment, water reuse in urban areas is often inhibited by the lack of a dual distribution system (i.e., purple pipe). The infrastructure costs for storing and transporting reclaimed water to the points of use are often prohibitive, thus making reuse uneconomic. An alternative to the conventional approach of transporting reclaimed water from a central treatment plant is the concept of decentralized (satellite) treatment at upstream locations with localized reuse and/or the recovery of wastewater solids.

Decentralized wastewater systems are used to treat wastewater at or near the point of waste generation and reuse (see Fig. 18–2). Decentralized treatment plants generally do not have solids processing facilities; solids are returned to the collection system for processing in a central treatment plant located downstream. Individual decentralized systems can be used for water reclamation and reuse for applications such as landscape irrigation, toilet flushing, cooling applications, and water features. Use of decentralized systems is predicated on the assumption that the existing collection system can be utilized for the

Figure 18–2

Schematic illustration of four types of satellite water reclamation systems: (a) interception type where wastewater to be reclaimed and recycled is intercepted before discharge to a centralized collection system, (b) extraction type (i.e., sewer mining) in which wastewater is extracted (i.e., pumped) from a centralized collection system for local reuse, (c) upstream type for treatment and reuse for a remote community or development with solids discharged to a centralized collection system, and (d) satellite system for individual home.



transport of solids and reduced flow. Onsite reclamation systems may obviate the need for large-scale dual piping systems, which are generally prohibitively expensive in urbanized areas and reduce the need to expand existing treatment plants to meet future growth projections. Two notable examples of the use of decentralized wastewater treatment systems are in the Los Angeles, CA area as illustrated on Fig. 18–3. Both the City of Los Angeles and the County Sanitation District's of Los Angeles County upstream treatment plants discharge screenings and biological solids to the large treatment plants located near the ocean where the treated effluent is discharged. The implementation of decentralized wastewater treatment will require a new approach to the management and reuse of wastewater (Tchobanoglous and Leverenz, 2013).

Low Impact Development

Low impact development (LID)(also known as low impact design) is a concept that has been applied to the management of wet-weather flow. With respect to the management of wet-weather flows, the goal of LID is to control both rainfall and stormwater runoff, at or near the source. Both engineered and vegetated natural systems are utilized to filter the rainfall runoff and replenish groundwater locally through infiltration (see Fig. 18–4).

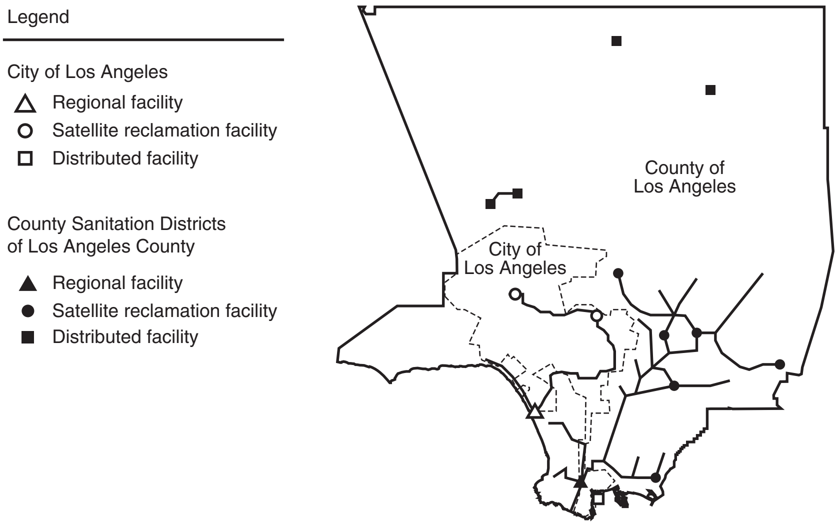


Figure 18–3

Diagram of satellite and centralized treatment systems in the City of Los Angeles and the County Sanitation Districts of Los Angeles County. Waste solids from the upstream plants are processed at the main downstream plants.

It should be noted that interventions such as the use of rain barrels, cisterns, vegetated roof covers, rain gardens, porous pavements, and grassed swales are all part of the LID strategy. By retaining and infiltrating a portion of the runoff, the quantity of flow discharged to stormwater or combined wastewater collection systems can potentially be reduced substantially.

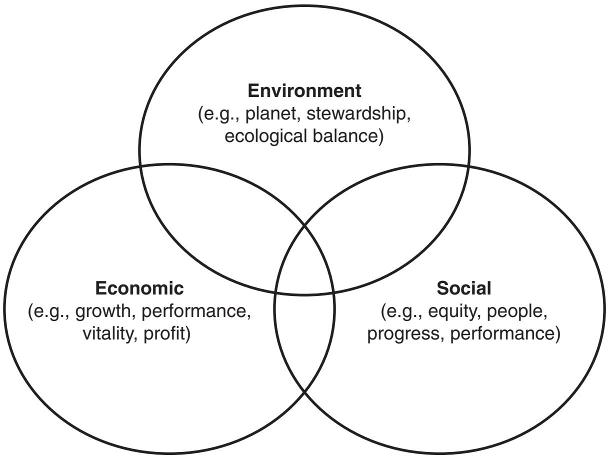
The effectiveness of LID will depend on the nature of the community and the magnitude of the rainfall event. For example, because the City of San Francisco is essentially built out, the impact of LID in terms of flowrate reduction has been relatively minimal

Figure 18–4

Typical examples of landscaping designed to limit the effects of wet weather flows: (a) swale, (b) around treatment unit, (c) open area at treatment plant, and (d) roadway runoff capture.

Figure 18–5

Relationship of the elements involved in the triple bottom line method of analysis.



(e.g., on the order of two to five percent). Yet in other cities, flowrate reductions as high as 20 percent have been reported. With some of the recent record (extreme) rainfall events, the impact of LID has been difficult to assess. The challenge is how to integrate LID most effectively into the management of wastewater. Some notable examples in the form of case studies have been reviewed by the U.S. EPA (2010).

Triple Bottom Line

The *triple bottom line* refers to a method of analysis in which engineers are encouraged to consider social and environmental bottom lines in addition to the financial bottom line in arriving at the most sustainable solution (see Fig. 18–5). While the concept is laudable, in practice it has proven to be difficult to implement uniformly. Assessing the pluses and minuses of the financial aspects of a project has proven to be much easier than assessing the social and environmental pluses and minuses. In practice, to implement social and environmental bottom lines effectively, dollar values must be assigned to the corresponding benefits and drawbacks. If dollar values cannot be assigned, social and environmental concerns must be considered even though they will have little practical impact on project implementation beyond the normal concern for these issues incorporated into any project analysis. The challenge moving forward is how to best incorporate the concepts embodied in the triple bottom line analysis in project planning, design, and implementation.

18–2 IMPACT OF POPULATION DEMOGRAPHICS, CLIMATE CHANGE AND SEA LEVEL RISE, UNCONTROLLABLE EVENTS, AND UNINTENDED CONSEQUENCES

In addition to the identifiable causes of the variability observed in the treatment of wastewater as described in Chap. 3 and 4, a number of other global and local events now routinely impact the design and operation of wastewater treatment plants (WWTPs). Four such event categories are considered in the following discussion: (1) the impact of population demographics, (2) the impact of climate change and sea level rise, (3) the impact of uncontrollable events, and (4) the impact of the law of unintended consequences. Consideration of these topics is especially important in meeting the challenges and sustainability issues identified in the previous section.

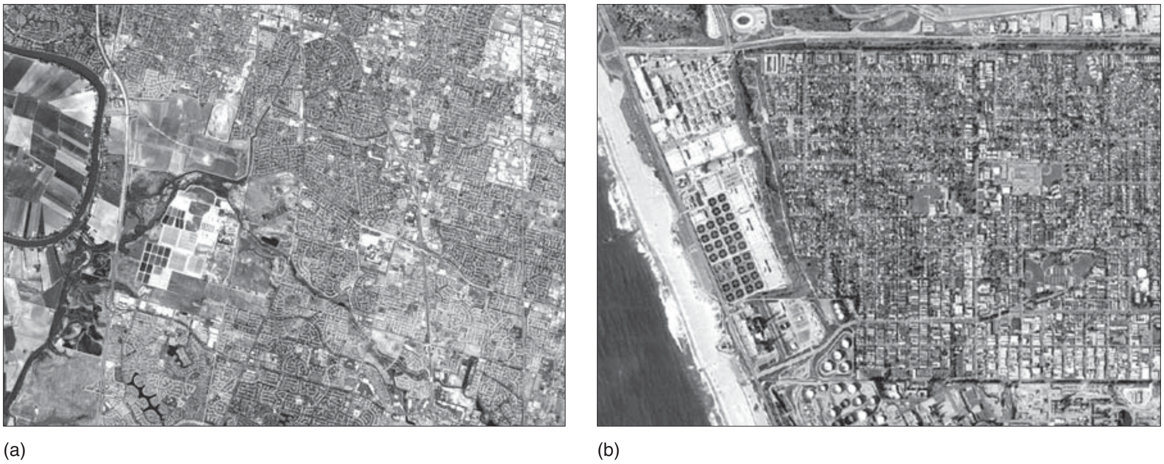

(a)
(b)

Figure 18–6

Impact of urbanization on wastewater treatment plants: (a) inland location, Sacramento, CA (coordinates 38.439 N, 121.480 W, view at altitude 10 km) and (b) coastal location, Los Angeles (coordinates 33.923 N, 118.429 W, view at altitude 2.75 km).

Impact of Population Demographics

Population demographics will continue to impact existing and future of WWTPs in a variety of ways including the impact of urban spread and urbanization along coastal areas.

Impact of Urban Spread and Higher Density Housing. Since the early twentieth century, treatment plants have typically been located at some remote location, distant from the city they serve, where the wastewater would flow by gravity and near some water body that could be used as to receive the treated wastewater. In some cases, pumping was required. What has happened in the 50 to 100 intervening years is that urban spread has essentially encircled most of these early WWTPs (see Fig. 18–6). Urbanization, especially near the boundaries of the WWTPs, has often resulted in a number of diverse complaints about the WWTPs, including odors, noise, excess birds, unsightly vistas, and truck traffic, among others. As a result, a number of corrective measures have had to be implemented to deal with citizen complaints. Perhaps the most common intervention has been to cover open treatment tanks and to install odor management facilities (see Fig. 18–7). Scheduling truck deliveries in off hours has been used in a number of locations. Moving forward, care must be taken to identify and deal with citizens issues and concerns both in building new WWTPs and in upgrading existing WWTPs.

Along with urbanization, higher density housing will be necessary to accommodate anticipated population growth, especially along coastal areas as described below. Although, higher density housing poses a variety of infrastructure challenges, it also offers new opportunities to implement cost-effective decentralized wastewater treatment plants and localized reuse. As discussed previously, treatment plants located in high-density urban areas could be either of the interception or extraction type, as illustrated on Fig. 18–2. In many cases, it may make more sense to integrate extraction type treatment facilities in upscale apartment buildings, as now done in New York City (e.g., the apartment buildings in the Battery Park area of New York City are classic examples). The challenge is to develop a rational cost-effective plan for co-locating high-density housing and wastewater treatment plants.

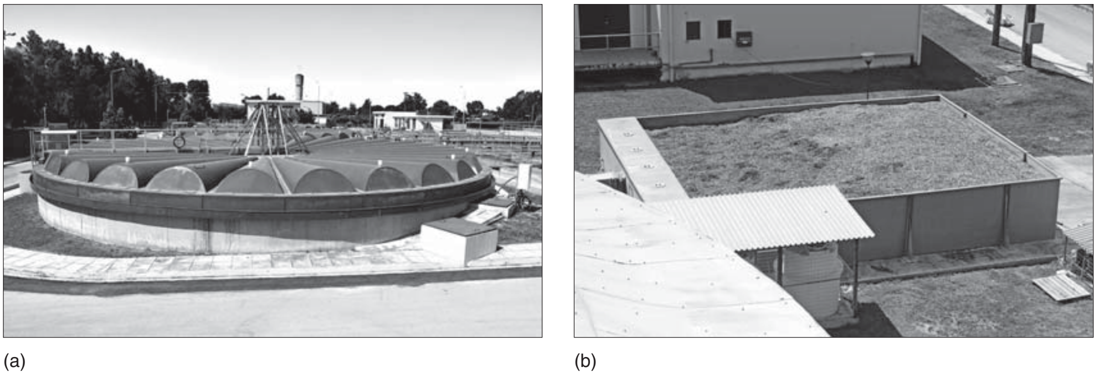


Figure 18–7

Typical odor control facilities for new and existing wastewater treatment plants: (a) covered primary sedimentation facilities (b) compost filter for odor control for odorous gases from covered primary sedimentation tanks.

Urbanization Along Coastal Areas. Currently, about 50 percent of the population of the United States lives within 80 km (50 mi.) of a coastal area; about the same ratio applies worldwide. Worldwide it is projected that up to 60 percent of the world's population will live near coastal areas by 2025 to 2030. The implications for water resources management of such a population shift from rural to coastal areas are significant. For example, withdrawing water from inland areas, transporting it to coastal population centers, treating and using it once, and then discharging it to coastal waters is, in the long run, unsustainable. Clearly, wastewater must be reused if the accumulation of large populations along coastal areas is to remain a viable option.

Although irrigation with treated wastewater has been occurring for decades, it is reaching logistical and economic constraints. In general, agricultural irrigation with reclaimed water is not feasible for most cities due to the long distance between the large sources of recycled water (e.g., coastal cities) and the major agricultural demand (rural areas). Further, the cost and disruption to construct a separate pipe system to convey recycled water back to agricultural areas and the need to provide winter water storage facilities are significant impediments for agricultural reuse. Thus, if significant amounts of wastewater are to be reused, the solution is to implement either IPR or the DPR of purified water in the existing water distribution system (see Fig. 18–8). In the future, because it is inevitable that DPR will become part of the water management portfolio, it is important that water and wastewater agencies begin to develop the necessary information that will allow DPR to become a reality (Haarhoff and van der Merwe, 1995; Leverenz et al., 2011; Tchobanoglous et al., 2011).

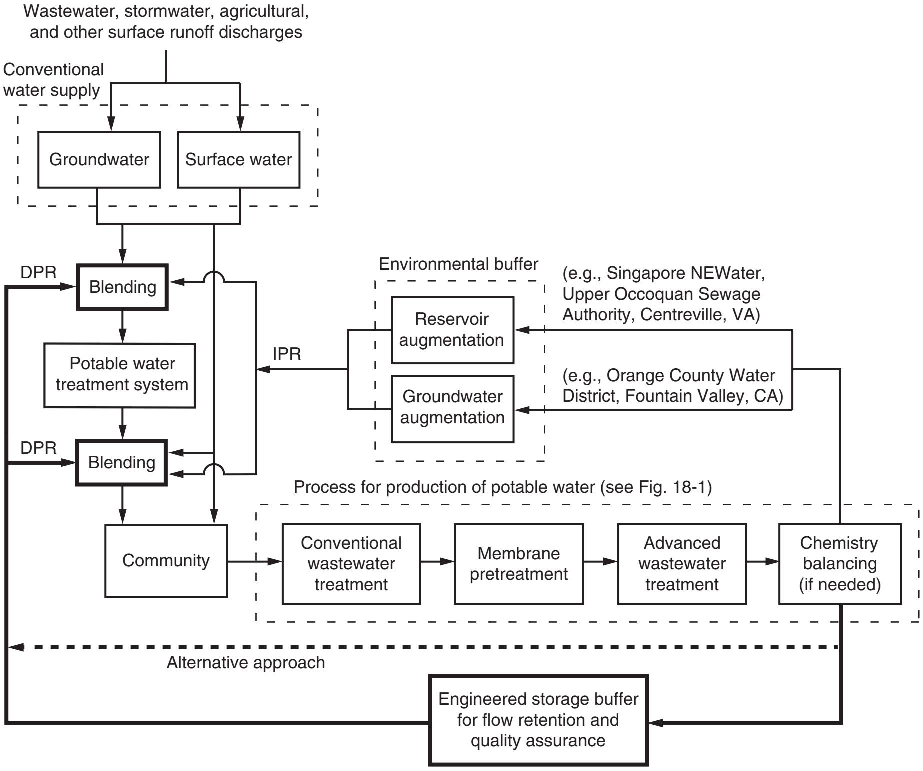
Impact of Climate Change and Sea Level Rise

Climate change and sea level rise have already had an impact on wastewater management facilities, but moving forward, even greater impacts should be anticipated and must be accounted for in the planning, design, and implementation of wastewater management facilities. Examples of the impacts of climate change and sea level rise that must be considered are reviewed in the following discussion.

Climate Change. The most immediate impacts of climate change are reflected in increased temperature, increased evaporation rates, earlier snowmelt, and reduced or increased

Figure 18–8

Definition sketch for direct and indirect potable reuse. The bold solid line corresponds to a system in which an engineered storage buffer is used to replace the environmental buffer used for indirect potable reuse. The bold dashed line corresponds to a DPR system in which an engineered storage buffer is not used. (Adapted from Tchobanoglous et al., 2011.)



rainfall events. Although the number of rainfall events has decreased in a number of locations, the intensity of rainfall events has increased. While the impacts of climate change on the water sector are perhaps most severe, the impacts of extreme rainfall events on wastewater management systems have also been significant, including (1) extreme flooding events, exceeding the capacity of the existing collection systems; (2) increased pumping costs; (3) the discharge of untreated stormwater runoff; (4) damage to collection system infrastructure; (5) flooding of wastewater treatment plants; (6) washout of biological treatment process at treatment plants with limited storage capacity in the collection system; and (7) flows beyond the capacity of the disinfection facilities. The impact of extreme increases in rainfall may necessitate significant increases in peak capacity and/or in system retention. Because such rainfall events can be expected to continue in the future, planning efforts must be undertaken to assess how to best adapt to these changes. Increasing temperatures are also of concern because as the wastewater in the collection system gets warmer, effluent temperature TMDL values may be exceeded. A discussion of the potential costs of adaptation of water and wastewater infrastructure to climate change may be found in NACWA (2009).

Sea Level Rise. Coupled with the effects of climate change, sea level rise has already impacted a number of facilities located on or near coastal areas. A significant impact of sea level rise is that it contributes to increased stormwater flooding along coasts and tidal rivers and estuaries. When tidal surges occur, the discharge from treatment plants located along the path of the surge is often blocked, leading to the release of untreated wastewater. Localized flooding in low lying areas is also exacerbated because stormwater tide-gates cannot open. Because sea level rise also increases the level of sea-water intrusion along with the hydrostatic pressure, the design of facilities for coastal areas must account for these occurrences. Corrective measures that have been considered include the construction of levees and sea walls, raising the elevation of equipment prone to flooding, relocating stormwater discharge locations, pumping effluent through existing gravity outfalls, improvements to the collection system to reduce extraneous flows and, in the extreme, relocating wastewater management facilities.

Impact of Uncontrollable Events

In addition to the future challenges identified in Sec. 18–1 and the impact of demographics, climate change, and sea level rise discussed previously, wastewater treatment plants are subject to the effects of uncontrollable events such as natural disasters and the price of chemicals and supplies.

Natural Disasters. Natural disasters are sudden events such as hurricanes, floods, cyclones, earthquakes, and brushfires caused by natural phenomena that result in the loss of life and extensive property damage. Natural disasters that come to mind readily with respect to their impact on wastewater management facilities are Hurricane Katrina, which occurred in 2005, and the resulting damage to the wastewater management facilities of the city of New Orleans, Louisiana; the string of earthquakes that struck New Zealand in 2011; and Superstorm Sandy that struck the East coast of the United States in 2012. The effects of Katrina were magnified by the tidal surge which breeched the levees that were supposed to protect against flooding. The lesson from Katrina and Superstorm Sandy is that when thinking about natural disasters with respect to the construction of new facilities or upgrading existing facilities, the unthinkable must be thought.

Chemical Costs. The impact of chemical costs that are difficult to control, especially for relatively small wastewater treatment plants, must be considered carefully. In many communities, treatment processes have been abandoned because of increases in the cost of chemicals. Because chemical costs are generally beyond the control of small municipalities, it is important to consider designs that will minimize the need for chemicals.

Impact of the Law of Unintended Consequences

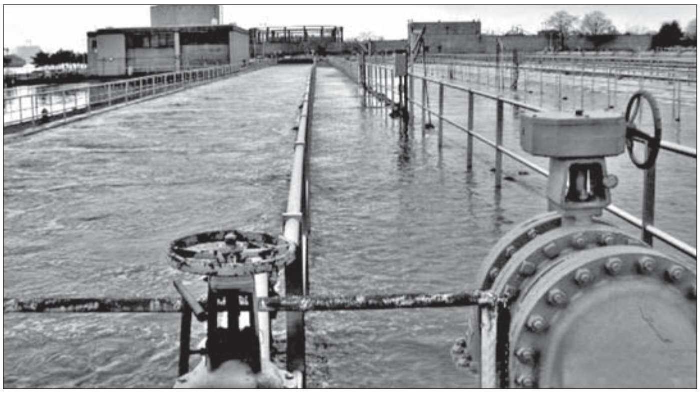
Unintended consequences are outcomes that are not anticipated or intended by a particular action. Even beyond all of the factors considered previously, unintended consequences must be anticipated, as resources and cost for wastewater management become more restrictive. The field of environmental engineering is littered with monuments to the law of unintended consequences. Some examples are considered below.

Treatment Plant Siting. The unintended consequences of locating treatment plants near coastal areas, as described previously, is a prime example of an unintended consequence. When the treatment plants were located originally, little or no thought was given to sea level rise or the subsequent development that would amplify the impact of tidal surges (see Fig. 18–9). To mitigate the unintended consequences will now require the construction of expensive levees or a sea wall and/or the installation of well points to depress the groundwater level or even the possibility of having to relocate one or more of the treatment plants, an unbelievably costly undertaking.

Location of Stormwater Storage Basins. Another example of an unanticipated consequence is related to the stormwater storage basins, located around the periphery of the City of San Francisco. When the storage basins were designed and constructed, the storage basins were equipped with discharge weirs that were located so that they would

Figure 18–9

Flooding at the NYC DEP Bronx WPCP on March 2001. (Courtesy of CU-CCSR and NASA-GISS.)



only be breeched under an extreme rainfall event when the storage capacity of the basins was exceeded (see Fig. 18–10). However, with sea level rise and the subsequent development which channelized the slough to which the basins discharge, flow from tidal surges now overtops the weirs, allowing the seawater to enter the storage basins. The presence of sea water containing high sulfate concentrations has led to excessive hydrogen sulfide corrosion. To remedy the situation, the overflow weirs will have to be raised and pumps will be needed to pump the excess flow. Here again, the designers did not consider the potential impacts of sea level rise or that the subsequent development would amplify the impact of tidal surges.

Water Conservation. In the twenty-first century, water conservation has become a goal for most water and wastewater management agencies. Water conservation is an important element of Leadership in Energy and Environmental Design (LEED) certification. In simple terms LEED is a rating system for buildings. Points can be accumulated for features such as energy conservation, water conservation, water reuse, and mitigation of stormwater. The impact of water conservation, as discussed in Chap. 3, has been significant with respect to wastewater flowrates and constituent concentrations. In the past, per capita wastewater flowrates greater than 450 L/capita·d (120 gal/capita·d) were common. In the not-so-distant future, it is reasonable to assume the per capita flowrates could decrease to below 150 L/capita·d (40 gal/capita·d). Such a decrease would have a

Figure 18–10

Impact of sea level rise on the operation of stormwater storage basins in San Francisco, CA. (Courtesy of City of San Francisco, CA.)

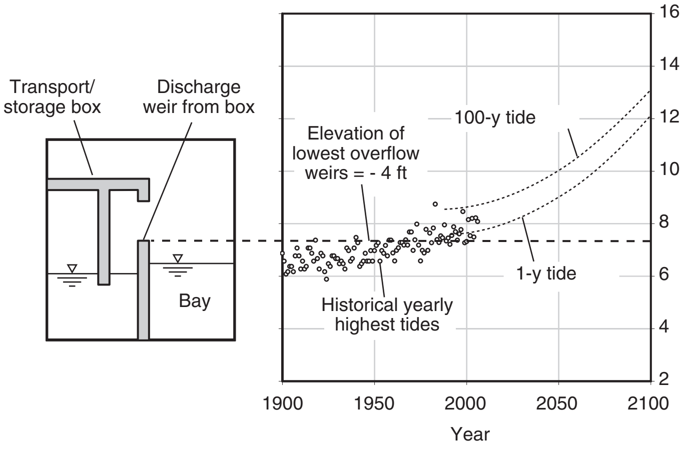
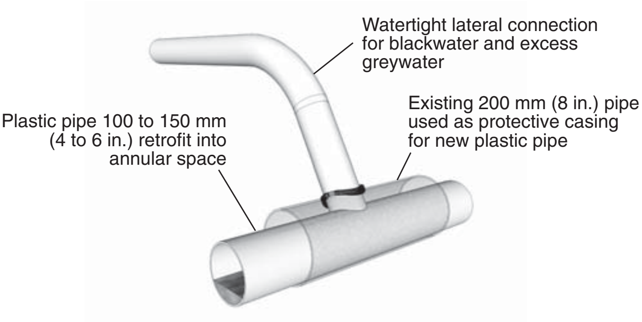


Figure 18–11

Use of a pipe within a pipe to accommodate reduced wastewater flows from residences.



significant impact on wastewater collection systems with respect to the release of odors, grease accumulations, and hydrogen sulfide corrosion.

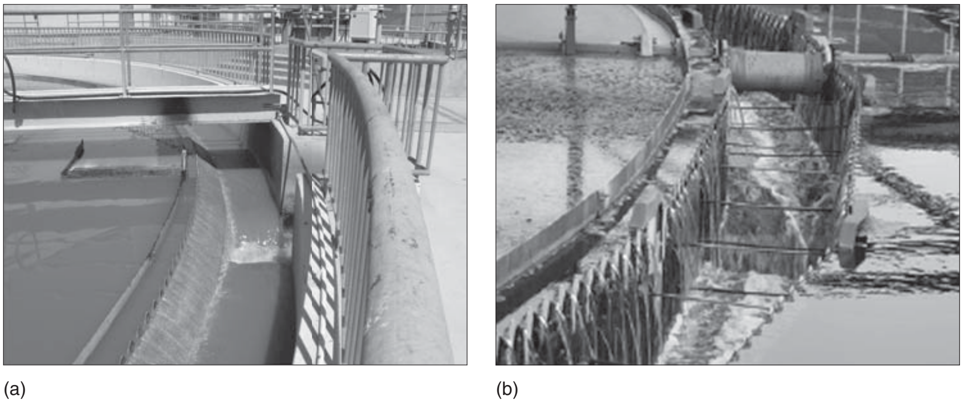
Most wastewater collection systems were designed to be self-cleaning at flowrates on the order of 380 to 450 L/capita·d (100 to 150 gal/capita·d). At reduced flowrates, wastewater solids and grease tend to accumulate and undergo anaerobic decomposition. In general, the release of odors locally has not been a significant problem. However, the transport of hydrogen sulfide by surface water friction has resulted in a significant increase in the rate of corrosion of downstream facilities. In some cities the rate of corrosion has increased by a factor of eight as compared to historic rates of corrosion. Grease accumulations resulting in flow blockages are another important issue associated with reduced wastewater flowrates.

In the future, with reduced rainfall and drought conditions associated with climate change, hydrogen sulfide corrosion and grease accumulation problems will continue to increase and exacerbate the problem of collection system maintenance. One mitigation measure that has been proposed is a pipe within a pipe (see Fig. 18–11). In such an arrangement, a smaller diameter plastic pipe would be placed within an existing collection system. Regardless of the mitigation measures, water conservation will continue to have a number of unintended consequences for the operation and maintenance of wastewater collection systems. Developing workable solutions will be a major change for wastewater management agencies.

Treatment Plant Hydraulics. In the past, when energy was $0.02–0.03/kWh, little attention or effort was devoted to minimizing treatment plant hydraulic headlosses. For example, as illustrated on Fig. 18–12, the water free-fall over the weirs in the primary

Figure 18–12

Examples of excessive headloss at a primary clarifier launder discharge box.


(a)

(b)

sedimentation tank to the collection box is more than 1 m (3.25 ft). Such a large loss of head results from the use of weir equations, based on having a free nappe. With a more nuanced analysis, the free-fall could probably be reduced by 50 percent, or more, resulting in a significant savings in energy when the cost for energy in many location is now approaching \$0.15/kW h, and higher during peak demand periods. In the future, it will be important to revisit all aspects of treatment plant hydraulics to achieve a effective utilization of energy.

18–3 UPGRADING TREATMENT PLANT PERFORMANCE THROUGH PROCESS OPTIMIZATION AND/OR OPERATIONAL CHANGES

Establishing proper control of treatment processes was always considered a primary duty of treatment plant operating personnel. In the past, the criterion of good control and operation was producing an effluent with water quality indicators that on the average did not exceed limits established by a regulatory authority, typically prescribed in a NPDES permit. In the twenty-first century, the operational challenge for existing WWTPs will be to improve treatment performance to meet more stringent discharge requirements, to meet new constituent discharge requirements as discussed in Sec 4–2 in Chap. 4, to minimize the use of resources, and to meet some of the challenges identified in Secs. 18–1 and 18–2.

In recent years, many secondary treatment plants were converted to biological nutrient removal and water reclamation facilities. Some plants are even required to produce an effluent suitable for the subsequent production of potable water. In a number of cases, because of funding limitations for plant expansion, plants operate in excess of their design capacity. At the same time, more emphasis is being placed on operational reliability, and, in some cases, plants are cited and fined for even the slightest violation of water quality standards. Finally, because of the need to improve plant efficiency and reduce capital and operating cost, privatization of facilities construction and operation has entered the wastewater field, resulting in cost competition between public and private operating entities.

Process Optimization

As a result of the social, economic, and technological changes discussed previously and throughout this book, the wastewater treatment plant operating staff is faced with many operational challenges. Often, these challenges can be addressed by optimizing process parameters and operational procedures, modernizing facilities, and retrofitting existing equipment and processes. Development of such improvements usually requires the use of sophisticated tools and protocols, rather than just intuitive approaches. Analysis of plant operating data is a first step in facility evaluation and performance optimization. Utilization of several methods can simplify this analysis including the use of (1) histograms, (2) linear correlation, (3) online process monitoring, (4) computer models, and (5) pilot scale testing. Each of these methods is described in Table 18–2 and expanded upon below. Plant optimization through process modification is considered in the following section.

Use of Histograms. A histogram is a graphical representation of the frequency with which an event occurs over a range of different values for the same variable. An example is the number of times in a year the TSS in the secondary effluent is from 0 to 9.99, 10 to 19.9, 20 to 29.9, and greater than 30 mg/L. Histograms are particularly useful if the

Table 18–2

Methods used to evaluate process performance

Method	Description
Histograms	Graphs displaying frequency of occurrence of parameters such as wastewater characteristics, flow, and cost of chemicals and electricity
Linear correlation	A statistical method used to evaluate data from historical records such as flowrates and water quality parameters
Online process monitoring	Instruments are used continuously or intermittently to record and track important operating parameters such as flowrates, dissolved oxygen concentrations, chlorine residual concentrations, and tank levels. Data from such monitoring can be used to identify trends in operation so that process changes can be implemented before a problem occurs
Computer models	Computer modeling is a useful tool to simulate existing process operations and the effects of possible changes such as modifications to operating strategies or the addition of new equipment or processes
Pilot scale testing	Pilot scale testing is useful in evaluating the performance of new or alternative technologies and in developing criteria that can be used for the design of full scale facilities

frequency of occurrence of undesirable values, such as effluent TSS, is high. Depending on the frequency of occurrence and critical nature of the parameter, significant changes may have to be made to the process operation or perhaps to the process unit infrastructure to achieve correction. Infrequent occurrences of undesirable values often mean that only minor adjustments to operation and maintenance practices may be necessary to achieve the desired results.

Another application of histograms is in establishing the priorities for the tasks that have to be performed to avoid an undesirable event. This type of histogram, called a Pareto chart, represents a graph displaying relative contribution of small problems causing one larger problem. The Pareto chart was named after Italian economist Vilfredo Pareto (1848–1923) who made the famous observation that 80 percent of the world wealth was owned by only 20 percent of people. This principle is valid in many situations and can be equated to process operations, as 80 percent of process problems are due to 20 percent of the causes. Pareto analysis is used to identify those 20 percent that led to majority of process troubles. Analysis includes following four steps:

1. Identifying the causes of a particular problem
2. Determining the frequency of each cause
3. Calculating the percentage of each cause of the total occurrences
4. Plotting the percentage in descending order

Use of a Pareto chart in analyzing the occurrence of tank overflows due to specific causes, such as instrumentation and electrical instrumentation failures, is illustrated on Fig. 18–13.

Use of Linear Correlation. If the relationship between two variables is linear, a linear *correlation* is said to exist. The strength of the relationship is defined by the correlation coefficient, which can vary between -1 to $+1$. Linear correlation is often used to evaluate relationship between a parameter that needs to be optimized and other variables

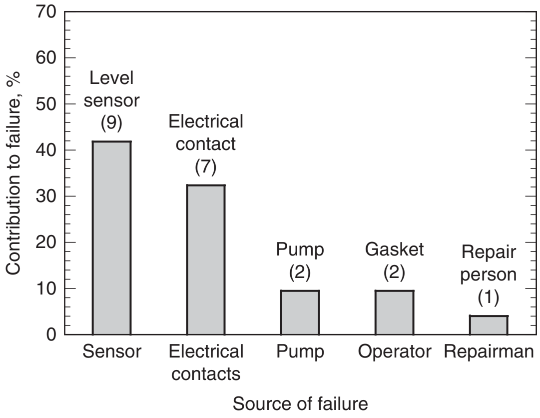


Figure 18–13

Pareto plot used to illustrate relative contribution of various causes to overall failure.

that may have some effect on that parameter. Many models used in process control are linear because they are well understood and are simple to analyze and solve. Usually, linear correlation is written as follows:

$$y = a_0 + a_1x_1 \tag{18–1}$$

where y = parameter to be optimized
a_0, a_1 = coefficients
x_1 = variable

The higher the probability that coefficient a_1 is not zero, the higher is the effect of the variable x_1 on the parameter y. This probability can be calculated using various statistical software packages. Usually, a 95 percent probability that a coefficient a_1 is not equal to zero is considered an indication that a variable x_1 has an effect on the optimized parameter y. The sign of the a_1 coefficient is used to determine in which direction the variable affects an optimized parameter. It is also important to note that lack of correlation between an optimized parameter and a variable does not necessarily mean that they are not related in general. In some cases, an appropriate relationship may be more complex than just a linear relationship (for example, the relationship may be log-normal as discussed in the Sec. 12–9 on UV disinfection), or the effect of their relationship was minimal within examined ranges.

Use of Online Process Monitoring. Since the 1950s, many industries have been using a variety of statistical analyses for making continuous process improvements. In the wastewater industry, statistical process control is not as popular because most effluent water quality characteristics are not distributed normally (i.e., bell-shaped). As a result, the methods required for implementation of statistical control for municipal wastewater treatment processes have to be more sophisticated than ones used in other industries.

Conventional analysis of historical data includes the analysis of data obtained through grab and composite sampling. Such data often do not reflect full dynamics of a treatment process. Online instrumentation, however, can be used to provide information that is more representative of the process dynamics. An example of online monitoring of dissolved oxygen (DO) for process control is shown on Fig. 18–14. The DO values vary significantly from the target value of 2 mg/L, indicating for much of the time the wastewater is overaerated. Installation of automatic DO control can prevent over- as well as underaeration resulting in savings in aeration energy and improved process control.

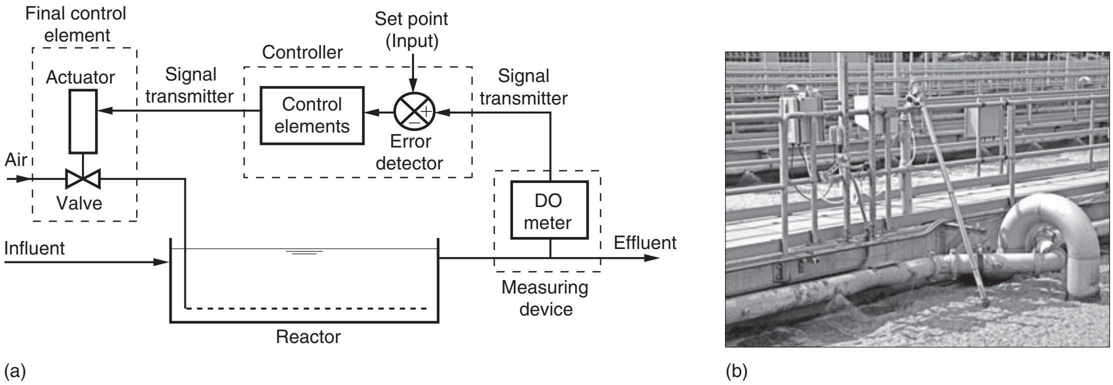


Figure 18–14

Dissolved oxygen (DO) control in an activated sludge biological treatment process: (a) general structure and components of an oxygen control loop and (b) typical installation of online meter used to monitor DO in activate sludge aeration tank.

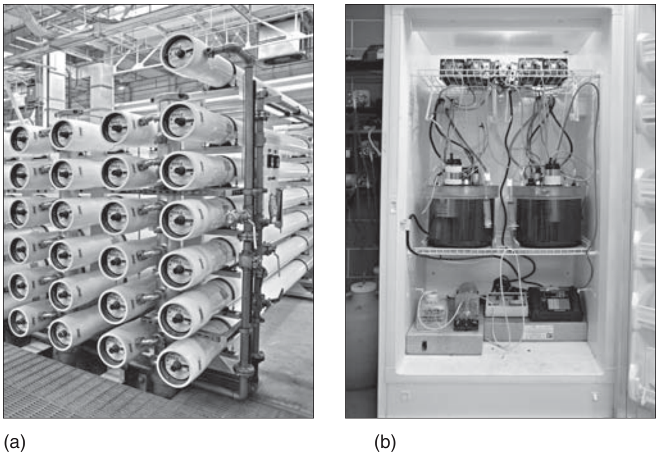
The use of online ammonium analyzers and nitrate analyzers have been increasing at the WWTPs with biological nutrient removal and addition of external carbon for denitrification. The analyzers are sometimes used to control the dosing of the external carbon in combination with the process control algorithm. In some treatment plants, online phosphorus analyzers have been used to control chemical dosing for the chemical phosphorus removal process.

Use of Computer Models. Computer modeling of treatment processes is an effective tool for development of facility improvements because it is now possible to understand the dynamics of treatment processes and to optimize operational parameters. The effect of changes in the operational modes on process performance can also be analyzed. For example, the results of computer modeling of the activated sludge process and the hydraulic modeling of clarifiers can be used to evaluate whether an existing activated sludge plant, designed for conventional treatment (i.e., BOD and TSS removal) can be modified to remove nutrients or/and to treat higher flow than it was designed for. Computer modeling is also very helpful in the evaluation of retrofit options, such as the installation of baffles in aeration tanks and clarifiers, changing of media size in filters, and so forth. In many cases, computer modeling can simplify physical modeling. The use of modeling in activated sludge design is discussed in Sec. 8–5 in Chap. 8.

To provide reliable results, a computer model requires calibration to verify the values that have been chosen for the coefficients included in the model. These coefficients often depend on wastewater characteristics and the design features of the particular process. Calibration involves modifying the coefficient values until model output matches the plant performance data collected using online water quality and flow monitors, composite samplers, and other data collection equipment. Sometimes to improve the quality of the model calibration, the process and equipment need to be operated under extreme conditions. Such experiments are called *stress testing* and require careful planning and execution to avoid threatening plant reliability. Poor planning of stress testing at the Chernobyl nuclear power station, for example, caused the worst nuclear disaster in the history of mankind.

Figure 18–15

Full and bench scale pilot plant facilities: (a) reverse osmosis module shown above the first row of modules is used to test different RO membranes in a side-by-side comparison with existing membrane modules and (b) laboratory scale reactors set up in the temperature control incubator at 25°C to study the anammox process.



Use of Pilot-Scale Testing. A full-scale treatment process is the ultimate tool for evaluation of proposed improvements [see Fig. 18–15(a)]. In recent years, more use has been made of laboratory (bench) and pilot-scale testing of proposed improvements before they are implemented at full scale. Pilot scale testing was discussed previously in Sec. 4–2 in Chap. 4; laboratory-scale biological test reactors are shown on Fig. 18–15(b). Pilot scale units can be operated under a variety of different conditions, and the quality of effluent and other parameters can be compared between old and new modes of operation. Such piloting not only provides the most accurate information about the benefits of proposed process improvements, but also provides an opportunity for training the operating staff in the operation of new processes.

Operational Changes to Improve Plant Performance

In some cases, process modifications can be made at little or no cost. Some of the common problems faced in the operation of treatment facilities relate to the changing nature of the wastewater to be treated, variations in flows and loads, utilizing and maintaining each process component to achieve its maximum capability, and maintaining quality control so that treated effluent and plant residuals meet exacting reuse or permit requirements. Examples of operational changes that can be used to resolve plant performance issues are presented in Table 18–3. An example would be the turndown of the aeration system in an activated sludge plant during low flow periods to reduce over-aeration and to save on energy costs.

Where a more comprehensive approach is needed, a capacity evaluation may be required to determine the capability of the critical elements in the treatment process. For example, short-circuiting may exist in the primary clarifiers resulting in the carryover of settleable material to the biological treatment units, thus causing overloading conditions and increasing the cost of treatment. Stress testing of the process components may be necessary to determine the limits of operating capability of the key components. In other cases, tracer testing to document hydraulic flow patterns (see Appendix H and I) or measuring dissolved oxygen transfer rates to determine aeration efficiency may be employed. These types of tests can also be used to determine if new physical facilities are required.

Table 18–3

Examples of operational changes that can be used to improve plant performance

Issue	Possible remedial action	See Sec. no.
General plant		
Odors from open tanks and channels	Reduce turbulence by controlling water levels to eliminate free falls and splashing	
	Add chemicals (such a chlorine, ferric chloride, or hydrogen peroxide) to influent wastewater	6–2, 6–3
	Modify process loading	
Wide influent flow variations	Conduct collection system I/I investigation to identify sources of extraneous flow	
Short circuiting in clarifiers and chlorine contact tanks	Conduct tracer tests	12–6
	Conduct stress tests	
Headworks		
Odorous grit	Adjust air flowrates in aerated grit chambers to obtain cleaner grit	
	Add lime to dewatered grit	
Odors and vectors in headworks	Cover grit and screenings receptacles	16–3
	Add lime to dewatered grit and screenings	
Inadequate grit removal	Analyze channel and grit chamber hydraulics	5–5
	Adjust flow distribution to grit chamber	
	Add temporary baffles to prevent short circuiting	
	Adjust airflowrate in aerated grit chambers	
Grit deposition in channels	Modify/adjust channel flow-through velocity	
Primary clarifiers		
Poor solids removal	Check for short circuiting/modify baffling	
	Improve flow distribution	5–6
	Add chemicals to influent	6–3
	Reduce return flows from other processes	15–2
Low solids concentration in primary sludge	Modify sludge pumping rate/install timers	
	Increase sludge blanket depth	
Aeration tanks		
Low DO	Determine DO profile in tank and adjust air flowrate	
	Conduct oxygen transfer test	5–11
	Assess diffuser fouling/clean fouled diffusers	
	Change conventional plug flow operation to step feed (if possible)	
	Check wastewater characteristics (rbCOD and nbVSS)	8–2
High DO	Turn off aeration equipment during low flow and loading periods	
	Install timers to control blower or mechanical aerator operation	
Filamentous organisms in mixed liquor	Examine mixed liquor microscopically to identify types of organisms	8–3
	Increase sludge wasting	
	Chlorinate return sludge	

(*continued*)

Table 18–3 (*Continued*)

Issue	Possible remedial action	See Sec. no.
Process is nitrifying	Check SRT	
Process is not nitrifying	Check SRT, alkalinity, and temperature	
Low pH	Check if nitrification is occurring	8–6
	Add alkalinity	8–6
Foaming	Identify nature of foam	8–3
	Change MLSS concentration	8–5
	Add defoaming agent to spray water	
Nocardia foaming	Use dilute chlorine solution spray on foam	
	Reduce oil and grease discharges to collection system	
Trickling filters		
Poor BOD removal	Reduce dosing rate	9–2
Solids washout at peak flows	Reduce recirculation rate	
Biological nitrogen removal process		
Inadequate removal	Check rbCOD, MLSS, and temperature/revise feed	
	Increase external carbon feed	
Final clarifiers		
Bulking sludge	Increase dissolved oxygen concentration	
	Increase F/M ratio	
	Modify return and waste activated sludge pumping rates	
	Chlorinate influent wastewater	
	Chlorinate return activated sludge	
Rising sludge	Increase return activated sludge pumping rate to reduce sludge blanket depth in clarifier	
	Conduct state point analysis of final clarifiers	8–10
Poor solids separation	Perform state point analysis	8–10
Disinfection		
High coliform count in effluent	Improve chlorine mixing	12–6
	Conduct tracer tests to determine short circuiting	12–6
	Check process for is partial nitrification	8–6
Solids processing		
Low solids concentration from dissolved air flotation thickener	Check air-solids ratio	
	Reduce solids loading rate	
	Add/increase polymer feed	
Poor anaerobic digester performance	Change frequency of solids feeding	13–9
	Increase concentration of feed solids	
	Check adequacy of mixing	
	Remove sludge and grit deposits	
	Increase SRT	

(*continued*)

Table 18–3 (*Continued*)

Issue	Possible remedial action	See Sec. no.
Poor aerobic digester performance	Check temperature/adjust SRT Increase concentration of feed solids Check adequacy of mixing Increase DO Check pH/adjust alkalinity	13–9
Odors from composting operations	Increase aeration by air addition or frequency of turning	14–5
Poor compost quality	Perform materials balance/adjust feed composition	14–5
Excessive moisture in compost mixture	Change compost mixture by adding amendment or bulking agent Improve sludge dewatering operations	14–5
Processing of return flows	Use flow equalization to input return flows more uniformly over the day or in the evening hours when excess aeration capacity is available	15–2
	Consider installation of separate treatment facilities for return flows.	15–3
	Consider the recovery of nutrients	15–4, 15–5

18–4 UPGRADING TREATMENT PLANT PERFORMANCE THROUGH PROCESS MODIFICATION

In many cases, to meet some of the many challenges identified in Secs. 18–1 and 18–2, it will be necessary to upgrade existing treatment plants. Upgrading of a wastewater treatment plant can encompass a large number or only a few factors. The scope of this text cannot include identifying and discussing all of the factors that may go into the decision to upgrade a facility; however, some common factors that necessitate upgrading existing treatment facilities include the need to:

1. Improve treatment plant performance
2. Reduce chemical, energy, and maintenance costs
3. Meet more stringent discharge requirements
4. Meet additional treatment capacity needs for population growth
5. Meet new constituent removal requirements for potable reuse
6. Meet new energy and resource recovery objectives

Because these factors have been discussed throughout this textbook, they are not discussed further, but are presented as a basis for understanding the material presented in this section. The various issues related to upgrading existing treatment plants are divided into two categories: (1) upgrading of physical facilities for liquid and solids treatment and (2) potential process modifications for meeting new requirements for constituent removal.

Upgrading Physical Facilities

Most treatment plants contain all of the essential elements necessary to meet treatment goals, but in some instances, the capacity of some of the components is underutilized or overloaded, hydraulic bottlenecks exist that constrain effective and efficient operation, and inadequacies in facilities design affect plant operations and maintenance. Some common issues related to upgrading liquid treatment and solids processing facilities are discussed below.

Upgrading Liquid Treatment Facilities. Upgrading existing facilities may be necessary to mitigate existing operating problems. An example might be the carryover of shredded screenings and hair from the screening process that is causing plugging of membrane filters. Replacement of coarse bar racks or comminutors may be necessary to reduce screenings pass-through and filter plugging. Covering primary sedimentation tanks to limit the release of odors and installing odor control facilities is a common plant upgrade (see Fig. 18–7). Another example is the installation of equalization basins or sidestream treatment processes to accommodate wet-weather flows. Examples of issues relating to the upgrading of physical facilities for the treatment of wastewater (liquid treatment facilities) are summarized in Table 18–4.

In some cases, a capacity evaluation is required in which a full-scale test program is required to determine the capability of the critical elements in the treatment process. For example, short-circuiting may exist in the primary clarifiers resulting in the carryover of settleable material to the biological treatment units, thus causing overloading conditions or increasing the cost of treatment. Stress testing of the process components may be necessary to determine the limits of operating capability of the key components. In other cases, tracer testing to document hydraulic flow patterns or measuring dissolved oxygen transfer rates to determine aeration efficiency may be employed.

Upgrading Solids Processing Facilities. Although a major focus of treatment plant design and operation is on the liquid treatment facilities because of the standards for treated effluent reuse and disposal, solids processing facilities are receiving increased attention because of the potential to recover energy and resources. Solids processing is often the most vexing problem for many plants because of operational difficulties, increasingly stringent requirements for reuse, and limited options for disposal. However, some of the new technologies described in this text can be used to upgrade the design and operation of solids processing facilities. Example upgrade options are described in Table 18–5. Frequent problems are often associated the return flows and loads from solids processing facilities such as thickening and dewatering. However, some of the new technologies described in this text can be used to upgrade the design and operation of solids processing facilities (see Chaps. 13 and 14). The separate treatment of return flows is considered in Chap. 15.

Upgrading to Meet New Constituent Removal Requirements

As discussed in Chaps. 1, 4, 8, 13, and 15 and in earlier sections of this chapter, standards for constituent removal have changed in recent years and will continue to change as more scientific information is developed and as the reuse of treated wastewater and biosolids becomes increasingly important. Examples of many of the current and future issues and upgrade options in resolving these issues are summarized in Table 18–6.

18–5 MANAGEMENT OF WET-WEATHER FLOWS

Although the management of wet-weather flows has, to date, received considerable attention, it is anticipated that even greater attention will be devoted to this aspect of wastewater management in the future. Sanitary sewer overflows (SSOs) are currently regulated under the Clean Water Act, with the U.S. EPA taking the position that all SSOs are illegal discharges and must be eliminated. It is widely recognized, however, that no matter how well a sanitary sewer collection system is operated and maintained, occasional unintentional discharges occur from almost every system. To date, efforts to develop and implement a national policy to regulate the control of SSOs have not been successful.

Table 18–4

Examples of upgrading of physical liquid treatment facilities to improve plant performance

Issue	Remedial action/upgrade option	See sec. no.
General plant		
Odors	Cover structures	16–3
	Add odor collection and treatment system	
	Reduce turbulence by eliminating free falls and sharp bends	
	Add chemical feed facilities	
Wide influent flow variations	Add upstream flow equalization	3–7
	Install variable speed drives on pumps	
	Install small capacity pumps for low flows	
Flow control/ distribution	Improve flow splitting	
	Add metering	
Return flows from sludge processing facilities upset biological process	Provide flow storage/equalization	15–2
	Provide sidestream treatment of return flows	15–3
	Modify operations/upgrade solids processing facilities to reduce load	
Headworks		
Inadequate screenings removal	Modify/replace screens to prevent screenings carryover	5–1
	Install fine screens	
	Replace comminutors	
Odorous, wet screenings	Install screenings press	5–2
	Replace screens with macerators	
	Enclose and ventilate screening equipment	16–3
Odorous grit	Install grit washer	
	Enclose and ventilate grit equipment	16–3
Inadequate grit removal	Add permanent baffles to prevent short circuiting	
	Replace/upgrade grit removal equipment	5–5
Primary clarifiers		
Inadequate solids removal in primary clarifiers	Add chemical treatment and flocculation	6–2, 6–3
	Add high rate clarification	5–7
	Install baffles at effluent weirs	
Aeration tanks		
Low DO	Install DO probes for DO monitoring	
	Replace coarse bubble with fine pore diffusers	5–11
	Change diffuser placement to a grid pattern	
High DO	Install variable speed drives on centrifugal blowers to provide turndown capability	5–11
	Install inlet guides vanes on centrifugal blowers to provide turndown capability	
	Install variable speed drives on positive displacement blowers	
	Install timers and two-speed motors on mechanical aerators	
	Install automatic DO control system	

(*continued*)

Table 18–4 (*Continued*)

Issue	Remedial action/upgrade option	See sec. no.
Unbalanced DO profile in plug flow aeration tanks	Change to step feed process	8–9
	Add DO control system	
Solids deposition	Increase mixing capacity	5–3
Nocardia foaming	Add selector	8–4
Trickling filter		
Plugging and ponding of rock filters	Install plastic packing	9–2
Odors and poor BOD removal	Increase airflow by improving natural draft or adding ventilation system	9–2
Biological treatment system		
Insufficient reactor and solids separation capacity	Add chemical treatment	6–2, 6–3
	Add high rate clarification to reduce loading on biological treatment system	5–7
	Add membrane bioreactor	8–12
Solids washout from high flow-rates	Add flow equalization	3–7
	Add high rate clarification process for excess flows	5–7
	Use contact stabilization process	
Secondary clarifiers		
Inadequate solids separation in secondary clarifiers	Modify flow distribution	8–11
	Modify circular clarifier center feedwell	8–11
	Add flocculating center feedwell	8–11
	Install baffles at effluent weirs	
	Add tube or plate settlers	5–4
	Modify effluent weir configuration	
	Modify sludge collector to improve solids withdrawal	
Disinfection		
Inadequate chlorine disinfection	Add/replace chlorine mixers	12–6
	Add/modify baffles to reduce short circuiting in chlorine contact tank	
	Add chlorine residual control system	
TSS in effluent	Add depth filters before disinfection	
Excessive chlorine residual	Add chlorine residual analyzer and automatic control system	12–6
	Add dechlorination facilities	
	Replace chlorination system with UV	

SSO Policy Issues

In 1994, a stakeholder process was initiated to consider technical and policy issues pertaining to SSOs. An SSO Subcommittee was formed to evaluate the need for national consistency in regulating SSOs, and defining public policy issues related to collection system operation and maintenance as well as public health and environmental impacts related to SSOs. SSO Subcommittee efforts continued until 1999 and several basic principles were identified as suggested NPDES Permit requirements. In 2001, a Notice of Proposed Rulemaking (NPRM)

Table 18–5

Examples of upgrading physical solids treatment facilities to improve plant performance

Issue	Remedial action/upgrade option	See sec. no.
Thickening		13–6
Low solids concentration in primary sludge	Add gravity thickening Add co-settling thickening	
Insufficient gravity thickening of waste activated sludge	Use alternative thickeners (dissolved air flotation, or centrifuge)	
Alkaline stabilization		13–8
Odor and vector problems in dewatered sludge	Add post-lime stabilization	
Anaerobic digesters		13–9
Excessive hydraulic loading	Add sludge thickening prior to digestion	
Inadequate mixing	Upgrade digester mixing system Install egg-shaped digester	
Poor digestion of mixed primary and biological sludge	Install separate digesters	
Inadequate solids destruction	Install two-phased anaerobic digestion process	
Aerobic digestion		13–10
Insufficient pathogen removal	Increase SRT by adding thickening or additional aerobic digester capacity Add ATAD process	
Inadequate mixing	Increase mixing energy	
Composting		
Excessive plastics and inert material in product to be reused	Install fine screens in plant influent Install sludge screens	5–1 13–5
Dewatering and Drying		14–2, 14–3
Excessive water in dewatered sludge cake	Add sludge thickeners Install high solids centrifuge dewatering Install filter press Add solar drying beds Add heat dryers	
Sludge Lagoons and drying beds		
Odors	Construct turbulence inducing structures	16–3
Land application of biosolids		14–10
Excessive attraction of vectors	Modify preapplication treatment methods or method of biosolids application	
Excessive pathogen levels	For Class A biosolids, use one of the six prescribed alternative treatment alternatives For Class B biosolids, use of three prescribed alternative monitoring or treatment alternatives	

Table 18–6

Potential process modifications for meeting new standards for constituent removal

Issue	Remedial action/upgrade option	See Sec. no.
TSS discharge standards	Investigate alternative solids separation facilities	5–9
	Chemical treatment to enhance settling	6–2, 6–3
	Addition of tube or plate settlers to final clarifiers	5–7
	Addition of depth filtration	11–5
	Addition of surface filtration	11–6
	Addition of membrane separation	8–12, 11–7
BOD/COD standards	Investigate alternative treatment facilities	
	Supplemental chemical treatment	6–6
	Nitrification	7–9, 8–6
	Combined aerobic biological processes	9–3
	Membrane biological reactors	8–12
	Adsorption	11–9
	Advanced oxidation	6–8
Removal of nitrogen and phosphorus	Investigate alternative removal facilities	
	Chemical treatment for phosphorus removal	6–4
	Activated sludge selector	8–4
	Suspended growth processes	
	Nitrification	8–6
	Nitrogen removal	8–7
	Phosphorus removal	6–4, 8–8
	Attached growth processes	9–7
	Ammonia stripping of digester supernatant	15–5
	Ion exchange for nitrogen removal	11–11
New disinfection standards	Add depth filtration (prior to disinfection)	11–5, 11–6
	Improve chlorine mixing and dispersion	12–6
	Add dechlorination system	12–5
	Replace chlorination with UV	12–9
VOC emission requirements	Investigate alternative advanced treatment systems	
	Adsorption	16–4
	Air stripping	16–4
	Advanced oxidation	6–8
Removal of residual solids for water reuse	Investigate alternative advanced treatment systems	
	Depth filtration	11–5
	Surface filtration	11–6
	Microfiltration	11–7
	Activated carbon adsorption	11–9
	Ion exchange	11–11
	Advanced oxidation	6–8

(*continued*)

Table 18–6 (*Continued*)

Issue	Remedial action/upgrade option	See Sec. no.
Removal of trace constituents	Investigate alternative treatment systems	
	Chemical precipitation and oxidation	6–6, 6–8
	Microfiltration and/or reverse osmosis	11–7
	Microfiltration and/or reverse osmosis with UV oxidation	6–8, 11–7
Part 503 biosolids regulations for Class A land application	Investigate alternative processes to further reduce pathogens (PFRP) including	14–5
	Thermophilic aerobic digestion	13–9
	Composting	14–5
	Heat drying	14–3
	Heat treatment	14–3
Part 503 biosolids regulations for Class B land application	Investigate alternative processes to significantly reduce pathogens (PSRP) including	
	Lime stabilization	13–8
	Anaerobic digestion	13–9
	Aerobic digestion	13–10
	Composting	14–5
	Air drying	14–3

intended to further development of these basic principles was signed by the EPA Administrator but rulemaking efforts never advanced.

SSO Guidance

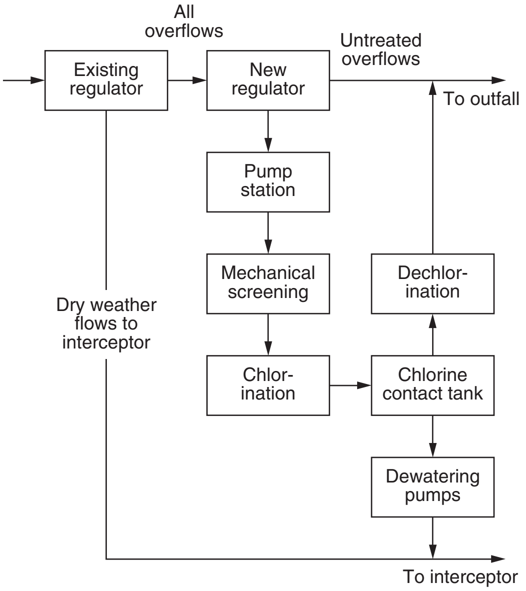
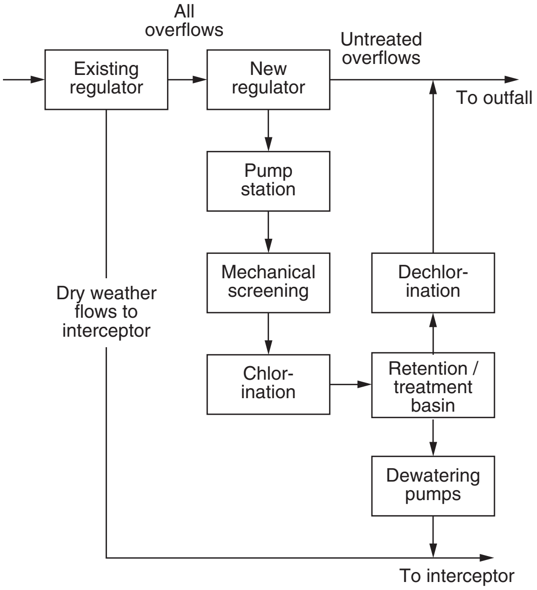
In 2005 EPA issued a guidance document that contains most of what was intended to be in the original rule relative to these basic principles, which have become known as CMOM – Capacity, Management, Operations, and Maintenance. Individual EPA Regional offices have required implementation of CMOM principles through the NPDES Permit process. During 2010, a series of listening sessions were held by EPA to obtain public input for use by EPA in considering whether and how to modify NPDES Regulations as they pertain to municipal sanitary sewer systems and SSOs. EPA is currently preparing a summary of the input received.

Wet-Weather Management Options

Wet weather treatment systems and processes are designed similar to conventional wastewater treatment processes in many respects. The key difference between wet-weather treatment and conventional wastewater treatment is the highly variable, intermittent nature of wet weather flow as compared to the continuous (24 h/d, 365 d/y) operation of a conventional wastewater treatment process. To withstand the rigors of constant ON/OFF, wet-dry cycles and widely varying flows and loads, wet weather treatment processes must be designed to be simple, rugged, and reliable. In assessing the applicability of any technology, system, or process for wet weather treatment, the designer must be assured that it can withstand the highly variable operating conditions outlined previously. A number of treatment technologies are identified and discussed in Table 18–7. Each of the processes

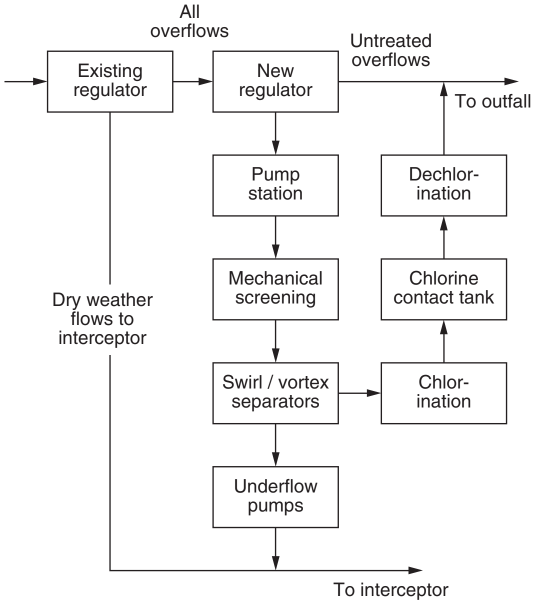
Table 18–7

Representative wet-weather treatment processes in various combinations

Unit processes	Description
Screening and disinfection  	Screening and disinfection facilities combine these two unit processes to control the two most prevalent constituents in CSOs in terms of contribution to water quality impacts—floatable debris and pathogens. Screening equipment must be selected to withstand repeated wet-dry cycles, and screens that avoid moving parts below the water surface are preferred. Screen opening sizes can range from approximately 12 to 20 mm (0.5 to 0.75 in.) down to 3 to 4 mm (0.1 to 0.15 in.). Screenings are normally containerized for disposal following each storm event but, in some cases, can also be macerated and introduced into a sewer tributary to a downstream WWTP. Disinfection is normally accomplished using sodium hypochlorite due to the proven nature of high-rate hypochlorite disinfection and the ability of the equipment to start and stop automatically in response to storm events. Hypochlorite contact times of 15 min are common, but times as short as five min have been used in combination with high-intensity, induction mixers to introduce the hypochlorite into the screened flow. When required, dechlorination is accomplished by applying liquid sodium bisulfite following disinfection
Detention/treatment  	Detention/treatment facilities employ gravity sedimentation (see Chap. 5) in combination with offline storage to reduce impacts of CSOs. Unit processes typically include: influent screening (to capture floating pollutants and remove large objects), gravity sedimentation, sodium hypochlorite disinfection, and de-chlorination using liquid sodium bisulfite (when required). High-rate disinfection practices, using induction mixers, are customarily used. High-rate disinfection practices improve disinfection efficiency, which is important as disinfection takes place concurrent with sedimentation. Ancillary processes typically include detention basin dewatering and flushing systems, and odor control. Dewatering and flushing systems are automated typically to reduce staffing requirements, with flushing gates and tipping buckets are used commonly to scour settled solids from the tank bottom following each storm event. Side water depths ranging from 3 to 6 m (10 to 20 ft), and surface overflowrates (SORs) on the order of 180 to 240 m/d (4500 to 6000 gal/ft^2·d) are used in conjunction with the peak design flow to size detention basins. These high SORs take into account the transient nature of wet weather flows, meaning that the peak design flow may occur for only minutes during each storm event that causes the facility to activate, and that for much of the storm the actual SOR is much lower. In addition to pollutant removal achieved by gravity sedimentation, a portion of the influent flow remains in the basin following each storm event. On an annual basis, the percent removal achieved by this captured volume is significant due to the prevalence of small storms that can be largely captured in the detention basin. Chemically-enhanced detention treatment is an emerging wet weather treatment process that involves the addition of a coagulant (typically a metal salt) and flocculent (polymer) to improve solids removal efficiency. When chemically-enhanced detention treatment is used, high-rate disinfection follows in a separate contact basin.

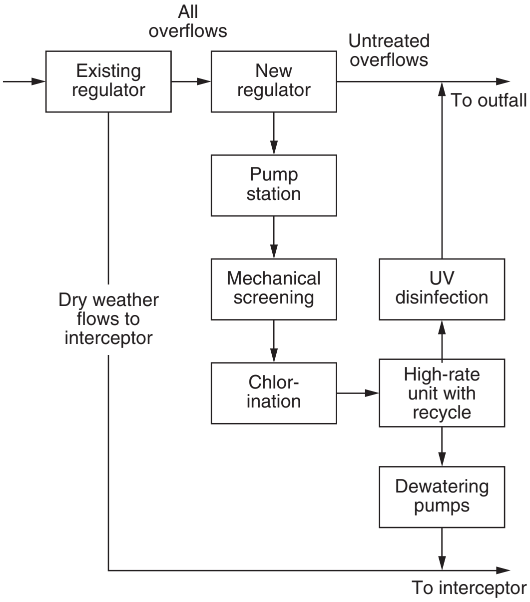
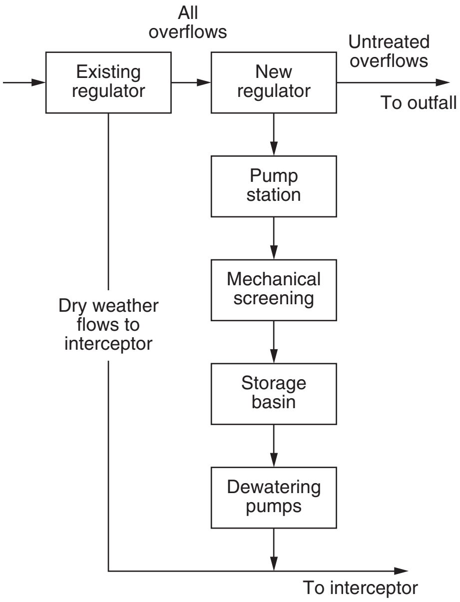
(*continued*)

Table 18–7 (*Continued*)

Unit processes	**Description**
Swirl/vortex 	Swirl/vortex treatment uses a combination of gravitational and rotational forces to enhance liquid-solids separation. Flow is introduced tangentially into a circular basin with specific geometry to direct the flow on a long spiral path. While some rotational motion occurs, gravitational forces acting on the solids as they travel along the spiral flow path are primarily responsible for solids separation. Solids that settle are concentrated at the bottom of the unit and from there are conveyed as a slurry to a sewer or interceptor and on to the WWTP. Swirl/vortex units are equipped with integral screens and/or baffles to capture floating pollutants. For configurations that do not include an integral screen, a separate screening process is provided typically upstream of the swirl/vortex unit. High-rate disinfection, using sodium hypochlorite, is often performed concurrent with swirl/vortex treatment, followed by de-chlorination using liquid sodium bisulfite (when required). Provisions to dewater the unit following each storm event are normally provided, but automated flushing mechanisms and odor control systems are uncommon. There are three specific geometric configurations commonly used for swirl/vortex treatment units: one developed by the U.S. EPA, which is in the public domain, and two proprietary configurations. Details on the geometry of the U.S. EPA design are found in EPA-R2—72-008. Each swirl/vortex unit operates on the basic principles outlined above. It is imperative to characterize the particle settling velocity specific to the wet weather flow to be treated using swirl/vortex technology. The use of settling column tests is highly recommended as a basis for establishing an appropriate surface overflowrate (SOR). For removals that approach a primary sedimentation level of efficiency (TSS removals of 50 percent or more) SORs on the order of 300 to 400 m/d (7500 to 10,000 gal/ft^2·d) are typical. Higher SORs, on the order of 600 to 1200 m/d (15,000 to 30,000 gal/ft^2·d), can be used with recognition that a swirl/vortex unit sized at those SORs will generally only remove floatable pollutants and heavier solids (grit). Chemically-enhanced swirl/vortex treatment is an emerging wet weather treatment process that involves the addition of a coagulant (typically a metal salt) and flocculent (polymer) to improve solids removal efficiency. When chemically-enhanced swirl/vortex treatment is used, high-rate disinfection follows in a separate contact basin.

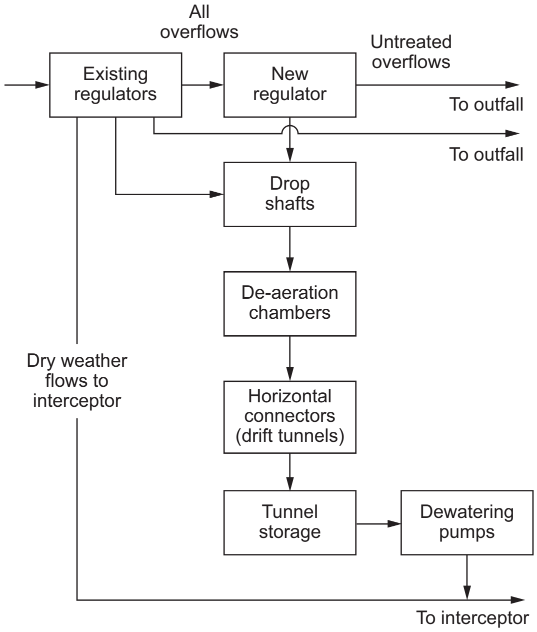
(*continued*)

Table 18–7 (*Continued*)

Unit processes	Description
Ballasted flocculation 	Ballasted flocculation is a physical-chemical treatment process that uses a recycled media and chemical addition to improve the settling characteristics of solids. By agglomerating and increasing the specific gravity of the solids particles, effective solids separation can be accomplished at surface overflowrates (SORs) many times higher than with conventional sedimentation, ranging from 0.80 to 3.25 m/min (20 to 80 $\text{gal/ft}^2\cdot\text{min}$). Appurtenant unit processes include fine screening (screens with clear openings of 6 mm (0.25 in.) or smaller are generally required), and may include grit removal. A separate grit removal process is typically included for wet weather applications with higher activation frequencies and volumes to mitigate contamination of the ballast with grit particles. Because coagulant (typically a metal salt) and flocculent (polymer) are added to enhance solids separation, disinfection and de-chlorination (if necessary) are provided downstream of ballasted flocculation. Because ballasted flocculation produces a high-quality effluent, UV disinfection provides an alternative to sodium hypochlorite disinfection. When used for wet weather treatment the ballasted flocculation process train incorporates an automated shut-down procedure to partially drain the system and recycle the ballast to the influent end of the process. There are two ballasted flocculation process systems that have been used successfully for wet weather treatment. One uses recycled sludge from the process as ballast and the other uses a fine "microsand," which is recovered using hydrocyclones. Both processes involve dosing the influent flow with coagulant upstream of initial rapid mixing, then adding polymer and ballast to increase the size and specific gravity of solids in the flow. As the flow passes through a series of mixing compartments the mixing intensity decreases, allowing the particles to agglomerate. Solids separation usually takes place in a circular clarifier equipped with plate or tube settlers to provide a greater effective surface area for settling. While there are many functional similarities between the two processes, both have specific advantages and disadvantages which must be carefully weighed relative to each unique application.
Storage basins 	Storage basins are used to hold excess wet weather flow temporarily during and shortly after storm events, and can be installed at the WWTP or at satellite locations. These basins fill during periods when conveyance and/or treatment capacity is not available and are drained when capacity is available. Basin volume is generally determined by defining the excess (or overflow) volume corresponding to a design storm event and annual performance is checked using a collection system model. Ancillary processes typically include detention basin dewatering and flushing systems, and odor control. Influent screens are normally provided to prevent large objects (sticks, bricks, etc) from entering the basin. Because the purpose of these screens is protection from overflows rather than treatment, screen openings are relatively large 25 to 75 mm (1 to 3 in.), which mitigates the quantities of screenings to be handled at satellite locations.

(*continued*)

Table 18–7 (*Continued*)

Unit processes	Description
Tunnel storage 	Tunnel storage is used to temporarily hold excess wet weather flow during and shortly after storm events in the same manner as storage basins. While tunnel storage often mitigates peak flows to be handled at a WWTP, it is typically implemented in the collection system. Tunnel volume is generally determined by defining the excess (or overflow) volume corresponding to a design storm event and annual performance is checked using a collection system model. Ancillary processes typically include a tunnel dewatering system and odor control. Storage tunnels are normally designed to be self-cleansing but, if low velocities are expected, flushing systems can be included. A means of preventing large objects from entering the tunnel, such as a coarse bar rack or baffle arrangement can be devised, and screens sized with 25 to 75 mm (1 to 3 in.) clear openings to prevent large objects (sticks, bricks, etc) from entering the dewatering pump station at the downstream end of the tunnel are normally provided.

described in Table 18–7 has performed satisfactorily under various wet-weather operating conditions. The challenge moving forward is how to integrate the technologies identified and others into existing wastewater management programs.

DISCUSSION TOPICS

18–1 Review three articles on asset management for wastewater treatment facilities and discuss the benefits and potential risks in implementing asset management as discussed in the articles.

18–2 A large development is planned at the outskirt of a city with 60,000 population, and population is expected to increase by 20,000 in the next 10 y. Currently the city has one wastewater treatment facility, which is nearing its plant capacity. Discuss the potential advantages and/or disadvantages of considering decentralized wastewater management based on the geographic and climatic characteristics of your location.

18–3 Assuming that the intensity of wet-weather events has continued to increase, what measures would you propose investigate to alleviate the potential washout of solids at a treatment plant with shallow secondary clarifiers [3 m (10 ft) side water depth].

18–4 Discuss advantages and disadvantages of converting a combined wastewater collection system into separate wastewater and stormwater collection systems.

18–5 Discuss advantages and disadvantages of converting separate wastewater and stormwater collection systems into a combined wastewater collection system.

18–6 Discuss the benefits and drawbacks of either of the actions proposed in Problems 18–4 and 18–5 with respect to the management of stormwater.

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Conversion Factors

Appendix A

Table A–1

Unit conversion factors, SI units to U.S. customary units and U.S. customary units to SI units

To convert, multiply in direction shown by arrows					
SI unit name	**Symbol**	→	←	**Symbol**	**U.S. customary unit name**
Acceleration					
meters per second squared	m/s^2	3.2808	0.3048	ft/s^2	feet per second squared
meters per second squared	m/s^2	39.3701	0.0254	in./s^2	inches per second squared
Area					
hectare (10,000 m^2)	ha	2.4711	0.4047	ac	acre
square centimeter	cm^2	0.1550	6.4516	in.^{2}	square inch
square kilometer	km^2	0.3861	2.5900	mi^2	square mile
square kilometer	km^2	247.1054	4.047×10^{-2}	ac	acre
square meter	m^2	10.7639	9.2903×10^{-2}	ft^2	square foot
square meter	m^2	1.1960	0.8361	yd^2	square yard
Energy					
kilojoule	kJ	0.9478	1.0551	Btu	British thermal unit
joule	J	2.7778×10^{-7}	3.6×10^{6}	kW·h	kilowatt-hour
joule	J	0.7376	1.356	ft·lb$_f$	foot-pound (force)
joule	J	1.0000	1.0000	W·s	watt-second
joule	J	0.2388	4.1876	cal	calorie
kilojoule	kJ	2.7778×10^{-4}	3600	kW·h	kilowatt-hour
kilojoule	kJ	0.2778	3.600	W·h	watt-hour
megajoule	kJ	0.3725	2.6845	hp·h	horsepower-hour
Force					
newton	N	0.2248	4.4482	lb$_f$	pound force
Flowrate					
cubic hectometers per day	hm^3	264.1720	3.7854×10^{3}	Mgal/d	million gallons per day
cubic meters per day	m^3/d	264.1720	3.785×10^{-3}	gal/d	gallons per day
cubic meters per day	m^3/d	2.6417×10^{-4}	3.7854×10^{3}	Mgal/d	million gallons per day

(*continued*)

Table A–1 (*Continued*)

To convert, multiply in direction shown by arrows					
SI unit name	**Symbol**	→	←	**Symbol**	**U.S. customary unit name**
cubic meters per second	m^3/s	35.3147	2.8317×10^{-2}	ft^3/s	cubic feet per second
cubic meters per second	m^3/s	22.8245	4.3813×10^{-2}	Mgal/d	million gallons per day
cubic meters per second	m^3/s	15850.3	6.3090×10^{-5}	gal/min	gallons per minute
liters per second	L/s	22,824.5	4.3813×10^{-2}	gal/d	gallons per day
liters per second	L/s	2.2825×10^{-2}	43.8126	Mgal/d	million gallons per day
liters per second	L/s	15.8508	6.3090×10^{-2}	gal/min	gallons per minute
Length					
centimeter	cm	0.3937	2.540	in.	inch
kilometer	km	0.6214	1.6093	mi	mile
meter	m	39.3701	2.54×10^{-2}	in.	inch
meter	m	3.2808	0.3048	ft	foot
meter	m	1.0936	0.9144	yd	yard
millimeter	mm	0.03937	25.4	in.	inch
Mass					
gram	g	0.0353	28.3495	oz	ounce
gram	g	0.0022	4.5359×10^{2}	lb	pound
kilogram	kg	2.2046	0.45359	lb	pound
megagram (10^3 kg)	Mg	1.1023	0.9072	ton	ton (short: 2000 lb)
megagram (10^3 kg)	Mg	0.9842	1.0160	ton	ton (long: 2240)
Power					
kilowatt	kW	0.9478	1.0551	Btu/s	British thermal units per second
kilowatt	kW	1.3410	0.7457	hp	horsepower
watt	W	0.7376	1.3558	$ft\text{-}lb_f/s$	foot-pounds (force) per second
Pressure (force/area)					
Pascal (newtons per square meter)	Pa (N/m^2)	1.4504×10^{-4}	$6.8948 \text{ x } 10^3$	$lb_f/in.^2$	pounds (force) per square inch
Pascal (newtons per square meter)	Pa (N/m^2)	2.0885×10^{-2}	47.8803	lb_f/ft^2	pounds (force) per square foot
Pascal (newtons per square meter)	Pa (N/m^2)	2.9613×10^{-4}	3.3768×10^3	in. Hg	inches of mercury (60°F)
Pascal (newtons per square meter)	Pa (N/m^2)	4.0187×10^{-3}	2.4884×10^2	in. H_2O	inches of water (60°F)
kilopascal (kilonewtons per square meter)	kPa (kN/m^2)	0.1450	6.8948	$lb_f/in.^2$	pounds (force) per square inch
kilopascal (kilonewtons per square meter)	kPa (kN/m$^{2)}$	0.0099	1.0133×10^2	atm	atmosphere (standard)

(*continued*)

Table A–1 (*Continued*)

To convert, multiply in direction shown by arrows					
SI unit name	**Symbol**	**→**	**←**	**Symbol**	**U.S. customary unit name**
Temperature					
degree Celsius (centigrade)	°C	1.8(°C) + 32	0.0555(°F) − 32	°F	degree Fahrenheit
degree kelvin	K	1.8(K) − 459.67	0.0555(°F) + 459.67	°F	degree Fahrenheit
Velocity					
kilometers per second	km/s	2.2369	0.44704	mi/h	miles per hour
meters per second	m/s	3.2808	0.3048	ft/s	feet per second
Volume					
cubic centimeter	cm^3	0.0610	16.3781	in.^{3}	cubic inch
cubic hectometer (100 m × 100 m × 100 m)	hm^3	8.1071×10^2	1.2335×10^{-3}	ac·ft	acre·foot
cubic hectometer	hm^3	264.1720	3.7854×10^3	Mgal	million gallons
cubic meter	m^3	35.3147	2.8317×10^{-2}	ft^3	cubic foot
cubic meter	m^3	1.3079	0.7646	yd^3	cubic yard
cubic meter	m^3	264.1720	3.7854×10^{-3}	gal	gallon
cubic meter	m^3	8.1071×10^{-4}	1.2335×10^3	ac·ft	acre·foot
liter	L	0.2642	3.7854	gal	gallon
liter	L	0.0353	28.3168	ft^3	cubic foot
liter	L	33.8150	2.9573×10^{-2}	oz	ounce (U.S. fluid)

Table A–2

Conversion factors for commonly used wastewater treatment plant design parameters

To convert, multiply in direction shown by arrows			
SI units	→	←	**U.S. units**
g/m^3	8.3454	0.1198	lb/Mgal
ha	2.4711	0.4047	ac
hm^3	264.1720	3.785×10^3	Mgal
kg	2.2046	0.4536	lb
kg/ha	0.8922	1.1209	lb/ac
kg/kW·h	1.6440	0.6083	lb/hp·h
kg/m^2	0.2048	4.8824	lb/ft^2
kg/m^3	8345.4	1.1983×10^{-4}	lb/Mgal
kg/m^3·d	62.4280	0.0160	lb/10^3 ft^3·d
kg/m^3·h	0.0624	16.0185	lb/ft^3·h
kJ	0.9478	1.0551	Btu
kJ/kg	0.4299	2.3260	Btu/lb
kPa (gage)	0.1450	6.8948	lb$_f$/in.^{2} (gage)
kPa Hg (60 °F)	0.2961	3.3768	in. Hg (60 °F)
kW/m^3	5.0763	0.197	hp/10^3 gal
kW/10^3 m^3	0.0380	26.3342	hp/10^3 ft^3
L	0.2642	3.7854	gal
L	0.0353	28.3168	ft^3
L/m^2·d	2.4542×10^{-2}	40.7458	gal/ft^2·d
L/m^2·h	0.5890	1.6978	gal/ft^2·d
L/m^2·min	0.0245	40.7458	gal/ft^2·min
m^3/m^2·min	24.5424	4.0746×19^{-2}	gal/ft^2·min
L/m^2·min	35.3420	0.0283	gal/ft^2·d
m	3.2808	0.3048	ft
m/h	3.2808	0.3048	ft/h
m/h	0.0547	18.2880	ft/min
m/h	0.4090	2.4448	gal/ft^2·min
m^2/10^{3}m^3·d	0.0025	407.4611	ft^2/Mgal·d
m^3	1.3079	0.7646	yd^3
m^3/capita	35.3147	0.0283	ft^3/capita
m^3/d	264.1720	3.785×10^{-3}	gal/d
m^3/d	2.6417×10^{-4}	3.7854×10^3	Mgal/d
m^3/h	0.5886	1.6990	ft^3/min
m^3/ha·d	106.9064	0.0094	gal/ac·d
m^3/kg	16.0185	0.0624	ft^3/lb
m^3/m·d	80.5196	0.0124	gal/ft·d
m^3/m·min	10.7639	0.0929	ft^3/ft·min

(*continued*)

Table A–2

(*Continued*)

To convert, multiply in direction shown by arrows			
SI units	→	←	**U.S. units**
$m^3/m^2 \cdot d$	24.5424	0.0407	$gal/ft^2 \cdot d$
$m^3/m^2 \cdot d$	0.0170	58.6740	$gal/ft^2 \cdot min$
$m^3/m^2 \cdot d$	1.0691	0.9354	$Mgal/ac \cdot d$
$m^3/m^2 \cdot h$	3.2808	0.3048	$ft^3/ft^2 \cdot h$
$m^3/m^2 \cdot h$	589.0173	0.0017	$gal/ft^2 \cdot d$
m^3/m^3	0.1337	7.4805	ft^3/gal
$m^3/10^3\ m^3$	133.6805	7.4805×10^{-3}	$ft^3/Mgal$
$m^3/m^3 \cdot min$	133.6805	7.4805×10^{-3}	$ft^3/10^3\ gal \cdot min$
$m^3/m^3 \cdot min$	1,000.0	0.001	$ft^3/10^3\ ft^3 \cdot min$
Mg/ha	0.4461	2.2417	ton/ac
mm	3.9370×10^{-2}	25.4	in.
ML/d	0.2642	3.785	Mgal/d
ML/d	0.4087	2.4466	ft^3/s

Table A–3

Abbreviations for SI units

Abbreviation	SI unit
°C	degree Celsius
cm	centimeter
g	gram
g/m^2	gram per square meter
g/m^3	gram per cubic meter (= mg/L)
ha	hectare (= 100 m × 100 m)
hm^3	cubic hectometer (= 100 m × 100 m × 100 m)
J	Joule
K	Kelvin
kg	kilogram
kg/capita·d	kilogram per capita per day
kg/ha	kilogram per hectare
kg/m^3	kilogram per cubic meter
kJ	kilojoule
kJ/kg	kilojoule per kilogram
kJ/kW·h	kilojoule per kilowatt-hour
km	kilometer (= 1000 m)
km^2	square kilometer
km/h	kilometer per hour
km/L	kilometer per liter
kN/m^2	kiloNewton per square meter

(*continued*)

Table A–3

(*Continued*)

Abbreviation	SI unit
kPa	kiloPascal
ks	kilosecond
kW	kilowatt
L	liter
L/s	liters per second
m	meter
m^2	square meter
m^3	cubic meter
mm	millimeter
m/s	meter per second
mg/L	milligram per liter (= g/m^3)
m^3/s	cubic meter per second
MJ	megajoule
N	Newton
N/m^2	Newton per square meter
Pa	Pascal (usually reported as kilopascal, kPa)
W	Watt

Table A–4

Abbreviations for US customary units

Abbreviation	US Customary Units
ac	acre
ac-ft	acre foot
Btu	British thermal unit
Btu/ft^3	British thermal unit per cubic foot
d	day
ft	foot
ft^2	square foot
ft^3	cubic foot
ft/min	feet per minute
ft/s	feet per second
ft^3/min	cubic feet per minute
ft^3/s	cubic feet per second
°F	degree Fahrenheit
gal	gallon
$gal/ft^2 \cdot d$	gallon per square foot per day
$gal/ft^2 \cdot min$	gallon per square foot per minute
gal/min	gallon per minute

(*continued*)

Table A–4

(*Continued*)

Abbreviation	US Customary Units
h	hour
hp	horsepower
hp-h	horsepower-hour
in.	inch
kWh	kilowatt-hour
lb_f	pound (force)
lb_m	pound (mass)
lb/ac	pound per acre
lb/ac·d	pound per acre per day
lb/capita·d	pound per capita per day
lb/ft^2	pound per square foot
lb/ft^3	pound per cubic foot
lb/in^2	pound per square inch
lb/yd^3	pound per cubic yard
Mgal/d	million gallons per day
mi	mile
mi^2	square mile
mi/h	mile per hour
min	minute
mo	month
ppb	part per billion
ppm	part per million
s	second
ton (2000 lbm)	ton (2000 pounds mass)
wk	week
y	year
yd	yard
yd^2	square yard
yd^3	cubic yard

Physical Properties of Selected Gases and the Composition of Air

Appendix B

B–1 PHYSICAL PROPERTIES OF SELECTED GASES

Table B–1

Molecular weight, specific weight, and density of gases found in wastewater at standard conditions (0°C, 1 atm)[a]

Gas	Formula	Molecular weight	Specific weight, lb/ft³	Density, g/L,
Air	–	28.97	0.0808	1.2928
Ammonia	NH_3	17.03	0.0482	0.7708
Carbon dioxide	CO_2	44.00	0.1235	1.9768
Carbon monoxide	CO	28.00	0.0781	1.2501
Hydrogen	H_2	2.016	0.0056	0.0898
Hydrogen sulfide	H_2S	34.08	0.0961	1.5392
Methane	CH_4	16.03	0.0448	0.7167
Nitrogen	N_2	28.02	0.0782	1.2507
Oxygen	O_2	32.00	0.0892	1.4289

[a] Adapted from Perry, R. H., D. W. Green, and J. O. Maloney: Perry's (eds) (1984) *Chemical Engineers' Handbook*, 6th ed., McGraw-Hill Book Company, New York.

B–2 COMPOSITION OF DRY AIR

Table B–2

Composition of dry air at 0°C and 1.0 atmosphere[a]

Gas	Formula	Percent by volume[b,c]	Percent by weight
Nitrogen	N_2	78.03	75.47
Oxygen	O_2	20.99	23.18
Argon	Ar	0.94	1.30
Carbon dioxide	CO_2	0.03	0.05
Other[d]	–	0.01	–

[a] Note: Values reported in the literature vary depending on the standard conditions.

[b] Adapted from North American Combustion Handbook, 2nd ed., North American Mfg., Co., Cleveland, OH.

[c] For ordinary purposes air is assumed to be composed of 79 percent N_2 and 21 percent O_2 by volume.

[d] Hydrogen, Neon, Helium, Krypton, Xenon.

Note: Molecular weight of air = $(0.7803 \times 28.02) + (0.2099 \times 32.00) + (0.0094 \times 39.95) + (0.0003 \times 44.00) = 28.97$ (see Table B–1 above).

B–3 DENSITY OF AIR AT OTHER TEMPERATURES

In SI units

The following relationship can be used to compute the density of air, ρ_a, at other temperatures at atmospheric pressure.

$$\rho_a = \frac{PM}{RT}$$

where P = atmospheric pressure, $1.0132^5 \times 10^5\ \text{N/m}^2$
M = molecular weight of air (see Table B-1), 28.97 g/ g mole
R = universal gas constant, 8314 N · m/ (mole air · K)
T = temperature, K (273.15 + °C)

For example, at 20°C, the density of air is:

$$\rho_{a,20°C} = \frac{(1.01325 \times 10^5\ \text{N/m}^2)(28.97\ \text{g/ mole air})}{[8314\ \text{N}\cdot\text{m/(mole air}\cdot\text{K)}][(273.15 + 20)\text{K}]}$$

$$= 1.204 \times 10^3\ \text{g/m}^3 = 1.204\ \text{kg/m}^3$$

In U.S. customary units

The following relationship can be used to compute the specific weight of air, γ_a, at other temperatures at atmospheric pressure.

$$\gamma_a = \frac{P\,(144\ \text{in}^2/\text{ft}^2)M}{RT}$$

where P = atmospheric pressure, 14.7 lb/in^2
M = molecular weight of air (see Table B–1), 28.97 lb/ lb mole air
R = universal gas constant, 1544 ft · lb/(lb mole air · °R)
T = temperature, °R (460 + °F)

For example, at 68°F, the specific weight of air is:

$$\gamma_a = \frac{(14.7\ \text{lb/in}^2)(144\ \text{in}^2/\text{ft}^2)(28.97\ \text{lb/lb mole air})}{[1544\ \text{ft}\cdot\text{lb/(lb mole air}\cdot°\text{R)}][(460 + 68)°\text{R}]} = 0.0752\ \text{lb/ft}^3$$

B–4 CHANGE IN ATMOSPHERIC PRESSURE WITH ELEVATION

In SI units

The following relationship can be used to compute the change in atmospheric pressure with elevation.

$$\frac{P_b}{P_s} = \exp\left[-\frac{gM(z_b - z_a)}{RT}\right]$$

where P_b = pressure at elevation z_b, N/m^2
P_s = atmospheric pressure at sea level, $1.01325 \times 10^5\ \text{N/m}^2$
g = acceleration due to gravity, 9.81 m/s^2
M = molecular weight of air (see Table B-1), 28.97 g/mole air
z_b = elevation b, m

z_a = elevation b, ft
R = universal gas constant, 8314 N · m/(mole air · K)
T = temperature, K (273.15 + °C)

In U.S. customary units

The following relationship can be used to compute the change in atmospheric pressure with elevation.

$$\frac{P_b}{P_s} = \exp\left[-\frac{gM(z_b - z_a)}{g_c RT}\right]$$

where P_b = pressure at elevation z_b, lb/in^2
P_s = atmospheric pressure at sea level, lb/in^2
g = acceleration due to gravity, 32.2 ft/s^2
M = molecular weight of air (see Table B–1), 28.97 lb$_m$/lb mole air
z_b = elevation b, ft
z_a = elevation b, ft
g_c = 32.2 ft · lb$_m$/lb · s^2
R = universal gas constant, 1544 ft · lb/(lb mole air · °R)
T = temperature, °R (460 + °F)

Physical Properties of Water **Appendix** C

The principal physical properties of water are summarized in SI units in Table C–1 and in U.S. customary units in Table C–2. They are described briefly below (Vennard and Street, 1975; Webber, 1971).

C–1 SPECIFIC WEIGHT

The specific weight of a fluid, γ, is its weight per unit volume. In SI units, specific weight is expressed in kilonewtons per cubic meter (kN/m^3). The relationship between γ, ρ, and the acceleration due to gravity g is $\gamma = \rho g$.

C–2 DENSITY

The density of a fluid, ρ, is its mass per unit volume. In SI units density is expressed in kilograms per cubic meter (kg/m^3). For water, ρ is 1000 kg/m^3 at 4°C. There is a slight decrease in density with increasing temperature.

C–3 MODULUS OF ELASTICITY

For most practical purposes, liquids may be regarded as incompressible. The bulk modulus of elasticity, E, is given by

$$E = \frac{\Delta p}{(\Delta V/V)}$$

where Δp is the increase in pressure, which when applied to a volume V, results in a decrease in volume ΔV. In SI units, the modulus of elasticity is expressed in kilonewtons per meter squared (kN/m^2).

C–4 DYNAMIC VISCOSITY

The viscosity of a fluid, μ, is a measure of its resistance to tangential or shear stress. In SI units, the dynamic viscosity is expressed in Newton seconds per square meter (N·s/m^2).

C–5 KINEMATIC VISCOSITY

In many problems concerning fluid motion, the viscosity appears with the density in the form μ/ρ, and it is convenient to use a single term, ν, known as the kinematic viscosity. In SI units, the kinematic viscosity is expressed in meters squared per second (m^2/s). The kinematic viscosity of a liquid diminishes with increasing temperature.

C–6 SURFACE TENSION

The surface tension of a fluid, σ, is the physical property that enables a drop of water to be held in suspension at a tap, a glass to be filled with liquid slightly above the brim and yet not spill, or a needle to float on the surface of a liquid. The surface-tension force across any imaginary line at a free surface is proportional to the length of the line and acts in a direction perpendicular to it. In SI units, surface tension per unit length is expressed in Newtons per meter (N/m). There is a slight decrease in surface tension with increasing temperature.

C–7 VAPOR PRESSURE

Liquid molecules that possess sufficient kinetic energy are projected out of the main body of a liquid at its free surface and become vapor. In a system open to the atmosphere, the vapor pressure, p_v, is the partial pressure exerted by the liquid vapor in the atmosphere. In a closed system, the vapor molecules are in equilibrium with the liquid; the pressure exerted by the vapor molecules is known as the saturated vapor pressure. In SI units, vapor pressure is expressed in kilonewtons per square meter (kN/m^2).

REFERENCES

Vennard, J.K., and R.L. Street (1975) *Elementary Fluid Mechanics*, 5th ed., Wiley, New York.

Webber, N.B. (1971) *Fluid Mechanics for Civil Engineers*, SI ed., Chapman and Hall, London.

Table C–1

Physical properties of water (SI units)[a]

Temperature, °C	Specific weight, γ, kN/m^3	Density[b], ρ, kg/m^3	Modulus of elasticity[b], $E/10^6$, kN/m^2	Dynamic viscosity, $\mu \times 10^3$, $N \cdot s/m^2$	Kinematic viscosity, $\nu \times 10^6$, m^2/s	Surface tension[c], σ, N/m	Vapor pressure, p_v, kN/m^2
0	9.805	999.8	1.98	1.781	1.785	0.0765	0.61
5	9.807	1000.0	2.05	1.518	1.519	0.0749	0.87
10	9.804	999.7	2.10	1.307	1.306	0.0742	1.23
15	9.798	999.1	2.15	1.139	1.139	0.0735	1.70
20	9.789	998.2	2.17	1.002	1.003	0.0728	2.34
25	9.777	997.0	2.22	0.890	0.893	0.0720	3.17
30	9.764	995.7	2.25	0.798	0.800	0.0712	4.24
40	9.730	992.2	2.28	0.653	0.658	0.0696	7.38
50	9.689	988.0	2.29	0.547	0.553	0.0679	12.33
60	9.642	983.2	2.28	0.466	0.474	0.0662	19.92
70	9.589	977.8	2.25	0.404	0.413	0.0644	31.16
80	9.530	971.8	2.20	0.354	0.364	0.0626	47.34
90	9.466	965.3	2.14	0.315	0.326	0.0608	70.10
100	9.399	958.4	2.07	0.282	0.294	0.0589	101.33

[a] Adapted from Vennard and Street (1975).

[b] At atmospheric pressure.

[c] In contact with the air.

Table C–2

Physical properties of water (U.S. customary units)[a]

Temperature, °F	Specific weight, γ, lb/ft³	Density[b], ρ, slug/ft³	Modulus of elasticity[b], $E/10^3$, lb$_f$/in.²	Dynamic viscosity, $\mu \times 10^5$, lb·s/ft²	Kinematic viscosity, $\nu \times 10^5$, ft²/s	Surface tension[c], σ, lb/ft	Vapor pressure, p_v, lb$_f$/in.²
32	62.42	1.940	287	3.746	1.931	0.00518	0.09
40	62.43	1.940	296	3.229	1.664	0.00614	0.12
50	62.41	1.940	305	2.735	1.410	0.00509	0.18
60	62.37	1.938	313	2.359	1.217	0.00504	0.26
70	62.30	1.936	319	2.050	1.059	0.00498	0.36
80	62.21	1.934	324	1.799	0.930	0.00492	0.51
90	62.11	1.931	328	1.595	0.826	0.00486	0.70
100	62.00	1.927	331	1.424	0.739	0.00480	0.95
110	61.86	1.923	332	1.284	0.667	0.00473	1.27
120	61.71	1.918	332	1.168	0.609	0.00467	1.69
130	61.55	1.913	331	1.069	0.558	0.00460	2.22
140	61.38	1.908	330	0.981	0.514	0.00454	2.89
150	61.20	1.902	328	0.905	0.476	0.00447	3.72
160	61.00	1.896	326	0.838	0.442	0.00441	4.74
170	60.80	1.890	322	0.780	0.413	0.00434	5.99
180	60.58	1.883	318	0.726	0.385	0.00427	7.51
190	60.36	1.876	313	0.678	0.362	0.00420	9.34
200	60.12	1.868	308	0.637	0.341	0.00413	11.52
212	59.83	1.860	300	0.593	0.319	0.00404	14.70

[a] Adapted from Vennard and Street (1975).

[b] At atmospheric pressure.

[c] In contact with the air.

Statistical Analysis of Data **Appendix** D

The statistical analysis of wastewater flowrate and constituent concentration data involves the determination of statistical parameters used to quantify a series of measurements. Commonly used statistical parameters and graphical techniques for the analysis of wastewater management data are reviewed below.

D–1 COMMON STATISTICAL PARAMETERS

Commonly used statistical measures include the mean, median, mode, standard deviation, and coefficient of variation, based on the assumption that the data are distributed normally. Although the terms just cited are the most commonly used statistical measures, two additional statistical measures are needed to quantify the nature of a given distribution. The two additional measures are the coefficient of skewness, and coefficient of kurtosis. If a distribution is highly skewed, as determined by the coefficient of skewness, normal statistics cannot be used. For most wastewater data that are skewed, it has been found that the log of the value is normally distributed. Where the log of the values is normally distributed, the distribution is said to be log normal. The common statistical measures used for the analysis of wastewater management data (Eqs. D–1 through D–9) are summarized in Table D–1.

D–2 GRAPHICAL ANALYSIS OF DATA

Graphical analysis of wastewater management data is used to determine the nature of the distribution. For most practical purposes, the type of the distribution can be determined by plotting the data on both arithmetic- and logarithmic-probability paper and noting whether the data can be fitted with a straight line. The three steps involved in the use of arithmetic, and logarithmic-probability paper are as follows.

1. Arrange the measurements in a data set in order of increasing magnitude and assign a rank serial number.
2. Compute a corresponding plotting position for each data point using Eqs. (D–10) and (D–11).

   $$\text{Plotting position (\%)} = \left(\frac{m}{n+1}\right) \times 100 \qquad \text{(D–10)}$$

   where m = rank serial number
   n = number of observations

   The term $(n + 1)$ is used to correct for a-small-sample bias. The plotting position represents the percent or frequency of observations that are equal to or less than the indicated value. Another expression often used to define the plotting position is known as Blom's transformation:

   $$\text{Plotting position (\%)} = \frac{m - 3/8}{n + 1/4} \times 100 \qquad \text{(D–11)}$$

3. Plot the data on arithmetic- and logarithmic-probability paper. The probability scale is labeled "Percent of values equal to or less than the indicated value."

Table D–1

Statistical parameters used for the analysis of wastewater management data[a]

Parameter		Definition
Mean value $\bar{x} = \frac{\sum f_i x_i}{n}$	(D–1)	**Terms** $\bar{x}$ = mean value f_i = frequency (for ungrouped data $f_i = 1$) x_i = the mid-point of the ith data range (For ungrouped data x_i = the *i*th observation)
Standard Deviation $s = \sqrt{\frac{\sum f_i (x_i - \bar{x})^2}{n - 1}}$	(D–2)	n = number of observations (Note $\sum f_i = n$) s = standard deviation
Coefficient of variation $C_v = \frac{100\, s}{\bar{x}}$	(D–3)	C_v = coefficient of variation, percent α_3 = coefficient of skewness α_4 = coefficient of kurtosis M_g = geometric mean s_g = geometric standard deviation
Coefficient of skewness $\alpha_3 = \frac{\sum f_i (x_i - \bar{x})^3/(n-1)}{s^3}$	(D–4)	$P_{15.9}$ and $P_{84.1}$ = values from arithmetic or logarithmic probability plots at indicated percent values, corresponding to one standard deviation
Coefficient of kurtosis $\alpha_4 = \frac{\sum f_i (x_i - \bar{x})^4/(n-1)}{s^4}$	(D–5)	**Median value** If a series of observations are arranged in order of increasing value, the middlemost observation, or the arithmetic mean of the two middlemost observations, in a series is known as the median.
Geometric mean $\log M_g = \frac{\sum f_i (\log x_i)}{n}$	(D–6)	**Mode** The value occurring with the greatest frequency in a set of observations is known as the mode. If a continuous graph of the frequency distribution is drawn, the mode is the value of the high point, or hump, of the curve. In a symmetrical set of observations, the mean, median, and mode will be the same value. The mode can be estimated with the following expression. Mode = 3(median) − 2($\bar{x}$).
Geometric standard deviation $\log s_g = \sqrt{\frac{\sum f_i (\log^2 x_g)}{n - 1}}$	(D–7)	**Coefficient of skewness** When a frequency distribution is asymmetrical, it is usually defined as being a skewed distribution.
Using probability paper $s = P_{84.1} - \bar{x}$ or $P_{15.9} + \bar{x}$	(D–8)	
$s_g = \frac{P_{84.1}}{M_g} = \frac{M_g}{P_{15.9}}$	(D–9)	**Coefficient of kurtosis** Used to define the peakedness of the distribution. The value of the kurtosis for a normal distribution is 3. A peaked curve will have a value greater than 3 whereas a flatter curve it will have a value less than 3.

[a] Adapted from Metcalf & Eddy (1991) and Crites and Tchobanoglous (1998).

If the data, plotted on arithmetic-probability paper, can be fit with a straight line, then the data are assumed to be normally distributed. Significant departure from a straight line can be taken as an indication of skewness. If the data are skewed, logarithmic probability paper can be used. The implication here is that the logarithm of the observed values is normally distributed. On logarithmic-probability paper, the straight line of best fit passes through the geometric mean, M_g, and through the intersection of $M_g \times s_g$ at a value of 84.1 percent and M_g/s_g at a value of 15.9 percent. The geometric standard deviation, s_g, can be determined using Eq. D–9 given in Table D–1. The use of arithmetic- and logarithmic-probability paper is illustrated in Example D–1.

EXAMPLE D–1 **Statistical Analysis of Wastewater Flowrate Data.** Using the following weekly flowrate data obtained from an industrial discharger for a calendar quarter of operation, determine the statistical characteristics and predict the maximum weekly flowrate that will occur during a full year's operation.

Week No.	Flowrate, m^3/wk	Week No.	Flowrate, m^3/wk
1	2900	8	3675
2	3040	9	3810
3	3540	10	3450
4	3360	11	3265
5	3770	12	3180
6	4080	13	3135
7	4015		

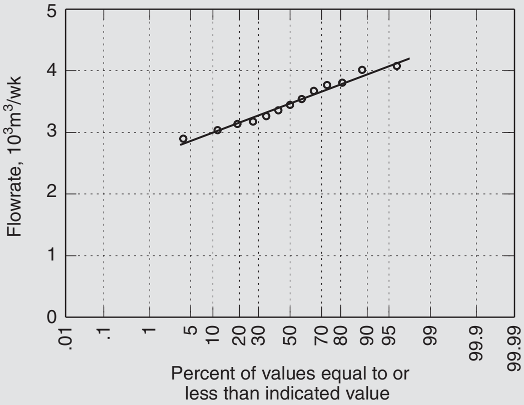
Solution

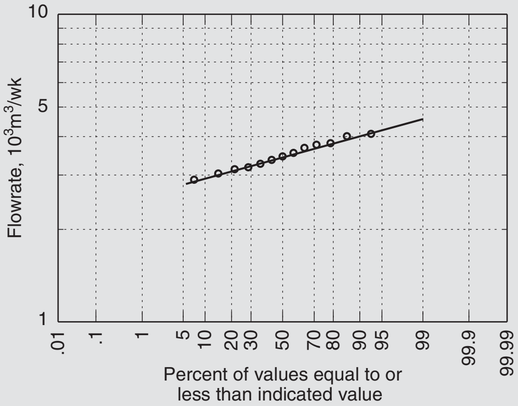
1. Plot the flowrate data using the log/probability method.
   a. Set up a data analysis table with three columns as described below.
      i. In column 1, enter the rank serial number starting with number 1
      ii. In column 2, arrange the flowrate data in ascending order
      iii. In column 3, enter the probability plotting position

Rank serial no., m	Flowrate, m^3/wk	Plotting position,[a] %
1	2900	7.1
2	3040	14.3
3	3135	21.4
4	3180	28.6
5	3265	35.7
6	3360	42.9
7	3450	50.0
8	3540	57.1
9	3675	64.3
10	3770	71.4
11	3810	78.6
12	4015	85.7
13	4080	92.9

[a] Plotting position = $[m/(n + 1)]100$.

b. Plot the weekly flowrates expressed in m^3/wk versus the plotting position. The resulting plots are presented below. Because the data fall on a straight line on both plots, the flowrate data can be described adequately by either distribution. This fact can be taken as indication that the distribution is not skewed significantly and that normal statistics can be applied.





2. Determine the statistical characteristics of the flowrate data.

a. Set up a data analysis table to obtain the quantities needed to determine the statistical characteristics.

Flowrate, m^3/wk	$(x - \bar{x})$	$(x - \bar{x})^2$	$(x - \bar{x})^3$ 10^{-6}	$(x - \bar{x})^4$ 10^{-9}
2900	−578	334,084	−193	11,161
3040	−438	191,844	−84	3680
3135	−343	117,649	−40	1384
3180	−298	88,804	−26	789
3265	−213	45,369	−9.6	206
3360	−118	13,924	−1.6	19.4
3450	−28	784	−0.02	0.06
3540	62	3844	0.24	1.48
3675	197	38,809	7.6	151
3770	292	85,264	25	727
3810	332	110,224	37	1215
4015	537	288,369	155	8316
4080	602	362,404	218	13,134
45,220		1,681,372	88.62	40,784

b. Determine the statistical characteristics using the parameters given in Table D–1.

i. Mean

$$\bar{x} = \frac{\sum x}{n}$$

$$\bar{x} = \frac{45{,}220}{13} = 3478 \text{ m}^3/\text{wk}$$

ii. Median (the middle-most value)
Median = 3450 m^3/wk (see data table above)

iii. Mode
Mode = 3(Median) − 2($\bar{x}$) = 3(3450) − 2(3478) = 3394 m^3/wk

iv. Standard deviation

$$s = \sqrt{\frac{\sum(x - \bar{x})^2}{n - 1}}$$

$$s = \sqrt{\frac{1{,}681{,}372}{12}} = 374.3 \text{ m}^3\text{/wk}$$

v. Coefficient of variation

$$C_V = \frac{100s}{\bar{x}}$$

$$C_V = \frac{100(374.3)}{3478} = 10.8\%$$

vi. Coefficient of skewness

$$\alpha_3 = \frac{[\sum(x - \bar{x})^3/(n - 1)]}{s^3}$$

$$\alpha_3 = \frac{(88.62 \times 10^6/12)}{(374.3)^3} = 0.141$$

vii. Coefficient of kurtosis

$$\alpha_4 = \frac{[\sum(x - \bar{x})^4/(n - 1)]}{s^4}$$

$$\alpha_4 = \frac{(40{,}784 \times 10^9/12)}{(374.3)^4} = 1.73$$

Reviewing the statistical characteristics, it can be seen that the distribution is somewhat skewed (α_3 = 0.141 versus 0 for a normal distribution) and is considerably flatter than a normal distribution would be (α_4 = 1.73 versus 3.0 for a normal distribution).

3. Determine the probable annual maximum weekly flowrate.
   a. Determine the probability factor:

$$\text{Peak week} = \frac{m}{n + 1} = \frac{52}{52 + 1} = 0.981$$

   b. Determine the flowrate from the figure given in Step 1b at the 98.1 percentile:

Peak weekly flowrate = 4500 m^3/wk

Comment The statistical analysis of data is important in establishing the design conditions for wastewater treatment plants. The application of statistical analysis to the selection of design flowrates and mass loadings rates is considered in the Sec. 3–6 in Chap. 3.

Dissolved Oxygen Concentration in Water as a Function of Temperature, Salinity, and Barometric Pressure

Appendix E

Table E–1

The air solubility of oxygen in mglL as functions of temperature and elevation in meters for 0–1800 m[a]

Temp., °C	Elevation above sea level, m									
	0	200	400	600	800	1,000	1,200	1,400	1,600	1,800
0	14.621	14.276	13.94	13.612	13.291	12.978	12.672	12.373	12.081	11.796
1	14.216	13.881	13.554	13.234	12.922	12.617	12.32	12.029	11.745	11.468
2	13.829	13.503	13.185	12.874	12.57	12.273	11.984	11.701	11.425	11.155
3	13.46	13.142	12.832	12.53	12.234	11.945	11.663	11.387	11.118	10.856
4	13.107	12.798	12.496	12.201	11.912	11.631	11.356	11.088	10.826	10.57
5	12.77	12.468	12.174	11.886	11.605	11.331	11.063	10.801	10.546	10.296
6	12.447	12.174	11.866	11.585	11.311	11.044	10.782	10.527	10.278	10.035
7	12.138	11.851	11.571	11.297	11.03	10.769	10.514	10.265	10.022	9.784
8	11.843	11.562	11.289	11.021	10.76	10.505	10.256	10.013	9.776	9.544
9	11.559	11.285	11.018	10.757	10.502	10.253	10.01	9.772	9.54	9.314
10	11.288	11.02	10.759	10.504	10.254	10.011	9.773	9.541	9.315	9.093
11	11.027	10.765	10.51	10.26	10.017	9.779	9.546	9.319	9.098	8.881
12	10.777	10.521	10.271	10.027	9.789	9.556	9.329	9.107	8.89	8.678
13	10.536	10.286	10.041	9.803	9.569	9.342	9.119	8.902	8.69	8.483
14	10.306	10.06	9.821	9.587	9.359	9.136	8.918	8.705	8.498	8.295
15	10.084	9.843	9.609	9.38	9.156	8.938	8.724	8.516	8.313	8.114
16	9.87	9.635	9.405	9.18	8.961	8.747	8.538	8.334	8.135	7.94
17	9.665	9.434	9.209	8.988	8.774	8.564	8.359	8.159	7.963	7.772
18	9.467	9.24	9.019	8.804	8.593	8.387	8.186	7.99	7.798	7.611
19	9.276	9.054	8.837	8.625	8.418	8.216	8.019	7.827	7.639	7.455
20	9.092	8.874	8.661	8.453	8.25	8.052	7.858	7.669	7.485	7.304
21	8.914	8.7	8.491	8.287	8.088	7.893	7.703	7.518	7.336	7.159
22	8.743	8.533	8.328	8.127	7.931	7.74	7.553	7.371	7.193	7.019
23	8.578	8.371	8.169	7.972	7.78	7.592	7.408	7.229	7.054	6.883
24	8.418	8.214	8.016	7.822	7.633	7.449	7.268	7.092	6.92	6.752
25	8.263	8.063	7.868	7.678	7.491	7.31	7.132	6.959	6.79	6.625
26	8.113	7.917	7.725	7.537	7.354	7.175	7.001	6.83	6.664	6.501
27	7.968	7.775	7.586	7.401	7.221	7.045	6.873	6.706	6.542	6.382
28	7.827	7.637	7.451	7.269	7.092	6.919	6.75	6.584	6.423	6.266
29	7.691	7.503	7.32	7.141	6.967	6.796	6.63	6.467	6.308	6.153
30	7.559	7.374	7.193	7.017	6.845	6.677	6.513	6.353	6.196	6.043
31	7.43	7.248	7.07	6.896	6.727	6.561	6.399	6.241	6.087	5.937
32	7.305	7.125	6.95	6.779	6.612	6.448	6.289	6.133	5.981	5.833
33	7.183	7.006	6.833	6.665	6.5	6.339	6.181	6.028	5.878	5.731
34	7.065	6.89	6.72	6.553	6.39	6.232	6.077	5.925	5.777	5.633
35	6.949	6.777	6.609	6.445	6.284	6.127	5.974	5.825	5.679	5.536
36	6.837	6.667	6.501	6.338	6.18	6.025	5.874	5.727	5.583	5.442
37	6.727	6.559	6.395	6.235	6.078	5.926	5.776	5.631	5.489	5.35
38	6.62	6.454	6.292	6.134	5.979	5.828	5.681	5.537	5.396	5.259
39	6.515	6.351	6.191	6.035	5.882	5.733	5.587	5.445	5.306	5.171
40	6.412	6.25	6.092	5.937	5.787	5.639	5.495	5.355	5.218	5.084

[a] From Colt, J. (2012) *Dissolved Gas Concentration in Water: Computation as Functions of Temperature, Salinity and Pressure*, 2nd ed., Elsevier, Boston, MA.

Table E–2

Standard air saturation concentration of oxygen as a function of temperature and salinity in mg/L. 0–40 g/kg (Seawater. I atm moist air)[a]

Temp., °C	Salinity, g/kg								
	0.0	5.0	10.0	15.0	20.0	25.0	30.0	35.0	40.0
0	14.621	14.120	13.635	13.167	12.714	12.276	11.854	11.445	11.050
1	14.216	13.733	13.266	12.815	12.378	11.956	11.548	11.153	10.772
2	13.829	13.364	12.914	12.478	12.057	11.649	11.255	10.875	10.506
3	13.460	13.011	12.577	12.156	11.750	11.356	10.976	10.608	10.252
4	13.107	12.674	12.255	11.849	11.456	11.076	10.708	10.352	10.008
5	12.770	12.352	11.946	11.554	11.174	10.807	10.451	10.107	9.774
6	12.447	12.043	11.652	11.272	10.905	10.550	10.205	9.872	9.550
7	12.138	11.748	11.369	11.002	10.647	10.303	9.970	9.647	9.335
8	11.843	11.465	11.098	10.743	10.399	10.066	9.743	9.431	9.128
9	11.559	11.194	10.839	10.495	10.162	9.839	9.526	9.223	8.930
10	11.288	10.933	10.590	10.257	9.934	9.621	9.318	9.024	8.739
11	11.027	10.684	10.351	10.028	9.715	9.411	9.117	8.832	8.556
12	10.777	10.444	10.121	9.808	9.505	9.210	8.925	8.648	8.379
13	10.536	10.214	9.901	9.597	9.302	9.016	8.739	8.470	8.209
14	10.306	9.993	9.689	9.394	9.108	8.830	8.561	8.299	8.046
15	10.084	9.780	9.485	9.198	8.920	8.651	8.389	8.135	7.888
16	9.870	9.575	9.289	9.010	8.740	8.478	8.223	7.976	7.736
17	9.665	9.378	9.099	8.829	8.566	8.311	8.064	7.823	7.590
18	9.467	9.188	8.917	8.654	8.399	8.151	7.910	7.676	7.448
19	9.276	9.005	8.742	8.486	8.237	7.996	7.761	7.533	7.312
20	9.092	8.828	8.572	8.323	8.081	7.846	7.617	7.395	7.180
21	8.914	8.658	8.408	8.166	7.930	7.701	7.479	7.262	7.052
22	8.743	8.493	8.250	8.014	7.785	7.561	7.344	7.134	6.929
23	8.578	8.334	8.098	7.868	7.644	7.426	7.215	7.009	6.809
24	8.418	8.181	7.950	7.726	7.507	7.295	7.089	6.888	6.693
25	8.263	8.032	7.807	7.588	7.375	7.168	6.967	6.771	6.581
26	8.113	7.888	7.668	7.455	7.247	7.045	6.849	6.658	6.472
27	7.968	7.748	7.534	7.326	7.123	6.926	6.734	6.548	6.366
28	7.827	7.613	7.404	7.201	7.003	6.811	6.623	6.441	6.263
29	7.691	7.482	7.278	7.079	6.886	6.698	6.515	6.337	6.164
30	7.559	7.354	7.155	6.961	6.773	6.589	6.410	6.236	6.066
31	7.430	7.230	7.036	6.847	6.662	6.483	6.308	6.138	5.972
32	7.305	7.110	6.920	6.735	6.555	6.379	6.208	6.042	5.880
33	7.183	6.993	6.807	6.626	6.450	6.279	6.111	5.949	5.790
34	7.065	6.879	6.697	6.521	6.348	6.180	6.017	5.857	5.702
35	6.949	6.768	6.590	6.417	6.249	6.085	5.925	5.769	5.617
36	6.837	6.659	6.486	6.316	6.152	5.991	5.834	5.682	5.533
37	6.727	6.553	6.383	6.218	6.057	5.899	5.746	5.597	5.451
38	6.620	6.450	6.284	6.122	5.964	5.810	5.660	5.514	5.371
39	6.515	6.348	6.186	6.027	5.873	5.722	5.575	5.432	5.292
40	6.412	6.249	6.090	5.935	5.784	5.636	5.492	5.352	5.215

[a] From Colt, J. (2012) *Dissolved Gas Concentration in Water: Computation as Functions of Temperature, Salinity and Pressure*, 2nd ed., Elsevier, Boston, MA.

Carbonate Equilibrium

Appendix F

The chemical species that comprise the carbonate system include gaseous carbon dioxide [$(CO_2)_g$], aqueous carbon dioxide [$(CO_2)_{aq}$], carbonic acid [H_2CO_3], bicarbonate [HCO_3^-], carbonate [CO_2^{-2}], and solids containing carbonates. In waters exposed to the atmosphere, the equilibrium concentration of dissolved CO_2 is a function of the liquid phase CO_2 mole fraction and the partial pressure of CO_2 in the atmosphere. Henry's law (see Chap. 2) is applicable to the CO_2 equilibrium between air and water; thus

$$x_g = \frac{P_T}{H} p_g \tag{F–1}$$

where x_g = mole fraction of gas in water, mole gas/mole water

$$= \frac{\text{mole gas}\,(n_g)}{\text{mole gas}\,(n_g) + \text{mole water}\,(n_w)}$$

P_T = total pressure, usually 1.0 atm

H = Henry's law constant, $\dfrac{\text{atm (mole gas/mole air)}}{\text{(mole gas/mole water)}}$

p_g = mole fraction of gas in air, mole gas/mole of air (Note: The mole fraction of a gas is proportional to the volume fraction.)

The concentration of aqueous carbon dioxide is determined using Eq (F–1). At sea level, where the average atmospheric pressure is 1 atm, or 101.325 kPa, carbon dioxide comprises approximately 0.03 percent of the atmosphere by volume (see Appendix B). Values of the Henry's law constant for CO_2 as a function of temperature are given in Table F–1. The values in Table F–1 were computed using Eq. (2–48) and the data given in Table 2–7 in Chap. 2.

Table F–1

Henry's Law constant for CO_2 as a function of temperature

T, °C	H, atm
0	794
10	1073
20	1420
30	1847
40	2361
50	2972
60	3691

Aqueous carbon dioxide [$(CO_2)_{aq}$] reacts reversibly with water to form carbonic acid.

$$(CO_2)_{aq} + H_2O \rightleftarrows H_2CO_3 \tag{F–2}$$

The corresponding equilibrium expression is

$$\frac{[H_2CO_3]}{[CO_2]} = K_m \tag{F–3}$$

The value of K_m at 25°C is 1.58×10^{-3}. Note that K_m is unitless. Because of the difficulty of differentiating between $(CO_2)_{aq}$ and H_2CO_3 in solution and the observation that very little H_2CO_3 is ever present in natural waters, an effective carbonic acid value ($H_2CO_3^*$) is used which is defined as:

$$H_2CO_3^* \rightleftarrows (CO_2)_{aq} + H_2CO_3 \tag{F–4}$$

Because carbonic acid is a diprotic acid it will dissociate in two steps - first to bicarbonate and then to carbonate. The first dissociation of carbonic acid to bicarbonate can be represented as

$$H_2CO_3^* \rightleftarrows H^+ + HCO_3^- \tag{F–5}$$

The corresponding equilibrium relationship is defined as

$$\frac{[H^+][HCO_3^-]}{[H_2CO_3^*]} = K_{a1} \tag{F–6}$$

The value of first acid dissociation constant K_{a1} at 25°C is 4.467×10^{-7} mole/L. Values of K_{a1} at other temperatures are given in Table F–2, which is repeated here from Table 6–16 in Chap. 6.

The second dissociation of carbonic acid is from bicarbonate to carbonate as given below

$$HCO_3^- \rightleftarrows H^+ + CO_3^{2-} \tag{F–7}$$

The corresponding equilibrium relationship is defined as:

$$\frac{[H^+][CO_3^{2-}]}{[HCO_3^-]} = K_{a2} \tag{F–8}$$

The value of the second acid dissociation constant K_{a2} at 25°C is 4.477×10^{-11} mole/L. Values of K_{a2} at other temperatures are given in Table F–2.

Table F–2

Carbonate equilibrium constants as function of temperature[a]

Temperature, °C	Equilibrium constant[b]	
	$K_{a1} \times 10^7$	$K_{a2} \times 10^{11}$
5	3.020	2.754
10	3.467	3.236
15	3.802	3.715
20	4.169	4.169
25	4.467	4.477
30	4.677	5.129
40	5.012	6.026

[a] Adapted from Table 6–20 in Chap. 6.

[b] The reported values have been multiplied by the indicated exponents.

Thus, the value K_{a2} at 20°C is equal to 4.169×10^{-11}.

The distribution of carbonate species as function of pH is illustrated on Fig. F–1.

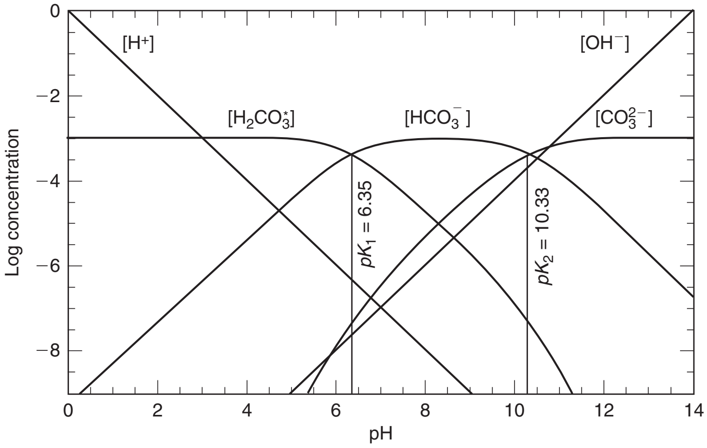


Figure F–1

Log concentration versus pH diagram for a 10^{-3} molar solution of carbonate at 25°C. By sliding the constituent curves up or down, pH values can be obtained at different concentration values. (LD)

To illustrate the use of the data presented in Tables F–1 and F–2, it will be helpful to estimate the pH of a water, assuming the atmosphere above the water contains 0.03 percent CO_2 by volume (see Appendix B), the bicarbonate (HCO_3^-) concentration in the water is 610 mg/L, and the temperature of the water is 20°C.

1. Use Eq. (F–1) to determine concentration of H_2CO_3 in the water. The value of Henry's constant from Table F–1 is 1420 atm, thus

   $$x_{H_2CO_3} = \frac{P_T}{H} p_g = \frac{(1\text{ atm})(0.00030)}{1420\text{ atm}} = 2.113 \times 10^{-7}$$

   Because one liter of water contains 55.6 mole [1000 g/(18 g/mole)], the mole fraction of H_2CO_3 is equal to:

   $$x_{H_2CO_3} = \frac{\text{mole gas}\,(\text{n}_\text{g})}{\text{mole gas}\,(\text{n}_\text{g}) + \text{mole water}\,(\text{n}_\text{w})}$$

   $$2.113 \times 10^{-7} = \frac{[H_2CO_3]}{[H_2CO_3] + 55.6\text{ mole/L}}$$

   Because the number of moles of dissolved gas in a liter of water is much less than the number of moles of water,

   $$[H_2CO_3] \approx (2.113 \times 10^{-7})(55.6\text{ mole/L}) \approx 11.75 \times 10^{-6}\text{ mole/L}$$

2. Use Eq. (F–6) to determine the pH of the water. The value of K_{a1} at 20°C from Table F–2 is 4.169×10^{-7}, thus

   $$[H^+] = \frac{K_{a1}[H_2CO_3^*]}{[HCO_3^-]}$$

   Substitute known values and solve for $[H^+]$

   $$[H^+] = \frac{(4.169 \times 10^{-7})(11.75 \times 10^{-6}\text{ mole/L})}{[(610\text{ mg/L})/61{,}000\text{ mg/mole})]} = 4.90 \times 10^{-10}$$

   $$\text{pH} = -\log[H^+] = -\log[4.90 \times 10^{-10}] = 9.31$$

   From Fig. F–1, the amount of carbonate present is essentially non-measurable.

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Appendix G

Moody Diagrams for the Analysis of Flow in Pipes

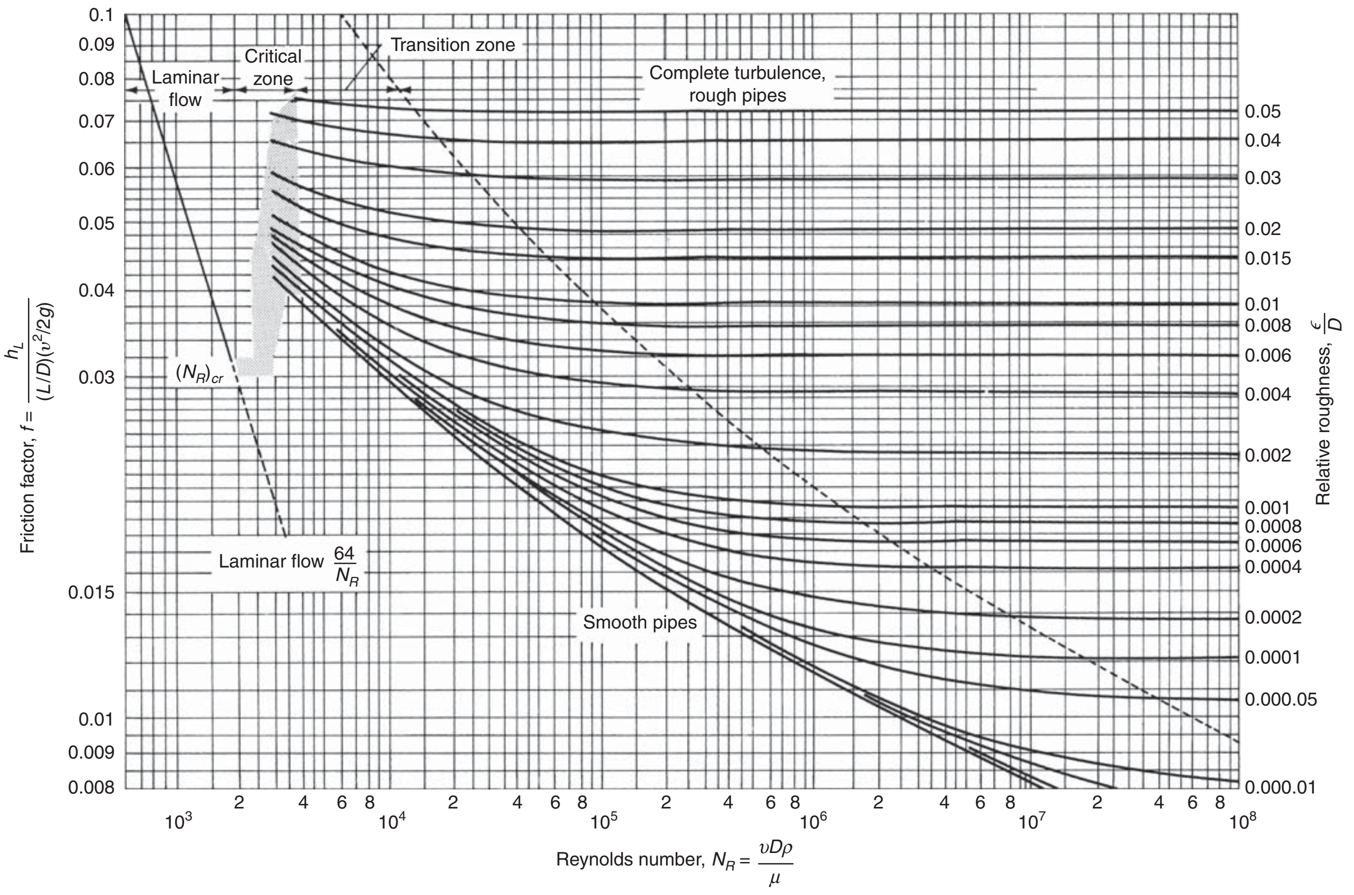
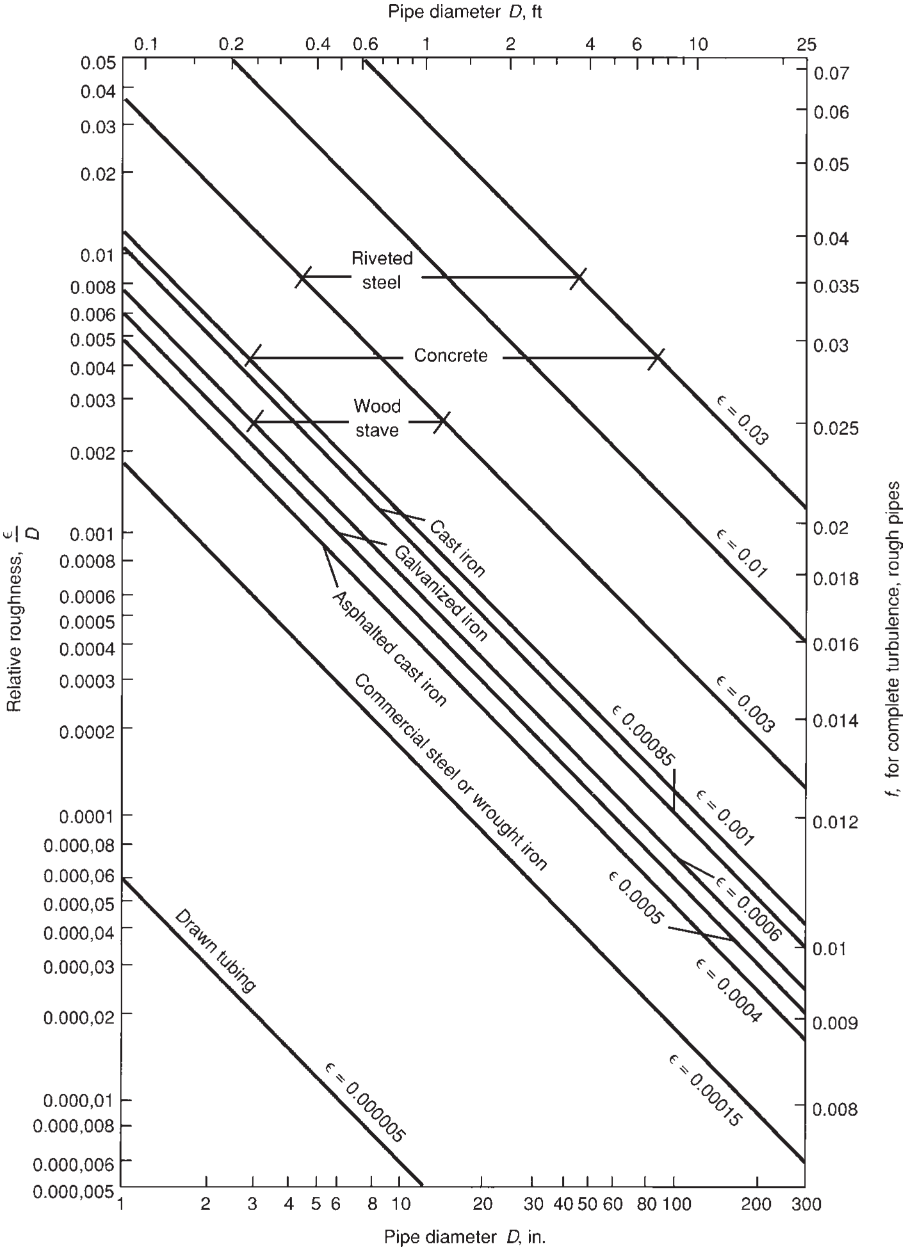


Figure G–1

Moody diagram for friction factor in pipes versus Reynolds number and relative roughness. [From Moody, L.F. (1944) Friction Factors for Pipe Flow, Transactions American Society of Civil Engineers vol. 66, p. 671.]

Figure G–2

Moody diagram for relative roughness as a function of diameter for pipes constructed of various materials. [Adapted from Moody, L.F. (1944) Friction Factors for Pipe Flow, Transactions American Society of Civil Engineers vol. 66, p. 671.]



Analysis of Nonideal Flow in Reactors using Tracers

Appendix H

The analysis of reactor hydraulic performance using tracers is considered in this appendix. Important applications of tracer studies include the assessment of: (1) short circuiting in sedimentation tanks and biological reactors, (2) the effective contact time in chlorine contact basins, (3) hydraulic approach conditions in UV reactors, and (4) patterns in constructed wetlands and other natural treatment systems. Tracer studies are also of critical importance in assessing the degree of success that has been achieved with corrective measures. Topics considered in this appendix include: (1) factors leading to nonideal flow in reactors, (2) important characteristics of tracers, (3) analysis of tracer response curves, and (4) practical interpretation of tracer measurements. The types of tracers used and the conduct of tracer tests are discussed in Sec. 12-3 in Chap. 12. The discussion of nonideal flow in this section will also serve as an introduction to the modeling of nonideal flow considered in Appendix I.

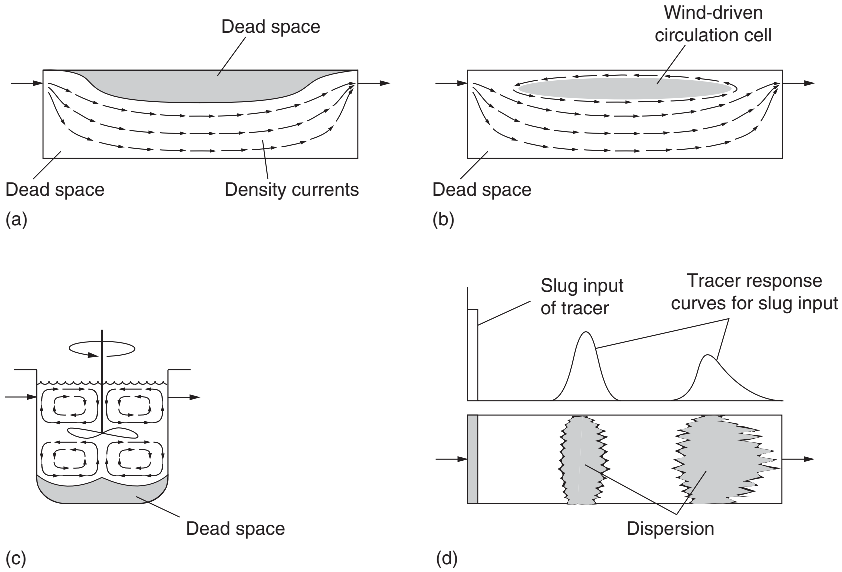
H–1 FACTORS LEADING TO NONIDEAL FLOW IN REACTORS

Nonideal flow occurs when a portion of the flow which enters the reactor during a given time period arrives at the outlet, in less than the theoretical detention time, ahead of the bulk flow which entered the reactor during the same time period. The theoretical detention time, τ, is defined as V/Q, where V is the volume and Q is the flowrate in consistent units. Nonideal flow is often identified as short circuiting. Factors leading to nonideal flow in reactors include:

1. Temperature differences. In complete-mix and plug-flow reactors, nonideal flow (short circuiting) can be caused by density currents due to temperature differences. When the water entering the reactor is colder or warmer than the water in the tank, a portion of the water can travel to the outlet along the bottom of or across the top of the reactor without mixing completely [see Fig. H–1(a)].
2. Wind driven circulation patterns. In shallow reactors, wind circulation patterns can be set up that will transport a portion of the incoming water to the outlet in a fraction of the actual detention time [see Fig. H–1(b)].
3. Inadequate mixing. Without sufficient energy input, portions of the reactor contents may not mix with the incoming water [see Fig. H–1(c)].
4. Poor design. Depending on the design of the inlet and outlet of the reactor relative to the reactor aspect ratio, dead zones may develop within the reactor which will not mix with the incoming water.
5. Axial dispersion in plug-flow reactors [see Fig. H–1(d)]. In plug-flow reactors the forward movement of the tracer is due to advection and dispersion. *Advection* is the term used to describe the movement of dissolved or colloidal material with the current velocity. For example, in a tubular plug-flow reactor (e.g., a pipeline), the early arrival of the tracer at the outlet can be reasoned partially by remembering that the velocity distribution in the pipeline will be parabolic. *Dispersion* is the term used to describe the axial and longitudinal transport of material brought about by velocity differences and dispersion. The distinction between molecular diffusion and dispersion is considered further in Appendix I.

Figure H–1

Definition sketch for short circuiting caused by (a) density currents caused by temperature differences, (b) wind circulation patterns, (c) inadequate mixing, (d) fluid advection and dispersion.



Ultimately, the inefficient use of the reactor volume due to short circuiting resulting from any of the factors described above, can result in reduced treatment performance. The subject of short circuiting in a series of complete-mix reactors was examined extensively in an early paper by MacMullin and Weber (1935); Fitch (1956) and Morrill (1932) examined the effects of short circuiting on the performance of sedimentation tanks.

H–2 TYPES OF TRACERS

Over the years, a number of tracers have been used to evaluate the hydraulic performance of reactors. Important characteristics for a tracer include (adapted in part from Denbigh and Turner, 1984):

1. The tracer should not affect the flow (should have essentially the same density as water when diluted).
2. The tracer must be conservative so that a mass balance can be performed.
3. It must be possible to inject the tracer over a short time period.
4. The tracer should be able to be analyzed conveniently.
5. The molecular diffusivity of the tracer should be low.
6. The tracer should not be adsorbed onto or react with the exposed reactor surfaces.
7. The tracer should not be adsorbed onto or react with the particles in wastewater.

Dyes and chemicals that have been used successfully in tracer studies at wastewater treatment plants are discussed in Sec. 12–6 in Chap. 12. In addition to the tracers considered in Sec. 12-3, lithium chloride is used commonly for the study of natural systems. Sodium chloride, used extensively in the past, has a tendency to form density currents unless mixed completely. Sulfur hexafluoride gas (SF_6) is used most commonly for tracing the movement of groundwater.

H–3 ANALYSIS OF TRACER RESPONSE CURVES

Because of the complexity of the hydraulic response of full-scale reactors, tracer response curves are used to analyze the hydraulics of reactors. Typical examples of tracer response curves are shown on Fig. H–2. Tracer response curves, measured using a short-term and continuous injection of tracer, are known as C (concentration versus time) and F (fraction of tracer remaining in the reactor versus time) curves, respectively. The fraction remaining is based on the volume of water displaced from the reactor by the step input of tracer. The terms used to characterize tracer response curves, the analysis of concentration versus time, and the development of residence time distribution (RTD) curves are described below.

Terms Used to Characterize Tracer Response Curves

Over the years, a number of different symbols and numerical values, as reported in Table H–1, have been used to characterize output tracer curves. The relationship of the terms in Table H–1 and a typical tracer response curve are illustrated on Fig. H–3.

Concentration Versus Time Tracer Response Curves

As noted previously, tracer response curves measured using a short-term or continuous injection of tracer are known as "C" curves (concentration versus time). To characterize C curves, such as those shown on Fig. H–3, the mean value is given by the centroid of the distribution. For C curves, the theoretical mean residence time is determined as follows.

$$\bar{t}_c = \frac{\int_0^\infty tC(t)dt}{\int_0^\infty C(t)dt} \tag{H–1}$$

where $\bar{t}_c$ = mean residence time derived from tracer curve, T

t = time, T

$C(t)$ = tracer concentration at time t, ML^{-3}

The variance, σ_c^2 used to define the spread of the distribution, is defined as

$$\sigma_c^2 = \frac{\int_0^\infty (t - \bar{t})^2 C(t)dt}{\int_0^\infty C(t)dt} = \frac{\int_0^\infty t^2 C(t)dt}{\int_0^\infty C(t)dt} - (\bar{t}_c)^2 \tag{H–2}$$

Figure H–2

Typical tracer response curves: (a) two different types of circular clarifiers (adapted from Dague and Baumann, 1961) and (b) open channel UV disinfection system (courtesy of Andy Salveson).

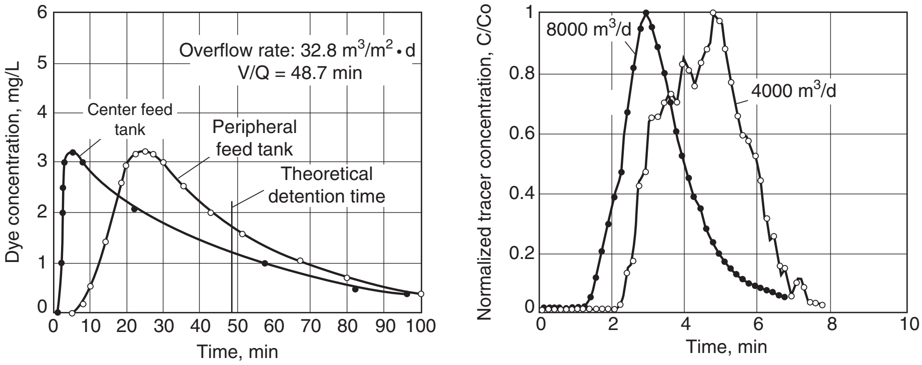


Table H–1

Various terms used to describe the hydraulic performance of reactors used for wastewater treatment[a]

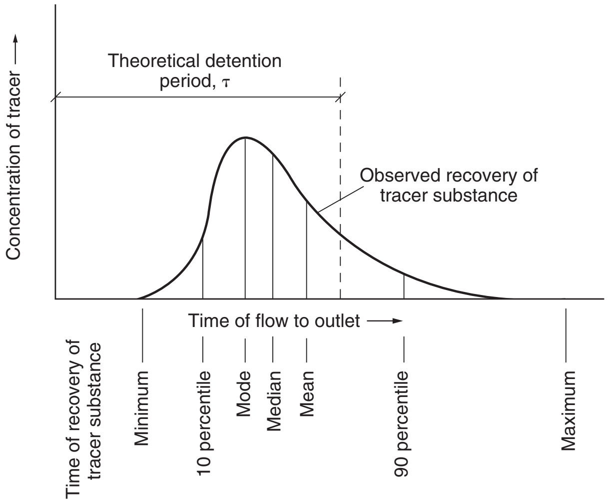
Term	Definition
τ	Theoretical hydraulic residence time (V, volume/Q, flowrate)
t_i	Time at which tracer first appears
t_p	Time at which the peak concentration of the tracer is observed (mode)
t_g	Mean time to reach centroid of the RTD curve
t_{10}, t_{50}, t_{90}	Time at which 10, 50, and 90 percent of the tracer had passed through the reactor
t_{90}/t_{10}	Morrill Dispersion Index, MDI
1/MDI	Volumetric efficiency as defined by Morrill (1932)
t_i/τ	Index of short circuiting. In an ideal plug-flow reactor, the ratio is one, and approaches zero with increased mixing
t_p/τ	Index of modal retention time. Ratio will approach 1 in a plug-flow reactor, and 0 in a complete-mix reactor. For values of the ratio greater than or less than 1.0 the flow distribution in the reactor is not uniform
t_g/τ	Index of average retention time. A value of one would indicate that full use is being made of the volume. A value of the ratio greater than or less than 1.0 indicates the flow distribution is not uniform
t_{50}/τ	Index of mean retention time. The ratio t_{50}/τ, is a measure of the skew of the RTD curve. In an effective plug-flow reactor, the RTD curve is very similar to a normal or Gaussian distribution (U.S. EPA, 1986). A value of t_{50}/τ, less than 1.0 corresponds to an RTD curve that is skewed to the left. Similarly, for values greater than 1.0 the RTD curve is skewed to the right
$t/\tau = \theta$	Normalized time, used in the development of the normalized RTD curve

[a] Adapted,in part, from Morrill (1932) and U.S. EPA (1986).

It will be recognized that the integral term in the denominator in Eqs. (H–1) and (H–2) corresponds to the area under the concentration versus time curve. It will also be recognized that the mean and variance are equal to the *first* and *second moments* of the distribution about the y axis.

Figure H–3

Definition sketch for the parameters used in the analysis of concentration versus time tracer response curves.



If the concentration versus time tracer response curve is defined by a series of discrete time step measurements, the theoretical mean residence time is typically approximated as

$$\bar{t}_{\Delta c} \approx \frac{\sum t_i C_i \Delta t_i}{\sum C_i \Delta t_i} \tag{H–3}$$

where $\bar{t}_{\Delta c}$ = mean detention time based on discrete time step measurements
t_i = time at *i*th measurement, T
C_i = concentration at ith measurement, ML^{-3}
Δt_i = time increment about C_i, T

The variance for a concentration versus time tracer response curve, defined by a series of discrete time step measurements, is defined as

$$\sigma^2_{\Delta c} \approx \frac{\sum t_i^2 C_i \Delta t_i}{\sum C_i \Delta t_i} - (\bar{t}_{\Delta c})^2 \tag{H–4}$$

Where $\sigma^2_{\Delta C}$ = variance based on discrete time measurements, T^2. The application of Eqs. (H–3) and (H–4) is illustrated in Example 12–8 in Chap. 12. Additional details on the analysis of tracer response curves may be found in Levenspiel (1998).

Residence Time Distribution (RTD) Curves

To standardize the analysis of output concentration versus time curves a single reactor for a pulse input of tracer, such as shown on Figs. H–2 and H–3, the output concentration measurements are often normalized by dividing the measured concentration values by an appropriate function such that the area under the normalized curve is equal to one. The normalized curves are known, more formally, as *residence time distribution (RTD) curves* (see Fig. H–4). When a pulse addition of tracer is used, the area under the normalized curve is known as an E curve (also known as the exit age curve). The most important characteristic of an E curve is that the area under the curve is equal to one, as defined by the following integral.

$$\int_0^\infty E(\bar{t}_{\Delta c})\, dt = 1 \tag{H–5}$$

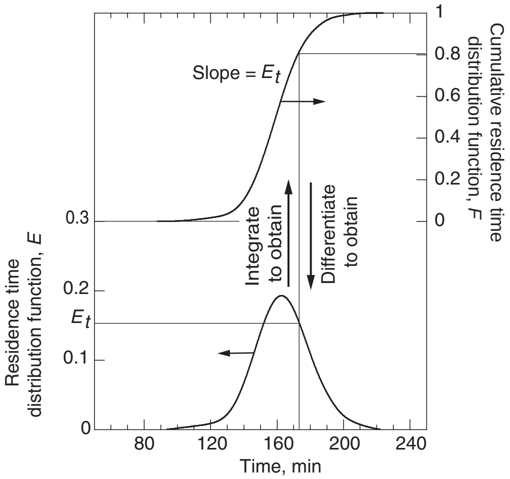


Figure H–4

Normalized residence time distribution curves. The curve on the bottom is known as the exit age curve, identified as the "*E* curve." The curve on top is the cumulative residence time curve, identified as the "*F* curve."

where $E(t)$ is the residence time distribution function. The $E(t)$ value is related to the $C(t)$ value as follows:

$$E(t) = \frac{C(t)}{\displaystyle\int_0^\infty C(t)dt} \tag{H–6}$$

As in Eqs. (H–1) and (H–2), the integral term in the denominator in Eqs. (H–6) corresponds to the area under the concentration versus time curve. Applying Eq. (H–6) to the expression, obtained previously, for the complete-mix reactor, the exit age curve $E(t)$ for a complete-mix reactor is obtained as follows:

$$E(t) = \frac{C(t)}{\displaystyle\int_0^\infty C(t)\,dt} = \frac{C_o e^{-t/\tau}}{\displaystyle\int_0^\infty C_o e^{-t/\tau}\,dt} = \frac{e^{-t/\tau}}{\tau} \tag{H–7}$$

and the corresponding value based on normalized time, $\theta = t/\tau$, is

$$E(\theta) = \tau E(t) = e^{-\theta} \tag{H–8}$$

The mean residence time for the $E(t)$ curve, given by Eq. (H–8), can be derived by applying Eq. (H–23); the resulting expression is given by

$$t_m = \frac{\displaystyle\int_0^\infty tE(t)dt}{\displaystyle\int_0^\infty E(t)dt} = \int_0^\infty tE(t)dt \tag{H–9}$$

In a similar manner when a step input is used, the normalized concentration curve is known as the cumulative residence time distribution curve and is designated as the F curve. The F curve is defined as

$$F(t) = \int_0^t E(t)dt = 1 - e^{-t/\tau} \tag{H–10}$$

where $F(t)$ is the cumulative residence time distribution function. As shown on Fig. H–4, the $F(t)$ curve is the integral of the $E(t)$ curve while the $E(t)$ curve is the derivative of the $F(t)$ curve. In effect, $F(t)$ represents the amount of tracer that has been in the reactor for less than the time t. The development of E and F RTD curves is illustrated in Example H–1. Additional details on the analysis of E and F curves may be found in Denbigh and Turner (1984), Fogler (2005), Levenspiel (1998), and in the chemical engineering literature.

EXAMPLE H–1 **Development of Residence Time Distribution (RTD) Curves from Concentration versus Time Tracer Curves** Use the concentration versus time tracer data given in Example 12–8 in Chap. 12 to develop E and F residence time distribution (RTD) curves for the chlorine contact basin. Using the E curve, compute the residence time and compare to the value obtained in Example 12–8. Plot the resulting E and F curves.

Solution

1. Using the values for time and concentration from Example 12–8, set up a computation table to calculate $E(t)$ values. The computation table is shown below. The $E(t)$ values are calculated by finding the sum of the $C \times \Delta t$ values, which corresponds to

the area under the C (concentration) curve, and dividing the original concentrations by the sum as illustrated below.

Area under C curve $= \Sigma C\Delta t$

$$E(t) = \frac{C}{\Sigma C\Delta t}$$

a. For each time interval, multiply the concentration by the time step ($t = 8$ min) and obtain the sum of the multiplied values. As shown in the computation table, the sum (which is the approximate area under the curve) is 817.778 μg/L-min.

b. Calculate the $E(t)$ values by dividing the original concentration values by the area under the curve.

c. Confirm that the $E(t)$ values are correct by multiplying each one by the time step, and calculating the sum. According to Eq. (H–6) written in summation form, the sum should be 1.00.

$$\Sigma E(t)\,\Delta t = \Sigma\left(\frac{C\Delta t}{\Sigma C\Delta t}\right) = \Sigma\left(\frac{C}{\Sigma C}\right) = 1$$

Time, t, min	Conc., C, μg/L	$C \times \Delta t$ μg/L·min	$E(t)$, min^{-1}	$E(t) \times \Delta t$, unitless	$t \times E(t) \times \Delta t$, min
88	0.000	0.000	0.00000	0.00000	0.000
96	0.056	0.445	0.00007	0.00054	0.077
104	0.333	2.666	0.00041	0.00326	0.333
112	0.556	4.445	0.00068	0.00544	0.627
120	0.833	6.666	0.00102	0.00815	0.960
128	1.278	10.222	0.00156	0.01250	1.638
136	3.722	29.778	0.00455	0.03641	5.005
144	9.333	74.666	0.01141	0.09130	13.133
152	16.167	129.336	0.01977	0.15816	24.077
160	20.778	166.224	0.02541	0.20326	32.512
168	19.944	159.552	0.02439	0.19510	32.794
176	14.111	112.888	0.01726	0.13804	24.358
184	8.056	64.445	0.00985	0.07880	14.573
192	4.333	34.666	0.00530	0.04239	8.141
200	1.556	12.445	0.00190	0.01522	3.040
208	0.889	7.111	0.00109	0.00870	1.830
216	0.278	2.222	0.00034	0.00272	0.518
224	0.000	0.000	0.00000	0.00000	0.000
Total	102.222	817.778	-	1.0000	163.616

[a] ppb = parts per billion

2. Determine the mean residence time using the following summation form of Eq. (H–28).

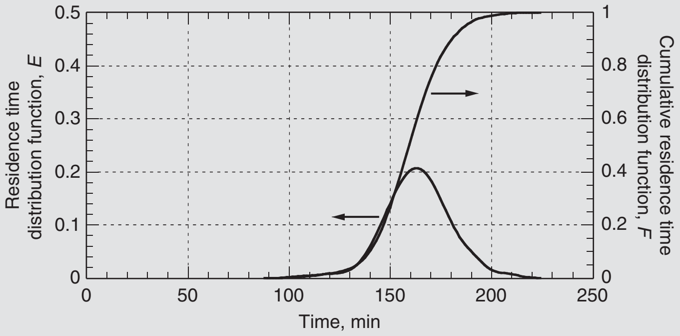
$$\bar{t} = \Sigma(t)E(t)\,\Delta t$$

The required computation is presented in the final column of the computation table given above. As shown, the computed mean residence time (163.6 min) is essentially the same as the value (163.4 min) determined in Example 12–8 in Chap.12.

3. Develop the F RTD curve. The values for plotting the F curve are obtained by summing cumulatively the $E(t)\Delta t$ values to obtain the coordinates of the F curve.

Time t, min	$E(t)$, min^{-1}	$E(t) \times \Delta t$, unitless	Cumulative total, F, unitless
88	0.00000	0.00000	0.0000
96	0.00007	0.00054	0.0005
104	0.00041	0.00326	0.0038
112	0.00068	0.00544	0.0092
120	0.00102	0.00815	0.0174
128	0.00156	0.01250	0.0299
136	0.00455	0.03641	0.0663
144	0.01141	0.09130	0.1576
152	0.01977	0.15816	0.3158
160	0.02541	0.20326	0.5191
168	0.02439	0.19510	0.7142
176	0.01726	0.13804	0.8522
184	0.00985	0.07880	0.931
192	0.00530	0.04239	0.9734
200	0.00190	0.01522	0.9886
208	0.00109	0.00870	0.9973
216	0.00034	0.00272	1.0000
224	0.00000	0.00000	1.0000
Total		1.0000	

4. Plot the resulting E and F curves. The plot of the E (column 3) and F (column 4) curves using the values determined above is shown below:



Comment The use of normalized RTD curves to obtain coefficients of dispersion is included in the discussion of the Hydraulic Characteristics of Nonideal Reactors. Additional details may be found in Denbigh and Turner (1984) and Levenspiel (1998).

Practical Interpretation of Tracer Measurements

In 1932, based on his studies of sedimentation basins, Morrill (1932) suggested that the ratio of the 90 percentile to the 10 percentile value from the cumulative tracer curve could be used as a measure of the dispersion index, and that the inverse of the dispersion index is a measure of the volumetric efficiency (Morrill, 1932). The dispersion index as proposed by Morrill is given by

$$\text{Morrill Dispersion Index, MDI} = \frac{P_{90}}{P_{10}} \tag{H–11}$$

where P_{90} = 90 percentile value from log-probability plot

P_{10} = 10 percentile value from log-probability plot

The percentile values are obtained from a log-probability plot of the time (log scale) versus the cumulative percentage of the total tracer which has passed out of the basin (on probability scale). The value of the MDI for an ideal plug-flow reactor is 1.0 and about 22 for a complete-mix reactor. A plug-flow reactor with an MDI value of 2.0 or less is considered by the U.S. EPA to be an effective plug-flow reactor (U.S. EPA, 1986). The volumetric efficiency is given by the following expression:

$$\text{Volumetric efficiency, \%} = \frac{1}{\text{MDI}} \times 100 \tag{H–12}$$

The determination of the Morrill dispersion index and the volumetric efficiency for the analysis of the flow pattern in a chlorine contact basin is illustrated in Example 12–8 in Chap. 12.

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Modeling Nonideal Flow in Reactors

Appendix I

The hydraulic characteristics of nonideal reactors can be modeled by taking dispersion into consideration. For example, if dispersion becomes infinite, the plug-flow reactor with axial dispersion is equivalent to a complete-mix reactor. Both the plug-flow reactor with axial dispersion and complete-mix reactors in series are considered in the following discussion. However, before considering nonideal flow in reactors it will be helpful to examine the distinction between the diffusion and dispersion as applied to the analysis of reactors used for wastewater treatment.

I–1 THE DISTINCTION BETWEEN MOLECULAR DIFFUSION AND DISPERSION

In addition to short-circuiting caused by the use of improper design of reactor inlets and outlets, inadequate mixing, thermal and density currents, and diffusion and dispersion can also result in nonideal flow. The distinction between the diffusion and dispersion is as follows.

DIFFUSION

Under quiescent flow conditions (i.e., no flow), the mass transfer of material is brought about by *molecular diffusion*, in which dissolved constituents and/or small particles move randomly. This random motion is known as *Brownian Motion*. Further, it should be noted that molecular diffusion can occur under either laminar of turbulent flow conditions, as it does not depend on the bulk movement of a liquid. The transfer of mass by molecular diffusion in stationary systems can be represented by the following expression, known as Fick's first law:

$$r = -D_m \frac{\partial C}{\partial x} \tag{I–1}$$

where r = rate of mass transfer per unit area per unit time, $ML^{-2}T^{-1}$

D_m = coefficient of molecular diffusion in the x direction, L^2T^{-1}

C = concentration of constituent being transferred, ML^{-3}

In the chemical engineering literature the symbol "J" is used to denote mass transfer in concentration units whereas the symbol "N" is used to denote the transfer of mass expressed as moles. The negative sign in Eq. (I–1) is used to denote the fact that diffusion takes place in the direction of decreasing concentration (Shaw, 1966). Adolf Fick (1829–1901), a physician and physiologist, derived the first, and second, laws of diffusion in the 1850s by direct analogy to the equations used to describe the conduction of heat in solids as proposed by Fourier (Crank, 1957). Determination of numerical values for the coefficient of molecular diffusion is illustrated in Sec. 1–10 in Chap. 1.

DISPERSION

The transfer of a constituent from a higher concentration to a lower concentration (e.g., blending) brought about by eddies formed by turbulent flow or by the shearing forces between fluid layers is termed *dispersion.* Under this definition eddies can vary in size from microscale to macroscale to large circulation patterns in the oceans. While microscale transport can only be brought about by molecular diffusion, macroscale transport is brought about by both molecular diffusion and dispersion (Crittenden et al., 2012). Under turbulent flow conditions, the longitudinal spreading of a tracer is caused by dispersion, in which case the coefficient of molecular diffusion term, D_m, in Eq. (I–1) is replaced by the "coefficient of dispersion", D.

While the magnitude of the molecular diffusion depends primarily on the chemical and fluid properties, turbulent or eddy diffusion and dispersion depend primarily on the flow regime. Typical observed ranges for the coefficient of molecular diffusion and dispersion are reported in Table I–1. In all cases, it is important to remember that regardless of whether the coefficient of molecular diffusion or the coefficient of turbulent or eddy diffusion, or dispersion is operative, the driving force for mass transfer is the concentration gradient.

I–2 PLUG-FLOW REACTOR WITH AXIAL DISPERSION

In the following analysis only the one-dimensional problem is considered. However, it should be noted that all dispersion problems are three dimensional, with the dispersion coefficient varying with direction and the degree of turbulence. Using the relationship given above [Eq. (I–1)] and referring to Fig. I–1(c), the one-dimensional materials mass balance for the transport of a conservative dye tracer by advection and dispersion is:

Accumulation = inflow − outflow

$$\frac{\partial C}{\partial t}A\Delta x = \left(\upsilon AC - AD\frac{\Delta C}{\Delta x}\right)\bigg|_{x} - \left(\upsilon AC - AD\frac{\Delta C}{\Delta x}\right)\bigg|_{x+\Delta x} \tag{I–3}$$

where $\partial C/\partial t$ = change in concentration with time, $ML^{-3}T^{-1}$, (g/m$^3\cdot$s)

A = cross-sectional area in x direction, L^2, (m^2)

Δx = differential distance, L, m

C = constituent concentration, ML^{-3}, (g/m^3)

D = coefficient of axial dispersion, L^2T^{-1}, (m^2/s)

υ = average velocity in x direction, LT^{-1}, (m/s)

In Eq. (I–3) the term υAC represents the transport of mass due to advection and the term $AD(\Delta C/\Delta x)$ represents the transport brought about by dispersion. Taking the limit of Eq. (I–3) as Δx approaches zero results in the following two expressions:

$$\frac{\partial C}{\partial t} = -D\frac{\partial^2 C}{\partial x^2} - \upsilon\frac{\partial C}{\partial x} \tag{I–4}$$

Table I–1

Typical range of values for molecular diffusion and dispersion[a]

Coefficient	Symbol	Range of values, cm²/s
Molecular diffusion	D_m	$10^{-10} - 10^{-7}$
Dispersion	D	$10^{-3} - 10^{0}$

[a] Adapted from Schnoor (1996), Shaw (1966), and Thibodeaux et al. (2012).

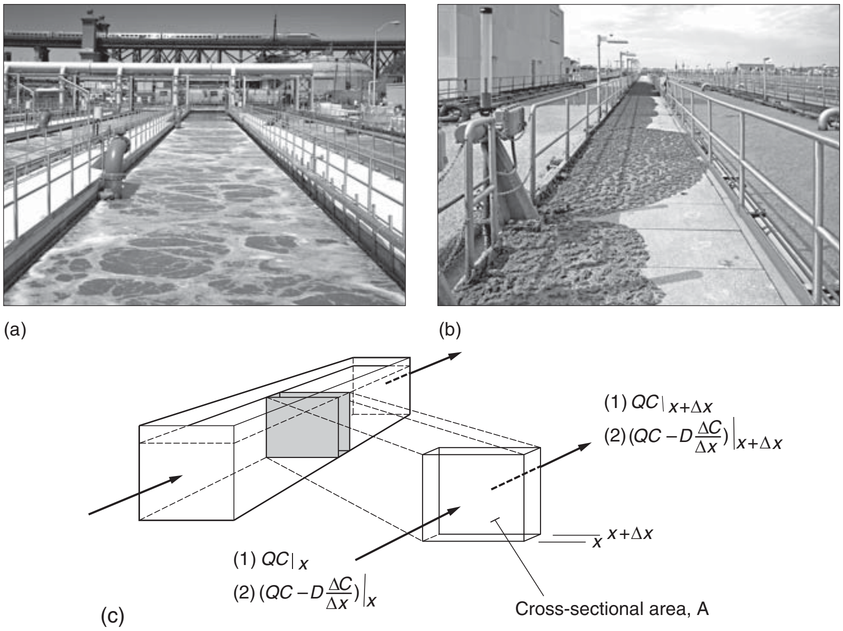


Figure I–1

Views of plug-flow reactors and definition sketch: (a) and (b) views of plug-flow activated sludge process reactors and (c) definition sketch for the hydraulic analysis of a plug-flow reactor with (1) advection only and (2) with advection and axial dispersion.

$$\frac{\partial C}{\partial t} = -D\frac{\partial^2 C}{\partial x^2} - \frac{\partial C}{\partial \tau} \tag{I–5}$$

In Eq. (I–5) the hydraulic detention time $\partial\tau$ has been substituted for the term $\partial x/\upsilon$. Equation (I–5) has been solved for small amounts of axial dispersion (see below). The solution for a unit pulse input leading to symmetrical output tracer response curves for small amounts of axial dispersion is given by Levenspiel (1998):

$$C_\theta = \frac{1}{2\sqrt{\pi(D/\upsilon L)}} \exp\left[-\frac{(1 - \theta)^2}{4(D/\upsilon L)}\right] \tag{I–6}$$

where C_θ = normalized tracer response, C/C_o, unitless
θ = normalized time, t/τ, unitless
t = time, T, (s)
τ = theoretical detention time, V/Q, T, (s)
D = coefficient of axial dispersion, L^2T^{-1}, (m^2/s)
υ = fluid velocity, LT^{-1}, (m/s)
L = characteristic length, L, (m)

The solution given by Eq. (I–6) is for what is known as a *closed system* in which it is assumed that there is no dispersion upstream or downstream of the boundaries of the reactor (e.g., a reactor with inlet and outlet weirs). A large reactor, such as a rectangular sedimentation basin, fed by a small diameter pipe, is an example of a closed system. If a tracer were added to the flow in the small diameter pipe, the tracer would be transported by advection with little or no dispersion. The same situation exists in the discharge pipe from the reactor. Nevertheless, for small amounts of dispersion, Eq. (I–6) can be used to approximate the performance of an open or closed reactor, regardless of the boundary conditions.

It will be noted that Eq. (I–6) has the same general form as the equation for the normal probability distribution. Thus, the corresponding mean and variance are

$$\bar{\theta} = \frac{\bar{t}_c}{\tau} = 1 \tag{I–7}$$

$$\sigma_\theta^2 = \frac{\sigma_c^2}{\tau^2} = 2\frac{D}{\upsilon L} \tag{I–8}$$

where $\bar{\theta}$ = normalized mean detention time, $\bar{t}_c/\tau$, unitless
$\bar{t}_c$ = mean detention (or residence) time derived from C curve [see Eq. (H–1), Appendix H], T, (s)
τ = theoretical detention (or residence) time, T, (s)
σ_θ^2 = variance of normalized tracer response C curve, T^2, (s^2)
σ_c^2 = variance derived from C curve [see Eq. (H–2), Appendix H], T^2, (s^2)

Defining the exact extent of axial dispersion is difficult. To provide an estimate of dispersion, the following unitless dispersion number has been defined:

$$d = \frac{D}{\upsilon L} = \frac{Dt}{L^2} \tag{I–9}$$

where d = dispersion number, unitless
D = coefficient of axial dispersion, L^2T^{-1}, (m^2/s)
υ = fluid velocity, LT^{-1}, (m/s)
L = characteristic length, L, (m)
t = travel time (L/υ), T, (s)

Normalized effluent concentration versus time curves, obtained using Eq. (I–6), for a plug-flow reactor with limited axial dispersion for various values of the dispersion number are shown on Fig. I–2.

When dispersion is large, the output curve becomes increasingly nonsymmetrical, and the problem becomes sensitive to the boundary conditions. In environmental problems, a wide variety of entrance and exit conditions are encountered, but most can be considered approximately open; that is, the flow characteristics do not change greatly as the boundaries are crossed. The solution to Eq. (I–5) for a unit pulse input in an *open system* with larger amounts of dispersion is as follows (Fogler, 2005; Levenspiel, 1998):

$$C_\theta = \frac{1}{2\sqrt{\pi\theta(D/\upsilon L)}} \exp\left[-\frac{(1-\theta)^2}{4\theta\,(D/\upsilon L)}\right] \tag{I–10}$$

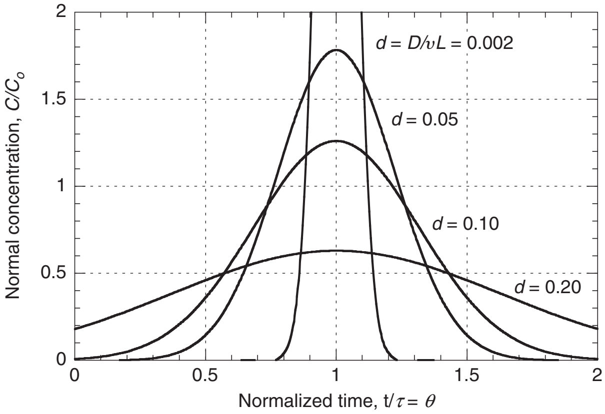


Figure I–2

Typical concentration versus time tracer response curves for a plug flow reactor with small amounts of axial dispersion subject to a pulse (slug) input of tracer.

The corresponding mean and variance are:

$$\bar{\theta} = \frac{\bar{t}_c}{\tau} = 1 + 2\frac{D}{\upsilon L} \tag{I–11}$$

$$\sigma_\theta^2 = \frac{\sigma_c^2}{\tau^2} = 2\frac{D}{\upsilon L} + 8\left(\frac{D}{\upsilon L}\right)^2 \tag{I–12}$$

where the terms are as defined previously for Eqs. (I–7), (I–8) and (I–9).

The mean as given by Eq. (I–11) is greater than the hydraulic detention time because of the forward movement of the tracer due to dispersion. Effluent concentration versus time curves, obtained using Eq. (I–12), for a plug-flow reactor with significant axial dispersion for various dispersion factors are shown on Fig. I–3. A more detailed discussion of closed and open reactors may be found in Fogler (2005) and Levenspiel (1998).

In the literature, the inverse of Eq (I–9), as given below, is often identified as the Peclet number of longitudinal dispersion (Kramer and Westererp, 1963).

$$P_e = \frac{\upsilon L}{D} = \frac{1}{d} \tag{I–13}$$

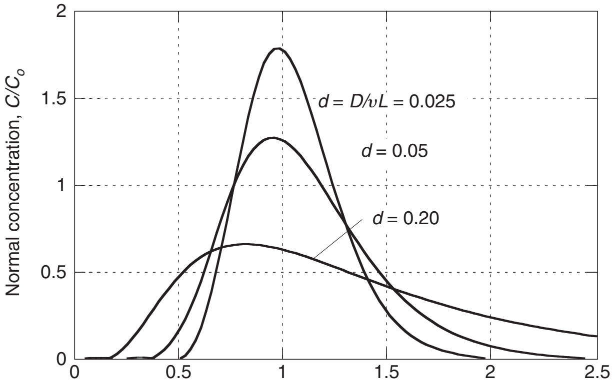
In effect, the Peclet number represents the ratio of the mass transport brought about by advection and dispersion. If the Peclet number is significantly greater than one, advection is the dominant factor in mass transport. If the Peclet number is significantly less than one, dispersion is the dominant factor in mass transport. Although beyond the scope of this presentation, it can also be shown that the number of complete-mix reactors in series required to simulate a plug-flow reactor with axial dispersion is approximately equal to the Peclet number divided by 2. Thus, for a dispersion factor of 0.025, the Peclet number is equal to 40 and the corresponding number of reactors in series needed to simulate the dispersion in a plug-flow reactor is equal to 20. This relationship will be illustrated in the following discussion dealing with complete-mix reactors in series. The Peclet number is also used to define transverse diffusion in packed bed plug-flow reactors (Denbigh and Turner, 1984).

For practical purposes, the following dispersion values can be used to assess the degree of axial dispersion in wastewater treatment facilities.

No dispersion	$d = 0$ (ideal plug-flow)
Low dispersion	$d = <0.05$
Moderate dispersion	$d = 0.05$ to 0.25
High dispersion	$d = >0.25$
	$d \to \infty$ (complete-mix)

Figure I–3

Typical concentration versus time tracer response curves for a plug flow reactor with large amounts of axial dispersion subject to a pulse (slug) input of tracer.



Typical dispersion numbers determined for actual treatment facilities are given in Table I–2. The considerable range in the dispersion numbers reported in Table I–2 for individual treatment processes is due, most often, to one or more of the following factors (Arceivala and Asolekar, 2007):

1. The scale of the mixing phenomenon
2. Geometry (i.e., aspect ratio) of the unit
3. Power input per unit volume (i.e., mechanical and pneumatic)
4. Type and disposition of the inlets and outlets of the treatment units
5. Inflow velocity and its fluctuations
6. Density and temperature differences between the inflow and the contents of the reactor.

Because of the wide range of dispersion numbers that can result for individual treatment processes, special attention must be devoted to the factors cited above in the design of treatment facilities. Determination of the dispersion number and the coefficient of dispersion using the tracer response curves is illustrated in Example I–1.

Evaluation of the coefficient of axial dispersion D for existing facilities is done experimentally using the results of tracer tests, as discussed previously. Because the systems encountered in wastewater treatment are large, experimental work is often difficult and expensive, and is, unfortunately, usually after the fact. To take into account axial dispersion in the design of treatment facilities both scaled models and empirical relationships have been developed for a variety of treatment units including oxidation ponds (Polprasert and Bhattasrai, 1992) and chlorine contact basins (Crittenden et al., 2012). An approximate value of D for water for large Reynolds numbers is (Davies, 1972)

$$D = 1.01\nu \mathrm{N_R}^{0.875} \tag{I–14}$$

where D = coefficient of dispersion, L^2T^{-1}, (m^2/s)
ν = kinematic viscosity, L^2T^{-1}, m^2/s (see Appendix C)
N_R = Reynolds number, unitless
$= 4\upsilon R/\nu$
υ = velocity in open channel, LT^{-1}, (m/s)
R = hydraulic radius = area/wetted perimeter, L, (m)

Table I–2

Typical dispersion numbers for various wastewater treatment facilities [a]

Treatment facility	Range of values for dispersion number
Rectangular sedimentation tanks	0.2–2.0
Activated sludge aeration reactors	
Long plug-flow	0.1–1.0
Complete-mix	3.0–4.0+
Oxidation ditch activated sludge process	3.0–4.0+
Waste stabilization ponds	
Single ponds	1.0–4.0+
Multiple cells in series	0.1–1.0
Mechanically aerated lagoons	
Long rectangular shaped	1.0–4.0+
Square shaped	3.0–4.0+
Chlorine contact basins	0.02–0.004

[a] Adapted from Arceivala and Asolekar (2007).

Values for N_R found in open channel flow in wastewater treatment plants are typically in the range from 10^3 to 10^4. The corresponding values for D range from 0.0004 to 0.003 m^2/s (4 to 30 cm^2/s).

EXAMPLE I–1 **Determination of the Dispersion Number and the Coefficient of Dispersion from Concentration versus Time Tracer Response Curves** Use the concentration versus time tracer data given in Example 12-8 in Chap. 12 to determine the dispersion number and the coefficient of dispersion for the chlorine contact basin described in Example 12–8. Compare the value of the coefficient of dispersion computed, using the tracer data, to the value computed using Eq. (I–14). Assume the following data are applicable:

1. The flowrate at the time when the tracer test was conducted = 240,000 m^3/d
2. Number of chlorine contact basin = 4
3. Number of channels per contact basins = 13
4. Channel dimensions
   a. Length = 36.6 m
   b. Width = 3.0 m
   c. Depth = 4.9 m
5. Temperature = 20°C

Solution

1. From Example 12-8, the mean value and variance for the tracer response C curve are:
   a. Mean, $\tau_{\Delta c} = 2.7$ h
   b. Variance, $\sigma^2_{\Delta c} = 280.5\ \text{min}^2$
2. Determine the theoretical detention time for the chlorine contact basin using the given data.

$$\tau = \frac{4 \times 13 \times (36.6\ \text{m} \times 3.0\ \text{m} \times 4.9\ \text{m})}{(240{,}000\ \text{m}^3/\text{d})}$$

3. Determine the normalized mean detention time using Eq. (I–7). Use the approximate value of $\bar{t}_{\Delta c}$ for $\bar{t}_c$.

$$\bar{\theta}_{\Delta c} = \frac{\bar{t}_{\Delta c}}{\tau} \approx \frac{2.7}{2.8} \approx 0.96 \approx 1.0$$

4. Determine the dispersion number using Eq. (I–8). Use the approximate value of $\sigma^2_{\Delta c}$ for σ^2_c.

$$\sigma^2_{\Delta c} = \frac{\sigma^2_{\Delta c}}{\tau^2} \approx 2\frac{D}{\upsilon L} = 2\text{d}$$

$$d \approx \frac{1}{2}\frac{\sigma^2_{\Delta c}}{\tau^2} \approx \frac{1}{2}\frac{280.5\ \text{min}^2}{(167.9\ \text{min})^2} = 0.00498$$

5. Using Eq. (I–9), determine the coefficient of dispersion.

$$D = d \times \upsilon \times L$$

$$\upsilon = \frac{Q}{A} = \frac{(240{,}000\ \text{m}^3/\text{d})}{(4 \times 13 \times 3.0\ \text{m} \ \times 4.9\ \text{m})} = 314.0\ \text{m/d} \ = 0.00363\ \text{m/s}$$

$$D = (0.00498)(0.00363\ \text{m/s})(36.6\ \text{m}) = 6.62 \times 10^{-4}\ \text{m}^2/\text{s}$$

6. Compare the value of the coefficient of dispersion computed in Step 5 with the value computed using Eq. (I–14).

   $$D = 1.01\nu N_R^{0.875}$$

   a. Compute the Reynolds number

   $$N_R = 4vR/\nu$$
   $$\nu = 1.002 \times 10^{-6}\ \text{m}^2/\text{s}$$
   $$N_R = \frac{(4)(0.00363\ \text{m/s})\ [(3.0\ \text{m} \times 4.9\ \text{m})/(2 \times 4.9\ \text{m} + 3.0\ \text{m})]}{(1.002 \times 10^{-6}\ \text{m}^2/\text{s})} = 1664$$

   b. Compute the coefficient of dispersion

   $$D = (1.01)(1.002 \times 10^{-6})\,(1664)^{0.875} = 6.66 \times 10^{-4}\ \text{m}^2/\text{s}$$

Comment Based on the computed value of the dispersion number (0.00498), the chlorine contact basin would be classified as having low dispersion (i.e., $d = < 0.05$). The coefficient of dispersion determined using the results of the tracer study and Eq. (I–14) are remarkably close, given the nature of such measurements.

I–3 COMPLETE-MIX REACTORS IN SERIES

When varying amounts of axial dispersion are encountered, the flow is sometimes identified as *arbitrary flow*. The output from a plug-flow reactor with axial dispersion (arbitrary flow) is often modeled using a number of complete-mix reactors in series, as outlined below. In some situations, the use of a series of complete-mix reactors may have certain advantages with respect to treatment. To understand the hydraulic characteristics of reactors in series (see Fig. I–4), assume that a pulse input (i.e., a slug) of tracer is injected into the first reactor in a series of equally sized reactors so that the resulting instantaneous concentration of tracer in the first reactor is C_o. The total volume of all the reactors is V, the volume of an individual reactor is V_i, and the hydraulic residence time V_i/Q is τ_i. The effluent concentration from the first reactor as is given by the following equation [see Eq. (1–13) in Chap. 1].

$$C_1 = C_o e^{-t/\tau_i} \tag{I–15}$$

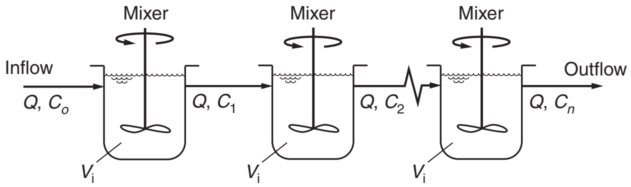
Writing a materials balance for the second reactor results in the following:

Accumulation = inflow − outflow

$$\frac{dC_2}{dt} = \frac{1}{\tau_i}C_1 - \frac{1}{\tau_i}C_2 \tag{I–16}$$

Figure I–4

Definition sketch for the analysis of complete-reactors in series.



Rearranging the terms in Eq. (I–16) and substituting Eq. (I–15) for C_1 results in

$$\frac{dC_2}{dt} + \frac{C_2}{\tau_i} = \frac{C_o}{\tau_i} e^{-t/\tau_i} \tag{I–17}$$

The general non-steady-state solution for Eq. (I–17) is obtained by first noting that Eq. (I–17) has the form of the standard first-order linear differential equation. The solution procedure outlined in Eqs. (I–18) through (I–23) involves the use of an integrating factor. It should be noted that Eq. (I–17) can also be solved using the separation of variables method. The solution to Eq. (I–17) is included here because these types of equations are encountered frequently in the field of environmental engineering and in this text. The first step in the solution is to rewrite Eq. (I–17) in the form

$$C_2' + \frac{C_2}{\tau_i} = \frac{C_o}{\tau_i} e^{-t/\tau_i} \tag{I–18}$$

where C_2' is used to denote the derivative dC_2/dt. In the next step, both sides of the expression are multiplied by the integrating factor $e^{\beta t}$, where $\beta = (1/\tau_i)$.

$$e^{\beta t}(C_2' + \beta C_2) = e^{\beta t}(\beta C_o e^{-\beta t}) = \beta C_o \tag{I–19}$$

The left hand side of the above expression can be written as a differential as follows:

$$(C_2 e^{\beta t})' = \beta C_o \tag{I–20}$$

The differential sign is removed by integrating the above expression

$$C_2 e^{\beta t} = \beta C_o \int dt \tag{I–21}$$

Integration of Eq. (I–21) yields

$$C_2 e^{\beta t} = \beta C_o t + \mathrm{K}\ \text{(constant of integration)} \tag{I–22}$$

Dividing by $e^{\beta t}$ yields

$$C_2 = \beta \mathrm{C}_o t e^{-\beta t} + \mathrm{K} e^{-\beta \mathrm{t}} \tag{I–23}$$

But when $t = 0$, $C_2 = 0$ and K is equal to 0. Thus,

$$C_2 = C_o \frac{t}{\tau_i} e^{-t/\tau_i} \tag{I–24}$$

Following the same solution procedure, the generalized expression for the effluent concentration for the ith reactor is

$$C_i = \frac{C_o}{(i-1)!} \left(\frac{t}{\tau_i}\right)^{i-1} e^{-t/\tau_i} \tag{I–25}$$

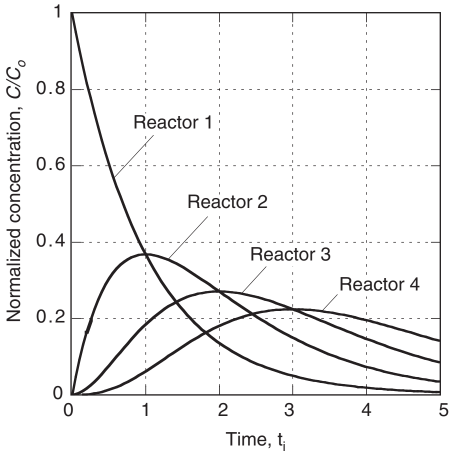
The effluent concentration from each of four complete-mix reactors in series is shown on Fig. I–5.

The corresponding expression based on the overall hydraulic residence time τ, where τ is equal to $n\tau_i$ is

$$C_i = \frac{C_o}{(i-1)!} \left(\frac{nt}{\tau}\right)^{i-1} e^{-nt/\tau} \tag{I–26}$$

Figure I–5

Effluent concentration curves for each of four complete-mix reactors in series subject to a slug input of tracer into the first reactor of the series.



Fraction Remaining

Equation (I–25) can also be used to obtain the fraction of tracer remaining in a series of complete-mix reactors at any time *t*. The fraction of tracer remaining, *F*, at time *t*, is equal to

$$F = \frac{C_1 + C_2 + \cdots + C_n}{C_o} \tag{I–27}$$

Using Eq. (I–25) to obtain the individual effluent concentrations, the fraction remaining in four equal sized reactors in series is given by

$$F_{4R} = \frac{C_o e^{-4t/\tau} + C_o(4t/\tau)\,\mathrm{e}^{-4t/\tau} + (C_o/2)\,(4t/\tau)^2 e^{-4t/\tau} + (C_o/6)\,(4t/\tau)^3 e^{-4t/\tau}}{C_o}$$

$$F_{4R} = \left[1 + (4t/\tau) + \frac{(4t/\tau)^2}{2} + \frac{(4t/\tau)^3}{6}\right] e^{-4t/\tau} \tag{I–28}$$

The fraction of a tracer remaining in a series of 2, 4, 6 and 75 complete-mix reactors in series is given on Fig. I–6.

Comparison of Nonideal Plug-Flow Reactor and Complete-Mix Reactors in Series

In many cases it will be useful to model the performance of plug-flow reactors with axial dispersion, as discussed previously, with a series of complete-mix reactors in series. To obtain

Figure I–6

Fraction of tracer remaining in a system comprised of reactors in series.

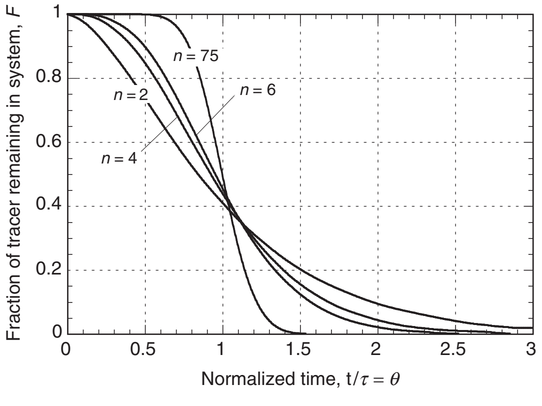
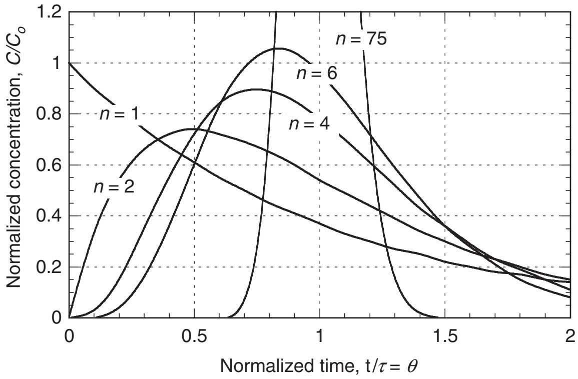


Figure I–7

Effluent tracer concentration curves for reactors in series, subject to a slug input of tracer into the first reactor of the series. Concentration values greater than one occur because the same amount of tracer is placed in the first reactor in each series of reactors.



the normalized residence time distribution (RTD) curve for n reactors in series, Eq. (I–26) can be written as follows by noting that the total volume is nV_i and $\tau = n\tau_i$:

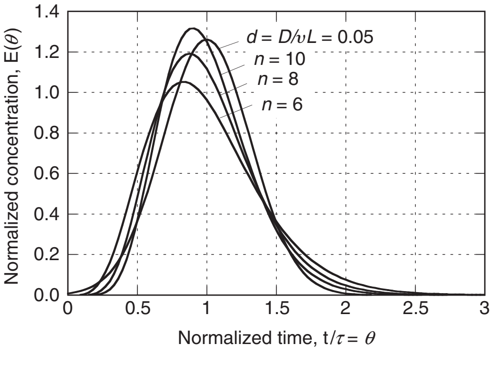
$$E(\theta) = \frac{C_n}{C_o/n} = \frac{n}{(n-1)!}(n\theta)^{n-1}e^{-n\theta} \tag{I–29}$$

In effect, in Eq. (I–29) it is assumed that the same amount of tracer is always added to the first reactor in series. Effluent residence time distribution curves, obtained using Eq. (I–57), for 1, 2, 4, 6, and 75 reactors in series are shown on Fig. I–7. It is interesting to note that a model comprised of four complete-mix reactors in series is often used to describe the hydraulic characteristics of constructed wetlands. As shown on Fig. I–7, the concentration increases as the number of reactors in series increases because the same amount of tracer is used regardless of the number of reactors in series. It should also be noted that the F curves shown on Fig. I–6 can also be obtained by integrating the E curves given on Fig. I–7.

A comparison of the residence time distribution curves obtained for a plug-flow reactor, with a dispersion number of 0.05, and to the residence time distribution curves obtained for six, eight and ten complete-mix reactors in series is shown on Fig. I–8. As shown, all three of the complete-mix reactors in series can be used, for practical purposes, to simulate a plug-flow reactor with a dispersion factor of 0.05. As noted previously, the number of reactors in series needed to simulate a plug-flow reactor with dispersion is approximately equal to the Peclet number divided by 2. Thus, for a dispersion factor of 0.05, the Peclet number is equal to 20 and the corresponding number of reactors in series needed to simulate the dispersion in a plug-flow reactor is equal to 10. As shown on Fig. I–8, the response curves computed using ten reactors in series and Eq. (I–10) with a dispersion number, $d = 0.05$ are essentially the same.

Figure I–8

Comparison of effluent response curves for a plug-flow reactor with a dispersion factor of 0.05 and reactor systems comprised of six, eight, and ten reactors in series subject to a pulse input of tracer.



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Name Index

Subject Index

Because a number of the subjects covered in this text can be referenced (i.e., indexed) under different alphabetical listings, it has been necessary to develop an approach to limit the degree of duplication, yet not affect the utility of the index. The approach used is as follows. Each subject with multiple subentries is indexed in detail under one letter of the alphabet. Where the same subject is indexed under another letter of the alphabet, inclusive page numbers are given and a *See also* citation is given to the location where the subject is indexed in detail. For example, Activated carbon adsorption, 1138–1162, is followed by (*See also* Adsorption). Where a subject is indexed in detail (e.g., Biochemical oxygen demand) and a commonly used abbreviation (e.g., BOD) is also cited under the same letter of the alphabet, the abbreviated entry is followed by a *See* citation. For example, the index entry BOD is followed by (*see* Biochemical oxygen demand). Where an abbreviation occurs under a different letter of the alphabet than where the detailed citation is given, inclusive page numbers are given followed by the *See also* citation. For older or unused terms, the *see* citation is used to direct the reader to the appropriate term used.

To access the number of data tables in the textbook more easily, an index entry followed by the capital letter T in parenthesis [e.g., (T)] is used to denote a data table related to the subject matter.