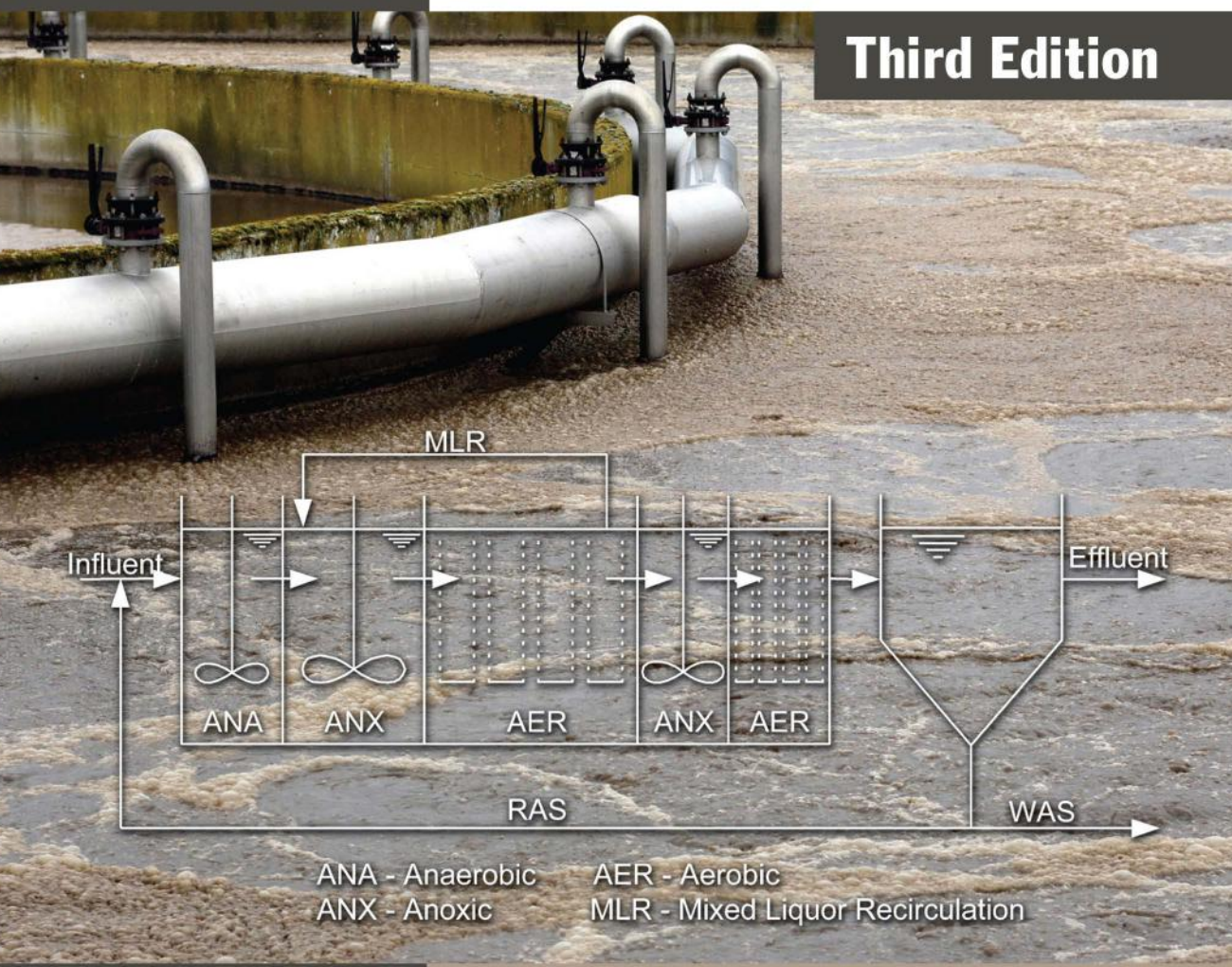
Biological Wastewater Treatment

Third Edition



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For Francisco, Jack, Matt, Rita, Ryan, Sophia, and all of the other children who will live in an increasingly crowded world. We hope that the material in this book will make it less polluted and more sustainable.

Disclaimer

This book has been prepared based on information presented in the technical and professional literature and the knowledge and experience of the authors. The authors' intention is to present, to the best of their ability, their profession's current understanding of the design and operation of biological wastewater treatment processes. The reader must recognize, however, that both the authors' understanding of the current state of the art and the profession's understanding of the principles on which the processes operate are unavoidably incomplete. This book was prepared primarily for instructional purposes, and it is the knowledge and experience of the designer and operator that determine its success, not the use of any particular design or operational procedure. Thus, while the information presented in this book may serve to supplement the expertise of a competent practitioner, it is not a replacement. It is the user's responsibility to independently verify and interpret information found in this book prior to its application. Consequently, use of the information presented in this book does hereby release the authors, the publisher, and the authors' employers from liability for any loss or injuries of any nature that may result from use of the information presented.

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PART II Theory: Modeling of Ideal Suspended Growth Reactors

Preface

The components in wastewater treatment processes may be conveniently categorized as physical, chemical, and biochemical unit operations. A thorough understanding of the principles governing their behavior is a prerequisite for process design. This "unit operations approach" to the study of process engineering has been widely accepted in the field of environmental engineering, just as in chemical engineering where it was developed, and environmental engineering textbooks now commonly use it. The purpose of this book is to present the theoretical principles and design procedures for the biochemical operations used in wastewater treatment processes. It follows in the tradition established with *Biological Wastewater Treatment: Theory and Applications* (1980) and its successor, *Biological Wastewater Treatment, Second Edition, Revised and Expanded* (1999).

The field of biological wastewater treatment has continued to evolve since 1999, and we have sought to capture our increased understanding in this new edition. Our knowledge of biological nutrient removal has increased markedly and much of that knowledge has been captured in new versions of the International Water Association (IWA) activated sludge models. We have revised our presentation of the microbiology and kinetics of nutrient removal to reflect that advance in knowledge and have updated the simulation of biological phosphorus removal with a newer version of the model. Our profession's increased understanding of anaerobic systems is reflected in the IWA anaerobic digestion model and, consequently, we have added a new chapter specifically devoted to the description and simulation of anaerobic bioreactors. We have also updated the modeling of attached growth systems to take advantage of solution techniques introduced—but not applied—in the second edition. Just as our basic understanding of biochemical operations has increased in the past decade, our application of those operations in practice has continued to evolve. All of the application chapters have been updated to reflect that evolution. Of particular significance is the increased application of submerged attached growth bioreactors and thus the chapter dealing with them was revised extensively. One realization during the past decade concerned the presence of trace organic compounds in the environment, much of which come from consumer products in wastewaters. Consequently, we have added information on the fate and effects of trace contaminants to the chapter dealing with xenobiotic organic chemicals. Finally, during the past decade, humankind began to realize the limitations associated with finite resources and began taking small steps toward increased sustainability. Consequently, because biochemical unit operations have much to offer for achieving a more sustainable world, we have added a chapter on designing systems for sustainability.

The book continues to be organized into six parts: Part I, Introduction and Background; Part II, Theory: Modeling of Ideal Suspended Growth Reactors; Part III, Applications: Suspended Growth Reactors; Part IV, Theory: Modeling of Ideal Attached Growth Reactors; Part V, Applications: Attached Growth Reactors; and Part VI, Future Challenges.

Part I seeks to do three things. First, it describes the various "named biochemical operations" in terms of their treatment objectives, biochemical environment, and reactor configuration. This helps to remove some of the confusion caused by the somewhat peculiar names given to some biochemical operations early in their history. Second, it introduces the format and notation that will be used to present the models describing the biochemical operations. Finally, it presents the basic stoichiometry and kinetics of the various microbial reactions that form the key for quantitative description of biochemical operations.

In Part II, the stoichiometry and kinetics are used in mass balance equations to investigate the theoretical performance of biological reactors containing microorganisms growing suspended in

the wastewater as it moves through the system. Part II is at the heart of the book because it provides the reader with a fundamental understanding of why suspended growth reactors behave as they do.

In Part III, the theory is applied to the various named suspended growth biochemical operations introduced in Part I. In that application, however, care is taken to point out when practical constraints must be applied to ensure that the system will function properly in the real world. In this way, the reader obtains a rational basis for the design of biological wastewater treatment operations that incorporates knowledge that has been obtained through practice. In other words, we have sought to make Part III as practical as possible.

Parts IV and V parallel Parts II and III in organization but focus on biochemical operations in which microorganisms grow attached to solid surfaces. This mode of growth adds complexity to the analysis, even though the operations are often simpler in application.

Finally, Part VI looks to the future, introducing the fate and effects of xenobiotic and trace contaminants in wastewater treatment systems and examining how the application of biochemical operations can lead to a more sustainable world.

Our plan in preparing this book was to provide a text for use in a graduate-level environmental engineering course of three semester-hours' credit for students who have had a course in environmental microbiology. In reality, the amount of information provided is more than can be covered comfortably. This provides latitude for the instructor but also makes the book a resource for the student wanting to know more than the minimum. Furthermore, it is our hope that our professional colleagues will find the book to be worthwhile as a reference and as a resource for self-guided study.

At this point, we would like to add a note of caution to the students using this book. Parts II and IV rely heavily upon modeling to provide a conceptual picture of how biochemical operations function. Although the models employed are based on our best current ideas, one must always remember that they are just someone's way of describing in simple terms very complex phenomena. Their purpose is to help the reader learn to think about the processes described by providing "experience." One should not fall into the trap, however, of substituting the models and their simulated experience for reality. Engineering requires the application of judgment in situations lacking sufficient information. The reader can use the background provided by this book to help gain sound judgment but should not hesitate to discard concepts when real-world experience indicates that they are incorrect or don't apply. Theories are constantly evolving, so be prepared to change your ideas as our knowledge advances.

As with any book, many people have had a hand in its preparation, either directly or indirectly. First, we would like to thank Henry C. Lim, coauthor of the first edition, whose approach to process engineering continues to permeate the work. His thoughts on the use of effectiveness factors in modeling attached growth systems remains an important component of this edition. Second, Dr. Grady owes a great deal to M. Henze of the Technical University of Denmark, W. Gujer of the Swiss Federal Institute of Aquatic Science and Technology, G. v. R. Marais of the University of Cape Town (now retired), and T. Matsuo of Toyo University for all that he learned through long discussions about the modeling of suspended growth biological reactors when he studied with them as members of the first IWA task group on mathematical modeling. Third, this book would not have been possible without the dedication and work of the hundreds of researchers (both fundamental and applied) who generated the knowledge upon which it is based. Once again we ask for the forbearance of those we did not cite. Fourth, we thank the thousands of practitioners (both designers and operators) who have had the foresight and faith to use biological processes to treat such a wide variety of wastewaters. Their observations and factual documentation of the performance and operational characteristics of these processes have provided both a sound basis for process design and operation and the development of new process options. It is the combination of thoughtful and creative research and application that has provided the factual basis for this book. Fifth, we thank the students at Purdue University, Clemson University, Virginia Tech, and the University of Michigan who tolerated draft versions of all editions of this book and provided helpful comments about how to make the material more understandable to them. Sixth, we would like to express our

appreciation to the many people directly involved in the preparation of this book. Among them are Rebecca E. Laura, who did the art work for the second edition, most of which has been carried over to this edition, and Jeremy Guest, who did the art work for the new simulations herein. We would also like to thank Dr. Benoit Chachaut for his work in implementing the models for packed towers and rotating biological contactors in MATLAB®. Finally, all of us acknowledge the many sacrifices made by our respective spouses to enable us to complete this project.

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Part I

Introduction and Background

As with any subject, the study of the biochemical operations used in wastewater treatment systems requires an understanding of the terminology used. The purpose of Chapter 1 is to provide that understanding by defining the nature of biochemical operations in terms of the biochemical transformations being performed, the environments within which the transformations are occurring, and the reactor configurations employed. Chapter 1 also provides descriptions of the major biochemical operations, including their process flow sheets. Engineering design is greatly facilitated by the application of mathematical models to quantitatively describe system performance. Construction of such models for biochemical operations must be based on a fundamental understanding of the microbiological events occurring in them. Chapter 2 provides that understanding, as well as an appreciation of the complex interactions occurring among the microorganisms that form the ecosystems in the operations. That appreciation is crucial to recognition of the simplified nature of the models, thereby encouraging their appropriate usage. Finally, construction of the models requires knowledge of the stoichiometry and kinetics of the major reactions occurring in biochemical operations. Chapter 3 provides that knowledge.

1 Classification of Biochemical Operations

The purpose of wastewater treatment is to remove pollutants that can harm the aquatic environment if they are discharged into it. Because of the deleterious effects of low dissolved oxygen (DO) concentrations on aquatic life, wastewater treatment engineers historically focused on the removal of pollutants that would deplete the DO in receiving waters. These so-called oxygen-demanding materials exert their effects by serving as a food source for aquatic microorganisms, which use oxygen in their metabolism and are capable of surviving at lower DO levels than higher life forms. Most oxygen-demanding pollutants are organic compounds, but ammonia nitrogen is an important inorganic one. Thus, early wastewater treatment systems were designed to remove organic matter and sometimes to oxidize ammonia nitrogen to nitrate nitrogen, and this is still the goal of many systems being built today. As industrialization and population growth continued, another problem was recognized—eutrophication, which is the accelerated aging of lakes, estuaries, and so on due to excessive plant and algal growth. This is the result of the discharge of nutrients such as nitrogen and phosphorus. Hence, engineers became concerned with the design of wastewater treatment systems that could remove these pollutants in an efficient and cost-effective manner. Most recently, we have become concerned about the discharge of toxic organic chemicals to the environment. Many of them are organic, and thus the processes used to remove oxygen-demanding materials are effective against them as well.

In addition to the categories listed above, pollutants in wastewaters may be characterized in a number of ways. For example, they may be classified by their physical characteristics (e.g., soluble or insoluble), by their chemical characteristics (e.g., organic or inorganic), by their susceptibility to alteration by microorganisms (e.g., biodegradable or nonbiodegradable), by their origin (e.g., biogenic or anthropogenic), by their effects (e.g., toxic or nontoxic), and so on. Obviously, these are not exclusive classifications, but overlap. Thus, we may have soluble, biodegradable organic material; insoluble, biodegradable organic material; and so on. The job of the wastewater treatment engineer is to design a process train that will remove all of them in an efficient and economical manner. This requires a sound understanding of process engineering, which must be built on a thorough knowledge of unit operations. Unit operations, which are the components that are linked together to form a process train, are commonly divided on the basis of the fundamental mechanisms acting within them (i.e., physical, chemical, and biochemical). Physical operations are those, such as sedimentation, that are governed by the laws of physics. Chemical operations are those in which strictly chemical reactions occur, such as precipitation. Biochemical operations are those that use living microorganisms to destroy or transform pollutants through enzymatically catalyzed chemical reactions. In this book we will examine the role of biochemical operations in wastewater treatment process trains and develop the methods for their design.

1.1 THE ROLE OF BIOCHEMICAL OPERATIONS

The most effective way to define the role of biochemical operations in wastewater treatment systems is to examine a typical process flow diagram, as shown in Figure 1.1. Four categories of pollutants are traced through the process, with the widths of the arrows depicting them being indicative of their mass flow rates. They are soluble organic matter (SOM), insoluble organic matter (IOM), soluble

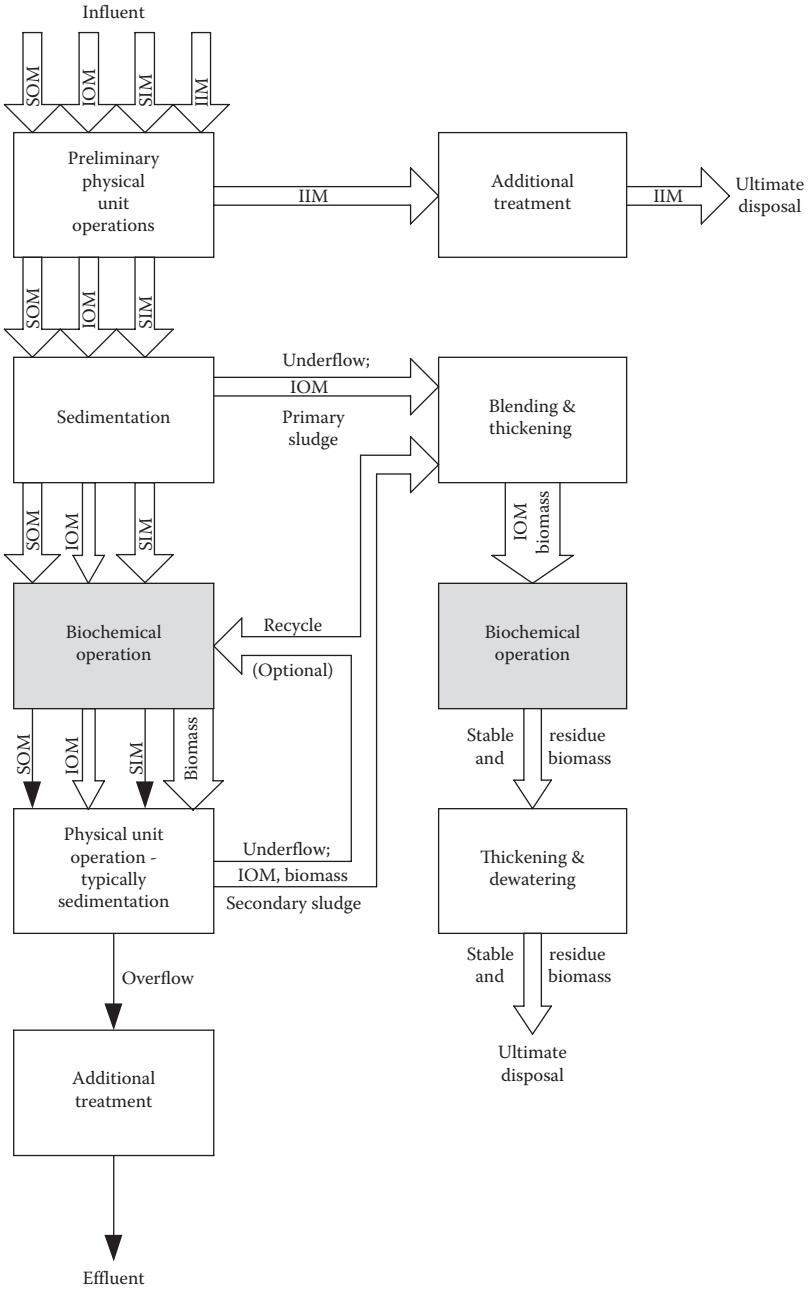

FIGURE 1.1 Typical process flow diagram for a wastewater treatment system illustrating the role of the biochemical operations. SOM = soluble organic matter; IOM = insoluble organic matter; SIM = soluble inorganic matter; IIM = insoluble inorganic matter.

inorganic matter (SIM), and insoluble inorganic matter (IIM). For the most part, the transformation rates of insoluble inorganic matter by microorganisms are too low to be of practical importance. Thus, insoluble inorganic matter is typically removed by preliminary physical unit operations and taken elsewhere for treatment and disposal. Wastewaters occur in large volume, but the pollutants are relatively dilute. Thus, engineers attempt to remove pollutants in the most efficient way, concentrating

them where possible to reduce the volumes that must be handled. For insoluble constituents this can be accomplished by the physical operation of sedimentation, which is why it is often one of the first unit operations in a treatment system. The effluent from a sedimentation basin (overflow) contains all of the soluble constituents in the influent, plus those insoluble ones that were too small to be removed. The bulk of the insoluble material, however, exits from the bottom of the vessel (underflow) as a thick suspension called "sludge." Both the overflow and the underflow require further treatment, and that is where biochemical operations come into play.

Most unit operations used for the destruction or transformation of soluble pollutants in the overflow are biochemical ones. This is because biochemical operations function more efficiently than chemical and physical ones when the concentrations of reacting constituents are low. In biochemical operations the soluble pollutants are converted either into an innocuous form, such as carbon dioxide or nitrogen gas, or into new microbial biomass, which can be removed by a physical operation because it is a particulate. In addition, as the microorganisms grow, they entrap insoluble organic matter that escaped removal upstream, thereby allowing it to be removed from the wastewater by the physical operation as well. Consequently, the effluent from the physical operation is relatively clean and often can be discharged with little or no additional treatment. A portion of the insoluble materials removed by the physical operation may be returned to the upstream biochemical operation while the remainder is transferred to another portion of the process train for further treatment.

The other major use of biochemical operations is in the treatment of sludges, as shown in Figure 1.1. Primary sludges are those resulting from sedimentation of the wastewater prior to the application of any biochemical operations. Secondary sludges are those produced by biomass growth in the biochemical operations and by entrapment of insoluble organic matter by that biomass. The nature of the materials in primary sludges tends to be very diverse because of the multitude of sources from which the materials arise, whereas secondary sludges are more uniform, being mainly microbial biomass. Sometimes the two sludges are blended and treated together as shown in the figure, but at other times they are treated separately. This is because the efficacy of a biochemical operation in treating sludge depends strongly on the nature of the materials in it.

In spite of the major role of biochemical operations in the treatment of wastewaters, if a visitor to a treatment facility were to ask the name of the particular biochemical operation being used, the answer generally would give little indication of its nature. In fact, the most common operation, activated sludge, was named before its biochemical nature was even recognized. Consequently, before starting the study of the various biochemical operations it would be beneficial to establish what they are and what they do.

1.2 CRITERIA FOR CLASSIFICATION

The classification of biochemical operations may be approached from three points of view: (1) the biochemical transformation, (2) the biochemical environment, and (3) the bioreactor configuration. If all are considered together, the result is a detailed classification system that will aid the engineer in choosing the operation most appropriate for a given need.

1.2.1 The Biochemical Transformation

1.2.1.1 Removal of Soluble Organic Matter

The major application of biochemical operations to the main wastewater stream is for the removal of soluble organic matter. This occurs as the microorganisms use it as a food source, converting a portion of the carbon in it into new biomass and the remainder into carbon dioxide. The carbon dioxide is evolved as a gas and the biomass is removed by liquid:solid separation, leaving the wastewater free of the original organic matter. Because a large portion of the carbon in the original organic matter is oxidized to carbon dioxide, removal of soluble organic matter is also often referred to as carbon oxidation.

Aerobic cultures of microorganisms are particularly suitable for the removal of organic matter in the concentration range between 50 and 4000 mg/L as biodegradable chemical oxygen demand (COD). At lower concentrations, carbon adsorption is often more economical, although biochemical operations are being used for treatment of contaminated groundwater that contains less than 50 mg/L of COD. Although they must often be followed by aerobic cultures to provide an effluent suitable for discharge, anaerobic cultures are frequently used for high strength wastewaters because they do not require oxygen, give less excess biomass, and produce methane gas as a usable product. If the COD concentration to be removed is above 50,000 mg/L, however, then evaporation and incineration may be more economical. Anaerobic cultures are also used to treat wastewaters of moderate strength (down to about 1000 mg/L as COD), and have been proposed for use with dilute wastewaters as well. It should be emphasized that the concentrations given are for soluble organic matter. Suspended or colloidal organic matter is often removed more easily from the main wastewater stream by physical or chemical means and then treated in a concentrated form. Mixtures of soluble, colloidal, and suspended organic matter are often treated by biochemical means, however.

1.2.1.2 Stabilization of Insoluble Organic Matter

Many wastewaters contain appreciable quantities of colloidal organic matter that are not removed by sedimentation. When they are treated in a biochemical operation for removal of the soluble organic matter, much of the colloidal organic matter is entrapped with the biomass and ultimately converted to stable end products that are resistant to further biological activity. The formation of such stable end products is referred to as stabilization. Some stabilization will occur in the biochemical operation removing the soluble organic matter, but most will occur in operations designed specifically for that purpose.

Insoluble organic matter comes from the wastewater itself and from the growth of microorganisms as they remove soluble organic matter. Because these solids can be removed from the wastewater by settling, they are normally concentrated by sedimentation before being subjected to stabilization by biochemical means. Stabilization is accomplished both aerobically and anaerobically, although anaerobic stabilization is more energy efficient. The end products of stabilization are carbon dioxide, inorganic solids, and insoluble organic residues that are relatively resistant to further biological activity and have characteristics similar to humus. In addition, methane gas is a product from anaerobic operations.

1.2.1.3 Conversion of Soluble Inorganic Matter

Since the discovery during the 1960s of the effects of eutrophication, engineers have been concerned about the removal of inorganic nutrients from wastewater. Two of the prime causes of eutrophication are nitrogen and phosphorus, and a number of biological nutrient removal processes have been developed to remove them. Phosphorus is present in domestic wastewater in an inorganic form as orthophosphate, condensed phosphates (e.g., pyrophosphate, tripolyphosphate, and trimetaphosphate), and organic phosphate (e.g., sugar phosphates, phospholipids, and nucleotides). Both condensed phosphates and organic phosphate are converted to orthophosphate through microbial activity. Orthophosphate, in turn, is removed through its uptake by specialized bacteria possessing unique growth characteristics that allow them to store large quantities of it in granules within the cell. Nitrogen is present in domestic wastewater as ammonia and as organic nitrogen (e.g., amino acids, protein, and nucleotides), which is converted to ammonia as the organic matter is biodegraded. Two groups of bacteria are required to convert the ammonia into an innocuous form. First, nitrifying bacteria oxidize it to nitrate in a process called nitrification. Then denitrifying bacteria convert the nitrate to nitrogen gas in a process called denitrification. The nitrogen gas escapes to the atmosphere. Other inorganic transformations occur in nature, but few are exploited on a large scale in biochemical operations.

1.2.2 The Biochemical Environment

The most important characteristic of the environment in which microorganisms grow is the terminal acceptor of the electrons they remove as they oxidize chemicals to obtain energy. There are three major types of electron acceptors: oxygen, inorganic compounds, and organic compounds. If dissolved oxygen is present or supplied in sufficient quantity so as to not be rate limiting, the environment is considered to be aerobic. Growth is generally most efficient in this environment and the amount of biomass formed per unit of waste destroyed is high. Strictly speaking, any environment that is not aerobic is anaerobic. Within the wastewater treatment field, however, the term anaerobic is normally reserved for the situation in which organic compounds, carbon dioxide, and sulfate serve as the major terminal electron acceptor and in which the electrode potential is very negative. Growth is less efficient under this condition. When nitrate and/or nitrite are present and serve as the primary electron acceptor in the absence of oxygen, the environment is called anoxic. The presence of nitrate and/or nitrite causes the electrode potential to be higher and growth to be more efficient than under anaerobic conditions, although not as high or as efficient as when oxygen is present.

The biochemical environment has a profound effect on the ecology of the microbial community. Aerobic operations tend to support complete food chains from bacteria at the bottom to rotifers at the top. Anoxic environments are more limited and anaerobic are most limited, being predominantly bacterial. The biochemical environment influences the outcome of the treatment process because the microorganisms growing in the three environments may have quite different metabolic pathways. This becomes important during the treatment of industrial wastewaters because some transformations can be carried out aerobically but not anaerobically and vice versa.

1.2.3 Bioreactor Configuration

The importance of classifying biochemical operations according to bioreactor type follows from the fact that the completeness of a given biochemical transformation will be strongly influenced by the physical configuration of the bioreactor in which it is being carried out. Therefore, it is important to get a clear picture of the many bioreactor types available.

Wastewater treatment bioreactors fall into two major categories, depending on the way in which microorganisms grow in them: suspended in the liquid under treatment or attached to a solid support. When suspended growth cultures are used, mixing is required to keep the biomass in suspension and some form of physical unit operation, such as sedimentation or membrane filtration, is used to remove the biomass from the treated effluent prior to discharge. In contrast, attached growth cultures grow as a biofilm on a solid support and the liquid being treated flows past them. However, because organisms can slough from the support, a physical unit operation is usually required before the treated effluent may be discharged.

1.2.3.1 Suspended Growth Bioreactors

The simplest possible continuous flow suspended growth bioreactor is the continuous stirred tank reactor (CSTR), which consists of a well-mixed vessel with a pollutant-rich influent stream and a treated effluent stream containing microorganisms. The liquid volume is constant and the mixing is sufficient to make the concentrations of all constituents uniform throughout the reactor and equal to the concentrations in the effluent. Consequently, these reactors are also called completely mixed reactors. The uniform conditions maintain the biomass in a constant average physiological state. Considerable operational flexibility may be gained by the addition of a physical unit operation, such as a sedimentation basin, which captures the biomass, as shown in Figure 1.1. As discussed previously, the overflow from the sedimentation basin is relatively free of biomass, while the underflow contains concentrated slurry. Most of that concentrated slurry is recycled to the bioreactor but a portion is wasted. Because the wasted biomass is organic, it must be treated in an appropriate process before release to the environment.

Connecting several CSTRs in series offers additional flexibility as feed may be added to any or all of them. Furthermore, biomass recycle may be employed about the entire chain or any portion of it. The behavior of such systems is complex because the physiological state of the biomass changes as it passes from bioreactor to bioreactor. Nevertheless, many common wastewater treatment systems use bioreactors with split influent and recycle streams. One advantage of multistage systems is that different environments may be imposed upon different stages, thereby allowing multiple objectives to be accomplished. This is very common in biological nutrient removal processes.

A batch reactor is a completely mixed reactor without continuous flow through it. Instead, a "batch" of material is placed into the vessel with the appropriate biomass and allowed to react to completion as the microorganisms grow on the pollutants present. As growth proceeds, reaction conditions change and, consequently, so does the growth environment. Batch processes can be very flexible and are particularly well suited for situations with low or highly variable flows. Furthermore, by changing the nature of the electron acceptor temporally, it is also possible to accomplish nutrient removal in a single bioreactor. Because their operation follows a sequence of events, they are commonly called sequencing batch reactors (SBRs).

A perfect plug-flow reactor (PFR) is one in which fluid elements move through in the same order that they enter, without intermixing. Thus, the perfect PFR and the CSTR represent the two extreme ends of the continuum of all possible degrees of mixing. Because of the lack of intermixing, perfect PFRs may be considered to contain an infinite number of moving batch cultures wherein changes occur spatially as well as temporally. Both, however, cause the biomass to go through cycles of physiological change that can have strong impacts on both community structure and activity. Because perfect PFRs are difficult to achieve in practice, plug-flow conditions are generally approximated with a number of CSTRs in series. In Chapter 4 we will examine ways of characterizing the mixing conditions in suspended growth bioreactors.

1.2.3.2 Attached Growth Bioreactors

There are three major types of attached growth bioreactors: packed towers, rotating discs, and fluidized beds. The microorganisms in a packed tower grow as a film on an immobile support, such as plastic media. In aerobic bioreactors the wastewater flows down the media in a thin film. If no recirculation of effluent is practiced, there is considerable change in the reaction environment from top to bottom of the tower as the bacteria remove the pollutants. The recirculation of effluent tends to reduce the severity of that change, and the larger the recirculation flow, the more homogeneous the environment becomes. The performance of this bioreactor type is strongly influenced by the manner in which effluent is recirculated. Organisms are continually sloughed from the support surface as a result of fluid shear. If they are removed from the effluent prior to recirculation, then pollutant removal is caused primarily by the activity of the attached biomass. On the other hand, if flow is recirculated prior to the removal of the sloughed-off microorganisms, the fluid stream will resemble that of a suspended growth bioreactor and pollutant removal will be by both attached and suspended biomass. In anaerobic packed towers, the media is submerged and flow may be either upward or downward.

The microorganisms in a rotating disc reactor (RDR) grow attached to plastic discs that are rotated in the liquid. In most situations, the horizontal shaft on which the discs are mounted is oriented perpendicular to the direction of flow and several reactors in series are used to achieve the desired effluent quality. Consequently, environmental conditions are uniform within a given reactor, but change from reactor to reactor down the chain. This means that both the microbial community structure and the physiological state change from reactor to reactor.

In fluidized bed biological reactors (FBBRs) the microorganisms grow attached to small particles, such as sand grains, which are maintained in a fluidized state by the upward velocity of the wastewater undergoing treatment. The effluent from such bioreactors generally contains little suspended biomass, but particles must continually be removed and cleaned to maintain a constant mass of microorganisms in the system. The cleaned particles are continually returned to the bioreactor

while the wasted biomass is sent to an appropriate treatment process. Recirculation of effluent around the bioreactor is usually needed to achieve the required fluidization velocity and thus the system often tends to behave as if it were completely mixed.

1.3 COMMON "NAMED" BIOCHEMICAL OPERATIONS

In almost all fields, certain operations have gained common names through years of use. Although such names are not always descriptive, they are recognized and accepted because of their historical significance. Such is the case in environmental engineering. In fact, some of the names bear little resemblance to the process objectives and are even applied to more than one reactor configuration. For purposes of discussion, 12 common names have been chosen and are listed in Table 1.1. To relate those names to the classification scheme presented above, Table 1.2 was prepared. It defines each name in terms of the bioreactor configuration, the treatment objective, and the reaction environment. Many other named biochemical operations are used, but they can all be related to those described in Table 1.2.

1.3.1 Suspended Growth Bioreactors

1.3.1.1 Activated Sludge

Four factors are common to all activated sludge processes: (1) a flocculent slurry of microorganisms (mixed liquor suspended solids [MLSS]) is used to remove soluble and particulate organic matter from the influent waste stream; (2) liquid:solid separation is used to remove the MLSS from the process flow stream, producing an effluent that is low in suspended solids; (3) concentrated solids are recycled from the liquid:solid separator back to the bioreactor; and (4) excess solids are wasted to control the solids retention time (SRT) to a desired value. Nitrification will also occur under appropriate conditions. The term mixed liquor suspended solids is used to denote the microbial slurry because it is a mixture of microorganisms, undegraded particulate substrate, and inert solids. Figure 1.2 illustrates the configuration traditionally employed, in which quiescent settling serves as the means of liquid:solid separation. The bioreactor containing the MLSS is commonly referred to as the aeration basin, and it is aerobic throughout, as indicated by the term AER in the figure. Mixing energy provided by the oxygen transfer equipment (and supplemental mixing equipment in some cases) maintains the MLSS in suspension. Quiescent settling occurs in a downstream secondary clarifier. The stream of concentrated solids being recycled from the clarifier to the bioreactor is called return activated sludge (RAS). Solids produced in the process (called waste activated sludge [WAS]) can be removed from the process at several locations to maintain the desired SRT. Two locations, from the clarifier underflow (referred to as the conventional method) and from the aeration basin (the Garrett[4] method), are illustrated in Figure 1.2.

TABLE 1.1
Common Biochemical Operations

Suspended Growth Reactors	Attached Growth Reactors
Activated sludge	Fluidized bed biological reactor
Biological nutrient removal	Rotating biological contactor
Aerobic digestion	Trickling filter
High-rate anaerobic processes	Packed bed
Anaerobic digestion	Integrated fixed film activated sludge systems
Fermenter	
Lagoon	

TABLE 1.2
Classification of "Named" Biochemical Operations

			Objective								
			Removal of Soluble Organic Matter			Stabilization of Insoluble Organic Matter			Conversion of Soluble Inorganic Matter		
Name	Acronym	Bioreactor Configuration	Aerobic	Anaerobic	Anoxic	Aerobic	Anaerobic	Anoxic	Aerobic	Anaerobic	Anoxic
			Suspended Growth Reactors								
Activated sludge		All with biomass recycle									
Completely mixed	CMAS	CSTR	X						N[a]		
Contact stabilization	CSAS	CSTRs in series	X								
Conventional	CAS	CSTRs in series or plug flow with dispersion	X						N		
Extended aeration	EAAS	CSTR, CSTRs in series, or plug flow with dispersion	X			X			N		
High purity oxygen	HPOAS	CSTRs in series	X								
Membrane bioreactor	MBRAS	CSTRs in series	X						N		
Selector	SAS	CSTRs in series	X						N		
Sequencing batch reactors	SBRAS	Completely mixed batch	X						N		
Step feed	SFAS	CSTRs in series or plug flow with dispersion, both with multiple feed points	X						N		
Biological nutrient removal	BNR	All with biomass recycle									
Biological phosphorus removal		CSTRs in series	X	X						P[b]	
Separate stage denitrification		CSTRs in series			X						D[c]
Separate stage nitrification		CSTR or CSTRs in series	X						N		
Sequencing batch reactors		Completely mixed batch	X	X	X				N	P	D

Single-sludge nitrogen removal		CSTRs in series with internal recirculation streams	X		X				N		D	
Single-sludge systems		CSTRs in series with internal recirculation streams	X	X	X				N	P	D	
Aerobic digestion												
Conventional	CAD	CSTR				X			N			
Anoxic/aerobic	A/AD	CSTRs in series				X		X	N		D	
Autothermal thermophilic	ATAD	CSTRs in series				X						
High-rate anaerobic processes												
Upflow anaerobic sludge blanket	UASB	Upflow sludge blanket reactor		X			X					
Anaerobic filter	AF	CSTR with biomass recycle		X			X					
Hybrid UASB/AF	UASB/AF	Upflow sludge blanket reactor followed by settler		X			X					
Anaerobic digestion	AD	CSTR					X					
Fermenters		CSTR with biomass recycle					X					
Lagoon		All without biomass recycle										
Completely mixed aerated	CMAL	CSTR	X			X			N			
Facultative/aerated	F/AL	Large shallow basins in series	X		X	X	X	X	N		D	
Anaerobic	ANL	Large deep basins		X			X					
Attached Growth Reactors												
Fluidized bed biological reactors	FBBR											
Aerobic		Fluidized bed with oxygenation cell	X						N			
Anaerobic		Fluidized bed		X								
Anoxic		Fluidized bed			X						D	

(*Continued*)

TABLE 1.2 (CONTINUED)
Classification of "Named" Biochemical Operations

Name	Acronym	Bioreactor Configuration	Objective: Removal of Soluble Organic Matter			Objective: Stabilization of Insoluble Organic Matter			Objective: Conversion of Soluble Inorganic Matter		
			Aerobic	Anaerobic	Anoxic	Aerobic	Anaerobic	Anoxic	Aerobic	Anaerobic	Anoxic
Attached Growth Reactors											
Rotating biological contactor	RBC	Rotating disc	X						N		
Trickling filter	TF	Packed tower with large media	X						N		
Packed bed	DFPB or UFPB	Submerged packed tower with small media	X		X				N		D
Integrated fixed film activated sludge systems	IFAS	Media of various types added to suspended growth bioreactor, with or without biomass recycle	X	X	X				N	P	D

[a] Nitrification.
[b] Phosphorus uptake or release. Requires both aerobic and anaerobic zones.
[c] Denitrification.

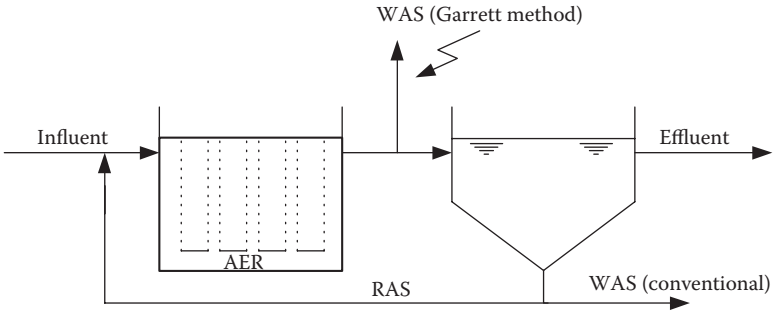

FIGURE 1.2 Typical activated sludge process.

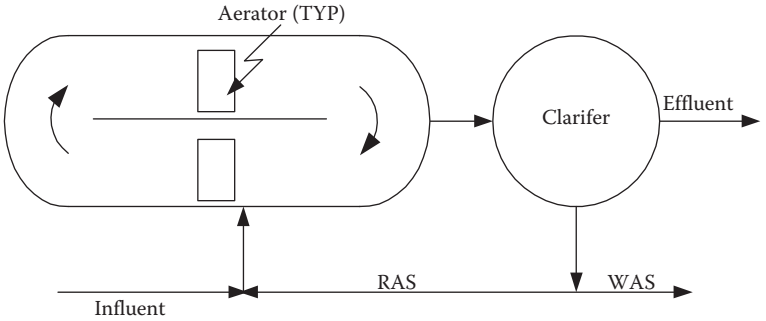

FIGURE 1.3 Oxidation ditch activated sludge process. An example of extended aeration activated sludge (EAAS).

While many different types of activated sludge systems exist, nine are listed in Table 1.2. This suggests that the term activated sludge is not very descriptive. As further indicated in Table 1.2, the primary treatment objective for all activated sludge processes is the removal of soluble organic matter and oxidation of the carbon contained in it. Under appropriate conditions, nitrification will also occur, and thus it is listed as an objective for those systems in which it is most likely. Extended aeration activated sludge (EAAS) systems are often used on wastewaters that have not been treated in a physical operation to remove suspended organic matter. In that case, the insoluble organic matter becomes trapped in the biofloc and undergoes some oxidation and stabilization. Thus that objective is marked for it. As illustrated in Figure 1.3, EAAS systems are often configured as closed loop bioreactors, typically referred to as oxidation ditches. The other activated sludge types can be used on wastewaters from which settleable solids either have or have not been removed. However, those wastewaters still contain colloidal organic matter, most of which will be removed along with the soluble organic matter. Even though the colloidal material is insoluble and will be partially stabilized during treatment, the main event governing system performance is removal of the soluble organic matter, which is listed as the main treatment objective.

The first uses of activated sludge were on a batch basis.[2] At the end of each aeration period, suspended solids (referred to as sludge) were present and they were left in the bioreactor when the clear wastewater was withdrawn after settling. As this batch procedure was repeated the quantity of suspended solids increased, giving more complete removal of organic matter within the allotted reaction time. Although this increase in suspended solids with the associated improvement in removal activity was due to the growth of a viable microbial culture, the reason was unknown to the early researchers, who characterized the sludge as being "activated," thereby giving the process its name.[6] Use of the batch process waned as larger facilities were required, but during the 1970s there was a resurgence of interest in the use of batch reactors because of the flexibility offered in small

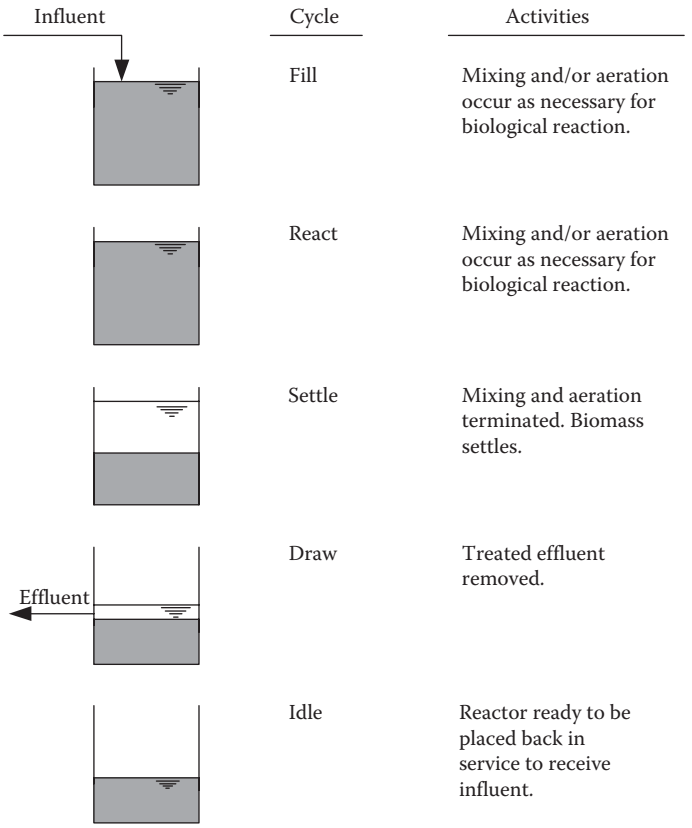

FIGURE 1.4 Sequencing batch reactor activated sludge (SBRAS) operating cycle.

installations. Now referred to as sequencing batch reactor activated sludge (SBRAS), many are in use treating both municipal and industrial wastewaters. Figure 1.4 illustrates the typical operating cycle for a modern SBRAS.

As the need to treat larger flows increased, the early batch operation was converted to continuous flow through the use of long aeration chambers similar to plug-flow reactors, followed by sedimentation and biomass recycle. Such systems are called conventional activated sludge (CAS). Various modifications of the plug-flow reactor were tried, among them the introduction of the wastewater at various points along the tank while continuing to add the RAS at the inlet end, in what has been called step feed activated sludge (SFAS). The result is that a gradient in MLSS concentration is produced with the highest concentrations at the inlet of the aeration basin and the lowest at the outlet. Figure 1.5 illustrates two ways in which this is typically accomplished in practice. Figure 1.5a depicts a single narrow basin with influent added at various points along its length, while Figure 1.5b shows a series of such basins (each often referred to as a pass) with influent added to each. A further extension of this concept is contact stabilization activated sludge (CSAS) where influent is added at a single downstream feed point. The result is that the portion of the aeration basin upstream of the feed point contains only RAS and the portion downstream MLSS.

In the middle 1950s various engineers began advocating the CSTR with cell recycle as an alternative to the CAS reactor because of its inherent stability. That stability, plus the advantages regarding the maintenance of the microbial community in a relatively constant physiological state, caused wide adoption of the completely mixed activated sludge (CMAS) process, particularly for the treatment of industrial wastewaters. Figure 1.6 illustrates two bioreactor configurations commonly used to achieve completely mixed conditions. The first (Figure 1.6a) has been used with diffused aeration

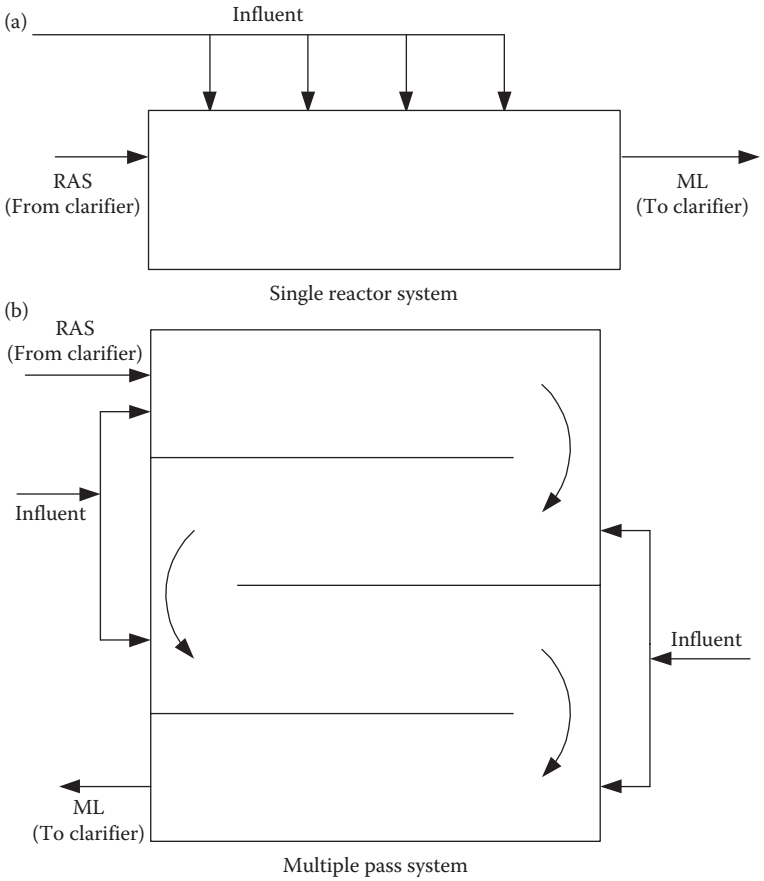

FIGURE 1.5 Step feed activated sludge (SFAS) process.

systems; complete mixing is achieved by distributing the influent along one side of a long, narrow bioreactor, with effluent being taken from the opposite side. Alternatively (Figure 1.6b), an essentially square shaped bioreactor has been used with influent and effluent locations positioned to achieve completely mixed conditions. Mechanical surface aeration is typically used with the latter because it provides good overall circulation of basin contents. Multiple inlets with each located near an aerator may be used when several aerators are present in the basin.

Experience with CMAS revealed that it tended to produce sludges that did not settle as well as sludges from systems containing concentration gradients. Consequently, today many bioreactor systems are in use that employ several small CSTRs in series before a large one, as illustrated in Figure 1.7, thereby achieving desired environmental conditions. Such systems are referred to as selector activated sludge (SAS) systems because they select for microbes with desired settling characteristics. Other innovations that require CSTRs in series, such as the use of high purity oxygen activated sludge (HPOAS), illustrated in Figure 1.8, have also been adopted.[1]

A recent development is membrane bioreactor activated sludge (MBRAS), in which a membrane filter is used to separate the treated effluent from the MLSS and concentrate the MLSS for return to the aeration basin. As illustrated in Figure 1.9a, in one system a pressurized membrane filtration unit is located outside of the aeration basin, like a clarifier, with mixed liquor pumped to it and RAS returned to the aeration basin by gravity (generally referred to as an external membrane bioreactor or MBR). Alternately the membranes may be immersed in a portion of the aeration basin, with effluent withdrawn through the membranes. The membranes may be in the main aeration basin or

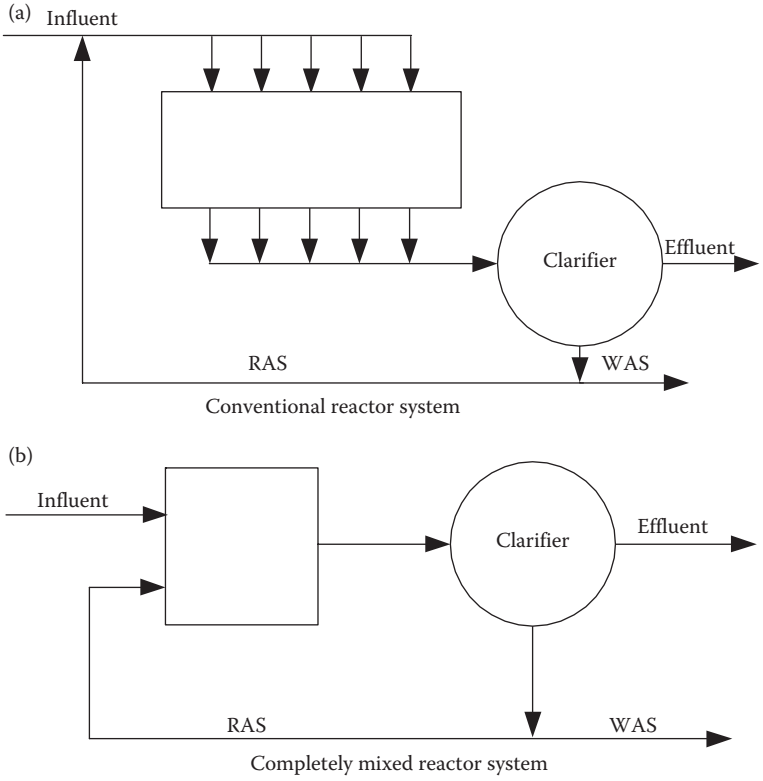

FIGURE 1.6 Completely mixed activated sludge (CMAS) process.

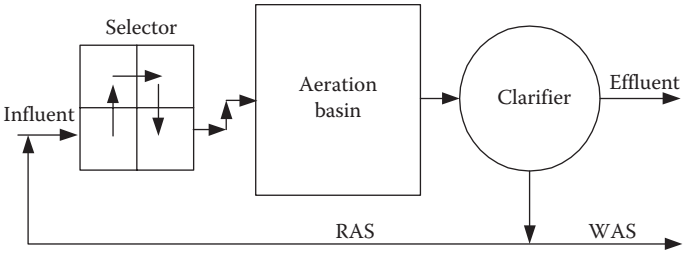

FIGURE 1.7 Selector activated sludge (SAS) process.

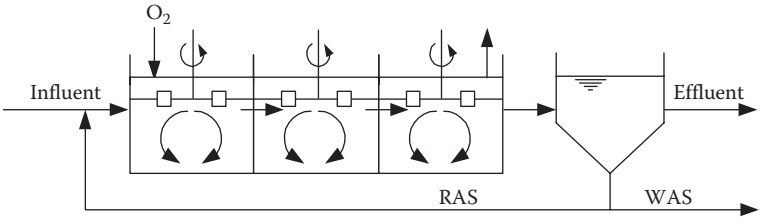

FIGURE 1.8 High purity oxygen activated sludge (HPOAS) process.

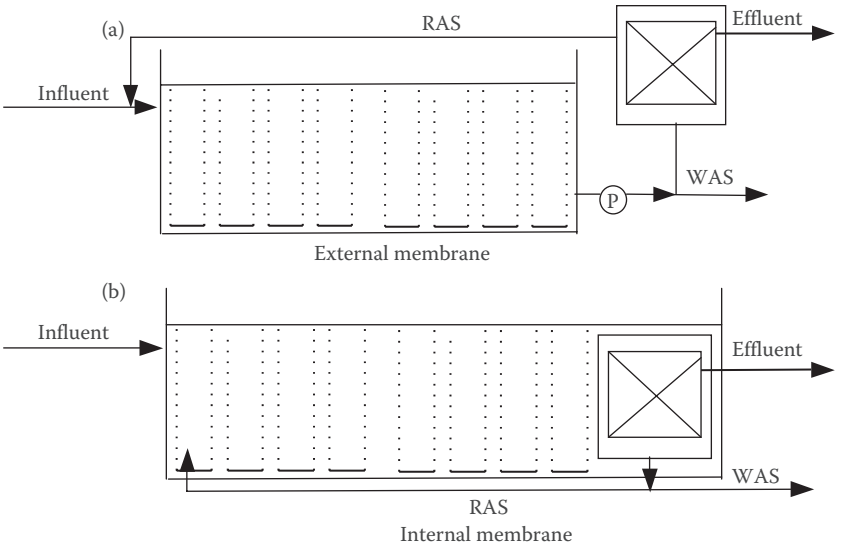

FIGURE 1.9 Membrane bioreactor activated sludge (MBRAS) process.

in a separate basin, but in either case mixed liquor is recirculated from the membrane section to the remainder of the aeration basin to ensure distribution of the solids throughout (Figure 1.9b). These systems are referred to as submerged or immersed MBRs. The use of membrane filters rather than gravity sedimentation for biomass separation and retention offers several advantages, including higher quality effluent (lower in particulate matter) and more compact systems.

The history of the activated sludge process is very interesting and the reader is encouraged to learn more about it by referring to Alleman and Prakasam[2] and Sawyer.[6] The theoretical performance of activated sludge systems is discussed in Chapters 5 and 6 while their design is covered in Chapter 11.

1.3.1.2 Biological Nutrient Removal

Biological nutrient removal (BNR) systems are among the most complicated biochemical operations devised for wastewater treatment, and like the activated sludge systems from which they were derived, they come in a number of configurations. Some of these configurations are listed in Table 1.2. The common feature of all BNR processes is that they are divided into zones containing different biochemical environments, as illustrated in Figure 1.10. Provision of these zones allows the BNR processes to remove nitrogen and/or phosphorus.

A biological phosphorus removal system is essentially an activated sludge system employing CSTRs in series, in which the first bioreactor is anaerobic to encourage the growth of specialized phosphorus-storing bacteria. The prototype biological phosphorus removal process, illustrated in Figure 1.11, is the A/O (anaerobic/oxic) process. It is also known as the Phoredox process.

Separate stage nitrification and denitrification systems usually employ single CSTRs or CSTRs in series with cell recycle to convert ammonia to nitrate and nitrate to nitrogen gas, respectively. They are usually used as downstream treatment additions to existing systems. A separate stage nitrification system is configured essentially like CAS, as illustrated in Figure 1.2. Because the influent wastewater has already been treated to remove the soluble and particulate organic matter, the aeration basin is smaller than a comparable nitrifying activated sludge process. A separate stage denitrification system consists of an anoxic zone followed by an aerobic zone, as illustrated in Figure 1.12, and receives an influent that has previously been nitrified. A supplemental carbon source, such as methanol, is generally required because the influent wastewater does not contain

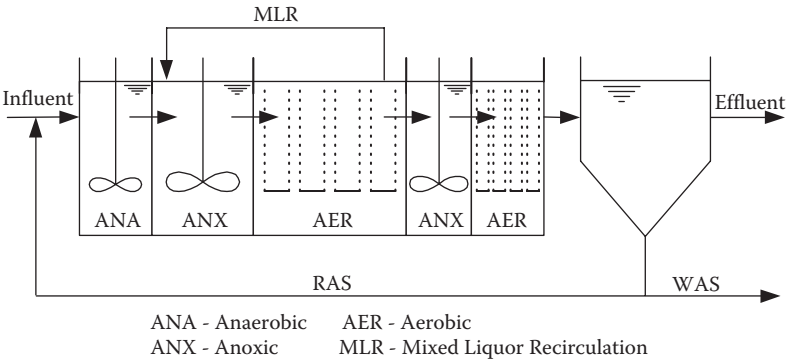

FIGURE 1.10 Single-sludge biological nutrient removal (BNR) process.

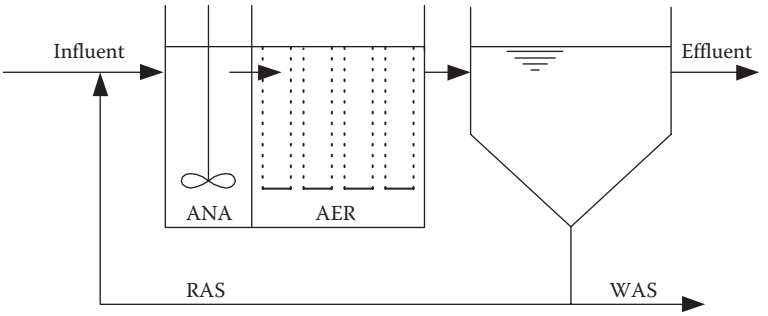

FIGURE 1.11 Anaerobic/oxic (A/O) or Phoredox biological phosphorus removal process.

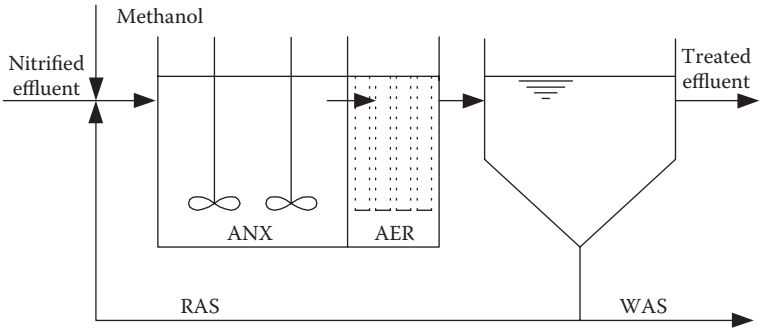

FIGURE 1.12 Separate stage suspended growth denitrification process.

sufficient biodegradable organic carbon relative to the influent nitrate. The bioreactor also contains a small aerobic zone to strip entrained nitrogen gas prior to the downstream clarifier.

Single-sludge nitrogen removal systems use the biodegradable organic matter in the influent wastewater as the carbon source for denitrification and incorporate internal mixed liquor recirculation (MLR) streams to supply nitrate to the anoxic zone. Figure 1.13 illustrates the simplest of these, the modified Ludzak-Ettinger (MLE) process. Nitrogen removal is limited in this process by the practical range of MLR flow rates. Additional nitrogen removal can be achieved when a second anoxic zone is included, as in the four-stage Bardenpho process illustrated in Figure 1.14.

Sequencing batch reactors can be made to remove phosphorus and nitrogen while they are achieving carbon oxidation by imposing anaerobic and anoxic periods during their cycles, but otherwise are similar to the SBRAS used exclusively for removal of soluble organic matter.

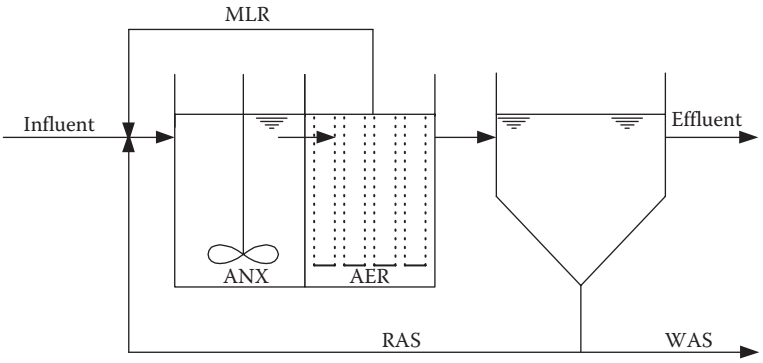

FIGURE 1.13 Modified Ludzak-Ettinger (MLE) process for single-sludge nitrogen removal.

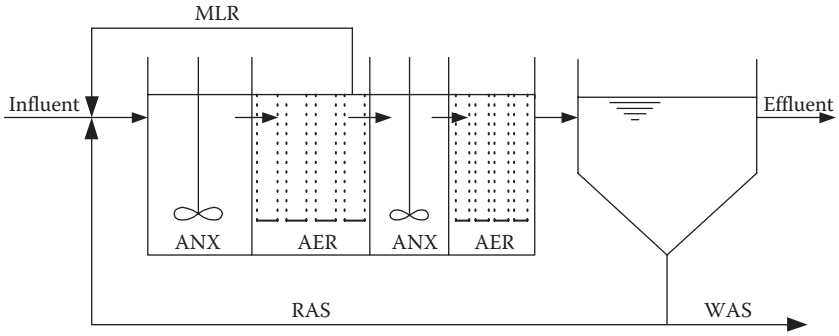

FIGURE 1.14 Four-stage Bardenpho process for single-sludge nitrogen removal.

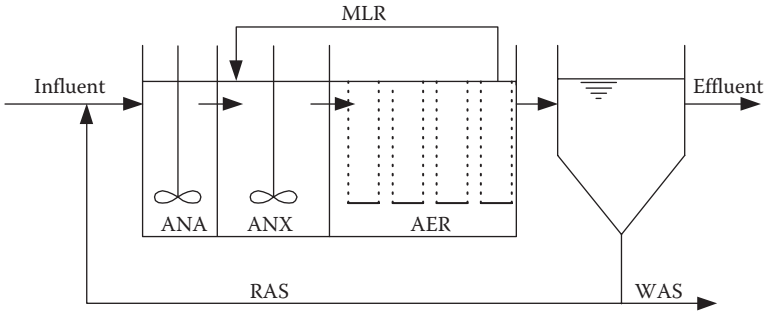

FIGURE 1.15 Anaerobic/anoxic/oxic (A^2/O) process for single-sludge nutrient removal.

The most complex BNR systems are the single-sludge systems that accomplish carbon oxidation, nitrification, denitrification, and phosphorus removal with a single biomass by recycling it through CSTRs in series in which some are aerobic, some anoxic, and some anaerobic. A prototype system is an extension of the A/O process in which an anoxic zone receiving MLR from the aerobic zone is placed between the anaerobic and aerobic zones. Termed A^2/O, for anaerobic/anoxic/oxic, it is illustrated in Figure 1.15. Several versions of these processes exist, such as the University of Capetown (UCT) process, illustrated in Figure 1.16.

The theoretical performance of BNR systems is covered in Chapters 6 and 7. Their design is discussed in Chapter 12, where several additional process configurations are described.

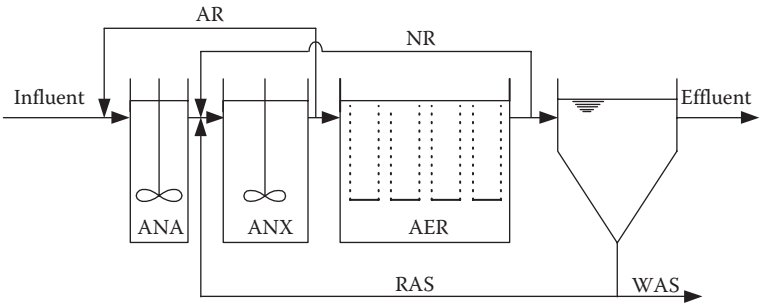

FIGURE 1.16 University of Cape Town (UCT) process for single-sludge nutrient removal.

1.3.1.3 Aerobic Digestion

Aerobic digestion is the name given to the aerobic destruction of insoluble organic matter in a suspended growth bioreactor. Generally, aerobic digesters employ a CSTR or CSTRs in series with a long SRT, allowing ample time for the conversion of much of the organic matter to carbon dioxide. Pathogens in the feed sludge are also inactivated as a result of the extended SRT provided. Although not the primary objective, nitrification also occurs to the extent that sufficient alkalinity is present. Aerobic digestion is often used to destroy part of the excess biomass formed during treatment of soluble industrial wastewater and at small "package plant" installations treating domestic wastewater. Conventional aerobic digestion (CAD) maintains the biomass in an aerobic state at all times. As illustrated in Figure 1.17, CAD systems can be operated on either a batch basis (with or without solids settling and decanting [Figure 1.17a]), or as a continuous flow system with solids settling and recycle (Figure 1.17b).

Sufficient alkalinity is generally not available to allow nitrification of all of the ammonia released from the destruction of biomass, resulting in depression of the pH as the available alkalinity is consumed. However, if the nitrate formed by nitrification is denitrified, sufficient alkalinity

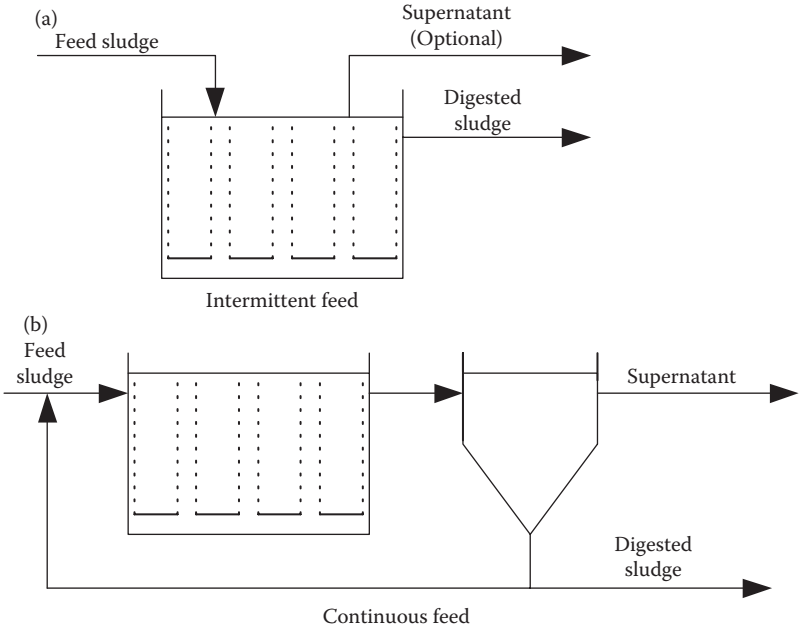

FIGURE 1.17 Conventional aerobic digestion (CAD) process.

is formed to allow complete nitrification while maintaining neutral pH. Anoxic/aerobic digestion (A/AD) cycles the biomass between aerobic and anoxic conditions to allow both nitrification and denitrification, thereby reducing costs of aeration and pH control. Figure 1.18 illustrates three A/AD configurations.

Autothermal thermophilic aerobic digestion (ATAD) systems take advantage of the heat released through the destruction of organic matter to elevate the temperature of the digester into the thermophilic range (45 to 65°C). As illustrated in Figure 1.19, the bioreactor is insulated to retain the

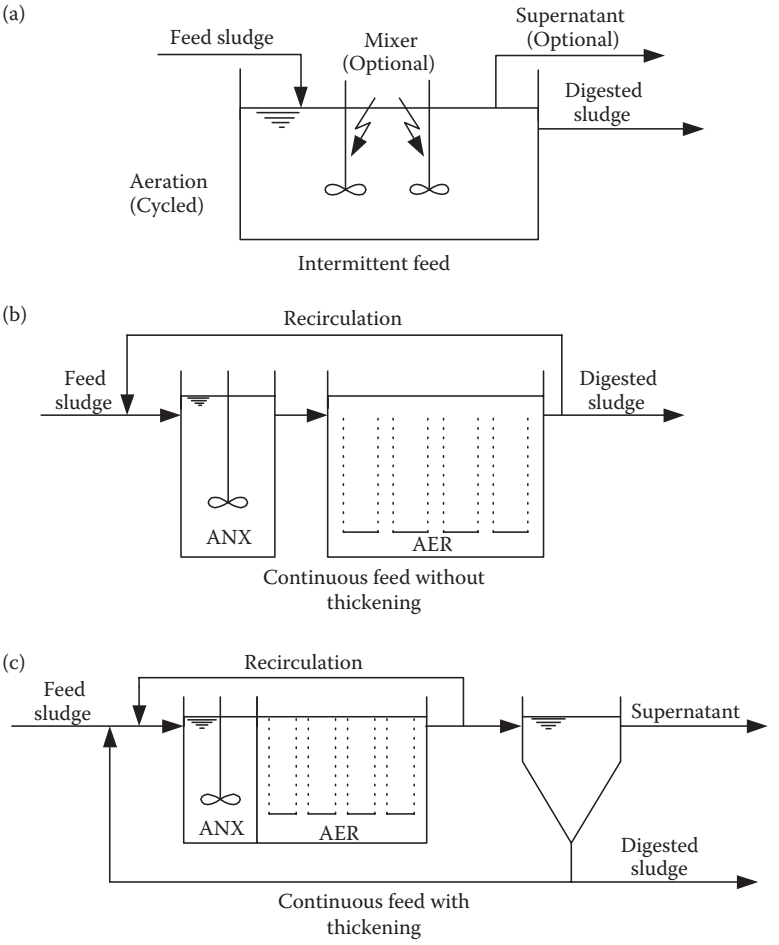

FIGURE 1.18 Anoxic/aerobic digestion (A/AD) process.

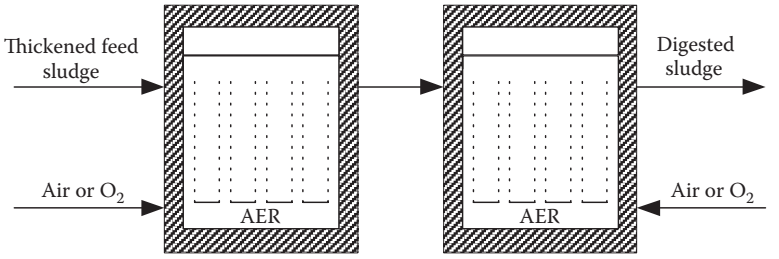

FIGURE 1.19 Autothermal thermophilic aerobic digestion (ATAD) process.

released heat. Special oxygen transfer devices are also required since feed sludge with a high solids concentration is used to minimize the amount of water that must be heated using the released heat. Increased temperatures allow for increased rates of organic matter destruction and pathogen inactivation. However, elevated temperature also prevents the growth of nitrifying bacteria, resulting in no nitrification of released ammonia.

Sometimes small treatment plants do not have primary sedimentation and allow aerobic digestion of the insoluble organic matter present in the influent to occur in the same bioreactor as the removal of soluble organic matter and the stabilization of the excess biomass formed in the process. In those cases the system is usually considered to be an extended aeration activated sludge process as discussed above. Aerobic digestion is discussed in Chapter 13.

1.3.1.4 High-Rate Suspended Growth Anaerobic Processes

Several processes are used to remove soluble organic matter under anaerobic conditions in CSTRs with cell recycle. They are also used to treat wastes containing a mixture of soluble and insoluble organic matter, just as the activated sludge process is. Two groups of microorganisms are involved. Acidogenic bacteria are responsible for the conversion of the influent organic matter into acetic acid, molecular hydrogen, and carbon dioxide. Other short chain volatile fatty acids may accumulate, as will a stable insoluble residue similar to humus. Methanogenic bacteria are responsible for the conversion of the acetic acid, hydrogen, and carbon dioxide to methane gas. High-rate suspended growth anaerobic processes are well suited as a pretreatment method for wastes containing more than 4000 mg/L of biodegradable COD, but less than 50,000 mg/L, because they are less expensive than either activated sludge or evaporation.[3] Their main advantages over activated sludge systems are lower power requirements, less production of excess solids, and the generation of methane gas. Further treatment is often required for the effluent from these processes, however, because many aerobically biodegradable soluble products remain.

The upflow anaerobic sludge blanket (UASB) reactor is distinguished by the absence of an external sedimentation chamber. Instead, the wastewater is introduced at the bottom of the reactor and flows upward at a velocity that matches the settling velocity of the biomass. In this way a sludge blanket is formed and maintained, as illustrated in Figure 1.20. A special zone is required to allow the gas formed to escape without carrying sludge particles with it. The biomass in these reactors is in the form of compact granules that contain mixed cultures of methanogenic and acidogenic bacteria.[7] Because of the good retention of biomass in UASBs, they are suitable for treating wastewaters with relatively low substrate concentrations. In fact, they have been demonstrated to be capable of effective treatment of municipal wastewater.[8]

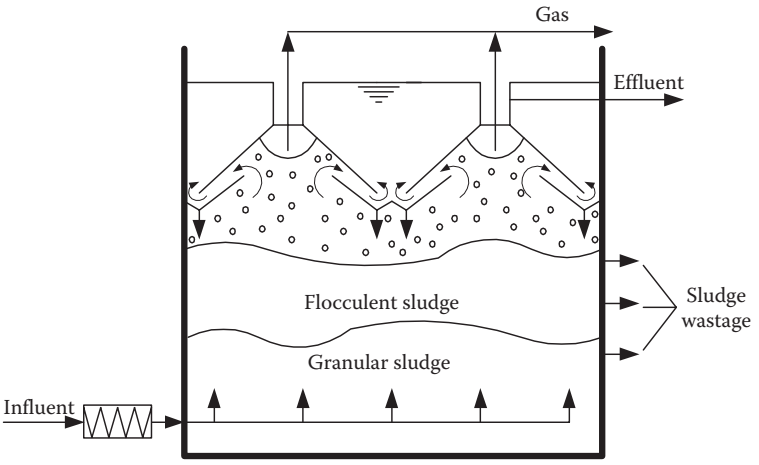

FIGURE 1.20 Upflow anaerobic sludge blanket (UASB) bioreactor.

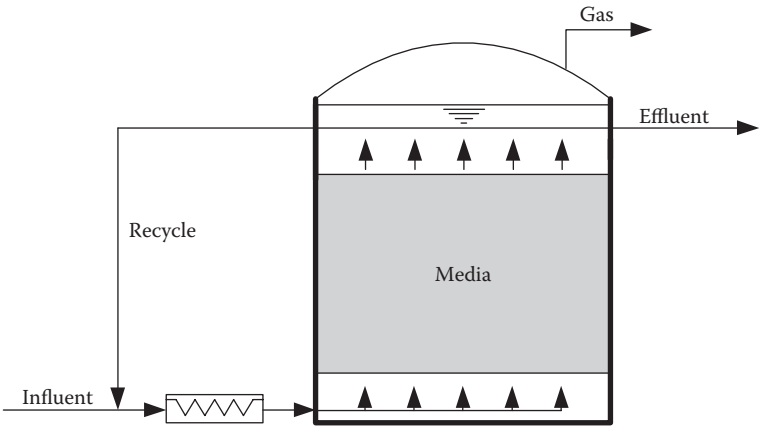

FIGURE 1.21 Anaerobic filter (AF).

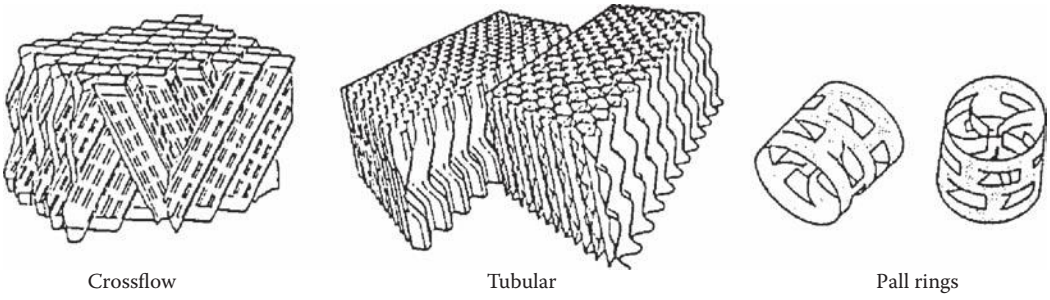

FIGURE 1.22 Typical media used in anaerobic filters and packed towers. (From J. C. Young, Factors affecting the design and performance of upflow anaerobic filters. *Water Science and Technology,* 24 (8): 133–56, 1991. Copyright © IWA Publishing; reprinted with permission.)

Another high-rate suspended growth anaerobic process is the anaerobic filter (AF). As illustrated in Figure 1.21, it consists of a reactor filled with media through which wastewater (generally) flows from bottom to top. The name suggests that the AF might be an attached growth, rather than a suspended growth, process. This is not correct, however, as the media is relatively open, as illustrated in Figure 1.22[12], and functions essentially like a tube settler to retain suspended biomass. It can also retain particulate matter contained in the influent wastewater better than a UASB.

Because the UASB and AF possess complementary advantages, they have been combined in the hybrid UASB/AF process. As illustrated in Figure 1.23, the influent wastewater first passes through the granular sludge blanket and then upward to the top of the reactor where the media is placed.

The theory of high-rate suspended growth anaerobic processes is discussed in Chapter 8 and their design is discussed in Chapter 14.

1.3.1.5 Anaerobic Digestion

By far the largest use of anaerobic cultures is in the stabilization of insoluble organic matter by anaerobic digestion (AD), which involves microbial communities similar to those found in high-rate suspended growth anaerobic processes. Anaerobic digestion is one of the oldest forms of wastewater treatment, yet because of the complex ecosystem involved it has continued to be the subject of research and new process development. Designers have historically favored the use of CSTRs (as illustrated in Figure 1.24) because of their uniform environmental conditions, and some utilize CSTRs with solids recycle because smaller bioreactors can be used. However, various configurations

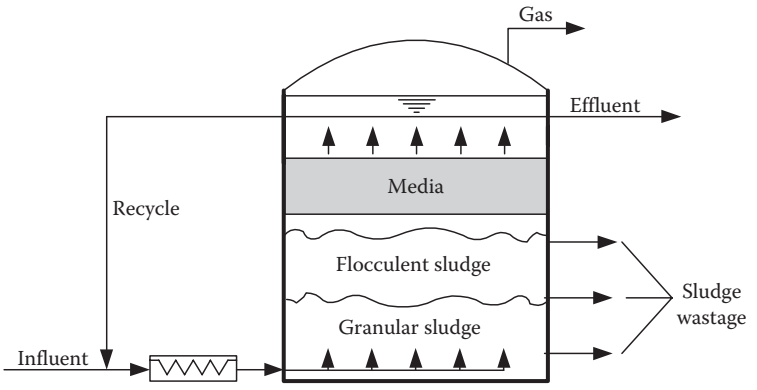

FIGURE 1.23 Hybrid UASB/AF process.

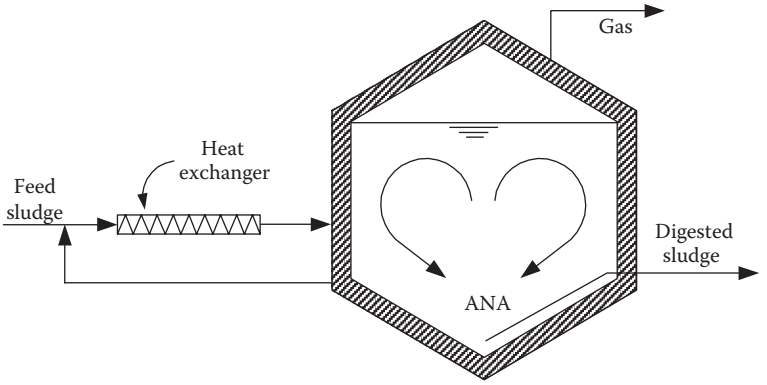

FIGURE 1.24 Anaerobic digestion (AD) process.

using CSTRs in series are being evaluated to increase the stabilization of biodegradable organic matter and pathogen destruction. Anaerobic digestion is discussed in Chapters 8 and 14.

1.3.1.6 Fermenters

Biological phosphorus removal systems require volatile fatty acids (VFAs) as a feed component. When wastewaters contain insufficient VFAs to remove the influent phosphorus, VFAs must be manufactured in the treatment process, or they must be purchased and added. The VFAs are among the intermediates formed in the anaerobic digestion of particulate organic matter. Use has been made of this fact to develop processes to produce VFAs from primary sludge (Figure 1.1) and elutriate them for addition to a BNR process.

Figure 1.25 illustrates a typical solids fermentation process. It consists of a fermentation bioreactor operated at a short SRT to allow hydrolysis and acidification of particulate organic matter while preventing the subsequent conversion of the VFAs to methane, followed by a tank in which liquid containing the VFAs is separated from the residual solids. The VFA-rich liquid stream is subsequently added to a BNR process. Fermenters are discussed in Chapters 8 and 14.

The various anaerobic processes have a long and interesting history, similar to that of the activated sludge process. Readers wanting to learn more about this history should consult McCarty.[5]

1.3.1.7 Lagoons

The term lagoon refers to suspended growth bioreactors that do not include biomass recycle from a downstream liquid:solid separator. Their name comes from their construction and appearance,

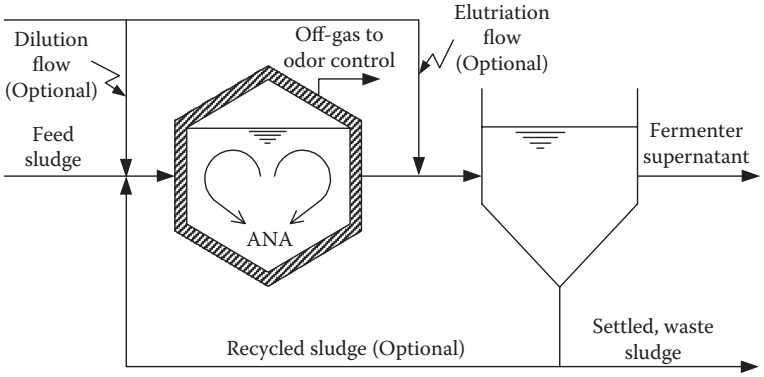

FIGURE 1.25 Solids fermentation process for the production of volatile fatty acids.

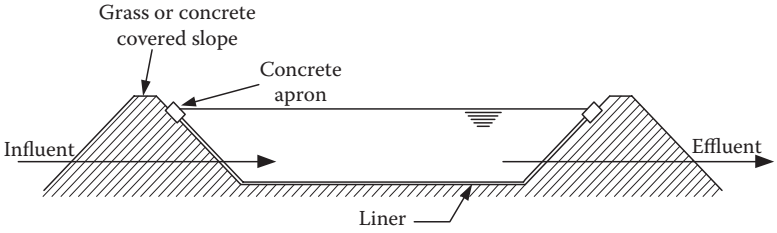

FIGURE 1.26 Schematic diagram of a lagoon (vertical dimension exaggerated).

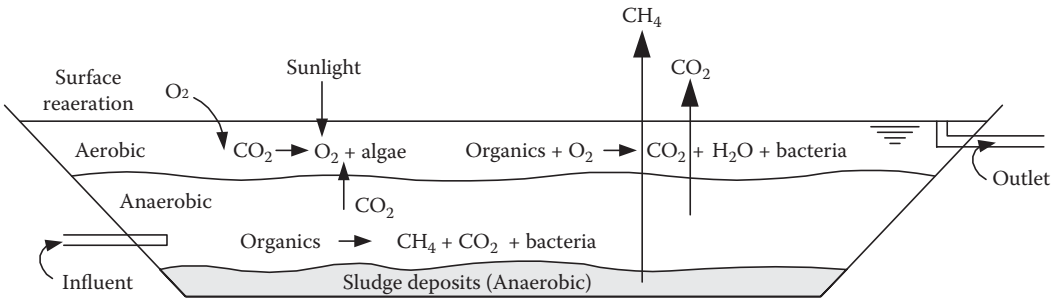

FIGURE 1.27 Facultative/aerated lagoon (F/AL; vertical dimension exaggerated).

illustrated in Figure 1.26. Historically, they have been constructed as large earthen basins that, because of their size, resemble typical "South Sea island lagoons." Originally, lagoons were not lined, but this has proven to be unacceptable because of the potential for leakage of the basin contents into groundwater. Consequently, current design practice requires them to be lined with an impermeable liner. A wide range of environmental conditions can exist in lagoons, depending on the degree of mixing imposed. If the lagoon is well mixed and aerated, it can be aerobic throughout, but with lesser degrees of mixing, solids will settle, leading to anoxic and anaerobic zones.

Three types of lagoons are characterized in Table 1.2. Completely mixed aerated lagoons (CMALs) can generally be classified as completely mixed reactors that are used for the removal of soluble organic matter, although stabilization of insoluble organic matter and nitrification can also occur. Facultative/aerated lagoons (F/ALs), illustrated in Figure 1.27, are mixed but not sufficiently to keep all solids in suspension. As a consequence, the upper regions tend to be aerobic whereas the

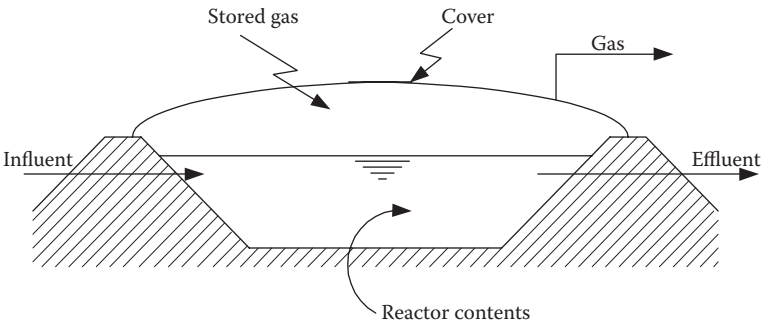

FIGURE 1.28 Anaerobic lagoon (ANL).

bottom contains anaerobic sediments. Anaerobic lagoons (ANLs), illustrated in Figure 1.28, are not purposefully mixed. Rather, any mixing that occurs is the result of gas evolution within them.

Lagoons represent one of the oldest forms of biological wastewater treatment, having been used in some form for more than 3000 years.[9] They have been used as the only means of treatment prior to discharge to surface waters and for pretreatment and/or storage prior to treatment in a conventional system or a wetland. A wide range of industrial and municipal wastewaters has been treated in lagoon systems. Each of the lagoon types is discussed in Chapter 15.

1.3.2 Attached Growth Bioreactors

1.3.2.1 Fluidized Bed Biological Reactors

Fluidized bed biological reactors (FBBRs) come under the broad category of submerged attached growth bioreactors. They can be operated with any of the three biochemical environments, and the nature of that environment determines what the bioreactor accomplishes. Fluidized bed systems for denitrification were among the earliest developed because all materials to be reacted were present in a soluble state. However, through the use of pure oxygen as a means of providing dissolved oxygen at high concentration, aerobic fluidized beds soon followed. Their chief purpose is removal of soluble organic matter, but they are also used for nitrification. Finally, anaerobic fluidized bed systems were developed for the treatment of soluble wastewaters.[7] The key characteristic of fluidized bed systems is their ability to retain very high biomass concentrations, thereby allowing small bioreactor volumes to be used. This is accomplished by using very small particles, which provide a large surface area per unit volume, as the attachment media for biofilm growth. Media frequently used include silica sand with a diameter of 0.3 to 0.7 mm and granular activated carbon with a diameter of 0.6 to 1.4 mm. Maintenance of the particles in a fluidized state by control of the upflow velocity ensures better mass transfer characteristics than can be achieved in other attached growth systems. Figure 1.29 provides a schematic of the process illustrating that the necessary upflow velocity is provided by a combination of influent and recirculation flows. Biomass accumulation is controlled by removing media from the top of the fluidized bed where the largest amount of biomass accumulates, passing it through a pump where biomass is sheared off, and then sending it to a separation device where the cleaned media is separated from the biomass. The major use of FBBRs has been for industrial wastewater treatment, although they have also been used to denitrify municipal wastewater. This type of attached growth bioreactor is discussed in Chapters 18 and 21.

1.3.2.2 Rotating Biological Contactor (RBC)

The rotating biological contactor is a modern application of an old idea for the removal of soluble organic matter and the conversion of ammonia to nitrate. Microorganisms growing attached to rotating discs, as illustrated in Figure 1.30, accomplish the desired objectives by the same mechanisms used in suspended growth systems, but in a more energy efficient manner because oxygen

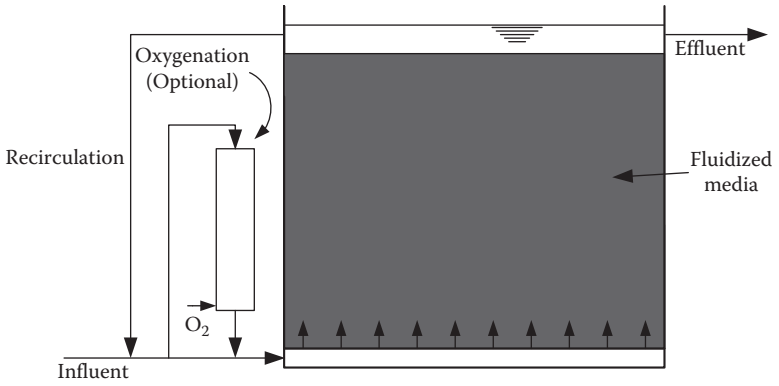

FIGURE 1.29 Fluidized bed biological reactor (FBBR).

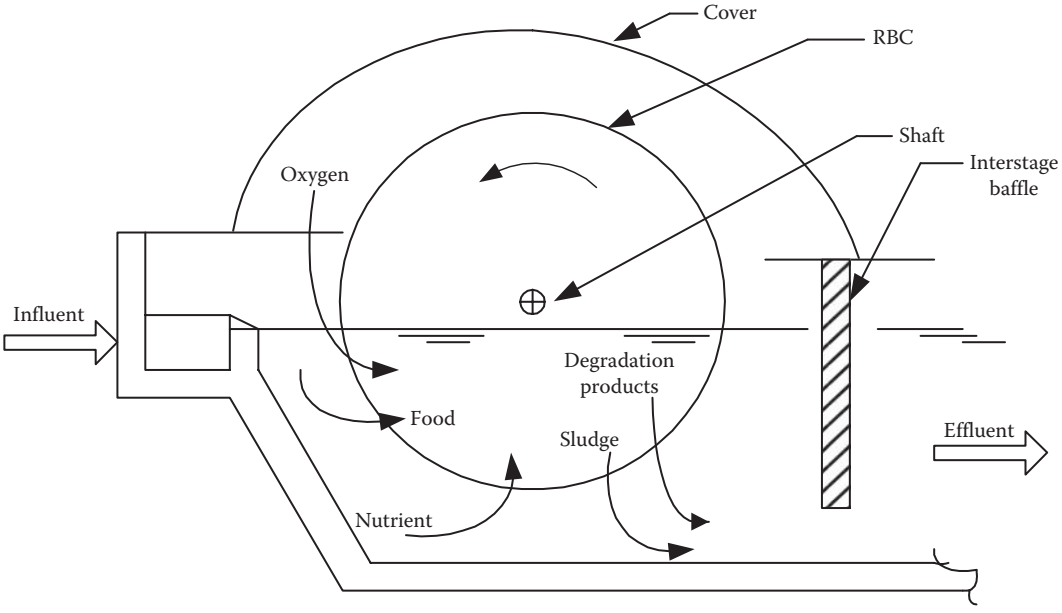

FIGURE 1.30 Schematic diagram of a rotating biological contactor (RBC).

transfer is accomplished by the rotation of the discs, which are only half submerged. The media is similar to the corrugated plastic sheet media used in AFs, as illustrated in Figure 1.22, and in trickling filters (described below). The RBC units are generally arranged in trains to provide stages to increase treatment efficiency. These bioreactors have been popular for the treatment of both domestic and industrial wastewaters, typically at smaller installations. The RBCs are discussed in Chapters 17 and 20.

1.3.2.3 Trickling Filter (TF)

As indicated in Table 1.2, trickling filter is the name given to an aerobic attached growth bioreactor in the shape of a packed tower. One of the first biochemical operations developed, initial experimentation with the use of gravel beds for wastewater treatment occurred at the Lawrence Experiment Station in Massachusetts in 1889.[11] This was followed by research in England in the late 1890s and early 1900s, research in the United States in the early 1900s, and initial full-scale applications in the United States in the late 1900s and early 1910s. The popularity of the

trickling filter increased throughout the first half of the twentieth century until, in the 1950s, it was the most popular biochemical operation in the United States. Then, during the 1960s and 1970s, the popularity of the activated sludge process increased due to better economic and performance characteristics.

Until the mid-1960s trickling filters were made of stone, which limited their height to around two meters for structural reasons. Now trickling filters are made of plastic media much like that used as packing in absorption and cooling towers (Figure 1.22) and are self-supporting to heights of around seven meters because of the greater void space and lighter weight of the media. These new media, coupled with improved process configurations and increased understanding of biofilm processes, have resulted in improved trickling filter economics and performance, causing a resurgence in use.[10,11] The primary use of trickling filters is for removal of soluble organic matter and oxidation of ammonia to nitrate. Traditionally, trickling filters have been used for municipal wastewater treatment in small to medium size installations desiring minimal operating expense. However, since the introduction of plastic media they have also found use as pretreatment devices preceding other biochemical operations. This is because they have the ability to reduce the waste concentration at relatively low operating cost, a bonus when aerobic treatment is being employed. Trickling filters cause relatively little degradation of insoluble organic matter and should not be used for that purpose.

Figure 1.31 presents a schematic of the trickling filter process illustrating its major components including: the media bed, the containment structure, the wastewater application (or dosing) system, the underdrain, and the ventilation system. Wastewater is applied to the top of the media, to which the biomass is attached and flows down over it, thereby allowing biological treatment to occur. The open structure of the media allows air to flow through the trickling filter, providing needed oxygen. Trickling filters are covered in Chapters 17 and 19.

1.3.2.4 Packed Bed

Packed bed bioreactors fall within the broad category of submerged attached growth bioreactors. They are a recently developed biochemical operation that utilizes submerged media with a particle size on the order of a few millimeters. Configured much like a granular media filter, as illustrated in Figure 1.32 they are designed and operated with flow either upward or downward. Several types of media are used, including rounded sand, fired clay, and plastic. Because of the small particle size, packed beds act as physical filters, thereby providing removal of particulate matter. Packed beds are used to oxidize organic matter, both soluble and particulate, and for conversion of soluble inorganic

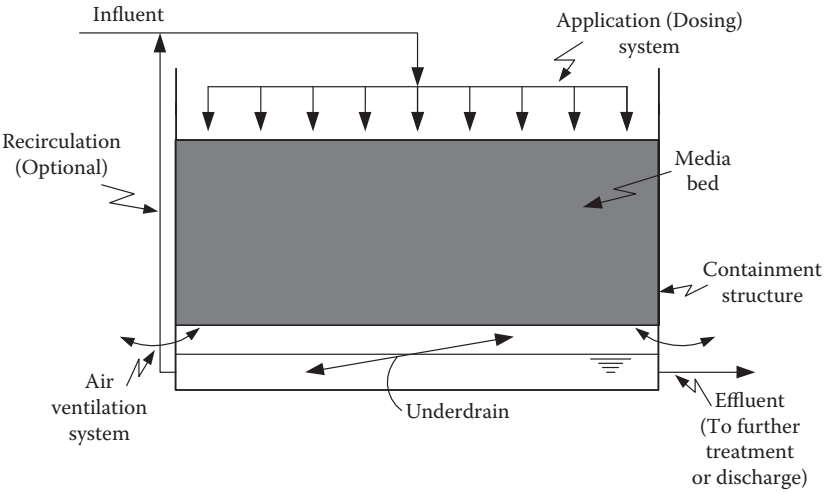

FIGURE 1.31 Schematic diagram of a trickling filter (TF).

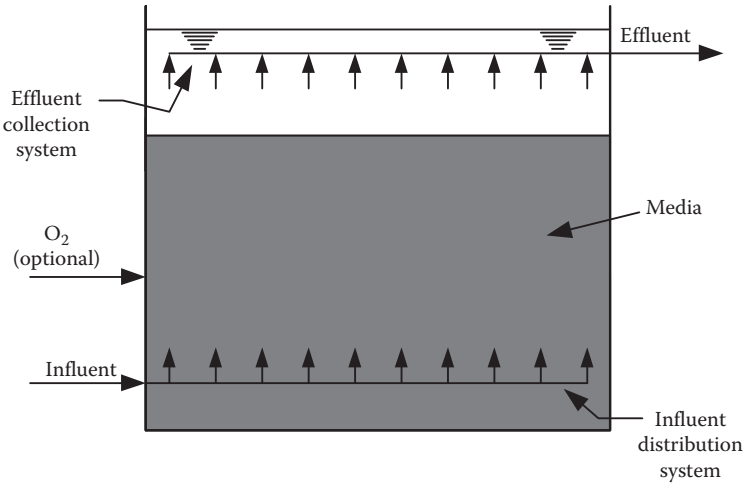

FIGURE 1.32 Submerged attached growth bioreactor (SAGB). Example shown is an upflow packed bed (UFPB) bioreactor; reverse direction of arrows for a downflow packed bed (DFPB) bioreactor.

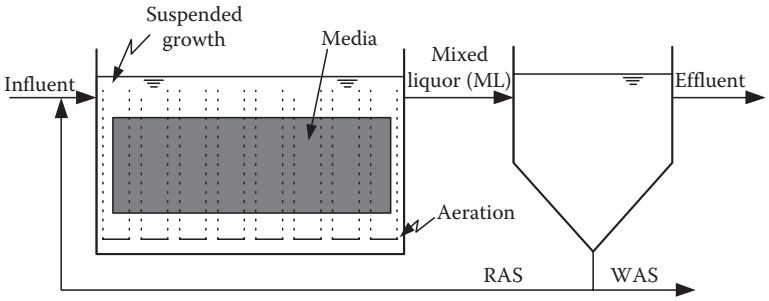

FIGURE 1.33 Integrated fixed film activated sludge (IFAS) process.

matter, particularly nitrification and denitrification. The bed is sparged with air when aerobic conditions are desired. When used for denitrification the bed is not sparged (except to dislodge accumulated nitrogen gas and for backwashing), and supplemental carbon (such as methanol) is often needed. They are discussed in Chapter 21.

1.3.2.5 Integrated Fixed Film Activated Sludge Systems

Integrated fixed film activated sludge (IFAS) systems, illustrated in Figure 1.33, are a recent development within which media of various types are added to an activated sludge process. Media can be fixed in place (such as sheet plastic trickling filter media) or be free to circulate in the bioreactor. Suspended biomass settling in the downstream clarifier can be recycled to build up a suspended biomass within the bioreactor and, when this is done, both the attached and suspended biomass contribute to wastewater treatment. Sections of the bioreactor can be aerated to create aerobic conditions, while others can be mixed and the process flow can be recirculated to create anaerobic or anoxic conditions. Thus, these systems are used for the removal of soluble and particulate organic matter (like the activated sludge processes) and of inorganic matter, including ammonia, nitrogen, and phosphorus (like BNR). Empirical knowledge developed over the past three decades allows these systems to be used for a variety of applications. However, the interaction of the attached biomass with the suspended biomass is still poorly characterized and represents an important research area for this technology. In some cases suspended biomass is not recirculated from the downstream

clarifier, resulting in a process that contains largely attached biomass—such systems are referred to as moving bed biological reactor (MBBR) systems. Integrated fixed film activated sludge systems are discussed in Chapter 21.

1.3.3 Miscellaneous Operations

There are many other biochemical operations in use or in development. Even though these systems are not listed in Tables 1.1 and 1.2, some are covered at appropriate points. However, many other new biochemical operations are not included in this book due to space constraints. Their exclusion should not be construed as a bias against their use. Rather, it was felt that once the fundamental principles of biochemical operations are learned, the reader will be able to apply them to understand and evaluate any biochemical wastewater treatment system.

1.4 KEY POINTS

1. Biochemical operations may be carried out in aerobic, anoxic, or anaerobic environments, and the choice of environment has a profound effect on both the ecology of the microbial community and the outcome of its activity.
2. Three major biochemical transformations may be performed with biochemical operations: removal of soluble organic matter, stabilization of insoluble organic matter, and conversion of soluble inorganic matter.
3. Bioreactors for biochemical operations may be divided into two major categories, depending on the manner in which the microorganisms grow: suspended in the wastewater undergoing treatment or attached to a solid support.
4. The major suspended growth bioreactors are: continuous stirred tank reactor, either alone or in series; batch reactor; and plug-flow reactor. The major attached growth bioreactors are: fluidized bed, packed tower, and rotating disc reactor.
5. The major suspended growth biochemical operations are: activated sludge, biological nutrient removal, aerobic digestion, high-rate suspended growth anaerobic processes, anaerobic digestion, fermenters, and lagoons. The major attached growth biochemical operations are: fluidized bed biological reactor, rotating biological contactor, trickling filter, packed bed, and integrated fixed film activated sludge systems.

1.5 STUDY QUESTIONS

1. List and define the three major biochemical transformations that may be performed with biochemical operations.
2. Describe each of the major bioreactor types that find use in biochemical operations.
3. List the 12 named biochemical operations and tell whether each uses a suspended or an attached growth culture.
4. Describe each of the named biochemical operations in terms of the biochemical transformation involved, the reaction environment used, and the bioreactor configuration employed. Include in your description a sketch of a typical process flow diagram.

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2 Fundamentals of Biochemical Operations

Before we begin the systematic study of biochemical operations it is necessary to develop a clear picture of what wastewater treatment engineers hope to accomplish through their use. Furthermore, if we are to develop the capability for their design, it is necessary to understand what is happening within them and to recognize the role of various types of microorganisms in those events.

2.1 OVERVIEW OF BIOCHEMICAL OPERATIONS

Biochemical operations only alter and destroy materials that microorganisms act upon; that is, those that are subject to biodegradation or biotransformation. If soluble pollutants are resistant to microbial attack, they are discharged from a biochemical operation in the same concentration that they enter it, unless they are acted on by chemical or physical mechanisms such as sorption or volatilization, as discussed in Chapter 22. Insoluble pollutants entering a suspended growth biochemical operation become intermixed with the biomass and, for all practical purposes, are inseparable from it. Consequently, engineers consider this mixture of biomass and insoluble pollutants as an entity, calling it mixed liquor suspended solids (MLSS), which follows from referring to the mixture of MLSS and wastewater undergoing treatment as "mixed liquor." If insoluble pollutants are biodegradable, their mass is reduced. On the other hand, if they are nonbiodegradable, their only means of escape from the system is through MLSS wastage and their mass discharge rate in the wasted MLSS must equal their mass input rate to the system. Attached growth processes usually have little impact on nonbiodegradable insoluble pollutants, although in some cases those pollutants are flocculated and settled along with the biomass discharged from the operation.

When wastewater treatment engineers design biochemical operations they use natural cycles to accomplish in a short time what nature would require a long time to accomplish, often with environmental damage. For example, as discussed in Chapter 1, if biodegradable organic matter were discharged to a stream, the bacteria in that stream would use it as a source of carbon and energy (electrons) for growth. In the process, they would incorporate part of the carbon into new cell material and the rest would be oxidized to carbon dioxide to provide the energy for that synthesis. The electrons removed during the oxidation would be transferred to oxygen in the stream, but if the supply of oxygen were insufficient, the dissolved oxygen (DO) concentration would be depleted, killing fish and causing other adverse effects. On the other hand, in a well-designed biochemical operation, microbial growth is allowed to occur in an environment where the appropriate amount of oxygen can be supplied, thereby destroying the organic matter and allowing the treated wastewater to be discharged without environmental harm.

The two major cycles employed in biochemical operations are the carbon and nitrogen cycles. Actually, most biochemical operations only use half of the carbon cycle (i.e., the oxidation of organic carbon) releasing carbon dioxide. While some biochemical operations use algae and plants to fix carbon dioxide and release oxygen, thereby using the other half of the carbon cycle, they are not as widely applied and will not be covered in this book. Almost all of the nitrogen cycle is used, however. It is illustrated in Figure 2.1. In domestic wastewaters, most nitrogen is in the form of ammonia (NH_3) and organic nitrogen, whereas industrial wastewaters sometimes contain nitrate (NO_3^-) nitrogen as well. Organic nitrogen is in the form of amino groups (NH_2^-), which are released as

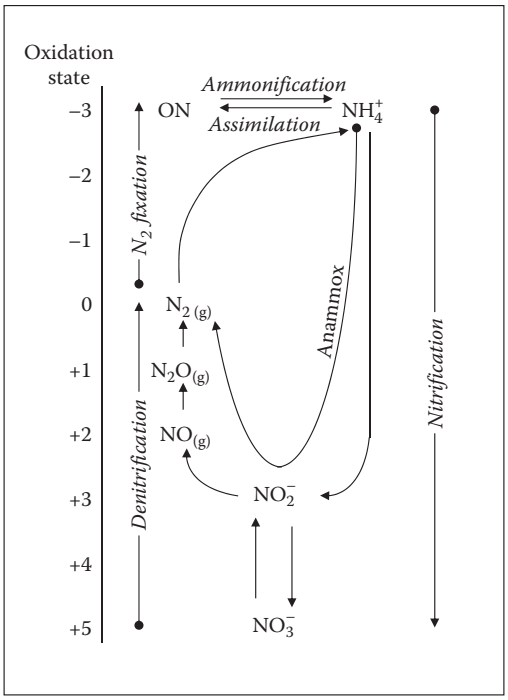

FIGURE 2.1 The nitrogen cycle. Nitrogen species are aligned with their oxidation state. Gaseous species are designated with (g). Metabolic processes are italicized. (From Water Environment Federation, Nutrients and their effect on the environment, *Nutrient Removal*, Manual of Practice No. 34, McGraw Hill, New York, 2011. With permission.)

ammonia, in the process called ammonification, as the organic matter containing them undergoes biodegradation. The form in which bacteria incorporate nitrogen during growth is as ammonia. If an industrial wastewater has insufficient ammonia or organic nitrogen to meet the growth needs of the bacteria, but contains nitrate or nitrite (NO_2^-) nitrogen, they will be converted to ammonia through assimilative reduction for use in cell synthesis. On the other hand, if a wastewater contains ammonia-N in excess of that needed for cell synthesis, nitrification can occur, in which the excess ammonia-N is oxidized to nitrate-N, going through the intermediate, nitrite. Discharge of nitrate to a receiving water is preferable to discharge of ammonia because nitrification in the receiving water can deplete the DO, just as degradation of organic matter can. In some cases, however, the discharge of nitrate can have a deleterious effect on the receiving water, and thus some effluent standards limit its concentration. In that case, biochemical operations that use denitrification to convert nitrate and nitrite to gaseous end products must be used to reduce the amount of soluble nitrogen in the effluent. More recently, *an*aerobic *amm*onia *ox*idation (anammox) has been found to be an important contributor to the global nitrogen cycle[55] and has important implications for treating high ammonia-laden waste streams.[123] This process also involves oxidation of ammonia but uses nitrite as the electron acceptor to produce nitrogen gas and nitrate. The only step in the nitrogen cycle not normally found in biochemical operations is nitrogen fixation, in which nitrogen gas is converted to a form that can be used by plants, animals, and microorganisms.

2.2 MAJOR TYPES OF MICROORGANISMS AND THEIR ROLES

Modern molecular biology has allowed scientists to investigate the relatedness among organisms by analysis of the nucleotide sequences within certain segments of their genes. Organization of this information into a phylogenetic tree has revealed that organisms fall into three primary groupings,

or domains: Archaea, Bacteria, and Eucarya.[128] Members of the domains Archaea and Bacteria are microscopic and procaryotic (i.e., they lack a nuclear membrane), whereas members of the domain Eucarya are eucaryotic (i.e., they have a nuclear membrane) and vary in size from microscopic (e.g., protozoa) to macroscopic (e.g., animals). The workhorses of biochemical operations belong to the domains Bacteria and Archaea, but protozoa and other microscopic Eucarya have a role as well. Thus it is important to have a clear picture of what various microorganisms do.

2.2.1 Bacteria

Bacteria can be classified in many ways; however, the most important from an engineering perspective is operational. Consequently, we will focus on it.

Like all organisms, members of the domain Bacteria derive energy and reducing power from oxidation reactions, which involve the removal of electrons. Thus, the nature of the electron donor is an important criterion for their classification. The two sources of electrons of most importance in biochemical operations are organic and inorganic compounds that are present in the wastewater or released during treatment. Bacteria that use organic compounds as their electron donor and their source of carbon for cell synthesis are called chemoheterotrophic bacteria, or simply heterotrophs. Since the removal and stabilization of organic matter are the most important uses of biochemical operations, it follows that heterotrophic bacteria predominate in the systems. Bacteria that use inorganic compounds as their electron donor and carbon dioxide as their source of carbon are chemoautotrophic bacteria, although most wastewater treatment engineers call them autotrophic bacteria or simply autotrophs. The most commonly encountered autotrophic bacteria in biochemical operations are those that use ammonia-N and nitrite-N as electron donors. Together, they are responsible for nitrification and are referred to as nitrifiers. Other autotrophic bacteria, such as anammox bacteria, are of increasing interest because of their utility in treating high strength, ammonia-laden wastewaters, such as recycle waters generated during biosolids dewatering processes.

Another important characteristic of bacteria is the type of electron acceptor they use. The most important acceptor in biochemical operations is oxygen, and bacteria that use only it are called obligately aerobic bacteria or simply obligate aerobes. Nitrifying bacteria are the most significant obligately aerobic bacteria commonly found in biochemical operations. At the other end of the spectrum are obligately anaerobic bacteria, which can only function in the absence of molecular oxygen. Between the two obligate extremes are the facultative bacteria, which use oxygen as their electron acceptor when it is present in sufficient quantity but shift to an alternative acceptor in its absence. Because the environment within flocs and biofilms in biochemical operations often varies from aerobic to anaerobic extremes, facultative bacteria tend to predominate in these systems. Some facultative bacteria are fermentative, meaning that they use organic compounds as their alternative terminal electron acceptor in the absence of oxygen, producing reduced organic end products. Others perform anaerobic respiration, in which an inorganic compound serves as the alternative acceptor. In Chapter 1, mention was made of anoxic environments in which oxygen is absent, but nitrate is present as an electron acceptor. Because of the prevalence of such environments in biochemical operations, the most significant facultative bacteria are those that perform denitrification (i.e., reduce nitrate-N or nitrite-N to nitrogen gas). Other facultative and obligately anaerobic bacteria reduce other inorganic compounds, but with the exception of protons (H^+), most are not of general importance in biochemical operations, although increasing use is being made of them for specialized needs.[115] Proton reduction, which occurs in anaerobic operations, yields hydrogen gas (H_2), which is an important electron donor for methane formation.

Gravity sedimentation is the most common method for removing biomass from the effluent from biochemical operations prior to its discharge. Since single bacteria are so small (~0.5–1.0 μm), it would be impossible to remove them by gravity if they grew individually. Fortunately, under the proper growth conditions, bacteria in suspended growth cultures grow in clumps or gelatinous assemblages called biofloc, which range in size from 0.05 to 1.0 mm.[97] Figure 2.2a shows a typical

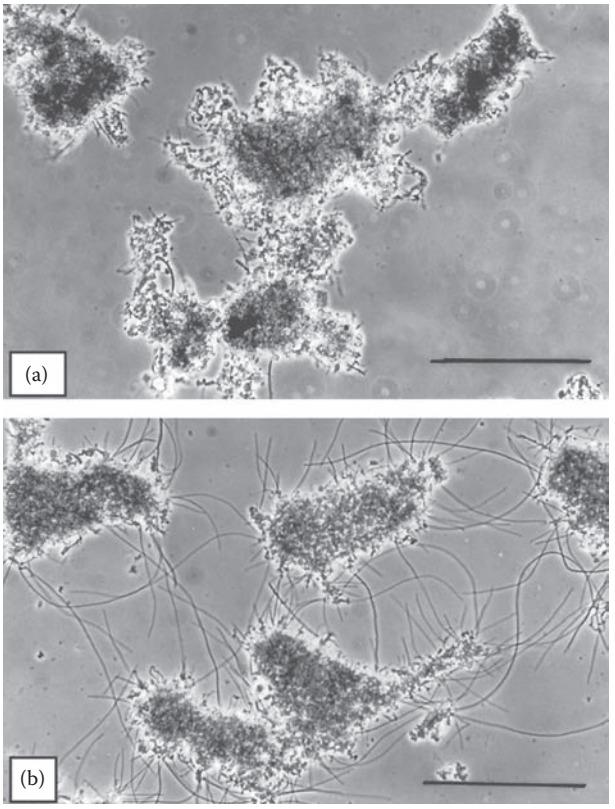

FIGURE 2.2 Photomicrographs of activated sludge floc: (a) Good settling biomass with optimal filaments; (b) poor settling biomass with excessive filaments. (Courtesy of M. G. Richard, Michael Richard Wastewater Microbiology LLC, Fort Collins, Colorado and David Jenkins, University of California, Berkeley.)

floc particle. The bacteria that are primarily responsible for this are called floc-forming bacteria, and a variety of species fall into this category.

Not all bacteria are beneficial in biochemical operations; some are a nuisance. In aerobic/anoxic systems, filamentous bacteria grow as long strands, which become intermeshed with biofloc particles and interfere with sedimentation. Although a small number of filaments can provide strength for the biofloc preventing its disruption by fluid shear forces, too many can act to hold the biofloc particles apart,[114] as shown in Figure 2.2b. When that occurs, sedimentation is very inefficient and the biomass will not compact into a sufficiently small volume to allow discharge of a clear effluent. Another type of nuisance bacteria forms copious quantities of foam in bioreactors that are being aerated for oxygen transfer. The foam can become so deep as to completely cover aeration and sedimentation basins, thereby disrupting treatment and posing a danger to plant personnel. In biological phosphorus removal (BPR) systems glycogen accumulating organisms (GAOs) often compete with phosphate accumulating organisms (PAOs) for the electron donor, thereby decreasing the efficiency of phosphorus removal. The most common nuisance organisms in anaerobic systems are the sulfate reducing bacteria. It is generally desirable to design anaerobic operations to produce methane because it is a valuable product. If a wastewater contains high concentrations of sulfate, however, sulfate reducing bacteria will compete for the electron donor, producing sulfide as a product. This not only decreases the amount of methane produced, but results in a product that is both dangerous and undesirable in most situations. Wastewater treatment engineers need to be aware of the growth characteristics of such nuisance organisms so that systems that discourage or prevent their growth can be designed.

Bacteria can also be classified according to their function in biochemical operations. Many act as primary degraders and attack the organic compounds present in the wastewater, beginning their degradation. If an organic compound is one normally found in nature (biogenic), the primary degraders will usually completely metabolize it in an aerobic environment, converting it to carbon dioxide, water, and new biomass. Such ultimate destruction is called mineralization and is the goal of most wastewater treatment systems. On the other hand, if an organic compound is synthetic and foreign to the biosphere (xenobiotic), it is possible that no single type of bacteria will be able to mineralize it. Instead, a microbial consortium may be required, with secondary degraders living on the metabolic products excreted by the primary degraders. The more complex the organic compounds found in a wastewater, the more important secondary degraders will be. Secondary degraders are common in anaerobic environments, however, even when biogenic compounds are being degraded because of the specialized needs of the bacteria involved. Other functions that are important in wastewater treatment systems are the production and elimination of nitrate-N through nitrification and denitrification, respectively. Consequently, it is not surprising that bacteria are classified according to those functions, as nitrifiers and denitrifiers. While the nitrifiers constitute a highly specialized group containing a limited number of species of aerobic, chemoautotrophic bacteria, the denitrifying bacteria constitute a diverse group of facultative heterotrophic bacteria containing many species. Finally, PAOs have the ability to store and release phosphate in response to cyclical environmental conditions. Because of this ability, they can contain quantities of phosphate well in excess of other bacteria.

As with the classification of pollutants in wastewaters, the classifications listed above are not exclusive, but overlap, with members of the domain Bacteria playing many roles. Nevertheless, these simple classification schemes are very helpful in describing the events occurring in biochemical operations and will be used throughout this book.

2.2.2 Archaea

Many Archaea are capable of growing in extreme environments, such as high temperatures (up to 90°C), high ionic strength, and highly reduced conditions. Members of this domain were first thought to be restricted to growth in such environments, but are now known to be abundantly distributed in a wide variety of environments[25] and may even play an important role in nitrification.[93] As our knowledge of the Archaea expands, it is likely that wastewater treatment engineers will find more applications for them. Currently, however, their major use in biological wastewater treatment is in anaerobic operations, where they play the important role of producing methane. Methane-producing Archaea, commonly called methanogens, are obligate anaerobes that bring about the removal of organic matter from the liquid phase by producing an energy rich gas of low solubility. This allows capture of the energy in the pollutants in a useful form. Because methanogens are very limited in the electron donors they can use, they grow in complex microbial communities with Bacteria, which carry out the initial attack on the pollutants and release the methanogens' electron donors as fermentation products.

2.2.3 Eucarya

Although fungi can use soluble organic matter in competition with Bacteria, they seldom compete well in suspended growth cultures under normal conditions, and thus do not usually constitute a significant proportion of the microbial community.[35] On the other hand, when the supplies of oxygen and nitrogen are insufficient or when the pH is low, fungi can proliferate, causing problems similar to those caused by filamentous bacteria. In contrast to suspended growth cultures, fungi commonly play an important role in attached growth cultures, making up a large part of the biomass.[121] Under certain conditions, however, they can also become a nuisance in such systems by growing so heavily as to block interstices and impede flow.

Protozoa play an important role in suspended growth cultures by grazing on colloidal organic matter and dispersed bacteria, thereby reducing the turbidity remaining after the biofloc has been removed by sedimentation.[16,63] Protozoa are also known to contribute to bioflocculation, but their contribution is thought to be less important than that of the floc-forming bacteria.[15] Although some protozoa can utilize soluble organic compounds for growth, it is doubtful that they can compete effectively with bacteria in that role and thus soluble organic compound removal is generally considered to be due to bacterial action. Despite this, they have been shown to serve as reliable indicators of process performance[68,101] and heavy metal contamination.[69] Protozoa also play a significant role in attached growth bioreactors where the protozoan community is usually richer than it is in suspended growth cultures. Nevertheless, their role appears to be similar to that in suspended growth cultures.

Other Eucarya in suspended growth cultures are usually limited to metazoa, like rotifers and nematodes, but their presence depends very much on the way in which the culture is grown and possibly the nature of the wastewater.[101] Although metazoa feed upon protozoa and biofloc particles, their contribution to biochemical operations using suspended growth cultures is largely unknown because little change in process performance can be attributed to their presence. In contrast, because attached growth bioreactors provide a surface upon which higher organisms can graze, it is not uncommon for such reactors to have highly developed communities of macroinvertebrates in addition to rotifers and nematodes.[35] The nature of those communities depends largely on the physical characteristics of the bioreactor and in some cases the presence of the higher community has no deleterious effect on system performance. In other cases, however, the grazing community can disrupt development of the primary biofilm that is responsible for the removal of the pollutants, leading to deterioration in system performance.

2.3 MICROBIAL ECOSYSTEMS IN BIOCHEMICAL OPERATIONS

An ecosystem is the sum of the interacting elements (both biological and environmental) in a limited universe. Consequently, each biochemical operation will develop a unique ecosystem governed by the physical design of the facility, the chemical nature of the wastewater going to it, and the biochemical changes wrought by the resident organisms. The microbial community that develops in that ecosystem will be unique from the viewpoint of species diversity, being the result of physiological, genetic, and social adaptation. Thus it is impossible to generalize about the numbers and types of species that will be present. Nevertheless, it would be instructive to consider the general nature of the community structures in biochemical operations and relate them to the environments in which the operations are performed. The objective of such an exercise is not the simple listing of the organisms present, but rather an understanding of the role that each important group plays in the operation. Indeed, the biochemical processes in aerobic and anoxic environments are quite different from those in anaerobic environments. Thus, the biochemical environment provides a logical way for dividing this discussion. Before beginning that discussion, however, we will briefly consider aggregation because it plays such an important role in many biochemical operations.

2.3.1 Aggregation and Bioflocculation

Although microbiologists have generated a large body of knowledge about microbial life from the study of pure microbial cultures grown in liquid suspension, the vast majority of microbial life on Earth exists as microbial communities growing as aggregates.[24] These aggregates take the form of biofilms or floc (planktonic biofilms) held together by extracellular polymeric substances (EPSs). Long before the general ubiquity of microbial aggregates was recognized, environmental engineers exploited them to retain the mixed microbial communities that are central to biological wastewater treatment. Attached growth processes rely upon biofilms attached to some form of solid support for retention of their microbial communities. Suspended growth processes, on the other hand,

retain their microbial communities in suspension by the generation of floc particles capable of being removed by gravity sedimentation or membrane processes and recycled to the bioreactor. Although from a macroscopic perspective biofilms and floc particles have different appearances, at the microscopic scale they have many similarities and are formed by similar mechanisms.

Aggregation is a complex phenomenon and its mechanisms are still poorly understood, in spite of numerous studies.[104] One point upon which there is agreement, however, is that EPSs are central to the aggregation of individual bacteria into both floc particles[14,122] and biofilms.[24] In addition to their role in the aggregation of microbial cells, EPSs serve several other functions, including the retention of water, the buildup of nutrients, the accumulation of enzymatic activity, and as a barrier against toxins.[60] As the name suggests, EPS comprises materials of high molecular weight and a number of biomolecules, including polysaccharides, proteins, glycoproteins, nucleic acids, phospholipids, and humic acids have all been found in EPS,[82] with the relative importance of each depending on the nature of the culture and its growth conditions. Extracellular polymeric substances serve as a barrier between cells and the bulk liquid in which the biomass exists, thereby causing concentration gradients within flocs and biofilms that generate a range of ecological zones.[24] In addition, EPS allows organisms to establish stable arrangements, thereby allowing them to function as synergistic consortia,[24] accomplishing things that they might not be able to accomplish individually. Because EPS is closely associated with microbial aggregates, it is bound to, but distinct from, the active biomass. Recently, the role of amyloid-like adhesins in binding both biofilms[56] and flocs[57] has been explored.

The mere presence of EPS is not sufficient to ensure bioflocculation. Rather, the milieu within which the microbes are growing also has an impact. For example, both ionic strength[137] and divalent cations[42,122] play important roles. Bacteria are negatively charged. Consequently, the ionic strength must be sufficiently large to allow individual cells to approach closely enough together for bridging by the EPS to occur, but not so large as to cause deflocculation. Furthermore, ionic strength will influence the conformation of the EPS. One study suggested that an ionic strength on the order of 0.005–0.050 resulted in optimum floc stability.[137] Because both the cell surface and EPS are negatively charged, divalent cations are thought to act as bridges between the two, allowing aggregation to occur. Consequently, the proper level of divalent cations is essential. One study found that the minimum concentration of calcium and magnesium required to obtain a biofloc with good settling properties was in the range of 0.7–2.0 meq/L of each (14–40 mg/L of calcium and 8–24 mg/L of magnesium).[42] However, the actual concentration required in a particular facility will depend on the ionic strength of the wastewater. Furthermore, the ratio of divalent to monovalent cations is also important because when that ratio is less than 0.5, deterioration of the settling characteristics results.[42] This is thought to be due to the competition between divalent and monovalent cations for binding sites on the cell surfaces and the EPS.

Empirical observations suggest that the solids retention time (SRT) in a suspended growth process must exceed a minimum value to achieve bioflocculation. This observation is consistent with both the role of protozoa and EPS production by bacteria. Because they grow slowly, protozoa can be lost from systems in which the biomass is retained for only short times. Alternately, the requirement for a minimum SRT could represent a balance between the rate of EPS production by the floc-forming bacteria and the rate of generation of new surface area by bacterial growth. Although EPS is produced on a continuous basis, at short SRTs the rate of generation of new bacteria may exceed the rate of EPS production and bioflocculation is incomplete.

Figure 2.3 presents data illustrating the impact of SRT on bioflocculation in a pilot completely mixed activated sludge (CMAS) system receiving a synthetic wastewater consisting of glucose, yeast extract, and inorganic nutrients.[6] The proportion of activated sludge suspended solids that did not settle under quiescent conditions is plotted as a function of the SRT. A high proportion of the activated sludge solids (10–30%) would not settle when the process was operated at SRTs between 0.25 and 0.5 days, but that fraction was significantly reduced when the process was operated at an SRT of 1 day and it remained low as the SRT was increased up to 12 days. Settling velocity also

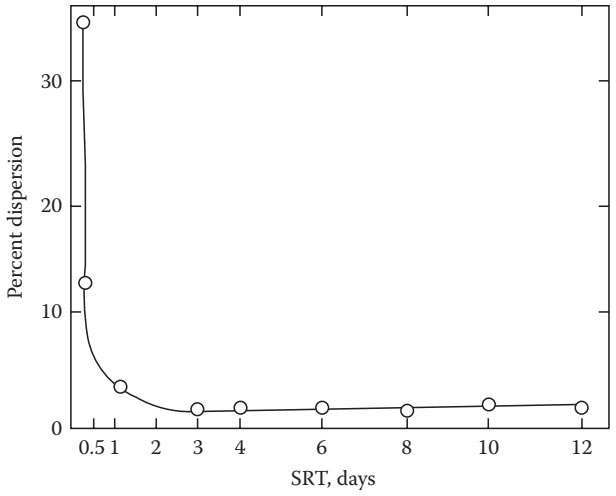

FIGURE 2.3 Effect of SRT on the amount of dispersed growth in activated sludge effluent. (Reprinted from Bisogni, J. J., and Lawrence, A. W., Relationships between biological solids retention time and settling characteristics of activated sludge. *Water Research,* 5:753–63, 1971. Copyright © Elsevier Ltd. With permission.)

TABLE 2.1
Characteristics of the Biomass Produced in the CMAS Reactors of Bisogni and Lawrence

SRT Range (Days)	Character of Solids
0.25–2	Predominantly dispersed growth
2–9	Well-formed average size floc of low to medium density
9–12	Pinpoint floc and irregularly shaped floc particles of low density that looked as though they had broken loose from larger floc particles (deflocculated)

Note: Adapted from Bisogni, J. J. and Lawrence, A. W., Relationships between biological solids retention time and settling characteristics of activated sludge. *Water Research,* 5:753–63, 1971.

TABLE 2.2
Relationship between SVI and Activated Sludge Settling Characteristics

SVI Range (mL/g)	Sludge Settling and Compaction Characteristics
<80	Excellent
80–150	Moderate
>150	Poor

increased with increasing SRT.[6] Microscopic analysis of the biomass produced at each operating SRT provided the results presented in Table 2.1.[6]

Sludge settleability and compaction are often quantified using the sludge volume index (SVI) measurement. It is performed by placing mixed liquor from a bioreactor into a one liter graduated cylinder and measuring the settled volume after 30 minutes of settling.[20] This volume is divided by the initial suspended solids concentration to obtain the SVI, and the result (with units of mL/g) represents the volume occupied by one gram of settled suspended solids. Table 2.2 summarizes

TABLE 2.3
Comparison of Sludge Settleability Indices

Index	Settling Device	Low-Speed Stirring	MLSS Concentration
Conventional SVI (SVI)	1 L graduated cylinder	No	Aeration basin MLSS concentration
Mallory SVI (SVI_m)	Mallory settleometer	No	Aeration basin MLSS concentration
Diluted SVI (DSVI)	1 L graduated cylinder	No	Mixed liquor diluted so that settled volume in graduated cylinder is less than or equal to 200 ml/L
Stirred SVI at 3.5 g SS/L ($SSVI_{3.5}$)	1L graduated cylinder	Yes	Tests at several MLSS concentrations; value at 3.5 g/L obtained by extrapolation

Note: Adapted from Jenkins, D., Richard, M. G., and Daigger, G. T., *Manual on the Causes and Control of Activated Sludge Bulking and Foaming*, 3rd ed., Lewis Publishers, Boca Raton, FL, 2004.

the typical relationship between SVI and biomass settling characteristics. An SVI of 150 mL/g is often considered to be the dividing line between a poorly settling (bulking) and a good settling (nonbulking) biomass. In addition to the conventional SVI, several other biomass settleability tests are available, but somewhat different results are obtained with each. Consequently, caution must be exercised when comparing reported SVI values to ensure that comparable settleability measurements were used. Table 2.3 compares the more common settleability tests. The diluted SVI (DSVI) or the stirred SVI at 3.5 g/L ($SSVI_{3.5}$) produce the most reproducible results.[17,53]

2.3.2 Aerobic/Anoxic Operations

2.3.2.1 Suspended Growth Bioreactors

Activated sludge, aerated lagoons, and aerobic digesters have similar microbial ecosystems, although they differ somewhat in the relative importance of various groups. The microorganisms in those operations are primarily Bacteria and microscopic Eucarya, and generally may be divided into five major classes: floc-forming organisms, saprophytes, nitrifying bacteria, predators, and nuisance organisms.[98] These are not distinct physiological groups and, in fact, any particular organism may fit into more than one category at a time or may change categories as the selective pressures within the community change.

Floc-forming organisms play a very important role in suspended growth biochemical operations because without them the biomass could not be separated from the treated wastewater nor would colloidal-sized organic pollutants be removed. Figure 2.2a shows typical, good settling biomass. The predominant floc-forming organisms in suspended growth cultures are bacteria.[15] A variety are capable of flocculation,[57,97] and they constitute between 10 and 40% of the volume of biomass in activated sludge floc.[57] Although *Zooglea*-like bacteria have been shown to play an important role in some systems,[106] studies using advanced molecular tools have shown that a broad range of microorganisms is responsible, including *Thauera*, *Azoarcus*, and *Aquaspirillum*-related bacteria, as well as a variety of filamentous bacteria belonging to the α-*proteobacteria*, β-*proteobacteria*, and γ-*proteobacteria*.[57] *Actinobacter*-like polyphosphate accumulating and amyloid-like adhesin-forming bacteria have been found to be prevalent in several BPR processes.[57] Nitrifying bacteria do not generate amyloid-like adhesins but adhere strongly to surfaces and other bacteria.[58]

Saprophytes are the organisms responsible for the degradation of organic matter. In wastewater treatment systems, they are primarily heterotrophic bacteria and include most of those considered to be floc formers. Nonflocculent bacteria are also involved, but are entrapped within the floc particles.

The saprophytes can be divided into primary and secondary degraders, as discussed previously, and the larger the number of electron donors, the more diverse the community will be.

Nitrification is the conversion of ammonia-N to nitrate-N and it may be performed by heterotrophic bacteria,[92] autotrophic bacteria,[92] or Archaea.[93] In spite of the fact that many heterotrophic species can form nitrite or nitrate from ammonia or reduced organic nitrogen compounds,[111,124] heterotrophic nitrification is not thought to routinely contribute to nitrogen oxidation in conventional wastewater nitrification, possibly because the kinetics are slower than they are in autotrophic nitrification.[111] Furthermore, ammonia oxidizing Archaea of the marine Crenarchaeota are significant contributors to nitrogen cycling in marine and terrestrial environments[89] and have been found in wastewater treatment plants,[93] although their relative contribution to nitrification activity there has yet to be determined. Therefore, nitrification in wastewater treatment systems is generally considered to be performed by autotrophic bacteria and we will make that assumption throughout this book.

Autotrophic oxidation of ammonia by ammonia oxidizing bacteria (AOB) can occur either in the presence or absence of dissolved oxygen. Anaerobic AOB, also known as anammox bacteria, are currently being explored for treatment of wastewaters containing high concentrations of ammonia and will be discussed in Section 2.3.3. Aerobic nitrification is very common in biological wastewater treatment reactors. It is a two-step process involving two groups of bacteria. Aerobic AOB oxidize ammonia-N to nitrite-N with hydroxylamine as an intermediate product. The nitrite-N is subsequently oxidized to nitrate-N in a single step by aerobic nitrite oxidizing bacteria (NOB). The microbial ecology of aerobic AOB and NOB in biological treatment reactors has been widely studied. While early studies concluded that aerobic AOB were primarily of the genus *Nitrosomonas* and aerobic NOB were primarily of the genus *Nitrobacter*, molecular tools have provided higher resolution of the ecology of both groups. Aerobic AOB that proliferate during the biological treatment of domestic wastewater are primarily of the β-*proteobacteria* and include the *Nitrosomonas* and *Nitrosospira* lineages.[102] The NOB are a diverse group that contain the α-*proteobacteria*, including *Nitrobacter* and *Nitrospira*. The latter are the more prevalent genus of NOB present in suspended growth treatment systems.[47] Independent of the genus present, aerobic AOB and NOB appear to grow in close physical association[84] and NOB cluster along nitrite gradients generated by AOB.[71]

The fact that aerobic AOB and NOB are autotrophic does not mean that they cannot incorporate exogenous organic compounds while obtaining their energy from inorganic oxidation, because they can.[48] The amount of such uptake will be small and will vary with the growth conditions, however, so that most equations depicting the stoichiometry of nitrification ignore it and use carbon dioxide as the sole carbon source. Nitrifying bacteria have several unique growth characteristics that are important to their impact on and survival in biochemical operations. The first is that their maximal growth rate is smaller than that of heterotrophic bacteria. Consequently, if suspended growth bioreactors are operated in a way that requires the bacteria to grow rapidly, the nitrifying bacteria will be lost from the system and nitrification will stop even though the removal of organic compounds will continue. Second, the amount of biomass formed per unit of nitrogen oxidized is small. As a result, they may make a negligible contribution to the MLSS concentration even when they have a significant effect on process performance.

The main predators in suspended growth bioreactors are the protozoa, which feed on the bacteria. About 230 species have been reported to occur in activated sludge and they may constitute as much as 5% of the biomass in the system.[97] Ciliates are usually the dominant protozoa, both numerically and on a mass basis. Almost all are known to feed on bacteria and the most important are either attached to or crawl over the surface of biomass flocs. Surveys of protozoa in activated sludge plants have revealed that selected species correlate with certain treatment performance patterns[68] or treatment process configurations.[63] As discussed earlier, it has been suggested that protozoa play a secondary role in the formation of biomass flocs and contribute to the absence of dispersed bacteria and colloidal organic material in stable communities.[15]

Nuisance organisms are those that interfere with proper operation of a biochemical reactor when present in sufficient numbers. In suspended growth bioreactors, most problems arise with respect

to removal of the biomass from the treated wastewater and are the result of filamentous bacteria and fungi. Although a very small number of filamentous bacteria is desirable to strengthen floc particles, too many are undesirable.[114] Even a small percentage by weight in the microbial community can make the effective specific gravity of the biomass flocs so low that the biomass becomes very difficult to remove by gravity settling. This leads to a situation known as bulking. A poor settling biomass is shown in Figure 2.2b. Because of the pioneering work of Eikelboom,[21] it is recognized that many types of filamentous organisms can be responsible for bulking and that different organisms are favored by different growth conditions. Effective bulking control is based on identification of the causative organism using a range of methods and elimination of the condition favoring its growth.[45] Table 2.4 ranks the most abundant filamentous organisms found in bulking sludges in the United States and Table 2.5 lists the suggested causes for some. In that table, the term "low F/M" refers to a low food to microorganism ratio; in other words, the system is being operated with a very low loading of organic matter to it. It should be noted that although *Gordonia* spp. (formerly *Nocardia* spp.) is a commonly found filamentous organism, it does not normally cause bulking because its filaments do not extend beyond the floc particle.[45]

The other major nuisance associated with suspended growth cultures is excessive foaming. The microbial ecology of this condition has been extensively studied and it now appears that most foaming incidents are caused primarily by *Actinobacteria*, including the mycolic acid formers (known as Mycolata, e.g., *Gordonia amarae*) and other species with hydrophobic cell surfaces (e.g., *Candidatus*

TABLE 2.4
Filament Abundance in Bulking and Foaming Activated Sludge in the United States

		Percentage of Treatment Plants with Bulking or Foaming Where Filament Was Observed	
Rank	**Filamentous Organism**	**Dominant**	**Secondary**
1	Nocardiaform organisms	31	17
2	Type 1701	29	24
3	Type 021N	19	15
4	Type 0041	16	47
5	*Thiothrix* spp	12	20
6	*Sphaerotilus natans*	12	19
7	*Microthrix parvicella*	10	3
8	Type 0092	9	4
9	*Haliscomenobacter hydrossis*	9	45
10	Type 0675	7	16
11	Type 0803	6	9
12	*Nostocoida limicola* (Types I, II, and III)	6	18
13	Type 1851	6	2
14	Type 0961	4	6
15	Type 0581	3	1
16	*Beggiatoa* spp	1	4
17	Fungi	1	2
18	Type 0914	1	1
—	All others	1	—

Note: Adapted from Jenkins, D., Richard, M. G., and Daigger, G. T., *Manual on the Causes and Control of Activated Sludge Bulking and Foaming*, 3rd ed., Lewis Publishers, Boca Raton, FL, 2004.

TABLE 2.5
Conditions Associated with Filamentous Organism Growth in Activated Sludge

Suggested Causative Conditions	Filamentous Organism
Low DO	*H. hydrossis*
	S. natans
	Type 1701
Low F/M	Type 0041
	Type 0675
	Type 1851
	Type 0803
Elevated low molecular weight organic acid concentration	Type 021N
	Thiothrix I and II
	N. limicola I, II and III
	Type 0914
	Type 0411
	Type 0961
	Type 0581
	Type 0092
Septic wastewater/sulfide	*Thiothrix* spp.
	Type 021N
	Type 0914
	Beggiatoa spp.
Nutrient deficiency	*S. natans*,
	Thiothrix I and II
	Type 021N
	H. hydrossis
Low pH	Fungi

Note: Adapted from Jenkins, D., Richard, M. G., and Daigger, G. T., *Manual on the Causes and Control of Activated Sludge Bulking and Foaming*, 3rd ed., Lewis Publishers, Boca Raton, FL, 2004.

"Microthrix parvicella").[45,113] Because *Actinobacteria* have very hydrophobic cell surfaces, they migrate to air bubbles where they stay, stabilizing the bubbles and causing foam.[45] There is still controversy concerning the conditions responsible for excessive foaming in suspended growth cultures. Interestingly, while foaming incidents correlate with increases in the abundance of foam-causing bacteria, the bacteria appear to have low metabolic activity during these incidents,[113] suggesting that our knowledge of the metabolic triggers of foaming remain unclear. Furthermore, warm temperatures may enhance the abundance of *Gordonia amarae*-like organisms during summer, making it a likely cause of seasonal foaming.

Although the ecosystems of activated sludge, aerated lagoons, and aerobic digestion are complex, they are not as complicated as those in suspended growth systems accomplishing biological nutrient removal (BNR). This is because BNR systems also contain anoxic and anaerobic reactors, which provide opportunities for the growth of microorganisms that do not ordinarily grow in totally aerobic systems.

The impact of having appropriately placed anoxic zones in a suspended growth system is to allow the proliferation of denitrifying heterotrophic bacteria. As discussed in Section 2.2.1, these organisms respire using nitrate-N and nitrite-N as electron acceptors when molecular oxygen is absent or present at very low concentrations.[12,66,111] Denitrification can be accomplished by a large number of bacterial

genera commonly found in wastewater treatment systems, thereby making the establishment of a denitrifying culture relatively easy. However, although many wastewater treatment engineers assume that all heterotrophic bacteria can switch from aerobic respiration to denitrification when oxygen is depleted and nitrate is present, in reality a large fraction of heterotrophic bacteria do not have the capacity to respire under both conditions.[19] Those that do perform heterotrophic denitrification are quite diverse and can be generally categorized as true denitrifiers (reduce both nitrate and nitrite), incomplete denitrifiers (reduce nitrate to nitrite), and nitrite reducers (reduce nitrite to gaseous by-products but cannot reduce nitrate).[19,96] Doubtless, the relative abundance of these groups depends on the nature of the biological reactor containing them as well as the wastewater undergoing treatment. On occasion, exogenous electron donors, especially methanol, are added to treatment plants required to meet stringent effluent nitrogen guidelines because the amount of organic matter in the wastewater is insufficient to do so. The bacteria that use methanol as electron donor (methylotrophs) during denitrification are distinct from those that denitrify on naturally occurring organic matter and decay products.[28] Consequently, treatment systems that initiate enhanced denitrification with methanol often experience a significant lag in performance while a microbial community that can use methanol establishes itself.[33]

As described in Section 1.3.1, the placement of an anaerobic zone at the influent end of an otherwise aerobic suspended growth system (illustrated in Figure 1.11) establishes the conditions required for proliferation of PAOs, thereby allowing development of a biomass that is rich in phosphorus. Although bacteria of the genus *Acinetobacter* were originally thought to be the major PAOs,[116] several other bacterial types have also been found to be capable of storing polyphosphate.[51,125] In fact, in one study, *Acinetobacter* was not the predominant PAO present and, instead, unidentified gram-positive bacteria were found.[125] Using molecular methods, researchers identified members of the *Rhodocyclus* group as important PAOs in acetate-enriched BPR reactors and the predominant species was called "*Candidatus* Accumulibacter phosphatis."[40] Although "*Ca*. A. phosphatis" has been found in full-scale systems around the world, not all PAOs belong to this group, or even *Rhodocyclus*.[36,130] The identity of the PAOs that do not belong to *Rhodocyclus* remains unclear and although some may be *Actinobacteria*,[5] it is possible that these PAOs exhibit different metabolism than the *Rhodocyclus*-related PAOs.[90]

Glycogen accumulating organisms (GAOs), originally called "G bacteria,"[9] often coexist with PAOs in BPR systems. Although GAOs store and use the same organic compounds as PAOs, they do not accumulate polyphosphate. Under certain conditions, excessive proliferation of GAOs can diminish organic compound availability to PAOs and prevent phosphorus removal goals from being achieved. As a result, excessive growth of GAOs makes them nuisance organisms and an understanding of their metabolism and physiology is important to their control. The GAOs are more diverse than PAOs, although their ecology has not yet been well defined. "*Candidatus* Competibacter phosphatis," *Defluviicoccus vanus*, *Actinobacteria*, and a range of unidentified tetrad-forming organisms are among the most commonly identified GAOs found in full-scale BPR bioreactors.[30,64,130]

The previous discussion has indicated the various types of organisms that can be present in suspended growth bioreactors. However, it is very important to recognize that the types that are present in any given system will depend on the reactor configuration and the biochemical environment imposed. In later chapters we will see how these conditions, which are under engineering control, can be used to select the type of microbial community required to accomplish a specific objective.

2.3.2.2 Attached Growth Bioreactors

Attached growth bioreactors are those in which the microorganisms grow as a biofilm on a solid support. In a fluidized bed bioreactor (FBBR), the biofilm grows on small particles of sand or activated carbon that are maintained in a fluidized state by the forces of water flowing upward. Submerged attached growth bioreactors contain similar support particles, as well as synthetic media, but the water being treated flows over them without displacing them. Thus, in both bioreactor types, the biofilm is surrounded by the fluid containing the electron donor being removed. In a trickling filter or rotating biological contactor, on the other hand, the biofilm grows on a large surface over which

the wastewater flows in a thin film (trickling filter) or which moves through the wastewater (rotating biological contactor). As a consequence, the fluid shear associated with the latter two is less than that associated with the first two. Hybrid designs, such as integrated fixed film activated sludge systems, which contain plastic media in an activated sludge basin, exhibit fluid shear forces between the other examples. Ultimately, the hydrodynamic environment in which attached growth bioreactors function has an impact on the type of microbial community involved and the relative proximity of microbial groups within the biofilm.

Our understanding of the microbial ecology of FBBRs and submerged attached growth bioreactors is improving. Similar to those in suspended growth bioreactors, the communities in these systems are primarily comprised of bacteria and protozoa. In contrast, trickling filters and rotating biological contactors contain more diverse microbial communities that include many other Eucarya, notably nematodes, rotifers, snails, sludge worms, and larvae of certain insects.[13] This more complex food chain allows more complete oxidation of organic matter, with the net result that less excess biomass is produced. This has the beneficial effect of decreasing the mass of solid material that must be disposed of.

Bacteria form the base of the food chain by acting on the organic matter in the wastewater being treated. Soluble materials are taken up rapidly, while colloidal-sized particles become entrapped in the EPS layer forming the biofilm. There they undergo attack by extracellular enzymes, releasing small molecules that can be metabolized. The bacterial community is composed of primary and secondary saprophytes, much like suspended growth bioreactors. Unlike suspended growth cultures, however, the species distribution is likely to change with position in the reactor. Attached growth reactors can also contain nitrifying bacteria, which tend to be found in regions of the biofilm where the organic compound concentration is low.[27] In addition, the nature of the nitrifying bacteria present tends to vary as the ammonia load varies.[26]

Quite extensive communities of Eucarya are known to exist in trickling filters.[13,15,121] Over 90 species of fungi have been reported and of these, more than 20 species are considered to be permanent members of the community. Their role is similar to that of the bacteria (i.e., saprophytic). Many protozoa have also been found, with large communities of Sarcodina, Mastigophora, and Ciliata being reported. Their roles are largely those of predators. During warm summer months, algae can flourish on the upper surfaces of the biomass. Usually green algae and diatoms predominate. Finally, trickling filters also contain a large metazoan community, consisting of annelid worms, insect larvae, and snails. These feed on the biofilm and in some cases have been responsible for extensive biofilm destruction.

Because of the diverse nature of the microbial community in attached growth bioreactors, the microbial interactions are extremely complex. Unfortunately, even less is known about the impact of these interactions on system performance than is known about them in suspended growth systems.

2.3.3 Anaerobic Operations

The microbial communities in anaerobic operations are primarily procaryotic, with members of both the Bacteria and the Archaea being involved. Although fungi and protozoa have been observed under some circumstances, the importance of eucaryotic organisms is questionable.[120] Thus, the emphasis here will be on the complex and important interactions between the Bacteria and the Archaea that are fundamental to the successful functioning of methanogenic communities. Because those interactions occur in both suspended and attached growth systems, no distinctions will be made between the two.

2.3.3.1 General Nature of Methanogenic Anaerobic Operations

The multistep nature of anaerobic biochemical operations involving methanogenesis is depicted in Figure 2.4. Before insoluble organic materials can be consumed, they must be solubilized, just as was necessary in aerobic systems. Furthermore, large soluble organic molecules must be reduced

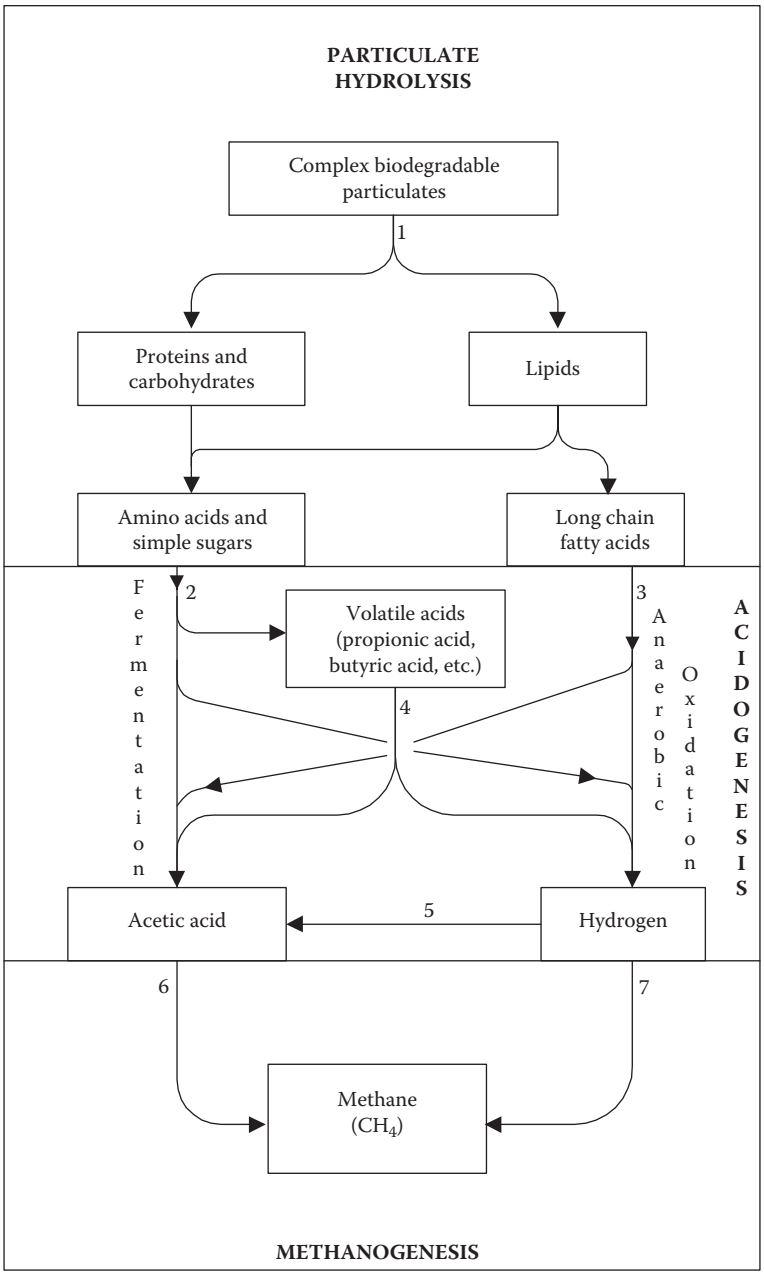

FIGURE 2.4 Multistep nature of methanogenic processes.

in size to facilitate transport across the cell membrane. The reactions responsible for solubilization and size reduction are usually hydrolytic and are catalyzed by extracellular enzymes produced by bacteria. They are all grouped together as hydrolysis reactions (reaction 1) in Figure 2.4, but in reality many enzymes are involved, such as cellulases, amylases, and proteases. They are produced by the fermentative bacteria that are an important component of the second step, acidogenesis.

Acidogenesis is carried out by members of the domain Bacteria. Amino acids and sugars are degraded by fermentative reactions (reaction 2) in which organic compounds serve as both electron donors and acceptors. The principle products of reaction 2 are intermediary degradative products

like propionic and butyric acids and the direct methane precursors, acetic acid and H_2. The H_2 production from fermentative reactions is small and originates from the dehydrogenation of pyruvate by mechanisms that are different from the production of the bulk of the H_2 produced.[31] In contrast, most of the H_2 produced comes from oxidation of volatile and long chain fatty acids to acetic acid (reactions 3 and 4) and arises from the transfer of electrons from reduced carriers directly to hydrogen ions, in a process called anaerobic oxidation.[31] Because of the thermodynamics of this reaction, it is inhibited by high partial pressures of H_2, whereas the production of H_2 from pyruvate is not.

The production of H_2 by anaerobic oxidation is very important to the proper functioning of anaerobic processes. First, H_2 is one of the primary electron donors from which methane is formed. Second, if no H_2 were formed, acidogenesis would not result in the oxidized product acetic acid being the major soluble organic product. Rather, the only reactions that could occur would be fermentative, in which electrons released during the oxidation of one organic compound are passed to another organic compound that serves as the electron acceptor, yielding a mixture of oxidized and reduced organic products. Consequently, the energy level of the soluble organic matter would not be changed significantly because all of the electrons originally present would still be in solution in organic form. When H_2 is formed as the reduced product, however, it can escape from the liquid phase because it is a gas, thereby causing a reduction in the energy content of the liquid. In actuality, the H_2 does not escape. It is used as an electron donor for methane production, but because methane is removed as a gas, the same thing is accomplished. Finally, if H_2 formation did not occur and reduced organic products were formed, they would accumulate in the liquid because they cannot be used for methane production. Only acetic acid, H_2, methanol, and methylamines can be used. As shown by reaction 5, some of the H_2 can be combined with carbon dioxide by H_2-oxidizing acetogens to form acetic acid,[135] but since the acetic acid can serve as a carbon and energy source for methanogens, the impact of this reaction is thought to be small.

The products of the acidogenic reactions, acetic acid and H_2, are used by methanogens, which are members of the domain Archaea, to produce methane gas. Two groups are involved: aceticlastic methanogens that split acetic acid into methane and carbon dioxide (reaction 6), and H_2-oxidizing methanogens that reduce carbon dioxide (reaction 7). It is generally accepted that about two-thirds of the methane produced in anaerobic digestion of primary sludge is derived from acetic acid, with the remainder coming from H_2 and carbon dioxide.[31,135] With the exception of the electrons incorporated into the cell material formed, almost all of the energy removed from the liquid being treated is recovered in the methane. Chemical oxygen demand (COD),[109] a common measure of pollutant strength, is a measure of the electrons available in an organic compound, expressed in terms of the amount of oxygen required to accept them when the compound is completely oxidized to carbon dioxide and water. One mole of methane requires two moles of oxygen to oxidize it to carbon dioxide and water. Consequently, each 16 grams of methane produced and lost to the atmosphere corresponds to the removal of 64 grams of COD from the liquid.[75] At standard temperature and pressure, this corresponds to 0.34 m^3 of methane for each kg of COD stabilized.[76]

2.3.3.2 Microbial Groups in Methanogenic Communities and Their Interactions

The hydrolytic and fermentative bacteria comprise a rather diverse group of facultative and obligately anaerobic Bacteria. In sewage sludge digesters the numbers of obligate anaerobes have been found to be over 100 times greater than the number of facultative bacteria.[50] This does not mean that facultative bacteria are unimportant, because their relative numbers can increase when the influent contains large numbers of them[43] or when the bioreactor is subjected to shock loads of easily fermentable compounds.[70] Nevertheless, it does appear that most important hydrolytic and fermentative reactions are performed by strict anaerobes, such as *Bacteroides*, *Clostridia*, *Bifidobacteria*, and members of the family *Porphyromonadaceae*,[62,107] although the nature of the electron donor will determine the species present.

The role of H_2 as an electron sink is central to the production of acetic acid as the major end product of acidogenesis. Reactions leading from long chain fatty acids, volatile acids, amino acids, and

carbohydrates to acetic acid and H_2 are thermodynamically unfavorable under standard conditions, having positive standard free energies.[135] Thus, when the H_2 partial pressure is high, these reactions will not proceed and instead fermentations occur with the results discussed above. Under conditions in which the partial pressure of H_2 is 10^{-4} atmospheres or less, however, the reactions are favorable and can proceed, leading to end products (acetic acid and H_2) that can be converted to methane. This means that the bacteria that produce H_2 are obligately linked to the methanogens that use it. Only when the methanogens continually remove H_2 by forming methane will the H_2 partial pressure be kept low enough to allow production of acetic acid and H_2 as the end products of acidogenesis. Likewise, methanogens are obligately linked to the bacteria performing acidogenesis because the latter produce the carbon and energy sources required by the former. Such a relationship between two microbial groups is called obligate syntrophy.

While the organisms responsible for the fermentative reactions are reasonably well characterized, less is known about the H_2-producing acidogenic bacteria. This is due in part to the fact that the enzyme system for H_2 production is under very strict control by H_2.[110] As a consequence, early studies that attempted to enumerate the H_2-forming bacteria underestimated them by allowing H_2 to accumulate during testing. However, because H_2 partial pressures are kept low in anaerobic biochemical operations, H_2-forming bacteria play an important role. Several microorganisms have been identified and studied, and include the obligate anaerobic *Clostridia*,[88] facultative anaerobes including the *Enterobactericiae*,[88] and other novel acidophilic H_2-producing populations.[131]

The major nuisance organisms in anaerobic operations are the sulfate-reducing bacteria, which can be a problem when the wastewater contains significant concentrations of sulfate. Sulfate-reducing bacteria are all obligate anaerobes of the domain Bacteria. They are morphologically diverse, but share the common characteristic of being able to use sulfate as an electron acceptor. 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 specialize in the oxidation of fatty acids, particularly acetate, to carbon dioxide while reducing sulfate to sulfide. An important genus in this group is *Desulfobacter.*

The H_2-oxidizing methanogens are classified into three orders within the domain *Archaea*: *Methanobacteriales*, *Methanococcales*, and *Methanomicrobiales*.[7] A wide variety of these microorganisms have been cultured from anaerobic digesters, including the genera *Methanobrevibacter* and *Methanobacterium* from the first order, and the genera *Methanospirillum* and *Methanogenium* from the third.[136] They are all strictly obligate anaerobes that obtain their energy primarily from the oxidation of H_2 and their carbon from carbon dioxide. Because of this autotrophic mode of life, the amount of cell material synthesized per unit of H_2 used is low. During their metabolism they also use carbon dioxide as the terminal electron acceptor,[34] forming methane gas in the process:

$$4H_2 + CO_2 \rightarrow CH_4 + 2H_2O. \tag{2.1}$$

Their range of electron donors is very restricted, usually being limited to H_2 and formate.[107] In some cases, short chain alcohols can also be used.[7]

In spite of the importance of the aceticlastic route to methane (reaction 6), fewer aceticlastic methanogens have been cultured and identified. All are of the order *Methanosarcinales*, which contains two families, *Methanosarcinaceae* and *Methanosaetaceae*.[7] *Methanosarcina*, of the first family, can be cultivated from anaerobic operations[136] and is among the most versatile genera of methanogens known, being able to use H_2 and carbon dioxide, methanol, methylamines, and acetic acid as substrates.[107,135] When acetic acid is the substrate, it is cleaved, with all of the methyl carbon ending up as methane and all of the carboxyl carbon as carbon dioxide:

$$*CH_3COOH \rightarrow *CH_4 + CO_2. \tag{2.2}$$

Methanosarcina grows relatively rapidly at high acetic acid concentrations, although it is very sensitive to changes in that concentration. Furthermore, H_2 exerts a regulatory effect on acetic acid utilization, shutting it down as the H_2 partial pressure increases. The family *Methanosaetaceae* contains a single genus, *Methanosaeta* (formerly *Methanothrix*), the members of which can use only acetic acid as their electron and carbon donor.[7] They grow much more slowly than *Methanosarcina* at high acetic acid concentrations, but are not influenced as strongly by that concentration and can compete effectively when it is low. As a consequence, the manner in which an anaerobic operation is designed and operated will determine the predominant aceticlastic methanogen. For example, *Methanosaeta* are typically found in mesophilic anaerobic digesters operated so as to maintain low acetate concentrations.[80]

2.3.3.3 Anaerobic Ammonia Oxidation

Although historically the most important application of anaerobic microbial communities in environmental engineering practice has been for the stabilization of waste biomass and primary sewage solids, the discovery of anaerobic ammonia oxidation has fostered considerable interest. This is because aerobic ammonia oxidation (nitrification) requires large amounts of oxygen, with the high energy costs associated with its transfer. Anaerobic ammonia oxidation has the potential to greatly reduce that cost. At this point the focus will be upon the microbiology involved. Potential applications of anaerobic ammonia oxidation are discussed in Section 23.3.3.

Anaerobic ammonia oxidation (anammox) has been studied in marine environments, where it plays a significant role in producing N_2 while oxidizing ammonia,[127] and in wastewater treatment plants from which the responsible bacteria were first enriched.[87] They are obligate anaerobes that oxidize ammonia using nitrite as the electron acceptor and possess a unique organelle where the oxidation occurs. They also grow very slowly.[117] The marine organisms have been found in a number of locations worldwide and all fall within the phylum Planctomycetes.[127] The organisms enriched from wastewater treatment plants are also Planctomycetes but are of different genera.[46] Based on studies in a sequencing batch reactor containing an enrichment culture obtained from a wastewater treatment plant,[127] the stoichiometry of their metabolism has been proposed to be[117]

$$\begin{aligned} NH_4^+ + 1.32\ NO_2^- + 0.066\ HCO_3^- + 0.13\ H^+ \rightarrow \\ 1.02\ N_2 + 0.26\ NO_3^- + 0.066\ CH_2O_{0.5}N_{0.15} + 2.03\ H_2O. \end{aligned} \tag{2.3}$$

This is consistent with observations during start-up of a full-scale facility performing the anammox reaction.[123]

2.3.4 The Complexity of Microbial Communities: Reality versus Perception

It is apparent from the preceding that the microbial communities in biochemical operations are very complex, involving many trophic levels and many genera and species within a trophic level. Unfortunately, most studies on community structure have been descriptive and the exact roles of many organisms have not even been defined much less quantified. As a consequence, wastewater treatment engineers have tended to view the communities in biochemical operations as if they were monocultures consisting only of procaryotes of a single species. This is slowly changing, but the models used by engineers still primarily reflect only the procaryotic portion of the community, and its divisions are usually limited to major groups, such as aerobic heterotrophs, floc-formers, denitrifiers, nitrifiers, PAOs, and so on. In the chapters to follow we will be exploring the performance of biochemical operations based on these divisions. While the resulting mathematical descriptions are adequate for establishing a fundamental understanding of system performance, and indeed, even for design, it is important to remember the complex nature of the microbial communities involved and to temper your acceptance of the models accordingly. As engineers and microbiologists continue to

work together to understand these fascinating systems, we will eventually be able to consider community structure in a quantitative way, resulting in better system design and performance.

2.4 IMPORTANT PROCESSES IN BIOCHEMICAL OPERATIONS

Regardless of the nature and complexity of the microbial community involved, there are certain fundamental processes that occur universally in biochemical operations. The relative importance of these processes, and hence the outcome from a biochemical operation, depends on the physical configuration of the operation and the manner in which it is operated. Our ability to select and design the appropriate biochemical operation for a specific task depends on our recognition of the importance of the various processes in it and our capability for quantitatively expressing the rates of those processes. In this section we will introduce those processes in qualitative terms; in Chapter 3 we will describe them quantitatively.

2.4.1 Biomass Growth, Substrate Utilization, and Yield

When reduced to their barest essentials, biochemical operations are systems in which microorganisms are allowed to grow by using pollutants as their carbon and/or energy source, thereby removing the pollutants from the wastewater and converting them to new biomass and carbon dioxide or other innocuous forms. Because of the role of enzymes in microbial metabolism, the carbon and/or energy source for microbial growth is often called the substrate, causing wastewater treatment engineers to commonly refer to the removal of pollutants during biomass growth as substrate utilization. If growth is balanced, which is the case for most (but not all) biochemical operations, biomass growth and substrate utilization are coupled, with the result that the removal of one unit of substrate results in the production of Y units of biomass, where Y is called the true growth yield, or often, simply the yield.* Because of the coupling between biomass growth and substrate utilization, the rates of the two activities are proportional, with Y as the proportionality factor. Consequently, the selection of one as the primary event (or cause) and the other as the secondary event (or effect) is arbitrary. Both selections are equally correct and benchmark papers have been published using both substrate removal[61] and biomass growth[39] as the primary event. The point of view taken in this book is that biomass growth is the fundamental event, and the rate expressions presented in Chapter 3 are written in terms of it. However, it should be emphasized that rate expressions for biomass growth and substrate utilization can be interconverted through use of the yield, Y.

Because of the central role that Y plays in the relationship between biomass growth and substrate utilization, it is an intrinsic characteristic. Consequently, a clear understanding of the factors that can influence its magnitude is important. The development of such an understanding requires consideration of the energetics of microbial growth, including energy conservation and energy requirements for synthesis.

2.4.1.1 Overview of Energetics

Microorganisms require four things for growth: carbon, inorganic nutrients, energy, and reducing power. As mentioned in Section 2.2.1, microorganisms derive energy and reducing power from oxidation reactions, which involve the removal of electrons from the substrate with their ultimate transfer to the terminal electron acceptor. Consequently, the energy available in a substrate depends on its oxidation state, which is indicative of the electrons available for removal as the substrate is oxidized. Highly reduced compounds contain more electrons and have a higher standard free energy than do highly oxidized compounds regardless of whether they are organic or inorganic. As we saw in Chapter 1, most biochemical operations are used for the removal of soluble organic matter and the stabilization of insoluble organic matter. Consequently, in this discussion we will focus on carbon

* Throughout this book, the term "yield" will be considered synonymous with "true growth yield."

oxidation by heterotrophic bacteria. Since COD is a measure of available electrons, compounds with a high COD:C ratio are highly reduced, whereas those with a low COD:C ratio are more oxidized. The carbon in methane is in the most highly reduced state possible, with a COD:C ratio of 5.33 mg COD/mg C, whereas the carbon in carbon dioxide is in the most highly oxidized state with a COD:C ratio of zero. Thus, all organic compounds will have a COD:C ratio between these extremes.

As heterotrophic bacteria oxidize the carbon in organic compounds through their catabolic pathways, they convert them to metabolic intermediates of the central amphibolic pathways that are in a higher oxidation state than either the starting compound or the biomass itself. Those metabolic intermediates are used in the anabolic pathways for cell synthesis, but since they are in a higher oxidation state than the cell material being synthesized from them, electrons must be available in an appropriate form for reducing them. Those electrons arise from the original substrate during its catabolism and are transferred to the anabolic pathways through the use of carriers such as nicotinamide adenine dinucleotide (NAD) and nicotinamide adenine dinucleotide phosphate (NADP), which alternate between the oxidized (NAD and NADP) and the reduced (NADH and NADPH) state. Thus NAD and NADP serve as electron acceptors for catabolic reactions, forming NADH and NADPH, which act as electron donors for biosynthetic reactions. The availability of NADH and NADPH is called reducing power.

Biosynthetic reactions also require energy in a form that can be used in coupled reactions to join the amphibolic intermediates into new compounds. That energy is provided primarily by adenosine triphosphate (ATP) and to a lesser degree by other nucleotides. Adenosine triphosphate is generated by phosphorylation reactions from adenosine diphosphate (ADP) and when the ATP is used to provide energy in biosynthetic reactions, ADP is released for reuse. The ATP can be formed from ADP by two types of phosphorylation reactions: substrate level and electron transport phosphorylation. During substrate level phosphorylation, ATP is formed directly by coupled reactions within a catabolic pathway. Only small amounts of ATP can be generated in this way. Much larger amounts can be generated during electron transport phosphorylation, which occurs as electrons removed during oxidation of the substrate (and carried in NADH) are passed through the electron transport (or terminal respiratory) chain, to the terminal electron acceptor, setting up a proton-motive force.[67] The magnitude of the proton motive force, and consequently, the amount of ATP that can be generated, depends on both the organism and the nature of the terminal electron acceptor.

An important concept to recognize about microbial energetics is that as a compound is degraded, all of the electrons originally in it must end up in the new cell material formed in the terminal electron acceptor or in the soluble organic metabolic intermediates excreted during growth. If a compound is mineralized, the amount of metabolic intermediates will be very small, so that essentially all electrons must end up either in the cell material formed or in the terminal acceptor. Because the yield is the amount of cell material formed per unit of substrate destroyed, because the amount of cell material formed depends on the amount of ATP generated, and because the amount of ATP generated depends on the electrons available in the substrate, the organism carrying out the degradation and the growth environment, it follows that the yield also depends on the nature of the substrate, the organism involved, and the growth environment.

2.4.1.2 Effects of Growth Environment on ATP Generation

The electron transport chains found in most Bacteria and Eucarya share common features. They are highly organized and are localized within membranes. They contain flavoproteins and cytochromes that accept electrons from a donor like NADH and pass them in discrete steps to a terminal acceptor. All conserve some of the energy released by coupling the electron transfer to the generation of proton motive force, which drives a number of processes, such as the synthesis of ATP from ADP and inorganic phosphate, active transport, and flagellar movement. The electron transport chain in Eucarya is located in the mitochondria and is remarkably uniform from species to species. The electron transport chain in Bacteria is located in the cytoplasmic membrane and exhibits considerable variety among individual species in the identity of the individual components and in the presence or

absence of sections of the chain. Nevertheless, the sequential organization of the components of the electron transport chain is determined by their standard oxidation-reduction potentials. Table 2.6 presents the potentials for the array of couples found in mitochondrial electron transport chains.[34] The couples in Bacteria are similar, but not necessarily identical. The transfer is in the direction of increasing redox potential until the final reaction with the terminal acceptor is catalyzed by the appropriate enzyme. When the environment is aerobic, oxygen serves as the terminal acceptor and the enzyme is an oxidase.

Adenosine triphosphate generation is associated with the transfer of electrons down the electron transport chain through electron transport phosphorylation, although it is not directly coupled to specific biochemical reactions that occur during that transfer.[3,67] Rather, the generation of ATP is driven by proton motive force through chemiosmosis. The elements of the electron transport chain are spatially organized in the cytoplasmic membrane of Bacteria and the mitochondrial membrane of Eucarya in such a way that protons (hydrogen ions, H^+) are translocated across the membrane as the electrons move down the electron transport chain (i.e., toward more positive E_0' values). In Bacteria the transfer is from the cytoplasm (inside the cell) to the periplasmic space (outside the cell); in Eucarya from inside the mitochondria to the outside. The transfer of electrons across the membrane establishes a proton gradient that causes a diffusive counterflow of protons back across the membrane through proton channels established by a membrane-bound ATPase enzyme. This proton counterflow drives the synthesis of ATP from ADP and inorganic phosphate. The number of ATP synthesized per electron transferred to the terminal acceptor depends on the nature and spatial organization of the electron transport chain because they determine the number of protons that are translocated per electron transferred down the chain. In mitochondria, 3 ATP can be synthesized per pair of electrons transferred. However, in Bacteria the number will depend on the organization of the electron transport chain in the particular organism involved. This explains why the amount of ATP synthesized from the oxidation of a given substrate depends on the organism performing the oxidation.

TABLE 2.6
The Standard Oxidation-Reduction Potentials of a Number of Redox Couples of Interest in Biological Systems

Redox Couple	E_0' (mV)
$H_2/2H^+ = 2e^-$	–420
Ferredoxin reduced/oxidized	–410
NADPH/$NADP^+$	–324
NADH/NAD^+	–320
Flavoproteins reduced/oxidized	–300 to 0
Cytochrome b reduced/oxidized	+30
Ubiquinone reduced/oxidized	+100
Cytochrome c reduced/oxidized	+254
Cytochrome a_3 reduced/oxidized	+385
$O^{2-}/½O_2 + 2e^-$	+820

Note: Data from Hamilton, W. A., Microbial energetics and metabolism. *Micro-Organisms in Action: Concepts and Applications in Microbial Ecology*, 75–100, eds. J. M. Lynch and J. E. Hobbie, Blackwell Scientific Publications, Palo Alto, CA, 1988.

TABLE 2.7
Standard Oxidation Reduction Potentials of Various Acceptor and Donor Redox Couples

Redox Couple	E_0' (mV)
Acceptor	
$½O_2/H_2O$	+820
NO_3^-/NO_2^-	+433
NO_2^-/NO	+350
Fumarate/succinate	+33
SO_4^{2-}/SO_3^{2-}	–60
CO_2/CH_4	–244
Donor	
$H_2/2H^+$	–420
$HCOOH/HCO_3^-$	–416
$NADH/NAD^+$	–320
Lactate/pyruvate	–197
Malate/oxaloacetate	–172
Succinate/fumarate	+33

Note: Data from Hamilton, W. A., Microbial energetics and metabolism. *Micro-Organisms in Action: Concepts and Applications in Microbial Ecology*, 75–100, eds. J. M. Lynch and J. E. Hobbie, Blackwell Scientific Publications, Palo Alto, CA, 1988.

In the absence of molecular oxygen, other terminal acceptors may accept electrons from the electron transport chain and the oxidation reduction potentials ($\Delta E_O'$) for them, as well as for various donors, are given in Table 2.7.[34] In order for ATP to be generated by electron transport phosphorylation, the oxidation-reduction potential for the donor redox couple must be smaller (more negative) than the potential for the acceptor redox couple, there must be at least one site of proton translocation in the electron transport chain between the final acceptor and the point where the donor contributes its electrons, and the associated free energy change ($\Delta G^{0\prime}$) must exceed 44 kJ ($\Delta G^{0\prime} = -2F \cdot \Delta E_O'$, where $F = 96.6$ kJ/(V · mol)). Nitrate and nitrite are important terminal electron acceptors in biochemical operations performing denitrification and the bacteria capable of using the nitrogen oxides as electron acceptors are biochemically and taxonomically diverse.[52] The enzyme nitrate reductase is responsible for the conversion of nitrate to nitrite. It can be either membrane bound or located in the periplasmic space between the cytoplasmic membrane and outer membrane, and couples with the electron transport chain through cytochromes. The enzymes nitrite reductase, nitric oxide reductase, and nitrous oxide reductase are involved in the reduction of nitrite to nitrogen gas in coordination with electron transport.[34,52,96] The number of ATPs synthesized per electron transported is less than the number associated with oxygen as the terminal acceptor because the available free energy change is less. Consequently, bacteria growing with nitrate as the terminal electron acceptor exhibit lower yields than bacteria growing under aerobic conditions.[11,79]

Under strictly anaerobic conditions (i.e., when neither oxygen nor the nitrogen oxides are present), many Bacteria generate their ATP through substrate level phosphorylation associated with fermentation reactions in which the oxidation of one organic substrate is coupled to the reduction of another. The second substrate is generally a product of the catabolic pathway leading from the oxidized substrate with the result that the fermentation pathway is internally balanced, with neither

TABLE 2.8
Types of Fermentations of Various Microorganisms

Type of Fermentation	Products	Organisms
Alcoholic	Ethanol, CO_2	Yeast
Lactic acid	Lactic acid	*Streptococcus*, *Lactobacillus*
Mixed acid	Lactic acid, acetic acid, ethanol, CO_2, H_2	*Escherichia*, *Salmonella*
Butanediol	Butanediol, ethanol, lactic acid, acetic acid, CO_2, H_2	*Aerobacter*, *Serratia*
Butyric acid	Butyric acid, acetic acid, CO_2, H_2	*Clostridium butyricum*
Acetone-butanol	Acetone, butanol, ethanol	*Clostridium acetobutylicum*
Propionic acid	Propionic acid	*Propionibacterium*

a net production nor a net requirement for reducing power. Several types of fermentation reactions are listed in Table 2.8. Because ATP generation occurs only by substrate level phosphorylation and a large part of the available electrons in the original substrate end up in the reduced organic products, bacteria receive relatively little energy in this mode of growth and thus have low yields per unit of substrate processed. As discussed in Section 2.3.3, however, the production of H_2 allows more oxidized products like acetate to be produced. As a result, more ATP can be produced by bacteria when they generate H_2, allowing them to have a higher biomass yield per unit of substrate processed.

Methanogens are obligately anaerobic Archaea that have very restricted nutritional requirements, with the oxidation of acetate and H_2 being their main sources of energy. Even though methane is produced from the reduction of carbon dioxide during the oxidation of H_2, methanogens lack the components of a standard electron transport chain and thus carbon dioxide does not function as a terminal electron acceptor in a manner analogous to nitrate or oxygen.[34] Rather, reduction of carbon dioxide to methane involves a complex sequence of events requiring a number of unique coenzymes.[129] However, there is a sufficient free energy change during methane formation for the theoretical production of two molecules of ATP and it appears that a normal chemiosmotic mechanism is involved,[34] although it involves a sodium motive force as well as a proton motive force.[129] Regardless of the exact mechanisms involved, it is important to recognize that ATP generation in Archaea is different from that associated with both respiration and fermentation in Bacteria and Eucarya. Furthermore, like bacteria growing in anaerobic environments, methanogens have low yields.

2.4.1.3 Factors Influencing Energy for Synthesis

Energy for synthesis represents the energy required by microorganisms to synthesize new cell material. In the absence of any other energy requirements, the energy required for synthesis is the difference between the energy available in the original substrate and the energy associated with the cell material formed, or in the common units of the environmental engineer, the difference between the COD of the original substrate and the COD of the biomass formed. Consequently, the energy for synthesis and the yield are intimately linked. If the efficiency of ATP generation were the same for all bacteria, it would be possible to theoretically predict the energy for synthesis, and hence the yield, from thermodynamic considerations.[77] However, as we saw above, the amount of ATP generated per electron transferred differs from microorganism to microorganism, which means that the efficiency of energy generation differs. This, coupled with the fact that the pathways of synthesis and degradation are not the same in all microorganisms, makes it difficult to use exactly the thermodynamic approaches for predicting yields that have been presented in the environmental engineering literature. Nevertheless, there are many instances in which it would be advantageous to have a theoretical prediction of the energy for synthesis or the yield prior to experimental work and a technique based on the Gibbs energy dissipation per unit of biomass produced appears to be best.[37] Regardless, thermodynamic concepts are most useful for understanding why different

substrates and different terminal electron acceptors have different energies of synthesis and yields associated with them.

During biomass growth, energy is required to synthesize the monomers needed to make the macromolecules that form the structural and functional components of the cell. This suggests that more energy would be required for a culture to grow in a minimal medium containing only a single organic compound as the carbon and energy source than in a complex medium in which all required monomers were supplied. Actually, such a conclusion is false.[112] For example, the energy needed to synthesize all of the amino acids needed by a cell amounts to only about 10% of the total energy needed to synthesize new cell material. This is because macromolecules are too large to be transported into the cell and must be formed inside even when all of the needed monomers are provided in the medium. Consequently, although the complexity of the growth medium has some effect on the energy required for synthesis, it is not large.

Of more importance are the oxidation state and size of the carbon source.[37] The oxidation state of carbon in biomass is roughly the same as that of carbon in carbohydrate.[112] If the carbon source is more oxidized than that, reducing power must be expended to reduce it to the proper level. If the carbon source is more reduced, it will be oxidized to the proper level during normal biodegradation and no extra energy will be required. Therefore, as a general rule, a carbon source at an oxidation state higher than that of carbohydrate will require more energy to be converted into biomass than will one at a lower oxidation state. Pyruvic acid occupies a unique position in metabolism because it lies at the end of many catabolic pathways and the beginning of many anabolic and amphibolic ones. As such, it provides carbon atoms in a form that can be easily incorporated into other molecules. Indeed, three-carbon fragments play an important role in the synthesis of many compounds. If the carbon source contains more than three carbon atoms, it will be broken down to size without the expenditure of large amounts of energy. If it contains less than three carbon atoms, however, energy must be expended to form three-carbon fragments for incorporation. Consequently, substrates containing few carbon atoms require more energy for synthesis than do large ones.

Carbon dioxide, which is used by autotrophic organisms as their chief carbon source, is an extreme example of the factors just discussed, being a single-carbon compound in which the carbon is in the highest oxidation state. Consequently, the energy for synthesis for autotrophic growth is very much higher than for heterotrophic growth. As a result, the amount of biomass that can be formed per unit of available electrons in the energy source is quite low.

2.4.1.4 True Growth Yield

The true growth yield (Y) is defined as the amount of biomass formed per unit of substrate removed when all energy expenditure is for synthesis. In this context, the substrate is usually taken to be the electron donor, although it can be defined differently. If the electron donor is an organic compound, it is common in environmental engineering practice to express Y in terms of the amount of soluble COD removed from the wastewater. This is because wastewaters contain undefined, heterogeneous mixtures of organic compounds and the COD is an easily determined measurement of their quantity. In addition, the COD is fundamentally related to available electrons, having an electron equivalent of eight grams of oxygen. Thus, a Y value expressed per gram of COD removed can be converted to a Y value per available electron when multiplying by eight. If the electron donor is an inorganic compound, such as ammonia or nitrite nitrogen, it is common to express Y in terms of the mass of the element donating the electrons. Furthermore, regardless of the nature of the electron donor, it has been common practice to express the amount of biomass formed on a dry weight basis (i.e., mass of total suspended solids, TSS) or on the basis of the dry weight of ash-free organic matter (i.e., mass of volatile suspended solids, VSS). When grown on a soluble substrate, microorganisms have an ash content of about 15%, and thus the value of Y when expressed as VSS will be slightly less than the value of Y when expressed as TSS. As will be discussed later, there are certain advantages to expressing biomass concentrations on a COD basis rather than on a TSS or VSS basis, and thus yields are sometimes expressed as the amount of biomass COD formed per unit of substrate COD

removed from the medium. Nevertheless, in engineering practice it remains more convenient to represent yield on a TSS or VSS basis, as discussed in Section 5.1.4, and this convention will be used throughout this book. It is often helpful to convert between the various ways of expressing yield. If we assume an empirical formula for the organic (i.e., ash-free) portion of biomass of $C_5H_7O_2N$, the COD of that organic portion can be calculated to be 1.42 g COD/g VSS.[44] Furthermore, if we assume the ash content of biomass to be 15%, the theoretical COD of biomass is 1.20 g COD/g TSS. Although theoretically based, these conversion factors find broad use and will be adopted herein.

The nature of the substrate influences the yield. Hadjipetrou et al.[32] summarized data from one species, *Aerobacter aerogenes*, which was grown in unrestricted batch growth in minimal media on a number of substrates, and found Y to vary from 0.40 to 0.56 mg biomass COD formed per mg substrate COD removed. Recognizing that the yield expressed on the basis of cell COD formed per unit of substrate COD removed is a measure of the amount of energy available in the substrate that was conserved through cell synthesis, it can be seen that 40–56% of the available energy was conserved while 60–44% was expended.

The species of organism will also affect Y, although the effect will not be as great as the effect of substrate. Payne[94] collected Y values for eight bacterial species growing aerobically on glucose in minimal media and found them to vary from 0.43 to 0.59 mg biomass COD formed per mg substrate COD removed. The data were from a number of different published reports and thus some of the variation may be due to differences in experimental conditions, rather than to species. Nevertheless, they clearly show that the microbial species has an impact.

The growth environment, including media complexity, type of terminal electron acceptor, pH, and temperature will all affect Y.[37] As explained above, biomass grown in complex media will have only slightly higher Y values than biomass grown in minimal media, whereas biomass grown with oxygen as the terminal electron acceptor will exhibit significantly higher yields than biomass grown with nitrate as the acceptor. The yield from fermentations will depend on the reduced end products and the method of expressing the yield. If Y is expressed on the basis of the amount of the original substrate removed, ignoring the COD returned to the medium as reduced end products, the value will be very small, on the order of 0.03–0.04 mg biomass COD formed per mg substrate COD removed. However, when expressed on the basis of the COD actually utilized (accounting for the COD remaining as reduced end products), the Y value is not much different from that obtained with aerobic cultures.[1] On the other hand, when methane is produced, so that most of the reduced end product is lost from the system as a gas, then the COD removed from the solution is actually much higher than the COD utilized by the microorganisms, making the yield per unit of COD removed about an order of magnitude lower than for aerobic growth. The pH of the medium has long been known to affect microbial growth, but the quantitative effects are unclear. The yield is likely, however, to have a maximum around pH 7 because that is optimal for so many physiological functions. Temperature also affects Y, as shown in Figure 2.5.[86] Although the significance of temperature is apparent, no generalizations can be made and most engineers assume that Y is constant over the normal physiological temperature range. A final factor that may influence Y is the composition of the microbial community. When it is heterogeneous, the waste products from one species serve as growth factors for another, thereby converting a seemingly minimal medium into a complex one. Consequently, it might be anticipated that the yields from mixed microbial cultures would be slightly higher than those from pure cultures growing on the same medium. A comparison of the two revealed this to be the case.[41]

2.4.1.5 Constancy of Y in Biochemical Operations

Biochemical operations use mixed microbial communities to treat wastewaters containing mixtures of substrates. Thus it is apparent that Y will depend on both the character of the wastewater and the particular community that develops on it. It is important that this variability be recognized by engineers designing biochemical operations, because then the estimated yield values will be interpreted in an appropriate way. As will be seen in Chapter 3, similar conclusions can be reached about the

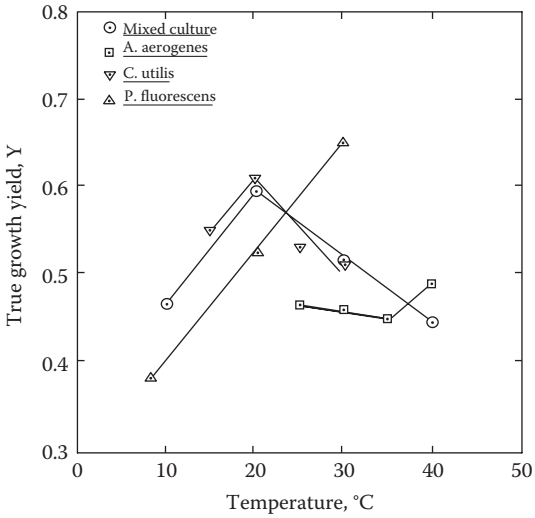

FIGURE 2.5 Effect of temperature on the true growth yield, Y. The units of Y are mg biomass formed per mg substrate COD removed. (Reprinted from Muck, R. E. and Grady Jr., C. P. L., Temperature effects on microbial growth in CSTR's. *Journal of the Environmental Engineering Division, ASCE,* 100:1147–63, 1974. With permission from the American Society of Civil Engineers.)

kinetic parameters associated with biochemical operations. This means that designers must utilize considerable judgment and allow for uncertainty. This situation does not prevent generalities from being made, however. For example, examination of a large number of yield values indicates that Y will generally lie within the range of 0.48–0.72 mg biomass COD formed per mg substrate COD utilized for aerobic heterotrophs degrading carbohydrates.[103] Under similar conditions, Y values for growth on a number of xenobiotic compounds, including substituted phenols, benzenes, and phthalate esters, lay within the range of 0.20–0.60 mg biomass COD formed per mg substrate COD removed.[29] One study[61] reported the range of yield values for aerobic nitrifying bacteria to be from 0.06 to 0.35 mg biomass COD per mg nitrogen oxidized, with values for NOB being lower than those for AOB. Likewise, another study[132] reported the Y value for aerobic NOB to be 0.12 mg biomass COD per mg nitrogen oxidized and the value for aerobic AOB to be 0.47. However, it is recognized that the traditional method for estimating yield may overestimate values and that the yields for both AOB and NOB may be similar at 0.07 and 0.08 mg biomass COD per mg nitrogen oxidized.[10] Although ranges such as these provide the engineer with an idea of the magnitudes to be expected, designs should only be based on estimates of Y obtained from laboratory- and pilot-scale studies with the particular wastewater to be treated.

2.4.2 Maintenance, Endogenous Metabolism, Decay, Lysis, and Death

The yield values in the preceding section are those that result when all energy obtained by the biomass is being channeled into synthesis. Energy for synthesis is not the only energy requirement for microorganisms, however. They must also have energy for maintenance.[99]

Cellular processes, whether mechanical or chemical, require energy for their performance, and unless a supply is available these essential processes will cease and the cell will become disorganized and die. Mechanical processes include motility, osmotic regulation, molecular transport, maintenance of ionic gradients, and in the case of some Eucarya, cytoplasmic streaming. While it might be argued that motility can be dispensed within some microorganisms, this argument would not hold for all because some require motility to find food. Osmotic regulation is quite important in all cells, even those protected by a rigid cell wall, and pump mechanisms, such as contractile

vacuoles, exist in cells to counteract the normal tendency of osmotic pressure to pump water into them. Cell membranes are permeable to many small molecules, such as amino acids, and because of the high concentrations within the cell these tend to diffuse into the medium. Active transport mechanisms operate to bring such molecules into the cell against the concentration gradient. Of a similar nature is the necessity for maintaining an ionic gradient across the cell membrane, which is closely linked to the proton motive force responsible for ATP synthesis. Maintenance of this gradient is thought to be a major consumer of maintenance energy.[118] Finally, cytoplasmic streaming and the movement of materials within Eucarya are often required for their proper functioning. They also require energy.

Chemical factors also contribute to maintenance energy needs. Microbial cells represent chemical organization and many of the components within them have higher free energies than the original compounds from which they were formed. In general, because of this organization, energy must be available to counteract the normal tendency toward disorder (i.e., to overcome entropy). The chemical processes contributing to the energy requirement for maintenance are those involved in resynthesis of structures such as the cell wall, flagella, the cell membrane, and the catabolic apparatus. For example, one study[72] suggested that energy for the resynthesis of proteins and nucleic acids was an important portion of the maintenance energy requirement for *Escherichia coli*.

A major point of controversy in the microbiological literature has concerned the impact on the maintenance energy requirement of the rate at which a culture is growing. Early investigations[99] suggested that the need for maintenance energy was independent of growth rate, but later research indicated the opposite.[118] Nevertheless, engineers generally consider maintenance energy needs to be independent of growth rate in biochemical operations for wastewater treatment and that is the approach that will be adopted in this book.

Given the existence of a need for maintenance energy, what energy sources can be used to supply it? The answer to that question depends on the growth conditions of the microorganisms. If an external (exogenous) energy supply is available, a portion of it will be used to meet the maintenance energy requirement and the remainder will be used for synthesis. As the rate of energy supply is decreased, less and less will be available for new growth and thus the net, or observed, yield will decline. When the point is reached at which the rate of energy supply just balances the rate at which energy must be used for maintenance, no net growth will occur because all available energy will be used to maintain the status quo. If the rate of energy supply is reduced still further, the difference between the supply rate and the maintenance energy requirement will be met by the degradation of energy sources available within the cell (i.e., by endogenous metabolism). This will cause a decline in the mass of the culture. Finally, if no exogenous energy source is available, all of the maintenance energy needs must be met by endogenous metabolism. When the point is reached at which all endogenous reserves have been exhausted, the cells deteriorate and die or enter a resting state.

The nature of the materials serving as substrates for endogenous metabolism depends on both the species of the microorganism and the conditions under which the culture was grown. For example, when *E. coli* is grown rapidly in a glucose-mineral salts medium it stores glycogen.[72] If those cells are then placed in an environment devoid of exogenous substrate, they utilize the glycogen as an endogenous energy source. Amino acids and proteins show little net catabolism until the glycogen is gone. When grown in tryptone medium on the other hand, *E. coli* accumulates little glycogen. As a result, endogenous metabolism utilizes nitrogenous compounds immediately. Other organisms use still other compounds, including ribonucleic acid (RNA) and the lipid poly-β-hydroxybutyrate (PHB).

The amount of biomass actually formed per unit of substrate used in a biochemical operation, referred to as the observed yield (Y_{obs}), is always less than the true growth yield (Y). One reason for this is the need for maintenance energy. The more energy that must be expended for maintenance purposes, the less available for synthesis, and the smaller the quantity of biomass formed per unit of substrate degraded. Other factors also contribute to the difference, however. For example, consider the effect of predation. In a complex microbial community such as that found in the activated sludge process, protozoa and other Eucarya prey on the bacteria, reducing the net amount of

biomass formed. To illustrate the effect of predation, assume that the value of Y for bacteria growing on glucose is 0.60 mg bacterial biomass COD formed per mg of glucose COD used. Thus, if 100 mg/L of glucose COD were used, 60 mg/L of bacterial biomass COD would result. Now assume that the value of Y for protozoa feeding on bacteria is 0.70 mg protozoan biomass COD formed per mg of bacterial biomass COD used. If the protozoa consumed all of the bacteria resulting from the glucose, the result would be 42 mg/L of protozoan biomass. As a consequence, if we observed only the net amount of biomass formed, without distinction as to what it was, we would conclude that 42 mg/L of biomass COD resulted from the destruction of 100 mg/L of glucose COD. Therefore, we would conclude that the observed yield was 0.42, which is less than the true growth yield for bacteria growing on glucose. Macroscopically, it is impossible to distinguish between the various factors acting to make the observed yield less than the true growth yield. Consequently, environmental engineers lump them together under the term "microbial decay," which is the most common way they have modeled their effect in biochemical operations.[61]

Another process leading to a loss of biomass in biochemical operations is cell lysis.[74] The growth of bacteria requires coordination of the biosynthesis and degradation of cell wall material to allow the cell to expand and divide. The enzymes responsible for hydrolysis of the cell wall are called autolysins and their activity is normally under tight regulation to allow them to act in concert with biosynthetic enzymes during cell division. Loss of that regulation, however, will lead to rupture of the cell wall (lysis) and death of the organism. When the cell wall is ruptured, the cytoplasm and other internal constituents are released to the medium where they become substrates for other organisms growing in the culture. In addition, the cell wall and cell membranes, as well as other structural units, begin to be acted upon by hydrolytic enzymes in the medium, solubilizing them, and making them available as substrates as well. Only the most complex units remain as cell debris, which is solubilized so slowly that it appears to be refractory in most biochemical operations.[78,81] The arguments for how lysis results in the loss of biomass are similar to those associated with predation, illustrated above. The yield exhibited by bacteria growing on the soluble products released by lysis is of the same magnitude as the yield associated with growth on other biogenic substrates. Consequently, if 100 mg/L of biomass is lysed, only 50–60 mg/L of new biomass will result from regrowth on the lysis products. Thus, the net effect of lysis and regrowth is a reduction in biomass within the system. In general, starvation itself does not initiate lysis, although the events that trigger it are not yet clear. Nevertheless, engineers seeking to model the decline in observed yield associated with situations in which the microbial community is growing slowly have focused on cell lysis as the primary mechanism.[18,38]

The final event impacting on the amount of active biomass in a biochemical operation is death. Traditionally, a dead cell has been defined as one that has lost the ability to divide on an agar plate[100] and studies based on this definition have shown that a large proportion of the microorganisms in slowly growing cultures are nonviable or dead.[100,119] In addition, a large number of studies using indirect evidence involving comparisons of substrate removal rates and enzyme activities have concluded that large portions of the MLSS in wastewater treatment systems are inactive.[126] However, a later study[73,74] using more sophisticated techniques for identifying dead bacteria, has suggested that a very low fraction of the cells present at low growth rates are actually dead. Instead, many are simply nonculturable by standard techniques, although they are still alive. Furthermore, the more recent work[73] suggests that dead cells do not remain intact for long, but rather lyse, leading to substrates and biomass debris, as discussed above. The presence of biomass debris acts to make the mass of viable microorganisms less than the mass of suspended solids in the system. Thus, it appears that direct consideration of cell death is not warranted.[73,74] Rather, the fact that only a portion of the MLSS in a biological wastewater treatment system is actually viable biomass can be attributed to the accumulation of biomass debris rather than to the presence of dead cells.

In summary, as a result of several mechanisms, biochemical reactors exhibit two important characteristics: the observed yield is less than the true growth yield and active, viable bacteria make up only a fraction of the "biomass." One simplified conceptualization of the events leading to these

characteristics is that bacteria are continually undergoing death and lysis, releasing organic matter to the environment in which they are growing. Part of that organic matter is degraded very, very slowly making it appear to be resistant to biodegradation and causing it to accumulate as biomass debris. As a consequence, only a portion of the biomass is actually viable cells. The remainder of the released organic matter is used by the bacteria as a food source, resulting in new biomass synthesis. However, because the true growth yield is always less than one, the amount of new biomass produced is less than the amount destroyed by lysis, thereby making the observed yield for the overall process less than the true growth yield on the original substrate alone.

2.4.3 Formation of Extracellular Polymeric Substances and Soluble Microbial Products

In suspended growth treatment systems the microorganisms grow as floc particles whereas in attached growth systems they form biofilms. Both floc particles and biofilms are created by a common mechanism; through the presence of extracellular polymeric substances (EPSs). Several types of EPS are involved in bioflocculation. Polysaccharides have received the most study and are generally thought to be of major importance.[14,24,122] Nevertheless, proteins also play an important role.[42] Possible sources of EPS are formation by microbial metabolism, release by cell lysis, and the wastewater itself.[122] Evidence for the role of the wastewater itself comes from the observation that flocculation in activated sludge systems treating industrial wastewaters, which contain a limited number of organic compounds, is often more difficult than in systems treating domestic wastewaters, which contain a rich variety of large molecular weight organic materials. Nevertheless, the most important sources of EPS are metabolism[60] and cell lysis. The EPS is produced by both protozoa[16] and bacteria,[8] although the relative contribution of the two is unknown. Nevertheless, the formation of bulk EPS is associated with cell synthesis and its rate of formation is considered to be proportional to the rate of active biomass growth.[60]

Much of the soluble organic matter in the effluent from a biological reactor is of microbial origin and is produced by the microorganisms as they degrade the organic substrate in the influent to the bioreactor. The major evidence for this phenomenon has come from experiments in which single soluble substrates of known composition were fed to microbial cultures and the resulting organic compounds in the effluent were examined for the presence of the influent substrate.[105] The bulk of the effluent organic matter was not the original substrate and was of higher molecular weight, suggesting that it was of microbial origin. These soluble microbial products (SMPs) are thought to arise from two processes, one growth associated and the other nongrowth associated. Growth associated SMP formation results directly from biomass growth and substrate utilization. As such, it is coupled to those events through another yield factor, the microbial product yield, Y_{MP}, and the biodegradation of one unit of substrate results in the production of Y_{MP} units of products. Values of Y_{MP} for a variety of organic compounds have been found to be less than 0.1.[29] Nongrowth associated SMP formation is related to decay and lysis and results in biomass associated products. They are thought to arise from the release of soluble cellular constituents through lysis and from hydrolysis of bound EPS.[60] The SMPs have a variety of biochemical forms, including humic and fulvic acids, polysaccharides, proteins, nucleic acids, organic acids, amino acids, and others.[4] They are thought to be biodegradable, although some at a very low rate.[60]

Although a number of researchers have studied the nature of both EPS and SMP it is not easy to generalize about them, perhaps because of the difficulties associated with their isolation and analysis. Nevertheless, a few researchers have attempted to model the contribution of such products to the organic matter discharged from wastewater treatment systems.[59,91,105] Even though SMPs are not included in most models of biological wastewater treatment, an awareness of their existence is necessary for an accurate understanding of the response of those systems. For example, one impact of SMPs is to make the concentration of soluble organic matter in the effluent from a biological reactor roughly proportional to the influent concentration.

2.4.4 Solubilization of Particulate and High Molecular Weight Soluble Organic Matter

Bacteria can only take up and degrade soluble organic matter of low molecular weight. All other organic material must be attacked by extracellular enzymes that release low molecular weight compounds that can be transported across cellular membranes. Many organic polymers, particularly those of microbial origin, such as cell wall components, proteins, and nucleic acids, are composed of a few repeating subunits connected by bonds that can be broken by hydrolysis. Consequently, the microbial process of breaking particulate and high molecular weight soluble organic compounds into their subunits is commonly referred to as hydrolysis, even though some of the reactions involved may be more complicated.

Hydrolysis reactions play two important roles in biochemical reactors for wastewater treatment. First, they are responsible for the solubilization of cellular components released as a result of cell lysis, preventing their buildup in the system. Because cell lysis occurs in all microbial systems, hydrolysis reactions are even important in bioreactors receiving only soluble substrate. Second, many biochemical operations receive particulate organic material, in which case hydrolysis is essential to bring about the desired biodegradation. In spite of its central position in the functioning of biochemical operations, relatively few studies have sought to understand the kinetics and mechanisms of hydrolysis.[85] Nevertheless, it has important impacts on the outcome of biochemical operations and must be considered for a complete understanding of their functioning.

2.4.5 Ammonification

Ammonification is the name given to the release of ammonia nitrogen as amino acids and other nitrogen containing organic compounds undergo biodegradation. It occurs as a normal result of the biodegradation process, during which amino groups are liberated and excreted from the cell as ammonia. The rate of ammonification will depend on the rate of nitrogen containing substrate utilization and the carbon to nitrogen ratio of that substrate. Ammonification is very important in wastewater treatment processes for nitrogen control because organic nitrogen is not subject to oxidation by nitrifying bacteria. They can only oxidize nitrogen to nitrate after it has been converted to ammonia and released to the medium.

2.4.6 Phosphorus Uptake and Release

If a suspended growth bioreactor system is configured as two zones in series with the first zone anaerobic and the second aerobic, PAOs will proliferate and store large quantities of inorganic phosphate as polyphosphate, thereby allowing phosphorus removal from the wastewater via biomass wastage. Although PAOs are often present in significant numbers in totally aerobic suspended growth cultures, they only develop the ability to store large quantities of phosphate when they are subjected to alternating anaerobic and aerobic conditions by being recycled between the two zones.[65] This follows from their unique capability to store carbon at the expense of phosphate under anaerobic conditions and to store phosphate at the expense of carbon under aerobic conditions. Multiple scenarios have been postulated to explain PAO metabolism and they differ primarily in the source of reducing power needed to form poly-β-hydroxyalkanoate (PHA), the organic acid storage molecule that ultimately fuels phosphate uptake and growth of PAOs. The current, most widely accepted conceptual model was developed by Arun et al.[2] (called the Mino model) and adapted by several others. This model proposed that glycogen was the source of reducing power that resulted in PHA formation, and this proposal has since been unequivocally confirmed.[95] Although some contend that the tricarboxylic acid (TCA) cycle plays a role during the anaerobic phase by producing reducing power under certain circumstances, we will present only the glycogen-fueled model. For more details about the features of PAO metabolism, the reader is referred to a comprehensive

review by Oehman et al.[90] In recognition of the important contribution that Mino and colleague's made in initially proposing the role of glycogen in PAO metabolism, we call the metabolic PAO model presented in Figure 2.6 the modified Mino PAO model. Filipe et al.[23] complemented the PAO metabolic model by elucidating an anaerobic model for GAOs, and Zeng et al.[133] proposed an aerobic metabolic GAO model. For this reason, we call the combined model, shown in Figure 2.7, the Filipe–Zeng GAO model. These models are described below.

2.4.6.1 The Modified Mino PAO Model

We will first consider the events occurring in the anaerobic zone. Because of fermentations that occur in sewers, much of the soluble organic matter in domestic wastewater is in the form of acetate, pyruvate, and other short chain fatty acids. Furthermore, when the wastewater enters an anaerobic bioreactor, additional quantities of fatty acids are formed by fermentative reactions performed by non-PAO facultative heterotrophs. To simplify the presentation of the model, we will use acetate as the model fatty acid. As indicated in Figure 2.6 (anaerobic), acetate is transported across the cell membrane using the energy contained in the proton motive force (represented by H^+).[108] Once inside, it is activated to acetyl-CoA by coupled ATP hydrolysis, yielding ADP. The majority of the ATP is synthesized in concert with the hydrolysis of stored polyphosphate (Poly-P_n), releasing a light metal cation (Me^+) bound phosphate from the cell.[90] The light metal cation is typically potassium or magnesium and its release helps maintain a charge balance. Maintenance of PAOs under anaerobic conditions is supported by the polyphosphate-derived ATP. A carbon storage molecule, PHA, is synthesized from acetyl-CoA using reducing power produced by the metabolism of glycogen[2,83,95] and possibly the TCA cycle.[90,134] Degradation of the carbohydrate storage polymer glycogen results in the production of pyruvate via glycolysis through the Entner–Doudoroff (ED) or Embden–Meyerhof–Parnas (EMP) pathway, depending on the type of PAO, thereby providing some of the ATP required to convert acetate to acetyl-CoA and some of the reducing power needed for PHA synthesis. Pyruvate, in turn, is converted to acetyl-CoA and carbon dioxide, with the electrons and protons released supporting the generation of reducing power required for PHA synthesis. Almost all the acetate carbon taken up is conserved in the synthesis of PHA.

When the wastewater and the associated biomass enter the aerobic zone, the wastewater is low in soluble organic matter, but the PAOs contain large PHA reserves. Furthermore, the wastewater is rich in inorganic phosphate, while the PAOs have low polyphosphate levels. Because they now have oxygen as an electron acceptor in the aerobic zone (or nitrate in an anoxic zone), the PAOs perform normal aerobic/anoxic metabolism for growth by using the stored PHA as their carbon and energy source, generating ATP through electron transport phosphorylation, as illustrated in Figure 2.6 (aerobic). Furthermore, polyphosphate synthesis is stimulated, thereby removing phosphate and associated light metal cations from solution and regenerating the stored polyphosphate in the cells. At the same time, glycogen is replenished by PHA degradation through gluconeogenesis. Because of the large amount of energy provided by the aerobic metabolism of the stored PHA, the PAOs grow and increase their capacity to take up all of the phosphate released in the anaerobic zone plus the phosphate originally present in the wastewater.

The continual cycling between the anaerobic and aerobic zones gives PAOs a competitive advantage over ordinary heterotrophic bacteria, because without the capability to make and use polyphosphate, the ordinary heterotrophs are not able to take up organic matter in the anaerobic zone. Because most of the carbon and energy in the wastewater are stored in PHA and glycogen, the ordinary heterotrophs are deprived of the materials needed for growth. While most systems that remove phosphate through the use of PAOs employ aerobic zones for the regeneration of the stored polyphosphate, some PAOs can use nitrate and moderate concentrations of nitrite as alternative electron acceptors,[49,54] allowing anoxic zones to be used as well. Although the use of anaerobic-anoxic BPR can be less expensive due to lower aeration costs, as well as other benefits,[90] the significantly slower rate of phosphorus uptake by denitrifying PAOs[22] can be a disadvantage to utilities that have to achieve the lowest possible effluent phosphorus concentration.

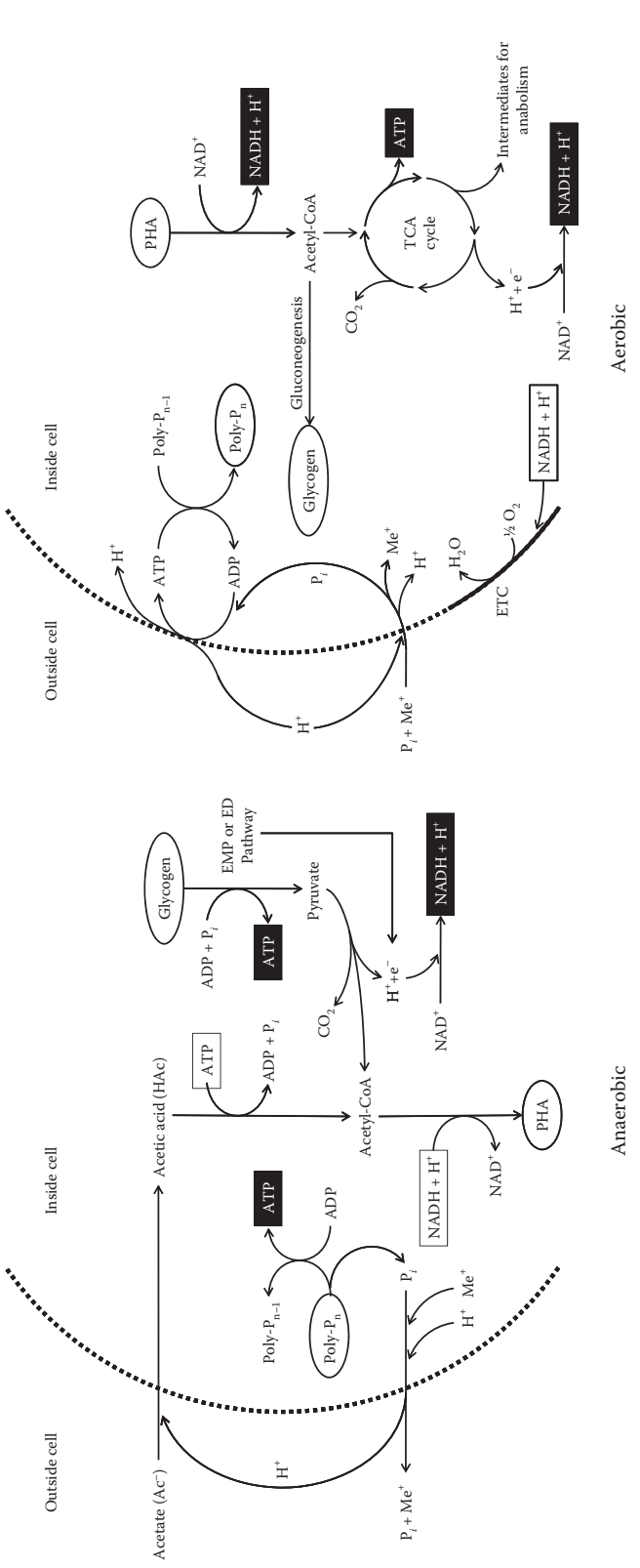

FIGURE 2.6 Schematic diagram depicting the modified Mino PAO model. Storage compounds are highlighted with ovals. The generation of ATP or reducing power is emphasized using black boxes. The consumption of ATP or reducing power is emphasized using white boxes.

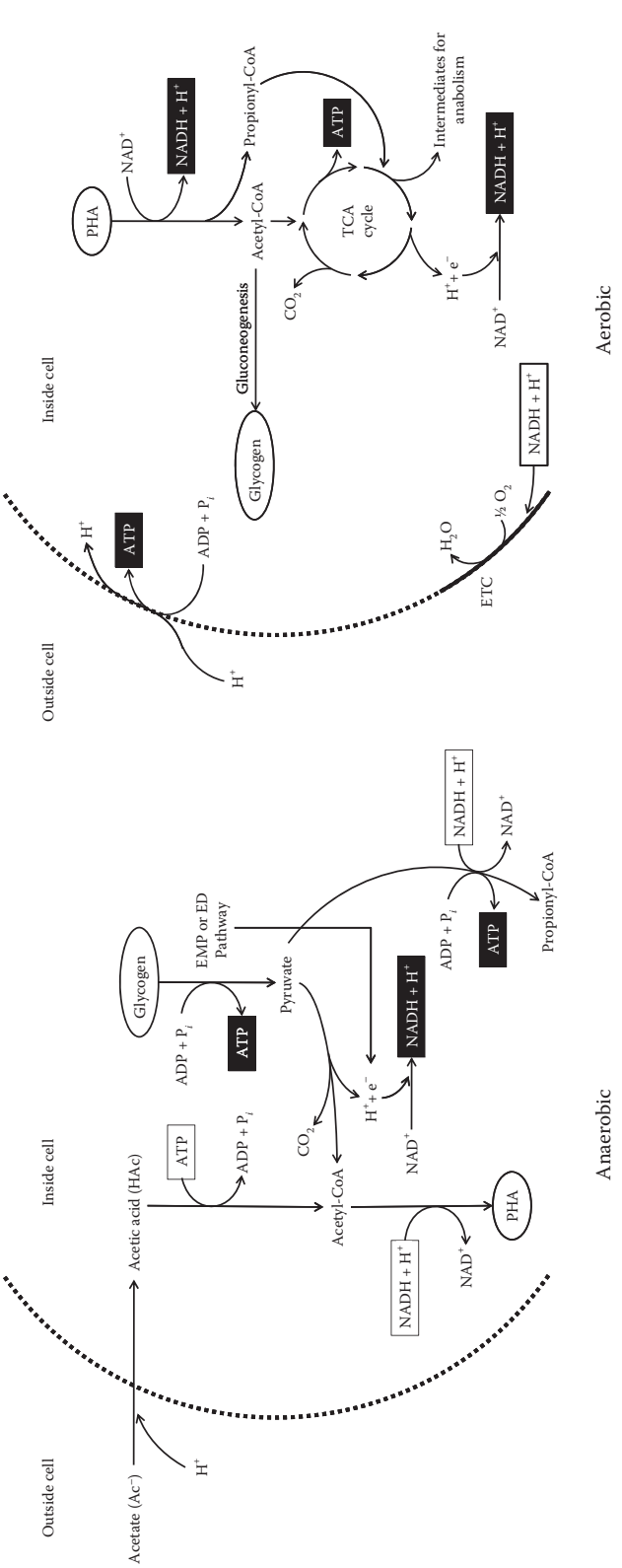

FIGURE 2.7 Schematic diagram depicting the Filipe–Zeng GAO model. Storage compounds are highlighted with ovals. The generation of ATP or reducing power is emphasized using black boxes. The consumption of ATP or reducing power is emphasized using white boxes.

2.4.6.2 Filipe–Zeng GAO Model

The metabolism of GAOs is similar to PAOs in many ways, but sufficiently distinct to deserve description (Figure 2.7).[23,133] Acetate uptake by GAOs is also fueled by proton motive force but through a slightly different process than by PAOs.[108] Inside the cell, many of the metabolic processes are the same except that polyphosphate storage molecules are not present, and anaerobic maintenance is fueled by glycogen. Therefore, glycogen degradation serves as the primary source of ATP to form acetyl CoA and reducing power to form PHA. A whole or partial TCA cycle may also participate in the generation of reducing power.[90] The stoichiometry of acetate-derived PHA in GAOs results in residual reducing power that is directed into the synthesis of PHA through two intermediates, acetyl-CoA (similar to PAOs) and propionyl-CoA (from the TCA cycle, which is not common in PAOs). Because of the prominent role of propionyl-CoA in GAO metabolism, an important PHA formed by GAOs is poly-β-hydroxyvalerate (PHV), whereas PAOs typically form little PHV. Under aerobic conditions, the process of consuming PHA to fuel growth and glycogen formation is quite similar to PAOs. Aerobic glycogen formation is believed to occur by gluconeogenesis of the acetyl CoA formed by the hydrolysis of PHA.[133]

2.4.7 Overview

A diagram depicting the overall sum of the events occurring in an aerobic bioreactor receiving a soluble substrate is shown in Figure 2.8. Bacteria consume the soluble substrate (S_{S1}) and grow, leading to more bacteria, with the relationship between substrate consumption and biomass growth being given by the true growth yield, Y. There will also be soluble microbial product (S_{MP}) and extracellular polymeric substance (S_{EPS}) formation associated with that substrate consumption and growth. Concurrently with growth, the biomass will be undergoing decay and lysis, releasing soluble (S_{S2}) and particulate (X_S) substrate to the medium. Cell debris (X_D), which is degraded so slowly that it appears to be nonbiodegradable, and biomass associated products (S_D) are also released. The

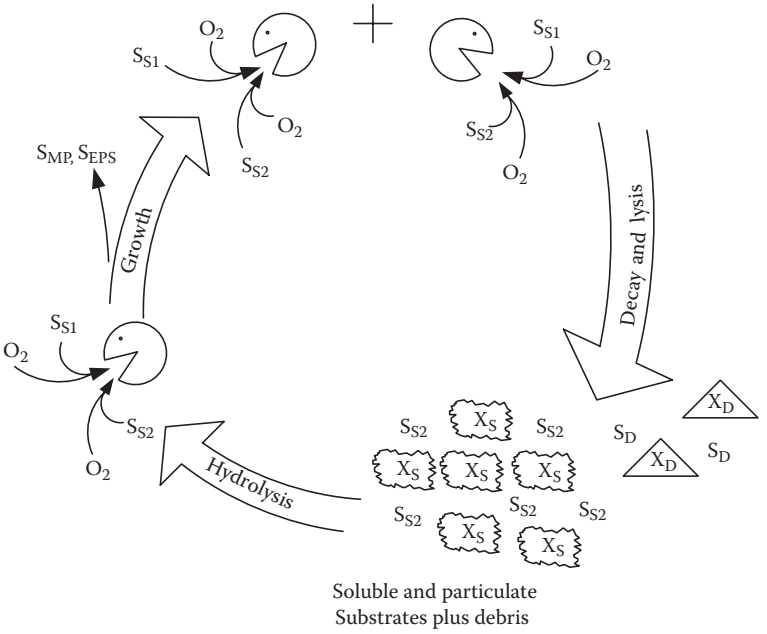

FIGURE 2.8 Overview of fundamental events occurring in an aerobic bioreactor receiving a soluble substrate (S_{S1}). (Adapted from Mason, C. A., Bryers, J. D., and Hamer, G., Activity, death and lysis during microbial growth in a chemostat. *Chemical Engineering Communications,* 45:163–76, 1986.)

particulate cell fragments (X_S) undergo hydrolysis, freeing more soluble substrate (S_{S2}) that can be used by the cells. Part of the microbial products may undergo biodegradation, but others may be degraded so slowly that they appear inert. As might be imagined by the previous discussion in this section, more complicated conceptualizations could be depicted. However, this one contains the essential elements required to model biological processes and it will be used in later chapters for that purpose.

2.5 KEY POINTS

1. Biochemical operations use the carbon and nitrogen cycles to remove organic and nitrogenous pollutants from wastewaters.
2. The microorganisms in biochemical operations can be classified in several ways. Among the most important are: the type of electron donor used, the type of electron acceptor employed, their physical growth characteristics, and their function.
3. The microorganisms in aerobic/anoxic suspended growth bioreactors may be divided into five overlapping groups: floc-forming organisms, saprophytes, nitrifying bacteria, predators, and nuisance organisms.
4. Attached growth bioreactors have more diverse microbial communities encompassing more trophic levels than suspended growth bioreactors.
5. Methanogenic anaerobic cultures are highly interdependent ecosystems with many complex interactions between Bacteria and Archaea. Acetic acid and H_2 play a central role in those interactions, being products of the Bacteria and substrates for the Archaea.
6. There are two major groups of methanogens: those that oxidize H_2 and those that cleave acetic acid. Both are essential to the proper functioning of anaerobic cultures receiving complex substrates.
7. In most situations, biomass growth and substrate utilization are coupled with the true growth yield, Y, serving as the coupling factor. The yield is the amount of biomass formed per unit of substrate removed. Its value depends on the nature of the substrate, the organism involved, and the growth environment.
8. Heterotrophic bacteria obtain their energy from the oxidation of organic carbon. Hence, chemical oxygen demand (COD), which is a measure of available electrons, is a convenient way in which to express the concentration of organic matter in wastewaters. When an organic compound is mineralized, all of the electrons available in it must end up either in the biomass formed or in the terminal electron acceptor. Consequently, COD is also a conceptually convenient technique for expressing the concentration of biomass, although in engineering practice biomass concentrations are usually expressed as total suspended solids (TSS) or volatile suspended solids (VSS). Theoretical conversion factors can be used to convert from one unit of expression to another.
9. Yield values for heterotrophic biomass cover a very broad range, but seldom exceed 0.75 mg biomass COD formed per mg substrate COD removed because of the energy required for synthesis.
10. As a result of maintenance energy needs and decay, death, and lysis, biochemical reactors exhibit two characteristics: the observed yield is less than the true growth yield and active viable bacteria make up only a fraction of the "biomass."
11. Soluble microbial product formation is associated with substrate utilization and with biomass decay and lysis. As a consequence, much of the soluble organic matter leaving a biochemical operation is of microbial origin.
12. Extracellular polymeric substances (EPSs) are composed of biomolecules and are key in achieving cellular aggregation, water retention, the accumulation of enzymatic activity and nutrients, and in protecting cells against toxins. The rate of EPS formation is proportional to the rate of active biomass growth.

13. Hydrolysis reactions are important for the biodegradation of particulate substrates and cellular components released by biomass death and lysis.
14. Ammonification is the release of ammonia-N as nitrogen containing organic compounds undergo biodegradation.
15. Phosphate accumulating organisms (PAOs) will only store large amounts of phosphorus as polyphosphate granules when they are cycled between substrate-rich anaerobic and substrate-poor aerobic/anoxic environments.

2.6 STUDY QUESTIONS

1. Draw a sketch of the nitrogen cycle, labeling all reactions. Then, explain the following terms and their importance in biochemical operations: ammonification, assimilation, nitrification, denitrification, and assimilative reduction.
2. Define or explain the following terms and their use in classifying the microorganisms in biochemical operations: electron donor, electron acceptor, heterotroph, autotroph, nitrifier, denitrifier, methanogen, obligate aerobe, obligate anaerobe, facultative anaerobe, biofloc, primary degrader, and secondary degrader.
3. Describe the roles of microorganisms in each of the following groups commonly found in aerobic/anoxic suspended growth bioreactors: floc-forming organisms, saprophytes, nitrifying bacteria, predators, and nuisance organisms.
4. Draw a sketch depicting the multistep nature of methanogenic anaerobic cultures and use it to describe the roles of the major groups of microorganisms involved.
5. Why is the maintenance of a low partial pressure of H_2 necessary to the proper functioning of a methanogenic anaerobic culture? What is the role of methanogens in the maintenance of the required conditions?
6. There are two major groups of methanogens. Describe them, list their growth characteristics, and contrast their roles in anaerobic cultures.
7. Why does the value of the true growth yield, Y, depend on the nature of the substrate, the microorganism involved, and the growth environment?
8. Why is it convenient to express the concentrations of organic substrates and biomass in COD units?
9. Give a "typical" yield value for heterotrophic biomass growing on carbohydrates and then explain why there is considerable variability associated with Y in biochemical operations.
10. Explain why the observed yield in a biochemical reactor is less than the true growth yield. While so doing, explain what is meant by the term "decay."
11. Why does cell lysis in a biochemical operation make the observed yield less than the true growth yield and the viability less than 100%?
12. What is the difference between growth associated and nongrowth associated product formation?
13. Why are hydrolysis reactions important to the performance of all biochemical operations, even those receiving only soluble substrate?
14. Describe the scenarios that have been postulated to explain the functioning of phosphate accumulating bacteria.

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3 Stoichiometry and Kinetics of Aerobic/Anoxic Biochemical Operations

Stoichiometry is concerned with the relationships between the quantities of reactants and products in chemical reactions. Kinetics is concerned with the rates at which reactions take place. Because stoichiometry quantitatively relates a change in one reactant (product) to the change in another, once the reaction rate of one reactant (product) is known, stoichiometry may be used to determine the reaction rate of another in the reaction. In this chapter we will first examine these relationships on a generalized basis. Then we will apply them to the major biochemical events discussed in Chapter 2 and examine the expressions that will be used to model the theoretical performance of biochemical operations in Parts II and IV.

3.1 STOICHIOMETRY AND GENERALIZED REACTION RATE

3.1.1 Alternative Bases for Stoichiometry

Stoichiometric equations are usually derived in molar units, but they are not the most convenient units for our purposes. This is because we must write mass balance equations for the various constituents being acted upon in a biochemical operation in order to model its performance. Thus, it would be more convenient if the stoichiometric equations for the reactions were written in mass units. Consequently, we need to know how to convert a molar-based stoichiometric equation into a mass-based one. Furthermore, we saw in Chapter 2 that microorganisms gain their energy from oxidation/reduction reactions in which electrons are removed from the electron donor and passed ultimately to the terminal electron acceptor. This suggests that it would also be convenient to write electron balances. Unfortunately, as we saw earlier, we usually don't know the exact composition of the electron donor in a wastewater, making this difficult to do. However, we can experimentally determine the chemical oxygen demand (COD), which is a measure of available electrons, of the various constituents. Thus, we can accomplish the same thing by writing a mass balance on COD for each of the constituents that undergo a change in oxidation state. Consequently, we also need to know how to convert molar- or mass-based stoichiometric equations into COD-based equations.

The general formula for a stoichiometric equation can be written as[76]

$$a_1A_1 + a_2A_2 + \cdots + a_kA_k \rightarrow a_{k+1}A_{k+1} + a_{k+2}A_{k+2} + \cdots + a_mA_m, \tag{3.1}$$

where A_1 through A_k are the reactants and a_1 through a_k are their associated molar stoichiometric coefficients, A_{k+1} through A_m are the products, and a_{k+1} through a_m are their molar stoichiometric coefficients. Two characteristics allow recognition of a stoichiometric equation as being molar-based. First, the charges are balanced. Second, the total number of moles of any given element in the reactants equals the number of moles of that element in the products.

When writing a mass-based stoichiometric equation it is common practice to normalize the stoichiometric coefficients relative to one of the reactants or products. Thus, each normalized mass-based stoichiometric coefficient represents the mass of the particular reactant used or product formed relative to the mass of the reference reactant used or product formed. If A_1 is the component

that we want to use as the basis for our mass-based stoichiometric equation, its stoichiometric coefficient would be 1.0 and the new mass-based stoichiometric coefficient for every other component (referred to as a normalized stoichiometric coefficient, Ψ_i) would be calculated from

$$\Psi_i = \frac{(a_i)(MW_i)}{(a_1)(MW_1)}, \tag{3.2}$$

where a_i and MW_i are the molar stoichiometric coefficient and molecular weight, respectively, of component A_i, and a_1 and MW_1 have the same meanings for the reference component. Thus, the equation becomes:

$$A_1 + \Psi_2 A_2 + \cdots + \Psi_k A_k \rightarrow \Psi_{k+1} A_{k+1} + \Psi_{k+2} A_{k+2} + \cdots + \Psi_m A_m. \tag{3.3}$$

Two characteristics can be used to identify this type of stoichiometric equation: the charges do not appear to be balanced and the total mass of reactants equals the total mass of products. In other words, the sum of the stoichiometric coefficients for the reactants equals the sum of the stoichiometric coefficients for the products. The latter characteristic makes a mass-based stoichiometric equation well suited for use in mass balance equations for biochemical reactors.

A similar approach can be used to write the stoichiometric equation in terms of compounds or components that change the oxidation state by taking advantage of COD units.[76] In this case, the normalized stoichiometric coefficients are referred to as COD-based coefficients and are given the symbol γ. The COD-based coefficient, γ_i, for component A_i would be calculated from

$$\gamma_i = \frac{(a_i)(MW_i)(COD_i)}{(a_1)(MW_1)(COD_1)}, \tag{3.4}$$

$$\gamma_i = \frac{\Psi_i (COD_i)}{(COD_1)}, \tag{3.5}$$

where COD_i and COD_1 are the COD per unit mass of component A_i and the reference component, respectively. They can be obtained by writing a balanced equation for the oxidation of the compound or component to carbon dioxide and water. Table 3.1 contains COD mass equivalents of several constituents that commonly change oxidation state in biochemical operations. Note that under oxidizing conditions, carbon dioxide has a COD of zero, since the carbon in it is already in the most oxidized state (+IV). Likewise, for bicarbonate and carbonate. Therefore, these oxidized forms of carbon do not appear in COD-based stoichiometric equations unless they serve as an electron acceptor, as would occur under methanogenic conditions. Furthermore, oxygen is equivalent to negative COD since COD is oxygen demand (i.e., it represents loss of oxygen). Finally, it should be noted that any reactant or product containing only elements that do not change oxidation state during biochemical oxidation/reduction reactions will have a unit COD of zero, causing them to drop out of the COD-based stoichiometric equation.

Example 3.1.1.1

Consider a typical molar-based stoichiometric equation for bacterial growth on carbohydrate (CH_2O) with ammonia as the nitrogen source:

$$CH_2O + 0.290\ O_2 + 0.142\ NH_4^+ + 0.142\ HCO_3^- \rightarrow$$
$$0.142\ C_5H_7O_2N + 0.432\ CO_2 + 0.858\ H_2O, \tag{3.6}$$

TABLE 3.1
COD Mass Equivalents of Some Common Constituents

Constituent[a]	Change of Oxidation State	COD Equivalent[b]
Biomass, $C_5H_7O_2N$	C to +IV	1.42 g COD/g $C_5H_7O_2N$, 1.42 g COD/g VSS, 1.20 g COD/g TSS
Oxygen (as e^- acceptor)	O (0) to O (–II)	–1.00 g COD/g O_2[c]
Nitrate (as e^- acceptor)	N (+V) to N (0)	–0.646 g COD/g NO_3^-, –2.86 g COD/g N
Nitrate (as N source)	N (+V) to N (–III)	–1.03 g COD/g NO_3^-, –4.57 g COD/g N
Sulfate (as e^- acceptor)	S (+VI) to S (–II)	–0.667 g COD/g $SO_4^=$, –2.00 g COD/g S
Carbon dioxide (as e^- acceptor)	C (+IV) to C (–IV)	–1.45 g COD/g CO_2, –5.33 g COD/g C
CO_2, HCO_3^-, $H_2CO_3^=$	No change in an oxidizing environment	0.00
Organic matter in domestic wastewater, $C_{10}H_{19}O_3N$	C to +IV	1.99 g COD/g organic matter
Protein, $C_{16}H_{24}O_5N_4$	C to +IV	1.50 g COD/g protein
Carbohydrate, CH_2O	C to +IV	1.07 g COD/g carbohydrate
Grease, $C_8H_{16}O$	C to +IV	2.88 g COD/g grease
Acetate, CH_3COO^-	C to +IV	1.08 g COD/g acetate
Propionate, $C_2H_5COO^-$	C to +IV	1.53 g COD/g propionate
Benzoate, $C_6H_5COO^-$	C to +IV	1.98 g COD/g benzoate
Ethanol, C_2H_5OH	C to +IV	2.09 g COD/g ethanol
Lactate, $C_2H_4OHCOO^-$	C to +IV	1.08 g COD/g lactate
Pyruvate, CH_3COCOO^-	C to +IV	0.92 g COD/g pyruvate
Methanol, CH_3OH	C to +IV	1.50 g COD/g methanol
$NH_4^+ \rightarrow NO_3^-$	N (–III) to N (+V)	3.55 g COD/g NH_4^+, 4.57 g COD/g N
$NH_4^+ \rightarrow NO_2^-$	N (–III) to N (+III)	2.67 g COD/g NH_4^+, 3.43 g COD/g N
$NO_2^- \rightarrow NO_3^-$	N (+III) to N (+V)	0.36 g COD/g NO_2^-, 1.14 g COD/g N
$S \rightarrow SO_4^=$	S (0) to S (+VI)	1.50 g COD/g S
$H_2S \rightarrow SO_4^=$	S (–II) to S (+VI)	1.88 g COD/g H_2S, 2.00 g COD/g S
$S_2O_3^= \rightarrow SO_4^=$	S (+II) to S (+VI)	0.57 g COD/g $S_2O_3^=$, 1.00 g COD/g S
$SO_3^= \rightarrow SO_4^=$	S (+IV) to S (+VI)	0.20 g COD/g $SO_3^=$, 0.50 g COD/g S
H_2	H (0) to H (+I)	8.00 g COD/g H

[a] Listed in the same order as the reactants in Table 3.2.
[b] A negative sign implies that the constituent is receiving electrons.
[c] By definition, oxygen demand is negative oxygen.

where $C_5H_7O_2N$ is the empirical formula for cell mass. Note that the charges are balanced and that the number of moles of each element in the reactants equals the number in the products. The molar-based stoichiometric equation tells us that the biomass yield is 0.142 moles of biomass formed per mole of carbohydrate used and that 0.290 moles of oxygen are required per mole of carbohydrate used to synthesize that biomass.

Convert this equation to a mass-based stoichiometric equation. To do this, we need the molecular weight of each reactant and product. These are CH_2O, 30; O_2, 32; NH_4^+, 18; HCO_3^-, 61; $C_5H_7O_2N$, 113; CO_2, 44; and H_2O, 18. Using these with the stoichiometric coefficients from Equation 3.6 in Equation 3.2 gives:

$$CH_2O + 0.309\ O_2 + 0.085\ NH_4^+ + 0.289\ HCO_3^- \rightarrow$$
$$0.535\ C_5H_7O_2N + 0.633\ CO_2 + 0.515\ H_2O. \qquad (3.7)$$

In this case, the charges are no longer balanced, but the sum of the stoichiometric coefficients for the reactants equals the sum for the products. The mass-based stoichiometric equation tells us that the biomass yield is 0.535 grams of biomass formed per gram of carbohydrate used and that 0.309 grams of oxygen are required per gram of carbohydrate used to synthesize that biomass.

Now convert the molar-based equation to a COD-based equation. To do this, use must be made of the unit CODs given in Table 3.1. In this case, the unit COD of ammonia is taken as zero because the nitrogen in cell material primarily exists as amino acids or nucleic acids and, therefore, is in the same oxidation state as the nitrogen in ammonia (i.e., –III); thus, it does not undergo a change of oxidation state. Carrying out the conversion represented by Equation 3.4 yields:

$$CH_2O\ COD + (-0.29)\ O_2 \rightarrow 0.71\ C_5H_7O_2N\ COD. \tag{3.8}$$

Note that only three constituents remain because they are the only ones that can be represented by COD in this case. Also note that like the mass-based equation, the sum of the stoichiometric coefficients for the reactants equals the sum of the stoichiometric coefficients for the products. Finally, note that the stoichiometric coefficient for oxygen carries a negative sign even though it is a reactant. That is because it is being expressed as COD. Thus, the COD-based stoichiometric equation tells us that the biomass yield is 0.71 grams of biomass COD formed per gram of carbohydrate COD used and that 0.29 grams of oxygen are required per gram of carbohydrate COD used to synthesize that biomass.

3.1.2 Generalized Reaction Rate

Stoichiometric equations can also be used to establish the relative reaction rates for reactants or products. Because the sum of the stoichiometric coefficients in a mass-based stoichiometric equation equals zero, its general form may be rewritten in the following way:[76]

$$(-1)A_1 + (-\Psi_2)A_2 + \cdots + (-\Psi_k)A_k + \Psi_{k+1}A_{k+1} + \cdots + \Psi_m A_m = 0, \tag{3.9}$$

where components 1 through k are reactants, components k + 1 through m are products, and reactant A_1 is the basis for the normalized stoichiometric coefficients. Note that the normalized stoichiometric coefficients are given negative signs for reactants and positive signs for products. Since there is a relationship between the masses of the different reactants used or products formed, it follows that there is also a relationship between the rates at which they are used or formed. If we let r_i represent the rate of formation of component i (where $i = 1 \rightarrow k$), it follows that:

$$\frac{r_1}{(-1)} = \frac{r_2}{(-\Psi_2)} = \frac{r_k}{(-\Psi_k)} = \frac{r_{k+1}}{(\Psi_{k+1})} = \frac{r_m}{(\Psi_m)} = r, \tag{3.10}$$

where r is called the generalized reaction rate. As above, the sign on Ψ_i signifies whether the component is being removed or formed. Consequently, if the stoichiometry of a reaction has been determined in mass units and the reaction rate has been determined for one component, then the reaction rates in mass units are known for all other components.

Equations 3.9 and 3.10 also hold true for COD-based stoichiometric equations. The normalized stoichiometric coefficients (Ψ_i) are simply replaced with appropriate COD-based coefficients (γ_i).

Example 3.1.2.1

Biomass is growing in a bioreactor at a rate of 1.0 g/(L·h) and the growth conforms to the stoichiometry expressed by Equation 3.7. At what rate are carbohydrate and oxygen being used in the bioreactor to support that growth?

Rewriting Equation 3.7 in the form of Equation 3.9 gives:

$$-CH_2O - 0.309\ O_2 - 0.085\ NH_4^+ - 0.289\ HCO_3^-$$
$$+ 0.535\ C_5H_7O_2N + 0.633\ CO_2 + 0.515\ H_2O = 0.$$

Use of Equation 3.10 allows determination of the generalized reaction rate:

$$r = \frac{r_{C_5H_7O_2N}}{0.535} = \frac{1.0}{0.535} = 1.87\ g\ CH_2O/(L \cdot h).$$

Note that the generalized reaction rate is expressed in terms of the constituent that serves as the basis for normalization of the stoichiometric equation (typically, the electron donor). The rates of carbohydrate and oxygen utilization can now also be determined from Equation 3.10:

$$r_{CH_2O} = (-1.0)(1.87) = -1.87\ g\ CH_2O/(L \cdot h)$$
$$r_{O_2} = (-0.309)(1.87) = -0.58\ g\ O_2/(L \cdot h).$$

Note: To facilitate learning, the reader is encouraged to prove that the units associated with each worked example produce the units for the answer given.

3.1.3 Multiple Reactions: The Matrix Approach

In Chapter 2 we learned that there are many important events occurring in biochemical operations. Consequently, multiple reactions will take place simultaneously, and all must be considered when mass balance equations are written for biochemical operations. Extension of the concepts above to multiple reactions simplifies the presentation of those mass balances and allows the fates of all reactants to be easily visualized.[65,76]

Consider a situation in which i components (where $i = 1 \rightarrow m$) participate in j reactions (where $j = 1 \rightarrow n$), in which case $\Psi_{i,j}$ represents the normalized mass-based stoichiometric coefficient for component i in reaction j. This situation gives a group of mass-based stoichiometric equations:

$$\begin{aligned}
(-1)A_1 + \cdots + (-\Psi_{k,1})A_k + (+\Psi_{k+1,1})A_{k+1} + \cdots + (+\Psi_{m,1})A_m &= 0\ r_1 \\
(-\Psi_{1,2})A_1 + \cdots + \quad (-1)A_k + (+\Psi_{k+1,2})A_{k+1} + \cdots + (+\Psi_{m,2})A_m &= 0\ r_2 \\
\cdot \qquad\qquad \cdot \qquad\qquad \cdot \qquad\qquad \cdot & \\
\cdot \qquad\qquad \cdot \qquad\qquad \cdot \qquad\qquad \cdot & \\
\cdot \qquad\qquad \cdot \qquad\qquad \cdot \qquad\qquad \cdot & \\
(-\Psi_{1,n})A_1 + \cdots + (+\Psi_{k,n})A_k + (+\Psi_{k+1,n})A_{k+1} + \cdots + \quad (-1)A_m &= 0\ r_n.
\end{aligned} \tag{3.11}$$

Note that A_1 does not necessarily represent the component chosen as the basis for the normalized stoichiometric coefficients. Rather, a different component may be selected for each reaction so that each resulting normalized stoichiometric coefficient has appropriate physical meaning. Nevertheless, because the equations are mass based, the sum of the normalized stoichiometric coefficients in each equation must equal zero, as indicated in Equation 3.11. This allows a continuity check to be made for each reaction. Furthermore, also note that any component A_i may be a reactant in one reaction

and a product in another. This means that the overall rate of formation of that component will be the net rate obtained by considering the sum of the rates for all reactions in which it participates:

$$r_i = \sum_{j=1}^{n} \Psi_{i,j} \cdot r_j. \tag{3.12}$$

If the net rate of formation is negative, the component is being consumed and if it is positive the component is being produced. The same approach can be used for COD-based stoichiometric equations by replacing $\Psi_{i,j}$ with $\gamma_{i,j}$. This approach will be applied in Part II when models are developed for biochemical reactors, and will be particularly useful when complex systems with several components and reactions are considered.

3.2 BIOMASS GROWTH AND SUBSTRATE UTILIZATION

3.2.1 Generalized Equation for Biomass Growth

It will be recalled from Section 2.4.1 that biomass growth and substrate utilization are coupled. Furthermore, we saw in Section 2.4.2 that environmental engineers account for maintenance energy needs through the decay reaction. This means that as long as the production of soluble microbial products is negligible, the only use of substrate is for biomass growth. Consequently, when a stoichiometric equation for biomass growth is written with the substrate as the basis, the stoichiometric coefficient for the biomass term will be the biomass true growth yield. With this in mind, the generalized equation for microbial growth can be written as

$$\begin{aligned}\text{Carbon source} + \text{energy source} + \text{electron acceptor} + \text{nutrients} \rightarrow \\ \text{biomass} + CO_2 + \text{reduced acceptor} + \text{end products.}\end{aligned} \tag{3.13}$$

For modeling purposes, it would be desirable to be able to write a quantitative equation in the same form for any situation, no matter what the carbon source, energy source, or electron acceptor. Using the concept of half reactions, McCarty[95,122] has devised a technique whereby this may be done.

3.2.1.1 Half-Reaction Approach

In the absence of significant soluble microbial product formation, all nonphotosynthetic microbial growth reactions consist of two components, one for synthesis and one for energy. The carbon in the synthesis component ends up in biomass, whereas any carbon associated with the energy component becomes carbon dioxide. Such reactions are also oxidation-reduction reactions and thus involve the transfer of electrons from a donor to an acceptor. For heterotrophic growth the electron donor is an organic substrate, whereas for autotrophic growth the electron donor is inorganic. To allow consideration of all of these factors, McCarty[95,122] has written three types of half reactions: one for cell material (R_c), one for the electron donor (R_d), and one for the electron acceptor (R_a). These are presented in Table 3.2 for a variety of substances. Reactions 1 and 2 represent R_c for the formation of biomass. Both are based on the empirical formula $C_5H_7O_2N$, but one uses ammonia nitrogen as the nitrogen source whereas the other uses nitrate. Reactions 3 through 6 are half-reactions R_a for the electron acceptors oxygen, nitrate, sulfate, and carbon dioxide, respectively. Reactions 7 through 17 are half-reactions R_d for organic electron donors. The first of these represents the general composition of domestic wastewater, while the next three are for wastes composed primarily of proteins, carbohydrates, and lipids, respectively. Reactions 11 through 17 are for specific organic compounds of interest in some biochemical operations. The last nine reactions represent possible autotrophic electron donors. Reactions 19 through 21 are for nitrification. To facilitate their combination, the half reactions all are written on an electron equivalent basis, with the electrons on the right side.

TABLE 3.2
Oxidation Half Reactions

Reaction Number	Half-Reactions
	Reactions for Bacterial Cell Synthesis (R_c)
	Ammonia as nitrogen source:
1.	$\frac{1}{20}C_5H_7O_2N + \frac{9}{20}H_2O = \frac{1}{5}CO_2 + \frac{1}{20}HCO_3^- + \frac{1}{20}NH_4^+ + H^+ + e^-$
	Nitrate as nitrogen source:
2.	$\frac{1}{28}C_5H_7O_2N + \frac{11}{28}H_2O = \frac{1}{28}NO_3^- + \frac{5}{28}CO_2 + \frac{29}{28}H^+ + e^-$
	Reactions for Electron Acceptors (R_a)
	Oxygen:
3.	$\frac{1}{2}H_2O = \frac{1}{4}O_2 + H^+ + e^-$
	Nitrate:
4.	$\frac{1}{10}N_2 + \frac{3}{5}H_2O = \frac{1}{5}NO_3^- + \frac{6}{5}H^+ + e^-$
	Sulfate:
5.	$\frac{1}{16}H_2S + \frac{1}{16}HS^- + \frac{1}{2}H_2O = \frac{1}{8}SO_4^= + \frac{19}{16}H^+ + e^-$
	Carbon dioxide (methanogenesis):
6.	$\frac{1}{8}CH_4 + \frac{1}{4}H_2O = \frac{1}{8}CO_2 + H^+ + e^-$
	Reactions for Electron Donors (R_d)
	Organic Donors (Heterotrophic Reactions):
	Domestic wastewater:
7.	$\frac{1}{50}C_{10}H_{19}O_3N + \frac{9}{25}H_2O = \frac{9}{50}CO_2 + \frac{1}{50}NH_4^+ + \frac{1}{50}HCO_3^- + H^+ + e^-$
	Protein (amino acids, proteins, nitrogenous organics):
8.	$\frac{1}{66}C_{16}H_{24}O_5N_4 + \frac{27}{66}H_2O = \frac{8}{33}CO_2 + \frac{2}{33}NH_4^+ + \frac{31}{33}H^+ + e^-$
	Carbohydrate (cellulose, starch, sugars):
9.	$\frac{1}{4}CH_2O + \frac{1}{4}H_2O = \frac{1}{4}CO_2 + H^+ + e^-$
	Grease (fats and oils):
10.	$\frac{1}{46}C_8H_{16}O + \frac{15}{46}H_2O = \frac{4}{23}CO_2 + H^+ + e^-$
	Acetate:
11.	$\frac{1}{8}CH_3COO^- + \frac{3}{8}H_2O = \frac{1}{8}CO_2 + \frac{1}{8}HCO_3^- + H^+ + e^-$

(*Continued*)

TABLE 3.2 (CONTINUED)
Oxidation Half Reactions

Reaction Number	Half-Reactions
Propionate:	
14. $\frac{1}{14}CH_3CH_2COO^- + \frac{5}{14}H_2O$	$= \frac{1}{7}CO_2 + \frac{1}{14}HCO_3^- + H^+ + e$
Benzoate:	
13. $\frac{1}{30}C_6H_5COO^- + \frac{13}{30}H_2O$	$= \frac{1}{5}CO_2 + \frac{1}{30}HCO_3^- + H^+ + e^-$
Ethanol:	
14. $\frac{1}{12}CH_3CH_2OH + \frac{1}{4}H_2O$	$= \frac{1}{6}CO_2 + H^+ + e^-$
Lactate:	
15. $\frac{1}{12}CH_3CHOHCOO^- + \frac{1}{3}H_2O$	$= \frac{1}{6}CO_2 + \frac{1}{12}HCO_3^- + H^+ + e^-$
Pyruvate:	
16. $\frac{1}{10}CH_3COCOO^- + \frac{2}{5}H_2O$	$= \frac{1}{5}CO_2 + \frac{1}{10}HCO_3^- + H^+ + e^-$
Methanol:	
17. $\frac{1}{6}CH_3OH + \frac{1}{6}H_2O$	$= \frac{1}{6}CO_2 + H^+ + e^-$
Inorganic Donors (Autotrophic Reactions):	
18. Fe^{++}	$= Fe^{+3} + e^-$
19. $\frac{1}{8}NH_4^+ + \frac{3}{8}H_2O$	$= \frac{1}{8}NO_3^- + \frac{5}{4}H^+ + e^-$
20. $\frac{1}{6}NH_4^+ + \frac{1}{3}H_2O$	$= \frac{1}{6}NO_2^- + \frac{4}{3}H^+ + e^-$
21. $\frac{1}{2}NO_2^- + \frac{1}{2}H_2O$	$= \frac{1}{2}NO_3^- + H^+ + e^-$
22. $\frac{1}{6}S + \frac{2}{3}H_2O$	$= \frac{1}{6}SO_4^= + \frac{4}{3}H^+ + e^-$
23. $\frac{1}{16}H_2S + \frac{1}{16}HS^- + \frac{1}{2}H_2O$	$= \frac{1}{8}SO_4^= + \frac{19}{16}H^+ + e^-$
24. $\frac{1}{8}S_2O_3^= + \frac{5}{8}H_2O$	$= \frac{1}{4}SO_4^= + \frac{5}{4}H^+ + e^-$
25. $\frac{1}{2}SO_3^= + \frac{1}{2}H_2O$	$= \frac{1}{2}SO_4^= + H^+ + e^-$
26. $\frac{1}{2}H_2$	$= H^+ + e^-$

Note: Adapted from McCarty, P. L., Stoichiometry of biological reactions. *Progress in Water Technology*, 7 (1): 157–72, 1975.

The overall stoichiometric equation (R) is the sum of the half reactions:

$$R = R_d - f_e \cdot R_a - f_s \cdot R_c. \tag{3.14}$$

The minus terms mean that half-reactions R_a and R_c must be inverted before use. This is done by switching the left and right sides. The term f_e represents the fraction of the electron donor that is coupled with the electron acceptor (i.e., the portion used for energy, hence the subscript e) and f_s represents the fraction captured through synthesis. As such they quantify the endpoint of the reaction. Furthermore, in order for Equation 3.14 to balance:

$$f_e + f_s = 1.0. \tag{3.15}$$

This equation is equivalent to stating that all electrons originally in the electron donor end up either in the biomass synthesized (f_s) or in the electron acceptor (f_e). This is an important fundamental concept that we will return to later.

3.2.1.2 Empirical Formulas for Use in Stoichiometric Equations

As can be seen by examining Table 3.2, it was necessary to assume empirical formulas for biomass and alternative organic electron donors in order to write the half reactions.

Various empirical formulas have been proposed to represent the organic composition of microbial cells. One of the oldest and most widely accepted in the field of wastewater treatment is the one introduced in Section 2.4.1 and used in Example 3.1.1.1, $C_5H_7O_2N$.[75] Other formulas consisting of the same elements have been used, but they all result in about the same COD per unit of biomass.[93] Another formula has been proposed that includes phosphorus, $C_{60}H_{87}O_{23}N_{12}P$.[94] While awareness of the need for phosphorus by biomass is essential, it is not necessary to include phosphorus in the empirical formula because the mass required is generally about one-fifth of the mass of nitrogen required. This allows the phosphorus requirement to be calculated even when the simpler empirical formula is used.

All empirical formulas for biomass seek to represent in a simple way material composed of a highly complex and integrated mixture of organic molecules. Furthermore, because the relative quantities of those molecules change as the growth conditions of the culture change,[68] it would be purely fortuitous if a single chemical formula for biomass applied to all cases. An estimate of the constancy of the overall elemental composition can be obtained by measuring the COD and heat of combustion of biomass grown under various conditions, because constancy of those parameters would imply that the ratios of the elements C, H, O, and N were relatively constant. Investigations of that sort have indicated that the elemental composition is indeed a function of the growth conditions.[53] Thus, while an empirical formula can be written for biomass, its applicability to all situations is doubtful and one should view with caution equations said to depict "the biochemical reaction" exactly. Nevertheless, the concepts stated in Equation 3.13 are still valid and many important relationships can be demonstrated through its use. Consequently, for illustrative purposes, the formula $C_5H_7O_2N$ will be used to represent biomass throughout this book. As discussed in Section 2.4.1, it has a COD of 1.42 mg COD/mg VSS (volatile suspended solids), or 1.20 mg COD/mg TSS (total suspended solids).

In a laboratory or research situation, the exact composition of the electron donor is usually known. For example, if glucose were the energy source, its empirical formula $C_6H_{12}O_6$ would be used in the stoichiometric equation. Furthermore, if a synthetic medium contained several organic electron donors, the half reaction for each could be written separately and then they could be combined to get R_d for the mixture by multiplying each half reaction by the fractional contribution (on an electron equivalent basis) of its electron donor in the medium and adding them together.

An actual wastewater presents a more difficult situation because the chemical composition of the electron donor is seldom known. One approach would be to analyze the waste for its carbon, hydrogen, oxygen, and nitrogen contents and construct an empirical formula from the results. A half reaction could then be written for that particular formula.[95,122] For example, as shown in Table 3.2, the empirical formula for the organic matter in domestic wastewater has been estimated to be $C_{10}H_{19}O_3N$. Alternatively, if the COD, organic carbon, organic nitrogen, and volatile solids content of a wastewater are known, they can be used to generate the half reaction.[95,122] Finally, if a wastewater contains predominately carbohydrate, protein, and lipid, knowledge of their relative concentrations can be used to write the equation for microbial growth because each can be represented by a generalized empirical formula: that is, CH_2O, $C_{16}H_{24}O_5N_4$, and $C_8H_{16}O$, respectively. As with other mixtures, the half reaction for each is multiplied by the fraction of the component in the wastewater and the three are added to get R_d.

The nature of the electron acceptor depends on the environment in which the biomass is growing. If the environment is aerobic, the acceptor will be oxygen. If it is anaerobic, the acceptor will depend on the particular reaction taking place. For example, if lactic acid fermentation is occurring, pyruvic acid is the acceptor, whereas carbon dioxide is the acceptor for methanogenesis. Finally, nitrate can serve as the electron acceptor under anoxic conditions. Half reactions have been written for all of these, as shown in Table 3.2.

3.2.1.3 Determination of f_s

Once the electron donor and the electron acceptor have been identified, either f_e or f_s must be determined before the balanced stoichiometric equation can be written. Generally, f_s is easier to estimate because it can be related to the true growth yield expressed on a COD basis. If f_e is the fraction of the electron donor transferred to the electron acceptor to provide the energy with which to synthesize new biomass, conservation of energy and Equation 3.15 tell us that the remainder of the electrons originally available in the donor must end up in the new biomass formed. If we accept $C_5H_7O_2N$ as being representative of biomass, we can see that carbon and nitrogen are the reduced elements that will house those electrons. Nitrogen in biomass is in the –III state (i.e., as amino nitrogen). If the nitrogen available for biomass synthesis is also in the –III state, as in ammonia, no electrons will be required to reduce it, and the electrons captured through synthesis will all be associated with the carbon. Consequently, the energy available in the carbon of the biomass is equal to the energy incorporated during synthesis, or f_s when expressed as a fraction of the electron donor. Thus, if we could measure the energy or electrons available in the biomass produced, we would have a measure of f_s.

In Section 2.4.1 the yield was defined as the amount of biomass formed per unit of substrate used. However, it was also pointed out that when the electron donor is an organic compound, it is often convenient to express the yield as mass of biomass COD formed per mass of substrate COD destroyed. The COD test is a measure of electrons available from carbon. Since COD is oxygen demand and oxygen has an equivalent weight of eight, there are eight grams of COD per electron equivalent, as can be seen by examining half-reaction 3 in Table 3.2. This allows interconversion of COD and electron equivalents. Consequently, the yield is also the number of electrons available from carbon in the new biomass per unit of electrons removed from the substrate, or the fraction of the electron donor captured through synthesis (i.e., f_s). Thus, when ammonia nitrogen serves as the nitrogen source for heterotrophic biomass synthesis:

$$f_s = Y_H \quad (NH_4^+ \text{ as nitrogen source, organic electron donor}), \tag{3.16}$$

where Y_H is expressed on a COD basis and the subscript H indicates that the true growth yield is for heterotrophic biomass growth. The utility of Equation 3.16 comes from the fact that the true growth yield, Y_H, can either be determined directly in COD units from data collected with full-, pilot-, or lab-scale bioreactors, or it can be determined in VSS or TSS units and converted to COD units using appropriate conversion factors. In either case, once Y_H is known in COD units, f_s for the

system under study can be determined from Equation 3.16. The techniques for determining Y_H will be discussed in Chapter 9.

As long as ammonia or amino nitrogen is available to the microorganisms, they will use it preferentially for biomass synthesis. If it is not available, the microorganisms will use nitrate-N. (If no nitrogen is available, cell synthesis cannot occur because an essential reactant is missing.) When nitrate is the nitrogen source, the nitrogen must be reduced from the +V state to the –III state before it can be assimilated. This requires some of the electrons available in the substrate and they are part of the energy required for synthesis (i.e., part of f_s). However, the electrons required to reduce the nitrogen are not measured in the COD test because that test does not oxidize nitrogen, but leaves it in the –III state. Thus, in this case, the true growth yield expressed on a COD basis is not an accurate estimate of f_s. Rather, Y_H will be smaller than f_s. This artifact can be corrected for, however, because we know the number of electrons required to reduce nitrate-N to the appropriate oxidation state. Assuming an empirical formula for biomass of $C_5H_7O_2N$, it can be shown that:

$$f_s = 1.40\ Y_H \quad (NO_3^- \text{ as nitrogen source, organic electron donor}). \tag{3.17}$$

Thermodynamics suggests that the true growth yield obtained for growth with nitrate as the nitrogen source will be smaller than the true growth yield obtained when ammonia is available.[64] For example, for carbohydrate as the electron and carbon donor, the value of Y_H would be about 20% smaller with nitrate as the nitrogen source.

There are often circumstances in which one needs to establish the stoichiometry of biomass growth and substrate utilization before experimentally determined values of Y_H are available. Thus, it would be advantageous to have a theoretical basis for estimating f_s or Y_H. This has led a number of workers to seek a thermodynamic approach for predicting yield values.[63,93] However, as discussed in Section 2.4.1, this is a difficult task because of the large number of factors that influence the yield. VanBriesen[144] reviewed several methods for theoretically estimating heterotrophic bacterial yields on a variety of organic electron donors and found them to be comparable in that they predicted yields within 15% of one another. Special assumptions must be made in the case of autotrophic metabolism, as explained by Heijnen et al.,[63,64] who developed the Gibbs energy dissipation method, which was among the methods reviewed.[144] Because one should fully understand these theoretical techniques before using them and because the presentation required to establish that understanding is beyond the scope of this book, readers are referred to the original works if they desire to use such an approach.

3.2.2 Aerobic Growth of Heterotrophs with Ammonia as the Nitrogen Source

The best way to illustrate the use of half reactions is by an example. We will develop the molar stoichiometric equation for aerobic growth of heterotrophs that was the starting point for Example 3.1.1.1.

Example 3.2.2.1

Write the stoichiometric equation for aerobic heterotrophic microbial growth on a carbohydrate using ammonia as the nitrogen source, under conditions such that the true growth yield (Y_H) is 0.71 mg of biomass COD formed per mg of carbohydrate COD removed.

To do this we must make use of Equations 3.14 through 3.16:

$$R = R_d - f_e \cdot R_a - f_s \cdot R_c$$

$$f_s = Y_H = 0.71$$

$$f_e = 1.00 - 0.71 = 0.29.$$

Therefore:

$$R = R_d - 0.29R_a - 0.71R_c.$$

The electron donor is carbohydrate and the acceptor is oxygen. Thus, from Table 3.2:

$$R_d = \frac{1}{4}CH_2O + \frac{1}{4}H_2O = \frac{1}{4}CO_2 + H^+ + e^-$$

$$R_a = \frac{1}{2}H_2O = \frac{1}{4}O_2 + H^+ + e^-.$$

Since ammonia is the nitrogen source, R_c is

$$R_c = \frac{1}{20}C_5H_7O_2N + \frac{9}{20}H_2O = \frac{1}{5}CO_2 + \frac{1}{20}HCO_3^- + \frac{1}{20}NH_4^+ + H^+ + e^-.$$

Applying Equation 3.14 gives:

$$\begin{aligned}
R_d &= 0.25\ CH_2O + 0.25\ H_2O = 0.25\ CO_2 + H^+ + e^- \\
-0.29\,R_a &= 0.0725\ O_2 + 0.29\,H^+ + 0.29e^- = 0.145\,H_2O \\
-0.71R_c &= 0.142\,CO_2 + 0.0355\,HCO_3^- + 0.0355\,NH_4^+ + 0.71H^+ + 0.71\,e^- \\
&= 0.0355\ C_5H_7O_2N + 0.3195\ H_2O \\
R &= 0.25\,CH_2O + 0.0725\ O_2 + 0.0355\,NH_4^+ + 0.0355\,HCO_3^- \\
&= 0.0355\ C_5H_7O_2N + 0.108\,CO_2 + 0.2145\ H_2O.
\end{aligned}$$

This can be normalized to one mole of carbohydrate by dividing through by 0.25, giving Equation 3.6, which was the starting point of Example 3.1.1.1:

$$\begin{aligned}
&CH_2O + 0.29\ O_2 + 0.142\ NH_4^+ + 0.142\,HCO_3^- \rightarrow \\
&0.142\ C_5H_7O_2N + 0.432\ CO_2 + 0.858\ H_2O.
\end{aligned} \tag{3.6}$$

Equation 3.6 was converted to a COD-based stoichiometric equation in Example 3.1.1.1. If we rearrange Equation 3.8 in the same form as Equation 3.9, the result is

$$0.29\ O_2 + 0.71\ C_5H_7O_2N\ COD = CH_2O\ COD. \tag{3.18}$$

We return to this equation to make three important points. First, note that the value of Y_H in Equation 3.18 is 0.71 mg biomass COD formed/mg substrate COD used. This is the same as the Y_H value used to develop Equation 3.6, as we would expect. Second, note that Equation 3.18 expresses the same information as Equation 3.15. In other words, since all of the electrons removed from the substrate must end up in either the electron acceptor or the biomass formed, we can state that the substrate COD removed must equal the biomass COD formed plus the oxygen used. Finally, since Equation 3.18 expresses the same information as Equation 3.15, we can see that the COD-based stoichiometric coefficient on oxygen is the same as f_e. The balance portrayed by Equations 3.15 and 3.18 is a very important one that we will make extensive use of throughout this book.

3.2.3 Aerobic Growth of Heterotrophs with Nitrate as the Nitrogen Source

As previously discussed, consideration must be given to the form of nitrogen available for cell synthesis when writing the stoichiometric equation for cell growth. Ammonia will be used preferentially,

and thus the half-reaction 1 in Table 3.2 should be used when ammonia is available, even if nitrate is serving as the terminal electron acceptor. Only when nitrate is present as the sole nitrogen source should the half-reaction 2 be used. In that case, when expressing the stoichiometric equation on a COD basis, it must be recognized that nitrogen changes the oxidation state from +V to –III. As an example, consider the case of the aerobic growth of heterotrophs on carbohydrate with nitrate as the nitrogen source. In this case, the true growth yield is 0.45 mg biomass COD/mg carbohydrate COD removed, reflecting the energy that must be used to reduce the nitrogen. Applying Equation 3.17 reveals that f_s is 0.63, giving the following molar stoichiometric equation:

$$CH_2O + 0.370\ O_2 + 0.090\ NO_3^- + 0.090\ H^+ \rightarrow$$
$$0.090\ C_5H_7O_2N + 0.550\ CO_2 + 0.730\ H_2O. \quad (3.19)$$

After conversion to a mass basis by the application of Equation 3.2 this becomes:

$$CH_2O + 0.395\ O_2 + 0.186\ NO_3^- + 0.003\ H^+ \rightarrow$$
$$0.339\ C_5H_7O_2N + 0.807\ CO_2 + 0.438\ H_2O. \quad (3.20)$$

Conversion of this equation to a mass of COD basis requires the application of Equation 3.5 using the unit CODs given in Table 3.1. Note that NO_3^- has a unit COD of –1.03 mg COD/mg NO_3^-. This is equivalent to saying that each mg of nitrate that is reduced to amino nitrogen in biomass accepts as many electrons as 1.03 mg of oxygen. The application of Equation 3.4 through Equation 3.20 gives:

$$CH_2O\ COD + (-0.370)\ O_2 + (-0.180)\ O_2 \text{ equivalents of } NO_3^- \rightarrow$$
$$0.450\ C_5H_7O_2N\ COD. \quad (3.21)$$

Equation 3.21 shows clearly that the COD (electron) balance would not be correct if the change in oxidation state of the nitrogen was not considered. Failure to recognize this can lead to problems when COD balances are performed on operating bioreactors.

It is often convenient to express the COD equivalence of nitrate as a nitrogen source on the basis of the nitrogen utilized for biomass synthesis, rather than on the basis of nitrate. In that case the conversion factor is –4.57 mg COD/mg N (or 4.57 mg O_2/mg N), as indicated in Table 3.1.

3.2.4 Growth of Heterotrophs with Nitrate as the Terminal Electron Acceptor and Ammonia as the Nitrogen Source

If nitrate were serving as the terminal electron acceptor under anoxic conditions, the amount needed could be calculated from the stoichiometric equation obtained when half-reaction 4 was used in place of half-reaction 3 as R_a in Equation 3.14. Exactly the same procedures would be followed for obtaining the molar- and mass-based stoichiometric equations. Consider the case when ammonia serves as the nitrogen source for cell synthesis. Because biomass yield coefficients are about 20% smaller for biomass growing under anoxic conditions relative to aerobic conditions,[25,63,64,97] we will assume a true growth yield of 0.57 mg biomass COD/mg substrate COD, which is 20% smaller than that used in Examples 3.1.1.1 and 3.2.2.1. Application of the appropriate techniques gives the molar-based stoichiometric equation:

$$CH_2O + 0.344\ NO_3^- + 0.115\ NH_4^+ + 0.114\ HCO_3^- + 0.344\ H^+ \rightarrow$$
$$0.114\ C_5H_7O_2N + 0.544\ CO_2 + 0.172\ N_2 + 1.058\ H_2O. \quad (3.22)$$

Converting this to a mass of carbohydrate basis by application of Equation 3.2 gives:

$$CH_2O + 0.711\,NO_3^- + 0.069\,NH_4^+ + 0.232\,HCO_3^- + 0.011\,H^+ \rightarrow$$
$$0.429\,C_5H_7O_2N + 0.798\,CO_2 + 0.161\,N_2 + 0.635\,H_2O. \quad (3.23)$$

Because the true growth yield was assumed to be 20% less than that used in Example 3.1.1.1, the quantities of biomass formed in Equations 3.22 and 3.23 are 20% less than those in Equations 3.6 and 3.7, respectively.

The conversion of Equation 3.23 to a COD basis requires inclusion of a conversion factor for the oxygen equivalence of nitrate nitrogen when it is being reduced to nitrogen gas, N_2, which is the case when nitrate serves as the terminal electron acceptor. An examination of Table 3.1 reveals that the unit COD for the reduction of NO_3^- to N_2 is −0.646 mg COD/mg NO_3^-. The sign is negative because the nitrate is accepting electrons. The source of this value may be seen from the half reactions in Table 3.2, which reveal that 1/5 mole of nitrate is equivalent to 1/4 mole of oxygen. Conversion to a mass basis reveals that each gram of nitrate that is reduced to N_2 can accept as many electrons as 0.646 grams of oxygen. Applying Equation 3.4 with the appropriate conversion factors to Equation 3.23 gives:

$$CH_2O\ COD + (-0.43)\,O_2\ \text{equivalents of}\ NO_3^- \rightarrow 0.57\,C_5H_7O_2N\ COD. \quad (3.24)$$

Comparison of Equation 3.24 to Equation 3.8 reveals that 20% fewer electrons ended up as biomass due to the lower yield associated with growth using nitrate-N as the electron acceptor.[25,63,64,97]

Often it is convenient to express the oxygen equivalence of nitrate as an electron acceptor on the basis of nitrogen rather than nitrate. In that case the conversion factor is −2.86 mg COD/mg N (or 2.86 mg O_2/mg N), as shown in Table 3.1.

It should be noted from the preceding that the COD conversion factor for nitrate as a nitrogen source is different from the COD conversion factor for nitrate as a terminal electron acceptor because the final oxidation state of nitrogen is different in the two cases. This becomes especially important when nitrate serves as both the nitrogen source and the terminal electron acceptor. The safest way to handle this situation is to keep the two uses of nitrate separate in writing the stoichiometric equation, and to apply the appropriate conversion factor for each when converting the equation to a COD basis.

3.2.5 Aerobic Growth of Autotrophs with Ammonia as the Electron Donor

Nitrifying bacteria are autotrophic microorganisms that obtain their energy from the oxidation of reduced nitrogen. As discussed previously, ammonia oxidizing bacteria (AOB) oxidize ammonia-N to nitrite-N and nitrite oxidizing bacteria (NOB) oxidize nitrite-N to nitrate-N. The molar stoichiometric equations for their growth can be obtained by the half-reaction technique discussed previously, which requires knowledge of f_s. For autotrophic biomass growth, yield is often expressed as the mass of biomass COD formed per mass of inorganic element oxidized;[20,66] for example, for AOB it would be mg of biomass COD formed per mg of ammonia-N oxidized. To convert this yield value to an electron equivalent basis for determining f_s it is necessary to know that AOB oxidize ammonia-N (–III) to nitrite-N (+III), for a six electron change. Thus, the equivalent weight for nitrogen in this case is 14/6 = 2.33 grams/equivalent, which means that:

$$f_s = 0.291\,Y_{AOB} \quad (NH_4^+ \text{ as nitrogen source and electron donor}). \quad (3.25)$$

For NOB, nitrite-N (+III) serves as the electron donor and is oxidized to nitrate-N (+V) for a two electron change. Ammonia-N, however, serves as the nitrogen source. Consequently:

$$f_s = 0.875\ Y_{NOB} \quad (NH_4^+ \text{ as nitrogen source, } NO_2^- \text{ as electron donor}), \tag{3.26}$$

for these organisms, where Y_{NOB} has units of mg biomass COD formed/mg nitrite-N oxidized. Often nitrifying bacteria are considered together as a group and nitrification is treated as a single reaction converting ammonia-N to nitrate-N. In that case, nitrogen undergoes an eight electron change so that:

$$f_s = 0.219\ Y_A \quad (NH_4^+ \text{ as nitrogen source and electron donor}), \tag{3.27}$$

where Y_A represents the true growth yield for autotrophic nitrifying biomass and has units of mg biomass COD formed/mg ammonia-N oxidized.

Application of the half-reaction technique using typical yield values and Equation 3.2 provides the mass-based stoichiometric equations for nitrification. For AOB, when NH_4^+ is the basis, the equation is

$$\begin{aligned} NH_4^+ + 2.457\ O_2 + 6.716\ HCO_3^- &\rightarrow 0.114\ C_5H_7O_2N \\ + 2.509\ NO_2^- + 1.036\ H_2O + 6.513\ H_2CO_3&. \end{aligned} \tag{3.28}$$

When NO_2^- is the basis, the equation for NOB is

$$\begin{aligned} NO_2^- + 0.001\ NH_4^+ + 0.014\ H_2CO_3 + 0.003\ HCO_3^- + 0.339\ O_2 &\rightarrow \\ 0.006\ C_5H_7O_2N + 0.003\ H_2O + 1.348\ NO_3^-&. \end{aligned} \tag{3.29}$$

Furthermore, combining the two reactions reveals that the overall stoichiometry is

$$\begin{aligned} NH_4^+ + 3.300\ O_2 + 6.708\ HCO_3^- &\rightarrow 0.129\ C_5H_7O_2N \\ + 3.373\ NO_3^- + 1.041\ H_2O + 6.463\ H_2CO_3&. \end{aligned} \tag{3.30}$$

From these it can be seen that a large amount of alkalinity (HCO_3^-) is used during the oxidation of ammonium ion to nitrate ion: 6.708 mg HCO_3^-/mg NH_4^- removed, which is equivalent to 8.62 mg HCO_3^-/mg NHO_4^- -N removed (the sum of ammonia-N consumed for use as an electron donor and as a nitrogen source). The vast majority of that alkalinity utilization is associated with neutralization of the hydrogen ions released during the oxidation of ammonia-N. Only a small part of the alkalinity is incorporated into the cell material. If the wastewater contains insufficient alkalinity and if pH control is not practiced, the pH will drop below the normal physiological range, retarding the activity of both the autotrophs and the heterotrophs, thereby hurting the system performance. The equations also tell us that considerable oxygen is required for nitrification: 3.30 mg O_2 is consumed per mg NH_4^+ removed (or 4.24 mg O_2/mg NH_4^+ -N removed). Most (98%) of the NH_4^+ + -N removed is oxidized as the electron donor, which is equivalent to 4.33 mg O_2/mg of NH_4^+ + -N actually oxidized to nitrate-N. Of that amount 3.22 mg O_2 is used by AOB and 1.11 by NOB. The oxygen requirement of the nitrifying bacteria can have a significant impact on the total amount of oxygen required by a biochemical operation. Finally, it can be seen that relatively little biomass is formed, reflecting the low yields associated with autotrophic growth. For every mg of NH_4^+ removed, only 0.129 mg

of biomass is formed, which is equivalent to 0.166 mg biomass/mg NH_4^+ -N removed. Most of that, 0.146 mg biomass/mg NH_4^+ -N removed, is due to the growth of AOB, and only 0.020 mg biomass/mg NH_4^+ -N removed is due to NOB. Overall, the growth of nitrifying bacteria has little impact on the quantity of biomass in a biochemical operation treating a wastewater with the characteristics of domestic wastewater, but has a large impact on the oxygen and alkalinity requirements.

3.2.6 Kinetics of Biomass Growth

Equation 3.8 was the COD-based stoichiometric equation for aerobic growth of heterotrophic biomass with ammonia as the nitrogen source. Recognizing that the stoichiometric coefficient on biomass is the same as the true growth yield, Y_H, and that both substrate (S_S) and active heterotrophic biomass ($X_{B,H}$) are measured in COD units, it may be rewritten in terms of the true growth yield as

$$(1)S_S + [-(1-Y_H)]S_O \rightarrow Y_H X_{B,H}, \tag{3.31}$$

where S_O is oxygen, which is expressed in COD units, and thus carries a negative sign as indicated in Table 3.1.* Putting this in the form of Equation 3.9, while retaining COD units, gives:

$$(-1)S_S + (-1)\,[-(1-Y_H)]S_O + Y_H X_{B,H} = 0. \tag{3.32}$$

This equation is based on substrate as the reference constituent. Alternatively, it could be rewritten with active heterotrophic biomass as the reference constituent and that is the convention used herein:

$$\left(-\frac{1}{Y_H}\right)S_S + (-1)\left[-\left(\frac{1-Y_H}{Y_H}\right)\right]S_O + X_{B,H} = 0. \tag{3.33}$$

The application of Equation 3.10 gives:

$$\frac{r_{SS}}{\left(-\frac{1}{Y_H}\right)} = \frac{r_{SO}}{(-1)\left[-\left(\frac{1-Y_H}{Y_H}\right)\right]} = \frac{r_{XB}}{1} = r, \tag{3.34}$$

where [r] = mg COD/(L · hr). Thus, once r_{XB} has been defined, the rates for soluble substrate (r_{SS}) and dissolved oxygen (r_{SO}) are also known.

Similar equations can be written for the growth of heterotrophs with nitrate as the terminal electron acceptor and for the aerobic growth of autotrophs. The derivation of such equations is left as an exercise for the reader.

Bacteria divide by binary fission. Consequently, the reaction rate for bacterial growth can be expressed as first order with respect to the active biomass concentration (X_B):

$$r_{XB} = \mu \cdot X_B, \tag{3.35}$$

where μ is the specific growth rate coefficient (hr^{-1}). It is referred to as a specific rate coefficient because it defines the rate of biomass growth in terms of the concentration of active biomass present;

* S represents soluble constituents and X represents particulate constituents, with the subscript denoting the particular constituent involved.

that is, the mass of biomass COD formed per unit time per unit of active biomass COD present. Equation 3.35 holds for any type of bacterial growth, regardless of the nature of the electron donor or acceptor, although much of the following is written in terms of heterotrophic biomass growth on an organic substrate. Consequently, subscripts are not used at this point to distinguish between heterotrophic and autotrophic biomass, although they will be used later when it is necessary to make that distinction. The substitution of Equation 3.35 into Equation 3.34 defines the rates of substrate removal and oxygen (electron acceptor) utilization associated with biomass growth. It is important to note that the equation for oxygen utilization is also true for other electron acceptors, such as nitrate, as long as the quantity is expressed in oxygen equivalents.

3.2.7 Effect of Substrate Concentration on μ

3.2.7.1 The Monod Equation

Originally, exponential growth of bacteria (i.e., growth in accordance with Equation 3.35) was considered to be possible only when all nutrients, including the substrate, were present in high concentration. In the early 1940s, however, it was found that bacteria grow exponentially even when one nutrient is present only in a limited amount.[102] Furthermore, the value of the specific growth rate coefficient, μ, was found to depend on the concentration of that limiting nutrient, which can be the carbon source, the electron donor, the electron acceptor, nitrogen, or any other factor needed by the organisms for growth. Since that time, the generality of this observation has been substantiated often, so that it can now be considered a basic concept of microbial kinetics.[42] Let us first consider the situation when only an organic substrate is growth limiting.

Figure 3.1 illustrates the relationship that is obtained when μ is measured as a function of a single limiting substrate concentration. A number of different types of experiments can be performed to develop such a relationship and they will be discussed in Chapter 9. The important thing to note at this time is that μ initially rises rapidly as the substrate concentration is increased, but then asymptotically approaches a maximum, which is called the maximum specific growth rate, $\hat{\mu}$.

The question of the best mathematical formula to express the relationship shown in Figure 3.1 has been the subject of much debate. No one yet knows enough about the mechanisms of biomass growth to propose a mechanistic equation that will characterize growth exactly. Instead,

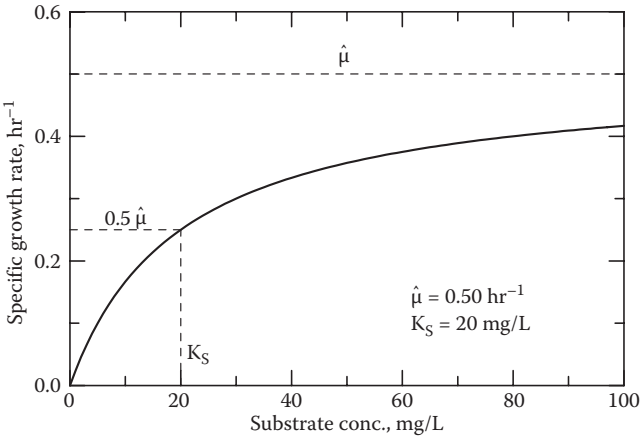

FIGURE 3.1 Typical plot of the relationship between the specific growth rate coefficient and the concentration of a noninhibitory substrate. The parameter values given were used to construct the curve with the Monod equation (Equation 3.36).

experimenters have observed the effects of various factors on growth and have then attempted to fit empirical equations to their observations. Consequently, all equations that have been proposed are curve-fits and the only valid arguments for use of one over another are goodness of fit, mathematical utility, and broad acceptance.

The equation with historical precedence and greatest acceptance is the one proposed by Monod[102]. Although his original work was done in batch reactors, it was later extended and refined by workers using continuous cultures of single bacterial species growing on defined media and it was concluded that the curve could be approximated adequately by the equation for a rectangular hyperbola.[42] Consequently, Monod proposed the equation:

$$\mu = \hat{\mu}\frac{S_S}{K_S + S_S}, \tag{3.36}$$

where K_S is the half-saturation coefficient. The K_S determines how rapidly μ approaches $\hat{\mu}$ and is defined as the substrate concentration at which μ is equal to half of $\hat{\mu}$, as shown in Figure 3.1. The smaller it is, the lower the substrate concentration at which μ approaches $\hat{\mu}$. Because of his pioneering efforts in defining the kinetics of microbial growth, Equation 3.36 is generally referred to as the Monod equation.

Because of the similarity of Equation 3.36 to the Michaelis-Menten equation in enzyme kinetics, many people have erroneously concluded that Monod proposed it on mechanistic grounds. While the Michaelis–Menten equation can be derived from consideration of the rates of chemical reactions catalyzed by enzymes, and thus has a mechanistic basis, the Monod equation is strictly empirical. In fact, Monod himself emphasized its empirical nature.[102]

The Monod equation has been found to fit the data for many pure cultures growing on single substrates, both organic and inorganic, and has been used extensively in the development of models describing the continuous cultivation of microorganisms. It has not been blindly accepted, however, and other workers have proposed alternative equations that fit their data better.[104,116,128] Nevertheless, it is still the most widely used equation.

Because the Monod equation was developed for pure cultures of bacteria growing on single organic substrates, two significant questions arise when its adoption is considered for modeling biochemical operations for wastewater treatment. The first concerns whether it can be used to express removal of a "substrate" that is really a mixture of hundreds of organic compounds measured by a nonspecific test like COD, since that is the nature of the organic matter in wastewater. Can the Monod equation adequately describe the effect of biodegradable COD on the specific growth rate of bacteria? The second question arises from consideration of the microbial communities present in wastewater treatment operations. As discussed in Chapter 2, those communities are highly complex, containing not only many bacterial species but higher life forms as well. Can the growth of such a heterogeneous assemblage be expressed simply as "biomass" by the Monod equation? Many researchers have investigated these questions and it is generally agreed that the answer to both is yes.[5,23,39,45,86] Nevertheless, it should be recognized that the manner in which the culture is grown will have a strong impact on its community structure, and that the values of $\hat{\mu}$ and K_S obtained from mixed culture systems are in reality average values resulting from many interacting species.[24,45,48] Consequently, it has been recommended that $\hat{\mu}$ and K_S be characterized by ranges, rather than by single values, just as was recommended for Y. It can be concluded that, however, the Monod equation is a reasonable model with which to describe the kinetics of microbial growth on complex organic substrates in wastewater treatment systems and, consequently, it is widely used. There are situations, however, in which it would be desirable to model the effects on microbial growth rates of individual organic compounds in complex mixtures. This situation is very complicated,[88] however, and consideration of it will be delayed until Chapter 22.

3.2.7.2 Simplifications of the Monod Equation

Examination of Equation 3.36 reveals that two simplifications can be made, and this is often done in the modeling of wastewater treatment systems. First, it can be seen that if S_S is much larger than K_S, the equation may be approximated as

$$\mu \approx \hat{\mu}. \tag{3.37}$$

This is called the zero-order approximation because under that condition the specific growth rate coefficient is independent of the substrate concentration (i.e., it is zero order with respect to S_S) and equal to the maximum specific growth rate coefficient. In other words, the bacteria will be growing as rapidly as possible. Second, if S_S is much smaller than K_S, the term in the denominator may be approximated as K_S and the equation becomes:

$$\mu \approx \frac{\hat{\mu}}{K_S} S_S. \tag{3.38}$$

This is called the first-order approximation because μ is first order with respect to S_S. Although Equation 3.38 is often easier to use than the Monod equation, care should be exercised in its use because serious error can result if S_S is not small relative to K_S. When COD is used as a measure of the total quantity of biodegradable organic matter, K_S can be relatively large, with the result that S_S in activated sludge reactors is often less than K_S. Consequently, Equation 3.38 is sometimes used to model such systems.

Garrett and Sawyer[44] were the first to propose the use of Equations 3.37 and 3.38 because they had observed that the specific growth rate coefficient for bacteria was directly proportional to the substrate concentration at low values and independent of it at high ones. Although they recognized that these two conditions were special cases of the Monod equation, others who adopted their first-order equation incorrectly considered it to be an alternative expression.

3.2.7.3 Inhibitory Substrates

On occasion, particularly in the treatment of synthetic (xenobiotic) organic compounds in industrial wastewaters, situations are encountered in which the specific growth rate of the microorganisms reaches a maximum and then declines as the substrate concentration is increased, as illustrated in Figure 3.2. Obviously, the Monod equation is not adequate for depicting this situation and, consequently, considerable effort has been expended to determine an appropriate equation.[40,106,125] As with normal, naturally occurring, noninhibitory (biogenic) substrate, many different models could be used to represent the observed relationship between the substrate concentration and μ, and from a statistical point of view there is little to recommend one over another.[40,125] Consequently, as with the Monod equation, it has been argued that model selection should be based on familiarity and ease of use, leading to a recommendation that an equation based on the enzymatic model of Haldane[58] should be used. Andrews[4] was the first to propose general use of such a function for depicting the effects of inhibitory organic substrates on bacterial growth rates, and thus it will be called the Andrews equation herein. Its form is

$$\mu = \hat{\mu} \frac{S_S}{K_S + S_S + S_S^2/K_I}. \tag{3.39}$$

Examination of Equation 3.39 reveals that it is similar to the Monod equation, containing only one additional parameter, K_I, the inhibition coefficient. Note that when K_I is very large the Andrews

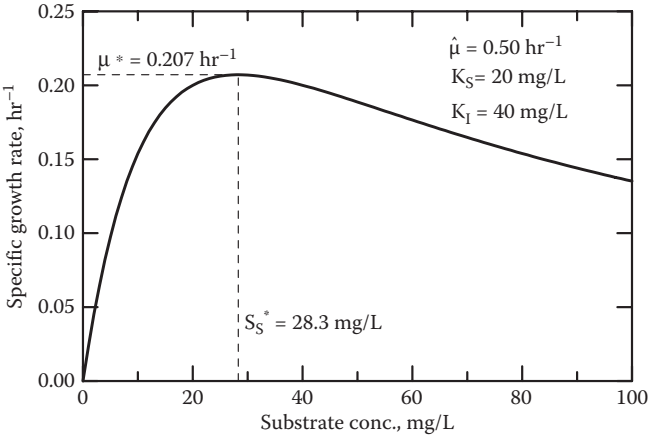

FIGURE 3.2 Typical plot of the relationship between the specific growth rate coefficient and the concentration of an inhibitory substrate. The parameter values given were used to construct the curve with the Andrews equation (Equation 3.39). Note that the values of $\hat{\mu}$ and K_S are the same as in Figure 3.1.

equation simplifies to the Monod equation, demonstrating that $\hat{\mu}$ and K_S have the same meaning in both equations. Unlike the situation for a noninhibitory substrate, however, $\hat{\mu}$ cannot actually be observed and thus is a hypothetical maximum specific growth rate that would be attained if the substrate were not inhibitory. Furthermore, since $\hat{\mu}$ cannot be observed, K_S also takes on a hypothetical meaning. The most outstanding characteristic of the curve in Figure 3.2 is that μ passes through a maximum, μ^*, at substrate concentration S_S^*, where

$$\mu^* = \frac{\hat{\mu}}{2(K_S/K_I)^{0.5} + 1} \tag{3.40}$$

and

$$S_S^* = (K_S \cdot K_I)^{0.5}. \tag{3.41}$$

Equation 3.40 is important because it demonstrates that the degree of inhibition is determined by K_S/K_I and not just by K_I alone. The larger K_S/K_I, the smaller μ^* is relative to $\hat{\mu}$, and thus, the greater the degree of inhibition. Furthermore, because they are measurable, μ^* and S_S^* are important in the determination of the kinetic parameters for inhibitory substrates. Equation 3.39 has been used widely in the modeling of various wastewater treatment systems, and will be adopted herein for depicting the effect of an inhibitory substrate on the specific growth rate of bacteria degrading it.

3.2.7.4 Effects of Other Inhibitors

Sometimes one compound may act to inhibit microbial growth on another compound. For example, some organic chemicals are known to inhibit the growth of nitrifying bacteria,[72,143] whereas others inhibit the growth of heterotrophic bacteria on biogenic organic matter.[150] In those cases it is necessary for the kinetic expression to depict the effect of the concentration of the inhibitor (S_i) on the relationship between μ and S_S. If the Monod equation can be used to relate μ to S_S in the absence of the inhibitor, then the effect of the inhibitor can be expressed as an effect on $\hat{\mu}$ and/or K_S.[62,149] Several types of inhibitors have been defined by analogy to enzyme inhibition, but all can be modeled by an extension of the Monod model proposed by Han and Levenspiel:[60]

$$\mu = \hat{\mu}\left(1 - \frac{S_i}{S_i^*}\right)^n \left[\frac{S_S}{S_S + K_S(1 - S_i/S_i^*)^m}\right], \tag{3.42}$$

where S_i^* is the inhibitor concentration that causes all microbial activity to cease and m and n are exponents that reflect the impact of increasing inhibitor concentrations on K_S and $\hat{\mu}$, respectively. Equation 3.42 has been used successfully to model the effects of various xenobiotic compounds on the removal of biogenic organic matter.[150] Its use will be discussed in Chapter 22.

3.2.8 Specific Substrate Removal Rate

In earlier sections it was stated that the basis for writing stoichiometric equations was arbitrary and that the reference component was the choice of the investigator. Thus, it is not surprising that many investigators[86,100,148] have selected substrate removal, rather than biomass growth, as their basic event and have written their rate equations accordingly. Combining Equations 3.34 and 3.35 yields:

$$r_{SS} = -\left(\frac{\mu}{Y}\right) X_B. \tag{3.43}$$

The term μ/Y has been called the specific substrate removal rate and given the symbol q.[54] (Note that the subscript H has been dropped from Y and X_B to emphasize the general nature of Equation 3.43.) Obviously, q will be influenced by S_S in exactly the same way as μ, and Equations 3.37 through 3.42 can all be written in terms of it. When this is done, the maximum specific substrate removal rate, $\hat{q}$, is used in place of $\hat{\mu}$, where

$$\hat{q} = \frac{\hat{\mu}}{Y}. \tag{3.44}$$

Both first- and zero-order approximations have been used for the relationship between q and S_S, just as they have for μ. In fact, the ratio of $\hat{q}$ over K_S has been called the mean reaction rate coefficient and given the symbol k_e:[37]

$$k_e = \frac{\hat{q}}{K_S}, \tag{3.45}$$

where k_e has units of L/(mg biomass COD · hr). All restrictions that apply to the approximate expressions for the effect of S_S on μ also apply to q.

3.2.9 Multiple Limiting Nutrients

In the broad sense, nutrients can be divided into two categories: complementary and substitutable.[12] Complementary nutrients are those that meet entirely different needs by growing microorganisms. For example, ammonia provides the nitrogen needed for protein synthesis while glucose provides carbon and energy. If either was missing from the growth medium and no substitute was provided, no growth would occur. Substitutable nutrients, on the other hand, are those that meet the same need. For example, ammonia and nitrate can both provide nitrogen whereas glucose and phenol can both provide carbon and energy. Thus, ammonia and nitrate are substitutable for each other, as are glucose and phenol. In this section we will consider simultaneous limitation of specific growth rate by two complementary nutrients. As stated previously, consideration of the effects of multiple carbon sources (i.e., multiple substitutable nutrients) is very complex,[88] and thus consideration of it will be delayed until Chapter 22.

In spite of its potential importance in the environment, relatively little is known about how microorganisms respond to simultaneous limitation by two or more complementary nutrients.[12]

Because the uncertainty increases greatly as the number of nutrients involved increases, we will limit our considerations to only two.

3.2.9.1 Interactive and Noninteractive Relationships

Consider two complementary nutrients, S_{S1} and S_{S2}. Both are required for biomass growth and both are present at low concentration in the environment in which the biomass is growing. Which will control the specific growth rate? Two different philosophies have been developed to answer this question and the models representing them have been classified as interactive and noninteractive.[9]

An interactive model is based on the assumption that two complementary nutrients can both influence the specific growth rate at the same time. If both are required for growth and each is present at a concentration equal to its half-saturation coefficient, then each alone can reduce μ to one-half of $\hat{\mu}$. However, since both effects are occurring simultaneously, the result would be to reduce μ to one-fourth of $\hat{\mu}$. The most common type of interactive model in use is the multiple Monod equation:[9,134]

$$\mu = \hat{\mu}\left(\frac{S_{S1}}{K_{S1} + S_{S1}}\right)\left(\frac{S_{S2}}{K_{S2} + S_{S2}}\right). \tag{3.46}$$

Any time the concentrations of S_{S1} and S_{S2} are such that both $S_{S1}/(K_{S1} + S_{S1})$ and $S_{S2}/(K_{S2} + S_{S2})$ are less than one, they both act to reduce μ below $\hat{\mu}$. This has two impacts. First, for a given value of S_{S1}, μ will be lower when S_{S2} is also limiting than it would be if S_{S2} were present in excess. Second, there is not a unique value of μ associated with a given value of S_{S1} or S_{S2} as there was with Equation 3.36. Rather, it depends on both.

A noninteractive model is based on the assumption that the specific growth rate of a microbial culture can only be limited by one nutrient at a time. Therefore, μ will be equal to the lowest value predicted from the separate single-substrate models:[140]

$$\mu = \min\left(\frac{\hat{\mu}S_{S1}}{K_{S1} + S_{S1}}, \frac{\hat{\mu}S_{S2}}{K_{S2} + S_{S2}}\right). \tag{3.47}$$

If $S_{S1}/(K_{S1} + S_{S1}) < S_{S2}/(K_{S2} + S_{S2})$, nutrient S_{S1} is rate limiting and vice versa. If $S_{S1}/(K_{S1} + S_{S1}) = S_{S2}/(K_{S2} + S_{S2})$, then both are rate limiting but that occurs only under special conditions. In the noninteractive conceptualization, the normal Monod equation (Equation 3.36) would apply for whichever nutrient was rate limiting and the concentration of the other would have no impact on μ.

Only limited experimental evidence is available to support one model over the other. Bae and Rittmann[10] have shown both theoretically and experimentally that the interactive model is more appropriate when the two limiting constituents are the electron donor and acceptor. Furthermore, Bader[9] has compared the mathematical characteristics of the two expressions. The noninteractive model, by its very nature, causes a discontinuity at the transition from one nutrient limitation to another. It also predicts significantly higher growth rates in the region where S_{S1}/K_{S1} and S_{S2}/K_{S2} are small. The interactive model does not cause discontinuities, but may err on the side of predicting lower growth rates when S_{S1}/K_{S1} and S_{S2}/K_{S2} are both small. Both functions become asymptotically the same if either nutrient is present in excess. Finally, the interactive model is mathematically preferable for modeling dynamic situations because it is continuous.

Equation 3.46, the interactive model, will be adopted for use herein. There are three reasons for this choice. First is the evidence provided by Bae and Rittmann.[10] Second, for the type of situation likely to be encountered in biochemical operations for wastewater treatment, the interactive model is more conservative. Third, it works well when one nutrient is the electron donor (i.e., the substrate) and the other is the electron acceptor (i.e., oxygen or nitrate),[126,134] a common occurrence in wastewater treatment systems.

A special case of multiple nutrients occurs when an increase in the concentration of one nutrient acts to diminish microbial activity. For example, consider the growth of heterotrophic bacteria under anoxic conditions. Because nitrate reduction can serve as an alternative to aerobic respiration, the enzymes involved in the transfer of electrons to nitrate and its reduced products are influenced negatively by the dissolved oxygen (DO) concentration and consideration must be given to this fact when expressing the kinetics of growth under anoxic conditions. Oxygen can have two effects; it can repress the synthesis of denitrifying enzymes and it can inhibit their activity.[31,82,113,138] Although there are exceptions, as a general rule the presence of oxygen in the medium (and/or its active utilization as the terminal electron acceptor) represses the synthesis of the nitrate reducing enzyme system. When oxygen is absent, or is present in amounts that are insufficient to meet the needs of the culture, derepression occurs, and the enzymes are synthesized. Complications occur, however, when the biomass is cycled between aerobic and anoxic conditions and this appears to alter the regulatory system so that some enzyme synthesis can continue at diminished rates even in the presence of DO.[133] The effect of oxygen on the activity of the enzymes depends on the bacterial species involved. In some, the activities are diminished in the presence of oxygen, whereas in others they are not. Nevertheless, it appears that inhibition of enzyme activity by oxygen is the primary mechanism influencing nitrate reduction rates in systems in which the bacteria are continually cycled between aerobic and anoxic conditions,[133] and that prior growth under anoxic conditions will provide an enzyme that can function at a diminished rate even in the presence of DO. One factor complicating the determination of the effects of oxygen on nitrate reduction in wastewater treatment systems is the necessity to grow the bacteria as flocculent cultures or as biofilms. Because diffusion is the only mechanism supplying oxygen to the bacteria in the interior of a floc particle or biofilm, some bacteria may be in an environment completely devoid of oxygen even when DO is present in the bulk liquid.[83]

Because of the complexity associated with the effects of DO on anoxic growth of heterotrophic bacteria and because all effects have not been clearly defined, relatively simple models have been used to express them.[13,65,66] A popular approach has been to use Equation 3.46 to depict the simultaneous effects of organic substrate, S_S, and nitrate, S_{NO}, on μ, but to add a third term that diminishes μ as the DO concentration, S_O, increases:

$$\mu = \hat{\mu}\left(\frac{S_S}{K_S + S_S}\right)\left(\frac{S_{NO}}{K_{NO} + S_{NO}}\right)\left(\frac{K_{IO}}{K_{IO} + S_O}\right). \tag{3.48}$$

The third term is the function most commonly used to depict the effects of a classical noncompetitive inhibitor as modeled in enzyme kinetics.[149] The parameter K_{IO} is the inhibition coefficient for oxygen.

3.2.9.2 Implications of Multiple Nutrient Limitation

Biochemical operations are designed on the premise that there is a functional relationship between the specific growth rate of biomass and the concentration of the growth-limiting nutrient in a bioreactor. Because of that relationship, if engineering control can be exerted over the specific growth rate, it will be possible to control the concentration of the growth-limiting nutrient leaving the bioreactor. This can only be achieved, however, if the nutrient the engineer wishes to control is the growth-limiting one. If the design objective is the removal of soluble organic matter, then all other nutrients must be supplied in excess. Or, if the goal is to remove nitrate-N by allowing it to serve as the terminal electron acceptor, then it should be made rate limiting at the appropriate place in the process. A clear definition of the objective to be met must be combined with knowledge of the concentrations of the various constituents in the wastewater to ensure that the resultant biochemical operation can indeed meet that objective.

Because oxygen is a gas of very low solubility, it must be supplied continuously to aerobic systems and the concentration actually in solution will depend on the relative rates of supply and

utilization. Furthermore, because oxygen transfer is one of the major costs associated with aerobic wastewater treatment, it is uneconomic to oversize the oxygen delivery system. As a consequence, it is not uncommon for the oxygen concentration to decrease sufficiently to make $S_O/(K_O + S_O) < 1.0$ (where K_O is the half-saturation coefficient for DO). Thus, it would be instructive to examine the impact of this occurrence. Figure 3.3 illustrates the simultaneous limitation of the specific growth rate of autotrophic nitrifying bacteria (modeled as a single step reaction) by ammonia, the electron donor, and oxygen (the electron acceptor) using typical parameter values. These bacteria were chosen because they are more sensitive to DO concentration than heterotrophic bacteria (i.e., $K_{O,H} < K_{O,A}$, where the subscripts H and A signify heterotrophic and autotrophic bacteria, respectively). Examination of Figure 3.3 reveals two things. First, if we could operate a bioreactor in a way that maintained a constant specific growth rate, decreasing the oxygen concentration in the bioreactor would cause the ammonia concentration to increase. Second, decreasing the oxygen concentration is analogous to decreasing $\hat{\mu}$ for the bacteria. This can also be seen by examining Equation 3.46. The consequence of this will be discussed in more detail in Chapter 6, but suffice it to say now that a decrease in $\hat{\mu}$ makes it more difficult for the autotrophic bacteria to compete for space in the bioreactor.

Both nitrogen and phosphorus are required for the synthesis of new biomass. If those proper quantities are not present, balanced biomass growth cannot occur and treatment performance will be impaired. Thus, care must be exercised to provide sufficient quantities. We have just seen that, however, if the concentrations of essential nutrients are very low in a bioreactor they can become rate limiting, which is undesirable when the treatment objective is removal of organic matter. This means that the concentration of nitrogen or phosphorus supplied to a bioreactor must be sufficiently high to meet the synthesis needs of the biomass as defined by stoichiometry while leaving enough residual in solution to prevent their concentrations from being rate limiting. Goel and Gaudy[49] determined that K_S for ammonia nitrogen during normal heterotrophic growth lies between 1.5 and 4.0 mg/L as N. Using 0.50 hr^{-1} as a representative value for $\hat{\mu}$, it can be shown that if the influent nitrogen concentration exceeds the stoichiometric requirement by 1.0 mg/L as N, nitrogen will not be rate limiting to heterotrophic biomass at the specific growth rates normally employed in wastewater treatment. Although some work has been done on kinetic limitation of heterotrophs by phosphorus, the results are not as clear as those with nitrogen. Attempts to measure the limiting phosphorus concentration in both pure and mixed microbial cultures found it to be too low to detect with the techniques available at the time.[127] Consequently, if the concentration of phosphorus in the influent exceeds the stoichiometric amount by a few tenths of a mg/L as P, phosphorus should not be rate

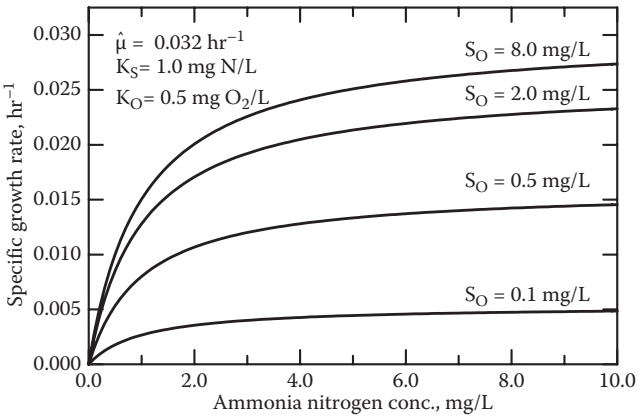

FIGURE 3.3 Double Monod plot showing the effects of both ammonia nitrogen and dissolved oxygen concentrations on the specific growth rate of autotrophic nitrifying bacteria. The parameter values given were used to construct the curves with Equation 3.46.

limiting. In some biochemical operations, the microorganisms pass through a growth cycle, and nutrients will be taken up in one phase and released in another. To prevent nutrient limitation during the phase of nutrient uptake, the amounts presented above should be in excess of the maximum quantity removed, not the net amount as determined by the final effluent.

3.2.10 Representative Kinetic Parameter Values for Major Microbial Groups

3.2.10.1 Aerobic Growth of Heterotrophic Bacteria

The values of the parameters $\hat{\mu}_H$ and K_S are very dependent on the organism and substrate employed. If an axenic bacterial culture is grown on each of several substrates under fixed environmental conditions, the values of $\hat{\mu}_H$ and K_S will vary from substrate to substrate. Likewise, if the same substrate is fed to each of several pure cultures, the values of $\hat{\mu}_H$ and K_S will depend on the species of organism. This makes it very difficult to generalize about parameter values and care should be exercised in the use of values considered to be "typical." However, it can be stated that readily biodegradable substrates are characterized by high values of $\hat{\mu}_H$ and low values of K_S, whereas slowly biodegradable substrates have low $\hat{\mu}_H$ values and high K_S values. For example, benzoic acid had $\hat{\mu}_H$ values between 0.61 and 0.64 hr^{-1} and K_S values between 4.2 and 5.8 mg/L as COD, whereas 2-chlorophenol had values of 0.020–0.025 hr^{-1} and 16–17 mg/L as COD for the two parameters.[30] Even lower K_S values have been reported for very easily degradable substrates, such as biogenic materials like carbohydrates and amino acids, with values as low as 0.2 mg/L for galactose and 0.5 mg/L for glutamic acid.[27] This means that degradation of many biogenic substrates may behave in a zero-order manner over a broad range of substrate concentrations.

Wastewaters usually contain complex mixtures of organic compounds and the total concentration of biodegradable soluble organic matter is commonly characterized by the COD concentration. When K_S is measured on such mixtures using the COD concentration, the values are generally one to two orders of magnitude higher than they are for single substrates expressed as COD. For example, poultry and soybean processing wastewater have been reported to have K_S values of 500 and 350 mg/L, respectively,[78] as five day biochemical oxygen demand (BOD_5), which is another measure of biodegradable organic matter. Thus, as a whole, overall removal of organic matter in wastewater treatment systems may behave in a first-order manner even though the removal of individual constituents may be zero order.[142]

Domestic wastewater is perhaps the most common example of a complex substrate, and because of its ubiquity, there has been considerable interest in characterizing its biodegradation kinetics. As one might expect from the discussion above, considerable variation in the parameter values has been reported, with $\hat{\mu}_H$ ranging from 0.12 to 0.55 hr^{-1} and K_S from 10 to 180 mg/L as COD.[65,66] An important characteristic of domestic wastewater is that the organic component can be divided into readily and slowly biodegradable fractions, thereby improving the ability of mathematical models to mimic process performance.[35] Use of this division should decrease the range of values observed. As a consequence, values of 0.25 hr^{-1} and 20 mg/L as COD have been adopted as representative of the $\hat{\mu}_H$ and K_S values for the readily biodegradable fraction.[66]

The microbial communities in wastewater treatment systems are complex, containing many microbial species, and the relative predominance of the species depends on the physical configuration of the system. Therefore, since the values of $\hat{\mu}_H$ and K_S are species dependent, it follows that their values in mixed culture systems will depend on the bioreactor configuration. For example, reactors that subject the microorganisms to variations in substrate concentrations from very high to very low tend to select species called r-strategists that can grow rapidly (higher $\hat{\mu}_H$) but tend to have low substrate affinity (high K_S), whereas reactors that maintain a low, uniform substrate concentration throughout select microorganisms called K-strategists that are good scavengers of substrate (low K_S).[6,27,34] This complicates kinetic analysis and requires that experiments to determine kinetic parameters be conducted with systems that mimic the physical configuration to be employed in the full-scale facility.

The biodegradation kinetics for many xenobiotic compounds can best be characterized by the Andrews equation (Equation 3.39). Dividing both the numerator and denominator by K_S yields:

$$\mu = \hat{\mu}\frac{S_S/K_S}{1 + S_S/K_S + (S_S/K_S)^2(K_S/K_I)}. \tag{3.49}$$

Expressing the equation in this manner emphasizes that the degree of substrate inhibition is determined by the ratio of K_S/K_I, rather than by K_I alone, as we saw with Equation 3.40. Furthermore, Equation 3.49 also makes it easy to see that the larger the ratio, the more inhibitory the substrate is. Both 1,3- and 1,4-dichlorobenzene are moderately inhibitory compounds and have ratios of 0.14 and 0.08, respectively.[52]

In Section 3.2.9, the undesirability of oxygen being rate limiting was discussed, suggesting that knowledge of the oxygen half-saturation coefficient for heterotrophs, $K_{O,H}$, is important. In spite of that, relatively little work has been done to estimate $K_{O,H}$ values for mixed microbial cultures, probably because population shifts occur in the community in response to changes in the DO concentration, making estimation of the value difficult. Nevertheless, limited pure culture data suggests that $K_{O,H}$ is very low. For example, values of 0.01, 0.08, and 0.15 mg O_2/L have been reported for *Sphaerotilus natans*[85] (a filamentous bacterium), *Candida utilis*[134] (a yeast), and *Citrobacter* sp.[85] (a floc-forming bacterium), respectively. This suggests that DO concentrations must be very low before they have serious impacts on the growth of heterotrophic bacteria, although they may influence the competition between filamentous and floc-forming bacteria. For depicting the impacts of DO on the general heterotrophic biomass growth, one group adopted a value of 0.2 mg O_2/L for $K_{O,H}$.[66]

It will be recalled from Section 3.2.8 that many investigators use substrate removal, rather than biomass growth, as the primary event with which to characterize biochemical operations. In that case, the primary kinetic parameter is the maximum specific substrate removal rate, $\hat{q}$, rather than the maximum specific growth rate. Equation 3.44 defined $\hat{q}$ as $\hat{\mu}/Y$. Thus, $\hat{q}$ will be influenced by variations in Y as well as variations in $\hat{\mu}$. Like $\hat{\mu}$, Y is influenced both by the substrate being degraded and the microorganism performing the degradation (see Section 2.4.1). However, it should be noted that Y is a reflection of the energy available in a substrate whereas $\hat{\mu}$ is a reflection of how rapidly a microorganism can process that energy and grow. Because they represent different characteristics, there is no correlation between the two parameters. For example, some substrates that are degraded very slowly (i.e., low $\hat{\mu}$) provide more energy to the degrading culture (i.e., higher Y) than do substrates that are degraded rapidly.[52] This suggests that deductions about the variability in $\hat{q}$ cannot be made from data on $\hat{\mu}$ alone and vice versa. Knowledge of the true growth yield is also important. Typical Y values are discussed in Section 2.4.1.

3.2.10.2 Anoxic Growth of Heterotrophic Bacteria

As we saw in Chapter 2, the only difference between aerobic and anoxic growth of heterotrophic bacteria on many biogenic organic substrates is the nature of the terminal electron acceptor and its impact on the amount of adenosine triphosphate (ATP) that the cells can generate. Thus, for substrates for which this is true, we might expect the kinetic parameters describing growth under the two conditions to be very similar and that is exactly what has been observed. When mixed microbial cultures were grown with excess oxygen or nitrate as the terminal electron acceptor and peptone as the rate-limiting substrate, the values of $\hat{\mu}_H$ and K_S were very similar, being 0.14 hr^{-1} and 67 mg/L as COD, respectively, under aerobic conditions and 0.13 hr^{-1} and 76 mg/L as COD under anoxic conditions.[97] Furthermore, as expected from the lower potential ATP formation under anoxic conditions, the anoxic yield was lower, being only 0.39 mg biomass COD/mg substrate COD versus 0.71 aerobically. Consequently, $\hat{q}_H$ was almost twice as large under anoxic conditions. Although data directly comparing kinetic parameters for biogenic substrates under aerobic and anoxic conditions are limited, experience with treatment systems suggest that these findings are generally

true.[97] Aromatic compounds, on the other hand, do not follow the patterns observed for more readily degradable substrates. Because different pathways are used to metabolize aromatic compounds under aerobic versus anoxic conditions, the kinetics are not related or similar.[32]

Anoxic growth conditions are generally imposed in biochemical operations for the purpose of reducing the nitrate concentration to low levels. Thus, there is a possibility that the terminal electron acceptor concentration will become rate limiting. Proper modeling of this situation requires knowledge of K_{NO}, the half-saturation coefficient for nitrate. As with oxygen, the half-saturation coefficient for nitrate as the terminal electron acceptor has been found to be low, with values around 0.1–0.2 mg/L as N being reported.[26,41,112] Consequently, values in that range have been adopted by investigators conducting modeling studies.[13,34]

Another parameter required to fully define the kinetics of microbial growth under anoxic conditions is K_{IO}, the oxygen inhibition coefficient used in Equation 3.48. If the cells are growing in a dispersed state so that all are exposed to the oxygen concentration in the bulk liquid, it appears that they do not denitrify when the DO concentration is above 0.1–0.2 mg/L.[121] However, when they grow as aggregates or films, the requirement for oxygen transport by diffusion allows the biomass in the interior to be free of oxygen even when the bulk liquid contains it. Consequently, anoxic growth will occur even when the DO concentration in the bulk liquid exceeds 0.2 mg/L.[121] Thus modelers have assumed values for K_{IO} ranging from 0.2[66] to 2.0[13] mg/L.

3.2.10.3 Aerobic Growth of Autotrophic Bacteria

The nitrifying bacteria are the most important aerobic autotrophs and for the nitrogen levels normally found in domestic wastewater the kinetics of their growth can be adequately represented by the Monod equation (Equation 3.36). Nitrifiers are not as diverse as the heterotrophic bacteria found in wastewater treatment systems and there is less variability in the uninhibited kinetic parameter values describing their growth. Nevertheless, nitrifier growth is quite vulnerable to inhibition by several factors and those factors can impose changes in kinetic parameter values, thereby having a large impact on nitrification performance.

The maximum specific growth rate coefficient for AOB has been reported to lie between 0.014[86] and 0.092[129] hr^{-1}, with a value of 0.032 hr^{-1} considered to be typical at 20°C.[123] The half-saturation coefficient for ammonia has been reported to be between 0.06 and 5.6 mg/L as N,[129] but is commonly assumed to be between 0.5 and 1.0 mg/L.[1,66,123]

Recently, designers have become interested in employing a two-step model that separates the growth kinetics of AOB and NOB to simulate treatment systems that employ partial nitrification in support of anammox bacteria and nitritation/denitritation (Sections 2.3.3 and 23.3.3). The maximum specific growth rate coefficient for NOB had been thought to be similar to that for AOB; however, recent estimates suggest that NOB grow more slowly, with the maximum specific growth rate ranging between 0.021 and 0.042 hr^{-1}.[21,79] The reported range of the half-saturation coefficient for NOB is slightly larger than that for AOB, being 0.06 to 8.4 mg/L as nitrite-N,[21,129] as is the value thought to be typical, 1.3 mg/L.[123]

The maximum specific growth rate coefficients for the autotrophic bacteria are considerably less than those for heterotrophic bacteria, reflecting their more restricted energy yielding metabolism and the fact that they must synthesize all cell components from carbon dioxide. This suggests that special consideration must be given to their requirements during the design of reactors in which both carbon oxidation and nitrification are to occur. Although the half-saturation coefficients for the autotrophs are less than the reported values for heterotrophs growing on complex substrates, they are similar to the values reported for heterotrophs growing on single organic compounds. As a consequence of their small size, the kinetics of nitrification will behave in a zero-order manner over a broad range of ammonia and nitrite concentrations. As will be seen later, this has a significant impact on bioreactor performance.

A major difference in the growth characteristics of heterotrophic and autotrophic biomass is the greater sensitivity of the latter to the concentration of DO. Whereas the value of the half-saturation

coefficient for oxygen is very low for heterotrophs, the values for AOB and NOB are sufficiently high in comparison to typical DO concentrations that dual nutrient limitation as expressed by Equation 3.46 should be considered to be the norm. Values of $K_{O,A}$ for AOB have been estimated at 0.74 and 0.99 mg/L of DO while estimates for NOB are higher at 1.4 and 1.75 mg/L of DO.[19,28] Measurements that considered the effects of diffusional resistance on half-saturation coefficients have suggested that the true values lie near the lower end of the range.[79,131,141]

Another difference between heterotrophic and autotrophic biomass is the greater sensitivity of the latter to changes in pH. Although all bacteria grow poorly outside of the normal physiological pH range of 6.0–8.0, nitrifying bacteria are particularly sensitive to pH, especially AOB, as shown in Figure 3.4.[117] There it can be seen that the rate reaches a maximum at a pH of about 8, and declines sharply for lower pH values. A wide range of pH optima has been reported,[136] but most workers agree that as the pH becomes more acidic the rate of ammonia oxidation declines.[112] Furthermore, if a culture is acclimated to a low pH, the effect is less severe than if the pH is suddenly shifted. Siegrist and Gujer[131] have modeled the effect in Figure 3.4 with Equation 3.50:

$$\hat{\mu}_A = \hat{\mu}_{Am}\left[1+10^{(6.5-pH)}\right]^{-1}, \tag{3.50}$$

where $\hat{\mu}_{Am}$ is the maximum specific growth rate at the optimum pH. It should be noted that this equation only predicts the decline in rate at low pH and does not predict the observed drop-off at pH above 8.5. This is not generally a problem with domestic wastewater treatment, however, because the release of hydrogen ions during nitrification acts to depress the pH so that values in excess of 8.5 are seldom encountered. There is less agreement concerning the effects of pH on NOB. For example, Boon and Laudelot[15] have suggested that their maximum specific growth rate is independent of pH over the range between 6.5 and 9, whereas others[135] have shown a strong pH dependence. One possible reason for this disagreement stems from the fact that the concentrations of the true substrates for both AOB and NOB are dependent on the pH, as discussed below.

The necessity for employing equations like 3.50 is due in part to the way in which the Monod equation is normally written for nitrifying bacteria. Although ammonia and nitrite are both ionizable

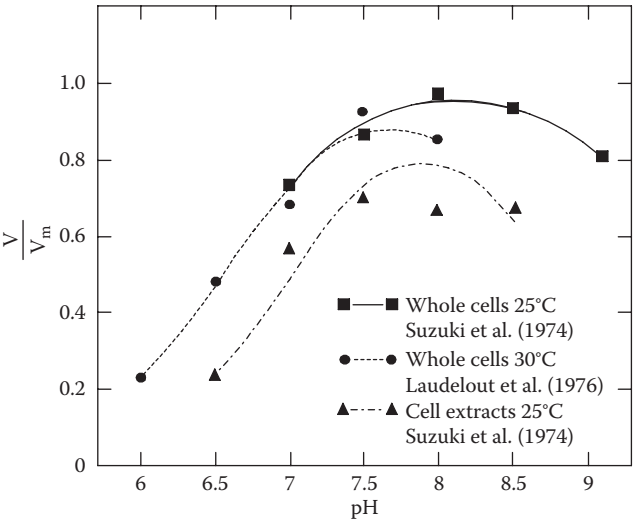

FIGURE 3.4 Effect of pH on the maximal activity of *Nitrosomonas*. The listed references are cited in reference 117. (Reprinted from Quinlan, A. V., Prediction of the optimum pH for ammonia-N oxidation by *Nitrosomonas europaea* in well-aerated natural and domestic-waste waters. *Water Research*, 18:561–66, 1984. Copyright © Elsevier Science Ltd. With permission.)

species, the Monod equation is normally written in terms of the total ammonia or nitrite concentration, without regard for the ionization state. However, the nonionized form of ammonia (free ammonia, FA) is thought to be the actual substrate for AOB,[117,139] and it is possible that undissociated (free) nitrous acid (FNA) is the substrate for NOB. For a given total ammonia concentration, the concentration of the nonionized form will change as the pH is changed, thereby making $\hat{\mu}$ as normally defined an apparent function of pH. A more direct approach would be to write the kinetic expression directly in terms of the true substrate and this has been done successfully for AOB and NOB by calculating the FA and FNA concentrations as a function of pH and applying the Andrews equation (Equation 3.39) to define the specific growth rate.[71] However, because this approach is more complex than combining Equation 3.36 with Equation 3.50 to reflect the effect of pH, the latter is more commonly used at nitrogen concentrations normally found in domestic wastewaters.

Free ammonia and undissociated nitrous acid become more of a problem at high nitrogen concentrations because they both act as inhibitory substrates as their concentrations are increased.[2,7,15] Furthermore, free ammonia can also inhibit nitrite oxidation to nitrate.[7] This suggests that there are complex relationships between the total ammonia and nitrite concentrations, the pH, and the activity of both groups of nitrifying bacteria. Indeed, the simple Monod equation is not adequate to depict the kinetics of nitrification when the concentration of ammonia exceeds that normally found in domestic wastewater (around 30–40 mg/L as N) and total ammonia-N or nitrite-N is used as the substrate in the equation. Again, use of the Andrews equation in combination with the FA and FNA concentrations as substrates has been used successfully to reflect the effect of either FA or FNA on nitrification rates.[70,71]

Because of the autotrophic nature of nitrifying bacteria, the concept developed that organic compounds display a general toxicity toward them. That this concept is fallacious has been demonstrated in pure[120] and mixed[69,73] cultures. Nitrification can proceed at rapid rates in the presence of organic matter, provided that other environmental factors, such as pH and DO concentration, are adequate. In fact, under some circumstances, the presence of biogenic organic matter can even enhance the rate of nitrification[73] and some nitrifiers can use simple organic compounds as a carbon source.[74] There are some organic compounds that are inhibitory, however, and act to decrease the specific growth rate of nitrifying bacteria. The most potent specific inhibitors of nitrification are compounds that chelate metals[73] and contain amine groups,[72] some of which are capable of decreasing the nitrification rate by 50% at concentrations of less than 1.0 mg/L. Furthermore, it appears that AOB are the weak link in the nitrification chain, being more susceptible than NOB to organic inhibitors.[72] Many inhibitors have been shown to act in a noncompetitive manner against nitrifiers,[87,110,111] allowing an equation like Equation 3.48 to be used to depict their effect:

$$\mu_A = \hat{\mu}_A \left(\frac{S_{NH}}{K_{NH} + S_{NH}} \right) \left(\frac{K_I}{K_I + S_i} \right), \tag{3.51}$$

where S_{NH} is the ammonia-N concentration, K_{NH} is the half-saturation coefficient for ammonia-N, S_i is the concentration of the inhibitor, and K_I is the inhibition coefficient. As might be expected, K_I is very small for some compounds,[111] denoting extreme inhibition. Although many inhibitors of AOB act in a noncompetitive manner, methane and ethylene act as competitive inhibitors.[80] This is because they are similar in size to ammonia and compete directly with it for the active site on the enzyme that initiates ammonia oxidation. Halogenated hydrocarbons act in a noncompetitive manner, but many are also reactive with the enzyme and can lead to products that damage the cell, thereby making their effects worse than simple inhibition. Finally, both AOB and NOB are known to be inhibited by light.[3,55] While light inhibition is not a major factor in full-scale wastewater treatment systems due to the opacity of activated sludge cultures, it is important to consider when performing laboratory experiments to determine kinetic parameters.

There have also been suggestions in the literature that the presence of heterotrophic bacteria is deleterious to the activity of nitrifying bacteria, but this has been shown to be false.[14,73] Any effect of

heterotrophs appears to be indirect, such as a decrease in DO concentration or an alteration of pH. Because of the sensitivity of autotrophs to these factors, care must be given to the design of facilities in which autotrophs and heterotrophs share the same space.

3.3 MAINTENANCE, ENDOGENOUS METABOLISM, DECAY, LYSIS, AND DEATH

As discussed in Section 2.4.2, a number of complex events interact to make the observed yield in biochemical operations less than the true growth yield and to cause only a fraction of the suspended solids to be active biomass. Even if our knowledge of all of those events was sufficient to allow mechanistically accurate kinetic models to be written, it is doubtful that they would be used in engineering practice because of their complexity. Consequently, as is common in engineering, simplified models have been adopted because of their utility and adequacy, and two will be reviewed in this section. The traditional approach has been in use for many years and has found many applications.[37,39,51,86,98] Its main attributes are its simplicity and familiarity. Its main weakness, however, is that while it can be used in environments in which a terminal electron acceptor is changing, it is not easy to do so. The second model, called the lysis:regrowth approach, handles changing electron acceptor conditions more easily.[34,35,65,66,91]

3.3.1 The Traditional Approach

In the traditional approach, all of the events leading to the reduction in yield and viability are expressed by the following stoichiometry:

$$\text{Biomass} + \text{electron acceptor} \rightarrow CO_2 + \text{reduced acceptor} + \text{nutrients} + \text{biomass debris.} \quad (3.52)$$

The important concepts incorporated into this expression are that active biomass is destroyed as a result of "decay" and that the electrons removed as a result of the oxidation of the carbon to carbon dioxide pass to the electron acceptor. Furthermore, not all of the biomass is totally oxidized and a portion is left as biomass debris.[77,96,98] Although the debris is ultimately biodegradable,[46,109] its rate of biodegradation is so low that for all practical purposes it is inert to further biological attack in most biochemical operations, causing it to accumulate, reducing the fraction of active biomass in the suspended solids. Finally, nitrogen is released as ammonia-N, although some remains in the biomass debris. Figure 3.5 illustrates how these events are related to microbial growth in an aerobic environment.

If Equation 3.52 is rewritten as a COD balance the result is

$$\text{Biomass COD} + [-(1-f_D)]O_2 \text{ equivalents of electron acceptor} \rightarrow f_D \text{ biomass debris COD,} \quad (3.53)$$

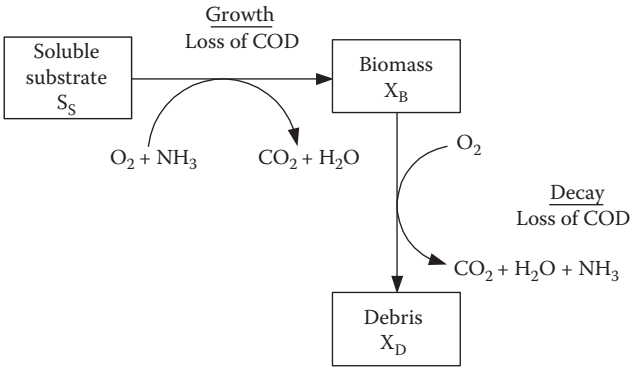

FIGURE 3.5 Schematic representation of the traditional approach to modeling biomass decay and loss of viability.

where f_D is the fraction of the active biomass contributing to biomass debris, X_D. For the type of biomass normally found in biochemical operations for wastewater treatment, it has a value of around 0.2.[34,96,99] Equation 3.53 shows that the utilization of oxygen or nitrate due to decay must equal the loss of active biomass COD minus the production of biomass debris COD.

Another important concept inherent in Equation 3.52 is that nitrogen is released as ammonia as biomass is destroyed. If Equation 3.52 were reformulated as a nitrogen-based stoichiometric equation it would read:

$$\text{Biomass N} \rightarrow \text{NH}_3\text{-N} + \text{biomass debris N}. \tag{3.54}$$

Since we have used biomass COD as the basic measurement of biomass, it would be convenient to write the nitrogen-based stoichiometric equation in a way that linked it to biomass COD. This can be done by introducing two conversion factors, $i_{N/XB}$ and $i_{N/XD}$, which are respectively, the mass of nitrogen per mass of COD in active biomass and the mass of nitrogen per mass of COD in biomass debris. Their use leads to

$$i_{N/XB} \cdot \text{biomass COD} \rightarrow \text{NH}_3\text{-N} + i_{N/XD} \cdot \text{biomass debris COD}. \tag{3.55}$$

Because the destruction of a unit mass of biomass COD leads to the generation of f_D units of biomass debris COD (Equation 3.53), Equation 3.55 tells us that the amount of ammonia-N released from the destruction of a unit mass of biomass COD is $(i_{N/XB} - i_{N/XD}f_D)$. If $C_5H_7O_2N$ is representative of biomass, then $i_{N/XB}$ has a value of 0.087 mg N/mg biomass COD. The nature of biomass debris is less well characterized than active biomass and thus there is no generally accepted empirical formula from which $i_{N/XD}$ can be calculated. However, because many nitrogenous compounds serve as energy reserves that are destroyed during endogenous metabolism, it is likely that the nitrogen content of biomass debris is less than that of biomass. As a result, a value of 0.06 mg N/mg COD has been recommended for $i_{N/XD}$.[65,66]

The rate expression for decay of biomass is first order with respect to the biomass concentration:

$$r_{XB} = -b \cdot X_B, \tag{3.56}$$

where b is the decay coefficient, with units of hr^{-1}. Employing the concept in Equation 3.10, the rate of production of biomass debris (r_{XD}) can be seen to be

$$r_{XD} = b \cdot f_D \cdot X_B, \tag{3.57}$$

and the rate of oxygen (electron acceptor) utilization associated with biomass decay is

$$r_{SO} = (1 - f_D)\, b \cdot X_B \ \text{(in COD units)} = -(1 - f_D)\, b \cdot X_B \ \text{(in } O_2 \text{ units)}. \tag{3.58}$$

The same equation would hold for utilization of nitrate expressed as oxygen equivalents, although the numerical value of the decay coefficient may well be different with alternative electron acceptors.[132] Finally, the rate of ammonia-N release (r_{SNH}) is

$$r_{SNH} = (i_{N/XB} - i_{N/XD} \cdot f_D)\, b \cdot X_B. \tag{3.59}$$

As might be expected from the discussion of parameter values in Section 3.2.10, the value of b is very dependent on both the species of organism involved and the substrate on which it is grown.

The latter effect is probably due to the nature of the energy reserves synthesized during growth. Because Equation 3.56 is an approximation describing very complex events, the value of b also depends to some extent on the rate at which the biomass is grown. Reported values for b for heterotrophic biomass in aerobic wastewater treatment systems are typically 0.01–0.03 hr^{-1}.[34,67] A large range of b values has been reported for autotrophic nitrifying bacteria,[29] with values ranging from 0.0002 to 0.007 hr^{-1}. A value of 0.003 hr^{-1} is considered typical at 20°C.[66]

3.3.2 The Lysis:Regrowth Approach

The most complete model depicting the loss of viability and biomass in biochemical operations was devised by Mason et al.[91] after an extensive review of the literature.[92] In that model, viable biomass can either die or be inactivated, leading to dead and nonviable biomass, respectively. Furthermore, all biomass can undergo lysis, although at different rates for different types, leading to soluble and particulate organic matter. The particulate organic matter is hydrolyzed to soluble organic matter, and the soluble organic matter from either source can be used by the viable biomass for new growth. Loss of viability is accounted for because the presence of dead biomass and particulate organic matter reduces the number of viable bacteria per unit mass of particulate material. Loss of biomass (i.e., decay) results from the fact that yield values are less than one so that the amount of biomass grown from the soluble substrate released is always less than the amount destroyed by lysis, as discussed in Section 2.4.1.

A conceptually similar, but less complex, model was developed by Dold et al.[35] for use in modeling wastewater treatment systems containing both aerobic and anoxic zones. Only one type of biomass is considered to be present: active, viable biomass. However, it is viewed as continually undergoing death and lysis, yielding particulate substrate and biomass debris. As in the model of Mason et al.,[91] particulate substrate is hydrolyzed to soluble substrate, and the soluble substrate is used by the viable biomass for growth, yielding new cell material. However, as above, because biomass yield values are always less than one, the amount of new biomass formed is always less than the amount destroyed by death and lysis, resulting in a net loss of biomass from the system (i.e., decay). A loss of viability results from the accumulation of biomass debris and particulate substrate.

The model of Dold et al.[35] is simpler than that of Mason et al.,[91] yet appears to be adequate for modeling many important wastewater treatment systems.[34] Furthermore, it can account for differences in decay observed as bacteria are cycled through aerobic, anoxic, and anaerobic conditions, whereas those differences cannot be easily accounted for by the traditional decay approach.[145] Finally, it has been adopted for use in a general model of single-sludge processes[65,66] that has been shown to adequately represent the dynamic performance of full-scale systems.[11] Thus, it will be used herein as an alternative to the traditional approach. The events in it are depicted in Figure 3.6.

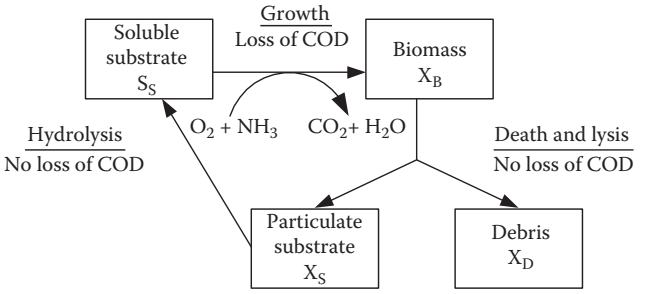

FIGURE 3.6 Schematic representation of the lysis:regrowth approach to modeling biomass decay and loss of viability.

The COD-based stoichiometry of the lysis:regrowth approach of Dold et al.[35] is

$$\text{biomass COD} \rightarrow (1 - f_D')\ \text{particulate substrate COD} + f_D'\ \text{biomass debris COD}, \tag{3.60}$$

where f_D' is the fraction of active biomass contributing to biomass debris. No COD is lost during death and lysis. Rather active biomass COD is simply converted into an equivalent amount of COD due to biomass debris and particulate substrate. As a consequence, no use of an electron acceptor is directly associated with the loss of biomass (i.e., "decay"). Electron acceptor utilization occurs as soluble substrate, which arises from hydrolysis of particulate substrate, is used by active biomass for growth. As with the traditional approach, cell debris is assumed to be resistant to microbial attack within the time constraints of biochemical operations.

The nitrogen in the biomass is divided between biomass debris and particulate substrate, with the latter being called particulate biodegradable organic nitrogen. The nitrogen-based stoichiometric equation depicting this is

$$\text{biomass N} \rightarrow \text{particulate biodegradable organic N} + \text{biomass debris N}. \tag{3.61}$$

Giving the same meanings to $i_{N/XB}$ and $i_{N/XD}$ as given above, Equation 3.61 can be rewritten in terms of biomass COD and biomass debris COD:

$$i_{N/XB} \cdot \text{biomass COD} \rightarrow \text{particulate biodegradable organic N} + i_{N/XD} \cdot \text{biomass debris COD}. \tag{3.62}$$

Thus, each unit of biomass COD lost to decay yields ($i_{N/XB} - i_{N/XD} \cdot f_D'$) units of particulate biodegradable organic nitrogen. This differs from the traditional approach that leads directly to soluble ammonia nitrogen.

As in the traditional approach, the rate of loss of biomass COD by death and lysis is considered to be first order with respect to the active biomass concentration:

$$r_{XB} = -b_L \cdot X_B, \tag{3.63}$$

where b_L has units of hr^{-1}, just as b does. In a manner similar to the traditional approach, the rate of production of biomass debris COD is

$$r_{XD} = b_L \cdot f_D' \cdot X_B. \tag{3.64}$$

And the rate of production of particulate substrate COD (r_{XS}) is

$$r_{XS} = (1 - f_D')\, b_L \cdot X_B. \tag{3.65}$$

Note the similarity of this equation to Equation 3.58, the equation for oxygen consumption in the traditional approach. This similarity arises from the retention in the particulate substrate of all electrons lost from active biomass, rather than their transfer to oxygen. Finally, the rate of production of particulate, biodegradable organic nitrogen (r_{XNS}) is

$$r_{XNS} = (i_{N/XB} - i_{N/XD} \cdot f_D')\, b_L \cdot X_B. \tag{3.66}$$

It is important to realize that b_L is conceptually and numerically different from b and that f_D' is numerically different from f_D. This follows from the cycling of COD that occurs in the lysis:regrowth approach. Biomass COD is lost, releasing particulate substrate COD, which is hydrolyzed to soluble substrate COD, which is degraded by active biomass yielding new biomass, which is lost by death and lysis giving particulate substrate COD, and so on. The net effect of the two approaches is the same because a given amount of biomass will be lost from a bioreactor regardless of how we conceptualize the actual events occurring. Since it is necessary for carbon to cycle around the system several times in the lysis:regrowth conceptualization to achieve the same loss of biomass that the traditional approach achieves in one pass, b_L must be numerically larger than b. Likewise, since the same amount of biomass debris is ultimately formed from the loss of a given amount of biomass by decay, f_D' must be numerically smaller than f_D. In fact, the values of the four parameters are related:[35]

$$f_D' \cdot b_L = f_D \cdot b. \tag{3.67}$$

Furthermore,

$$f_D' = \left(\frac{1 - Y}{1 - Y \cdot f_D}\right) f_D. \tag{3.68}$$

It was stated above that f_D has a value around 0.2. Given the Y values associated with the biomass for which f_D was estimated, Equation 3.68 suggests that the value of f_D' is around 0.08.[34] The values of f_D and f_D' are not likely to vary greatly, and thus those values will be adopted herein. However, it should be noted that the relationship between b_L and b also depends on Y:[34]

$$b_L = \frac{b}{[1 - Y(1 - f_D')]}. \tag{3.69}$$

Although it is common during parameter evaluation studies to measure both Y and b, neither f_D nor f_D' is commonly measured. Since Y can influence the relationship between b_L and b, it is recommended that Equation 3.69 be used instead of Equation 3.67 to convert measured b values to b_L values.[66]

An important assumption implicit in the lysis:regrowth approach is that within a given culture, cell lysis occurs all of the time with the same value of the rate coefficient b_L, regardless of the rate at which the bacteria are growing. The validity of this assumption has been confirmed by measuring the release of nucleic acids as direct evidence of cell lysis.[114]

For autotrophic growth, the relationship between b_L and b is different.[65,66] This is because autotrophic organisms do not use organic matter for growth. Thus, death and lysis will not lead to additional autotrophic biomass growth. (The amount of autotrophic biomass that will grow from the nitrogen released is negligible.) Rather, heterotrophic biomass will grow on the organic matter released. As a consequence, the lysis:regrowth and traditional approaches are the same for autotrophic biomass, with the result that the two parameter values are equal.

3.3.3 Endogenous Respiration with Storage

A modified approach to modeling biomass loss and oxygen utilization during decay assumes that heterotrophic bacteria form storage products when grown under conditions that alternate between feast (large amounts of available substrate per cell) and famine (starvation conditions).[147] An electron acceptor is used when the stored material is ultimately consumed to support growth and results

in endogenous respiration. This model assumes that substrate is not used directly to support growth but must, first, be incorporated into intracellular storage materials.[146] This approach also accommodates rate differences that are observed under aerobic and anoxic conditions.[47] In general, this approach improves modeling predictions for systems that experience significant substrate storage, but is not better for most conventional wastewater treatment applications.[47] Furthermore, many synthetic organic compounds found in industrial wastewaters are not transformed directly into intracellular storage products. Thus, the approach cannot be considered to be of general utility.

3.4 SOLUBLE MICROBIAL PRODUCT FORMATION

As discussed in Section 2.4.3, soluble microbial products are thought to arise from two processes, one growth associated and the other nongrowth associated.[124]

Growth associated product formation results directly from biomass growth and substrate utilization. If soluble microbial product formation was occurring in an appreciable amount, it would be necessary to modify the stoichiometric equation for microbial growth to account for it. Letting S_{MP} represent the concentration of soluble microbial products in COD units and Y_{MP} the microbial product yield in units of product COD formed per unit of substrate COD used, Equation 3.31 can be rewritten to account for soluble microbial product formation:

$$(1)S_S + [-(1-Y_H-Y_{MP})]\, S_O \rightarrow Y_H X_{B,H} + Y_{MP} S_{MP}. \tag{3.70}$$

This shows that less electron acceptor is used when soluble products are formed because part of the COD of the substrate remains in the medium as those products. Rewriting this equation in the form of Equation 3.9 with biomass as the reference constituent gives:

$$\left(-\frac{1}{Y_H}\right)S_S + (-1)\left[-\left(\frac{1-Y_H-Y_{MP}}{Y_H}\right)\right]S_O + X_{B,H} + \frac{Y_{MP}}{Y_H}S_{MP} = 0. \tag{3.71}$$

This tells us that the rate of soluble microbial product formation (r_{SMP}) is

$$r_{SMP} = (Y_{MP}/Y_H)r_{XB}. \tag{3.72}$$

Combining Equation 3.72 with Equation 3.35 for r_{XB} gives:

$$r_{SMP} = (Y_{MP}/Y_H)\mu \cdot X_{B,H}. \tag{3.73}$$

The fact that r_{SMP} is proportional to μ shows that it is growth associated.

Nongrowth associated product formation, also called biomass associated product formation,[124] occurs as a result of cell lysis and decay. Rewriting Equation 3.53 to incorporate soluble product formation into the COD-based stoichiometry of the traditional approach to decay gives:

$$\text{biomass COD} + [-(1-f_D-f_{MP})]\, O_2 \text{ equivalents of electron acceptor} \rightarrow$$

$$f_D \text{ biomass debris COD} + f_{MP} \text{ soluble product COD}, \tag{3.74}$$

where f_{MP} is the fraction of active biomass contributing to biomass associated products. Using this with Equation 3.56 gives the rate of production of biomass associated product (r_{SMP}):

$$r_{SMP} = b \cdot f_{MP} \cdot X_{B,H}. \tag{3.75}$$

By analogy to biomass debris formation, a similar approach could be used to account for soluble microbial product formation in the lysis:regrowth approach, giving a parameter f_{MP}' that is smaller than f_{MP} in the same way that f_D' is smaller than f_D.

Combining Equations 3.73 and 3.75 suggests that the specific rate of soluble microbial product formation is linearly related to the specific growth rate. While such a relationship may be adequate for slowly growing cultures like those found in activated sludge systems, it is not adequate for more rapidly growing systems[61] and thus Equations 3.73 and 3.75 cannot be considered to be of general applicability to all systems. Although a relatively large body of research on soluble microbial product formation has been conducted,[61,84,124] it is still not sufficient to allow consensus on the rate expressions to be used. Thus, in spite of its known importance, soluble microbial product formation will not be incorporated into the models in Parts II and IV.

Insufficient information is available to provide typical values for f_{MP} and f_{MP}', but, as indicated in Section 2.4.3, Y_{MP} values have been found to be less than 0.1.[52]

3.5 SOLUBILIZATION OF PARTICULATE AND HIGH MOLECULAR WEIGHT ORGANIC MATTER

The conversion of particulate and high molecular weight organic matter into forms small enough for bacteria to take up and degrade is an important step in biochemical operations for wastewater treatment because such materials are commonly present in wastewaters and also arise from lysis reactions as discussed previously. In spite of that, relatively few studies have focused on those reactions. Perhaps this is because many types of particulate materials are attacked by distinctly different mechanisms, even though they are collectively referred to as hydrolysis.

The stoichiometry of hydrolysis is thought to be very simple, with organic material simply changing form. Consequently, most investigators have assumed that COD is conserved (i.e., that no energy is consumed). This is indicated in Figure 3.6. Because no energy is consumed, no electrons are removed and no electron acceptor is used. Thus, the stoichiometric equation is simply:

$$\text{particulate substrate COD} \rightarrow \text{soluble substrate COD.} \tag{3.76}$$

This means that the rate of formation of soluble substrate COD is equal to the rate of loss of particulate substrate COD.

In the face of complex situations in which reactions are ill defined, it is common for engineers to choose the simplest possible reaction rate expression, and that is what a number of investigators have done, assuming that hydrolysis is first order with respect to the concentration of particulate substrate, X_S.[16,17,36,57,91] This approach, however, ignores the effect that the biomass concentration will have on the rate.

One group[66] performed an extensive literature survey before adopting a kinetic expression for the hydrolysis of particulate organic matter patterned after that of Dold et al.,[35] which is based on the work of Stenstrom:[137]

$$r_{XS} = -k_h \left[\frac{X_S/X_{B,H}}{K_X + (X_S/X_{B,H})} \right] X_{B,H}. \tag{3.77}$$

In this expression k_h is the hydrolysis coefficient (hr^{-1}) and K_X is a half-saturation coefficient (mg particulate substrate COD/mg active biomass COD). An important characteristic of this expression is that even though the rate is first order with respect to the heterotrophic biomass concentration, it is controlled by the ratio of particulate substrate concentration to heterotrophic biomass concentration, rather than by the particulate substrate concentration alone. This is necessary because the reaction is thought to be surface mediated, depending on the presence of extracellular enzymes whose quantity

will be proportional to the biomass concentration.[35] Actually, although not indicated in Equation 3.77, hydrolysis is a function of particle size, which decreases as hydrolysis progresses.[33] Because hydrolysis is a surface reaction, its rate increases as the particle surface area increases. Furthermore, for a given mass of particles, the particle surface area increases as the particle size decreases. Consequently, hydrolysis kinetics is more complex than can be captured by Equation 3.77. However, few models used in practice today capture this more complex reality. Equation 3.77 readily simplifies into distinct first-order expressions when $X_S \ll X_{B,H}$ and when $X_S \gg X_{B,H}$.[103] Consequently, for the purposes of this book, Equation 3.77 is considered sufficient to capture the effect of hydrolysis during most conventional wastewater treatment applications.

Data on the values of k_h and K_X are very limited. Based primarily on the recommendations of Dold and Marais,[34] one group[65] adopted a value of 0.092 hr^{-1} for k_h and a value of 0.15 for K_X. These values are adopted for use in this book.

The influence of an electron acceptor concentration on the hydrolysis rate has been controversial. Several researchers have shown that the rate of hydrolysis is influenced by the electron acceptor concentration,[34,65,66] even though no electron acceptor is used in the reaction. Later studies clarified that it is actually hydrolytic enzyme synthesis that is influenced by the electron acceptor condition, but that hydrolytic enzymes are quite stable and functional for hours to days under aerobic, anoxic, and anaerobic conditions.[50] Therefore, reactor configuration determines whether electron acceptor conditions will influence hydrolysis,[103] and we will see in Chapter 6 that different approaches are taken when incorporating hydrolysis into process models.

The most commonly employed activated sludge models consider the influence of electron acceptor conditions using the approaches introduced in Section 3.2.9. Under aerobic conditions, an interactive, dual nutrient limitation expression has been adopted in a manner similar to that in Equation 3.46. Under anoxic conditions, an expression similar to that in Equation 3.48 has been found to be appropriate, with nitrate stimulating anoxic hydrolysis and oxygen inhibiting it. In both expressions, the effect of the particulate substrate should be given by Equation 3.77. Under anaerobic conditions of short duration, hydrolysis is assumed to stop. While this would not be true for long-term anaerobic conditions, it is consistent with observations in biochemical operations that cycle bacteria between aerobic and anoxic conditions.[34] The approach for modeling hydrolysis under long-duration anaerobic conditions, such as those encountered during biomass digestion, will be discussed in Chapter 8.

As seen in Equation 3.61, biomass decay results in the formation of particulate biodegradable organic nitrogen. In addition, organic nitrogen will be associated with the particulate organic matter in the wastewater. All of this material will be converted into soluble, biodegradable organic nitrogen, S_{NS} (i.e., the nitrogen associated with amino acids and other soluble nitrogen containing organic substrates), as the particulate substrate is hydrolyzed. The rate of generation of S_{NS} (r_{SNS}) is numerically equivalent to the rate of loss of particulate organic nitrogen (r_{XNS}), which is proportional to the hydrolysis rate of particulate organic matter:[34]

$$r_{SNS} = -r_{XNS} = -(X_{NS}/X_S)r_{XS}, \tag{3.78}$$

where X_{NS} is the concentration of particulate, biodegradable organic nitrogen.

3.6 AMMONIFICATION AND AMMONIA UTILIZATION

Ammonification is the conversion of soluble organic nitrogen into ammonia-N that occurs as bacteria consume soluble organic matter containing nitrogen. Actually, the true rate of ammonification is difficult to measure because ammonia-N is being consumed by the bacteria as they grow, and thus the only measurable event is the net accumulation or loss of ammonia in the medium. If the amount of nitrogen available in the organic substrate is just sufficient to meet the biosynthetic needs of the new biomass, there will be no net change in the ammonia-N concentration in the medium.

On the other hand, if that amount exceeds the need, the ammonia concentration in the medium will increase, whereas if that amount is less than the need, the ammonia concentration will decrease. However, it should be recognized that whether organic nitrogen is incorporated directly into new biomass depends on its form. The nitrogen in simple compounds like amino acids may be incorporated directly as the amino acids are used for protein synthesis, while nitrogen in complex synthetic organic chemicals may be released to the medium as ammonia.

In an effort to make this complex situation mathematically tractable, most modelers assume that all nitrogen goes through the medium before being used. Thus, ammonification is assumed to release all organic nitrogen to the medium as ammonia, and nitrogen utilizing reactions are assumed to obtain their ammonia from the medium. Whether ammonia accumulates or is removed depends on the relative rates of its production and utilization.

Because ammonification occurs as heterotrophic biomass destroys nitrogen containing soluble organic matter, it is likely that its rate is proportional to the rate of soluble substrate removal. Relatively little work has been done to investigate the rate of ammonification in complex substrates in which only a part of the soluble organic matter contains nitrogen, and thus it is uncertain whether a direct proportionality can be assumed between soluble substrate removal and ammonification. Consequently, ammonification has been represented as a reaction that is first order with respect to both the heterotrophic biomass concentration and the concentration of soluble, biodegradable organic nitrogen:[34,65,66]

$$r_{SNS} = -k_a \cdot S_{NS} \cdot X_{B,H}, \tag{3.79}$$

where k_a is the ammonification rate coefficient (L/[mg biomass COD · hr]). Very little information is available about its value. As discussed above, the assumed stoichiometry of ammonification is such that all nitrogen removed from nitrogen containing soluble organic matter is released as ammonia, although some may ultimately be used for biomass synthesis. Thus, the rate of production of ammonia nitrogen (S_{NH}) through ammonification is

$$r_{SNH} = -r_{SNS.} \tag{3.80}$$

Ammonia is removed from a solution by two reactions. First, it is used in the synthesis of new biomass as seen in Equation 3.6 and others presented in Section 3.2. Second, it is used as a substrate by autotrophic biomass. The rate expression for the second use is the same as any other substrate, as discussed in Section 3.2.6. The rate expression for the first use can be determined from the generalized rate expression and the stoichiometry of growth. Since $i_{N/XB}$ is the mass of nitrogen per unit of biomass COD, the rate of ammonia removal through biomass growth is simply:

$$r_{SNH} = -i_{N/XB} \cdot \mu \cdot X_B. \tag{3.81}$$

Equation 3.81 is true for both heterotrophic and autotrophic growth, and thus the general symbol for biomass, X_B, has been used in it.

3.7 PHOSPHORUS UPTAKE AND RELEASE

Biological phosphorus removal is a complex process that is dependent on the growth of specialized phosphate accumulating organisms (PAOs), which store phosphorus as polyphosphate (Poly-P), as discussed in Section 2.4.6. To review, under anaerobic conditions PAOs cleave phosphate groups from Poly-P (releasing the phosphate to the medium), thereby obtaining the energy required to take up acetate and store it as poly-β-hydroxyalkanoate (PHA). They also obtain energy from the degradation of glycogen. When the PHA-rich biomass is transferred to aerobic or anoxic conditions,

the PAOs then metabolize the PHA providing the energy required for their growth, the formation of glycogen, and the uptake of phosphate for the formation of Poly-P. Given the appropriate ratio of COD to phosphorus in the wastewater undergoing treatment (to be discussed in Chapter 12), the PAOs are able to take up all of the phosphate released in the anaerobic zone plus the additional phosphate present in the wastewater, thereby achieving a net removal of phosphate. In this section we will present the kinetics of phosphate uptake and release, PHA utilization and formation, and PAO growth. Although kinetic expressions have been proposed for glycogen utilization and storage and the competition between glycogen accumulating organisms (GAOs) and PAOs,[90,152,154] they have not yet been well validated with field experience and are not included here.

Under anaerobic conditions, PAOs do not grow, but store acetate as PHA through glycogen degradation and the cleavage of Poly-P with the associated release of soluble phosphate (Figure 2.6). The rate of PHA storage in PAOs, r_{XPHA}, can be modeled with an interactive, dual limiting nutrient expression:

$$r_{XPHA} = \hat{q}_{PHA}\left(\frac{S_A}{K_A + S_A}\right)\left[\frac{X_{PP}/X_{B,PAO}}{K_{PP} + (X_{PP}/X_{B,PAO})}\right]X_{B,PAO}, \tag{3.82}$$

where $\hat{q}_{PHA}$ is the maximum specific rate of PHA formation, hr^{-1}; S_A is the acetate concentration in COD units; K_A is a half-saturation coefficient for acetate in COD units; X_{PP} is the Poly-P concentration in the biomass, expressed as a liquid phase P concentration; K_{PP} is the half-saturation coefficient for Poly-P, expressed as g Poly-P/g $X_{B,PAO}$ in COD units; and $X_{B,PAO}$ is the concentration of PAO biomass in COD units. Equation 3.82 shows that increasing concentrations of acetate and PAOs, and an increasing fraction of polyphosphate in the PAO biomass, all support a faster rate of PHA formation in PAOs. Although not shown here, it is also possible to include a limiting nutrient term for alkalinity that slows the rate of PHA formation if alkalinity drops below a threshold value.

Acetate provides most, but not all, of the acetyl-CoA needed to form PHA under anaerobic conditions, with the rest coming from glycogen consumption.[152,154] However, it is commonly assumed that all acetate consumption (as COD) goes to PHA formation (as COD), allowing a simple relationship between the rates:

$$r_{SA} = -r_{XPHA}, \tag{3.83}$$

where r_{SA} is the acetate consumption rate. Furthermore, Y_{PO_4} units of soluble phosphate, S_{PO_4}, are released for each unit of PHA formed as COD, increasing the soluble phosphate concentration by an amount equal to the decrease in stored Poly-P concentration. Thus, the rates of change of soluble phosphate (r_{SPO_4}), stored polyphosphate (r_{XPP}), and stored PHA (r_{PHA}) are all related:

$$r_{SPO_4} = -r_{XPP} = Y_{PO_4} \cdot r_{XPHA}. \tag{3.84}$$

The value of Y_{PO_4} selected by Henze et al.[67] was 0.40 mg P/mg COD, reflecting the average stoichiometry for the process. The values of K_A (4.0 mg COD/L) and K_{PP} (0.01 mg P/L) were chosen to be small relative to typical values for S_A and $X_{PP}/X_{B,PAO}$ to make the parenthetical terms in Equation 3.82 serve as switching functions that change rapidly from one to zero, thereby turning the reaction on and off. Estimates for $\hat{q}_{PHA}$ can vary widely and are influenced by pH, temperature, and solids retention time. Despite this variation, specific PHA formation rates for PAOs are typically reported to be around 3.0 hr^{-1} at 20°C.

Under aerobic conditions, the PAOs grow by using the stored PHA as a carbon and energy source (Figure 2.6). This is assumed to be their only substrate for growth, even though they are capable of growth on soluble substrates. The small amount of soluble substrate present in the aerobic zone of a biological phosphorus removal process is assumed to be consumed by non-PAO heterotrophic

bacteria, a simplifying but reasonable assumption.[67] Furthermore, to ensure that the rate expression reflects PAO growth only under aerobic conditions, a switching function for oxygen is included to make the rate go to zero when oxygen is absent. Considering all of these factors, the rate of PAO growth (r_{XBPAO}) under aerobic conditions can be described by

$$r_{XBPAO} = \hat{\mu}_{PAO}\left[\frac{X_{PHA}/X_{B,PAO}}{K_{PHA} + (X_{PHA}/X_{B,PAO})}\right]\left(\frac{S_{PO_4}}{K_{PO_4} + S_{PO_4}}\right)\left(\frac{S_O}{K_O + S_O}\right)X_{B,PAO}, \tag{3.85}$$

where $\hat{\mu}_{PAO}$ is the maximum specific growth rate coefficient for PAOs, X_{PHA} is the stored PHA concentration in mg/L as COD, S_{PO_4} is the soluble phosphate concentration in mg/L as P, K_{PO_4} is the half-saturation coefficient for soluble phosphate, S_O is the DO concentration, and K_O is the half-saturation coefficient for DO. It should be noted that the expression for the effect of PHA concentration on biomass growth is written in terms of the amount of PHA available per unit of PAO biomass COD because the PHA is not free in the medium, but is stored in the biomass. As a result, K_{PHA}, the half-saturation coefficient for PHA, has units of mg PHA COD/mg PAO COD. Because of biomass lysis, phosphate will continually be released to the medium. Consequently, S_{PO_4} will never reach a zero concentration and phosphorus will always be available for growth. The values chosen for the half-saturation coefficients by Henze et al.[67] were 0.01 mg PHA COD/mg PAO COD, 0.01 mg P/L, and 0.20 mg O_2/L, for K_{PHA}, K_{PO_4}, and K_O, respectively. As with anaerobic storage of PHA, a limiting nutrient term for alkalinity can be added to Equation 3.85 to reduce the rate of PAO growth if alkalinity becomes limiting and the pH drops. Furthermore, a limiting nutrient term can be added to consider situations when ammonia-N becomes rate limiting, for instance when complete nitrification occurs. Neither of these terms is included in Equation 3.85 for simplicity.

The stoichiometry of the PAO aerobic growth reaction on a COD basis is the same as that in Equation 3.33, except that PHA is the growth substrate. Consequently, the relationship between r_{XBPAO}, r_{XPHA}, and r_{SO} (all in COD units) will be the same as the relationship between r_{XB}, r_{SS}, and r_{SO} in Equation 3.34, or

$$\frac{r_{XPHA}}{\left(-\frac{1}{Y_{PAO}}\right)} = \frac{r_{SO}}{(-1)\left[-\left(\frac{1-Y_{PAO}}{Y_{PAO}}\right)\right]} = \frac{r_{XBPAO}}{1}, \tag{3.86}$$

where Y_{PAO} is the yield coefficient for PAOs growing on stored PHA. The value assumed for it is 0.63 mg PAO COD/mg PHA COD.[67] It should be noted that the rates of PHA loss and oxygen consumption expressed by Equation 3.86 are those associated only with PAO growth.

Storage of polyphosphate also occurs under aerobic conditions and the energy for it also comes from PHA utilization. Thus, the rate expression includes all of the parenthetical terms in Equation 3.85. However, it has been observed that storage of Poly-P stops if its content in the PAOs becomes too high.[67] Thus, it is necessary to include a term that decreases the rate of Poly-P storage as the Poly-P concentration per unit of PAOs approaches a maximum value of K_{PMAX}. Considering these factors, the rate of Poly-P storage under aerobic conditions, r_{XPP}, can be expressed as

$$r_{XPP} = \hat{q}_{PP}\left[\frac{X_{PHA}/X_{B,PAO}}{K_{PHA} + (X_{PHA}/X_{B,PAO})}\right]\left(\frac{S_{PO_4}}{K_{PO_4} + S_{PO_4}}\right)\left(\frac{S_O}{K_O + S_O}\right)$$
$$\cdot\left[\frac{K_{PMAX} - (X_{PP}/X_{B,PAO})}{K_{IPP} + K_{PMAX} - (X_{PP}/X_{B,PAO})}\right]X_{B,PAO}, \tag{3.87}$$

where $\hat{q}_{PP}$ is the maximum specific rate of Poly-P storage, which has a typical value of 0.06 mg P/(mg PAO COD·h) at 20°C. The K_{IPP} is the inhibition coefficient for Poly-P storage, with an

assumed value of 0.02 mg P/mg PAO COD. All other terms were defined following Equation 3.85. Soluble phosphate is removed from the medium in direct proportion to the amount incorporated into Poly-P. Furthermore, PHA is lost and oxygen is utilized proportionally as well. The relationship between the rates under aerobic conditions is determined from the stoichiometry as

$$\frac{r_{XPHA}}{(-Y_{PHA})} = \frac{r_{SO}}{(-1)(-Y_{PHA})} = \frac{r_{SPO_4}}{-1} = \frac{r_{XPP}}{1}, \tag{3.88}$$

where Y_{PHA} is the PHA requirement for Poly-P storage, which has a typical value of 0.20 mg PHA COD/mg P.[67] The rates of PHA loss and oxygen consumption in this expression are those associated with only Poly-P storage. The total rates of each under aerobic conditions due to Poly-P storage and PAO growth must be obtained by adding the expressions from Equations 3.86 and 3.88. Furthermore, Equation 3.88 does not give the total rate of soluble phosphate loss since polyphosphate formation is not the only mechanism for removing soluble phosphate from the liquid. Rather, phosphorus is also a required nutrient for biomass synthesis. If $i_{P/XB}$ is the mass of phosphorus incorporated into cell material per unit of PAO COD formed, the total rate of removal of soluble phosphorus by the PAOs will be

$$r_{SPO_4} = -(i_{P/XB} \cdot r_{XBPAO}) - r_{XPP}, \tag{3.89}$$

where r_{XBPAO} is given by Equation 3.85. Cellular biomass contains about 2.5% phosphorus on a mass basis, so on a biomass COD basis, $i_{P/XB}$ has a value around 0.02 mg P/mg biomass COD. If non-PAO heterotrophs and autotrophs are growing in the system, they will also consume soluble phosphate for incorporation into biomass with the same stoichiometry.

The processes for anoxic growth of PAOs and the uptake of soluble phosphate parallel those that occur under aerobic conditions except that nitrate is the electron acceptor. The rate of PAO growth under anoxic conditions, $r_{XBPAO,anx}$, uses the terms in Equation 3.85 except that the DO term becomes growth inhibiting in the presence of oxygen, a limiting nutrient term for nitrate is added, and an efficiency factor, $\eta_{g,PAO}$, is added to account for the decreased PAO activity under denitrifying conditions, either because the kinetics are truly slower or because not all PAOs can denitrify. The rate of PAO growth under anoxic conditions is described by

$$r_{XBPAO,anx} = \hat{\mu}_{PAO} \cdot \eta_{g,PAO} \left[\frac{X_{PHA}/X_{B,PAO}}{K_{PHA} + (X_{PHA}/X_{B,PAO})} \right] \left(\frac{S_{PO_4}}{K_{PO_4} + S_{PO_4}} \right) \cdot \left(\frac{S_{NO}}{K_{NO} + S_{NO}} \right) \left(\frac{K_O}{K_O + S_O} \right) X_{B,PAO}, \tag{3.90}$$

where all variables and parameters are as described previously. Henze et al.[67] used a value of 0.60 for the PAO anoxic growth efficiency factor ($\eta_{g,PAO}$) and a value of 0.50 mg N/L for the nitrate half-saturation coefficient, K_{NO}.

The stoichiometry of the PAO anoxic growth reaction on a COD basis is the same as under aerobic conditions, as given by Equation 3.86, except that a conversion factor is needed to convert the rate of nitrate consumption (r_{SNO}) to COD units and considering that it is the electron acceptor, as defined in Table 3.1:

$$\frac{r_{XPHA,anx}}{\left(-\frac{1}{Y_{PAO}}\right)} = \frac{r_{SNO}}{(-1)\left[-\left(\frac{1-Y_{PAO}}{2.86 \cdot Y_{PAO}}\right)\right]} = \frac{r_{XBPAO,anx}}{1}. \tag{3.91}$$

Similarly, the rate of Poly-P storage under anoxic conditions ($r_{XPP,anx}$) is less than the rate under aerobic conditions. The rate expression includes all of the parenthetical terms in Equation 3.87, except that the DO term becomes an inhibition term and $\eta_{g,PAO}$ is used to account for the reduced rate:

$$r_{XPP,anx} = \hat{q}_{PP} \cdot \eta_{g,PAO} \left[\frac{X_{PHA}/X_{B,PAO}}{K_{PHA} + (X_{PHA}/X_{B,PAO})} \right] \left(\frac{S_{NO}}{K_{NO} + S_{NO}} \right) \left(\frac{S_{PO_4}}{K_{PO_4} + S_{PO_4}} \right)$$
$$\cdot \left(\frac{K_O}{K_O + S_O} \right) \left[\frac{K_{PMAX} - (X_{PP}/X_{B,PAO})}{K_{IPP} + K_{PMAX} - (X_{PP}/X_{B,PAO})} \right] X_{B,PAO}. \quad (3.92)$$

All terms were defined following Equations 3.85, 3.87, or 3.90. The stoichiometry of the anoxic reactions is similar to that of the aerobic reactions, as shown in Equation 3.88, except for the need to include the COD equivalent of nitrate. Thus, the relationship between the rates of PHA loss ($r_{XPHA,anx}$), nitrate utilization (r_{SNO}), soluble phosphate removal ($r_{SPO_4,anx}$), and Poly-P formation ($r_{XPP,anx}$) under anoxic conditions is given by

$$\frac{r_{XPHA,anx}}{(-Y_{PHA})} = \frac{r_{SNO}}{(-1)(-2.86 \cdot Y_{PHA})} = \frac{r_{SPO_4,anx}}{-1} = \frac{r_{XPP,anx}}{1}. \quad (3.93)$$

As for the aerobic condition, the total rates of PHA loss and nitrate consumption under anoxic conditions can be determined by adding the expressions associated with PAO growth and Poly-P storage, as given by Equations 3.91 and 3.93, respectively.

3.8 SIMPLIFIED STOICHIOMETRY AND ITS USE

In Chapters 5 and 6 we will use the concepts developed in Section 3.1.3 to construct mathematical models that incorporate the various events discussed in this chapter. There are many circumstances, however, in which the use of stoichiometric concepts would be very useful even without the development of rigorous equations. For example, an examination of Equation 3.13 expressing biomass growth and Equation 3.52 expressing biomass decay by the traditional approach reveals that they could be combined into a single equation that incorporates both reactions. Since biomass is a product in Equation 3.13 and a reactant in Equation 3.52, the effect would be to reduce the net amount of biomass formed. Likewise, since nutrients are reactants in Equation 3.13 and products in Equation 3.52, the net amount of nutrients used would also be reduced. The electron acceptor, on the other hand, is a reactant in both equations, so the effect of combining them would be to increase the amount of electron acceptor required. Consideration of what is occurring when the equations are combined, in combination with a recollection of the discussion of yield in Section 2.4.1, reveals that the net stoichiometric coefficient on biomass in a mass-based combined equation is the observed yield. In other words, it is the yield when maintenance energy needs and decay are taken into account. By making use of the fact that the observed yield is a function of the growth conditions imposed on the biomass, the combined equation may be used to show how the nutrient and electron acceptor requirements change as the growth conditions are changed.[130]

3.8.1 Determination of the Quantity of Terminal Electron Acceptor Needed

Although other stoichiometric equations can be used, the COD-based equation is the most useful for determining the quantity of terminal electron acceptor required for the growth of heterotrophs. Writing the combined stoichiometric equation for heterotrophic biomass growth in COD units

illustrates a very important point that will be used throughout this book. When ammonia serves as the nitrogen source, the sum of the oxygen (or oxygen equivalents of nitrate) used and the biomass (active plus debris) formed (in COD units) must equal the COD removed from solution. This follows from the fact that COD is a measure of available electrons. In other words, all of the electrons available in a substrate being biodegraded are either removed and transferred to the terminal electron acceptor or they are incorporated into the biomass formed. As discussed in Section 3.2.1, when ammonia serves as the nitrogen source, no electrons are transferred to nitrogen during biomass synthesis. When nitrate serves as the nitrogen source, however, some of those electrons must be used to reduce nitrogen from the +V state to the –III state, and thus those electrons are incorporated into the biomass even though they will not be measured in the COD test. This is because nitrogen does not accept or give up electrons in the COD test. Consequently, if biomass is represented by $C_5H_7O_2N$, its COD must be multiplied by 1.4 for the balance to work, as suggested by Equation 3.17. Thus, to generalize:

$$\text{COD removed} = O_2 \text{ equivalents of terminal electron acceptor used} + \alpha_N \text{ (COD of biomass formed)}, \tag{3.94}$$

where

$$\alpha_N = 1.0 \text{ } NH_4^+ \text{ as nitrogen source},$$

$$\alpha_N = 1.4 \text{ } NO_3^- \text{ as nitrogen source}.$$

Equation 3.94 is generally applicable and is much easier to use for determining the amount of terminal electron acceptor required than the writing of a molar- or mass-based stoichiometric equation. Thus, it is widely employed and will be used frequently herein. Ammonia will be assumed to be the nitrogen source throughout this book, unless specifically stated otherwise. Thus, α will generally be set equal to 1.0.

3.8.2 Determination of Quantity of Nutrient Needed

The amount of nitrogen required for heterotrophic biomass growth can also be calculated from the combined stoichiometric equation. Since the only use of nitrogen in the equation is for synthesis of biomass, the equation may be used to establish a relationship that is very useful for estimating nutrient requirements. If the ammonium ion requirement is expressed per unit of biomass COD formed, it is found to be 0.112 mg of NH_4^+ per mg of biomass COD formed. Or, expressed as the amount of nitrogen required, it is 0.087 mg of N per mg of biomass COD formed. Actually, this can be considered to be a generality that is independent of the source of the nitrogen, provided that $C_5H_7O_2N$ represents the composition of biomass. This suggests that once the observed yield has been determined, the amount of nitrogen required can be estimated easily, allowing adequate amounts to be provided if they are not naturally present in the wastewater. Likewise, each time a mg of biomass COD is destroyed, 0.087 mg of ammonia-N will be released to the medium, and this fact must be considered in operations such as aerobic digestion that are designed to destroy biomass. As with the determination of the electron acceptor requirement, the main purpose of traditional stoichiometric equations has been to provide simplified relationships such as these for engineering use. Thus, the conversion factor is generally used in lieu of writing a new balanced stoichiometric equation for each situation.

As mentioned in Section 3.2.1, the phosphorus requirement for normal microbial growth can be estimated as one-fifth of the nitrogen requirement on a mass basis. Consequently, about 0.017 mg of phosphorus will be required for each mg of heterotrophic or autotrophic biomass COD formed, and an equal amount will be released for each mg destroyed. If PAOs are in the system, the amount released by destruction of the biomass will be different and will depend on the amount of Poly-P stored.

TABLE 3.3
Approximate Micronutrient Requirements for Bacterial Growth

Micronutrient	Approximate Requirement[a] µg/mg Biomass COD Formed	Approximate Requirement[a] µg/mg Biomass TSS Formed
Potassium	10	12
Calcium	10	12
Magnesium	7	8.4
Sulfur	6	7.2
Sodium	3	3.6
Chloride	3	3.6
Iron	2	2.4
Zinc	0.2	0.24
Manganese	0.1	0.12
Copper	0.02	0.024
Molybdenum	0.004	0.005
Cobalt	<0.0004	<0.0005

[a] Estimates based on the judgment of the authors after considering information from Eckenfelder, W. W., Jr., and Musterman, J. L., Activated sludge treatment of industrial waters. *Activated Sludge Process Design and Control: Theory and Practice*, 127–266, eds. W. W. Eckenfelder and P. Grau, Technomic Publishing Co., Inc., Lancaster, PA, 1992; Ribbons, D. W., Quantitative relationships between growth media constituents and cellular yields and composition. *Methods in Microbiology*, Vol. 3A, 297–304, eds. J. R. Norris and D. W. Ribbons, Academic Press, NY, 1970; Wood, D. K., and Tchobanoglous, G., Trace elements in biological waste treatment. *Journal, Water Pollution Control Federation*, 47:1933–45, 1975.

The provision of sufficient nutrients is essential if efficient wastewater treatment is to be achieved, because without them the microorganisms will not be able to perform their synthesis reactions. Although nitrogen and phosphorus are the nutrients needed in greatest quantity (macronutrients), many other elements are required by the microorganisms but are not normally included in the stoichiometric equation because of the complicating effect they would have. The need for them should not be ignored nor should their presence be taken for granted because severe problems can result if sufficient quantities are not available.[18,153] Table 3.3 lists the major micronutrients required for bacterial growth.[107] There is little agreement in the literature concerning their quantities in biomass. One reason is that different bacteria have different requirements. Another is that bacteria tend to adsorb cations, thereby making it difficult to determine exactly the quantity actually incorporated into biomass. The values listed in Table 3.3 are the authors' best estimates of the quantities required based on examination of several sources.[38,119,153]

3.9 EFFECTS OF TEMPERATURE

Temperature can exert an effect on biological reactions in two ways: by influencing the rates of enzymatically catalyzed reactions and by affecting the rate of diffusion of substrate to the cells. The importance of both has not always been recognized and this has led to some confusion in the quantification of temperature effects. For example, temperature effects observed in the laboratory are often more pronounced than those observed in the field. This is due in part to the fact that full-scale reactors are apt to be diffusion controlled. Consequently, the temperature coefficients given below are provided simply to give an idea of the importance of temperature to various microbial processes. For system design, actual temperature effects should always be measured in prototype systems that simulate the anticipated mixing regime.

3.9.1 Methods of Expressing Temperature Effects

There are three techniques in common use to quantify the effects of temperature on biochemical operations. The oldest is that of Arrhenius,[8] who first applied it in 1889 to quantify the effects of temperature on the enzymatic hydrolysis of sugar. It is

$$k = A \cdot e^{-u/RT}, \tag{3.95}$$

where k is the temperature dependent rate coefficient, A is a constant, u is the temperature coefficient, R is the gas constant, and T is the absolute temperature. The value of u may be obtained by plotting ln k versus 1/T and determining the slope. For normal SI units, the units of u are kJ/mole and a positive value means that k increases as the temperature is increased.

Although microorganisms have been found in extreme environments that can grow at temperatures approaching either the freezing point or the boiling point of water, most microorganisms exhibit a relatively narrow temperature range over which they can function. Within that range, most reaction rate coefficients increase as the temperature is increased, but then eventually decrease as the heat begins to inactivate cellular enzymes and denature other critical proteins and cellular structures in cells.[89] Figure 3.7 schematically shows the typical relationship between specific growth rate and temperature, and emphasizes the fact that specific growth rate increases with temperature gradually to an optimum value, then drops rapidly once the temperature increases past the optimum temperature. The Arrhenius equation and its variants are only applicable over the range where the coefficient increases with increasing temperature. Microorganisms are grouped into three categories depending on the temperature range over which they grow. Of chief concern in biochemical operations are mesophilic organisms, which grow well over the range of 10–35°C. The two other groups, psychrophilic and thermophilic, have ranges on either side and find use under special conditions. Unless otherwise specified, all parameter values given in this book will be for mesophilic microorganisms.

If a rate coefficient is known at one temperature, it may be calculated at another through rearrangement of the Arrhenius equation:

$$\ln (k_1/k_2) = \frac{u(T_1 - T_2)}{(R \cdot T_1 \cdot T_2)}. \tag{3.96}$$

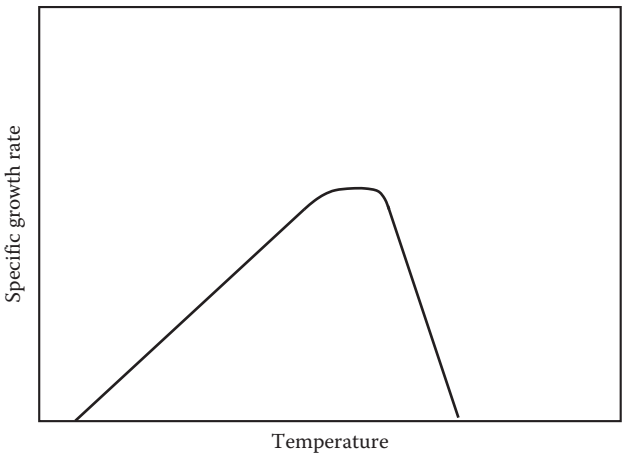

FIGURE 3.7 Typical plot of the effect of temperature on bacterial specific growth rate.

Because the mesophilic temperature range is small when T is expressed in K, the term $(R \cdot T_1 \cdot T_2)$ does not vary appreciably and may be considered to be constant. Consequently, a more commonly used expression is[108]

$$k_1 = k_2 \cdot e^{C(T_1 - T_2)}, \tag{3.97}$$

where

$$C = \frac{u}{(R \cdot T_1 \cdot T_2)} \approx 0.0015\ u, \tag{3.98}$$

for the normal mesophilic temperature range. Note that when Equation 3.97 is used, the temperature may be expressed in °C because only the temperature difference enters into the equation. In that case the units of C are $°C^{-1}$. The value of C may be determined by plotting ln(k) versus T, giving a slope equal to C.

Finally, a third equation has found considerable use in the environmental engineering literature:[115]

$$k_1 = k_2 \cdot \theta^{(T_1 - T_2)}. \tag{3.99}$$

Actually, Equations 3.97 and 3.99 are the same since

$$C = \ln(\theta). \tag{3.100}$$

Thus, Equation 3.99 is also a variant of the Arrhenius equation and the coefficient θ may also be estimated by plotting ln(k) versus T, giving a slope equal to ln(θ). The θ is dimensionless.

The temperature coefficients for the three variants of the Arrhenius equation may be interconverted by

$$\ln(\theta) = C \approx 0.0015\ u, \tag{3.101}$$

in which the temperature is expressed in °C or K.

As stated above, a limitation on Equations 3.95, 3.97, and 3.99 is that they only characterize the temperature range over which the temperature dependent coefficient rises with increases in temperature. This limitation can be easily accommodated during design by limiting their utility to the appropriate temperature range. During mathematical modeling, however, particularly of industrial wastewater treatment processes in which the temperature may exhibit broad ranges, it would be advantageous to express both the positive and negative effects of temperature as a continuous function. This can be done with the expression of Ratkowsky et al.:[118]

$$k = \left\{b \cdot (T - T_{min}) \cdot \left[1 - e^{(c \cdot (T - T_{max}))}\right]\right\}^2, \tag{3.102}$$

in which b, c, T_{min}, and T_{max} are empirically derived coefficients. This expression has been applied successfully to suspended growth wastewater treatment systems.[70,71]

3.9.2 Effects of Temperature on Kinetic Parameters

3.9.2.1 Biomass Growth and Substrate Utilization

It will be recalled from Equations 3.35 and 3.43 that biomass growth and substrate utilization are proportional to each other, with the yield being the proportionality coefficient. It will also be recalled from Figure 2.4 that temperature can influence the value of the yield. This suggests that temperature

can influence growth and substrate utilization in quantitatively different ways. Nevertheless, because of the uncertainty associated with the impact of temperature on Y, most engineers assume it to be independent of temperature, thereby allowing the same temperature coefficient to be used for both growth and substrate utilization.

Two parameters are required to characterize biomass growth, $\hat{\mu}$ and K_S. The first is clearly a rate coefficient, and as such, its value increases with increasing temperature up to the optimum temperature. The second describes how substrate concentration influences the specific growth rate, and thus the impact of temperature on it is less clear, with it increasing under some circumstances and decreasing under others. Consequently, there is no consensus about its relationship to temperature, and each situation must be experimentally determined.

Most studies of the impact of temperature have been done on the aerobic growth of heterotrophs. Two studies[22,105] have reviewed the literature, and have reported values of u for $\hat{\mu}$ ranging from 21.3 to 167.4 kJ/mole. The average value for the larger data base[22] (18 values) was 59.8 kJ/mole, which converts to C and θ values of 0.090 $°C^{-1}$ and 1.094, respectively. Very few studies reporting the effects of temperature on K_S were cited, and there was no consensus among them as to whether it increased or decreased with increasing temperature.

Very few studies have been done to quantify the effects of temperature on microbial growth under anoxic conditions. van Haandel et al.[145] recommend that a θ value of 1.20 (C = 0.182 $°C^{-1}$, u = 121 kJ/mole) be used for $\hat{q}$. This value is near the upper range for the aerobic values reported above, which suggests that it may be high. Until more data are available, it may be prudent to adopt a value more consistent with aerobic growth and substrate utilization since the two processes are mechanistically similar. No values have been reported for the effect on K_S under anoxic conditions.

Temperature is a critical consideration for nitrifying bacteria because their $\hat{\mu}$ values are low even under the best of circumstances. Characklis and Gujer[22] reported four temperature coefficients for $\hat{\mu}$ for nitrification, with an average u value of 71.8 kJ/mole (C = 0.108 $°C^{-1}$, θ = 1.114). However, there appears to be little consensus about the relative effects of temperature on the two major types of nitrifiers. For example, Characklis and Gujer[22] reported an average u of 74.3 kJ/mole (C = 0.111 $°C^{-1}$, θ = 1.118) for $\hat{\mu}$ of AOB and 44.0 kJ/mole (C = 0.066 $°C^{-1}$, θ = 1.068) for NOB. In contrast, Hall and Murphy[59] reported a u value of 62.4 kJ/mole (C = 0.094 $°C^{-1}$, θ = 1.098) for $\hat{q}$ for AOB and 71.1 kJ/mole (C = 0.107 $°C^{-1}$, θ = 1.112) for NOB. Nevertheless, there still seems to be a general consensus that the temperature coefficient for NOB is smaller than it is for AOB. In contrast to heterotrophs, for which temperature appears to have variable effects on K_S, increases in temperature cause the half-saturation coefficient for nitrifiers to increase. The most widely cited data is that of Knowles et al.,[81] for which u associated with the K_S for AOB was 78.7 kJ/mole (C = 0.118 $°C^{-1}$, θ = 1.125) and u associated with the K_S for NOB was 97.3 kJ/mole (C = 0.146 $°C^{-1}$, θ = 1.157).

3.9.2.2 Maintenance, Endogenous Metabolism, Decay, Lysis, and Death

Most studies have used the traditional decay concept to quantify the impacts of maintenance on microbial systems, and thus all temperature data are available in terms of the rate coefficient b (Equation 3.56). However, because b and b_L are proportional to each other (Equation 3.67), the resulting temperature coefficients should also be applicable to b_L.

Because the factors contributing to decay of heterotrophs are the same as those contributing to growth, it is logical to expect temperature to have similar effects on b and $\hat{\mu}$ and that has been observed, with data from three studies giving u values for b equal to 1.1 times the u values for $\hat{\mu}$ for a given culture.[105] Thus, from the effects of temperature on $\hat{\mu}$ reported earlier, a typical u value for b might be expected to be 65.8 kJ/mole (C = 0.120 $°C^{-1}$, θ = 1.104). Others,[35] however, have used much smaller values for the effects of temperature on decay, with a u value of 19.1 kJ/mole (C = 0.029 $°C^{-1}$, θ = 1.029).

In spite of the importance of temperature to nitrification, few studies have systematically studied the effects of temperature on the decay coefficient for nitrifying bacteria. Based on laboratory studies at two temperatures, Melcer et al.[101] determined the value of θ for the autotrophic decay coefficient (b_A) to be 1.029 (C = 0.029 $°C^{-1}$, u = 19.1 kJ/mole).

3.9.2.3 Solubilization of Particulate and High Molecular Weight Soluble Organic Matter

As might be anticipated from the discussion in Section 3.5, relatively little work has been done on the effects of temperature on the hydrolysis of particulate substrate. However, because it is an enzymatic step, the hydrolysis coefficient, k_h, is likely to rise as the temperature is increased. From a comparison of experimental data to simulation results from a complex system model, van Haandel et al.[145] concluded that a u value of 38.8 kJ/mole (C = 0.058 °C^{-1}, θ = 1.060) was appropriate for both aerobic and anoxic environments. No information was given for the effect of temperature on K_X, the half-saturation coefficient for hydrolysis.

3.9.2.4 Phosphorus Uptake and Release

Temperature has been shown to be an important parameter in biological phosphorus removal processes, although the nature of its impact is complicated by changes in the competition between PAOs and GAOs at different temperatures. Garcia-Usach et al.[43] determined θ values for several PAO-related parameters. Estimates of θ for $\hat{q}_{PHA}$, $\hat{q}_{PP}$ and $\hat{\mu}_{PAO}$ were 1.035 (C = 0.034 °C^{-1}, u = 22.7 kJ/mole), 1.053 (C = 0.052 °C^{-1}, u = 34.7 kJ/mole), and 1.058 (C = 0.056 °C^{-1}, u = 37.6 kJ/mole), respectively. Whang and Park[151] estimated C to be 0.042 °C^{-1} (θ = 1.043, u = 28.1 kJ/mole) for the specific rate of acetate uptake under anaerobic conditions, which is directly proportional to $\hat{q}_{PHA}$, suggesting that those temperature coefficients also apply to $\hat{q}_{PHA}$.

3.9.2.5 Other Important Microbial Processes

Insufficient data are available to allow quantification of the effects of temperature on other processes, such as phosphorus release, but it is likely that appropriate temperature coefficients will be developed for them in the future.

3.10 KEY POINTS

1. Stoichiometric equations may be written on a mass basis rather than a molar basis. When this is done, the total mass of reactants equals the total mass of products. When a stoichiometric equation is written on a mass of COD basis, only constituents containing elements that change oxidation state are included. The COD of the reactants must equal the COD of the products.
2. When nitrate serves as the terminal electron acceptor, nitrogen changes oxidation state from +V to 0. Consequently, the oxygen equivalence of nitrate is –2.86 mg COD/mg N (2.86 mg O_2/mg N). When nitrate serves as the nitrogen source for biomass growth, the nitrogen is reduced to the amino level (i.e., from the +V to the –III state). In that case, the oxygen equivalence is –4.57 mg COD/mg N (4.57 mg O_2/mg N).
3. If the general form of the mass-based stoichiometric equation is written as

$$(-1)A_1 + (-\Psi_2)A_2 + \cdots + (-\Psi_k)A_k + \Psi_{k+1}A_{k+1} + \cdots + \Psi_m A_m = 0,$$

then, r, the generalized reaction rate is given by

$$\frac{r_1}{(-1)} = \frac{r_2}{(-\Psi_2)} = \frac{r_k}{(-\Psi_k)} = \frac{r_{k+1}}{(\Psi_{k+1})} = \frac{r_m}{(\Psi_m)} = r.$$

Furthermore, if there are j reactions (where $j = 1 \rightarrow n$) involving i components (where $i = 1 \rightarrow m$), the overall reaction rate for component i will be given by

$$r_i = \sum_{j=1}^{n} \Psi_{i,j} \cdot r_j.$$

If r_i is negative, the component i is being consumed, whereas if it is positive, the component is being produced.

4. Knowledge of the yield is required before the stoichiometric equation for microbial growth can be written. If McCarty's half-reaction approach is used to write the stoichiometric equation, f_s, the fraction of the electron donor captured through synthesis, is directly related to the yield expressed as mass of biomass COD formed per unit mass of substrate COD used.
5. When the electron donor is an organic compound, ammonia serves as the nitrogen source, and the yield is expressed as biomass COD formed per unit of substrate COD used, f_s and Y are equal. For other circumstances, f_s may be either greater than or smaller than Y.
6. Bacteria divide by binary fission. Thus, their rate of growth is first order with respect to the concentration of active biomass present:

$$r_{XB} = \mu \cdot X_B.$$

The rate coefficient, μ, is called the specific growth rate coefficient. It is influenced by the substrate concentration. If the substrate is noninhibitory, the most commonly used expression is that of Monod:

$$\mu = \hat{\mu}\frac{S_S}{K_S + S_S}.$$

If the substrate is inhibitory to its own biodegradation, the Andrews equation is commonly used:

$$\mu = \hat{\mu}\frac{S_S}{K_S + S_S + S_s^2/K_I}.$$

7. When the substrate concentration is large relative to K_S, the Monod equation may be simplified to an expression that is zero order with respect to the substrate concentration. When the substrate concentration is small relative to K_S, the specific growth rate coefficient is approximately first order with respect to the substrate concentration.
8. Complementary nutrients are those that meet different needs by growing microorganisms whereas substitutable nutrients are those that meet the same need. The effects of limitation by two complementary nutrients may be depicted by interactive and noninteractive models. The interactive approach is more appropriate for modeling wastewater treatment systems.
9. Biochemical operations can be designed most easily when the nutrient the system is being designed to control acts as the growth limiting nutrient for the biomass in the system.
10. The kinetic parameters in the Monod and Andrews equations depend strongly on the species of microorganism and the substrate upon which the microorganisms are growing. Since wastewater treatment operations use mixed cultures, and wastewaters contain many compounds, the parameters used to describe such operations should be characterized by ranges rather than by single values.
11. Nitrifying bacteria have lower maximum specific growth rate coefficients than heterotrophic bacteria, and are more sensitive to pH and to low dissolved oxygen concentrations.
12. Three approaches have been used to model the loss of viability and biomass in biochemical operations: the traditional decay approach, the lysis:regrowth approach, and the endogenous respiration/storage approach. In the traditional approach, loss of active biomass leads directly to the use of an electron acceptor and the production of biomass debris, which

accumulates and acts to reduce the viability. In the lysis:regrowth approach, active biomass is lost by lysis, which releases particulate substrate and biomass debris. Electron acceptor consumption occurs only after soluble substrate, which is formed by hydrolysis of the particulate substrate, is used for new biomass growth. Because the yield is always less than one, the amount of new biomass formed is always less than the biomass lost by lysis, leading to a loss of biomass in the bioreactor. The endogenous respiration approach assumes that substrate is not used directly to support growth but must, first, be incorporated into intracellular storage materials before it is used to support growth. This stored material is also used for maintenance purposes, leading to a loss of mass. Rates change with the electron acceptor condition.

13. Both the traditional and the lysis:regrowth approaches to modeling decay and loss of viability depict the rate of active biomass loss as being first order with respect to the active biomass concentration, as is the generation of biomass debris. However, the decay coefficient in the traditional approach is smaller than the coefficient in the lysis:regrowth approach, although the fraction of the biomass leading to debris is larger.
14. Two types of soluble microbial products are formed by microbial cultures, growth associated and biomass associated. The rate of growth associated product formation is directly proportional to the rate at which biomass is growing and utilizing substrate. There is not yet consensus concerning the rate of biomass-associated product formation.
15. For modeling purposes, solubilization of particulate and high molecular weight organic matter is assumed to occur by hydrolysis, with conservation of COD. The rate expression adopted to describe hydrolysis is similar to the Monod equation, except that it is controlled by the particulate substrate to biomass ratio rather than by the particulate substrate concentration:

$$r_{XS} = -k_h \left[\frac{X_S/X_{B,H}}{K_X + (X_S/X_{B,H})} \right] X_{B,H}.$$

This is necessary because the reaction is thought to be surface mediated.

16. Even though organic nitrogen may be used directly in biomass synthesis, it is simpler to model the flow of nitrogen in biochemical operations by assuming that nitrogen is released to the medium as ammonia and then taken up for biomass synthesis as needed. The release as ammonia, called ammonification, is assumed to be first order with respect to both the biomass and soluble, biodegradable organic nitrogen concentrations. The uptake of ammonia for growth is assumed to be proportional to the rate of growth.
17. During biological phosphorus removal, the uptake of acetate, the formation of PHA, and the release of soluble phosphate by PAOs under anaerobic conditions are all coupled. The rate of PHA formation is controlled by both the acetate concentration in solution, S_A, and the polyphosphate concentration in the biomass, X_{PP}:

$$r_{XPHA} = \hat{q}_{PHA} \left(\frac{S_A}{K_A + S_A} \right) \left[\frac{X_{PP}/X_{B,PAO}}{K_{PP} + (X_{PP}/X_{B,PAO})} \right] X_{B,PAO}.$$

Under aerobic conditions, the PAOs grow by using the stored PHA as a carbon and energy source, storing polyphosphate in the process:

$$r_{XBPAO} = \hat{\mu}_{PAO} \left[\frac{X_{PHA}/X_{B,PAO}}{K_{PHA} + (X_{PHA}/X_{B,PAO})} \right] \left(\frac{S_{PO4}}{K_{PO4} + S_{PO4}} \right) \left(\frac{S_O}{K_O + S_O} \right) X_{B,PAO}.$$

The rate of phosphorus storage is coupled to the rate of biomass growth and thus is expressed by a similar equation. An additional term is required, however, to reflect the fact that there is a limit to the amount of polyphosphate that the PAOs can accumulate. Phosphate accumulating organisms can also grow and store polyphosphate under denitrifying conditions, but at a rate that is a fraction of aerobic rates.

18. The COD-based stoichiometric equation states that the COD removed by a biological reaction must equal the oxygen equivalents of the terminal electron acceptor used plus α_N times the COD of the biomass formed. The value of α_N depends on the nature of the nitrogen source. It is 1.0 when ammonia is the source and 1.4 when nitrate is the source.
19. If $C_5H_7O_2N$ can be considered to be representative of the elemental composition of biomass, then 0.087 mg of nitrogen is required to synthesize the mg of biomass COD. Conversely, each time the mg of biomass COD is destroyed by decay, 0.087 mg of nitrogen is released. Although not shown in the empirical equation for biomass, approximately 0.017 mg of phosphorus will be required (released) each time the mg of biomass COD is formed (destroyed).
20. Within a relatively narrow physiological range, the maximum specific growth rate coefficient, $\hat{\mu}$, increases as the temperature is increased, but then decreases for further temperature increases. Furthermore, for rapidly growing cultures the effect of temperature on the traditional decay coefficient, b, appears to be closely correlated with the effect on $\hat{\mu}$. No conclusions can be drawn about the effects of temperature on the half-saturation coefficient, K_S.
21. Three expressions are commonly used to relate the rate coefficients in biological operations (k) at different temperatures (T):

$$k = A \cdot e^{-u/RT}$$

$$k_1 = k_2 \cdot e^{C(T_1 - T_2)}$$

$$k_1 = k_2 \cdot \theta^{(T_1 - T_2)}.$$

The temperature coefficients for the three equations may be interconverted by

$$\ln(\theta) = C \approx 0.0015u.$$

3.11 STUDY QUESTIONS

1. Prove the COD mass equivalents given in Table 3.1 using the following methods:
   a. Half reactions given in Table 3.2.
   b. Balanced stoichiometric reactions that yield CO_2, H_2O, and NH_3, as appropriate.
   c. The change in redox state and the theoretical electron equivalency for O_2.
2. Why must the yield be known before the stoichiometric equation for microbial growth can be written? Which type of yield, the true growth yield or the observed yield, is most appropriate for doing this? Why? How is knowledge of the yield used to write the stoichiometric equation using McCarty's half-reaction approach?
3. Using the half-reaction technique, write the molar stoichiometric equation for microbial growth for each of the following situations:
   a. Aerobic growth on domestic wastewater with ammonia nitrogen as the nitrogen source. The yield is 0.60 mg biomass COD formed/mg substrate COD removed.
   b. Growth on a carbohydrate with nitrate as the terminal electron acceptor and ammonia as the nitrogen source. The yield is 0.50 mg biomass COD formed/mg substrate COD used.

   c. Growth on a carbohydrate with nitrate as the terminal electron acceptor and nitrogen source. The yield is 0.40 mg biomass COD formed/mg substrate COD used.
   d. Normalize them with respect to the electron donor.
4. Convert the molar stoichiometric equation from Study Question 3a into a mass-based equation with the electron donor as the reference component.
5. Convert the molar stoichiometric equation from Study Question 3a into a COD-based equation with the electron donor as the reference component.
6. Write the rate expression for bacterial growth and relate it to the rates of substrate and oxygen utilization for heterotrophic biomass growth on an organic substrate. Then state the Monod and Andrews equations relating the specific growth rate coefficient to the substrate concentration. Finally, draw sketches depicting the effects represented by both equations and use them to define the parameters in the equations.
7. State the zero- and first-order approximations of the Monod equation. Under what circumstances may they be used?
8. Explain the difference between complementary and substitutable nutrients. Then differentiate between interactive and noninteractive models for describing the effects of two complementary nutrients. Finally, state why the interactive approach was adopted herein.
9. Draw a sketch depicting the effects of two interactive, complementary nutrients on the specific growth rate of biomass and use it to explain why it is easier to design a biochemical operation to achieve a desired concentration of a given nutrient if that nutrient serves as the sole growth limiting nutrient for the biomass.
10. Even though it is best to characterize the kinetic parameters in the Monod and Andrews equations by ranges rather than by unique values, it is possible to state several generalities about the sizes of those parameters. Use such generalities to contrast and compare the growth characteristics of heterotrophic and autotrophic biomass.
11. Discuss the effects that organic compounds and heterotrophic biomass can have on the growth of nitrifying bacteria.
12. Describe the major groups of microorganisms participating in anaerobic operations and contrast their growth characteristics as described by their kinetic parameters.
13. Describe in detail the traditional and lysis:regrowth approaches to modeling the loss of biomass and viability observed in biochemical operations. In your description, contrast the routes of carbon, nitrogen, and electron flow, and explain how they influence the magnitudes of the kinetic parameters used to characterize the events.
14. Write the rate equations for loss of active biomass as depicted by the traditional and lysis:regrowth approaches. Then explain the relationships between the kinetic and stoichiometric parameters used in the two approaches.
15. Write the rate equation for the hydrolysis of particulate substrate, compare it to the Monod equation, and explain any differences.
16. Discuss the fate of nitrogen in biochemical operations and state the rate equations used to model that fate.
17. State the rate equations that have been proposed to represent acetate uptake, PHA formation, and phosphorus release by PAOs under anaerobic conditions in a biological phosphorus removal system. Then state the rate equations depicting PAO growth, soluble phosphorus uptake, and polyphosphate formation under aerobic conditions. Use those equations in a discussion of the events occurring in such systems and explain why the various terms were included in the rate expressions.
18. An aerobic culture is growing on a mixture of organic matter, such as that found in domestic wastewater, with ammonia as the nitrogen source. How many mg of nitrogen (N), phosphorus (P), and oxygen (O_2) must be provided per mg of COD removed for each of the following situations, where Y_{obs} is the observed yield? What quantities of micronutrients

will be required? Do not derive the stoichiometric equations. Rather, answer the question using generalizations presented in the text.

a. $Y_{obs} = 0.70$ mg biomass COD formed/mg substrate COD removed.
b. $Y_{obs} = 0.57$ mg biomass COD formed/mg substrate COD removed.
c. $Y_{obs} = 0.36$ mg biomass COD formed/mg substrate COD removed.

19. Demonstrate why the value of α_N in Equation 3.94 is 1.40 when nitrate serves as the nitrogen source and biomass is represented by $C_5H_7O_2N$.
20. Three techniques are often used to describe the effects of temperature on microbial cultures. Describe each of them and tell how you would plot data to determine the values of the temperature coefficients in the equations.
21. The data below describe the effects of temperature on the traditional decay coefficient, b. Use that data to determine the temperature coefficient by each of the three techniques. Use 20°C as the reference temperature. Discuss the utility of each technique for describing the effects of temperature on this parameter.

T°C	b = hr^{-1}
10	0.0037
20	0.0095
30	0.0229
40	0.0372

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Part II

Theory: Modeling of Ideal Suspended Growth Reactors

The primary function of a mathematical model is to reduce a complex system to the minimum terms essential for its description so that those terms may be manipulated, thereby helping us to understand how the system will respond under a variety of conditions. Generally, mathematical models do not describe a system completely, but if the terms are chosen with care, the model response will be qualitatively similar to the real system. In Part I we considered in detail the major events occurring in biochemical operations. Now the mathematical descriptions of those events will be incorporated into mass balance equations for the major reacting components in order to develop mathematical models describing a number of reactor configurations representing suspended growth systems. Chapter 4 presents the techniques for describing both ideal and nonideal reactors in mathematical terms. Chapter 5 establishes several fundamental principles governing the performance of suspended growth biochemical operations by considering the situation of aerobic heterotrophic microbial growth on a soluble organic substrate in a single ideal reactor. Chapter 6 extends the concepts of Chapter 5 by adding additional reactions, such as autotrophic growth of nitrifying bacteria and anoxic growth of heterotrophic bacteria, in a single ideal reactor. In Chapter 7, other aerobic and anoxic reactor configurations are included to demonstrate how the engineer can control system performance through selection of the appropriate reaction environment. Chapter 8 introduces anaerobic acidogenic and methanogenic bioreactors and the complex interactions that occur in the microbial communities within them. Finally, Chapter 9 describes techniques whereby the kinetic and stoichiometric parameters used in the models may be evaluated. In investigating reactor performance through modeling, it will be assumed that the reactors are ideal, with respect to both fluid flow and the response of the microbial culture. In other words, we will investigate how the reactors would respond if the mathematical models described them exactly. In Part III, any significant deviations from ideality are discussed and incorporated into the application of the models to design.

4 Modeling Suspended Growth Systems

In Chapter 1 we saw that many types of biochemical operations are used in wastewater treatment. Understanding how each performs would be a difficult task if we had to approach each as a unique entity. Fortunately, there is a great deal of commonality among them because only a finite number of events occur within them, as we saw in Chapter 2. This suggests that the major differences among them arise from their reactor configurations. Because the kinetics of the events within biochemical operations are reasonably well established, reactor engineering principles can be applied to see how reactor configuration influences their outcome, thereby setting the stage for understanding how the various types of biochemical operations perform.

Reactor engineering is the application of mathematical modeling to the analysis and design of chemical and biochemical reactors. In this chapter we will review briefly some of the basic concepts of reactor engineering. For more complete coverage, the reader should consult texts on the subject.[6,8]

4.1 MODELING MICROBIAL SYSTEMS

Microbial systems are extremely complex and the models describing them can be very complicated. Luckily, relatively simple models have proven to be satisfactory for describing the performance of many biochemical operations and we will focus on them because our intent is to gain an appreciation for the basic manner in which the systems function. For that purpose, additional complexity is more likely to confuse than to clarify. All of the models that we will consider are transport phenomena models in that they are based on the conservation of mass, momentum, and energy.[2] Furthermore, they are primarily phenomenological because the rate expressions in them, which were presented in Chapter 3, seek to depict in simple terms the basic mechanisms involved. At the same time, the models are also empirical because the ultimate justification for use of those rate expressions is observation and experience rather than derivation from first principles.

Many simplifying assumptions have been made by the developers of the rate expressions and models used in this text. Although many are implicit, it is important that the more common ones be stated explicitly.[2,7] The first assumption is that all of the organisms of a given species are the same within a given reactor in a system. In reality, individual microorganisms will differ in their physiological states because they will be at different points in their life cycles. However, little is known about the impact of those states and the mathematics of considering them would be complex. Thus, they will not be considered. The second assumption is that stochastic phenomena can be neglected; that is, that any random differences among the cells of a given type can be ignored. This assumption causes few problems because the enormous number of cells present in most biochemical operations causes the random deviations to be canceled out. The third assumption is that within a functional class (e.g., aerobic heterotroph, autotroph, etc.), all microorganisms are treated as if they were the same species. While this is never true in wastewater treatment systems, the problems associated with modeling dozens (if not hundreds) of distinct species are obvious. However, distinctions will be made between the important groups discussed in Chapters 2 and 3. The fourth assumption is an extension of the third in that within a given class of microorganism, individuals are ignored. That is, our focus will be on the mass of organisms present, rather than on the individual members forming that mass.

This assumption works reasonably well as long as growth is balanced because then changes in mass and numbers are proportional. Furthermore, the organisms will be considered to be distributed evenly throughout the culture within a given vessel. Finally, even though the microorganisms form a distinct phase within the reactor, the reactions within suspended growth reactors will be treated as if they were homogeneous. This assumption allows the transport of reactants from the liquid phase to the solid (biomass) phase to be ignored, thereby greatly simplifying the modeling task. Although transport is indeed important, particularly in flocculent systems like activated sludge,[1] imposition of this assumption causes no great difficulties as long as it is recognized that parameters like the half saturation coefficient in the Monod equation (Equation 3.36) are influenced by transport effects,[10] thereby making them dependent on the physical characteristics of the biomass, such as floc size.

4.2 MASS BALANCE EQUATION

Transport phenomena models are based on the conservation of mass, momentum, and energy. However, for most suspended growth bioreactor models, only mass balances are required and thus our focus will be on them. Furthermore, because the elemental composition of many reactants and products in biochemical operations are unknown, it is usually more convenient to work in mass units than in molar units, as discussed in Section 3.1.1.

The starting point for a mass balance on any system is the specification of the control volume or the boundary of the system. When the reaction conditions, including the composition, are uniform over the whole reactor volume, then the entire reactor may be taken as the control volume. Otherwise, a differential element should be used. Having picked an appropriate control volume, a mass balance must be written for each reactant or product around the control volume by keeping track of the amounts entering, leaving, being generated, and being consumed.

The mass balance for a given constituent takes the form:

$$\begin{aligned}\text{New rate of accumulation in the control volume} \\ &= \text{rate of flow into the control volume} \\ &- \text{rate of flow out of the control volume} \\ &+ \text{net rate of generation in the control volume}\end{aligned} \tag{4.1}$$

or simply

$$\text{Accumulation} = \text{input} - \text{output} + \text{generation}. \tag{4.2}$$

Each term in the mass balance equation has the units of mass/time. The generation term represents the sum of all reactions in which the constituent of interest participates, and incorporates the reaction terms we saw in Chapter 3, as expressed in Equation 3.12. If it is positive, the constituent is being produced in the control volume; if it is negative, the constituent is being destroyed. Furthermore, as we also saw in Chapter 3, the reaction term for any given constituent may be a function of the concentrations of several constituents. If so, it will be necessary to solve several mass balance equations simultaneously to determine the concentrations of the constituents in the control volume.

4.3 REACTOR TYPES

As we saw in Chapter 1, suspended growth biochemical operations employ a number of different types of reactors. Most of them are continuous flow, which means that liquid flows through them continuously, bringing in reactants, and carrying away products. On occasion, however, environmental

engineers use batch reactors, which have no flow through them while the reaction is occurring. Rather, they are loaded, allowed to react, and then unloaded.

4.3.1 Ideal Reactors

4.3.1.1 Continuous Stirred Tank Reactor

A continuous stirred tank reactor (CSTR), also known as a continuous flow stirred tank reactor (CFSTR), backmix reactor or completely mixed reactor, is used frequently, particularly for experimental studies. As shown in Figure 4.1, it has a feed stream called the influent and an exit stream called the effluent. It is usually equipped with baffles and is mixed sufficiently to make mixing perfect; that is, mixing is homogeneous and instantaneous so that any reactant carried into the reactor by the feed is dispersed evenly throughout the reactor without any time delay. Thus, samples taken from all parts of the reactor have the same composition. In addition, the effluent composition is the same as the reactor composition.

Under the above assumptions, it is logical to pick the entire reactor volume as the control volume for writing mass balance equations. The application of Equation 4.2 on reactant A around the entire reactor volume, V, yields

$$V \cdot \frac{dC_A}{dt} = F_O \cdot C_{AO} - F \cdot C_A + r_A \cdot V, \tag{4.3}$$

where F_O and F are the volumetric flow rates of the influent and effluent, C_{AO} and C_A are the concentrations of A in the influent and effluent (or reactor), respectively, t is time, and r_A is the reaction rate of A. The generalized concentration symbol C has been used here to emphasize the point that the mass balance equation may be written for any constituent, regardless of its state (i.e., soluble or particulate). For the systems with which we deal, it is safe to assume that the flow rates of the influent and effluent are equal (i.e., $F_O = F$). If the influent flow rate and concentration are constant, it is common for such reactors to achieve steady state, which means that the time rate of change of any constituent is zero. This allows the accumulation term to be set equal to zero, thereby simplifying Equation 4.3:

$$-r_A = \frac{F}{V}(C_{AO} - C_A). \tag{4.4}$$

Since the right side of Equation 4.4 is the difference between the mass flow rates of A in the influent and effluent per unit volume of reactor and the left side is the net generation rate of A in the reactor,

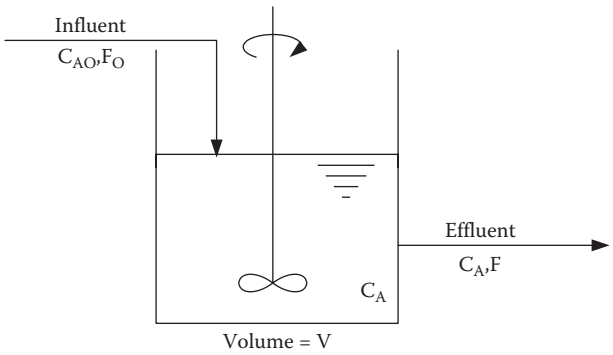

FIGURE 4.1 Continuous stirred tank reactor (CSTR).

Equation 4.4 states that the difference in the mass flow rates of A into and out of the reactor is due to the generation of A in the reactor. Note that if A is consumed, C_A will be less than C_{AO} and the generation rate will be negative. A negative generation rate is often called a consumption rate.

A steady-state mass balance equation like Equation 4.4 can be used for several things. First, because it allows calculation of the generation rate of A, it may be used to obtain a reaction rate expression experimentally. By varying the influent flow rate (F), the reactor volume (V), or the influent concentration (C_{AO}), and measuring the corresponding effluent concentration it is possible to determine how the reaction rate varies with the concentration of A (as well as the concentration of any other reactants). On the other hand, if the reaction rate expression is known, a rearranged form of Equation 4.4 can be used to determine the reactor volume required to achieve a desired effluent concentration:

$$V = \frac{F(C_{AO} - C_A)}{-r_A}. \tag{4.5}$$

Finally, Equation 4.4 can also be rearranged to allow calculation of the flow rate that can be handled by a reactor of volume V:

$$F = \frac{-r_A \cdot V}{C_{AO} - C_A}. \tag{4.6}$$

Thus, it can be seen that the mass balance equation for a CSTR is a powerful tool.

4.3.1.2 Plug-Flow Reactor

A plug-flow reactor (PFR) can be either a simple tube or one packed with a catalyst or some other type of packing. A feed containing the reactants is fed continuously to the reactor inlet while the effluent containing the products and unreacted reactants is removed from the outlet. The name of the reactor comes from the assumption that the flow pattern inside has a uniform velocity and concentration in the radial direction and no axial mixing, so that each fluid element moves through the reactor in the same order relative to all other elements, just like ping-pong balls rolling through a tube. The PFR is also known as a tubular or piston-flow reactor.

Because of the assumption of plug flow and because the reaction takes place all along the reactor length, the concentrations of reactants and products vary with the axial distance only. Therefore, it is appropriate to consider as the control volume an infinitesimal volume, ΔV, in which the concentration may be considered uniform. This control volume is shown in Figure 4.2.

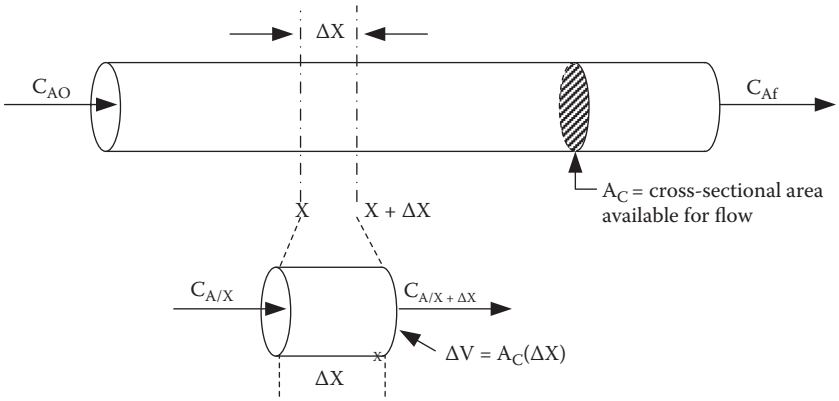

FIGURE 4.2 Plug-flow reactor (PFR).

The mass balance on component A around the control volume is

$$\frac{\partial(A_c \cdot \Delta x \cdot C_A)}{\partial t} = F \cdot C_A|_x - F \cdot C_A|_{x+\Delta x} + r_A \cdot A_c \cdot \Delta x, \tag{4.7}$$

or

$$A_c \frac{\partial C_A}{\partial t} = -\frac{F \cdot C_A|_{x+\Delta x} - F \cdot C_A|_x}{\Delta x} + r_A \cdot A_c, \tag{4.8}$$

where A_c is the cross-sectional area of the reactor, x is the distance from the reactor entrance, Δx is the length of the infinitesimal volume, and $F \cdot C_A|_x$ and $F \cdot C_A|_{x+\Delta x}$ the mass flow rates of A evaluated at the distances x and $x + \Delta x$ from the reactor entrance. In the limit as $\Delta x \to 0$, the first term on the right side in Equation 4.8 becomes the partial derivative of $F \cdot C_A$ with respect to distance and Equation 4.8 reduces to

$$A_c \frac{\partial C_A}{\partial t} = -\frac{\partial(F \cdot C_A)}{\partial x} + r_A \cdot A_c. \tag{4.9}$$

At a constant flow rate, the reactor will achieve steady state, at which there is no change with respect to time in the concentration at any point within the reactor, reducing the equation to

$$0 = -F \frac{dC_A}{dx} + r_A \cdot A_c, \tag{4.10}$$

or

$$r_A = \frac{F}{A_c} \cdot \frac{dC_A}{dx}. \tag{4.11}$$

Hence, it is theoretically possible to calculate the rates by determining the concentration gradient through the reactor. In practice, however, this is difficult to do so PFRs are seldom used to generate rate data in this fashion.

The main purpose for which the mass balance equation is used is to determine the size of reactor required to achieve a desired conversion from a feed stream of given composition and flow rate. If the rate expression is known, Equation 4.11 can be rearranged and integrated over the length of the reactor, L, to give

$$\int_0^L \frac{A_c \cdot dx}{F} = \frac{A_c \cdot L}{F} = \frac{V}{F} = \int_{C_{AO}}^{C_{Af}} \frac{dC_A}{r_A}. \tag{4.12}$$

Hence, the required ratio of reactor volume to feed flow rate, V/F, can be obtained.

4.3.1.3 Batch Reactor

The CSTR and the PFR are both continuous flow reactors. As stated previously, however, environmental engineers sometimes use batch reactors, which do not receive flow throughout the entire operational cycle. For the simplest cycle, they are rapidly charged with feed, allowed to react, and

the treated effluent removed. Under such conditions, a batch reactor has no input or output terms in the mass balance equation during the reaction period. As in a CSTR, a batch reactor is assumed to be perfectly mixed so that the entire reactor content is homogeneous at any given time. Hence, the rate is independent of position in the reactor and the concentration varies only with time. Therefore, it is appropriate to take the whole reactor volume as the control volume. The mass balance on reactant A around a batch reactor is

$$\frac{d(C_A \cdot V)}{dt} = 0 - 0 + r_A \cdot V. \tag{4.13}$$

For constant volume, this simplifies to

$$r_A = \frac{dC_A}{dt}, \tag{4.14}$$

which states that the reaction rate is equal to the time rate of change of the reactant concentration.

Equation 4.14 suggests that by running a batch reactor experiment, measuring the concentrations of appropriate reactants over time, and taking the time derivatives of the concentration history, one can obtain numerical values of the rate, r_A, associated with various concentrations of reactants. Having obtained data of rate versus reactant concentration, it is then possible to deduce the reaction rate expression, which can be used in the mass balance equations for any other batch or continuous flow reactor to determine the holding time or size required to achieve a given degree of reaction.

An analogy between a batch and a plug-flow reactor can be seen by comparing Equation 4.14 to Equation 4.11. Noting that the term F/A_c in Equation 4.11 is the axial velocity through the PFR and that distance divided by velocity is time, it can be seen that Equations 4.14 and 4.11 are the same when distance x is translated into the time required to reach it. Thus, it can be seen that each element of fluid passing through a PFR may be thought of as an infinitesimal batch reactor.

4.3.2 Nonideal Reactors

Ideal continuous flow reactors are useful for experimental purposes and for understanding how factors like the flow rate and reactor volume influence performance. In fact, a significant portion of this book will be devoted to modeling the performance of ideal reactors as a way of gaining understanding about biochemical operations. However, it is important to recognize that few wastewater treatment reactors are ideal. There are several reasons for this. One is size. The larger reactors are, the harder it is to achieve ideal mixing conditions. Another is the effect of other requirements. For example, the need for oxygen transfer to aerobic systems imparts considerable axial mixing in reactors that have been designed to approximate plug flow. Yet another is practicality. Site requirements and other such factors may prevent the economical construction of a reactor with a configuration that even approximates ideality.

4.3.2.1 Residence Time Distribution

Since most full-scale reactors in biochemical operations are nonideal, how might their mixing characteristics be identified and represented? One method is through measurement of the residence time distribution (RTD).

Examination of Equations 4.4 and 4.12 reveals that both contain the term V/F. This term represents the average length of time that an element of fluid (and therefore a dissolved constituent) stays in a reactor of constant volume receiving a constant flow rate of fluid with constant density. Hence, it has been given the name mean hydraulic residence time, or simply hydraulic residence time. It will

be given the symbol τ for use in equations and the acronym HRT for use in the text. It represents the time required to process one reactor volume of feed:

$$\tau = \frac{V}{F}. \tag{4.15}$$

Although the HRT is the mean residence time of fluid elements in a reactor, it is not the actual residence time of all elements. Rather, different elements of fluid reside in a reactor for different lengths of time, depending on the routes they follow, and the distribution of those times depends on the reactor's mixing characteristics.[5,12] Let the function, $F(t)$, be the fraction of the elements in the effluent stream having residence times less than t. With this definition, it is apparent that $F(0) = 0$ and $F(\infty) = 1$. In other words, none of the fluid can pass through the reactor in zero time and all must come out eventually. This function, which is shown in Figure 4.3, is known as the cumulative distribution function, or F curve. Another function is the point distribution function, $E(t)$, which is related to the cumulative distribution function by

$$E(t) = \frac{dF(t)}{dt}. \tag{4.16}$$

Therefore, it follows that $E(t)dt$ is the fraction of the effluent that has a residence time between t and t + dt, and thus is the residence time distribution function of the fluid in the reactor. The area under the RTD curve (also called the E curve) between the limits of 0 and ∞ is unity since the entire fraction of fluid must have residence times between 0 and ∞. A typical RTD (E) curve is shown in Figure 4.4.

The mixing characteristics of the two ideal continuous flow reactors represent the extremes between which all others lie. In a perfect PFR, all fluid elements stay in the reactor for exactly the mean residence time. Thus, there is no distribution of residence times; they are all the same. In

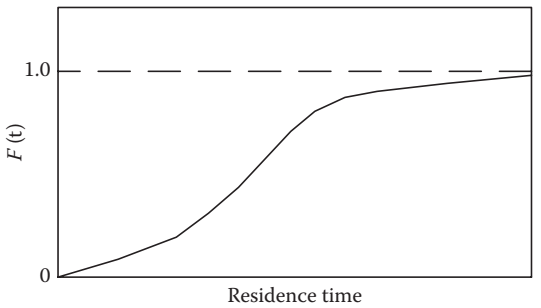

FIGURE 4.3 Cumulative residence time distribution function, $F(t)$.

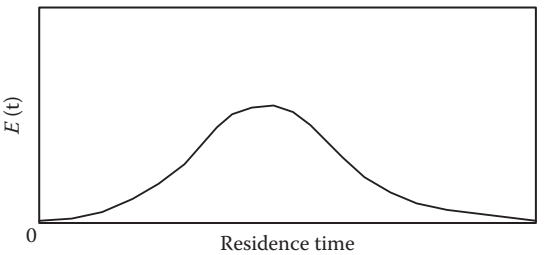

FIGURE 4.4 Point residence time distribution function, $E(t)$.

contrast, in a CSTR all fluid elements have equal probability of leaving the reactor at any moment, regardless of how long they have been in it. This means that the residence time of a fluid element in a CSTR is not fixed but is subject to statistical fluctuations.[7] In particular, the RTD is a negative exponential:

$$E(t) = \frac{1}{\tau} e^{-t/\tau}. \tag{4.17}$$

Integrating Equation 4.17 from $t = 0 \rightarrow \infty$ confirms that the area under the $E(t)$ curve for a CSTR is indeed unity. Furthermore, Equation 4.17 shows that the most probable residence time for fluid elements in a CSTR is zero. The mean residence time, τ, is also the standard deviation of the distribution of residence times.

4.3.2.2 Experimental Determination of Residence Time Distribution

The RTD for a given reactor and flow rate may be determined experimentally by introducing an inert tracer into the reactor input and observing the time response of the tracer concentration in the reactor effluent. The two most convenient types of tracer inputs are step and impulse signals.[8]

Imagine a reactor receiving influent at constant flow rate, F. At time zero a continuous flow of a soluble tracer is added to that stream, instantly changing the tracer concentration from zero to S_{TO} and maintaining it at that concentration thereafter (i.e., a step change in concentration). (Note that the symbol S_T is used to represent the concentration of the tracer; S implies that the tracer is a soluble constituent and the subscript T identifies it as tracer. This convention, with the main symbol representing the state of the constituent and the subscript its identity, is used throughout this book.) We immediately start measuring the output tracer concentration, S_T, and plot S_T/S_{TO} versus time. How does the resulting curve relate to the RTD? To answer that, imagine that we could divide the effluent stream into two fractions, one that has spent less time than t in the reactor, $F(t)$, and one that has spent more, $1 - F(t)$, where t is measured from the time that the tracer was introduced. Any flow that has been in the reactor less than t will have the tracer in it, and thus, the fraction $F(t)$ will have a tracer concentration $S_T = S_{TO}$. Conversely, any fluid that has been in the reactor more than t was present before the tracer was introduced and thus has no tracer in it. Consequently, the fraction $1 - F(t)$ has $S_T = 0$. Since mass flow rate is the product of liquid flow rate and concentration, the total mass flow rate of tracer in the effluent at time t, $F \cdot S_T(t)$, will be

$$F \cdot S_T(t) = F(t) \cdot F \cdot S_{TO} + [1 - F(t)]F \cdot 0. \tag{4.18}$$

That is

$$F(t) \cdot F \cdot S_{TO} = F \cdot S_T(t) \tag{4.19}$$

or

$$F(t) = \frac{S_T(t)}{S_{TO}}. \tag{4.20}$$

Equation 4.20 states that the cumulative distribution function, $F(t)$, is identical to the normalized tracer concentration response due to the step input. Thus, imposition of a step input of tracer to a reactor is a convenient way of experimentally determining the cumulative distribution function. According to Equation 4.16, it may be differentiated to obtain the RTD curve.

Differentiation of experimental data is risky, and thus it would be better to have a direct way of determining the RTD function. A similar analysis can be used to establish that the point distribution function, *E*(t), is identical to the normalized effluent concentration curve resulting from an impulse input into the feed.[11] An impulse input is one in which a slug of tracer is added instantaneously to the reactor feed. Consequently, in this case, the measured output concentration is normalized by dividing each measured concentration, S_T, by the total area under the curve of S_T versus time. The area can be obtained by Simpson's rule or by graphical integration. Furthermore, the area is equivalent to the total mass of tracer added divided by the flow rate to the reactor, if both of those quantities are known. Equation 4.16 also suggests that the *F*(t) function can be obtained from the impulse input response by integration. Hence, either a step input or an impulse input may be used to obtain both *E*(t) and *F*(t). However, the more commonly used technique is to obtain the *E*(t) function from an impulse input, and then obtain the *F*(t) function by integration if it is needed.

Use of a tracer test on a reactor will allow the determination of its mixing characteristics. If the RTD function resulting from the test conforms to either of the ideal reactor types, the performance of the reactor can be predicted or simulated by application of mass balance equations for all relevant constituents with appropriate reaction rate expressions. However, if the RTD function deviates from the ideal, more involved techniques must be used.

4.4 MODELING NONIDEAL REACTORS

The use of RTDs for the prediction of reactor performance is a complex subject and the reader is encouraged to consult other sources for more complete coverage.[6,8,11] The techniques commonly used in environmental engineering are relatively straightforward, however, and thus we will look at them briefly.

4.4.1 Continuous Stirred Tank Reactors in Series Model

The simplest way to model a reactor with a nonideal flow pattern is as a series of CSTRs and this technique will be used extensively in this book. The basis for doing this may be seen by considering the response of such a system to a step input of tracer. Consider a chain of N CSTRs with total volume V. As shown in Figure 4.5, the CSTRs are of equal size, each with volume V/N, and receive a flow F. This gives each CSTR a mean HRT of τ/N, where τ is the mean HRT of the entire chain (i.e., V/F). At time zero the feed to the first tank is switched to one with a tracer concentration S_{TO}. The effluent from the first tank is the feed to the second tank; the effluent of the second is the feed to the third, and so on. If we write and solve the mass balance equations for each (with no reaction term since the tracer is assumed to be inert), we obtain the following expression for the tracer concentration leaving the N^{th} reactor, S_{TN}:

$$\frac{S_{TN}}{S_{TO}} = 1 - \left[1 + [(t \cdot N)/\tau] + \frac{[(t \cdot N)/\tau]^2}{2!} + \cdots + \frac{[(t \cdot N)/\tau]^{N-1}}{(N-1)!}\right] e^{-(t \cdot N)/\tau}. \tag{4.21}$$

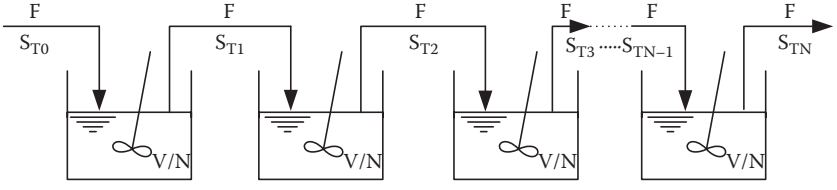

FIGURE 4.5 CSTRs in series.

Tracer concentration profiles [i.e., $F(t)$ curves] for various numbers of tanks in series are shown in Figure 4.6 while the corresponding $E(t)$ curves are shown in Figure 4.7. The curve for N = 1 is the classical response of a single CSTR. More significantly, however, the curve for N = ∞ is the classical response for a PFR; that is, a step change in effluent concentration after one HRT. This suggests that the step response for N CSTRs in series will lie somewhere between that of a single CSTR and a PFR, with the pattern depending on the number of tanks in the chain. Furthermore, this implies that a real reactor, which has the response of neither a CSTR nor a PFR, can be simulated as N CSTRs

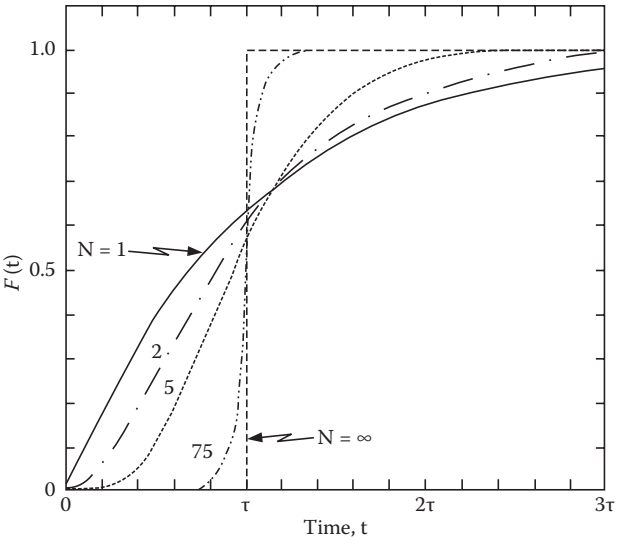

FIGURE 4.6 Responses of N CSTRs in series to a step tracer input. The τ in the abscissa is the total hydraulic residence time (HRT) for N CSTRs in series.

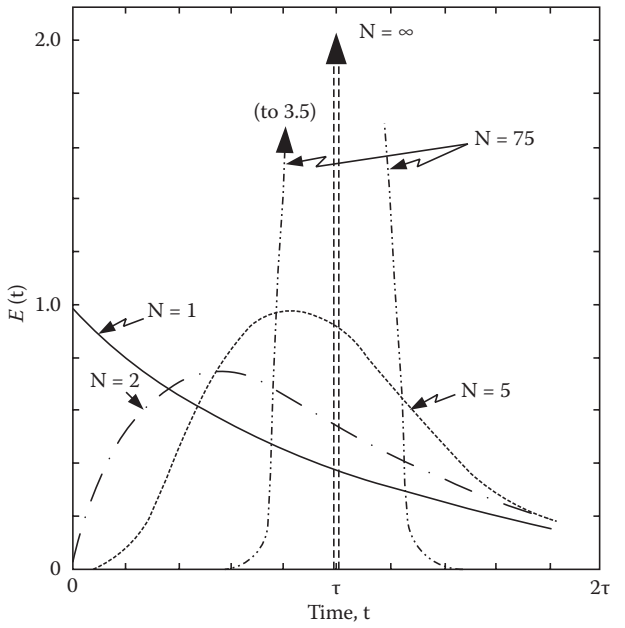

FIGURE 4.7 Responses of N CSTRs in series to an impulse tracer input. The τ in the abscissa is the total hydraulic residence time (HRT) for N CSTRs in series.

in series. The easiest way of determining the appropriate value for N, which is sufficiently accurate in many cases, is to plot either the *F*(t) or the *E*(t) curve for the reactor in question and compare it to the curves in Figures 4.6 and 4.7, thereby selecting the value of N that corresponds most closely.

The tanks in series model has been used frequently in environmental engineering practice. For example, Murphy and Boyko[9] found that many conventional activated sludge systems had RTDs that were equivalent to three to five CSTRs in series.

4.4.2 Axial Dispersion Model

An alternative approach to modeling a nonideal reactor is to superimpose some degree of backmixing upon a plug flow of fluid. The magnitude of the backmixing is assumed to be independent of the position in the reactor and is expressed by the axial dispersion coefficient, D_L, which is analogous to the coefficient of molecular diffusion in Fick's Law of diffusion. Modeling the superimposed backmixing by axial dispersion requires adding another transport term to the mass balance equation for the differential element of the PFR in Figure 4.2. In addition to the advective transport term used in Equation 4.9, a term for transport by axial dispersion must be included, thereby increasing the number of terms in the resulting partial differential equation:

$$\frac{\partial C_A}{\partial t} = D_L \frac{\partial^2 C_A}{\partial x^2} - \frac{F}{A_c}\frac{\partial C_A}{\partial x} + r_A. \tag{4.22}$$

The equation is usually rewritten to include the term D_L/vL (the dispersion number),

$$\frac{\partial C_A}{\partial \theta} = \frac{D_L}{vL}\frac{\partial^2 C_A}{\partial z^2} - \frac{\partial C_A}{\partial z} + \tau \cdot r_A, \tag{4.23}$$

in which v is the longitudinal velocity through the basin (F/A_c), L is the basin length, z is the dimensionless distance along the basin (x/L), and θ is dimensionless time (t/τ), where τ is the mean HRT for the reactor. When the dispersion number is zero, there is no axial dispersion and therefore plug flow, whereas when it is infinitely large, complete backmixing exists and the reactor behaves as a CSTR.

The effect of the value of the dispersion number on the RTD may be seen by solving Equation 4.23 with the appropriate initial and boundary conditions for a step input of an inert tracer.[8] The solution, expressed in the form of an *F*(t) curve, is shown in Figure 4.8. To characterize the mixing

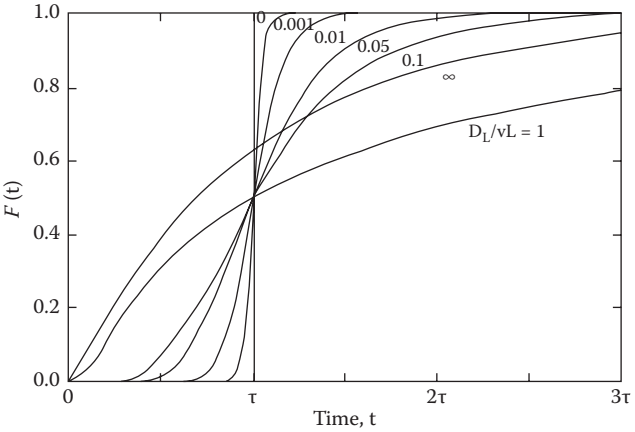

FIGURE 4.8 Step input tracer response for the axial dispersion model.

pattern in a reactor by this technique, we need a way to select the appropriate value for the dispersion number. One way is to evaluate the derivative of the $F(t)$ function at one mean residence time:

$$\left.\frac{dF(t)}{dt}\right|_{t=\tau} = \left(\frac{1}{2\tau(\pi)^{0.5}}\right)\left(\frac{vL}{D_L}\right)^{0.5} \tag{4.24}$$

or

$$\frac{D_L}{vL} = \frac{1}{4\pi\tau^2\left(\left.\frac{dF(t)}{dt}\right|_{t=\tau}\right)^2}. \tag{4.25}$$

Thus, the dispersion number for a nonideal reactor can be approximated by determining the slope of the $F(t)$ curve at the time equal to one HRT and using Equation 4.25. There are also several ways of determining the dispersion number using the $E(t)$ curve and they are discussed in detail elsewhere.[8]

After the dispersion number has been determined, the performance of the reactor can be determined by using Equation 4.23 with the appropriate reaction rate expressions. This often requires numerical techniques. Because of the complexity of such an approach, CSTRs in series have been used more frequently to model system performance in environmental engineering practice. However, the dispersion number is often used to characterize the flow patterns in reactors for other purposes.[4]

4.4.3 Representation of Complex Systems

The reactor systems used for some suspended growth biochemical operations are quite simple, whereas those used for others are complex, as seen in Table 1.2. Luckily, the concepts presented in this chapter suggest that many of the complex systems can be modeled simply as CSTRs in series, although those CSTRs may be of different size or may contain different biochemical environments (i.e., aerobic or anoxic). In Chapter 7 we will use this technique to investigate the theoretical performance of several systems.

Sometimes the flow patterns in reactor systems are so complex that the RTDs cannot be modeled by the techniques discussed above. In that case, it may be necessary to model the system as a complex network of flow regions with various modes of flow between and around them. This technique was proposed by Cholette and Cloutier[3] and is based on the premise that just as adding different numbers of tanks in series changes the RTD, so will adding different types of mixers in series and parallel. The main difference is that by adding different types of regions and flows it is possible to reproduce almost any RTD. As one might expect, however, the techniques for doing this can be quite complex. Thus the reader is referred to other sources for information about their use.[3,8]

4.5 KEY POINTS

1. The models used to depict suspended growth biochemical operations are phenomenological mass transport models that contain a number of implicit assumptions.
2. The mass balance equation for any constituent in the control volume of a suspended growth reactor is

$$\text{Accumulation} = \text{Input} - \text{Output} + \text{Generation}.$$

3. Reactors may be characterized by the distribution of residence times within them, with continuous stirred tank reactors (CSTRs) and plug-flow reactors (PFRs) representing the two extremes of ideality. Most reactors are nonideal, however, with residence time distributions between the two extremes.
4. The residence time distribution for a reactor may be determined directly from the normalized concentration response of the reactor to the imposition of an impulse input of tracer to the feed. It may also be obtained by differentiating the normalized concentration response to the imposition of a step input of tracer.
5. Within environmental engineering practice, the two most common techniques for modeling the performance of a nonideal reactor is as a series of CSTRs or as a PFR with axial dispersion.

4.6 STUDY QUESTIONS

1. List and discuss the significance of the six assumptions commonly made during the modeling of suspended growth reactors.
2. What is meant by the terms "hydraulic residence time" and "residence time distribution"? Describe a continuous stirred tank reactor (CSTR) and a plug-flow reactor (PFR) in terms of their hydraulic residence times and residence time distributions. What are the physical characteristics of each reactor that are responsible for the nature of its residence time distribution?
3. A reactor with a volume of 4.0 L receives a constant flow of water at a rate of 3.0 L/hr. At time zero an impulse of dye is added to the reactor that instantaneously places a dye concentration of 1200 mg/L into it. Effluent samples taken 1, 2, 3, 4, and 6 hours after addition of the dye have concentrations of 567, 268, 126, 60, and 13 mg/L, respectively. Is the reactor a perfect CSTR?
4. Explain how you would use the data from a tracer study to choose the appropriate model for characterizing the performance of a nonideal reactor.
5. Given in Table SQ4.1 are the response data from a reactor ($\tau = 30$ min) to which a step input of tracer was applied. If you desired to model this reactor as a series of CSTRs, how many CSTRs would you use? If you desired to model the reactor with an axial dispersion model, what dispersion number would you use?

TABLE SQ4.1
Step Input Tracer Response of a Reactor

t min	S_T/S_{TO}	t min	S_T/S_{TO}
0	0	45	0.88
5	0.006	50	0.96
10	0.017	55	0.98
15	0.084	60	0.99
20	0.220	65	0.99
25	0.380	70	0.99
30	0.550	75	0.99
35	0.700	80	0.99
40	0.800	90	1.00

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5 Aerobic Growth of Heterotrophs in a Single Continuous Stirred Tank Reactor Receiving Soluble Substrate

By returning to Chapter 1 and studying Table 1.2 it can be seen that the single continuous stirred tank reactor (CSTR) is the simplest reactor configuration used in biochemical operations, finding application in activated sludge, aerated lagoons, aerobic digestion, anaerobic digestion, and biological nutrient removal. It has also found extensive use in microbiological and environmental engineering research, so that much of our knowledge about microbial growth has come from it.

Because of its simplicity, the CSTR provides an ideal system with which to study the modeling of microbial growth. Consequently, in this chapter we will develop models describing the growth of microbial cultures in a single CSTR, either with or without biomass recycle, and use them to gain an understanding of how such systems behave. For simplicity, the models will be confined to the growth of aerobic heterotrophic biomass in an environment that contains ample nutrients so that the soluble, biodegradable organic substrate (expressed in chemical oxygen demand [COD] units) is the growth limiting material. Furthermore, the traditional approach to modeling biomass decay will be used because it is frequently found in the literature, making it important that the reader be familiar with it. With minor modifications, the same models can be applied to anoxic heterotrophic growth, as well as to aerobic autotrophic growth, making the principles illustrated of general importance.

5.1 BASIC MODEL FOR A CONTINUOUS STIRRED TANK REACTOR

A schematic diagram of a single CSTR is shown in Figure 5.1. A bioreactor with volume V receives a flow at rate F containing only soluble, noninhibitory, biodegradable organic substrate at concentration S_{SO} (in COD units) and sufficient inorganic nutrients to make the organic substrate the growth limiting material. The influent flow and concentrations are constant, as are pH, temperature, and other environmental conditions. Within the bioreactor, the active heterotrophic biomass uses the soluble substrate as its food source, thereby growing to concentration $X_{B,H}$ while reducing the substrate concentration to S_S. Biomass decay accompanies the growth so that microbial debris at concentration X_D is also present. Two effluent streams are discharged from the bioreactor, but because the bioreactor is completely mixed, the concentrations of all soluble constituents in them are the same as the concentrations in the bioreactor. One stream, with flow rate F_w, flows directly from the bioreactor and carries biomass and cell debris at concentrations equal to those in the bioreactor. The other, with flow $F - F_w$, passes through a biomass separator before discharge, making it free of suspended material. All particulate material removed by that separator is retained in the bioreactor, thereby increasing the concentration of biomass in the bioreactor and increasing the overall rate at which reactions occur. By maintaining a high biomass concentration in the bioreactor, the volume needed to achieve a given amount of substrate removal is reduced, thereby saving space.

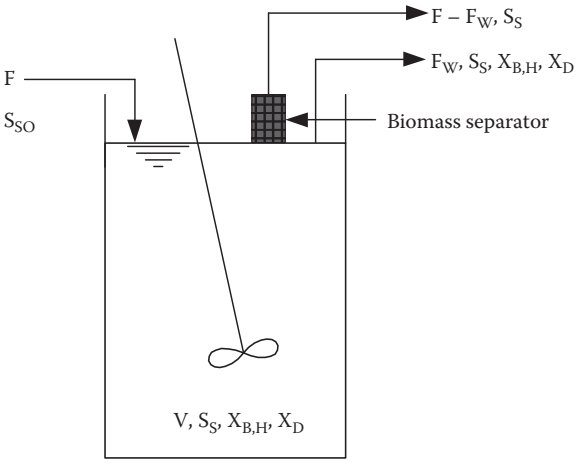

FIGURE 5.1 Schematic diagram of a CSTR. The biomass separator returns biomass from the effluent to the reactor. The stream F_w removes biomass from the reactor at a concentration equal to that in the reactor.

5.1.1 Methods of Solids Separation and Wastage

Solids separation devices retain a higher concentration of biomass in a bioreactor than would occur without them and are a key feature of several suspended growth reactor configurations. The technique that most closely mirrors the situation pictured in Figure 5.1 is the submerged membrane bioreactor, illustrated in Figure 1.9b. Membranes are typically configured as small hollow tubes that contain very small (generally 0.2 or 0.02 μm) pores. Water and dissolved constituents pass through the pores of the membrane into the center of the hollow tubes and are pumped out as effluent, whereas biomass is too large to fit through the pores, allowing it to be retained in the bioreactor. Another separation device that mirrors Figure 5.1 is an internal, upflow clarifier. This system employs a quiescent zone in the bioreactor into which flow enters from the bottom. By designing the system so that the upflow velocity of the effluent fluid is less than the settling velocity of the biofloc, it is possible to discharge a clarified effluent while retaining the biomass in the reactor. In both cases, wastage is ideally taken directly from the mixed bioreactor.

In contrast to the situation depicted in Figure 5.1, most full-scale suspended growth bioreactor systems use gravity sedimentation in a separate tank as the means of biomass separation. The most common technique is to use an external clarifier with the return of a concentrated solids stream, called solids (or biomass) recycle, to the bioreactor. Figure 5.2 illustrates two such systems that differ in how biomass wastage is implemented. The system shown in Figure 5.2a is called the Garrett configuration.[2] Its distinguishing characteristic is that it wastes biomass directly from the bioreactor, just as is done in Figure 5.1. The system shown in Figure 5.2b is the configuration most often found in practice and, thus, is called the conventional configuration. Its distinguishing characteristic is that biomass is wasted from the concentrated solids recycle stream. Wastage strategies and factors pertinent to each will be discussed further in Section 5.4. Although the configurations in Figure 5.2 are more common, the configuration depicted in Figure 5.1 will be assumed in deriving the equations presented in this chapter because it allows several fundamental concepts to be introduced in a simple manner.

On a laboratory scale, gravity sedimentation is commonly used along with the Garrett configuration for wasting. However, it is often beneficial to use reactors that mimic closely the ideal settling situation depicted in Figure 5.1. In one design, called a porous pot reactor, the bioreactor has porous walls through which clear liquid flows, leaving the biomass inside. The wastage flow is taken directly from the vessel. In another design, tangential-flow membrane filtration is used as a biomass separator, with a high recirculation flow being removed from the bioreactor, passed over the filter,

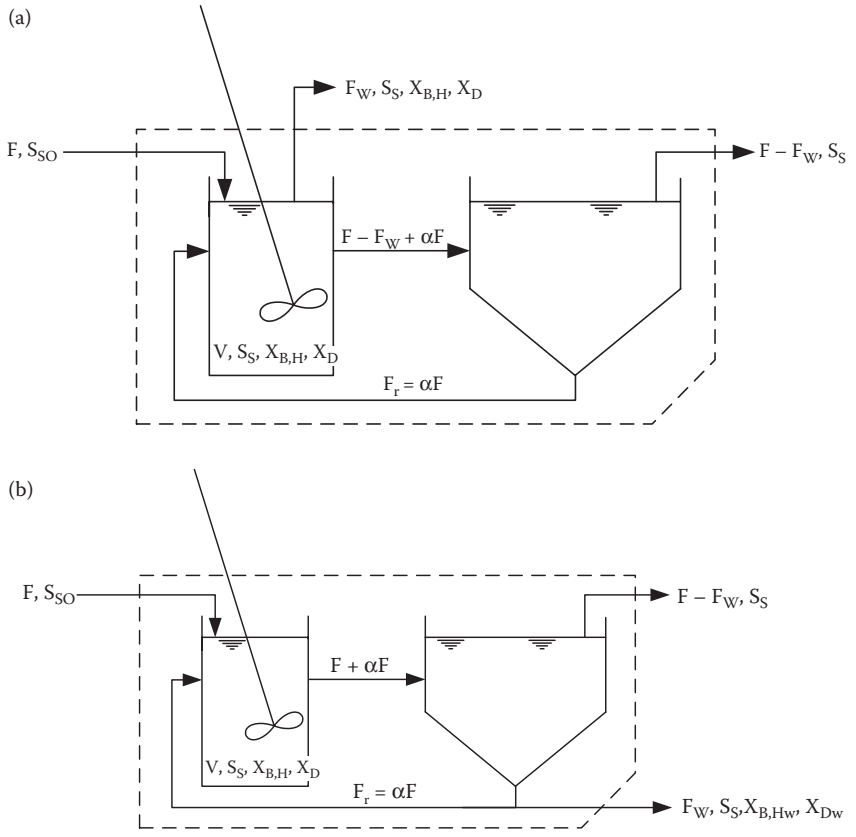

FIGURE 5.2 Schematic diagrams of two CSTRs with biomass recycle from sedimentation basins. (a) The Garrett configuration in which biomass is wasted directly from the reactor. (b) Conventional configuration in which biomass is wasted from the sludge recycle flow.

and returned to the bioreactor while a small percentage of liquid passes through the filter for discharge as the biomass-free effluent. Again, biomass wastage is through a stream removed directly from the bioreactor (i.e., the Garrett technique).

5.1.2 Definitions of Residence Times

A residence time defines the average amount of time a component stays in a system. Two types of components are present in the CSTR in Figure 5.1: soluble, denoted by the symbol S and particulate, denoted by the symbol X. Because their residence times are not necessarily the same, one must be defined for each.

Dissolved components are intimately associated with the fluid and cannot be easily separated from it. Thus, their residence time in a reactor is equal to the mean hydraulic residence time (HRT), which was defined by Equation 4.15:

$$\tau = \frac{V}{F}. \tag{4.15}$$

Particulate components can be separated from a fluid by physical means, such as filtration or settling, and engineers use this characteristic to control their discharge from a bioreactor. The bioreactor shown in Figure 5.1 has two discharge streams, one containing particulate material at a

concentration equal to that in the bioreactor and one free of it. The stream containing particulate material, F_w, is called the wastage stream because it provides the means for wasting particulate material from the bioreactor. The other stream, called the main effluent stream, is free of particulate material because it has passed through an ideal separator. Although in reality gravity settled streams contain small concentrations of particulate material due to inefficiencies in the separation device, we will consider them to be totally free of such material for now. In Part III we will consider how to account for the presence of particulate material in the effluent from real systems.

The importance of the biomass separator is that it makes the residence time of particulate materials greater than the residence time of soluble materials. Thus, we must define a second residence time, called the solids retention time or mean cell residence time, which represents the average length of time a particulate component stays in a bioreactor. It will be given the symbol Θ_c for use in equations and the acronym SRT for use in the text. By analogy to Equation 4.15, the SRT is defined as the mass of a particulate component contained in the bioreactor divided by the mass discharged from the bioreactor per unit time:

$$\Theta_c = \frac{V \cdot X}{F_W \cdot X_W}. \tag{5.1}$$

For the case illustrated in Figure 5.1, in which the concentration of particulate material in the wastage stream, X_w, is equal to the concentration in the bioreactor, X, Equation 5.1 may be simplified to

$$\Theta_c = \frac{V}{F_W}. \tag{5.2}$$

However, it should be remembered that the basic definition of SRT involves mass flow rates rather than volumetric flow rates. The comparison of Equation 5.2 to Equation 4.15 reveals that:

$$\Theta_c \geq \tau. \tag{5.3}$$

In other words, the closer F_w approaches F, the closer the SRT approaches the HRT, so that in the limiting case where no biomass separator is employed, the SRT and the HRT are the same.

5.1.3 Format for Model Presentation

To describe any reactor, mass balance equations must be written around a control volume for all components of importance. As discussed in Section 4.3.1, the appropriate control volume for a CSTR is the entire reactor, since it is homogeneous throughout. If we let C_A represent the mass-based concentration of component A in the reactor in Figure 5.1, and consider the reactor to have constant volume, the mass balance equation is

$$\frac{dC_A}{dt}V = F \cdot C_{AO} - F_W \cdot C_A - (F - F_W)C'_A + r_A \cdot V, \tag{5.4}$$

where C_{AO} is the concentration of A in the influent stream and C'_A is its concentration in the stream passing through the biomass separator. For soluble components, C'_A is the same as the concentration in the CSTR, whereas for particulate components, C'_A is zero because the biomass separator is assumed to be ideal. The reaction term r_A represents the sum of all reactions in which component A participates, as we saw in Equation 3.12. Furthermore, as we saw in Chapter 3, r_A may be

a function of the concentrations of several components. If so, it will be necessary to solve several mass balance equations simultaneously to determine the concentration of any single component in a bioreactor.

For the situation depicted in Figure 5.1, mass balance equations must be written for at least three components: S_S, $X_{B,H}$, and X_D. In addition, we will be interested in the amount of oxygen that must be supplied through an oxygen transfer system. Thus, we will also need a mass balance equation for it, making a total of four. Because of the advantages of expressing the concentrations in COD units, as discussed in Section 3.1.1, the mass balance equations will be written in those units.

Consideration of the number of mass balance equations required for this simple situation and reflection on the number of events and components that could be considered, as discussed in Chapters 2 and 3, make it clear that a system is needed for providing the required information. The matrix format[4,5,7] discussed in Section 3.1.3 provides such a system.

Table 5.1 presents all of the information required to compile the reaction rate terms, r_i, for insertion into COD-based mass balance equations for the situation described above. It contains the information in Equation 3.11, with the entries in the body of the table representing the stoichiometric coefficients in COD units for each component participating in each reaction (recall that oxygen is negative COD). The entries in the right column represent the process rates for the reactions, and the subscript H on the coefficients signifies that they are applicable to heterotrophic biomass. Only two reactions are considered in this case, growth and decay.

The stoichiometric equation for growth with active biomass as the reference component was given in COD units by Equation 3.33:

$$\left(-\frac{1}{Y_H}\right)S_S + (-1)\left[-\left(\frac{1-Y_H}{Y_H}\right)\right]S_O + X_{B,H} = 0. \tag{3.33}$$

Examination of Equation 3.33 reveals that the coefficients in Table 5.1 correspond to the coefficients in the equation, and thus their sum should equal zero, as it does. The term process rate, r_j, in Table 5.1 refers to the generalized reaction rate for process j as defined in Equation 3.10. For the growth process it can be obtained by substituting Equation 3.35a:

$$r_{XB} = \mu_H \cdot X_{B,H}, \tag{3.35a}$$

for r_{XB} in Equation 3.34a:

$$\frac{r_{XB}}{1} = r. \tag{3.34a}$$

TABLE 5.1
Process Kinetics and Stoichiometry for Aerobic Growth of Heterotrophic Bacteria; Traditional Model for Biomass Decay

	Component[a]				Process
Process	$X_{B,H}$	X_D	S_S	S_O[b]	**Rate, r_j**
Growth	1		$-(1/Y_H)$	$[(1-Y_H)/Y_H]$	$\mu_H \cdot X_{B,H}$
Decay	−1	f_D		$(1 - f_D)$	$b_H \cdot X_{B,H}$

[a] All components and coefficients are expressed as COD.

[b] Coefficients must be multiplied by −1 to express them as oxygen.

As we saw in Section 3.2.7, the specific growth rate coefficient, μ_H, is a function of the substrate concentration, S_S, but to simplify upcoming explanations, the substitution will be made later.

The COD-based stoichiometric equation for the traditional approach of modeling decay was given by Equation 3.53, which when written in symbolic form is

$$(-1)X_{B,H} + (1 - f_D)S_O + f_D \cdot X_D = 0. \tag{3.53a}$$

Inspection of Equation 3.53a reveals its correspondence to the coefficients in Table 5.1. Again, summing the coefficients reveals that continuity has been maintained. The process rate expression for biomass decay may be obtained by multiplying Equation 3.56a

$$r_{XB} = -b_H \cdot X_{B,H} \tag{3.56a}$$

by the stoichiometric coefficient for biomass in Table 5.1 (i.e., –1), giving the result shown in the table.

As stated by Equation 3.12, the overall rate expression for each component to be inserted into its mass balance equation is obtained by summing the products of the process rate expressions times the stoichiometric coefficients appearing in the column under the component. For more complex situations it is more convenient to express this in matrix format as illustrated by Mason et al.,[8] but for this simple case, the overall rate expression for each component can be obtained by examination of Table 5.1. Both reactions influence active biomass, $X_{B,H}$. Consequently, the overall reaction rate (r_{XB}) for it in COD units is obtained by multiplying the generalized reaction rate for each process by the corresponding stoichiometric coefficient in the column under $X_{B,H}$ and summing them:

$$r_{XB} = \mu_H \cdot X_{B,H} - b_H \cdot X_{B,H}. \tag{5.5}$$

Only one of the processes influences biomass debris, and that is decay, which generates it. Consequently, the overall reaction rate for biomass debris (r_{XD}) in COD units is

$$r_{XD} = b_H \cdot f_D \cdot X_{B,H}. \tag{5.6}$$

Likewise, soluble organic substrate is only influenced by one process, growth, so that:

$$r_{SS} = -\left(\frac{\mu_H}{Y_H}\right)X_{B,H}. \tag{5.7}$$

Finally, oxygen is influenced by both processes. Thus, its reaction rate (r_{SO}) includes two terms, which in COD units gives:

$$r_{SO} = \left(\frac{1 - Y_H}{Y_H}\right)\mu_H \cdot X_{B,H} + (1 - f_D)b_H \cdot X_{B,H} \quad \text{(as COD)}. \tag{5.8}$$

Note that as obtained from application of Equation 3.12, the overall rate expression for oxygen as COD carries a positive sign. This does not mean that oxygen is generated, but rather that oxygen demand is generated, which means that oxygen is used. To convert the rate expression to more familiar units of oxygen, it must be multiplied by –1, giving:

$$r_{SO} = -\left[\left(\frac{1 - Y_H}{Y_H}\right)\mu_H \cdot X_{B,H} + (1 - f_D)b_H \cdot X_{B,H}\right] \quad \text{(as } O_2\text{)}. \tag{5.9}$$

This makes it clearer that oxygen is being utilized in the system.

5.1.4 Alternative Methods of Expressing Biomass Concentrations and Yields

While the use of COD-based mass balance equations is necessary for maintaining electron balances, COD is not a convenient means of measuring the concentrations of biomass and other particulate material in bioreactors. Rather, particulate matter is typically measured in mass units, either as total suspended solids (TSS) or as volatile (i.e., organic) suspended solids (VSS). Thus, we must use conversion factors to convert biomass measured in mass units to equivalents of biomass COD. Let subscript T represent the measurement of biomass in TSS units and subscript V its measurement in VSS units. Thus, the symbol $X_{B,H,T}$ represents the concentration of heterotrophic biomass expressed in TSS units and the symbol $X_{B,H,V}$ its concentration in VSS units. The conversion factor from TSS units to COD units is denoted as $i_{O/XB,T}$ and the conversion factor from VSS units to COD units is denoted as $i_{O/XB,V}$. Thus, in symbolic form,

$$X_{B,H} = i_{O/XB,T} \cdot X_{B,H,T} \tag{5.10}$$

and

$$X_{B,H} = i_{O/XB,V} \cdot X_{B,H,V}. \tag{5.11}$$

As discussed in Section 2.4.1, if we assume an empirical formula for the organic (i.e., ash-free) portion of biomass of $C_5H_7O_2N$, the COD of that organic portion can be calculated as 1.42 g COD/g VSS.[6] In other words,

$$i_{O/XB,V} = 1.42 \text{ g COD/g VSS}. \tag{5.12}$$

Furthermore, if we assume the ash content of biomass to be 15%, the theoretical COD of biomass is 1.20 g COD/g TSS, or

$$i_{O/XB,T} = 1.20 \text{ g COD/g TSS}. \tag{5.13}$$

As discussed in Chapter 9, values of $i_{O/XB,V}$ and $i_{O/XB,T}$ can be measured for biomass grown on a particular wastewater as part of the treatability studies, but for the purposes of this book, the values given by Equations 5.12 and 5.13 will be assumed throughout. Similar conversions will be assumed for biomass debris, X_D, allowing its expression in TSS ($X_{D,T}$) or VSS ($X_{D,V}$) units.

The biomass yield (Y_H) in Table 5.1 is in COD units (i.e., mg biomass COD formed/mg substrate COD used), as are the equations obtained from it. However, if biomass is to be expressed in TSS or VSS units, the yield must also be expressed in those units; that is, as mg biomass TSS (or VSS) formed/mg substrate COD used, $Y_{H,T}$ (or $Y_{H,V}$). This is accomplished by the use of the conversion factors $i_{O/XB,T}$ or $i_{O/XB,V}$ as shown in Equation 5.14:

$$Y_H = i_{O/XB,T} \cdot Y_{H,T} = i_{O/XB,V} \cdot Y_{H,V}. \tag{5.14}$$

Substitution of Equation 5.10 for $X_{B,H}$ and Equation 5.14 for Y_H in all COD-based mass balance equations and reaction rate expressions allows biomass concentrations to be expressed in TSS units, while maintaining COD as the basis for the mass balances. Performance of these substitutions results in the modified matrix in Table 5.2. Note that biomass and debris are denoted by $X_{B,H}$ and X_D, respectively, in the component heading to indicate that the stoichiometry and kinetics in the table are COD-based. Making the appropriate substitutions, the reaction rate expressions for biomass, debris, soluble substrate, and oxygen are, respectively:

$$r_{XB} = \mu_H \cdot X_{B,H,T} \cdot i_{O/XB,T} - b_H \cdot X_{B,H,T} \cdot i_{O/XB,T}, \tag{5.15}$$

TABLE 5.2
Process Kinetics and Stoichiometry for Aerobic Growth of Heterotrophic Bacteria; Traditional Model for Biomass Decay with Biomass Expressed in TSS Units

	Component[a]				Process
Process	$X_{B,H}$	X_D	S_S	S_O[b]	Rate, r_j
Growth	1		$-1/(Y_{H,T}\cdot i_{O/XB,T})$	$[1-(Y_{H,T}\cdot i_{O/XB,T})]/(Y_{H,T}\cdot i_{O/XB,T})$	$\mu_H\cdot X_{B,H,T}\cdot i_{O/XB,T}$
Decay	−1	f_D		$(1-f_D)$	$b_H\cdot X_{B,H,T}\cdot i_{O/XB,T}$

[a] All components and coefficients are expressed as COD.
[b] Coefficients must be multiplied by −1 to express them as oxygen.

$$r_{XD} = b_H \cdot f_D \cdot X_{B,H,T} \cdot i_{O/XB,T}, \tag{5.16}$$

$$r_{SS} = -\left(\frac{\mu_H}{Y_{H,T}}\right) X_{B,H,T}, \tag{5.17}$$

and

$$r_{SO} = \left(\frac{1 - Y_{H,T}\cdot i_{O/XB,T}}{Y_{H,T}}\right)\mu_H \cdot X_{B,H,T} + (1-f_D) b_H \cdot X_{B,H,T} \cdot i_{O/XB,T} \quad \text{(as COD)}. \tag{5.18}$$

All reaction rate terms have units of mass of COD/(volume·time), which is why the terms for biomass and debris are denoted as r_{XB} and r_{XD}, respectively. Also note that the COD conversion terms canceled out of Equation 5.17. These reaction rate terms will be used to derive expressions depicting the performance of the bioreactor in Figure 5.1 with biomass and debris measured in TSS units.

5.1.5 Concentrations of Soluble Substrate and Biomass

At steady state the derivative in Equation 5.4 is zero, allowing the generalized mass balance equation to be simplified:

$$F \cdot C_{AO} - F_W \cdot C_A - (F - F_W)C'_A + r_A \cdot V = 0. \tag{5.19}$$

This equation must be applied twice, once for biomass and once for soluble substrate. We will start with biomass because that equation leads to an important concept.

5.1.5.1 Mass Balance on Biomass

Application of Equation 5.19 to active heterotrophic biomass COD and substitution of the appropriate terms for biomass concentration as TSS, reveals that the terms for the influent and main effluent streams fall out because those streams contain no biomass (i.e., C_{AO} and C'_A are both zero). Furthermore, the reaction rate, r_A, for active biomass is given in Equation 5.15 and insertion of it gives:

$$-F_W \cdot X_{B,H,T} \cdot i_{O/XB,T} + (\mu_H \cdot X_{B,H,T} \cdot i_{O/XB,T} - b_H \cdot X_{B,H,T} \cdot i_{O/XB,T})V = 0. \tag{5.20}$$

Substitution and simplification yield:

$$\mu_H = \frac{1}{\Theta_c} + b_H. \tag{5.21}$$

Several things are important about this equation. Because $X_{B,H,T}$ has canceled out of it, the mass balance on active biomass does not lead to an equation for calculating its concentration. Rather, it leads to an equation that shows that at steady state the specific growth rate of the active biomass, μ_H, is determined by its rate of loss from the bioreactor, as reflected by the SRT and the loss to decay. This is an important concept. Furthermore, because the engineer can control the SRT through manipulation of the wastage rate, F_w, he/she has control over how rapidly the biomass grows. Furthermore, we saw in Section 3.2.7 that μ_H is related to the substrate concentration. This suggests that control of the bioreactor SRT also allows the engineer to control the concentration of substrate being discharged from it. Finally, because the HRT does not appear in Equation 5.21, the specific growth rate of the biomass is independent of it. The only time the HRT has any impact on the steady-state specific growth rate of biomass is when there is no biomass separator and all effluent leaves via the wastage stream, making the SRT equal to the HRT.

In order to calculate the substrate concentration in the CSTR and its effluent, the functional relationship between μ_H and S_S must be known. Since we have assumed that the substrate is non-inhibitory and is the sole rate limiting constituent, the Monod equation, as given by Equation 3.36 would be appropriate:

$$\mu_H = \hat{\mu}_H \frac{S_S}{K_S + S_S}. \tag{3.36}$$

Substitution of it for μ_H in Equation 5.21 and rearrangement gives:

$$S_S = \frac{K_S(1/\Theta_c + b_H)}{\hat{\mu}_H - (1/\Theta_c + b_H)}. \tag{5.22}$$

Thus, the mass balance on biomass in a steady-state CSTR led to an equation for calculating the soluble substrate concentration, which is a unique characteristic of bioreactors. Because μ_H is a function of only the SRT, so is the steady-state substrate concentration. Furthermore, Equation 5.22 states that the substrate concentration in a steady-state CSTR is independent of the concentration in the influent, since S_{SO} does not appear in the equation. The reason for this will become clear shortly. This has been shown to be the case for a pure culture growing on a single substrate when S_S represents the concentration of that substrate in COD units.[3] However, it should be recognized that even though this simple model did not include the reaction, the biomass will form soluble microbial products as it metabolizes the substrate, as discussed in Section 2.4.3. Since the amount of product formed will depend on the amount of substrate used, the total amount of soluble organic matter in the bioreactor and in the effluent, as measured by the COD test, will be roughly proportional to the influent COD.[1,10] This effect should be recognized and remembered, even though it is not explicitly stated in models and design equations.

The examination of Equation 5.21 reveals that as the SRT becomes very large (so that $1/\Theta_c \rightarrow 0$), the specific growth rate of the active biomass approaches the specific rate of decay. This means that in a single CSTR, substrate must always be present to drive the required growth reaction. Consequently, there is a minimum substrate concentration (S_{Smin}) that can be achieved and it is given by the limit of Equation 5.22 as Θ_c approaches infinity:

$$S_{Smin} = \frac{K_S \cdot b_H}{\hat{\mu}_H - b_H}. \tag{5.23}$$

Equation 5.23 shows that the minimum attainable substrate concentration depends on the kinetic parameters describing the biodegradation, and therefore on both the nature of the substrate undergoing biodegradation and the type of culture degrading it. If a CSTR is being considered for treatment of a wastewater to a desired concentration, then that concentration should be compared to S_{Smin}. If it is less than S_{Smin}, some other bioreactor configuration must be used because a single CSTR will be insufficient.

The maximum rate at which biomass can grow on a given substrate (μ_{Hmax}) is when the surrounding substrate concentration is equal to the concentration in the influent to the bioreactor:

$$\mu_{Hmax} = \hat{\mu}_H \frac{S_{SO}}{K_S + S_{SO}}. \tag{5.24}$$

Consequently, the minimum SRT (Θ_{cmin}) at which biomass can grow on a given influent can be calculated by setting μ_H in Equation 5.21 equal to μ_{Hmax} in Equation 5.24 and rearranging:

$$\Theta_{cmin} = \frac{K_S + S_{SO}}{S_{SO}(\hat{\mu}_H - b_H) - K_S \cdot b_H}. \tag{5.25}$$

The minimum SRT is also called the point of washout because at shorter SRTs all organisms will be washed out of the bioreactor before growth occurs. At washout, no biomass is produced and no substrate is utilized because the substrate concentration in the bioreactor (and its effluent) equals the concentration in the influent (i.e., the process has failed).

In theory, there is no minimum HRT for a CSTR because as long as biomass can be separated from the effluent and returned to the bioreactor to keep the SRT greater than Θ_{cmin}, growth can be maintained. In practice, however, it is dangerous to make the HRT less than Θ_{cmin}. If the HRT were less than Θ_{cmin} and something happened to the biomass separator so that all effluent contained biomass at a concentration equal to that in the bioreactor, the SRT would then equal the HRT and the process would fail. Furthermore, the bioreactor would be impossible to restart once failure occurred, unless biomass was added from an external source. Thus, to be safe, the HRT should be kept larger than Θ_{cmin}.

Example 5.1.5.1

A culture of microorganisms is being grown in a CSTR with a volume of 8 liters (L). The medium contains a noninhibitory concentration of m-cresol as the sole carbon and energy source, and all inorganic nutrients are provided in excess. When the concentration of m-cresol in the feed is 200 mg/L as COD, the kinetic parameters have the values shown in Table E5.1.

TABLE E5.1
Kinetic Parameters and Stoichiometric Coefficients for a Culture Growing on m-Cresol at S_{SO} = 200 mg/L as COD

Symbol	Units	Value
$\hat{\mu}_H$	hr^{-1}	0.20
K_S	mg/L as COD of m-cresol	3.5
$Y_{H,T}$	mg biomass TSS/mg m-cresol COD	0.28
$i_{O/XB,T}$	mg COD/mg TSS	1.20
b_H	hr^{-1}	0.01
f_D	mg debris/mg biomass	0.20

a. What is the maximum permissible flow rate through the bioreactor?
When the flow is maximum, the HRT will be minimum, but τ_{min} should not be less than Θ_{cmin}. Thus, calculate Θ_{cmin} with Equation 5.25:

$$\Theta_{cmin} = \frac{3.5 + 200}{200(0.20 - 0.01) - (3.5)(0.01)}$$

$$\Theta_{cmin} = 5.36\,\text{hrs.}$$

Therefore,

$$\tau_{min} = 5.36\ \text{hrs}$$

and the maximum permissible flow is

$$F = 8.0\ \text{L}/5.36\ \text{hrs} = 1.49\ \text{L/hr.}$$

b. The flow through the bioreactor is 1.0 L/hr and the wastage flow is 0.05 L/hr. What is the concentration of m-cresol (in COD units) in the effluent?
First, we must calculate the SRT using Equation 5.2:

$$\Theta_c = 8.0\ \text{L}/(0.05\ \text{L/hr}) = 160\ \text{hrs.}$$

Then we must use Equation 5.22 to find the m-cresol concentration:

$$S_S = \frac{3.5(1/160 + 0.01)}{0.20 - (1/160 + 0.01)}$$

$$S_S = 0.31\,\text{mg/L as COD.}$$

c. What is the minimum m-cresol concentration (in COD units) that can be obtained from a CSTR?
This may be determined by using Equation 5.23:

$$S_{Smin} = \frac{(3.5)(0.01)}{0.20 - 0.01}$$

$$S_S = 0.18\,\text{mg/L as COD.}$$

If a lower concentration were desired, some other bioreactor configuration would have to be used.

The suspended solids in a bioreactor receiving only soluble substrate will contain two components, active biomass, $X_{B,H,T}$, and biomass debris, $X_{D,T}$. Thus, we must be able to calculate the concentration of each.

5.1.5.2 Mass Balance on Soluble Substrate

The only source of active biomass in the system is from growth due to substrate utilization and its concentration may be calculated from a mass balance on soluble substrate (in COD units). Substituting the appropriate terms into Equation 5.19, obtaining the reaction rate term from Equation 5.17, and recalling that the concentration of soluble constituents in the effluent from the biomass separator is the same as that in the reactor gives:

$$F \cdot S_{SO} - F_W \cdot S_S - (F - F_W)S_S - (\mu_H / Y_{H,T})X_{B,H,T} \cdot V = 0. \tag{5.26}$$

Rearrangement gives:

$$X_{B,H,T} = \frac{Y_{H,T}(S_{SO} - S_S)}{\mu_H \cdot \tau}. \tag{5.27}$$

Substitution of Equation 5.21 for μ_H gives:

$$X_{B,H,T} = \frac{Y_{H,T}(S_{SO} - S_S)}{(1/\Theta_c + b_H)\tau} = \left(\frac{\Theta_c}{\tau}\right)\frac{Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}. \tag{5.28}$$

This states that the active biomass concentration depends on both the SRT and the HRT. Furthermore, rearrangement of Equation 5.28 shows that for a fixed SRT (which determines S_S), the product $X_{B,H,T} \cdot \tau$ is constant:

$$X_{B,H,T} \cdot \tau = \frac{\Theta_c \cdot Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}. \tag{5.29}$$

Alternatively, substituting Equation 4.15 for τ and rearranging yields:

$$X_{B,H,T} \cdot V = F\frac{\Theta_C \cdot Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_C}. \tag{5.30}$$

In other words, at a fixed SRT and a fixed flow rate, a fixed mass of substrate will be removed per unit time, generating a fixed mass of microorganisms. Thus, Equation 5.30 shows that a bioreactor with a small volume (short HRT) will contain a higher concentration of active biomass than one with a large volume (large HRT), although they will both contain the same mass. Likewise, if the bioreactor volume is fixed and the flow is increased at constant SRT, the mass of microorganisms in the bioreactor must increase to keep it consistent with the mass of substrate being removed. As a consequence, the biomass concentration must increase proportionally. Equation 5.29 demonstrates why the steady-state performance of a CSTR equipped with a biomass separator is independent of the HRT. If the HRT is changed for any reason, the concentration of biomass will also change to maintain the mass of organisms sufficient to produce an effluent substrate concentration consistent with the SRT. Likewise, if the influent substrate concentration, S_{SO}, is changed, the concentration of active biomass will change until the effluent substrate concentration is consistent with the SRT, thereby making S_S independent of S_{SO}.

The situation concerning the active biomass concentration is different when the bioreactor has no biomass separator so that all effluent contains biomass and the SRT is equal to the HRT. In that case, the biomass concentration depends solely on the HRT, as can be seen by substituting τ for Θ_c in Equation 5.28:

$$X_{B,H,T} = \frac{Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \tau}. \tag{5.31}$$

The major disadvantage of a steady-state CSTR without a biomass separator is that its performance depends on its HRT, and this follows directly from its inability to maintain a constant mass of microorganisms as the HRT is changed. However, S_S is still independent of S_{SO} because a change in S_{SO} will cause the active biomass concentration to change until S_S is consistent with the HRT (SRT) at steady state.

5.1.5.3 Mass Balance on Biomass Debris

The TSS-based concentration of biomass debris ($X_{D,T}$) in the bioreactor can be obtained from a mass balance on it, recognizing that its concentration in the influent and the effluent from the biomass separator is zero:

$$-F_W \cdot X_{D,T} + f_D \cdot b_H \cdot X_{B,H,T} \cdot V = 0. \tag{5.32}$$

Rearrangement gives:

$$X_{D,T} = f_D \cdot b_H \cdot \Theta_c \cdot X_{B,H,T} = \left(\frac{\Theta_c}{\tau}\right)\left[\frac{f_D \cdot b_H \cdot \Theta_c \cdot X_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right]. \tag{5.33}$$

Debris adds to the total suspended solids concentration in the bioreactor, but does not add to the degradative capability because it has no biological activity associated with it. If the bioreactor has no biomass separator, the SRT will equal the HRT and Equation 5.33 should be modified appropriately in calculating $X_{D,T}$.

5.1.5.4 Total Biomass Concentration

The total biomass concentration in a bioreactor on a TSS basis, $X_{total,T}$, is the sum of the active biomass and biomass debris concentrations. Adding Equations 5.28 and 5.33 gives:

$$X_{total,T} = \left(\frac{\Theta_c}{\tau}\right)\left[\frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right]. \tag{5.34}$$

Examination of it reveals that like the $X_{B,H,T}{\cdot}\tau$ product, the $X_{total,T}{\cdot}\tau$ product (and hence, the $X_{total,T}{\cdot}V$ product) is fixed if the SRT is fixed. Thus, as far as theory is concerned, once the SRT has been chosen to give a desired effluent substrate concentration, any combination of bioreactor size and total biomass concentration may be used as long as it gives the proper $X_{total,T}{\cdot}\tau$ product. There are practical limits, of course, and these will be discussed in Chapter 11.

As discussed above for the active biomass concentration, when a bioreactor has no biomass separator, thereby making the SRT equal to the HRT, the total biomass concentration depends solely on the HRT as can be seen by substituting τ for Θ_c in Equation 5.34:

$$X_{total,T} = \left[\frac{(1 + f_D \cdot b_H \cdot \tau)Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \tau}\right]. \tag{5.35}$$

It should be noted that the total biomass concentration can be calculated on a COD basis (X_{total}) by using Y_H instead of $Y_{H,T}$ in Equations 5.34 and 5.35. Furthermore, similar statements can be made about the concentrations of active biomass and debris.

5.1.5.5 Active Fraction

The active fraction of the biomass, f_A, is defined as the concentration of active biomass divided by the total biomass concentration. Division of Equation 5.28 by Equation 5.34 and rearrangement yields:

$$f_A = \left[\frac{1}{(1 + f_D \cdot b_H \cdot \Theta_c)}\right]. \tag{5.36}$$

Examination of it shows that the active fraction declines as the SRT is increased. This occurs because of the buildup of biomass debris in the bioreactor.

5.1.5.6 Observed Yield

As discussed in Sections 2.4.2 and 3.8, the observed yield in a biochemical operation is always less than the true growth yield because some of the energy in the substrate must go to meet the maintenance energy needs of the culture. The observed yield associated with a bioreactor is equal to the actual net mass of biomass formed per unit mass of substrate destroyed, taking into consideration the amount of biomass lost to decay. When engineers attempt to measure the concentration of biomass formed, it is difficult to distinguish the active biomass from the biomass debris. Thus, the total biomass concentration is generally used for purposes of defining the observed yield. At steady state, the total biomass formed in the bioreactor must equal the mass wasted from it. Thus, the observed yield on a TSS basis, $Y_{Hobs,T}$, is given by

$$Y_{Hobs,T} = \frac{F_w \cdot X_{total,T}}{F(S_{SO} - S_S)}. \tag{5.37}$$

Substitution of Equation 5.34 for $X_{total,T}$ and simplification gives:

$$Y_{Hobs,T} = \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T}}{1 + b_H \cdot \Theta_c}. \tag{5.38}$$

Thus, it can be seen that the larger the SRT of the bioreactor, the smaller the observed yield will be. This is because longer SRTs provide greater opportunity for biomass decay and greater need for maintenance energy, thereby leaving less energy for synthesis of new biomass.

Example 5.1.5.2

Continue with the problem begun in Example 5.1.5.1.

a. What is the active biomass concentration in the bioreactor when the influent flow is 1.0 L/hr and the wastage flow is 0.05 L/hr?
From Example 5.1.5.1, for this condition the SRT is 160 hr and S_S is 0.31 mg/L as COD. Furthermore, the HRT is 8 hr. Thus, using Equation 5.28:

$$X_{B,H,T} = \frac{0.28(200 - 0.31)}{(1/160 + 0.01)8}$$

$$X_{B,H,T} = 430\,\text{mg/L as TSS}.$$

b. What is the total biomass concentration under the same conditions?
Use Equation 5.34 to calculate this:

$$X_{total,T} = \left(\frac{160}{8}\right)\left\{\frac{[1 + (0.20)(0.01)(160)](0.28)(200 - 0.31)}{1 + (0.01)(160)}\right\}$$

$$X_{total,T} = 568\,\text{mg/L as TSS}.$$

c. What is the active fraction of the biomass?
This may be calculated from its definition or from Equation 5.36. Using the definition gives:

$$f_A = 430/568 = 0.76.$$

Using Equation 5.36 gives:

$$f_A = 1/[1 + (0.2)(0.01)(160)] = 0.76.$$

d. What is the observed yield?
This may be calculated from Equation 5.38:

$$Y_{Hobs,T} = \frac{[1+(0.20)(0.01)(160)]\,0.28}{1+(0.01)(160)}$$

$$Y_{Hobs,T} = 0.14 \text{ mg biomass TSS produced/mg substrate COD destroyed.}$$

This is only 50% of the true growth yield, showing the impact of decay and maintenance energy requirements on the net production of biomass and debris.

5.1.6 Excess Biomass Production Rate, Oxygen Requirement, and Nutrient Requirements

The two major costs associated with the treatment of wastewaters in aerobic CSTRs are from the disposal of the excess biomass produced and the provision of ample oxygen. Thus it is important to be able to determine the amount of excess biomass produced and the quantity of oxygen that must be supplied. In addition, because of the negative impacts of nutrient limitations it is important to be able to determine the nutrient requirements as well.

5.1.6.1 Excess Biomass Production Rate

Excess biomass is removed from the bioreactor via the wastage stream and the mass that must be disposed of per unit time is simply the concentration in that stream times its flow rate. At steady state, this must equal the net production rate. Letting $W_{total,T}$ represent the total biomass wastage rate on a TSS basis gives:

$$W_{total,T} = F_w \cdot X_{total,T}. \tag{5.39}$$

Combining the equation relating F_w to the SRT (Equation 5.2) with Equation 5.34 for $X_{total,T}$ gives:

$$W_{total,T} = F\left[\frac{(1+f_D \cdot b_H \cdot \Theta_c)Y_{H,T}(S_{SO}-S_S)}{1+b_H \cdot \Theta_c}\right]. \tag{5.40}$$

Since S_S depends only on the SRT, it can be seen that $W_{total,T}$ depends on the SRT, the flow rate of the wastewater, and the concentration of substrate in it. Furthermore, it can be seen that the excess biomass wastage rate will decrease as the SRT is increased. This is due to the increased importance of decay at long SRTs. It will be recalled from Section 1.2.1 that one use of biochemical operations is the stabilization of insoluble organic matter. The decrease in the amount of excess biomass brought about by decay is one example of stabilization. As the SRT is increased, more and more of the active biomass is oxidized and converted to debris, meaning that less excess biomass must be disposed of.

Comparison of Equation 5.40 with Equation 5.38 and substitution of the latter into the first reveals that the excess biomass wastage rate is just the observed yield times the substrate removed, which is consistent with the fact that the mass of biomass wasted must equal the mass produced at steady state:

$$W_{total,T} = F \cdot Y_{Hobs,T}(S_{SO}-S_S). \tag{5.41}$$

Thus it can be seen that knowledge of the observed yield makes it easy to estimate the amount of excess biomass that must be disposed of.

5.1.6.2 Oxygen Requirement

The rate at which the microorganisms utilize oxygen in the bioreactor is equal to the overall rate expression for oxygen as developed from Table 5.2 and expressed in Equation 5.18 as oxygen demand. Consequently, oxygen must be supplied at the same rate. If we multiply the rate of oxygen demand (i.e., utilization) by the bioreactor volume to give the mass per unit time required (RO), the result is

$$RO = \left[\left(\frac{1 - Y_{H,T} \cdot i_{O/XB,T}}{Y_{H,T} \cdot i_{O/XB,T}}\right)\mu_H + (1 - f_D)b_H\right] X_{B,H,T} \cdot i_{O/XB,T} \cdot V. \tag{5.42}$$

Substitution of Equation 5.21 for μ_H, Equation 5.28 for $X_{B,H,T}$, Equation 4.15 for the HRT, and simplification gives:

$$RO = F(S_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T} \cdot i_{O/XB,T}}{1 + b_H \cdot \Theta_c}\right]. \tag{5.43}$$

Because the stoichiometric coefficients in Table 5.2 are from a COD balance for each reaction, Equation 5.43 simply represents a COD balance across the bioreactor. This can be seen in the following way. A COD balance states that the amount of oxygen that must be supplied to a bioreactor must equal the total COD in minus the total COD out, including the COD of the biomass and the biomass debris:

$$RO = F \cdot S_{SO} - F \cdot S_S - F_w \cdot X_{total,T} \cdot i_{O/XB,T}. \tag{5.44}$$

Examination of Equation 5.44 reveals that the last term is just $W_{total,T}$, the excess biomass wastage rate, converted to COD units by the use of $i_{O/XB,T}$. Substitution of Equation 5.40 for $W_{total,T}$ yields an equation identical to Equation 5.41 after rearrangement. Furthermore, substitution of Equation 5.41 for $W_{total,T}$ gives an equation for the oxygen requirement in terms of only the observed yield and the amount of substrate removed:

$$RO = F(S_{SO} - S_S)(1 - Y_{Hobs,T} \cdot i_{O/XB,T}). \tag{5.45}$$

Substitution of Equation 5.38 for $Y_{Hobs,T}$ also gives an expression identical to Equation 5.43. Thus, if the observed yield is known, the oxygen requirement is also known.

All of the above follows directly from the COD-based stoichiometry discussed in Sections 3.1.1 and 3.8.1. Equation 3.94 stated that the COD removed equals the oxygen equivalents of terminal electron acceptor used plus the COD of the biomass formed, which is the same as Equation 5.44. Thus, all of the above equations have their roots in the basic stoichiometry discussed earlier. The ability to calculate the steady-state oxygen requirement directly from such simple equations is the key advantage to using COD-based stoichiometry.

Equation 5.43 shows clearly that the oxygen requirement in a CSTR increases as the SRT is increased. This, too, is indicative of the increased stabilization that occurs as the SRT is increased. Increased stabilization implies that more of the electrons in a material end up being transferred to the terminal electron acceptor. Thus, as less excess biomass is produced, more oxygen must be used.

5.1.6.3 Nutrient Requirement

The amount of nutrient required can also be determined directly from the stoichiometry of biomass growth as discussed in Section 3.8.2. There it was seen that the amount of nitrogen required to

form active biomass that can be represented by the empirical formula $C_5H_7O_2N$ is 0.087 mg N/mg biomass COD. If we assume that the nitrogen content of biomass debris is the same as that of active biomass, then the amount of nitrogen required per unit of substrate COD removed, NR, is just 0.087 times the observed yield, or

$$NR = 0.087 \cdot Y_{Hobs,T} i_{O/XB,T}. \tag{5.46}$$

If we accept that $i_{O/XB,T}$ has a value of 1.20 mg COD/mg TSS, Equation 5.46 can be simplified to

$$NR = 0.10 \cdot Y_{Hobs,T}. \tag{5.47}$$

In other words, if the organic portion of biomass can be represented by $C_5H_7O_2N$ and biomass contains 15% ash, the nitrogen requirement is 0.10 mg N/mg biomass TSS formed. Furthermore, as seen earlier, the phosphorus requirement is about one-fifth of the nitrogen requirement on a mass basis and thus it may be calculated by replacing 0.087 in Equation 5.46 with 0.017 mg P/mg biomass COD or the 0.10 in Equation 5.47 with 0.02 mg P/mg biomass TSS. The requirements for micronutrients may be determined in a similar manner by using appropriate factors from Table 3.3. Nutrients should be added in slight excess of the theoretical amounts to ensure that the organic substrate is rate limiting, as discussed in Section 3.2.9.

Example 5.1.6.1

Continue with the problem begun in Example 5.1.4.1.

a. How many mg/hr of dry solids would have to be disposed of when the flow through the bioreactor is 1.0 L/hr and the wastage rate is 0.05 L/hr?
Using Equation 5.40:

$$W_{total,T} = 1.0\left\{\frac{[1+(0.20)(0.01)(160)](0.28)(200-0.31)}{1+(0.01)(160)}\right\}$$

$$W_{total,T} = 28.4 \text{ mg/hr as dry solids.}$$

b. How many mg/hr of oxygen must be supplied to the bioreactor?
This can be obtained from the use of Equation 5.44:

$$RO = (1.0)(200) - (1.0)(0.31) - (28.4)(1.20) = 166 \text{ mg/hr.}$$

It can also be obtained from the fundamental system parameters by using Equation 5.43:

$$RO = (1.0)(200-0.31)\left\{1-\frac{[1+(0.20)(0.01)(160)](0.28)(1.2)}{1+(0.01)(160)}\right\}$$

$$RO = 166 \text{ mg/hr.}$$

Finally, it can be obtained directly from the observed yield through the use of Equation 5.45:

$$RO = 1.0(200 - 0.31)[1.0 - ((0.14)(1.2))]$$

$$RO = 166 \text{ mg/hr.}$$

Since the oxygen demand associated with the influent substrate is 1 L/hr × 200 mg/L = 200 mg/hr, these calculations show that it is necessary to supply sufficient oxygen to meet 83% of that demand. The remainder is associated with the excess biomass formed and wasted from the system.

c. How many mg/L of nitrogen and phosphorus should the influent contain?
The nitrogen requirement can be calculated with Equation 5.47:

$$NR = (0.10)(0.14)$$

$$NR = 0.014 \text{ mg N/mg substrate COD removed.}$$

Since the substrate COD removed was 200 – 0.31 = 199.69 mg/L, the biomass will require 2.8 mg/L of nitrogen. If we allow an extra 0.5 mg/L to prevent nitrogen from being rate limiting, the influent should contain approximately 3.3 mg/L as N.

The phosphorus requirement will be about one-fifth of the nitrogen requirement. Thus, the biomass will use about 0.56 mg/L. If we allow an extra 0.25 mg/L to prevent phosphorus from being rate limiting, the influent should contain approximately 0.81 mg/L as P.

5.1.7 Process Loading Factor or F/M Ratio

Before the widespread use of SRT as the basic independent variable for design and control of a CSTR, most designers used the process loading factor, also called the food to microorganism (or F/M) ratio. The process loading factor, U, is defined as the mass of substrate applied per unit time divided by the mass of microorganisms contained in the bioreactor.[11] Because it is difficult to distinguish active biomass from biomass debris, the mass of microorganisms has generally been defined in terms of the total biomass concentration, $X_{total,T}$:

$$U \equiv \frac{F \cdot S_{SO}}{V \cdot X_{total,T}}. \tag{5.48}$$

The SRT is an important design parameter because at steady state it is related in a simple way to μ_H, the specific growth rate coefficient of the biomass, which in turn controls the substrate concentration in the bioreactor and its effluent. The relationship of the process loading factor to μ_H can be obtained by rearranging the mass balance on substrate (Equation 5.26) and substituting into it the fact that the active biomass concentration is equal to the total biomass concentration times the active fraction, giving:

$$\frac{f_A \cdot \mu_H}{Y_{H,T}} = \frac{F(S_{SO} - S_S)}{V \cdot X_{total,T}}. \tag{5.49}$$

For the conditions generally found in bioreactors used in wastewater treatment, $S_S \ll S_{SO}$. Therefore:

$$U \approx \frac{f_A \cdot \mu_H}{Y_{H,T}} = f_A \cdot q_{H,T}. \tag{5.50}$$

This shows that the active fraction must be known before the specific growth rate or the specific substrate removal rate of the biomass can be determined exactly from the process loading factor. Conversely, Equation 5.21 showed that at steady state the specific growth rate can be determined directly from the SRT without requiring such knowledge. Since the active fraction depends on the characteristics of the bioreactor and requires knowledge of the SRT (Equation 5.36), it is

simpler to work directly with SRT as the fundamental design and operational parameter for a CSTR at steady state. Furthermore, when a wastewater contains particulate substrates, the active fraction cannot be calculated easily from an expression like Equation 5.36 and thus its determination becomes a major problem. Consequently, the relationship of the process loading factor to the specific growth rate of the biomass is more difficult to determine than it is for the simplified situation under consideration here. This has led to the use of the SRT instead of the process loading factor by many designers. Nevertheless, because the process loading factor is related to the specific growth and substrate removal rates, there are situations in which it provides valuable information, particularly in tanks-in-series systems where the applied substrate varies from tank to tank but the active fraction does not. In Chapters 11 and 12 we will consider such situations.

5.1.8 First-Order Approximation

As discussed in Section 3.2.7, occasions arise in which the steady-state substrate concentration in a CSTR is much less than the half-saturation coefficient, K_S, so that the Monod equation may be simplified into a first-order equation:

$$\mu_H \approx \left(\frac{\hat{\mu}_H}{K_S}\right) S_S. \tag{3.38}$$

This situation often arises when the SRT is long, thereby making it difficult to evaluate the maximum specific growth rate, $\hat{\mu}_H$, and K_S independently from steady-state data. In that case, the mean reaction rate coefficient, k_e, is often used. As defined in Section 3.2.8, k_e has units of L/(mg biomass COD·hr). Because we are now using TSS as the unit system for measuring biomass, it would be convenient to define a new mean reaction rate coefficient, $k_{e,T}$, with units of L/(mg biomass TSS·hr). The relationship between k_e and $k_{e,T}$ is given by

$$k_e = \frac{k_{e,T}}{i_{O/XB,T}}. \tag{5.51}$$

Recalling Equations 3.44 and 3.45, substituting them into Equation 3.38 along with Equation 5.51 shows that:

$$\mu_H = Y_{H,T} \cdot k_{e,T} \cdot S_S. \tag{5.52}$$

The use of Equation 5.52 does not alter any of the mass balances nor does it alter the fact that μ_H is controlled by the SRT as expressed in Equation 5.21. All that it does is simplify the relationship between S_S and the SRT. Substitution of Equation 5.52 into Equation 5.21 and rearrangement give:

$$S_S = \frac{1/\Theta_c + b_H}{Y_{H,T} \cdot k_{e,T}}. \tag{5.53}$$

Comparison of Equation 5.53 with Equation 5.22 shows the effect that the first-order approximation has. None of the other equations are affected, except through the effect on S_S. It should be emphasized that care should be exercised in the application of the first-order approximation of the Monod equation to ensure that the basic assumption (i.e., $S_S \ll K_S$) is valid.

5.1.9 Effect of Solids Retention Time on the Performance of a Continuous Stirred Tank Reactor as Predicted by Model

The major value of the model in this chapter is as an aid to understanding how CSTRs behave under a variety of conditions. We saw during derivation of the equations that the most important operational variable is the SRT. Consequently, we will use graphs generated with the parameter values in Table 5.3 to see how SRT influences bioreactor performance. Furthermore, although the equations were developed with biomass and yield represented on a TSS basis, we convert these variables to a COD basis when performing simulations to allow a COD balance to be achieved and to simplify discussion of that balance when interpreting the simulation results.

We will see in Chapter 11 that the use of sedimentation as the means of biomass separation places both upper and lower limits on the SRT allowable in an operating bioreactor. However, to fully demonstrate the potential impact of SRT on bioreactor performance, the figures presented here were generated without regard to those limits. Consequently, most operating systems will not experience the broad range of conditions exhibited in the figures.

Figure 5.3 shows the effect of SRT on the soluble substrate concentration. For the parameter values in Table 5.3, the minimum SRT is 2.2 hours, and at that SRT the effluent substrate concentration is equal to the influent concentration. This SRT is also called the washout point for the bioreactor because biomass can no longer exist in it. As the SRT is increased, however, growth can be established in the bioreactor and substantial substrate removal occurs even when the SRT is low. For example, an SRT of slightly more than four hours is all that is required to reduce the substrate concentration from 500 to 50 mg/L as COD. This demonstrates an important characteristic of biological reactors: they are able to achieve substantial removal of soluble substrate at very short SRTs. However, incremental removal of substrate declines sharply as the SRT is increased, although such increases make the bioreactor more stable. For example, compare the differences in substrate concentration resulting from a 10% change in SRT at SRTs of 4, 40, and 400 hours. The minimum attainable substrate concentration (S_{Smin}) for the parameter values used to generate Figure 5.3 is 0.76 mg/L as COD, and examination of the graph shows that this value is approached very slowly. When methods like COD and five-day biochemical oxygen demand are used to measure effluent substrate concentration on full-scale systems, the SRT generally has little measurable effect beyond certain values. That is simply because the potential change is small compared to the error associated with the test method.

TABLE 5.3
Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used to Generate Figures Demonstrating CSTR Performance

Symbol	Units	Value
$\hat{\mu}_H$	hr^{-1}	0.50
K_S	mg/L as COD	50
$Y_{H,T}$	mg biomass TSS formed/mg COD removed	0.50
b_H	hr^{-1}	0.0075
f_D	mg debris COD/mg biomass COD	0.20
$i_{O/XB,T}$	mg COD/mg TSS	1.20
F	L/hr	1.00
S_{SO}	mg/L as COD	500
X_{IO}	mg/L as COD	0[a]
$X_{B,HO}$	mg/L as COD	0[a]

[a] Unless specified otherwise.

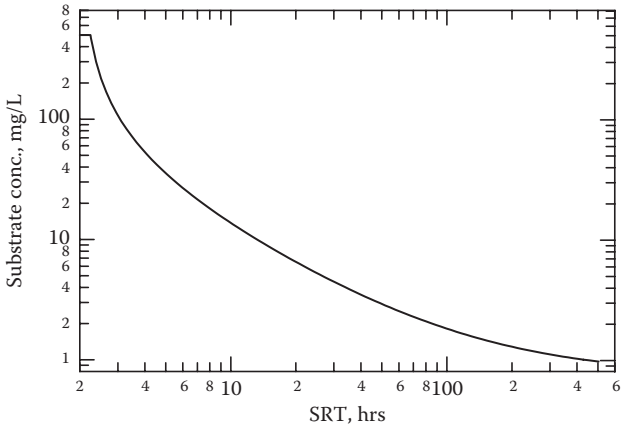

FIGURE 5.3 Effect of SRT on the concentration of soluble substrate (as COD) in a CSTR receiving a soluble substrate. Kinetic parameters and stoichiometric coefficients are listed in Table 5.3.

The dashed curve in Figure 5.4 shows the impact of SRT on the total mass of biomass (as COD) in the CSTR. It will be recalled from Equations 5.22 and 5.34 that the biomass concentration depends on the HRT, whereas the substrate concentration is independent of it. Thus, the HRT must be considered to convert the mass values in Figure 5.4 into concentrations. However, since the influent flow rate was fixed for the simulations, as shown in Table 5.3, consideration of different HRTs is equivalent to consideration of different bioreactor volumes. Thus, the total biomass concentration in a CSTR with a volume of V liters being operated at a particular SRT can be obtained by dividing the mass associated with that SRT by V. For example, when the SRT is 100 hours, the bioreactor contains 20 g COD (17 g as TSS) of total biomass (active plus debris). Thus, if the reactor volume were 2 liters the concentration would be 10 g/L as COD (8.3 g/L as TSS), whereas if the volume were 4 liters the concentration would be 5 g/L as COD (4.2 g/L as TSS). The dashed curve is important because it demonstrates clearly that the mass, and thus the concentration, of biomass in a reactor of volume V increases as the SRT is increased. It is this increase in the mass of organisms that allows more substrate to be removed as the SRT is increased, even though the HRT is kept the same; more biomass can accomplish more in the same available time. A fixed mass of organisms is

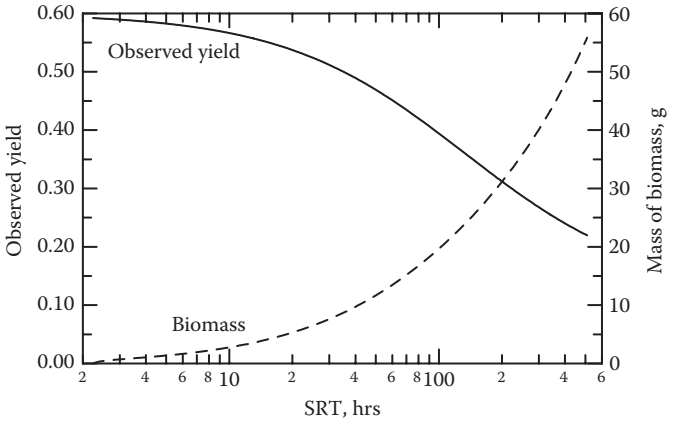

FIGURE 5.4 Effect of SRT on the observed yield (mg biomass COD/mg substrate COD) and the total mass of biomass (as COD) in a CSTR receiving a soluble substrate. Kinetic parameters and stoichiometric coefficients are listed in Table 5.3. To express the observed yield and the mass of biomass in TSS units, divide by $i_{O/XB,T}$ = 1.2 mg COD/mg TSS.

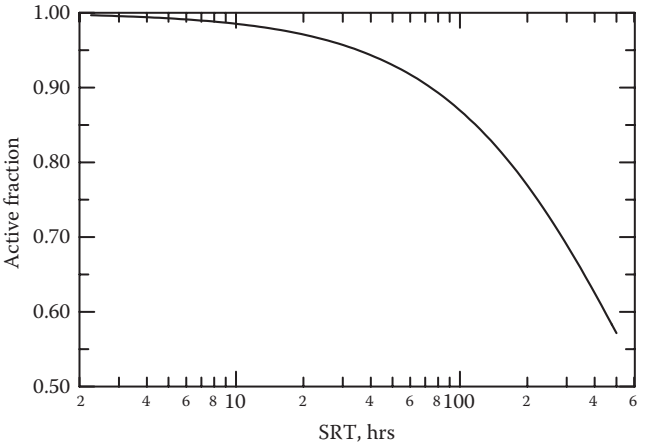

FIGURE 5.5 Effect of SRT on the active fraction (COD basis) of the biomass in a CSTR receiving a soluble substrate. Kinetic parameters and stoichiometric coefficients are listed in Table 5.3.

required to accomplish a given amount of substrate removal; only the concentration is influenced by the bioreactor volume.

The solid curve in Figure 5.4 is important because it illustrates the decrease in the observed yield that occurs as the SRT of a bioreactor is increased. This is because of the increased importance of decay as the SRT is increased, as indicated in Equation 5.38. Only when the SRT is very short and the biomass is growing very rapidly will most substrate utilization go for growth, allowing the observed yield to approach the true growth yield, which was 0.60 mg biomass COD/mg substrate COD (0.5 mg biomass TSS/mg substrate COD) for this case. For all other situations a significant amount of energy must be expended for maintenance and other purposes associated with decay, thereby lowering the observed yield.

Figure 5.4 presents the total mass of biomass, but we know that biomass decay will decrease the active fraction through the buildup of biomass debris as the SRT is increased. Figure 5.5 shows that effect. At low SRT, when the specific growth rate is high, the impact of decay will be small so that little debris will be generated, making the active fraction large. As the SRT is increased, however, the buildup of biomass debris in the bioreactor becomes significant and the active fraction drops, until at high SRTs only a small percentage of the "biomass" is actually contributing to substrate removal. Nevertheless, the amount of active biomass continues to increase as the SRT is increased, as can be seen by multiplying the active fraction from Figure 5.5 times the mass of biomass from Figure 5.4. As a consequence, increases in SRT are generally worthwhile, although a point of diminishing return will be reached. One price associated with a high SRT is the cost of moving inactive biomass continually around the system, yet little can be done about it because it is impossible to separate active biomass from debris. In fact, it is possible that accumulated debris contributes to improved settling properties at increased SRT.

A benefit associated with increased SRTs is that less excess biomass must be disposed of because more of it is oxidized through decay, maintenance energy needs, and so on as discussed above for the observed yield. This effect is illustrated in Figure 5.6. When the SRT is small, even though the observed yield is high, substrate removal is incomplete, as shown in Figure 5.3, and thus little excess biomass is synthesized. However, as the SRT is increased beyond the minimum SRT, more excess biomass is generated because of increased growth and substrate removal, all with a relatively high observed yield. As the SRT is increased further, the effluent substrate concentration becomes small relative to the influent concentration so that the term $S_{SO} - S_S$ becomes essentially constant. This occurs at an SRT of about 10 hours for the parameter values used to generate the graphs. Beyond that point, further increases in SRT increase the importance of decay, causing the observed yield and the

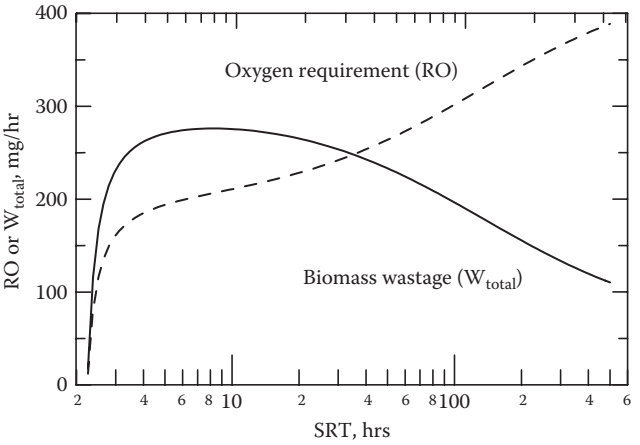

FIGURE 5.6 Effect of SRT on the biomass wastage rate (as COD) and oxygen requirement in a CSTR receiving a soluble substrate. Kinetic parameters and stoichiometric coefficients are listed in Table 5.3. To express the biomass wastage rate in TSS units, divide by $i_{O/XB,T}$ = 1.2 mg COD/mg TSS.

net production of biomass to decline as shown by the decreasing mass that must be wasted. Those same events determine the shape of the oxygen consumption curve. At very short SRTs, substrate removal is incomplete and decay is of little importance, with the result that most of the electrons available in the influent substrate are either associated with the effluent substrate or the biomass formed. Thus, relatively little oxygen is required. As the SRT is increased slightly to the point where substrate removal is essentially complete (SRT ≈ 10 hr) but decay is not yet important, all of the oxygen use is associated with substrate removal and biomass growth (i.e., energy for synthesis). Further increases in the SRT result in more decay, and almost the entire increase in oxygen requirement as the SRT is increased past 10 hours is associated with that. In other words, the decreased mass of excess biomass associated with longer SRTs is at the expense of an increased oxygen requirement. This suggests that the choice of SRT is often governed by the relative costs of supplying oxygen versus disposing of excess sludge.

Although Figures 5.3–5.6 were developed with parameter values representative of aerobic growth of heterotrophic bacteria, it is important to recognize that the shapes of the curves are representative of microbial growth in general, including aerobic growth of autotrophs and anoxic growth of heterotrophs. All that is necessary is an appropriate change in electron donor or acceptor and adjustment of the parameter values.

5.2 EXTENSIONS OF THE BASIC MODEL

The simple model developed in Section 5.1 was for a system receiving only soluble substrate. However, most wastewaters contain soluble organic matter that is nonbiodegradable. Furthermore, all domestic, and many industrial, wastewaters contain suspended matter that escapes removal by sedimentation prior to entrance of the wastewater into the biochemical operation, and the impacts of those solids must be accounted for in any models depicting fully the operation of CSTRs.

Suspended material may be classified in many ways, and one of the most commonly used methods is to split it into organic and inorganic, which have traditionally been measured as volatile and fixed suspended solids (FSS), respectively. Unfortunately, this division is not the best for describing biochemical operations, in part because around 40% of the volatile suspended solids in domestic wastewater are nonbiodegradable, and therefore inert to biological attack.[9] A more appropriate division would be inert, biodegradable, and biomass, because each influences biochemical operations in a different way. By definition, biodegradable suspended matter and biomass are both organic. Inert

suspended matter may be either organic or inorganic and, therefore, includes both nonbiodegradable VSS and FSS, which is just the difference between the TSS and VSS concentrations. In keeping with the convention adopted earlier in this chapter and considering the fact that both organic and inorganic suspended solids contribute to the total concentration of TSS that must be handled in the treatment system, the concentrations of all particulate constituents will be considered here and reported in TSS units. It should be noted that the conversion factor from mass units to COD units will depend on the nature of the suspended matter and may well be different for each type. Even though a conversion factor of 1.20 g COD per g of dry solids has been adopted herein for biomass and microbial debris, it is impossible to generalize about the value for other particulate constituents. They must be determined on a case by case basis.

5.2.1 Soluble, Nonbiodegradable Organic Matter in Influent

Soluble, nonbiodegradable organic matter will not be acted on by the biomass in a biochemical operation, although its concentration can be influenced by physical/chemical phenomena such as adsorption and volatilization. If the material is not adsorbed onto biomass and if it is not volatilized, then its concentration in the effluent from a bioreactor, S_I, will be the same as its concentration in the influent, S_{IO}:

$$S_I = S_{IO}. \tag{5.54}$$

If the material is volatile and/or adsorbable, appropriate reaction terms must be inserted into the mass balance equation, yielding an equation for the effluent concentration. Such expressions will be discussed in Chapter 22 where the fate of synthetic organic chemicals will be considered. The total concentration of soluble organic matter will be the sum of S_I and S_S, plus any soluble microbial products, as discussed earlier. Except in special cases, the presence of soluble, nonbiodegradable organic matter will be ignored in this book since it has no impact on the systems to be discussed.

5.2.2 Inert Suspended Solids in Influent

Like soluble, nonbiodegradable organic matter, inert suspended solids undergo no reaction in a biological reactor. Consequently, the mass leaving the bioreactor must equal the mass entering it if steady state is to be achieved. Unlike soluble, nonbiodegradable organic matter, however, the concentration of inert suspended solids in the bioreactor depends on the magnitude of the SRT relative to the HRT, as can be seen by performing a mass balance across the bioreactor. This follows from the fact that the suspended solids only leave the bioreactor through the wastage stream. Reference to Figure 5.1 and construction of the mass balance in which $X_{IO,T}$ and $X_{I,T}$ represent the concentrations of inert suspended solids in TSS units in the influent and bioreactor, respectively, yields:

$$F \cdot X_{IO,T} - F_w \cdot X_{I,T} = 0. \tag{5.55}$$

or

$$X_{I,T} = \left(\frac{F}{F_w}\right) X_{I,TO}. \tag{5.56}$$

Division of both F and F_w by V and invocation of the definitions of HRT and SRT reveals:

$$X_{I,T} = \left(\frac{\Theta_c}{\tau}\right) X_{I,TO}. \tag{5.57}$$

Thus, the concentration of inert suspended solids in the bioreactor is greater than the concentration in the influent, with the concentration factor being Θ_c/τ. If no effluent is removed through the biomass separator so that all leaves through the wastage flow, then the SRT and HRT are the same and the concentration of inert suspended solids in the bioreactor is the same as the concentration in the influent. Note that when inert suspended solids are represented in COD units, reflecting only the organic fraction, the designations for inert suspended solids and influent inert suspended solids are X_I and X_{IO}, respectively, and Equation 5.57 is still applicable after appropriate substitution.

If a treatment system receives inert suspended solids, the suspended solids in the bioreactor will include them in addition to active biomass and biomass debris. This combination of suspended solids is called mixed liquor suspended solids (MLSS) and will be given the symbol $X_{M,T}$ when expressed on a TSS basis. The concentration of MLSS is the sum of $X_{I,T}$, as given by Equation 5.57, and $X_{total,T}$, as given by Equation 5.34:

$$X_{M,T} = \left(\frac{\Theta_c}{\tau}\right)\left[X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right]. \tag{5.58}$$

Looking at the bracketed term it can be seen that the contribution of biomass related solids (right term) decreases as the SRT is increased, whereas the contribution of inert suspended solids (left term) does not. As a consequence, the percentage of active biomass and biomass debris in the MLSS decreases as the SRT is increased. Note that the MLSS can be calculated on a COD basis (X_M) if X_{IO} and Y_H are used in Equation 5.58 instead of $X_{IO,T}$ and $Y_{H,T}$, respectively.

The active fraction of the MLSS is the active biomass concentration divided by the MLSS concentration. Use of the appropriate equations in this definition gives:

$$f_A = \frac{1}{1 + f_D \cdot b_H \cdot \Theta_c + X_{IO,T}\left[\dfrac{(1 + b_H \cdot \Theta_c)}{Y_{H,T}(S_{SO} - S_S)}\right]}. \tag{5.59}$$

This reduces to Equation 5.36 when $X_{IO,T}$ is zero. As one might expect, it tells us that the active fraction will be smaller the larger $X_{IO,T}$ is relative to S_{SO}. It also tells us that the active fraction is not affected by the SRT to HRT ratio, even though the MLSS concentration is. Thus, maintenance of a high active fraction requires minimization of the amount of inert suspended solids entering a biological reactor.

In Section 5.1.7, it was seen that the process loading factor (F/M ratio) is less convenient than the SRT as a design and control parameter for CSTRs at steady state because of the necessity of knowing the active fraction of biomass in the MLSS before determining the microbial specific growth rate. This is particularly true when a wastewater contains inert suspended solids because of the impact of those solids on the active fraction, as defined in Equation 5.59.

As discussed in Section 5.1.5, the observed yield is defined as the mass of biomass formed per unit mass of substrate removed. As such, the presence of inert suspended solids has no impact on it and it is still given by Equation 5.38.

The mass rate at which solids must be disposed of will be increased by the presence of inert suspended solids. Since nothing happens in the biochemical operation to reduce the amount of inert suspended solids present, the mass rate of solids disposal will just be increased by the rate at which inert suspended solids enter the system. Letting $W_{M,T}$ represent the mass wastage rate of MLSS in TSS units, often referred to as the solids wastage rate, gives:

$$W_{M,T} = F \cdot X_{IO,T} + W_{total,T}, \tag{5.60}$$

$$W_{M,T} = F\left[X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right]. \quad (5.61)$$

Since nothing happens to inert suspended solids in the bioreactor, they will have no impact on the oxygen requirement. Thus, it is still given by Equations 5.42–5.45.

Example 5.2.2.1

Continue with the problem begun in Example 5.1.5.1.

a. What will be the MLSS concentration in the bioreactor expressed as TSS if the influent contains 21 mg/L as TSS of inert suspended solids.
From the previous examples, we know that $\tau = 8$ hr, $\Theta_c = 160$ hr, $S_{SO} = 200$ mg/L as COD, and $S_S = 0.31$ mg/L as COD. Insertion of these into Equation 5.58 gives:

$$X_{M,T} = \left(\frac{160}{8}\right)\left\{21 + \frac{[1 + (0.20)(0.01)(160)](0.28)(200 - 0.31)}{1 + (0.01)(160)}\right\}$$

$X_{M,T} = 988$ mg/L as TSS.

Comparison of this value to the total biomass concentration in Example 5.1.5.2 shows that the inert suspended solids increased the total suspended solids concentration by 420 mg/L as TSS. This could also have been determined through the application of Equation 5.57:

$$X_{I,T} = (160/8)(21) = 420 \text{ mg/L as TSS.}$$

b. For the situation in part a, what will be the MLSS concentration in COD units if 80% of the inert suspended solids are organic and the nature of these solids is such that they have a COD of 1.0 g COD/g TSS?
From part a we can determine that the total biomass concentration is 568 mg/L as TSS (988 – 420). This can be verified from part b of Example 5.1.5.2. Since the COD conversion factor for biomass is 1.2 mg COD/mg TSS, the total biomass concentration in COD units is 682 mg/L as COD. Above, we calculated that the inert suspended solids concentration is 420 mg/L as TSS, which includes both inorganic and organic inert suspended solids. Because the inert suspended solids are 80% organic, the inert organic suspended solids concentration in TSS units is

$$X_{I,T} = (420)(0.8) = 336 \text{ mg/L as TSS.}$$

Furthermore, because the COD of the inert organic suspended solids is 1.0 g COD/g TSS, their concentration in the bioreactor in COD units is 336 mg/L as COD. Thus, the MLSS concentration, expressed as COD, is

$$X_M = 682 + 336 = 1018 \text{ mg/L as COD.}$$

Note that it was necessary to apply the conversion to each type of suspended solids separately because each has a different value of COD/TSS.

c. What fraction of the MLSS is made up of active biomass?
When the MLSS contains inert organic suspended solids that have a unit COD different from that of biomass, the answer to this question depends on the way in which the concentration is measured. In Example 5.1.5.2 the active biomass concentration was found to be

430 mg/L as TSS. Because the MLSS concentration is 998 mg/L as TSS (from part a of this example), the active fraction is

$$f_A = 430/998 = 0.43.$$

The same value may also be obtained using Equation 5.59. Since the unit COD of active biomass is 1.2 mg COD/mg TSS, its concentration in COD units is 516 mg/L as COD. From part b of this example we found the COD of the MLSS to be 1018 mg/L. Thus the active fraction on a COD basis is

$$f_A = 516/1018 = 0.51.$$

The dependence of the active fraction on the unit system used to measure the components makes it important to specify that unit system. Regardless of the unit system, however, it is clear that the presence of an apparently insignificant concentration of inert suspended solids in the influent to a bioreactor can have a significant impact on the active fraction of the MLSS, in this case decreasing it from 0.76 to 0.43 (when the concentrations are expressed as TSS).

d. What is the solids wastage rate?
Again, the answer to this question depends on the unit system used. Using Equation 5.60 with $W_{total,T}$ in TSS units as given in Example 5.1.6.1:

$$W_{M,T} = (1.0)(21) + 28.4 = 49.4 \text{ mg/hr as dry solids.}$$

Using the same equation with X_{IO} and W_{total} in COD units and recalling that only 80% of the total inert suspended solids are organic:

$$W_M = (1.0)[(1.0)(0.80)(21)] + (1.2)(28.4) = 50.9 \text{ mg/hr as COD.}$$

The same values can be obtained from application of Equation 5.61 provided X_{IO} and Y_H are expressed in the appropriate units. The significant point, however, is that the addition of 21 mg/hr as TSS of inert suspended solids to the bioreactor caused an increase of 21 mg/hr as TSS in the amount of solids wasted.

5.2.3 Biomass in Influent

The effect of the presence of active biomass in the influent to a CSTR can be seen by performing a new mass balance on biomass using Equation 5.19, and including influent biomass at concentration $X_{B,H,TO}$. Performance of the steps that led to Equation 5.21 gives:

$$\mu_H = \frac{1}{\Theta_c} + b_H - \frac{X_{B,H,TO}}{\tau \cdot X_{B,H,T}}, \tag{5.62}$$

which reduces to Equation 5.21 when $X_{B,H,TO}$ is zero. It is important to recognize that by definition, the SRT is given by Equation 5.1. The presence of active biomass in the influent to the bioreactor does not change that definition. Rather, Equation 5.62 shows that the presence of active biomass in the influent reduces the specific growth rate of the biomass in the bioreactor relative to the SRT and that greater amounts reduce it more. In other words, if biomass is present in the influent, the biomass in the reactor does not have to grow as fast to maintain itself as it does when the influent contains no biomass. This means that if two bioreactors have the same HRT and SRT but one receives active biomass in the influent, it will produce an effluent with a lower substrate concentration.

The effluent substrate concentration cannot be found by substituting the equation for μ_H into the Monod equation as was done before because the result will contain $X_{B,H,T}$, which is unknown. Thus, we must use a different approach. The mass balance on substrate is unchanged, and thus Equation 5.27 is still valid. Substitution of Equation 5.62 for μ_H into it gives:

$$X_{B,H,T} = \left(\frac{\Theta_c}{\tau}\right)\left[\frac{X_{B,H,TO}}{1 + b_H \cdot \Theta_c} + \frac{Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right]. \tag{5.63}$$

Comparison of Equation 5.63 with Equation 5.29 reveals that the active biomass concentration will be higher than in a bioreactor receiving no biomass in the influent. Furthermore, it can be seen that the term in the bracket has been divided into two components. The one on the right is the contribution of new growth to the active biomass, whereas the term on the left is the contribution of the influent biomass. Note that the latter is less than the input biomass concentration because of decay. Substitution of Equation 5.63 into Equation 5.62, with subsequent substitution of the resulting equation into Equation 3.36 yields a quadratic equation for the substrate concentration:

$$\begin{aligned}&[\hat{\mu}_H - (1/\Theta_c + b_H)]S_S^2 \\ &- [\hat{\mu}_H(X_{B,H,TO}/Y_{H,T} + S_{SO}) + (K_S - S_{SO})(1/\Theta_c + b_H)]S_S \\ &+ S_{SO} \cdot K_S(1/\Theta_c + b_H) = 0.\end{aligned} \tag{5.64}$$

The easiest way to use the equations is to calculate the substrate concentration first, and then use that result to calculate the biomass concentration.

Any biomass debris in the influent will behave as inert suspended solids, thereby making its concentration in the bioreactor greater than that in the influent by a factor equal to the ratio of the SRT to HRT. In addition, biomass debris will be generated from decay of the active biomass in the influent as well as by decay of the active biomass formed in the bioreactor. Consequently, the total biomass debris concentration on a TSS basis will be

$$X_{D,T} = \left(\frac{\Theta_c}{\tau}\right)\left[X_{D,TO} + \frac{f_D \cdot b_H \cdot \Theta_c \cdot X_{B,H,TO}}{1 + b_H \cdot \Theta_c} + \frac{f_D \cdot b_H \cdot \Theta_c \cdot Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right], \tag{5.65}$$

where $X_{D,TO}$ is the concentration of biomass debris (in TSS units) in the influent. The first term in the brackets is the contribution of biomass debris in the influent, the second term represents the debris formed through decay of the active biomass in the influent, and the last term is the formation of biomass debris from growth and decay of new biomass in the bioreactor.

The MLSS concentration was defined in Equation 5.58 as the combination of inert suspended solids and biomass. Because biomass debris in the influent behaves in the same way as inert suspended solids (and, in fact, would be difficult to distinguish from it) the definition of MLSS can be extended to include influent biomass debris:

$$X_{M,T} = \left(\frac{\Theta_c}{\tau}\right)\left[X_{D,TO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)X_{B,H,TO}}{1 + b_H \cdot \Theta_c} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right]. \tag{5.66}$$

The first term in the brackets is the contribution of influent biomass debris, the second term is the contribution of the active biomass in the influent, and the third is the contribution of new biomass growth on the soluble substrate. Note that Equations 5.63, 5.65, and 5.66 all reduce to the equations in Section 5.1.5 when the influent is free of biomass.

One significant impact of having biomass in the influent to a CSTR is to reduce S_{Smin}, the minimum substrate concentration attainable. As was done in Section 5.1.5, S_{Smin} can be calculated by letting the SRT become very large so that $1/\Theta_c$ approaches zero. If the assumption is made that S_{Smin} is negligible with respect to S_{SO}, which will generally be the case, then it is possible to show that:

$$S_{Smin} = \frac{K_S \cdot b_H}{\hat{\mu}_H\left(1 + \frac{X_{B,H,TO}}{Y_{H,T} \cdot S_{SO}}\right) - b_H}. \tag{5.67}$$

Comparison of Equation 5.67 to Equation 5.23 shows clearly that S_{Smin} will be smaller when the influent contains biomass. Furthermore, if we let Ω represent the value of S_{Smin} when the influent contains active biomass expressed as a fraction of the value in the absence of active biomass, it can be shown that:

$$\Omega = \frac{1 - \frac{b_H}{\hat{\mu}_H}}{1 - \frac{b_H}{\hat{\mu}_H} + \frac{X_{B,H,TO}}{Y_{H,T} \cdot S_{SO}}}. \tag{5.68}$$

For many situations this can be further simplified by noting that $b_H/\hat{\mu}_H$ is often much less than one, allowing it to be dropped from the equation:

$$\Omega \approx \frac{1}{1 + \frac{X_{B,H,TO}}{Y_{H,T} \cdot S_{SO}}}. \tag{5.69}$$

Equations 5.68 and 5.69 show clearly that the degree of reduction in S_{Smin} will depend on the magnitude of the influent biomass concentration relative to the influent substrate concentration, with larger values producing lower S_{Smin} values. This suggests that one way to meet a desired S_{Smin} concentration when a normal CSTR cannot is to add active biomass to the influent.

Another impact of having biomass in the influent to a CSTR is to prevent washout, because if the influent contains active biomass, so will the bioreactor, no matter how small the SRT is made. Thus, a minimum SRT can no longer be defined in the same sense that it was defined for a bioreactor without biomass in the influent. However, the degree of substrate removal that will occur at very small SRTs depends on the influent biomass concentration, as can be seen by an examination of Equation 5.64. Consider the special situation in which the SRT has been selected so that

$$\frac{1}{\Theta_c} + b_H = \hat{\mu}_H. \tag{5.70}$$

Under that condition, Equation 5.64 reduces to

$$\left[\frac{X_{B,H,TO}}{Y_{H,T}} + S_{SO} + K_S - S_{SO}\right] S_S = S_{SO} \cdot K_S \tag{5.71}$$

or

$$S_S = \frac{S_{SO} \cdot K_S}{X_{B,H,TO}/Y_{H,T} + K_S}. \tag{5.72}$$

Consequently, the larger the influent biomass concentration, the greater the degree of substrate removal, even when the SRT is very small. Equations 5.64 and 5.69 can be used to evaluate the potential impacts of purposeful addition of biomass (bioaugmentation) on process performance.

In Equation 5.59 we saw how the presence of inert suspended solids in the influent to a CSTR influenced the active fraction of the suspended solids. Thus, it would be instructive to see how the entrance of biomass into a CSTR influences it. Application of the definition of active fraction gives:

$$f_A = \frac{1}{1 + f_D \cdot b_H \cdot \Theta_c + X_{D,TO}\left[\dfrac{1 + b_H \cdot \Theta_c}{X_{B,H,TO} + Y_{H,T}(S_{SO} - S_S)}\right]}. \tag{5.73}$$

Comparison of it to Equations 5.36 and 5.59 reveals some interesting things. First, if the influent biomass is all active, so that $X_{D,TO}$ is zero, the active fraction is the same as that in a bioreactor that does not receive any solids. In other words, Equation 5.73 reduces to Equation 5.36. Second, if the influent biomass is all debris and none is active, Equation 5.73 reduces to Equation 5.59, as we would expect. Finally, if the influent biomass contains both active biomass and debris, which would be the more usual case, the active fraction in the bioreactor will depend on the active fraction of the influent biomass, but will be different from it.

The impact of influent biomass on the active fraction is another drawback to the use of the process loading factor (F/M ratio) as a design tool for bioreactors. As when inert suspended solids are present, SRT is more easily related to the biomass specific growth rate and, therefore, to process performance.

The presence of biomass in the influent to a CSTR increases the mass of solids that must be disposed of. The basic definition of the MLSS mass wastage rate is the wastage flow rate, F_w, multiplied by the MLSS concentration. Using Equation 5.66 for $X_{M,T}$ and using the definitions for SRT and HRT gives:

$$W_{M,T} = F\left[X_{D,TO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)X_{B,H,TO}}{1 + b_H \cdot \Theta_c} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right]. \tag{5.74}$$

Alternatively, if the MLSS concentration is known, the wastage rate can be determined by multiplying the wastage flow rate by the MLSS concentration; that is, by invoking the definition used to derive Equation 5.74. The cell debris entering the bioreactor is unaffected by microbial activity, so the mass of influent debris leaving in the waste solids equals the mass entering. However, the active biomass entering undergoes decay so that the mass discharged in the waste biomass stream is less than that in the influent. If we could mark the influent biomass in order to distinguish it from biomass generated in the bioreactor, we would find that its contribution was given by the second term in the brackets. Finally, new biomass is formed by growth and substrate utilization, and it too must be wasted from the bioreactor. Its contribution is given by the last term in the brackets.

As discussed in Section 5.1.5, the observed yield is defined as the mass of new biomass formed per unit mass of substrate removed. An examination of the last term in Equation 5.74 reveals that the amount of new biomass formed when biomass is in the influent to the bioreactor is given by the same expression as that in a bioreactor not receiving biomass. Thus, the presence of biomass in the influent has no impact on the observed yield, which is still given by Equation 5.38. However, it should be noted that the observed yield is difficult to measure in such a bioreactor because it is impossible to distinguish new biomass from that entering in the influent.

The basic equation for the oxygen requirement in a CSTR receiving both soluble substrate and biomass is the same as Equation 5.42 because that equation came from the overall rate

expression for oxygen as given by Equation 5.9, which is independent of the influent characteristics. The effects of the influent characteristics are incorporated through substitution of the appropriate equations for μ_H and $X_{B,H,T}$. Using Equation 5.62 for the former and Equation 5.63 for the latter yields:

$$RO = F\left\{\frac{(1-f_D)b_H \cdot \Theta_c \cdot X_{B,H,TO} \cdot i_{O/XB,T}}{1+b_H \cdot \Theta_c} + (S_{SO}-S_S)\left[1-\frac{(1+f_D \cdot b_H \cdot \Theta_c)Y_{H,T} \cdot i_{O/XB,T}}{1+b_H \cdot \Theta_c}\right]\right\}. \quad (5.75)$$

Comparison of this expression to Equation 5.43 reveals that the second term in the braces is the oxygen requirement associated with soluble substrate removal. The first term is the requirement associated with the input of active biomass. When $X_{B,H,TO}$ is zero, Equation 5.75 reduces to Equation 5.43. The input of biomass debris has no effect on the oxygen requirement because the debris passes through the bioreactor without reaction. The oxygen requirement can also be calculated by performing a mass balance on COD across the bioreactor, as discussed previously for the soluble substrate case, with proper consideration of the input of active biomass on that balance.

Example 5.2.3.1

Continue with the problem begun in Example 5.1.5.1. The conditions are the same as those used previously ($\tau = 8$ hr, $\Theta_c = 160$ hr, $S_{SO} = 200$ mg/L as COD), except that 21 mg/L as TSS (25 mg/L as COD) of active biomass is added to the influent. No biomass debris is added.

a. What is the effluent m-cresol concentration?
   To determine this we must use Equation 5.64 with the kinetic parameters and stoichiometric coefficients given in Table E5.1:

$$\begin{aligned}&[0.20-((1/160)+0.01)]S_S^2\\&-[0.20((21/0.28)+200)+(3.5-200)((1/160)+0.01)]S_S\\&+(200)(3.5)((1/160)+0.01)=0\end{aligned}$$

$$S_S = 0.22 \text{ mg/L as COD.}$$

   Thus, the presence of 21 mg/L as TSS of active biomass in the influent to the CSTR decreased the effluent m-cresol concentration by 0.09 mg/L as COD. This is approaching the minimum (0.18 mg/L) that could be obtained in such a bioreactor if no biomass was entering in the influent (as shown in Example 5.1.5.1).

b. What is the minimum m-cresol concentration that could be obtained in this bioreactor?
   This must be calculated with Equation 5.67:

$$S_{Smin} = \frac{(3.5)(0.01)}{0.20\left[1+\frac{21}{(0.28)(200)}\right]-0.01}$$

$$S_{Smin} = 0.13 \text{ mg/L as COD.}$$

   Note that this is less than the value for S_{Smin} when no active biomass is in the influent. If it was necessary to decrease S_S below 0.13 mg/L, either another bioreactor configuration would have to be used or more active biomass would have to be added to the influent.

c. What is the MLSS concentration?
Insertion of the appropriate values into Equation 5.66 gives:

$$X_{M,T} = \left(\frac{160}{8}\right)\left\{0 + \frac{[1+(0.20)(0.01)(160)]21}{1+(0.01)(160)} + \frac{[1+(0.20)(0.01)(160)](0.28)(200-0.22)}{1+(0.01)(160)}\right\}$$

$$X_{M,T} = 781 \text{ mg/L as TSS.}$$

Thus, the addition of 21 mg/L as TSS of active biomass to the influent increased the suspended solids concentration in the bioreactor by 213 mg/L as TSS. This is less than the impact of 21 mg/L of inert suspended solids considered in Example 5.2.2.1 because the active biomass underwent decay in the bioreactor, whereas the inert suspended solids did not undergo any reaction.

d. What is the wastage rate of MLSS from the bioreactor?
This may be calculated with Equation 5.74 or by multiplying the MLSS concentration by the wastage flow rate. Since we already know the MLSS concentration, the latter approach is easier:

$$W_{M,T} = (0.05)(781) = 39.1 \text{ mg/hr of dry solids.}$$

The addition of 21 mg/hr as TSS of active biomass to the bioreactor increased the mass of solids to be disposed of by 10.7 mg/hr as dry solids or 12.8 mg/hr as COD. The remainder of the influent biomass was destroyed by decay in the bioreactor.

e. How much oxygen must be supplied to the bioreactor?
This may be determined with Equation 5.75:

$$RO = 1.0\left\{\frac{(1-0.20)(0.01)(160)(21)(1.2)}{1+(0.01)(160)} + (200-0.22)\left[1-\frac{[1+(0.20)(0.01)(160)](0.28)(1.2)}{1+(0.01)(160)}\right]\right\}$$

$$RO = 178.1 \text{ mg/hr.}$$

Comparison of this value to the oxygen requirement calculated in Example 5.1.6.1 shows that the oxygen requirement was increased by 12.1 mg/hr. Only 0.1 mg/hr of this was due to increased substrate removal, the remainder being due to decay of the added biomass.

Frequently the waste solids stream from a bioreactor is directed to a CSTR to allow stabilization of the wasted biomass before ultimate disposal. This represents an extreme case of a CSTR receiving biomass in its influent because the influent MLSS concentration is generally high whereas the concentration of influent soluble substrate is very low. In that case, we are seldom concerned with the concentration of soluble substrate in the effluent and the term $(S_{SO} - S_S)$ can be considered to be negligible compared to the other terms in the performance equations, allowing their simplification. If this is done to Equations 5.63, 5.65, 5.66, 5.74, and 5.75, the resulting expressions are

$$X_{B,H,T} = \left(\frac{\Theta_c}{\tau}\right)\left(\frac{X_{B,H,TO}}{1+b_H \cdot \Theta_c}\right), \tag{5.76}$$

$$X_{D,T} = \left(\frac{\Theta_c}{\tau}\right)\left(X_{D,TO} + \frac{f_D \cdot b_H \cdot \Theta_c \cdot X_{B,H,TO}}{1 + b_H \cdot \Theta_c}\right), \tag{5.77}$$

$$X_{M,T} = \left(\frac{\Theta_c}{\tau}\right)\left[X_{D,TO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)X_{B,H,TO}}{1 + b_H \cdot \Theta_c}\right], \tag{5.78}$$

$$W_{M,T} = F\left[X_{D,TO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)X_{B,H,TO}}{1 + b_H \cdot \Theta_c}\right], \tag{5.79}$$

and

$$RO = F\left[\frac{(1 - f_D)b_H \cdot \Theta_c \cdot X_{B,H,TO} \cdot i_{O/XB,T}}{1 + b_H \cdot \Theta_c}\right]. \tag{5.80}$$

The equation for the active fraction, Equation 5.73 is unchanged. As usual, the term F in Equations 5.76–5.80 (either directly or through the HRT, τ) is the flow rate entering the bioreactor. However, it should be noted that for this situation, the flow often arises from another bioreactor (e.g., as its waste solids stream). Furthermore, bioreactors receiving only biomass in their feed often do not employ a biomass separator, but discharge their entire effluent stream to a solids dewatering device. In that case the SRT and HRT are equal, allowing further simplification of Equations 5.76–5.80.

Example 5.2.3.2

Continue with the problem begun in Example 5.1.5.1. If the waste solids stream from the bioreactor receiving only the soluble substrate is directed to a CSTR with an SRT of 480 hr and an HRT of 24 hr, what will be the MLSS concentration, the active fraction, and the oxygen requirement in the CSTR? In addition, how many mg/hr of dry solids must be sent to ultimate disposal from it?

a. What is the MLSS concentration?
This may be determined from Equation 5.78. The value of $X_{B,H,TO}$ is the same as the value of $X_{B,H,T}$ calculated in Example 5.1.5.2, which was 430 mg/L as TSS. The value of $X_{D,TO}$ is the same as the biomass debris concentration in the bioreactor in Example 5.1.5.2. This value was not calculated in the example, but is the difference between $X_{total,T}$ and $X_{B,H,T}$, or 138 mg/L as TSS. These values must be substituted into Equation 5.78, along with the other appropriate values:

$$X_{M,T} = \left(\frac{480}{24}\right)\left\{138 + \frac{[1 + (0.20)(0.01)(480)]430}{1 + (0.01)(480)}\right\}$$

$$X_{M,T} = 5{,}666 \text{ mg/L as TSS.}$$

b. What is the active fraction when biomass is measured on a TSS basis?
This may be determined in either of two ways: from its definition or from Equation 5.73, which is applicable in this case also. By definition, the active fraction is the concentration of active biomass divided by the MLSS concentration. The active biomass concentration may be calculated with Equation 5.76:

$$X_{B,H,T} = \left(\frac{480}{24}\right)\left[\frac{430}{1 + (0.01)(480)}\right]$$

$$X_{B,H,T} = 1483 \text{ mg/L as TSS.}$$

Dividing by 5666 mg/L as TSS gives an active fraction of 0.26. Most of the MLSS is biomass debris that accumulated as the active biomass underwent decay.

c. What is the oxygen requirement?
The oxygen requirement can be calculated with Equation 5.80. For this equation, the flow rate, F, into the bioreactor must be known. Since the flow entering the CSTR in this example is the wastage flow from the bioreactor in Example 5.1.5.1, its flow rate is 0.05 L/hr. Therefore:

$$RO = 0.05\left[\frac{(1-0.20)(0.01)(480)(430)(1.2)}{1+(0.01)(480)}\right]$$

$$RO = 17.1 \text{ mg/hr.}$$

The input rate of COD into the bioreactor is (0.05 L/hr)(568 mg/L as TSS)(1.2 mg COD/mg TSS) = 34.1 mg COD/hr. Thus, approximately half of the original oxygen demand was satisfied in the bioreactor. This represents 50% stabilization of the waste solids. The remainder of the oxygen demand remains in the sludge that goes to ultimate disposal, but much of it is in the form of biomass debris and degrades very slowly.

d. How many mg/hr of solids go to ultimate disposal?
Equation 5.79 may be used to determine this:

$$W_{M,T} = 0.05\left\{138+\frac{[1+(0.20)(0.01)(480)]430}{1+(0.01)(480)}\right\}$$

$$W_{M,T} = 14.2 \text{ mg/hr as dry solids.}$$

The COD of the solids going to ultimate disposal is 17.0 mg COD/hr. Note that the sum of this value and the oxygen requirement equals the input rate of COD into the bioreactor. This follows from the requirement for a COD balance across the bioreactor. All electrons in the waste solids going to the bioreactor must either be transferred to oxygen or remain in the unreacted solids. This serves as a convenient continuity check on the computations.

5.2.4 Biodegradable Solids in Influent

With the exception of some industrial wastes, most wastewaters contain particulate organic matter, much of which is biodegradable. This is true even when the biochemical operation is preceded by a sedimentation basin because much of the particulate organic matter is colloidal in size, making it too small for removal by settling. As a consequence, consideration must be given to the fate of particulate, biodegradable organic matter in order for models to accurately reflect the responses of biochemical operations treating many wastewaters.

An important characteristic of particulate organic matter is that it is too large to be transported across cell membranes. Thus, it must be acted on by extracellular enzymes to release soluble constituents that can be taken up and used as substrate by the biomass. As discussed in Sections 2.4.4 and 3.5, the solubilization reactions, commonly referred to as hydrolysis, are quite complex and have received little research attention. It is clear, however, that a specific reaction term must be included for conversion of particulate substrate into soluble substrate. The basic model presented in Section 5.1 does not include such a reaction term, and thus it is not adequate for considering the impact of particulate biodegradable organic matter. Furthermore, inclusion of terms for hydrolysis complicates the situation sufficiently that explicit equations of the type presented in this chapter are difficult to obtain. Consequently, the impacts of particulate substrate will not be considered further here. Rather, Chapter 6 will present a more complex model that considers the fate of particulate

substrate. Nevertheless, the basic concepts presented in this chapter, such as the importance of SRT, the buildup of inert solids, and so on are valid for all wastewaters and all suspended growth cultures, regardless of the nature of the electron donor. Thus, the concepts presented herein serve as a foundation for consideration of more complex situations.

5.2.5 Effects of Influent Solids on the Performance of a Continuous Stirred Tank Reactor as Predicted by Model

In Section 5.2.2 it was seen that the impact of inert solids in the influent to a CSTR is to reduce the active fraction of the MLSS. This effect is illustrated in Figure 5.7. The solid curve is the same as the one in Figure 5.5 whereas the dashed one represents the case in which 84 mg/L as TSS (100 mg/L as COD) of inert organic suspended solids are added to the influent. Comparison of the two curves reveals that only moderate amounts of inert suspended solids in the influent to a CSTR can decrease the active fraction to less than 50%, especially at longer SRTs. As a result, final settlers and pumps for the recycle of biomass must be made larger to handle solids that contribute nothing to the process. Thus, it is generally more economic to reduce the concentration of inert solids prior to biological reactors.

As discussed in Section 5.2.3, one significant effect of having biomass in the influent to a CSTR is to prevent washout, thereby allowing substrate removal to occur at SRTs below the normal minimum. This is illustrated in Figure 5.8, where the impact on the soluble substrate concentration of having 42 mg/L as TSS (50 mg/L as COD) of active biomass in the influent is shown. The most dramatic effect is at SRTs near the minimum. Instead of having a discontinuity at the point of washout like the curve for the bioreactor without biomass in the influent, the concentration in the bioreactor receiving biomass slowly approaches the influent concentration as the SRT is made smaller and smaller. Under those conditions the microorganisms are growing and removing substrate at a very rapid specific rate, but the residence time in the bioreactor is too short to allow more complete removal to occur. Of course, higher concentrations of biomass in the influent will allow more substrate to be removed at short SRTs. The main importance of the effect illustrated in Figure 5.8 is as an explanation of why washout does not occur in circumstances where it might be expected. This is especially important during laboratory studies in which investigators attempt to measure $\hat{\mu}$ by observing the SRT at washout. Contamination of

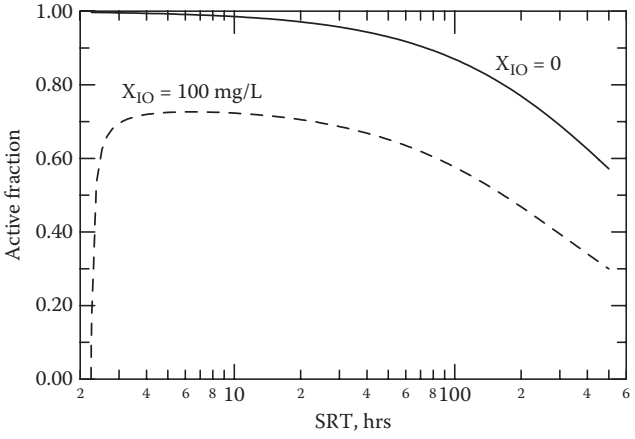

FIGURE 5.7 Effect of 100 mg/L (as COD) of influent inert organic solids on the active fraction (COD basis) of the biomass in a CSTR receiving a soluble substrate. Kinetic parameters and stoichiometric coefficients are listed in Table 5.3.

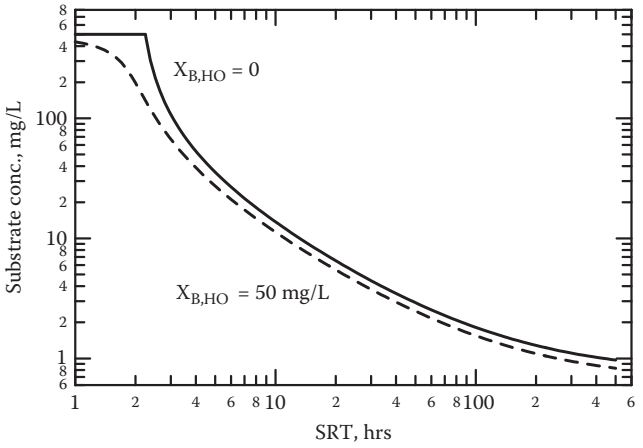

FIGURE 5.8 Effect of 50 mg/L (as COD) of active biomass in the influent to a CSTR on the soluble substrate concentration (as COD) in the reactor. Kinetic parameters and stoichiometric coefficients are listed in Table 5.3.

feed lines, thereby introducing biomass with the feed, can prevent the expected response and lead to error in the determination.

Another effect of influent biomass discussed previously is to reduce S_{Smin}, the minimum attainable substrate concentration from a single CSTR. It was seen in Equations 5.68 and 5.69 that the degree of reduction depends on the magnitude of the influent biomass concentration relative to the influent substrate concentration and this effect is illustrated in Figure 5.9. For the parameter values given in Table 5.3, S_{Smin} in the absence of influent biomass is 0.76 mg/L as COD, yet the presence of influent biomass can decrease that value significantly, as shown in the figure. This fact may be useful as engineers seek to reduce the concentrations of specific pollutants to very low levels. For example, although industrial wastewater treatment systems generally receive influent from several production areas, one may be the primary source of a targeted pollutant. If that waste stream was pretreated in a small bioreactor without biomass recycle prior to discharge to the main bioreactor, it would do two things: (1) provide a source of bacteria capable of degrading the targeted pollutant coming from the other production areas, and (2) reduce the concentration of the targeted pollutant

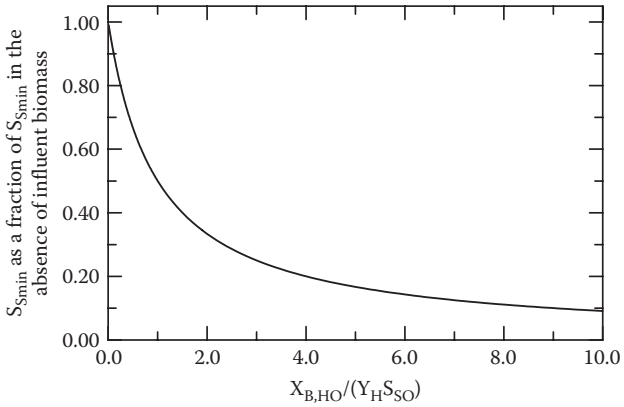

FIGURE 5.9 Effect of the influent biomass concentration (in COD units) relative to the influent substrate concentration (in COD units) on the minimum substrate concentration attainable in a CSTR. The values of S_{Smin} are expressed as fractions of the S_{Smin} value attainable in a similar bioreactor receiving no influent biomass.

in the influent to the main bioreactor. The combined effect of these two contributions would be to make the influent biomass to substrate concentration ratio large for the targeted pollutant, thereby allowing the main bioreactor to achieve a lower effluent substrate concentration than would be possible otherwise.

Reexamination of Figures 5.4 and 5.5 reveals that a CSTR with an SRT of 200 hours and an HRT of 10 hours would have a total biomass concentration of 3100 mg/L as COD (2600 mg/L as TSS) with an active fraction of 0.76 if it were treating a wastewater with the characteristics in Table 5.3. What would be the fate of the excess biomass from that bioreactor if it were sent for treatment to another CSTR? Since the concentration of soluble substrate in the waste biomass stream is negligible, Equations 5.76–5.80 describe the performance of the CSTR receiving the waste biomass. We will assess this scenario by holding the ratio of the SRT to the HRT constant at 10; in other words, as the SRT increases the HRT increases proportionally. Results for this case are shown in Figure 5.10, where the biomass concentrations are presented in COD units to facilitate comparison to the oxygen requirement. There it can be seen that because of the buildup of debris in the bioreactor, the total biomass concentration will not go to zero as the SRT is increased, but will approach a limit, although the active biomass will become quite small. Furthermore, it can be seen that there is a point of diminishing return with regard to further increases in SRT because the active biomass declines rapidly at first, but then more slowly as the SRT is increased further. This is characteristic of the first-order expression chosen to depict decay. It should be remembered that the model used assumes that debris is totally inert, whereas it will undergo some destruction given sufficient time, as discussed in Sections 2.4.2 and 3.3.1. Thus, it should be recognized that the residual stable biomass concentration will probably be less than that depicted by the model. Just as with the CSTR

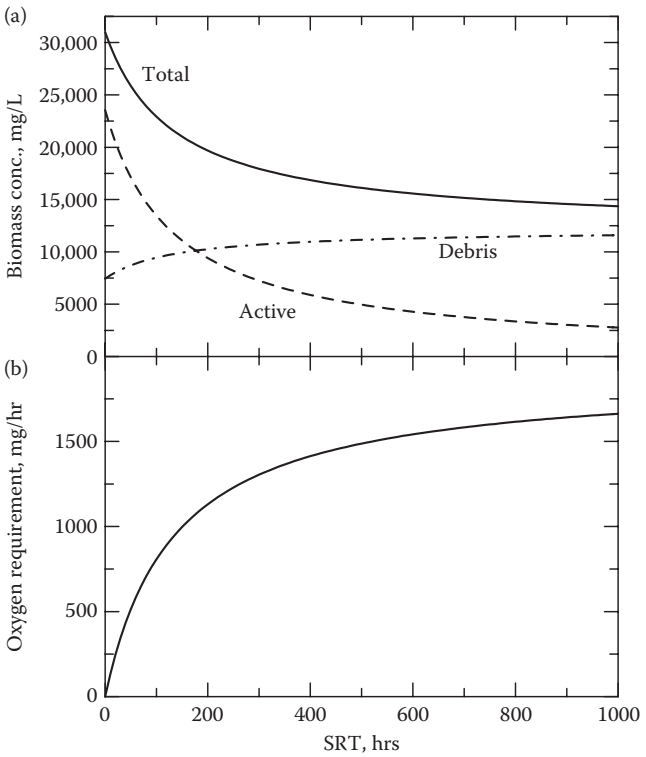

FIGURE 5.10 Effect of SRT on the performance of a CSTR receiving 1.0 L/hr of feed containing only 3100 mg/L as COD of biomass with an active fraction of 0.76. $\Theta_c/\tau = 10$. Kinetic parameters and stoichiometric coefficients are listed in Table 5.3. To express the biomass concentration in TSS units, divide by $i_{O/XB,T} = 1.2$ mg COD/mg TSS.

whose performance was depicted in Figure 5.6, in an aerobic process the destruction of biomass occurs at the expense of oxygen. Thus, the oxygen requirement is the mirror image of the total biomass curve. For simplicity's sake, the influent flow rate to the bioreactor was taken as 1.0 L/hr, making the mass input rate of biomass equal to 3100 mg/hr as COD. Thus, it can be seen that about 50% of the oxygen demand of the influent biomass must ultimately be satisfied at longer SRTs. In other words, the final residual solids are highly stabilized.

5.3 EFFECTS OF KINETIC PARAMETERS

The SRT of a CSTR is the primary control variable available to a designer or operator; however, it is not the only factor affecting the performance of such a bioreactor. Examination of the equations for the performance of a CSTR reveals that the values of each of the kinetic parameters and stoichiometric coefficients will influence them as well. The primary effects of $\hat{\mu}_H$ and K_S are on the substrate concentration. A higher value of $\hat{\mu}_H$ and a lower value of K_S allow the biomass to grow faster at a given substrate concentration, thereby giving a lower reactor substrate concentration for any given value of the SRT. The Monod parameters also exert a strong effect on the minimum SRT, so that organisms with high $\hat{\mu}_H$ and low K_S values can grow in CSTRs with short SRTs. The effect of the Monod parameters on biomass concentration is strongest at short SRTs where the effect on the substrate concentration is strongest. They have almost no effect at longer SRT values, however. In contrast to the Monod parameters, the primary effect of the decay coefficient is on the biomass concentration and the oxygen requirement at longer SRTs. A high decay coefficient means that the bioreactor will be more efficient in oxidizing the substrate to carbon dioxide; consequently, the biomass concentration will be low and the oxygen requirement high. This effect will be especially pronounced at long SRTs. Changes in the true growth yield will also primarily affect the biomass concentration and the oxygen requirement. High yields will result in more biomass, but the culture will require less oxygen because more of the electrons in the substrate will be retained in the biomass synthesized.

Seldom does a situation occur in which only one parameter changes. Usually all will change. For example, as discussed in Section 3.9, temperature can affect all of them and the response of the system will depend on how the changes interact. Thus, one must learn to think in terms of the group of parameter values that characterize a particular substrate and culture rather than thinking of individual coefficients.

5.4 BIOMASS WASTAGE AND RECYCLE

In developing the models in the preceding sections a portion of the flow from the bioreactor passed through a biomass separator, as shown in Figure 5.1. Alternative means of biomass separation and wastage were also introduced. Here, more details are given about the different options for biomass separation and wastage.

5.4.1 Garrett Configuration

The key features of the Garrett configuration are that for a CSTR of fixed volume, the SRT is controlled solely by the flow rate of the waste solids stream, F_w, and the performance of the bioreactor is independent of the solids recycle flow rate, F_r. This means that the recycle flow rate can be chosen to give proper operation of the settler, thereby ensuring that all biomass is returned to the bioreactor. In Chapter 11 we will consider the question of the choice of that flow rate.

The rationale behind the Garrett configuration can be seen by considering the system boundary to be the dashed line in Figure 5.2a. The flows in and out of that boundary are the same as those in Figure 5.1, and thus the Garrett configuration corresponds to the ideal bioreactor configuration used to derive the equations in Sections 5.1 and 5.2. Because the bioreactor is completely mixed, the

concentration of the constituents in the effluent stream leaving it and flowing to the settler are the same as those in the bioreactor. Furthermore, if the settler is operated properly, the solids blanket will be small so that the mass of solids in the settler is small relative to the mass in the bioreactor. This has two effects. First, there will be little reaction in the settler so that the concentrations of soluble constituents in the recycle stream are the same as those in the bioreactor. Because all soluble concentrations are the same, the recycle of soluble constituents around the system has no impact on system performance. This can be shown by performing mass balances around both the bioreactor and the settler. Furthermore, since the settler is perfect, all biomass entering it is returned to the bioreactor, and thus the recycle stream has no impact on the biomass concentration in the bioreactor either. (In Chapter 11 we will see how to correct system operation for the fact that the effluent stream from all real-world settlers contains a small amount of biomass.) Second, the SRT is still defined by Equation 5.1, and since the concentration of biomass in the wastage stream is the same as the concentration in the bioreactor, Equation 5.2 is also true. This means that the Garrett configuration conforms to the ideal situation considered in Sections 5.1 and 5.2, and thus the equations developed there are directly applicable. Because the recycle flow rate does not appear in any of those equations and because the SRT is not influenced by the recycle flow rate, it can be seen that bioreactor performance is independent of it.

5.4.2 Conventional Configuration

The conventional configuration shown in Figure 5.2b can also be defined by a dashed system boundary and thus by use of the arguments above, the recycle flow rate need not appear in the system descriptive equations, provided they are written in terms of the system SRT. However, because the solids in the recycle flow (and hence in the wastage flow) are at a higher concentration than the solids in the bioreactor, Equation 5.2 is no longer valid, although Equation 5.1 is valid. Furthermore, the waste solids concentration in Equation 5.1 is a function of the recycle flow rate. Thus, the key to understanding the impact of the recycle flow rate on the performance of a bioreactor with the conventional configuration is an understanding of the effect that the recycle flow rate has on the waste solids concentration.

Examination of Figure 5.2b reveals that the concentration of any constituent in the wastage flow will be the same as the concentration in the solids recycle stream since the wastage flow comes from it. The MLSS in a bioreactor can generally be considered to be homogeneous and to settle without segregation so that the ratio of active biomass concentration to MLSS concentration in the recycle stream is the same as the ratio in the bioreactor. Thus, we may simply perform a mass balance on MLSS across the settler to determine the value of X/X_w for use in Equation 5.1 for defining the SRT in terms of the recycle flow rate.

Assuming that the settler is perfect and that no reaction occurs in it, a steady-state mass balance across it states that the mass of MLSS entering the settler must equal the mass leaving in the combined recycle and wastage streams:

$$(F + F_r)X_{M,T} = (F_w + F_r)X_{M,T,r}, \tag{5.81}$$

where $X_{M,T,r}$ is the MLSS concentration in the recycle stream, in TSS units. Recognizing that the solids concentration in the wastage flow, $X_{M,T,w}$, is equal to $X_{M,T,r}$, Equation 5.81 may be rearranged and substituted into Equation 5.1 to give:

$$\Theta_c = \frac{V}{F_w}\left(\frac{F_w + F_r}{F + F_r}\right). \tag{5.82}$$

This shows that the SRT of a system with a conventional configuration depends on both the influent and recycle flow rates in addition to the wastage flow rate and the bioreactor volume. Rearrangement of Equation 5.82 gives:

$$F_w = \frac{V \cdot F_r}{(F + F_r)\Theta_c - V}. \tag{5.83}$$

This shows that any time the recycle flow rate is changed, the wastage flow rate must be adjusted to maintain a constant SRT. In contrast, if the Garrett configuration were used, no adjustment of the wastage flow would be required. Furthermore, Equation 5.83 also shows that the wastage flow from the conventional configuration must be adjusted any time the influent flow changes. Such a change is not required with the Garrett configuration, as shown by Equation 5.2.

Once the impact of the recycle flow rate on the SRT is recognized, all of the system performance equations from Sections 5.1 and 5.2, which were written in terms of the SRT, may be used because the recycle flow rate has no impact on them at a fixed SRT. Its impact is indirect, through influencing the SRT if proper adjustment of the wastage rate is not done. In Chapter 11 we will consider selection of the appropriate recycle flow rate for optimum settler performance.

5.4.3 Membrane Bioreactors

Membrane bioreactors use membrane filters in place of secondary clarifiers to separate the biomass from the liquid effluent. As discussed in Section 1.3.1, the membranes can be submerged in a section of the bioreactor where treatment is occurring, but they can also be placed in a separate chamber that is external to the bioreactor. The choice between these two options depends on practical considerations, such as access to the membranes for cleaning, so that the membranes are generally placed in a separate chamber in larger facilities (those greater than about 2000 to 4000 m^3/day in capacity). In both cases, wastage directly from the bioreactor is preferred for a variety of reasons including process simplicity, greater control of nuisance organisms, and the fact that little difference exists between the membrane section influent and effluent MLSS concentration. Thus, in general, the membrane bioreactor system would be modeled as described in Section 5.4.1 for the Garrett configuration.

5.5 KEY POINTS

1. Two retention times are important to the performance of continuous stirred tank reactors (CSTRs): the hydraulic residence time (HRT) and the solids retention time (SRT). The former represents the average length of time a fluid element stays in the bioreactor, whereas the latter is the average length of time a solid particle stays there. The SRT can never be less than the HRT, but can be greater if the bioreactor contains a separator that removes particulate material from the effluent and returns it to the bioreactor.
2. The matrix format is a convenient way to present the stoichiometry and kinetics for multiple parallel reactions acting on several components.
3. The specific growth rate of the biomass in a CSTR is controlled by the SRT of the bioreactor, but is independent of its HRT. Since there is a functional relationship between the specific growth rate of biomass and the concentration of substrate surrounding that biomass, control of the SRT allows control of the substrate concentration in a CSTR.
4. The concentration of biomass in a CSTR depends on the HRT, the SRT, and the amount of substrate removed. For a fixed SRT, the product of the HRT and the biomass concentration is a constant because the removal of a fixed mass of substrate generates a fixed mass of microorganisms.

5. The observed yield and active fraction of the biomass in a CSTR both decline as the SRT is increased because of the increased importance of death and decay at longer SRTs.
6. The oxygen requirement in an aerobic CSTR comes from a COD balance across the bioreactor; that is, it must equal the total COD in minus the total COD out, including the COD of the biomass and the biomass debris. Consequently, any factor that decreases the observed yield, such as an increase in the SRT, will increase the oxygen requirement.
7. The SRT is preferable to the process loading factor (F/M ratio) as a design and control parameter for CSTRs because the latter requires knowledge of the active fraction before it can be related to the specific growth rate of the biomass whereas the former does not.
8. As the SRT is increased past the point of washout, the soluble substrate concentration in a CSTR declines rapidly, but as the SRT is increased further, it has less and less impact on soluble substrate removal.
9. When the SRT is small, the excess biomass production rate is small because of incomplete substrate removal. When the SRT is large, the excess biomass production rate is small because of the importance of biomass decay. It will reach a maximum at intermediate values of the SRT. The oxygen requirement increases as the SRT is increased because more and more of the substrate and the biomass formed from it are oxidized.
10. Both soluble nonbiodegradable COD and inert solids will be unaffected by biomass in a CSTR. The concentration of the former in the bioreactor will be the same as its concentration in the influent, whereas the concentration of the latter will be greater by the factor Θ_c/τ because it becomes enmeshed in the MLSS and is lost from the system only through solids wastage.
11. The presence of active biomass in the influent to a CSTR decreases the specific growth rate of the biomass in the bioreactor, thereby reducing the effluent substrate concentration and increasing the biomass concentration associated with any given SRT. It also decreases the minimum substrate concentration that can be attained in the bioreactor.
12. When the influent to a CSTR contains biomass, but no soluble substrate, simplified equations can be derived for describing its performance.
13. Biodegradation of particulate substrates requires their solubilization by hydrolysis reactions. Such reactions are not included in the models in this chapter, and thus those models cannot be used to predict the performance of CSTRs receiving such substrates.
14. Changes in the Monod parameters, $\hat{\mu}_H$ and K_S, are reflected primarily in the substrate concentration, whereas changes in the true growth yield, Y_H, and decay coefficient, b_H, have their largest effects on the biomass concentration.
15. Two methods are used to waste biomass from a CSTR. The Garrett configuration wastes biomass directly from the bioreactor. The conventional configuration wastes biomass from the recycle flow returning biomass from the settler to the bioreactor. The Garrett configuration is easier to operate because it is not necessary to adjust the wastage flow rate each time a change is made in the recycle flow rate.

5.6 STUDY QUESTIONS

1. A CSTR with a volume of 1000 L receives a flow of l00 L/hr. Ninety percent of the effluent exits through a biomass separator that removes all particulate material and returns it to the bioreactor. The remainder exits directly from the bioreactor. What is the HRT? What is the SRT?
2. Using the traditional approach to decay, write the matrix representing the stoichiometry and kinetics for aerobic growth of heterotrophic biomass on a soluble, noninhibitory substrate. Then use the information in the matrix to write the reaction rate term for active biomass and explain how you did it.

3. Derive Equation 5.22 showing that the effluent substrate concentration from a CSTR is a function of only the SRT and is independent of the influent substrate concentration. Then explain what happens in the bioreactor to allow the effluent substrate concentration to be independent of the influent concentration.
4. Using derivations as needed, explain why the product of the biomass concentration and the HRT is a constant for a fixed SRT and influent concentration. Then explain why the steady-state performance of a CSTR equipped with a biomass separator is independent of its HRT.
5. The two major costs associated with the treatment of wastewaters in aerobic CSTRs are from the disposal of the excess biomass produced and the provision of ample oxygen. Describe how the SRT influences each of these and use that information as the basis of a discussion of the factors that must be considered by an engineer in choosing the design SRT for a CSTR.
6. A feed containing a soluble substrate with a biodegradable COD of 1000 mg/L is flowing at a rate of 100 L/hr into a 1000 L aerobic CSTR that contains a mixed community of microorganisms. The kinetic parameters and stoichiometric coefficients characterizing the culture are given in Table SQ5.1.
   a. Determine the following for an SRT of 100 hr.
      (1) The concentration of soluble biodegradable COD in the effluent.
      (2) The total biomass concentration in the bioreactor in TSS units.
      (3) The active fraction of the biomass.
      (4) The g/hr of oxygen that must be supplied to the bioreactor.
      (5) The g TSS/hr of excess biomass that must be disposed of.
      (6) The mg/hr of nitrogen that will be used for synthesis of new biomass.
      (7) The process loading factor.
   b. Would it be possible to achieve an effluent biodegradable COD concentration of 1.5 mg/L with this substrate and culture? Why?
   c. What SRT would be required to reduce the soluble substrate concentration to 20 mg/L as COD? What concentration of active biomass would be present in the bioreactor?
   d. If 200 mg/L as TSS of inert suspended solids that are 75% (by weight) organic are added to the influent, what fraction of the MLSS will be inert suspended solids for the condition in part a? What will be the active fraction of the MLSS on a TSS basis? What will be the MLSS concentration in TSS units? What will it be in COD units?
7. Explain through derivation why the concentration of inert solids in a CSTR is greater than the concentration in the influent by the factor Θ_c/τ.
8. Explain why the SRT is preferable to the process loading factor as a means of controlling a CSTR when the influent contains either inert solids or biomass.

TABLE SQ5.1
Kinetic Parameters and Stoichiometric Coefficients for a Mixed Microbial Community Growing on a Mixture of Organic Compounds

Symbol	Units	Value
$\hat{\mu}_H$	hr^{-1}	0.40
K_S	mg/L as COD	120
Y_H	mg biomass TSS/mg substrate COD	0.45
$i_{O/XB,T}$	mg COD/mg TSS	1.20
b_H	hr^{-1}	0.004
f_D	mg debris/mg biomass	0.20

9. Rework Study Question 6a, but include 90 mg/L as TSS of biomass with an active fraction of 80% in the influent. Calculate how much nitrogen must be added in units of mg/hr in order to meet the nitrogen requirement for the reactor.
10. A CSTR is being used to degrade biomass. A stream containing biomass at a concentration of 650 mg/L as TSS is flowing at a rate of 2.0 L/hr into a CSTR with a volume of 48 L that is being operated with an SRT of 240 hr. The active fraction of the biomass in the influent is 0.72; the decay coefficient, b_H, is 0.007 hr^{-1}; the COD coefficient, $i_{O/XB,T}$, is 1.2 mg COD/mg TSS; and the fraction of biomass contributing to biomass debris, f_D, is 0.20.
    a. What is the biomass concentration in the bioreactor (as TSS)?
    b. What is its active fraction (on a TSS basis)?
    c. What is the oxygen consumption rate in mg/hr?
    d. How many mg/hr (as TSS) of biomass go to ultimate disposal?
11. Explain why the model presented in this chapter is not adequate for describing the performance of a CSTR receiving biodegradable particulate substrate.
12. Describe both the Garrett and conventional configurations for wasting biomass from CSTRs. Then explain (using derivations as needed) why operation of the Garrett configuration is simpler.
13. Using sketches as appropriate, describe the effects of SRT on the performance of a CSTR receiving a soluble substrate and explain conceptually why those effects occur.
14. Using Figures 5.4 and 5.5, generate a graph of active biomass as a function of SRT and discuss the implications of the curve to the selection of the SRT for a bioreactor.
15. Describe the major impacts of the presence of active biomass in the influent to a CSTR on the removal of soluble substrate. How can those effects be used to advantage in the treatment of wastewaters containing specific pollutants that must be removed to very low levels?
16. Using sketches as appropriate, describe the effects of SRT on the performance of a CSTR receiving only biomass and explain conceptually why those effects occur.

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6 Multiple Microbial Activities in a Single Continuous Stirred Tank Reactor

In Chapter 5 we investigated the growth of aerobic heterotrophic bacteria in a single continuous stirred tank reactor (CSTR) receiving a soluble substrate. Through the development of a simple model we saw that the solids retention time (SRT) is an important determinant of bioreactor performance because it is related to the specific growth rate of biomass at steady state. We also saw that there is a minimum SRT below which biomass growth cannot occur, as well as a minimum substrate concentration that can be achieved no matter how large the SRT. Finally, we saw how stoichiometry can be applied to determine the amount of electron acceptor required and the amount of excess biomass produced. All of these characteristics are of fundamental importance and apply to all types of biomass, both heterotrophic and autotrophic, in all types of environments, whether aerobic, anoxic, or anaerobic. Thus, even though the concepts in Chapter 5 were developed in the context of aerobic growth of heterotrophs, they are broadly applicable.

In spite of the broad utility of the concepts, the model developed in Chapter 5 has two characteristics that restrict its applicability. First, it is limited to soluble, readily biodegradable substrates, whereas most wastewaters contain particulate contaminants and soluble constituents of large molecular weight that must be reduced in size before they can be taken into the bacteria for biodegradation. If a model is to accurately depict the response of bioreactors receiving such wastewaters, it must include hydrolysis reactions. Second, the biomass is assumed to be in a constant biochemical environment with no limitation by the electron acceptor. In many systems, however, limitations or alterations in the supply of electron acceptor cause shifts between aerobic and anoxic conditions, with short periods of anaerobiosis as well, and during these shifts the concentration of the electron acceptor may be limiting. Thus it would be desirable for a model to handle such situations.

To encourage practicing engineers to use modeling more extensively during the analysis of alternative wastewater treatment systems, in 1983 the International Water Association (IWA) appointed a task group to review models for suspended growth cultures and to produce one capable of depicting the performance of wastewater treatment systems receiving both soluble and particulate substrates in which organic substrate removal, nitrification, and denitrification were all occurring. In other words, they were to consider most of the processes discussed in Section 2.4. The result of their deliberations was the IWA Activated Sludge Model (ASM) No. 1.[21,22] Because ASM No. 1 is the result of the deliberations of several researchers with diverse opinions, because it is capable of mimicking the performance of pilot-[12] and full-scale[4] systems, and because it is relatively straightforward in its approach, it will be adopted herein for investigating more fully the performance of suspended growth bioreactors. In this chapter, ASM No. 1 will be used to introduce an approach to modeling complex systems and to illustrate the impact in a single CSTR of the processes and events not covered in Chapter 5. In Chapter 7 it will be used to investigate the performance of multiple bioreactor systems performing organic substrate removal, nitrification, and denitrification.

Because of the success of ASM No. 1, the IWA task group was asked to produce a consensus model capable of mimicking the performance of systems capable of performing phosphorus removal in addition to organic substrate removal, nitrification, and denitrification. The result was ASM No. 2,[23] which was later modified to include denitrifying phosphate accumulating organisms (PAOs),

resulting in ASM No. 2d.[24] Use will be made of this model in Chapter 7 to simulate the performance of nitrogen and phosphorus removal systems, but it will not be explained in the same detail as ASM No. 1 because of the large number of components and processes involved. The major rate expressions associated with phosphorus removal in the model were presented in Section 3.7. Finally, additional major changes were incorporated into ASM No. 3.[18] This version of the model has more limited application than ASM No. 2d and will, therefore, receive only limited attention here.

Modeling is now used extensively in biological wastewater treatment, with similar concepts being applied to develop models for anaerobic wastewater and sludge treatment processes.[6,17,47] We will describe and demonstrate one of these models, the IWA Anaerobic Digestion Model (ADM), in Chapter 8. The reader is encouraged to consult the primary literature for further information on this and other models.

6.1 INTERNATIONAL WATER ASSOCIATION ACTIVATED SLUDGE MODELS

International Water Association ASM No. 1 is presented in matrix format in Table 6.1, where it can be seen to incorporate 8 processes and 13 components. Examination of the table reveals the utility of the matrix format because application of the principles discussed in Section 5.1.3 allows immediate identification of the fate of each component and construction of the overall reaction rate term for it. It should be noted that components 1 through 8 are expressed as chemical oxygen demand (COD), whereas components 9 through 12 are given as nitrogen. The alkalinity is in molar units. These units are given in Table 6.2 along with the definition of each component. The use of COD, rather than total suspended solids (TSS), in the model for particulate components simplifies the equations and generates results that are more easily validated with a COD mass balance. Therefore, when shifting from the simple model to an ASM model, the reader will need to translate components into the correct units using the methods described in Chapter 5.

6.1.1 Components in Model No. 1

Components 1 through 5 and component 12 are all particulate. Inert particulate organic material is denoted by X_I. The fact that there are no entries listed under it in Table 6.1 shows that it is neither generated nor destroyed in a biochemical reactor. However, if it is present in the influent, it will accumulate as discussed in Section 5.2.2, with the degree being determined by the ratio of the SRT to the hydraulic residence time (HRT). Component 2, X_S, is a slowly biodegradable substrate. Although it is treated as a particulate constituent, its concentration in the influent to the bioreactor must be determined experimentally, as will be discussed in Chapter 9. It is destroyed by hydrolysis reactions, but is generated during biomass decay, which is modeled by the lysis:regrowth concept, as discussed in Section 3.3.2. Active heterotrophic biomass is depicted by $X_{B,H}$, whereas active autotrophic biomass (nitrifying bacteria) is given as $X_{B,A}$. One of the simplifying assumptions of ASM No. 1 is that nitrification occurs in a single step. Consequently, $X_{B,A}$ incorporates both ammonia oxidizing bacteria (AOB) and nitrite oxidizing bacteria (NOB). Active heterotrophic biomass is generated by growth on readily biodegradable substrate, whereas $X_{B,A}$ is generated by growth on ammonia-N. Both $X_{B,H}$ and $X_{B,A}$ are lost by decay, leading to slowly biodegradable substrate and biomass debris, X_D. The latter is inert, and behaves in a manner similar to X_I. Component 12, X_{NS}, is particulate biodegradable organic nitrogen. It is also formed by decay reactions since the slowly biodegradable substrate arising from biomass decay contains proteins and other nitrogen-containing organic compounds of high molecular weight. It is destroyed by hydrolysis.

Components 6 through 11 are all soluble, and with the exception of S_I, which is inert, are the constituents upon which the biomass acts. The presence of S_I in the matrix is simply to remind us that wastewaters contain nonbiodegradable soluble COD that passes through the bioreactor unaffected by biological activity, as discussed in Section 5.2.1. Readily biodegradable substrate is denoted by S_S. It is removed by the growth of heterotrophic biomass under aerobic or anoxic conditions and is

TABLE 6.1
Process Kinetics and Stoichiometry for Multiple Events in Suspended Growth Cultures as Presented by IWA Task Group on Mathematical Modeling for Activated Sludge Model No. 1

Component[a] → i		1	2	3	4	5	6	7	8	9	10	11	12	13	
j	Process ↓	X_I	X_S	$X_{B,H}$	$X_{B,A}$	X_D	S_I	S_S	S_O[b]	S_{NO}	S_{NH}	S_{NS}	X_{NS}	S_{ALK}	Process Rate, r_j, $ML^{-3}T^{-1}$
1	Aerobic growth of heterotrophs			1				$-\frac{1}{Y_H}$	$\frac{1-Y_H}{Y_H}$		$-i_{N/XB}$			$-\frac{i_{N/XB}}{14}$	$\hat{\mu}_H\left(\frac{S_S}{K_S+S_S}\right)\left(\frac{S_O}{K_{O,H}+S_O}\right)X_{B,H}$
2	Anoxic growth of heterotrophs			1				$-\frac{1}{Y_H}$		$\frac{1-Y_H}{2.86Y_H}$	$-i_{N/XB}$			$\frac{1-Y_N}{14(2.86Y_N)}-\frac{i_{N/XB}}{14}$	$\hat{\mu}_H\left(\frac{S_S}{K_S+S_S}\right)\left(\frac{K_{O,H}}{K_{O,H}+S_O}\right)\left(\frac{S_{NO}}{K_{NO}+S_{NO}}\right)\eta_g X_{B,H}$
3	Aerobic growth of autotrophs				1				$\frac{4.57-Y_A}{Y_A}$	$\frac{1}{Y_A}$	$-i_{N/XB}-\frac{1}{Y_A}$			$-\frac{i_{N/XB}}{14}-\frac{1}{7Y_A}$	$\hat{\mu}_A\left(\frac{S_{NH}}{K_{NH}+S_{NH}}\right)\left(\frac{S_O}{K_{O,A}+S_O}\right)X_{B,A}$
4	Death and lysis of heterotrophs		$1-f'_D$	−1		f'_D							$i_{N/XB}-f'_D i_{N/XD}$		$b_{L,H}X_{B,H}$
5	Death and lysis of autotrophs		$1-f'_D$		−1	f'_D							$i_{N/XB}-f'_D i_{N/XD}$		$b_{L,A}X_{B,A}$
6	Ammonification of soluble organic nitrogen										1	−1		$\frac{1}{14}$	$k_a S_{NS}X_{B,H}$
7	Hydrolysis of particulate organics		−1					1							$k_h\frac{X_S/X_{B,H}}{K_X+(X_S/X_{B,H})}\left[\left(\frac{S_O}{K_{O,H}+S_O}\right)+\eta_h\left(\frac{K_{O,H}}{K_{O,H}+S_O}\right)\left(\frac{S_{NO}}{K_{NO}+S_{NO}}\right)\right]X_{B,H}$
8	Hydrolysis of particulate organic nitrogen											1	−1		$r_7(X_{NS}/X_S)$
	Observed conversion rates, $ML^{-3}T^{-1}$									$r_i=\sum_{j=1}^{n}\psi_{ij}r_j$					

Note: Adapted from Henze, M., Grady, C. P. L., Jr., Gujer, W., Marais, G. v. R., and Matsuo, T., Activated sludge model No. 1. *IAWPRC Scientific and Technical Reports*, No. 1, International Water Association, London, 1987; Henze, M., Grady, C. P. L., Jr., Gujer, W., Marais, G. v. R., and Matsuo, T., A general model for single-sludge wastewater treatment systems. *Water Research*, 21:505–15, 1987.

[a] All organic compounds (1–7) and oxygen (8) are expressed as COD; all nitrogenous components (9–12) are expressed as nitrogen.

[b] Coefficients must be multiplied by −1 to express as oxygen.

TABLE 6.2
Definitions of Component Symbols in Table 6.1

Component Number	Component Symbol	Definition
1	X_I	Inert particulate organic matter, mg/L as COD
2	X_S	Slowly biodegradable substrate, mg/L as COD
3	$X_{B,H}$	Active heterotrophic biomass, mg/L as COD
4	$X_{B,A}$	Active autotrophic biomass, mg/L as COD
5	X_D	Debris from biomass death and lysis, mg/L as COD
6	S_I	Inert soluble organic matter, mg/L as COD
7	S_S	Readily biodegradable substrate, mg/L as COD
8	S_O	Oxygen, mg/L as COD
9	S_{NO}	Nitrate nitrogen, mg/L as N
10	S_{NH}	Ammonia nitrogen, mg/L as N
11	S_{NS}	Soluble biodegradable organic nitrogen, mg/L as N
12	X_{NS}	Particulate biodegradable organic nitrogen, mg/L as N
13	S_{ALK}	Alkalinity, molar units

generated by hydrolysis of slowly biodegradable organic matter. Its concentration in the wastewater entering a bioreactor must be determined experimentally, and the procedures for doing so will be discussed in Chapter 9.

Component 8 is oxygen, which is removed by aerobic growth of heterotrophic and autotrophic bacteria. The COD-based stoichiometric term for oxygen associated with heterotroph growth is the same as that in Table 5.1. For similar reasons, the oxygen term associated with autotroph growth contains the factor 4.57 because ammonia is the substrate for autotrophic nitrifying bacteria and its concentration in the matrix (S_{NH}, component 10) is expressed as nitrogen, whereas oxygen is expressed as COD. Furthermore, Y_A, the true growth yield for autotrophic bacteria, has units of mg of biomass COD formed per mg of nitrogen oxidized. Since the stoichiometric expression for oxygen in process 3 (autotrophic growth) comes from a COD-based stoichiometric equation, a factor must be included for the COD equivalents of ammonia-N in order to have consistent units. The oxidation state of nitrogen is changed from –III to + V as nitrifiers form nitrate from ammonia. The amount of oxygen required to accept the electrons removed during this oxidation is 4.57 g O_2/g N, as indicated in Table 3.1. Thus, that factor must be included. Unlike the model in Table 5.1, which used the traditional approach to decay, no oxygen utilization is associated directly with biomass loss in Table 6.1 because it is modeled with the lysis:regrowth approach. Rather, the oxygen utilization associated with biomass loss occurs because of the use of readily biodegradable substrate generated by hydrolysis of the slowly biodegradable substrate formed by death and lysis.

Components 9–12 all involve nitrogen species. Component 9, S_{NO}, is nitrate-N. It is formed by aerobic growth of autotrophic bacteria and is lost as it serves as the electron acceptor for anoxic growth of heterotrophic bacteria. In the latter role, the oxidation state of the nitrogen is changed from +V to zero and the factor 2.86 in the stoichiometric coefficient represents the oxygen equivalence of this change in units of g COD/g N, as shown in Table 3.1. Examination of column 10 shows that ammonia-N is involved in several reactions. Since ammonia is the preferred form of nitrogen for biomass growth, the term $-i_{N/XB}$ is included in rows 1–3 to represent the amount of nitrogen incorporated into new biomass. No provision is made in this model for reduction of nitrate-N to ammonia-N for incorporation into biomass in the event insufficient ammonia is present. This restriction should be recognized. Other models[26,49] allow the use of nitrate-N for biomass synthesis. The second stoichiometric coefficient in column 10 for aerobic growth of autotrophic bacteria represents the use of ammonia as a substrate and is analogous to the coefficients used for readily biodegradable substrate

(column 7) removal by heterotrophic biomass in rows 1 and 2. Ammonia is formed by ammonification of soluble organic nitrogen, S_{NS}, which is the last nitrogen-based soluble constituent. It, in turn, is formed by hydrolysis of particulate organic nitrogen.

In Section 3.2.10 the sensitivity of autotrophic biomass to low pH was discussed. Furthermore, in Section 3.2.5 it was seen that alkalinity is destroyed during their growth. If the wastewater contains insufficient alkalinity, the growth of autotrophic biomass will cease because a needed nutrient (carbon) is missing and because the pH will drop, inhibiting their activity. Thus, the destruction of alkalinity by autotrophic bacterial growth is an important event that must be considered by the engineer. Another factor influencing alkalinity is denitrification, which produces it, and in properly configured systems this production can offset somewhat its destruction. The coefficients in column 13 account for the changes in alkalinity, S_{ALK}, associated with nitrogen conversions and biomass growth in the bioreactor. Although the IWA task group did not attempt to model the effects of those changes on pH, they noted that when the alkalinity falls below 50 mg/L as $CaCO_3$, the pH becomes unstable and can fall well below 6,[54] thereby hindering nitrification.

6.1.2 Reaction Rate Expressions in Model No. 1

The growth of heterotrophic bacteria with the associated use of substrate and an electron acceptor is given by processes 1 and 2 in Table 6.1 for the situation in which ammonia serves as the nitrogen source for synthesis of new cell material. Although ASM No. 1 can be modified to include assimilative nitrate reduction for heterotrophic growth in situations when the amount of ammonia-N provided is insufficient,[26] this process is not routinely included in the model. Process 1 is aerobic growth and its rate is given by Equation 3.35 with substitution of Equation 3.46 for μ, the specific growth rate coefficient. In this case, reactant 1 is readily biodegradable substrate and reactant 2 is dissolved oxygen (DO). The main purpose of the DO term is to turn off aerobic growth as the DO concentration becomes low and to allow anoxic growth to begin if nitrate is present, as suggested by the rate term for process 2. The specific growth rate coefficient in that rate term is of the same form as Equation 3.48, with $K_{O,H}$ serving as the inhibition coefficient K_{IO}. By using $K_{O,H}$ in that capacity, the oxygen terms in the rate expressions for processes 1 and 2 complement each other, with one approaching zero as the other approaches one (their sum always equals one). It should be noted that both heterotrophic rate expressions go to zero under totally anaerobic conditions (i.e., in the absence of both oxygen and nitrate). Given long-term acclimation to anaerobic conditions, fermentative reactions would allow growth of facultative bacteria. ASM No. 1 does not consider this process although ASM No. 2/2d does. This limitation in ASM No. 1 should be recognized and the model should not be used to simulate bioreactors in which fully anaerobic conditions develop. Comparison of the rate term for process 2 to Equation 3.48 reveals the presence of an additional empirically derived parameter, η_g. It is a correction factor for growth under anoxic conditions. As we saw in Section 3.2.10, the $\hat{\mu}_H$ and K_S values for a limited number of pure culture studies differed when the cultures were grown on the same substrate under totally aerobic and totally anoxic conditions. Under anoxic conditions, $\hat{\mu}_H$ was always smaller than under aerobic conditions, whereas the relative changes in K_S varied. On the other hand, similar studies conducted with activated sludge cultures, which allow for interactions among species that would not be observed in pure culture studies, suggested that the growth parameters are similar under anoxic and aerobic conditions.[38] Therefore, it is difficult to generalize about how $\hat{\mu}_H$ and K_S vary with different redox states within activated sludge cultures. Consequently, the IWA activated sludge models assume that these parameters are constant across the different growth conditions. Additionally, ASM No. 1 assumes that the heterotrophic growth yield is independent of the redox state, whereas numerous studies have shown that it is lower under anoxic conditions.[38,41,42,52] To further complicate things, it is reasonable to assume that the entire biomass will be capable of aerobic growth, but only a portion of it will be able to grow under anoxic conditions. Collectively, these observations and assumptions point to slower anoxic growth relative to aerobic growth and the purpose of η_g is to correct for this condition

by capturing the various factors contributing to it. Because only a portion of the biomass will be capable of denitrification, η_g takes on values less than one, with those values depending somewhat on the system configuration.[20]

Process 3 in Table 6.1 is aerobic growth of autotrophic bacteria, which is modeled with Equation 3.46 in which reactant 1 is ammonia-N and reactant 2 is DO. Two important things should be noted about the way this process is modeled. The first is that nitrification is considered to occur in a single step, with nitrate-N arising directly from ammonia-N. This is a simplification, because, as we saw in Section 2.3.2, nitrification is a two-step process, with nitrite as an intermediate. As discussed in Section 3.2.10, the kinetic parameters for AOBs and NOBs are similar. Consequently, under balanced growth conditions nitrite is used as fast as it is formed so that its concentration is usually very low and of little importance. Therefore, to reduce the number of equations and to simplify the model, ASM No. 1 uses a one-step approach. However, it should be noted that nitrite can accumulate in suspended growth cultures due to an imbalance between the two bacterial populations, particularly during start-up, following severe temperature changes, or under low DO conditions when residual ammonia exists.[1,45,51] Furthermore, nitrogen removal technologies have been developed that intentionally select for AOBs over NOBs to achieve significant efficiencies by short circuiting nitrogen removal through nitrite (nitritation/denitritation; Section 23.3.3) or by involving innovative metabolisms such as anammox (Sections 2.3.3 and 23.3.3). Although modifications to the IWA activated sludge models have been developed that separate the AOB and NOB growth steps,[26,29,48,50] ASM No. 1 does not and is only appropriate for bioreactors at steady state or for those receiving dynamic loads no more severe than the diurnal flows and concentrations normally entering domestic wastewater treatment systems. The second important characteristic of the process rate expression for nitrification is that no consideration is given to substrate and product inhibition, which are known to occur at high nitrogen concentrations, as discussed in Section 3.2.10. These factors were not considered in ASM No. 1 because of a lack of adequate kinetic relationships and because substrate and product inhibition do not normally occur at the pH and nitrogen levels commonly found in domestic wastewater. Consequently, ASM No. 1 should not be applied to simulate the treatment of wastewaters containing nitrogen concentrations greatly in excess of those levels. Newer models are available that allow the impact of wastewaters with high nitrogen concentrations to be considered.[26]

Loss of heterotrophic biomass, process 4, is modeled by the lysis:regrowth approach discussed in Section 3.3.2. A primary reason for adopting this approach is that no use of electron acceptor is directly associated with it, thereby making it easier to express the effects of alternative electron acceptors in the overall model. In ASM No. 1, the loss of heterotrophic biomass is assumed to continue at the same rate, regardless of the electron acceptor available, and thus it is modeled by Equation 3.63. Similarly, the formation of biomass debris, slowly biodegradable substrate, and particulate biodegradable organic nitrogen are modeled by Equations 3.64–3.66, respectively. The nature of the electron acceptor will influence the rates of utilization of these constituents, however, as reflected in the other process rate expressions.

Process 5, loss of autotrophic biomass, is also modeled by the lysis:regrowth approach, although the amount of autotroph regrowth is not really significant, as discussed previously. Rather, heterotrophs grow on the organic substrate resulting from death and lysis of the autotrophic bacteria. As a consequence, the magnitude of the loss coefficient for autotrophs is the same as that for the traditional decay approach.

As nitrogen-containing organic compounds undergo biodegradation, the nitrogen in them is released as ammonia, as discussed in Section 3.6. This release is reflected in process 6, which is modeled with Equations 3.79 and 3.80. These expressions are approximate because of a lack of information about ammonification, as discussed in Section 3.6. They should be satisfactory, however, within the constraints established for the model.

An important contribution of ASM No. 1 is consideration of the fate of particulate and other slowly biodegradable substrate, as reflected in process 7, which is hydrolysis. Although the fate of

such material in suspended growth cultures is important, relatively little research has been done from which a rate expression can be developed, as discussed in Section 3.5. Nevertheless, based on the limited literature available, ASM No. 1 uses Equation 3.77 as the basic rate expression. Comparison of it to the expression in Table 6.1, however, reveals that it was extended to include the effects of the electron acceptor. First, it will be noted that another correction factor, η_h, is included to reflect the assumed retardation of hydrolysis under anoxic conditions. Like η_g, this correction factor is empirical and the rationale for its use is the same. Second, the rate of hydrolysis is assumed to go to zero in the total absence of oxygen or nitrate, although this assumption was later changed in ASM No. 3. In fact, hydrolysis is known to occur in anaerobic bioreactors (see Section 2.3.3). Because death and lysis are thought to continue at the same rate regardless of the nature of the electron acceptor, there will be an accumulation of slowly biodegradable substrate when suspended growth cultures are subjected to short periods without either oxygen or nitrate. In spite of the lack of certainty associated with the rate expression for process 7, the patterns of oxygen and nitrate-N utilization predicted by ASM No. 1 have been found to mimic well the performance of both pilot-[12,13] and full–scale[4] suspended growth systems with a number of configurations. Thus, although significant modifications, which are discussed below, have been made to ASM No. 1 since its introduction, use of the model depicted in Table 6.1 is still satisfactory for gaining a basic understanding of the response of suspended growth cultures to the major process variables.

The final process in Table 6.1 is conversion of the particulate, biodegradable organic nitrogen, X_{NS}, into soluble, biodegradable organic nitrogen, S_{NS}. This rate is assumed to be proportional to the rate of hydrolysis of slowly biodegradable organic matter, as modeled with Equation 3.78.

Two events discussed in Chapters 2 and 3 are not included in Table 6.1: soluble microbial product formation and phosphorus uptake and release. The impact of soluble microbial product formation is minor for most facilities, and acts primarily to raise the concentration of soluble organic matter in the effluent from a bioreactor, as discussed previously. It was excluded for the same reason it was excluded from the simple model in Chapter 5. Phosphorus uptake and release will occur only when anaerobic zones are included in systems to allow a selective advantage for PAOs, as discussed previously. As seen above, however, some of the rate expressions in Table 6.1 have questionable validity under anaerobic conditions. For this reason and because a model for the growth of only PAOs requires a matrix larger than the one in Table 6.1,[23,24,49,55,56] this process was not included here. International Water Association ASM No. 2d utilizes an expanded matrix to incorporate phosphorus removal by PAOs.[23] We will use it in Chapter 7 to see how reactor conditions affect phosphorus removal.

6.1.3 Representative Parameter Values in Model No. 1

The model depicted in Table 6.1 contains a large number of kinetic and stoichiometric parameters that must be evaluated for use in simulations. Techniques for conducting those evaluations will be discussed in Chapter 9. Although the model should be calibrated for each situation under study, it is acceptable to use "typical" parameter values to demonstrate fundamental principles concerning suspended growth cultures, provided the reader recognizes that the conclusions are general and not directly applicable to any specific situation. Typical parameter values for domestic sewage at neutral pH and 20°C were compiled for ASM No. 1.[21,22] Consideration of those values, as well as the values given in Chapter 3, has resulted in the list given in Table 6.3. They will be used here and in Chapter 7 to demonstrate several things about suspended growth cultures that could not be demonstrated with the simple model in Chapter 5.

6.1.4 Model Nos. 2 and 2d

Activated Sludge Model No. 2 incorporates all of the events included in ASM No. 1 plus biological phosphorus removal. The latter is very complex[49,55,56] and a large number of components must be

TABLE 6.3
Typical Parameter Values at Neutral pH and 20°C for Domestic Wastewater

Symbol	Units	Value
	Stoichiometric Coefficients	
Y_H	mg biomass COD formed/mg COD removed	0.60
f'_D	mg debris COD/mg biomass COD	0.08
$i_{N/XB}$	mg N/mg COD in active biomass	0.086
$i_{N/XD}$	mg N/mg COD in biomass debris	0.06
Y_A	mg biomass COD formed/mg N oxidized	0.24
	Kinetic Parameters	
$\hat{\mu}_H$	hr^{-1}	0.25
K_S	mg/L as COD	20
$K_{O,H}$	mg/L as O_2	0.10
K_{NO}	mg/L as N	0.20
$b_{L,H}$	hr^{-1}	0.017
η_g	dimensionless	0.8
η_h	dimensionless	0.4
k_a	L/(mg biomass COD·hr)	0.0067
k_h	mg COD/(mg biomass COD·hr)	0.092
K_X	mg COD/mg biomass COD	0.15
$\hat{\mu}_A$	hr^{-1}	0.032
K_{NH}	mg/L as N	1.0
$K_{O,A}$	mg/L as O_2	0.75
$b_{L,A}$	hr^{-1}	0.004

included to model it adequately, as was seen in Sections 2.4.6 and 3.7. Consequently, ASM No. 2 is considerably more complex than ASM No. 1.[23] Activated sludge model No. 2d is an important extension of ASM No. 2 that incorporates denitrifying PAOs, whose existence has been clearly demonstrated.[27,33,35,39] Activated sludge model No. 2d is much larger than ASM No. 1 and includes 19 components and 21 process rate equations that require 22 stoichiometric coefficients and 45 kinetic parameters.[24] Because of its size, ASM No. 2d will not be described in detail herein. However, because we will use it in Chapter 7, some of its major characteristics will be presented.

Some processes that were explicitly modeled in ASM No. 1 were simplified in ASM No. 2d to minimize its size. For example, processes 6 and 8 in Table 6.1, ammonification of soluble organic nitrogen and hydrolysis of particulate organic nitrogen, were eliminated. Their functions were made implicit by assuming that they occurred in stoichiometric proportion to soluble substrate removal and hydrolysis of slowly biodegradable organic matter. This accomplished the same thing as ASM No. 1, but with fewer process rate expressions. Organic phosphorus conversion to soluble phosphate was handled in a similar manner.

The events occurring under anaerobic conditions are quite different in the two models. Activated sludge model No. 1 assumed that growth and hydrolysis stopped under anaerobic conditions, although microbial death and lysis continued. This was adequate for the processes ASM No. 1 depicts, but is inadequate for biological phosphorus removal. Consequently, ASM No. 2d includes fermentation, uptake of acetate for formation of poly-β-hydroxyalkanoates (PHAs), release of soluble phosphate from hydrolysis of polyphosphate, and both chemical precipitation and dissolution of phosphate. The inclusion of fermentation required the partitioning of readily biodegradable substrate into two components, readily fermentable substrate and fermentation products, represented

by acetate. Acetate is produced from readily fermentable substrate under anaerobic conditions and is taken up by the PAOs, as depicted by Equation 3.83, forming PHA, as given by Equation 3.82. Under anoxic conditions (i.e., when nitrate-N is present as an electron acceptor), fermentation decreases and the common heterotrophic biomass consumes any residual acetate. As with the PAOs under aerobic conditions, growth of denitrifying PAOs under anoxic conditions is assumed to be supported exclusively by internal organic storage products like PHA and not residual acetate from fermentation. The nitrate flux, however, can be a determining factor in defining the degree to which denitrifying PAO activity occurs. For example, anoxic reactors receiving a nitrate mass load larger than the rate at which ordinary denitrifying heterotrophs can assimilate it are likely to support denitrifying PAOs.[27]

The scope of activities of the common heterotrophic bacteria was expanded. Under anaerobic conditions, they ferment readily fermentable substrate, producing acetate. However, they cannot grow. They can only grow under aerobic and anoxic conditions, and can use both readily fermentable substrate and acetate for that purpose. Unlike in ASM No. 1, low concentrations of ammonia, phosphate, and alkalinity can all reduce the growth rate of common heterotrophs under aerobic or anoxic conditions in ASM No. 2d. Finally, because heterotrophic growth cannot occur under anaerobic conditions, ASM No. 2d is not capable of modeling a totally anaerobic system. It can only mimic the performance of an anaerobic zone in a system with aerobic and anoxic zones.

6.1.5 Model No. 3

Activated Sludge Model No. 3[18] was a significant shift from ASM Nos. 1, 2, and 2d. It was designed to reduce the matrix to a small, manageable size (the most basic form that does not include phosphorus removal has 13 components and 12 process rate equations that require 15 stoichiometric coefficients and 21 kinetic parameters) and to include parameters that can be easily estimated in conjunction with field measurements. A significant change in ASM No. 3 is that it assumes that all organic substrate is stored intracellularly prior to being degraded to support growth under both aerobic and anoxic conditions. No direct consumption of substrate to support growth without an intermediate storage step is allowed in this model, thereby making the concentration of internal storage materials a key variable. Consequently, hydrolysis is deemphasized and its rate is assumed to be the same under different electron accepting conditions. Furthermore, ASM No. 3 does not use the interrelated and cyclic lysis:regrowth concept used in the other models. Rather, it assumes that the processes of maintenance, decay, endogenous respiration, lysis, predation, death, and so on are all lumped together and reflected in the traditional decay coefficient. Furthermore, unlike the previous versions of the model, decay rates are assumed to be slower under anoxic conditions, due in part to the fact that scavenging protozoa are less active in the absence of oxygen. Heterotrophic and autotrophic growth are modeled as distinct processes. Although ASM No. 3 is appropriate for modeling treatment plants that experience intracellular storage of substrate, there are many organic compounds, particularly in industrial wastewaters, that are not subject to intracellular storage. Both ASM Nos. 1 and 2d, on the other hand, are capable of adequately describing the steady-state and dynamic behavior of many treatment processes receiving a variety of wastewaters.[16]

6.1.6 Application of International Water Association Activated Sludge Models

Activated Sludge Models No. 1 and No. 2d are considerably more complex than the one used in Chapter 5 (Table 5.1). As a consequence, it is impossible to attain analytical solutions for the concentrations of the various constituents in a bioreactor. Rather, matrix solutions and numerical techniques must be used, depending on the complexity of the system under study. Several organizations have developed computer codes for solving the simultaneous mass balance equations for the constituents in the models, allowing their application to a variety of bioreactor configurations. Table 6.4 lists several computer codes implementing the IWA activated sludge models, and some provide

TABLE 6.4
Computer Codes Implementing IWA Activated Sludge Models

Code Name	Features	Contact Information
ASIM	A flexible modeling tool developed by Willie Gujer that implements ASM Model Nos. 1, 2d, and 3.	Educational Version: Eawag: Swiss Federal Institute of Aquatic Science and Technology P.O. Box 611 CH-8600 Dübendorf Switzerland http://www.asim.eawag.ch/index.htm Commercial Version: Holinger AG http://www.holinger.com/index.php?id=748&L=10
BioWin	Implements ASM Model Nos. 1, 2d, and 3, as well as other unit operations. Recent upgrades include a general purpose simulator.	EnviroSim Associates, Ltd. 7 Innovation Drive, Suite 205 Flamborough Ontario L9H 7H9 Canada www.envirosim.com
GPS-X	A general purpose simulator that implements ASM Model Nos. 1, 2d, and 3, as well as other unit operations.	Hydromantis, Inc. 1 James Street South Suite 1601 Hamilton Ontario L8P 4R5 Canada http://www.hydromantis.com/
WEST	Implements ASM Model Nos. 1, 2d, and 3, as well as other unit operations.	DHI Agern Allé 5 DK-2970 Hørsholm Denmark http://www.dhigroup.com/

academic packages. The code ASIM[17] allows multiple aerobic, anoxic and/or anaerobic reactors in series using both continuous flow and batch configurations. It performs steady-state and dynamic simulations and includes a menu-driven user interface. BioWin, GPS-X, and WEST are used commercially to a greater degree than ASIM. They include the elements in ASIM and much more (see each distributor's Web site for details). For instance, all three have user-friendly graphical interfaces with drag-and-drop capabilities, and all allow user-defined models to be added. The activated sludge simulations in this chapter and most of the multiple bioreactor simulations in Chapter 7 were done using SSSP (Simulation of Single-Sludge Processes),[7] which was a DOS-based application developed for implementation of ASM No. 1 on personal computers. Although no longer maintained, it was capable of both steady-state and dynamic simulations.

6.2 EFFECT OF PARTICULATE SUBSTRATE

A major limitation of the model presented in Chapter 5 is that it does not consider the biodegradation of particulate organic material, which is an important class of organic substrate in many wastewaters. Thus a suitable application of the model in Table 6.1 would be to see how the nature of the substrate influences the performance of a single CSTR, such as that depicted in Figure 5.1. To do this, two situations were considered, one in which all influent organic matter was soluble, and the other in which it was all particulate. The total concentration was the same in both cases, 500 mg/L

as COD, as was the flow rate, 1000 m^3/day (1.0 m^3 = 1000 L), giving a total COD mass input rate of 500 kg/day. The SRT/HRT ratio was held constant at 20 while the SRT was varied. In other words, the reactor volume was increased in proportion to the increase in SRT. This was done to make it easier to visualize the fate of particulate material, as well as the relative importance of growth and decay as SRT is changed. When the SRT/HRT ratio is held constant, if particulate material does not undergo reaction, it has a constant concentration in the system regardless of the SRT. Furthermore, as growth occurs on particulate substrate, the concentration of suspended matter (particulate substrate plus biomass) will decrease because the yield is less than one. In addition, growth associated with increased soluble substrate removal in the bioreactor is reflected by an increase in biomass concentration, whereas an increase in the importance of decay is reflected by a decrease. The biomass separator was assumed to be perfect so that it removed all undegraded particulate substrate from the effluent and returned it to the bioreactor. Thus, undegraded particulate substrate was removed only through the wastage flow.

The parameter values used to describe the reactions are those given in Table 6.3, with the exception of $\hat{\mu}_A$, the maximum specific growth rate for autotrophic bacteria, which was set equal to zero to eliminate them from consideration and limit the reactions to those associated with heterotrophs, as was done in Chapter 5. The influent contained sufficient ammonia-N for heterotrophic growth, but no nitrate-N, thereby eliminating nitrate-N, or reactions associated with it, from consideration as well. These selections simplified the model to include only processes 1, 4, and 7 and components 2, 3, 5, 7, and 8, as shown in Table 6.5. Finally, the DO concentration was held constant at 4.0 mg/L, which together with the $K_{O,H}$ value in Table 6.3, made the oxygen term in the rate expressions approach 1.0.

6.2.1 Steady-State Performance

Figure 6.1 shows the effect of SRT on the mixed liquor suspended solids (MLSS) concentration expressed as COD, the active fraction on a COD basis, and the oxygen requirement for bioreactors receiving the two types of substrate at a constant flow and concentration. The MLSS in the bioreactor receiving soluble substrate is composed of heterotrophic biomass and biomass debris, whereas the MLSS in the bioreactor receiving particulate substrate contains heterotrophic biomass, biomass debris, and unreacted particulate substrate, with the relative quantities depending on the SRT. The curves for the soluble feed are similar to those in Chapter 5 and can serve as a reference point with

TABLE 6.5
Process Kinetics and Stoichiometry Adopted for Considering the Impact of Particulate Substrate on the Performance of an Aerobic CSTR; Lysis:Regrowth Model for Biomass Loss

	Component[a]					
Process	X_S	$X_{B,H}$	X_D	S_S	S_O[b]	**Process Rate, r_j**
Growth		1		$-\frac{1}{Y_H}$	$\frac{1-Y_H}{Y_H}$	$\hat{\mu}_H\left(\frac{S_S}{K_S+S_S}\right)\left(\frac{S_O}{K_{O,H}+S_O}\right)X_{B,H}$
Death and lysis	$(1-f'_D)$	−1	f'_D			$b_{L,H}\cdot X_{B,H}$
Hydrolysis	−1			1		$k_h\left(\frac{X_S/X_{B,H}}{K_X+(X_S/X_{B,H})}\right)\left(\frac{S_O}{K_{O,H}+S_O}\right)X_{B,H}$

[a] All components and coefficients are expressed as COD.
[b] Coefficient must be multiplied by −1 to express as oxygen.

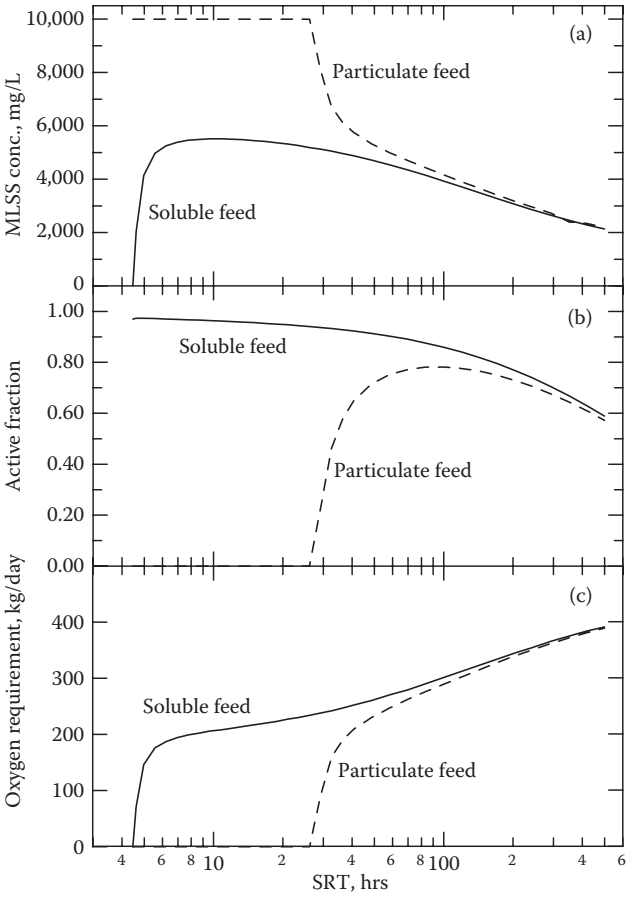

FIGURE 6.1 Effect of the nature of the organic matter in the feed on the performance of a CSTR with $\Theta_c/\tau = 20$. Influent biodegradable COD = 500 mg/L in each case. Flow = 1000 m^3/day. Particulate substrate was assumed to be removed by the biomass separator and retained in the reactor. Parameter values are listed in Table 6.3. The value of $\hat{\mu}_A$ was set equal to zero. The MLSS concentration is expressed in COD units; to convert to TSS units, divide by $i_{O/XB,T}$ = 1.2 mg COD/mg TSS.

which to see the effect of particulate feed. The most obvious effect is that a longer SRT is required to get biomass growth on particulate substrate. As indicated by the oxygen requirement, washout occurs at an SRT of about 4.5 hours when the substrate is soluble, but at an SRT of about 26 hours when it is particulate. This reflects the fact that hydrolysis reactions are slow. At SRTs below 26 hours, nothing happens to the particulate substrate and it acts like inert material, giving a MLSS concentration equal to the influent concentration times the SRT/HRT ratio ($500 \times 20 = 10{,}000$) and an active fraction of zero; that is, the MLSS is composed entirely of unreacted particulate substrate. As the SRT is increased past 26 hours, degradation of the particulate substrate begins, causing the MLSS concentration to drop and the active fraction and oxygen utilization to increase. In this region, the MLSS is composed of unreacted particulate substrate, active heterotrophic biomass, and some biomass debris. Eventually, at longer SRTs, bioreactor performance becomes independent of the feed type and is essentially the same for each. This occurs when both substrates are almost completely degraded so that system performance is governed primarily by biomass death and lysis, and the MLSS in each bioreactor is composed primarily of heterotrophic biomass and biomass debris.

Two significant points arise from Figure 6.1. The first is that long SRTs are required to achieve substantial degradation of particulate substrates. The second is that the use of the process loading

factor (F/M ratio) with a particulate substrate can be confusing. It will be recalled from Equation 5.50 that the active fraction must be known before the process loading factor can be related to the specific growth rate of the biomass. Figure 6.1b, however, shows that the active fraction varies in a complex manner as the SRT is changed when the substrate is particulate. This is because the active fraction is the active biomass concentration divided by the MLSS concentration, which includes the undegraded particulate substrate. The SRT, on the other hand, is still related to the biomass specific growth rate by Equation 5.21, and thus is more descriptive of process performance. Because most wastewaters contain some particulate substrate, SRT is preferable to process loading factor as a basic design and operational parameter for suspended growth bioreactors.

6.2.2 Dynamic Performance

So far we have only considered the steady-state performance of a CSTR; that is, the performance that results when a bioreactor receives a constant influent flow at a constant concentration. Most wastewaters are subject to time dependent variability, however, and thus it would be beneficial to investigate the impact of the nature of the substrate under those conditions.

Because of variations in human activities, municipal wastewater treatment systems experience diurnal variations in the flow and concentration of the wastewater entering them. Figure 6.2 shows typical variations experienced over a 24-hour period, beginning at midnight as time zero. The patterns correspond to those observed at a large municipal plant in South Africa over a period of one week,[13] but the values have been normalized to a daily average flow of 100 m^3/day and a flow-weighted average concentration of 100 mg/L. The patterns are also typical of those experienced in the United States[10] and will be adopted herein for demonstration purposes. As with kinetic and stoichiometric parameters, however, the necessity for determining the actual variations associated with a given wastewater cannot be overemphasized.

To determine the effect of the type of substrate on the dynamic response of a CSTR, the bioreactor in Figure 5.1 was subjected to the variations in flow and concentration shown in Figure 6.2. The flow values were adjusted to give a daily average flow of 1000 m^3/day and the biodegradable COD was adjusted to give a flow-weighted daily average concentration of 265 mg/L, a value commonly seen in U.S. domestic wastewater.[40] The SRT was set at 240 hours, a value sufficient to make the steady-state performance for the two substrate types essentially the same, as shown in Figure 6.1, while the HRT based on the daily average flow rate was set at 6 hours. As with the steady-state response, two situations were considered, one in which the organic matter was entirely soluble and one in which it

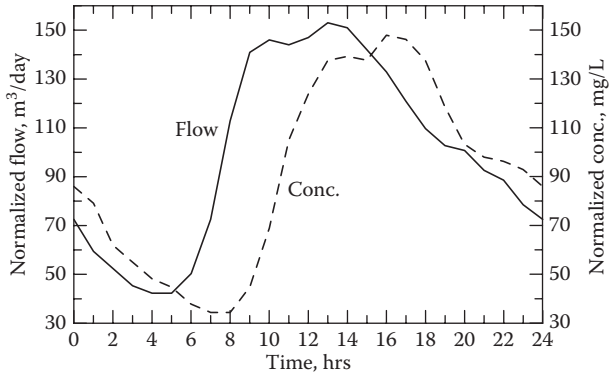

FIGURE 6.2 Typical diurnal patterns of wastewater flow and concentration for a community with little night time activity. (After Dold, P. L., and Marais G. v. R., Evaluation of the general activated sludge model proposed by the IAWPRC task group. *Water Science and Technology*, 18 (6): 63–89, 1986.) The flow has been normalized to an average of 100 m^3/day. The concentration has been normalized to give a flow-weighted average of 100 mg/L.

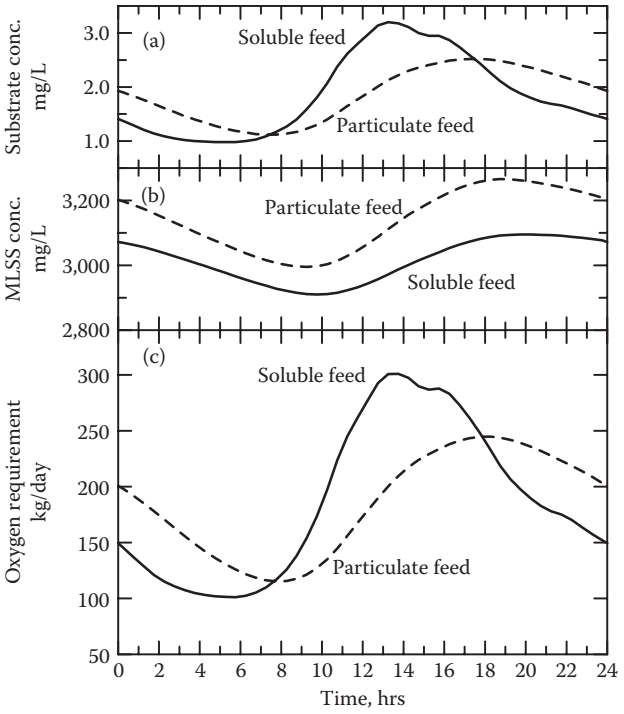

FIGURE 6.3 Effect of the nature of the organic matter in the feed on the response of a CSTR to the diurnal flow pattern in Figure 6.2. Flow-weighted average influent biodegradable COD = 265 mg/L in each case. Daily average flow = 1000 m^3/day; SRT = 240 hrs; average HRT = 6 hrs. Parameter values are listed in Table 6.3. The value of $\hat{\mu}_A$ was set equal to zero. The MLSS concentration is expressed in COD units; to convert to TSS units, divide by $i_{O/XB,T}$ = 1.2 mg COD/mg TSS.

was entirely particulate. All other constituents were assumed to be constant at concentrations sufficient to not limit the reactions. As before, the matrix in Table 6.5 described the system.

The response of the bioreactor is shown in Figure 6.3. Before considering the effect of the type of substrate on the bioreactor performance, we will examine only the soluble substrate case in order to understand why the bioreactor behaves as it does. Examination of the figure shows that the soluble substrate concentration has greater relative variations throughout the day than does the MLSS concentration. This is a direct result of the fact that the residence time of the MLSS in the bioreactor (i.e., the SRT) is much longer than the residence time of the soluble substrate (i.e., the HRT). As a result, variations in the MLSS concentration are dampened. In fact, the mass of MLSS in the bioreactor is almost constant throughout the day.

If we consider the mass of MLSS to be approximately constant throughout the day, we can then see why the soluble substrate concentration varies. Examination of Figure 6.2 reveals that the mass of substrate entering the bioreactor per unit time (flow × concentration) varies throughout the day. This means that the mass of substrate available to a unit mass of microorganisms also varies throughout the day. However, the rate at which a microorganism can remove substrate is controlled by the concentration of substrate surrounding it. Thus, as the mass flow rate of substrate into the bioreactor increases, the substrate concentration must rise to allow the microorganisms to remove substrate faster. Conversely, as the mass flow rate of substrate decreases, the microorganisms will drive the substrate concentration lower until the rate of substrate removal is decreased to be consistent with the rate of input. Thus, the variation in substrate concentration is a direct consequence of the necessity for the microorganisms to vary their activity in response to the changing input rate of substrate. The variation in the oxygen requirement directly reflects that variation in activity.

The soluble substrate curves in Figure 6.3a also demonstrate an important point about the growth characteristics of the biomass in a CSTR receiving a time varying input; the specific growth rate is not constant over time. It will be recalled that the specific growth rate is controlled by the soluble substrate concentration, as expressed by the Monod equation (Equation 3.36). Since the soluble substrate concentration is varying over time, so is the specific growth rate. This means that the bacteria are in a continually changing state. It will also be recalled that for the reactor configuration in Figure 5.1, the SRT is determined solely by the reactor volume and the wastage flow rate. Consequently, the SRT can be held constant, even though the specific growth rate of the bacteria is varying. In other words, the specific growth rate of the microorganisms in a CSTR that is not at steady state is not fixed by the SRT. This can also be seen by performing a mass balance on heterotrophic biomass in the reactor. Such an exercise using the kinetics and stoichiometry in Table 6.5 reveals:

$$\mu_H = \frac{1}{\Theta_c} + b_{L,H} + \frac{1}{X_{B,H}} \frac{dX_{B,H}}{dt}. \tag{6.1}$$

Since the heterotrophic biomass concentration varies over time in response to the changing input, so will the specific growth rate. It is important to recognize that the constant relationship between SRT and specific growth rate depicted in Equation 5.21 is only valid for steady-state conditions.

Even though the specific growth rate is not determined solely by the SRT for a nonsteady-state CSTR, the SRT is still a good indicator of average performance, with longer SRTs giving lower average substrate concentrations. Nevertheless, the average substrate concentration leaving a CSTR receiving a dynamic input will always be greater than the concentration leaving a steady-state CSTR. For the conditions imposed in Figure 6.3, the flow-weighted average output concentration for the soluble substrate case is 2.11 mg/L as opposed to 1.80 mg/L for a steady-state CSTR with the same SRT. The higher average substrate concentration results from the nonlinear nature of the Monod equation (Equation 3.36) describing the relationship between substrate concentration and microbial activity. This is one reason why it is advantageous to practice equalization prior to a biochemical reactor.

Now consider the impact of the type of substrate on the dynamic behavior of the CSTR. Figure 6.3a shows the effluent soluble substrate concentrations resulting from the two feed types. When the influent feed is all particulate, soluble substrate arises from hydrolysis of the particulate substrate, which is a slow reaction. Thus, the response is dampened and the flow-weighted average concentration is lower (1.92 mg/L vs. 2.11 mg/L). Figure 6.3b shows that there will be little difference in the MLSS concentration or its variation for the two substrate types. The slightly higher concentration in the bioreactor receiving particulate substrate is because of a slight buildup of that substrate in the system caused by the slow hydrolysis reactions, but the effect is small. Generally, one would not expect to be able to distinguish much difference in the amount of MLSS in the two systems.

The major impact of particulate substrate on the dynamic response of a CSTR is in the utilization of oxygen, as shown in Figure 6.3c. As might be expected from the previous discussion of the slow nature of hydrolysis, the impact of the presence of particulate substrate is to dampen the system response, thereby reducing the peak oxygen requirement. In addition, the need for hydrolysis to make substrate available causes a time lag in the occurrence of the maximum and minimum oxygen consumption rates. Examination of Figure 6.2 reveals that the minimum mass input rate occurs at about 6 hours and the maximum at about 13 hours, which correspond closely to the times of the minimum and maximum oxygen requirements in the bioreactor receiving soluble substrate, thereby demonstrating the rapidity with which biomass can respond to soluble substrate. In contrast, the need for hydrolysis of particulate substrate, in combination with the fact that its concentration does not vary rapidly because its residence time in the system is the SRT, delays the minimum response by about 3 hours and the maximum by about 5 hours. Since both systems use about the same amount of oxygen in a 24-hour period, this suggests that consideration must be given to the physical state of the substrate during design of the oxygen transfer system for a suspended growth bioreactor.

6.3 NITRIFICATION AND ITS IMPACTS

We saw in Section 3.2.6 that the kinetics of growth of autotrophic nitrifying bacteria could be represented in the same manner as that of heterotrophic bacteria. Consequently, the general conclusions derived in Chapter 5 about biomass growth in CSTRs are equally applicable to them. We also saw in Section 3.2.10 that the values of their kinetic coefficients were quite different from those for heterotrophs. This means that the specifics of their behavior in a given reactor environment will differ somewhat from that of the heterotrophs. In this section we will investigate some of the characteristics of nitrifying bacteria that require special recognition and see how autotrophic and heterotrophic bacteria might influence one another when grown in the same bioreactor. To simplify this demonstration, AOBs and NOBs will be considered collectively as "nitrifying bacteria."

6.3.1 Special Characteristics of Nitrifying Bacteria

Comparison of the typical $\hat{\mu}$ values for heterotrophic and autotrophic bacteria in Table 6.3 shows that the value for autotrophs is almost an order of magnitude lower than that for heterotrophs, suggesting that the minimum SRT for nitrifying bacteria is almost an order of magnitude larger. As a consequence, they can be lost from bioreactors under conditions that allow heterotrophic bacteria to grow freely. This situation is aggravated by the fact that nitrifying bacteria are more sensitive to low temperatures and low DO concentrations, as will be discussed later in this section. Therefore, special consideration must be given to the choice of the SRT in systems containing autotrophic bacteria, and it cannot be assumed that conditions suitable for the removal of soluble organic matter are suitable for the conversion of ammonia-N to nitrate-N.

Another characteristic of nitrifying bacteria is that their half-saturation coefficient (K_{NH}) is very low; the typical value given in Table 6.3 is 1.0 mg/L as N. It will be recalled that the half-saturation coefficient is the substrate concentration allowing bacteria to grow at half of their maximal rate. This means that CSTRs containing autotrophic bacteria will have low ammonia-N concentrations even when the bacteria are growing relatively fast and that very low concentrations will result whenever the SRT is large enough to ensure stable growth. It also means that the ammonia-N concentration will rise rapidly as the SRT is decreased to the point of washout. As a consequence, nitrification has gained the reputation of being an all-or-none phenomenon. In other words, since the nitrogen concentration entering municipal wastewater treatment systems is on the order of 30 to 40 mg/L, the percentage of nitrification approaches 100 whenever the SRT is long enough to give stable growth and rapidly falls to zero as washout occurs. This is illustrated in Figure 6.4,[43] which includes data from the literature as well as the results of steady-state simulations with a model similar to the one employed herein.

There are some occasions in which the all-or-none phenomenon will not occur or will not be as drastic as that shown in Figure 6.4. One is when the bioreactor is subjected to diurnal loading, like that shown in Figure 6.2. It will be recalled from Section 6.2.2 that the flow-weighted average effluent substrate concentration from a CSTR receiving a diurnal input is not as low as the concentration from the same reactor receiving a constant input. The degree of difference between the two responses depends on the bioreactor SRT, as shown in Figure 6.5,[43] which was prepared by simulation with a model similar to the one presented herein. It also shows that complete nitrification can still be achieved in a CSTR receiving a diurnal input but that longer SRTs are required. Another situation that can lead to incomplete nitrification is when the influent contains such a high ammonia-N concentration that both substrate and product inhibition can occur.[2,44] In that case the attainment of complete nitrification requires the use of multireactor systems in which denitrification can be used to reduce the product concentration (i.e., nitrite and nitrate). Although progress has been made in the modeling of substrate and product inhibition during nitrification,[26] pilot studies should always be run on wastes containing ammonia-N concentrations substantially higher than those found in domestic wastewater. A final circumstance in which a curve of a different shape may be obtained is when

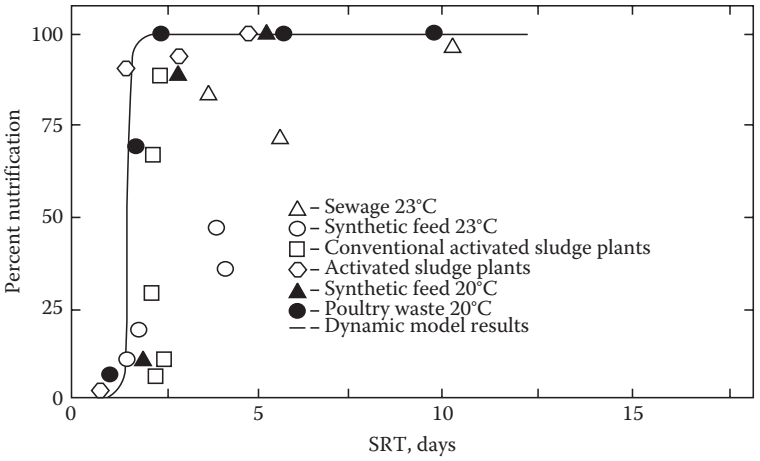

FIGURE 6.4 Effect of SRT on the steady-state nitrification performance of a CSTR. (Adapted from Poduska, R. A., and Andrews, J. F., Dynamics of nitrification in the activated sludge process, *Proceedings of the 29th Industrial Waste Conference*, Purdue University Engineering Extension Series No. 145, 1005–25, Purdue University, West Lafayette, IN, 1974, where data sources are given.)

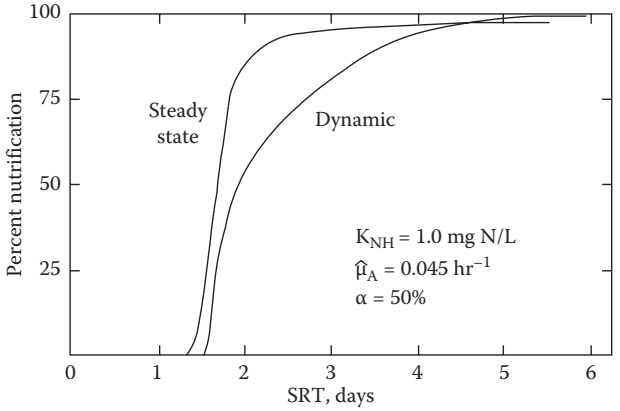

FIGURE 6.5 Comparison of steady-state and dynamic performance of nitrification in a CSTR. (Adapted from Poduska, R. A., and Andrews, J. F., Dynamics of nitrification in the activated sludge process, *Proceedings of the 29th Industrial Waste Conference*, Purdue University Engineering Extension Series No. 145, 1005–25, Purdue University, West Lafayette, IN, 1974.)

the wastewater contains chemicals inhibitory to the autotrophic bacteria. Such a curve is shown in Figure 6.6.[8] Inhibitory compounds can reduce $\hat{\mu}_A$ and increase K_{NH}, both of which reduce the percentage of nitrification achieved at a given SRT, thereby making the curve approach 100% more gradually. Both effects are illustrated in Figure 6.6, which were obtained with an industrial waste.[8]

Another important characteristic of nitrifying bacteria is their extreme sensitivity to the DO concentration, which necessitated the inclusion of a term for oxygen in the rate expression in Table 6.1. An examination of the values of the oxygen half-saturation coefficient for autotrophs ($K_{O,A}$) and heterotrophs ($K_{O,H}$) in Table 6.3 shows that $K_{O,A}$ is much larger than $K_{O,H}$. This means that as the DO concentration (S_O) is decreased, the term $S_O/(K_O + S_O)$ in the rate expression for autotrophs will become small more rapidly than the corresponding term in the rate expression for heterotrophs. Consequently, autotrophs will be affected by decreases in the oxygen concentration much more drastically than will heterotrophs.

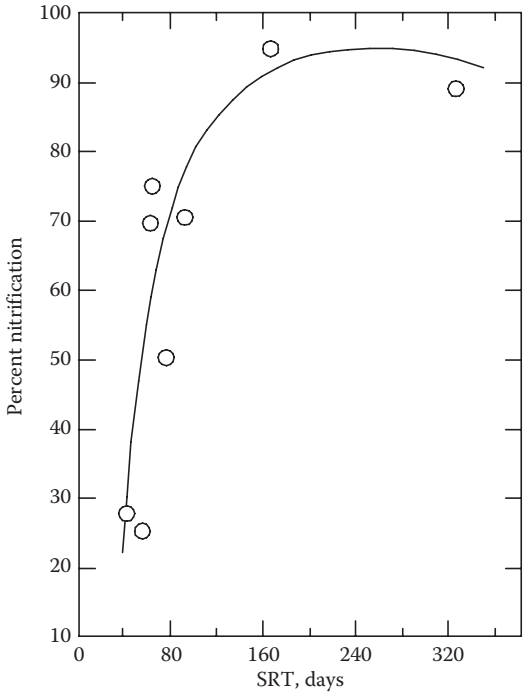

FIGURE 6.6 Effect of SRT on the performance of nitrification in a CSTR receiving a wastewater containing inhibitory compounds. (Adapted from Bridle, T. R., Climenhage, D. C., and Stelzig, A., Start-up of a full scale nitrification-denitrification treatment plant for industrial waste. *Proceedings of the 31st Industrial Waste Conference*, *1976, Purdue University*, 807–15, Ann Arbor Science Publishers, Ann Arbor, MI, 1977.)

The importance of DO concentration to the growth of autotrophs is illustrated in Figure 6.7 where the minimum SRT required for their growth is plotted as a function of the oxygen concentration. The curve was generated with Equation 5.25, with an adjustment of $\hat{\mu}_A$ for the effects of oxygen as given by the rate expression in Table 6.1. The parameter values from Table 6.3 were used, as indicated in the figure. Examination of the figure shows that oxygen concentrations above 2.0 mg/L have little effect on the minimum SRT, and thus it is seldom necessary to maintain the concentration in excess of that value to get satisfactory nitrification. However, oxygen concentrations below

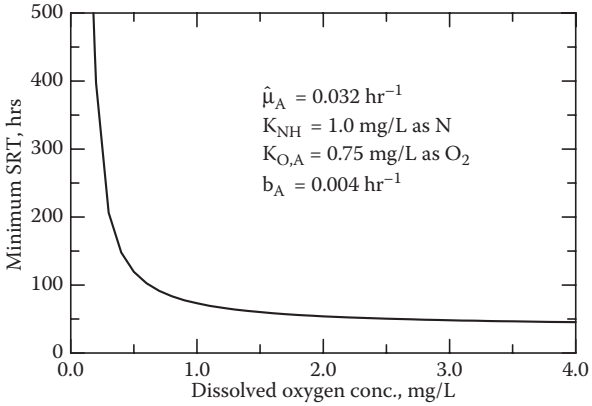

FIGURE 6.7 Effect of dissolved oxygen concentration on the minimum SRT required for nitrification in a CSTR receiving an ammonia-N concentration of 30 mg/L. Parameter values are for a temperature of 20°C.

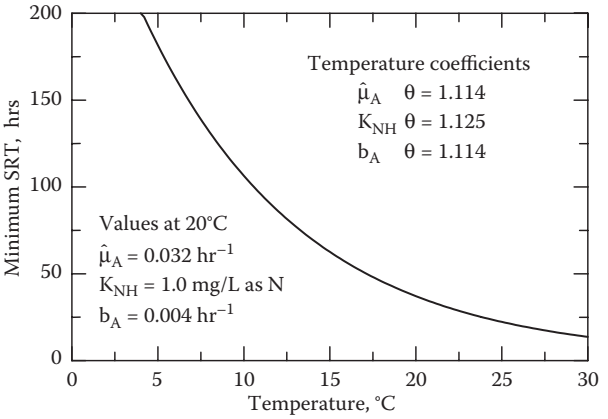

FIGURE 6.8 Effect of temperature on the minimum SRT required for nitrification in a CSTR receiving an ammonia-N concentration of 30 mg/L. The dissolved oxygen concentration is not limiting.

2.0 mg/L begin to have a strong effect and those below 0.5 mg/L have a drastic effect. A low DO concentration also diminishes the percent nitrification that can ultimately be achieved. Although not reflected in the current IWA activated sludge models, a low DO concentration can also disrupt the balance between AOBs and NOBs and lead to nitrite accumulation, since K_{OA} for NOB is higher than K_{OA} for AOB,[9] as discussed in Section 3.2.10. Consequently, care in the specification of the oxygen transfer system is an important component of the design of a suspended growth bioreactor in which nitrification is to occur.

Nitrifying bacteria are also very sensitive to temperature, as reflected by the temperature coefficients in Section 3.9.2. Because $\hat{\mu}_A$ is small even at 20°C, low temperatures require bioreactors to have very long SRTs in order for nitrifying bacteria to grow. This is illustrated in Figure 6.8, which was generated with Equation 5.25 using kinetic parameters corrected for temperature with Equation 3.99 and the coefficients shown in the figure. The DO concentration was assumed to be high enough to have no effect. Examination of the figure reveals that very large increases in the SRT are required to compensate for drops in temperature. This suggests that the choice of SRT for a nitrifying bioreactor must be made for the lowest temperature expected.

The model in Table 6.1 includes alkalinity as a component because of the sensitivity of nitrifying bacteria to low pH and the effect that the oxidation of ammonia-N has on alkalinity, and hence on pH. Equations 3.28 and 3.29 showed that 8.64 g of HCO_3^- alkalinity are destroyed for each g of nitrogen oxidized from ammonia-N to nitrate-N or 7.08 g of alkalinity (as $CaCO_3$) per g of N. Domestic wastewaters often contain between 30 and 40 mg/L of nitrogen that can be oxidized during nitrification. This means that almost 300 mg/L of alkalinity (as $CaCO_3$) can be destroyed. Since many wastewaters do not contain this much alkalinity, it is apparent why consideration must be given to pH control in nitrifying bioreactors. Equation 3.50 may be used to determine the effect of pH on $\hat{\mu}_A$, with its resulting impact on the minimum SRT required to achieve nitrification. Such an activity is left as an exercise for the reader.

6.3.2 Interactions between Heterotrophs and Autotrophs

From the discussion of the preceding section, it is apparent that there are several interactions between heterotrophic and autotrophic bacteria growing together in a CSTR. The most important concerns the DO concentration, which is why it is explicitly included in the model in Table 6.1. Because heterotrophic bacteria have larger $\hat{\mu}$ and smaller K_O values than autotrophs, they are capable of surviving in bioreactors with SRTs and oxygen concentrations that would cause autotrophic bacteria to wash out. In other words, the heterotrophs control the DO concentration if the supply rate is

insufficient to meet the needs of both groups. As a result, organic substrate removal can occur unimpeded under conditions that will not allow nitrification. Looked at another way, it is possible for a designer to choose an SRT and DO concentration that will allow only organic substrate removal or a combination that will allow both nitrification and organic substrate removal, but it is impossible to choose conditions that will allow nitrification without oxidation of any organic matter present. Since the designer can choose the conditions to be imposed on the bioreactor, he/she has control over the events that will occur in it, consistent with the kinetic characteristics of the organisms involved.

Simultaneous growth of heterotrophic and autotrophic bacteria need not be detrimental to the nitrifiers; it can be beneficial. As discussed previously, nitrifying bacteria are sensitive to inhibition by some organic compounds. If those compounds are biodegradable, it is possible to choose the SRT so that their concentrations in the bioreactor are too low to inhibit the nitrifiers, thereby allowing them to grow along with the heterotrophs destroying the inhibitors.[32] Alternatively, if the inhibiting compounds can be degraded under anoxic conditions, it could be possible to use an anoxic/aerobic sequential treatment scheme (e.g., Figure 1.13) to degrade the inhibitor before the nitrification stage occurs.[3] These examples show that it is not necessary to perform nitrification in a separate bioreactor following destruction of the organic matter.

As stated before, domestic wastewater in the United States often contains about 265 mg/L of biodegradable COD and around 40 mg/L of reduced nitrogen.[40] Comparison of the Y_H and Y_A values in Table 6.3 in light of these concentrations reveals that the amount of autotrophic biomass formed in a bioreactor receiving such a feed will be small relative to the amount of heterotrophic biomass formed. Indeed, biomolecular methods coupled with modeling have shown that nitrifier abundance in domestic activated sludge plants is typically less than 8% of the total bacterial community.[11,46] Furthermore, when consideration is given to the fact that the MLSS in a suspended growth bioreactor contains inert particulate organic matter as well as biomass debris, it is likely that the contribution of the autotrophic bacteria to the MLSS concentration will be even smaller. Using modeling, we can demonstrate that this is indeed the case, as illustrated in Figure 6.9a. The curves in this figure were generated with the model in Table 6.1 using as input a feed with the characteristics listed in Table 6.6, which are considered to be representative of a domestic wastewater in the United States following primary sedimentation.[7] The curve without nitrification was obtained by setting $\hat{\mu}_A$ equal to zero during the simulation.

Unlike their effect on the MLSS concentration, however, nitrifying bacteria have a major impact on the amount of oxygen required in a CSTR receiving a feed like domestic wastewater. This can be seen by considering the stoichiometry of microbial growth as reflected by the coefficients in Table 6.1. For the true growth yields in Table 6.3, the oxygen utilization associated with heterotrophic growth (exclusive of biomass death and lysis) is 1.00 – 0.60 or 0.40 g O_2/g COD removed, whereas that associated with autotrophic growth is 4.57 – 0.24 or 4.33 g O_2/g N oxidized. Death and lysis increase the utilization for the heterotrophs, perhaps to around 0.65 g O_2/g COD removed for a typical SRT, but have little impact on the autotrophic oxygen utilization because the value of the true growth yield (0.24) is small relative to 4.57. Given the concentrations of the two substrates in the influent, these values suggest that the autotrophs require around 173 mg of oxygen (40×4.33) for each liter of influent whereas the heterotrophs require around 172 mg (265×0.65). Thus, even though the concentration of reduced nitrogen entering the bioreactor is much less than the concentration of biodegradable COD entering, the large change in the oxidation state of nitrogen associated with the production of nitrate, in combination with the differences in biomass yield, means that similar amounts of oxygen are required by the two groups of microbes. This can be seen clearly in Figure 6.9b. Failure to consider this effect during bioreactor design will lead to an inability to transfer sufficient oxygen, with resulting impairment of nitrification performance, as discussed previously.

Because of the important effect of nitrification on the oxygen requirement in a bioreactor, there are circumstances in which it would be advantageous to be able to calculate it explicitly. This can be done by using the simple model in Chapter 5 with minor modification. For the simple decay

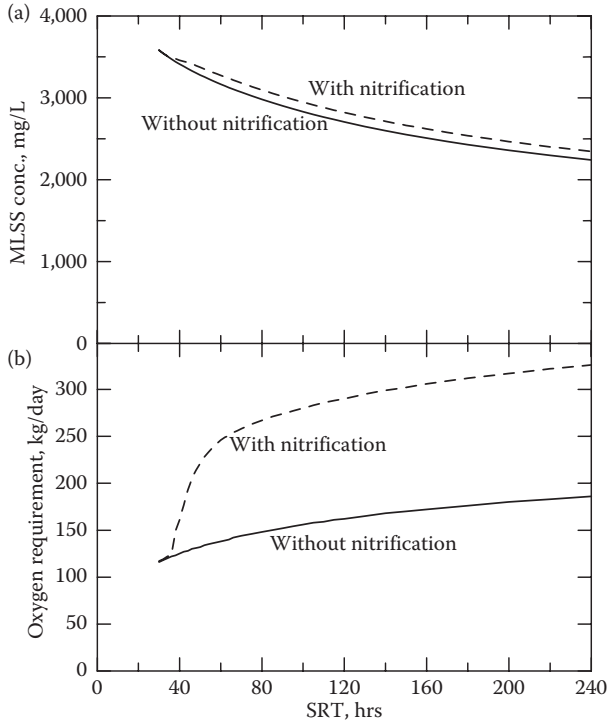

FIGURE 6.9 Effects of SRT and nitrification on the MLSS concentration and oxygen requirement in a CSTR with $\Theta_c/\tau = 20$. Parameter values are given in Table 6.3 and the influent conditions are given in Table 6.6. The value of $\hat{\mu}_A$ was set equal to zero to eliminate nitrification in one case. The dissolved oxygen concentration was fixed at 4.0 mg/L. The MLSS concentration is expressed in COD units; to convert to TSS units, divide by $i_{O/XB,T} = 1.2$ mg COD/mg TSS.

TABLE 6.6
Characteristics Considered to Be Representative of a United States Domestic Wastewater following Primary Sedimentation

Component	Concentration
Inert particulate organic matter	35.0 mg/L as COD
Slowly biodegradable substrate	150.0 mg/L as COD
Readily biodegradable substrate	115.0 mg/L as COD
Oxygen	0.0 mg/L as O_2
Soluble nitrate nitrogen	0.0 mg/L as N
Soluble ammonia nitrogen	25.0 mg/L as N
Soluble biodegradable organic nitrogen	6.5 mg/L as N
Particulate biodegradable organic nitrogen	8.5 mg/L as N
Alkalinity	5.0 mmol/L

Note: Data from Bidstrup, S. M., and Grady, C. P. L., Jr., SSSP—Simulation of single-sludge processes. *Journal, Water Pollution Control Federation,* 60: 351–61, 1988.

model, an equation analogous to Equation 5.43 can be used by recognizing that the factor 1.0 must be replaced with 4.57, as discussed in Section 6.1.1:

$$RO_A = F(S_{NHO} - S_{NH})\left[4.57 - \frac{(1 + f_D \cdot b_A \cdot \Theta_c)Y_{A,T} \cdot i_{O/XB,T}}{1 + b_A \cdot \Theta_c}\right]. \quad (6.2)$$

Where $Y_{A,T}$ is the autotrophic true growth yield on a TSS/N basis and S_{NHO} represents the ammonia-N concentration in the feed that is available to the nitrifiers; that is, the amount remaining after consideration of the amount used by the heterotrophs for biomass synthesis.

6.3.3 Effects of Nitrification in Bioreactors Receiving Only Biomass

In Section 5.2.5 and Figure 5.10, we investigated the performance of a CSTR receiving only biomass in the feed. Because the model in Chapter 5 considered only the growth of heterotrophic biomass, the discussion was limited to the fate of heterotrophic bacteria alone under such circumstances. However, if the SRT of the bioreactor is sufficiently long, autotrophic bacteria will also grow, with an important impact on system performance. As expected from the preceding discussion, the major impacts of autotrophs on a bioreactor receiving only biomass in its feed are on the oxygen requirement and the pH. If f_D is 0.20, the destruction of 100 g of biomass COD will lead to the release of 20 g of biomass debris COD and the consumption of 80 g of oxygen. As reflected by $i_{N/XB}$ and $i_{N/XD}$, the nitrogen contents of biomass and biomass debris are around 0.086 g N/g biomass COD and 0.06 g N/g debris COD, respectively. Consequently, in addition to producing 20 g of biomass debris COD, destruction of 100 g of biomass will also lead to the release of 7.4 g of ammonia-N (100[0.086 – (0.20 × 0.06)]), which can act as substrate for autotrophic bacteria. If the oxygen requirement associated with oxidation of the ammonia-N is 4.33 g O_2/g N, growth of the autotrophs will increase the reactor's oxygen requirement by 40%. This can have a substantial impact on the amount of oxygen that must be supplied. Furthermore, oxidation of the nitrogen released by destruction of the biomass will result in the destruction of 52.4 g of alkalinity (as $CaCO_3$) for each 100 g of biomass COD destroyed. Since it is not unusual for the feed stream entering such a reactor to contain 4000 mg/L of biomass COD and for 50% of it to be destroyed, one can see that the occurrence of nitrification in the reactor will destroy substantial quantities of alkalinity. Consequently, it is often necessary to control pH in such systems. Failure to do so will create an unstable environment in which the pH oscillates, thereby hindering system performance.

6.4 DENITRIFICATION AND ITS IMPACTS

In the context of biochemical operations, anoxic growth of heterotrophic bacteria is simply an alternative mode of growth in response to the absence of oxygen and the presence of nitrate-N as the terminal electron acceptor. Because the resulting reduction of nitrate-N to N_2 removes nitrogen from the wastewater undergoing treatment, the process is also referred to as denitrification. Generally, it occurs when the appropriate conditions are purposely established in bioreactors, but as the rate expression in Table 6.1 suggests, it will occur any time the oxygen concentration is low, the nitrate concentration is high, and organic matter is present as an electron donor. This means that the potential for its occurrence exists in all biochemical reactors. Thus it should be included as a possible reaction in any complete system model.

6.4.1 Characteristics of Denitrification

In Section 2.4.1 we saw that four enzymes are involved in the reduction of nitrate-N to N_2: nitrate reductase, nitrite reductase, nitric oxide reductase, and nitrous oxide reductase. Nevertheless, just as nitrification was treated as a single-step process in ASM No. 1, so is denitrification. This simplification

will not seriously restrict the results of simulations for wastewaters with the characteristics of domestic wastewater. However, the simplification can lead to serious errors when simulation of the treatment of industrial or agricultural wastewaters with high nitrogen concentrations is being performed. Consequently, models have been developed that incorporate multiple denitrification steps.[26,53] Furthermore, in response to an increasing number of treatment technologies that target the generation of nitrite as an intermediate, the IWA activated sludge models have been expanded to include denitrification via nitrite,[14,28,29] although those modifications will not be elaborated upon here.

Denitrification is usually practiced to remove nitrogen from a wastewater, but that is not the only reason for including it in a treatment system.[13] In the previous section we saw that the inclusion of nitrification in a biochemical reactor requires that substantial amounts of oxygen be supplied. The electron accepting potential of that oxygen is not entirely lost, however, because much of it still resides in the nitrate produced. For example, 4.57 g O_2 are needed to convert a g of ammonia-N to nitrate-N through nitrification, but each g of nitrate-N has the electron accepting capacity of 2.86 g O_2 as shown in Table 3.1. Thus, if there were ways to use that electron accepting capacity in the oxidation of some of the organic matter in the wastewater, almost 63% of the energy expended in converting the ammonia-N to nitrate-N could be recovered. The development of treatment systems in which this can be done has been a major environmental engineering success, and systems that employ this approach will be discussed in more detail in Chapter 12. We also saw in the previous section that nitrification results in the destruction of substantial quantities of alkalinity, 7.08 g (as $CaCO_3$) per g of ammonia-N oxidized. Denitrification, on the other hand, destroys hydrogen ions and produces carbon dioxide, as shown in Equation 3.22. This has the net effect of increasing alkalinity. As can be seen from the stoichiometric coefficient for alkalinity production in Table 6.1, approximately 1/14 moles of alkalinity will be produced for each gram of nitrate-N reduced to N_2. This amounts to approximately 3.5 g of alkalinity (as $CaCO_3$) per g of nitrate-N. Thus, about half of the alkalinity lost through the oxidation of ammonia-N to nitrate-N can be recovered during the subsequent reduction of nitrate-N to N_2. Through appropriate configuration or operation of the bioreactor it is possible to make use of this to reduce the amount of chemical that must be purchased for pH control.[13] We will see later how this can be done.

6.4.2 Factors Affecting Denitrification

Even though denitrification is usually practiced as part of a treatment system within which the nitrate is formed by nitrification, it is easier to understand the unique characteristics of denitrification if it is considered in isolation. Such a consideration is not artificial because some industrial wastewaters contain high concentrations of nitrate and biological treatment is one option open to the engineer for its removal. Thus, in this section we will consider the growth of biomass in a CSTR receiving an influent containing nitrate. We will assume that the influent also contains sufficient ammonia-N to meet the biosynthetic needs of the biomass so that the only role of the nitrate-N is as the terminal electron acceptor. This means that the stoichiometry depicted in Table 6.1 is applicable.

Comparison of process 1 in Table 6.1 (aerobic growth of heterotrophs) with process 2 (anoxic growth of heterotrophs) reveals that the COD-based stoichiometry is the same. This was also seen in Chapter 3 when Equation 3.24 was compared to Equation 3.8. Thus, for the utilization of a given amount of organic matter in a CSTR, the amount of electron acceptor used (on a COD basis) will be the same for both growth conditions. This suggests that the oxygen requirement curves that we have looked at in preceding sections can also be expressed as nitrate-N requirement curves, as long as the nitrate-N is expressed on a COD or O_2 equivalent basis through use of the factor 2.86 g O_2/g N (or –2.86 g COD/g N) as shown in Table 3.1. This, in turn, tells us that the amount of nitrate-N that will be removed by growth on a given amount of influent substrate will depend on the SRT of the bioreactor. The longer the SRT, the greater the fraction of the electrons in the substrate that will go to the acceptor and the greater the amount of nitrate-N that will be reduced. Thus, one factor affecting the removal of nitrate-N through denitrification is the SRT of the reactor in which the biomass is growing.

When considering denitrification it is necessary to reverse your thinking. During aerobic growth of heterotrophs, the goal is removal of the organic substrate and the electron acceptor is supplied in excess. However, during anoxic growth of heterotrophs, the goal is the removal of the electron acceptor, and sufficient organic substrate (i.e., electron donor) must be supplied to accomplish this. Consequently, engineers focus on the ΔS/ΔN ratio, the amount of substrate COD that must be supplied to remove a given amount of nitrate-N, rather than on the electron acceptor requirement as was done for aerobic bioreactors. Consideration of the factors discussed in the preceding paragraph suggests that ΔS/ΔN varies with SRT in a manner opposite to the way in which the oxygen requirement varies with the SRT in an aerobic bioreactor. Returning to Chapter 5, in which the traditional approach to decay was used, Equation 5.43 allowed the calculation of the amount of oxygen (electron acceptor) used:

$$RO = F(S_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T} \cdot i_{O/XB,T}}{1 + b_H \cdot \Theta_c}\right]. \tag{5.43}$$

Since RO represents the amount of electron acceptor used, and since each g of nitrate-N can accept as many electrons as 2.86 g of oxygen, the equation can be rewritten as

$$2.86 \cdot F(S_{NOO} - S_{NO}) = F(S_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T} \cdot i_{O/XB,T}}{1 + b_H \cdot \Theta_c}\right]. \tag{6.3}$$

Recognition that $F(S_{NOO} - S_{NO})$ is ΔN and that $F(S_{SO} - S_S)$ is ΔS, gives the following expression for ΔS/ΔN using the traditional approach to decay:

$$\frac{\Delta S}{\Delta N} = \frac{2.86(1 + b_H \cdot \Theta_c)}{1 + (b_H \cdot \Theta_c) - (Y_{H,T} \cdot i_{O/XB,T})(1 + f_D \cdot b_H \cdot \Theta_c)}. \tag{6.4}$$

Because of the cycling of carbon in the lysis:regrowth approach to decay, it is not possible to derive an analogous expression using that approach, but the same principles apply and thus Equation 6.4 is important for showing that ΔS/ΔN varies with SRT in a manner opposite to that of the oxygen requirement. In other words, the amount of electron donor required to remove a given amount of nitrate-N will decrease as the SRT is increased. This happens because the increased importance of decay at longer SRTs allows a greater fraction of the electrons available in the donor to go to the acceptor rather than into biomass. Furthermore, it can be seen that the magnitude of this effect depends on the values of the kinetic and stoichiometric coefficients describing the system. The effect of SRT is illustrated in Figure 6.10[15] with data collected from a CSTR receiving methanol as the electron donor. Although ΔS was expressed as the mass of methanol rather than as the mass of COD, it is clear that the observations confirm the theory.

The calculation of the ΔS/ΔN ratio above is based on the assumption that nitrate-N is the only available electron acceptor. In practice, however, it is difficult to totally eliminate the entrance of oxygen into a bioreactor. Since oxygen is the preferred electron acceptor, any entering the bioreactor will increase the amount of electron donor that must be added to reduce a given amount of nitrate-N. In other words, it will increase the ΔS/ΔN ratio.

The impact of the input of oxygen into a denitrification reactor can best be seen through the use of the model in Table 6.1 because it considers both aerobic and anoxic growth of the heterotrophic bacteria and the effect of DO on each. To illustrate this point, a wastewater like that in Table 6.6 was assumed to be entering a CSTR operated at an SRT of 240 hours. The kinetic and stoichiometric coefficients describing the biomass in the bioreactor were assumed to have the values given in Table 6.3, except for $\hat{\mu}_A$, which was set equal to zero to ensure only heterotrophic reactions.

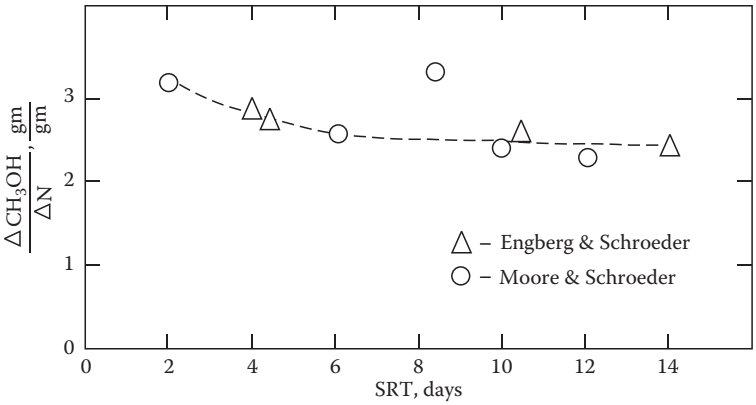

FIGURE 6.10 Effect of SRT on the ΔS/ΔN ratio. (Reprinted from Engberg, D. J., and Schroeder, E. D., Kinetics and stoichiometry of bacterial denitrification as a function of cell residence time. *Water Research*, 9, 1051–54, 1975. Copyright © Elsevier Science Ltd. With permission.)

Two situations were considered. In one, the influent nitrate-N concentration was set equal to 50 mg/L, which gave a mass input rate equivalent to 143 kg/day of oxygen. Given the values of the stoichiometric coefficients in Table 6.3, the amount of biodegradable COD entering the bioreactor (265 kg/day) was in excess of that needed to meet the required ΔS/ΔN ratio. This case is called the excess COD case. In the other, the influent nitrate-N concentration was set equal to 60 mg/L, giving a mass input rate equivalent to 172 kg/day of oxygen, which was more than could be completely removed by the available COD. This is called the limiting COD case. Simulations were then conducted in which the rate of oxygen transfer into the bioreactor was set at various values and the results are shown in Figure 6.11. For the excess COD case, the input of a significant amount of oxygen could be tolerated without having an effect on the effluent nitrate-N concentration because the oxygen simply acted to allow more removal of COD. Eventually, however, a point was reached at which nitrate-N removal deteriorated because the total input of electron acceptor (i.e., nitrate-N plus oxygen) exceeded the amount of electrons available from the donor. For the limiting COD case, the entrance of even a small amount of oxygen caused the nitrate-N concentration to increase

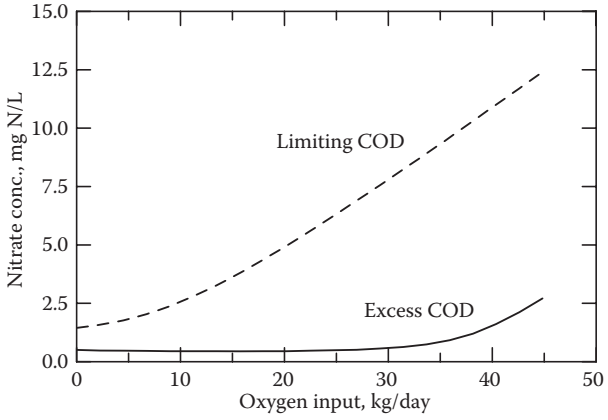

FIGURE 6.11 Effect of oxygen input rate on denitrification in a CSTR operated at an SRT of 240 hrs. Parameter values are given in Table 6.3 and the influent conditions are given in Table 6.6. For the limiting COD case the influent nitrate-N concentration was 60 mg/L whereas in the excess COD case it was 50 mg/L. The influent flow was 1000 m^3/day.

because any oxygen entering the bioreactor accepted electrons that otherwise would have gone for nitrate-N reduction. Although the magnitudes of the values given in Figure 6.11 are specific for the reactor conditions and parameter values used in the simulations, the results show clearly the importance of controlling the entrance of oxygen into a bioreactor in which denitrification is occurring.

The $\Delta S/\Delta N$ ratio is calculated from process stoichiometry, and simply tells us how much of one reactant will be removed in proportion to another. If the mass input rate of biodegradable COD into a denitrifying bioreactor is greatly in excess of that needed to remove the nitrate-N present, then the effluent biodegradable COD will be high (because there is insufficient acceptor to which to transfer electrons) and the nitrate-N concentration will be low and rate controlling. In other words, the term $S_S/(K_S + S_S)$ in the rate expression for process 2 in Table 6.1 will approach 1.0 and the term $S_{NO}/(K_{NO} + S_{NO})$ will be small. Conversely, if the input rate of biodegradable COD is less than that needed to remove the nitrate-N, then the effluent nitrate-N concentration will be high (because there is insufficient donor to provide the needed electrons) and the biodegradable COD concentration will be low and rate controlling. In other words, $S_{NO}/(K_{NO} + S_{NO})$ will approach 1.0 and $S_S/(K_S + S_S)$ will be small. The only way for both concentrations to be low and for both to simultaneously influence the rates of activity in the bioreactor is for the influent concentrations to closely match the required $\Delta S/\Delta N$ ratio. Given the influence of the entrance of small amounts of oxygen, as discussed above, and the variability associated with the values of the stoichiometric and kinetic coefficients in mixed microbial communities such as those used in wastewater treatment, this is difficult to achieve.

Figure 6.12[57] presents results from a study in which the relative amounts of biodegradable COD and nitrate-N (expressed as the carbon:nitrogen ratio) in the influent to a CSTR were varied. There it can be clearly seen that there was only a small range of influent ratios over which the effluent concentrations of both constituents were low. To overcome this problem, it is common practice to add the organic substrate in slight excess of the amount required to remove the nitrate-N, and then to pass the effluent from the anoxic bioreactor through a small aerobic bioreactor in which any residual electron donor can be removed with oxygen as the electron acceptor. Although not depicted in the model presented here, another advantage of adding organic substrate in excess of the amount required to remove the nitrate-N is that such practice minimizes the emission of nitrous oxide, a potent greenhouse gas.[25,31]

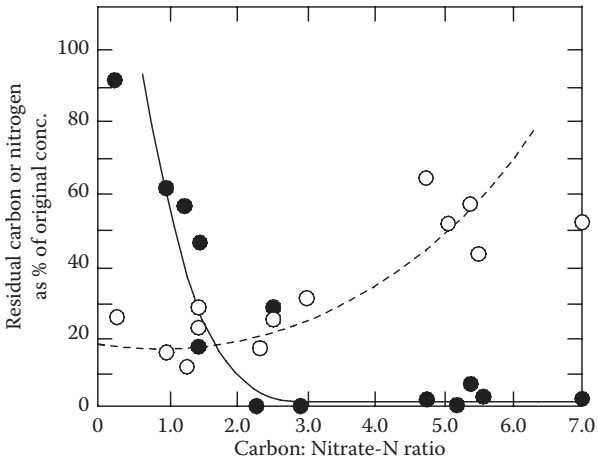

FIGURE 6.12 Effect of S_{SO}/S_{NOO} (expressed as C/N ratio) on the removal of carbon (○) and nitrogen (●) in a CSTR operated under anoxic conditions. (From Wuhrmann, K., Discussion of "Factors affecting biological denitrification of wastewater" by R. N. Dawson and K. L. Murphy. *Advances in Water Pollution Research, Jerusalem*, *1972*, 681–82, 1973. Reproduced with permission from Dr. K. L. Mechsner.)

6.5 MULTIPLE EVENTS

The reason for developing the model in Table 6.1 was to allow engineers to simulate biochemical reactors in which all of the listed processes are occurring. Thus, it would be instructive to use it to investigate such a situation in a single CSTR. By now, however, it is apparent that the conditions required for anoxic growth of heterotrophs and aerobic growth of autotrophs are mutually exclusive, since both are controlled by the DO concentration but in the opposite manner. Consequently, if a CSTR is receiving a constant input and is operating at steady state with a constant DO concentration, it is impossible for significant amounts of both nitrification and denitrification to occur. However, observations of treatment systems receiving diurnal variations in flow and concentration suggested that when the input was low, resulting in a high DO concentration, nitrification occurred, but when the input was high, driving the DO to very low concentrations, nitrification ceased and denitrification began, destroying part of the nitrate-N formed during the aerobic period. This made the daily average effluent total nitrogen concentration lower than in a system receiving adequate oxygen, suggesting that it might be possible to purposefully design a system in which the DO concentration varied sufficiently to allow both reactions to occur. In addition to reducing the amount of nitrogen discharged, this would reduce the amount of alkalinity destroyed and the amount of oxygen required, as discussed previously. Let us now consider each situation for a typical domestic wastewater.

6.5.1 Effects of Diurnal Variations in Loading

Figure 6.13 shows the simulated performance of a CSTR containing a biomass with the kinetic and stoichiometric coefficients in Table 6.3 and receiving an input with the variations shown in

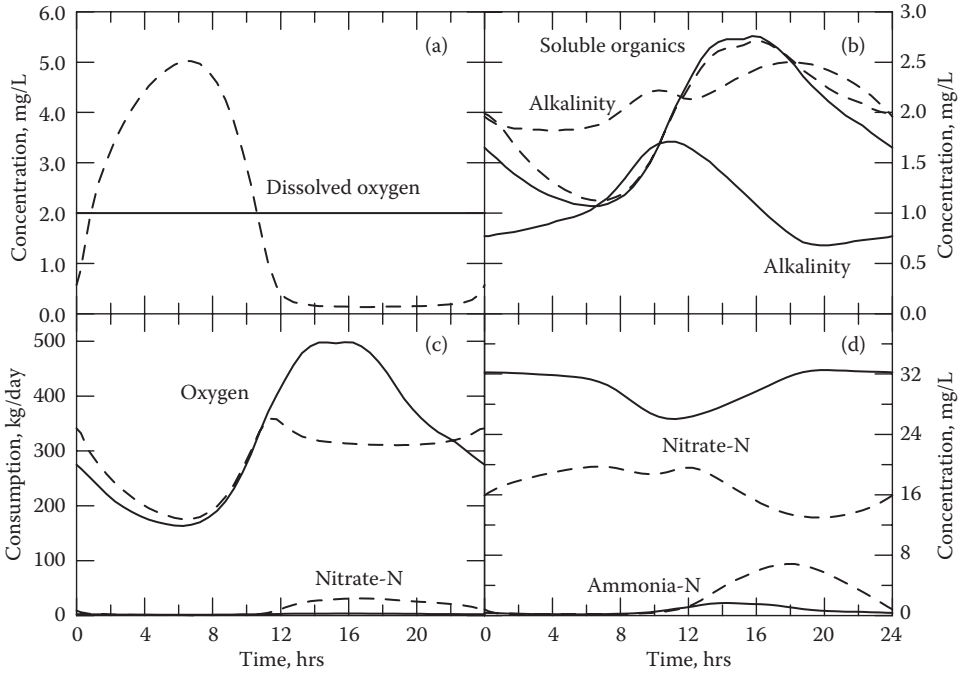

FIGURE 6.13 Response of a CSTR to a diurnal input. The input patterns are like those in Figure 6.2, but the flow-weighted average concentrations are those given in Table 6.6. The daily average flow rate was 1000 m^3/day. The solid curves are for a reactor in which the dissolved oxygen concentration was held constant at 2.0 mg/L. The dashed curves are for a reactor in which the mass transfer coefficient for oxygen was held constant at 3.83 hr^{-1}. The average HRT = 12 hrs and the SRT = 240 hrs. The parameter values are given in Table 6.3.

Figure 6.2. The average daily flow rate was 1000 m^3/day and the reactor volume was 500 m^3, giving an average HRT of 12 hours. The flow-weighted average concentrations of the various influent components were as shown in Table 6.6. The influent alkalinity, however, was assumed to have a constant concentration because it is influenced primarily by the characteristics of the carriage water. Two situations were considered. In one, the DO concentration was held constant at 2.0 mg/L throughout the entire 24-hour period. Given the value of $K_{O,H}$ in Table 6.3, this effectively eliminated denitrification. In the other, the mass transfer rate for oxygen was sufficient to maintain a DO concentration of 2.0 mg/L if the bioreactor received the daily average flow and concentrations. As shown in Figure 6.13a, however, because of the variable input, this resulted in excessive oxygen concentrations during periods of low loading but inadequate concentrations during high loading. Under high loading, significant denitrification occurred, as shown in Figure 6.13c. Comparison of the solid and dashed curves in parts b and d of Figure 6.13 shows that while the lack of oxygen had only a minor effect on the concentration of soluble organic matter in the effluent, it significantly lowered the nitrate-N concentration and raised the ammonia-N concentration. The raising of the ammonia-N concentration was due to retardation of nitrification, but the decrease in the nitrate-N concentration was caused by a combination of less production from nitrification and more consumption by denitrification. Nevertheless, it is apparent that less nitrogen was discharged from the system that had limited oxygen transfer capacity. Furthermore, less alkalinity was destroyed.

While it is not desirable to design a supposedly aerobic system with inadequate oxygen transfer capacity, the results from the simulation show clearly that all events can occur in a single biomass provided that the SRT of the system is sufficiently long to allow the nitrifying bacteria to grow during the aerobic period. This suggests that it should be possible to design a system in a way that maximizes nitrogen removal by controlling the periods with and without oxygenation. Such a system would have minimal power input as well as minimal alkalinity destruction.

6.5.2 Intermittent Aeration

Batchelor[5] was among the first to use simulation to investigate the possibility of using intermittent aeration to achieve both nitrification and denitrification in a single CSTR receiving a constant influent. His study was conducted with a model that was conceptually similar to the one in Table 6.1, although it differed somewhat with respect to both the kinetics and stoichiometry employed. In addition, the values of the kinetic and stoichiometric parameters were slightly different from those in Table 6.3. Nevertheless, the results of his simulations illustrate some important concepts regarding the major variables influencing the performance of such systems.

The situation considered by Batchelor[5] was of a wastewater containing 200 mg/L of biodegradable COD and 30 mg/L of ammonia-N, entering a CSTR with an SRT of 240 hours and an HRT of 4.3 hours. The bioreactor was operated with a cycle time (the time between initiations of aeration) of 0.5 hour while the aeration fraction (the fraction of time that the bioreactor was aerobic) was varied. In each case the effluent nitrogen concentrations from the bioreactor achieved a stable oscillation and the average concentrations of ammonia-N, nitrate-N, and total nitrogen were calculated. Figure 6.14 shows clearly that at low aeration fractions the average ammonia-N concentration rises, whereas at high aeration fractions the average nitrate-N concentration rises. Furthermore, an optimum exists at which the discharge of nitrogen is minimized. This result is consistent with our previous discussions. At aeration fractions in excess of the optimum, nitrification is complete, but the anoxic period is insufficient to allow much reduction of nitrate-N. Conversely, at aeration fractions less than the optimum, the aerobic period becomes insufficient for growth of the nitrifying bacteria. If the aeration fraction is made so low that the product of the aeration fraction and the SRT is below the minimum SRT for the nitrifiers, they will wash out and no nitrogen removal will be achieved, other than that associated with the incorporation

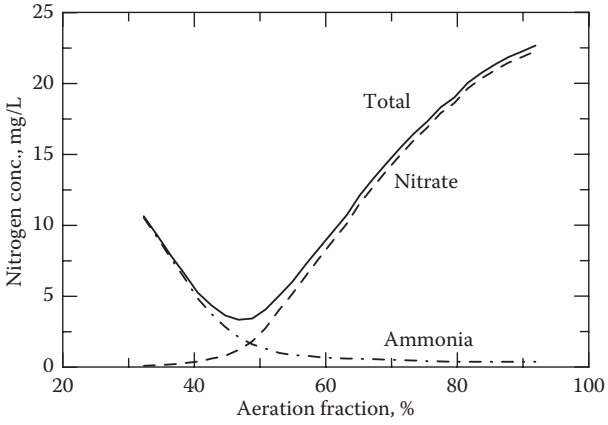

FIGURE 6.14 Effect of aeration fraction on the concentrations of ammonia-N, nitrate-N, and total-N in a CSTR operated with intermittent aeration at a cycle time of 0.5 hr. The HRT = 4.3 hrs and the SRT = 240 hrs. The influent biodegradable COD = 200 mg/L and the influent nitrogen concentration = 30 mg/L. (Adapted from Batchelor, B., Simulation of single-sludge nitrogen removal. *Journal of Environmental Engineering*, 109:1–16, 1983.)

of nitrogen into the heterotrophic biomass via synthesis. For the kinetic parameters used by Batchelor,[5] the minimum SRT for nitrification was 52.6 hours. Since the system SRT was 240 hours, this suggests that total process failure would occur at an aeration fraction of 0.22 (52.6 ÷ 240). The fraction of the SRT that is aerobic is called the aerobic SRT and it is now recognized as an important parameter in the design of systems in which both nitrification and denitrification are occurring.

The optimal aeration fraction in Figure 6.14 is 47%, but since the minimum allowable aeration fraction depends on the system SRT and the degree of denitrification also depends on the system SRT, we might expect the optimum aeration fraction to depend on the SRT as well. That this is the case is shown by the dark circles in Figure 6.15. The vertical bars in the figure indicate the range of aeration fractions at each SRT that results in total nitrogen concentrations within

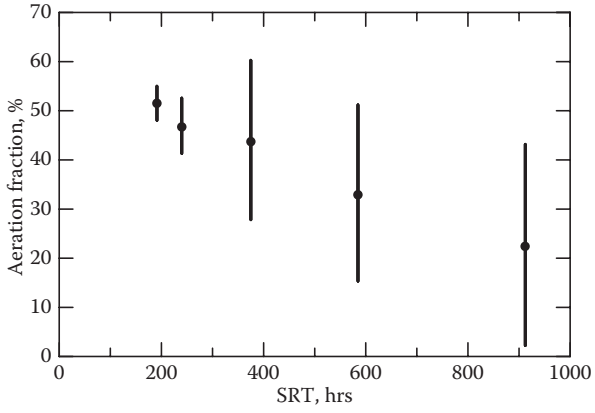

FIGURE 6.15 Effect of SRT on the optimal aeration fraction (dots) and the range of near optimal aeration fractions (lines) in a CSTR operated with intermittent aeration. The influent biodegradable COD = 200 mg/L and the influent nitrogen concentration = 30 mg/L. The dissolved oxygen concentration was fixed at 2.0 mg/L when the aeration was on and 0.0 mg/L when it was off. (Adapted from Batchelor, B., Simulation of single-sludge nitrogen removal. *Journal of Environmental Engineering*, 109:1–16, 1983.)

1.0 mg/L of the minimum. That range is seen to broaden as the SRT is increased, suggesting that longer SRTs provide more latitude in operation of the bioreactor to achieve optimal nitrogen removal.

Both Figure 6.14 and Figure 6.15 were prepared from simulations conducted with a cycle time of 0.5 hour. However, we might also expect the cycle time to influence the degree of nitrogen removal. During aerobic periods, the ammonia concentration will fall and the nitrate-N concentration will rise as nitrification occurs. Conversely, during anoxic periods, denitrification will cause the nitrate-N concentration to fall while the continued influx of ammonia in the absence of nitrification will cause the ammonia concentration to rise. The longer the cycle time, the greater the amplitude of the cycles because a longer time will be available for each reaction. Since the effluent concentration is characterized by the daily average concentration, longer cycle times result in higher average nitrogen concentrations. On the other hand, as the cycle time is shortened, a point will be reached at which it is difficult for the biomass to shift rapidly enough from aerobic to anoxic metabolism and vice versa. Consequently, there is also an optimum cycle time associated with each SRT and aeration fraction and it is significantly influenced by the nature of the wastewater.[19] Kinetic models that predict the shift in performance between denitrification and aerobic metabolism during intermittent aeration have been successfully developed for pure cultures,[34,37] and are now being used to optimize the cycle time needed to achieve maximum nitrogen removal with activated sludge cultures.[19] Experimental studies applied to domestic wastewaters have shown that cycle times on the order of 20 to 45 minutes work well.[30,36]

One event not included in the model used to generate Figures 6.14 and 6.15, nor in ASM Nos. 1, 2d, or 3, is the production of nitrous oxide when microbial cultures are exposed to alternating aerobic and anoxic conditions. However, because of the importance of nitrous oxide as a greenhouse gas, environmental engineers should be aware of this possibility. Both heterotrophic denitrifying and autotrophic ammonia-oxidizing bacteria can produce nitrous oxide. It is an intermediate in denitrification that can be released under anoxic conditions when the quantity of electron donor is insufficient to drive denitrification completely to nitrogen gas.[31] Ammonia oxidizing bacteria, on the other hand, can produce nitrous oxide under aerobic conditions as they recover their activity following a period of anoxia.[58] Although more is known about nitrous oxide production during denitrification and one modification of ASM No. 1 includes it,[26] the relative importance of the two means of nitrous oxide production is unknown. Nor is it certain how design and operational conditions influence each. Consequently, the reader should consult the current literature on this topic.

6.5.3 Closure

The important point to gain from the preceding is that engineers can exert control over the environment in biochemical reactors, thereby allowing processes to occur in a single system that would not otherwise occur together. This suggests that the engineer has considerable latitude in system design. The complexity of the interactions, however, makes it impossible to intuitively predict the outcome of all possible systems that the engineer might conceive. This, in turn suggests why it is necessary to work with models like those in Table 6.1. Through their application, engineers can explore large numbers of possible bioreactor systems to see how system layout and environment influence the outcome of the possible reactions.

It is apparent that the number of options available during the design of a single CSTR is very limited. Thus, most biochemical operations employ reactors with spatial gradients in them, either through the use of large length to width ratios or compartmentalized systems. Since both can be modeled as tanks in series systems, in the next chapter we will apply ASM Nos. 1 and 2d to study the performance of a number of such systems.

6.6 KEY POINTS

1. International Water Association Activated Sludge Model (ASM) No. 1 incorporates 8 processes acting on 13 components. The processes are aerobic and anoxic growth of heterotrophic biomass, aerobic growth of autotrophic bacteria, decay of both heterotrophic and autotrophic biomass, ammonification of soluble organic nitrogen, and hydrolysis of both particulate organic substrate and particulate organic nitrogen.
2. The 13 components incorporated into ASM No. 1 include six particulate and seven soluble ones. The particulate ones are inert organic matter, slowly biodegradable substrate, heterotrophic biomass, autotrophic biomass, biomass debris, and organic nitrogen. The soluble ones are inert organic matter, readily biodegradable substrate, dissolved oxygen (DO), nitrate-N, ammonia-N, organic nitrogen, and alkalinity.
3. In ASM No. 1, biomass growth is expressed by a double Monod equation in which both the electron donor and electron acceptor are considered, biomass loss is modeled by the lysis:regrowth approach, ammonification is considered to be first order with respect to the soluble organic nitrogen concentration, and hydrolysis is modeled by a surface mediated reaction term in which the rate is controlled by the slowly biodegradable substrate to biomass ratio and the electron acceptor concentration.
4. Activated sludge model No. 2d includes biological phosphorus removal in addition to carbon oxidation, nitrification, and denitrification. It assumes that phosphate accumulating organisms are a unique portion of the microbial community and that they can function under both aerobic and anoxic conditions. Because biological phosphorus removal is a complicated process, ASM No. 2d is considerably more complex than ASM No. 1.
5. Because hydrolysis is a slow reaction, long solids retention times (SRTs) are required to obtain substantial degradation of particulate substrate.
6. The major impact of the presence of particulate substrate in the feed to a continuous stirred tank reactor (CSTR) receiving a dynamic input is to dampen the variability in the oxygen requirement, thereby decreasing the peak requirement and delaying the occurrence of the maximum and minimum values.
7. Because the half-saturation coefficient for nitrifying biomass is very small relative to the concentration of ammonia-N in most wastewaters, nitrification behaves in an all-or-none manner in which nitrification is either almost complete or washout occurs. Furthermore, the washout SRT is very sensitive to the DO concentration and the temperature.
8. Because of the kinetics and stoichiometry of nitrification, the growth of autotrophic bacteria in a CSTR in which heterotrophs are growing can have a major impact on the amount of oxygen required while having little effect on the mixed liquor suspended solids (MLSS) concentration.
9. The release of ammonia-N as the result of biomass destruction in a CSTR receiving a feed containing only biomass provides a significant amount of substrate for nitrifying bacteria. Growth of those bacteria will have a substantial impact on the oxygen requirement in the bioreactor as well as on the amount of alkalinity destroyed.
10. Denitrification can have several benefits: it can remove nitrogen by converting nitrate-N to nitrogen gas, it can recover approximately 63% of the energy expended during nitrification by using the resulting nitrate-N as a terminal electron acceptor for organic substrate removal, and it can recover about half of the alkalinity destroyed during nitrification.
11. The $\Delta S/\Delta N$ ratio is an important characteristic of denitrifying bioreactors because it determines how much electron donor must be provided to convert a given amount of nitrate-N to nitrogen gas. However, because the ratio is influenced by factors such as the SRT and the presence of oxygen, it is not generally possible to operate a CSTR so that the effluent concentrations of both the electron donor and nitrate-N are low. Consequently, the

electron donor is usually provided in excess and any residual is removed in a small aerobic bioreactor with oxygen as the electron acceptor.

12. If a CSTR with an SRT sufficient for nitrification receives influent that follows typical diurnal patterns of flow and concentration, it is possible for denitrification to occur during periods of peak loading if the input rate of oxygen is insufficient to maintain significant concentrations of DO.
13. By subjecting a CSTR to intermittent aeration, it is possible to have both nitrification and denitrification occur in a single bioreactor receiving a constant input. Under those circumstances, the fraction of time that the system is aerobic is an important determinant of system performance, and there will be an optimum fraction that minimizes the effluent nitrogen concentration.

6.7 STUDY QUESTIONS

1. What are the 13 components and the 8 processes considered in IWA ASM No. 1? Construct a matrix that indicates which processes act on which components, using a plus sign to indicate when the concentration of a component is increased by the process and a negative sign to indicate when it is decreased.
2. What are the eight processes considered in ASM No. 1 and how is each modeled?
3. Simplify the matrix representing ASM No. 1 for the following situations:
   a. A totally aerobic bioreactor receiving only soluble constituents in the feed.
   b. A totally aerobic bioreactor receiving only particulate constituents in the feed.
   c. A totally anoxic bioreactor receiving particulate organic constituents and soluble nitrate and ammonia nitrogen.
   d. A totally aerobic bioreactor receiving ammonia-N as the only electron donor.
4. Draw a sketch comparing the steady-state performance of CSTRs receiving soluble and particulate substrates, and use it to contrast the usefulness of the process loading factor and SRT as design and operational parameters for bioreactors receiving particulate substrate.
5. Draw a sketch contrasting the response to typical diurnal loading patterns of the oxygen requirement in CSTRs receiving soluble and particulate substrates? Why do the two reactors behave differently?
6. What is meant by the statement that nitrification behaves in an all-or-none fashion? Why does it do so? What are the impacts of temperature and dissolved oxygen concentration on nitrification?
7. What are the effects of nitrification in a CSTR receiving an influent with the characteristics of typical domestic wastewater? Why do they occur?
8. Using a computer code for IWA ASM Nos. 1, 2, or 2d and typical temperature coefficients for the growth of heterotrophic and autotrophic biomass, investigate the effects of temperature (10° to 35°C) and SRT (48 to 360 hr) on the effluent concentrations of ammonia-N and readily biodegradable substrate from a CSTR receiving an influent with the characteristics given in Table 6.6. Use your results to discuss what is likely to happen to the performance of such a bioreactor throughout the year as the temperature changes and how SRT can be used to influence that performance.
9. What are the effects of nitrification in a CSTR receiving an influent containing heterotrophic biomass as its major constituent? What are the implications of those effects to the design and operation of such a reactor?
10. What is meant by the term $\Delta S/\Delta N$ ratio, how is it used in characterizing the performance of an anoxic CSTR, and why does it decline as the SRT of the bioreactor is increased?

11. Why is it difficult to operate an anoxic CSTR in a way that will ensure that the effluent concentrations of both the organic substrate and nitrate-N are low?
12. When a single CSTR receiving influent at a constant rate is subjected to intermittent aeration, the fraction of time that the system is aerobic will affect the effluent nitrogen concentration. Draw a sketch showing the effect of aeration fraction on the effluent nitrogen concentrations from such a system. Why does it look as it does?
13. For the situation described in Study Question 12, why does the optimal aeration fraction decrease as the SRT of the bioreactor is increased? Why does the range of near optimal aeration fractions increase?

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7 Multiple Microbial Activities in Complex Systems

Chapters 5 and 6 introduced us to the response of microbial cultures in single continuous stirred tank reactors (CSTRs) and demonstrated the importance of solids retention time (SRT) in determining that response. Although such reactors have found extensive use in research and in the treatment of some industrial wastewaters, more complex reactor systems are commonly used in practice and thus it is important to consider them. Consequently, in this chapter we will use the International Water Association (IWA) Activated Sludge Model (ASM) No. 1 to investigate the responses of complex bioreactor systems when they contain biomass reacting in accordance with the processes discussed in Chapter 6 and listed in Table 6.1. Since the events occurring in these various systems are all the same, the responses will demonstrate how bioreactor configuration affects system performance, thereby providing a foundation from which engineering decisions can be made about the appropriate system configuration for a given objective. In addition, we will introduce the concepts of biological phosphorus removal (BPR) through the use of ASM No. 2d.

7.1 MODELING COMPLEX SYSTEMS

Table 1.2 showed that several activated sludge types use CSTRs in series or plug-flow reactors with dispersion. Actually, for purposes of modeling system response, the latter can be represented adequately as CSTRs in series,[22] and thus only that configuration need be considered. Thus, this chapter will focus on the theoretical performance of a number of systems containing CSTRs linked in various ways.

7.1.1 Representing Complex Systems

Conventional activated sludge (CAS) systems use long rectangular bioreactors with influent and biomass recycle introduced together at one end. Even though such bioreactors are often called plug-flow reactors, the mixing patterns in them are equivalent to three to five CSTRs in series.[25] High purity oxygen activated sludge (HPOAS), on the other hand, actually uses a series of bioreactors in order to achieve efficient utilization of the oxygen supplied, with four or five in series being common configurations. Because five CSTRs in series can be considered representative of both of these important activated sludge variations, that configuration will be investigated first.

Step feed activated sludge (SFAS) is an important variation on the CAS process in which influent is introduced at several points along a rectangular aeration basin while biomass recycle is introduced only at the beginning.[13] It was originally developed as a means for distributing the oxygen demand more evenly throughout the basin, but is now used primarily for its effect on the concentration of mixed liquor suspended solids (MLSS) entering the final settler.[18,37] Since the reactor is the same as in CAS, it is also possible to represent this configuration as five CSTRs in series with all biomass recycle to the first bioreactor, but with the influent distributed evenly among the bioreactors. This configuration will be studied in Section 7.3 to see how the simple act of changing the feed distribution influences system performance.

Table 1.2 indicates that contact stabilization activated sludge[39] (CSAS) also uses CSTRs in series. While it is true that the system uses two CSTRs, they are only in series with respect to the biomass recycle flow, not the main wastewater flow. Rather, the main wastewater flow passes through only one bioreactor (the contact tank) while the biomass recycle flow passes through another bioreactor (the stabilization tank) before entering the contact tank. Thus, this system can be represented as two CSTRs in series with all biomass recycle entering the first bioreactor and all influent entering the second. This unique configuration responds in a way that is quite distinct from either of the first two configurations, and thus it will be considered in Section 7.4.

One observation from Chapter 6 was that aerobic and anoxic growth of heterotrophic biomass cannot both occur to significant degrees in a single bioreactor maintained under constant environmental conditions because the two processes have mutually exclusive requirements. However, we saw in Section 6.5.2 that removal of part of the nitrogen in a wastewater can be achieved by imposing time variant changes in the dissolved oxygen (DO) concentration in a bioreactor. While such a scheme can be used in small systems, it is not practically feasible for large ones because of problems associated with manipulation of large air flows. An alternative approach that achieves moderate total nitrogen removal and has been widely employed is to use two bioreactors in series, but with one maintained under anoxic conditions and the other under aerobic conditions so that the wastewater and associated biomass encounter both environments during their passage through the system. One such system is the modified Ludzack-Ettinger[24] (MLE) process, in which the anoxic bioreactor is first. It is often used to treat domestic wastewaters, which are devoid of nitrate but rich in organic and ammonia nitrogen. Because nitrate is only formed under aerobic conditions, a recirculation stream returns mixed liquor from the second bioreactor to the first, thereby providing nitrate, which the heterotrophic biomass can use as terminal electron acceptor during growth on the organic matter in the wastewater. Systems of this type, which are called "single-sludge systems" because they allow carbon oxidation, nitrification, and denitrification to occur in a single biomass that is passed through different biochemical environments, have become quite important in domestic wastewater treatment for biological nutrient removal (BNR). Thus, it is important that we investigate their performance through simulation.

One characteristic of the MLE process is that the effluent contains appreciable quantities of nitrate-N since nitrification occurs in the last bioreactor in the chain. One way to overcome this is to add a third bioreactor that is maintained under anoxic conditions, thereby allowing additional denitrification to occur. Unfortunately, discharge of an actively denitrifying biomass to a final settler results in biomass settling problems because the evolving nitrogen gas attaches to the biomass flocs and makes them rise, rather than settle. One way to eliminate this is to aerate the biomass for a short period prior to settling, thereby stopping denitrification. The four-stage Bardenpho[3] process is one of several BNR systems that use these concepts. It can be represented as four tanks in series with all influent and biomass recycle entering the first and with mixed liquor recirculation (MLR) from the second to the first. As suggested above, the first and third bioreactors are anoxic whereas the second and fourth are aerobic. It will be studied in Section 7.6.

As shown in Table 1.2, BPR processes also employ CSTRs in series to create the unique environment required for selection of phosphate accumulating organisms (PAOs). The simplest of those processes uses two bioreactors in series, with all influent wastewater and all biomass recycle entering the first bioreactor.[31] As will be recalled from Section 2.4.6, PAOs only develop their unique ability to store large quantities of phosphate when they are subjected to alternating anaerobic and aerobic/anoxic conditions, although PAO growth is slower under anoxic conditions compared to aerobic conditions. Because of this and because of the necessity to capitalize on the readily biodegradable substrate in the influent wastewater, the first bioreactor in a two-stage BPR system is anaerobic and the second is typically aerobic. Such systems are becoming increasingly important in domestic wastewater treatment. Consequently, it is important that we also investigate their performance through simulation.

The ability to represent several important biochemical operations as systems made of tanks in series simplifies the task of modeling their performance. All that is required is that a mass balance equation like Equation 4.2 be written for each component in each bioreactor, including the settler, giving a set of equations that must be solved simultaneously. If the influent to the system has time invariant characteristics, the derivative terms in the equations may be set to zero, reducing the differential equations to algebraic ones. Otherwise, they remain as differential equations. In either case, the presence of biomass recycle and MLR requires numerical techniques to be used for their solution. As mentioned in Section 6.1.6, several computer codes are available for solving the equations that result from the application of ASM No. 1 to a single CSTR with biomass recycle. They are also applicable to CSTRs in series. For example, the code SSSP (Simulation of Single-Sludge Processes),[5] which was used in Chapter 6, allows simulation of systems containing up to nine CSTRs with any distribution of feed or biomass recycle among them and with MLR from any bioreactor to any other bioreactor upstream of it. Because of its ability to simulate all of the systems discussed above that do not involve BPR, it was used to investigate their performance. Simulation of BPR processes is more complex than simulation of the other processes because more reactions and components are involved. As discussed in Section 6.1.4, ASM No. 1 was expanded to include biological phosphorus removal by PAOs capable of growth under both aerobic and anoxic conditions, resulting in ASM No. 2d. Both ASIM and GPS-X (Table 6.4) were used to perform BPR process simulations in this chapter.

An examination of Table 1.2 reveals that biological nutrient removal can also be accomplished in sequencing batch reactors (SBRs) and they are finding increasing use for that purpose.[16,31] The utility of SBRs stems from their characteristic of varying temporally in a manner that is analogous to the spatial variations in a perfect plug-flow reactor. As a consequence, by changing the aeration pattern over time, carbon oxidation, nitrification, denitrification, and phosphorus removal can be accomplished in the same way as in the tanks-in-series systems discussed above. Therefore, it is important that we consider the performance of SBRs and their similarity to the other systems. Oles and Wilderer[28] have developed a computer code, SBRSIM, for simulation of SBR performance. It is based on ASM No. 1. In Section 7.8 we will review briefly the operating strategies for SBRs and examine the performance of a system that is analogous to the MLE process.

7.1.2 Significance of Solids Retention Time

The solids retention time, SRT, was used as the fundamental variable in Chapters 5 and 6 because it is functionally related to the steady-state specific growth rate of the biomass in a CSTR. No such relationship holds for a chain of CSTRs, however. This is because the SRT is defined with respect to the entire system, while the biomass in each bioreactor has a unique specific growth rate in response to the concentration of the particular nutrient limiting growth in that bioreactor. Although one might question the significance of the SRT in such systems, it is still a variable of fundamental importance because it represents a net average specific growth rate for the system. This follows from the fact that SRT is defined as the mass of biomass contained in the system divided by the mass wasted per unit time (Equation 5.1), whereas specific growth rate is defined as the mass of biomass formed per unit time divided by the biomass present (Equation 3.35). Since at steady state, the mass of biomass wasted must equal the net mass formed (i.e., the growth minus the loss due to lysis, death, and decay), it follows that the SRT is inversely proportional to the net average specific growth rate of the entire system. Consequently, it determines important system characteristics, such as the electron acceptor requirement and the excess biomass production rate. It also influences strongly the concentrations of various components in the effluent, although they will also be influenced by the system configuration, as we will see.

If SRT is to continue to be a variable of fundamental importance, it is necessary to have a means of controlling it in practice and of setting it during simulations. This is best done through modification of the Garrett[12] configuration discussed in Section 5.4.1. As before, biomass should be wasted directly from the bioreactors, rather than from the biomass recycle flow, but in this case it should be

from each bioreactor in proportion to its volume. The efficacy of this procedure can be seen in the following. Consider a system containing N bioreactors of volume V_i, containing MLSS at concentration $X_{M,T,i}$, with wastage from each at flow rate $F_{w,i}$. In that case, the SRT is defined as

$$\Theta_c = \frac{\sum V_i \cdot X_{M,T,i}}{\sum F_{w,i} \cdot X_{M,T,i}}. \tag{7.1}$$

If wastage is from each bioreactor in proportion to its volume, then

$$F_{w,i} = F_w\left(\frac{V_i}{V_T}\right), \tag{7.2}$$

where F_w is the total wastage flow rate and V_T is the total system volume. Substitution of Equation 7.2 for $F_{w,i}$ in Equation 7.1 and rearrangement gives:

$$\Theta_c = \frac{V_T \sum V_i \cdot X_{M,T,i}}{F_w \sum V_i \cdot X_{M,T,i}} \tag{7.3}$$

or

$$\Theta_c = \frac{V_T}{F_w}, \tag{7.4}$$

which is the same as Equation 5.2. Thus, if wastage is from each bioreactor in proportion to its volume, the SRT will be controlled solely by the total bioreactor volume and the total wastage flow rate, regardless of the MLSS concentration in each bioreactor. Because of its simplicity, this technique will be used in all simulations in this chapter, even though the wastage streams will not be shown in the schematic diagrams to simplify them.

7.1.3 Importance of the Process Loading Factor

In Section 5.1.7, we argued that for a single CSTR at steady state the SRT is a better control parameter than the process loading factor or F/M ratio (U). This does not mean that the process loading factor is unimportant, however. Rather, for a tanks-in-series system it is quite significant because the value in each bioreactor is proportional to the specific growth rate of the biomass in that bioreactor. This can be seen by reviewing Equation 5.49 in Section 5.1.7. For the general situation under consideration here, it can be rewritten as

$$\mu_{H,i} = \frac{Y_{H,T}F(S_{SOi} - S_{Si})}{V_i \cdot f_A \cdot X_{M,T,i}}, \tag{7.5}$$

where $\mu_{H,i}$ is the specific growth rate of the heterotrophic biomass in bioreactor i, S_{SOi} is the substrate concentration entering that bioreactor, and S_{Si} is the concentration in (and leaving) it. If for any given bioreactor, the assumption can be made that $S_{Si} \ll S_{SOi}$, then

$$\mu_{H,i} = \frac{Y_{H,T} \cdot U_i}{f_A}, \tag{7.6}$$

where U_i is the process loading factor in bioreactor i. The active fraction, f_A, changes little from tank to tank and can be considered to be constant for a given system, as can $Y_{H,T}$. Consequently, the value of the process loading factor in each bioreactor tells us how the state of the microbial community is changing as it moves from bioreactor to bioreactor in the chain. This is important information that cannot be gained from the overall system SRT. In Chapter 11 we will see how control of the process loading factor in individual bioreactors can be used to influence the competition between members of the microbial community to produce activated sludges with desired characteristics.

In SBRs, influent is added for only a portion of the cycle, called the fill period. Equation 7.5 can be used to define an "instantaneous" process loading factor for an SBR by letting F be the volumetric flow rate during the fill period and by ignoring S_{Si}, which is time dependent. Short fill periods require a large F for a given delivered volume, thereby giving a large instantaneous process loading factor. Conversely, long fill periods require a small F, giving low instantaneous process loading factors. While only an approximation, the calculated instantaneous process loading factor in an SBR can be used in a manner analogous to the individual reactor process loading factors in a tanks-in-series system.

7.2 CONVENTIONAL AND HIGH PURITY OXYGEN ACTIVATED SLUDGE

7.2.1 Description

Figure 7.1 presents a schematic diagram of the system used to simulate conventional and high purity oxygen activated sludge systems. All influent and all biomass recycle enters the first bioreactor and passes from bioreactor to bioreactor down the chain. For the purposes of this chapter, the bioreactors were considered to be of equal volume, but different residence time distributions can be attained by using bioreactors of different size.[22] The influent flow rate used in the simulations was 1000 m^3/day and the volume of each bioreactor was 50 m^3, giving a total system volume of 250 m^3 and a system hydraulic residence time (HRT) of 6 hours, which is a value commonly used in practice. The biomass recycle flow rate was fixed at 500 m^3/day, unless otherwise specified, giving a recycle ratio, α (F_r/F), of 0.5, which is also commonly used in practice. All bioreactors are aerobic and the DO concentration was controlled at 2.0 mg/L in each, thereby eliminating denitrification. This concentration was chosen because it is easily and economically achieved in practice and allows unhindered nitrification, as shown in Figure 6.7.

To allow direct comparison to the results in Chapter 6, the characteristics of the influent were assumed to be those listed in Table 6.6 and the values of the kinetic and stoichiometric coefficients of the biomass growing on that influent were assumed to be those in Table 6.3.

7.2.2 Effect of SRT on Steady-State Performance

The effect of SRT on the steady-state performance of the tanks-in-series reactor system is given by the solid curves in Figure 7.2, which show the concentrations in the last bioreactor of the chain, and

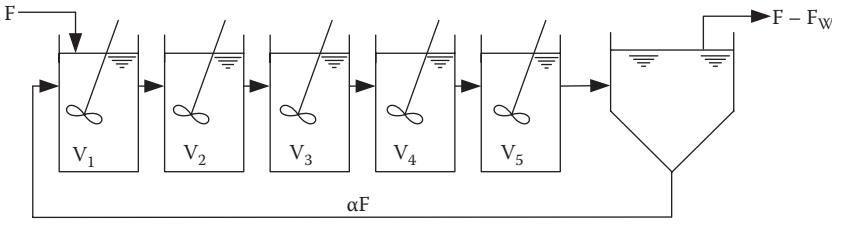

FIGURE 7.1 Schematic diagram of five CSTRs in series with all influent and all biomass recycle to the first reactor. Although not shown, solids wastage is directly from all reactors. This configuration simulates CAS and HPOAS systems.

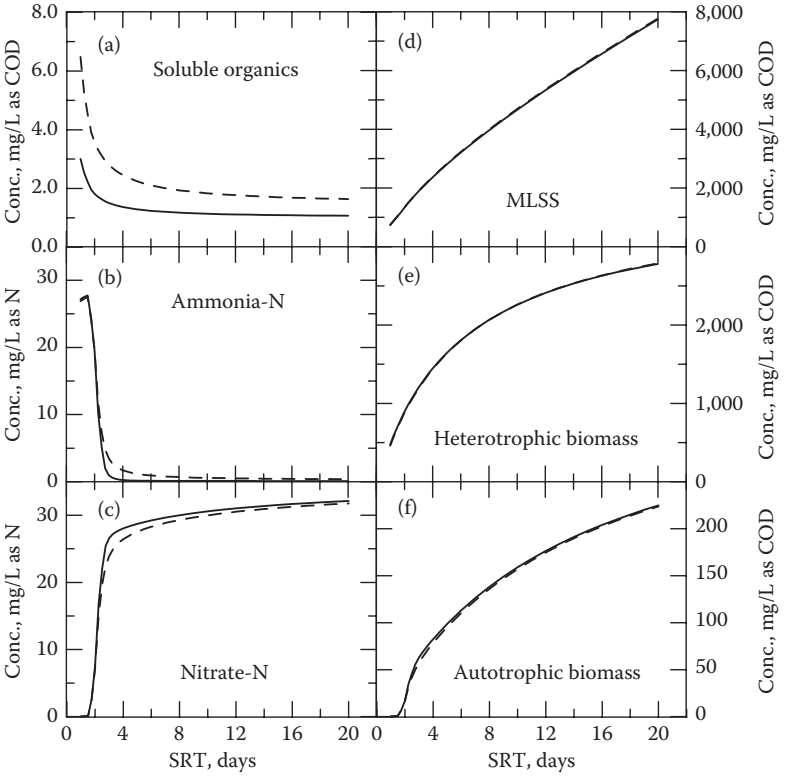

FIGURE 7.2 Effect of SRT on the steady-state concentrations of various constituents in the last reactor of the CAS or HPOAS system depicted in Figure 7.1. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Influent flow = 1000 m^3/day, influent concentrations are given in Table 6.6, biomass recycle flow = 500 m^3/day, and volume of each reactor = 50 m^3. Parameters are listed in Table 6.3. The DO concentration was held constant at 2.0 mg/L. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

thus represent the concentrations entering the final settler. For comparison, the dashed curves show the performance of a single CSTR with a volume equal to the total system volume of the chain (i.e., the systems have the same HRT). Because the influent to the systems contains particulate organic matter, simulations were only performed at SRTs in excess of one day.

The most striking thing about the curves in Figure 7.2 is that although the biomass and MLSS concentrations in the two systems are essentially the same, the tanks-in-series system achieves slightly better effluent quality, with lower concentrations of soluble organic substrate (expressed as chemical oxygen demand, COD) and ammonia-N and slightly higher concentrations of nitrate-N. This is because microbial growth and substrate utilization behave in a first-order manner when the substrate concentration is low, and first-order reactions achieve a higher extent of reaction in chains of reactors than in single CSTRs.[22] However, it should be noted that in contrast to previous simulation studies,[8] the effect of bioreactor configuration on soluble organic substrate concentration is not large. This is because ASM No. 1 includes the generation of soluble organic substrate due to biomass death and lysis, which was not considered in previous studies. This small effect of bioreactor configuration on organic substrate removal is consistent with experimental observation.[38] It will also be noted that the effect of bioreactor configuration on nitrification is larger than the effect on soluble substrate removal. This follows from the fact that most nitrogen released as a result of heterotrophic biomass death and lysis will be used for heterotrophic biomass regrowth, with only a small fraction being available as the substrate for nitrifier growth. It is doubtful that such a large

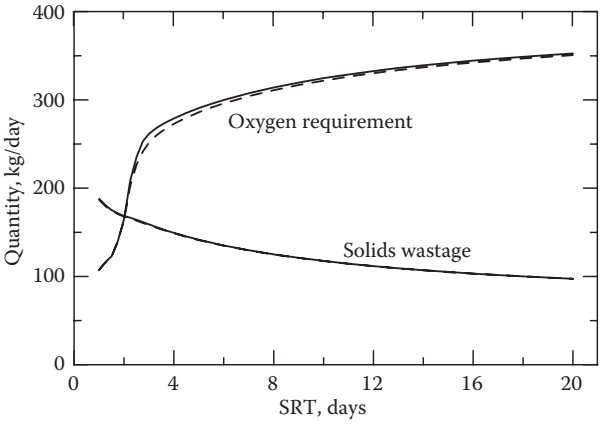

FIGURE 7.3 Effect of SRT on the total steady-state oxygen requirement and solids wastage rate (in COD units) for the CAS or HPOAS system depicted in Figure 7.1 operating under the conditions listed in Figure 7.2. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. To express the solids wastage rate in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

effect would be seen in practice, however, because a single CSTR will offer more protection to the nitrifiers against high concentrations of inhibitory organic matter.

Since microbial death and lysis is a first-order reaction, one might expect the tanks-in-series system to exhibit more biomass loss than the single CSTR, thereby resulting in lower concentrations of heterotrophic and autotrophic biomass. The fact that this does not occur may be surprising at first, but in reality, there is a logical reason. The residence time of the biomass in the system is equal to the SRT and is independent of the residence time of the fluid within the system (i.e., the HRT). Because the biomass recycle flow rate is of the same magnitude as the influent flow rate, the biomass is recycled around the system many times within one SRT, causing it to approach a completely mixed condition even though the fluid passes through a cascade. As long as the SRT is large with respect to the system HRT, the tendency of the biomass recycle to make the biomass well mixed will overpower the cascade effect of the fluid in the chain in determining the extent of reaction of the biomass by death and lysis. Thus, the biomass concentration in the chain will be very similar to the concentration in a CSTR with equal SRT. The tanks-in-series system will only experience more biomass loss when the HRT approaches the SRT so that the biomass experiences a cascade effect during its residence time.

Figure 7.3 shows the effect of SRT on the total system oxygen requirement and the solids wastage rate. Again, for comparison, the response of a single CSTR is shown as dashed curves. The lack of difference between the two systems at longer SRTs is consistent with the arguments in the preceding paragraph concerning the homogeneous nature of the MLSS caused by biomass recycle. The slight difference between the oxygen requirements in the two types of bioreactors at lower SRTs is due primarily to the greater degree of nitrification in the tanks-in-series system. The similarity in total oxygen requirement and solids wastage rate means that these characteristics can be estimated with the simple equations for a single CSTR in Chapter 5, even when a tanks-in-series reactor is to be used. Furthermore, these similarities tell us that the mass of MLSS in the two systems will be similar at a fixed SRT. Consequently, the simple equations for a CSTR can also be used to estimate the $X_{M,T}\cdot\tau$ product. This ability to use the simple equations is very helpful during process design, as will be seen in Chapters 10 and 11.

7.2.3 Dynamic Performance

Figures 7.2 and 7.3 present only the steady-state responses of the chain of CSTRs, but, as we saw in Chapter 6, biochemical operations are often subjected to diurnal changes in loading. Thus, it

is important to consider how bioreactor configuration influences dynamic response. Because 10 days is a commonly used value for the SRT in activated sludge systems, the dynamic response of the system in Figure 7.1 was studied at that SRT by imposing the diurnal loading pattern shown in Figure 6.2. The average flow rate to the system was 1000 m^3/day, the same as in the steady-state simulations, and the flow-weighted average concentrations of the various constituents were the same as those in Table 6.6. Because the alkalinity of a wastewater is determined primarily by the characteristics of the carriage water, rather than by the waste constituents, the alkalinity was assumed to be constant at the value in Table 6.6.

The output response of the tanks-in-series system to repeated application of the loading pattern in Figure 6.2 is shown in Figure 7.4 by the solid curves. For comparison, the response of a single CSTR with the same SRT and average HRT is shown by the dashed curves. Only the soluble constituents are shown because the particulate constituents are relatively constant due to the length of the SRT relative to the HRT. Examination of the curves shows that the variability in the effluent concentrations of the two reactants, organic substrate and ammonia-N, was much less for the chain of CSTRs than for the single CSTR. This is a direct result of the hydraulic differences between the two systems. Any change in the influent to a single CSTR is seen instantly in the effluent because the bioreactor and the effluent have the same concentration. The chain experiences time delays, however, as the fluid moves from tank to tank. This gives a greater opportunity for degradation or

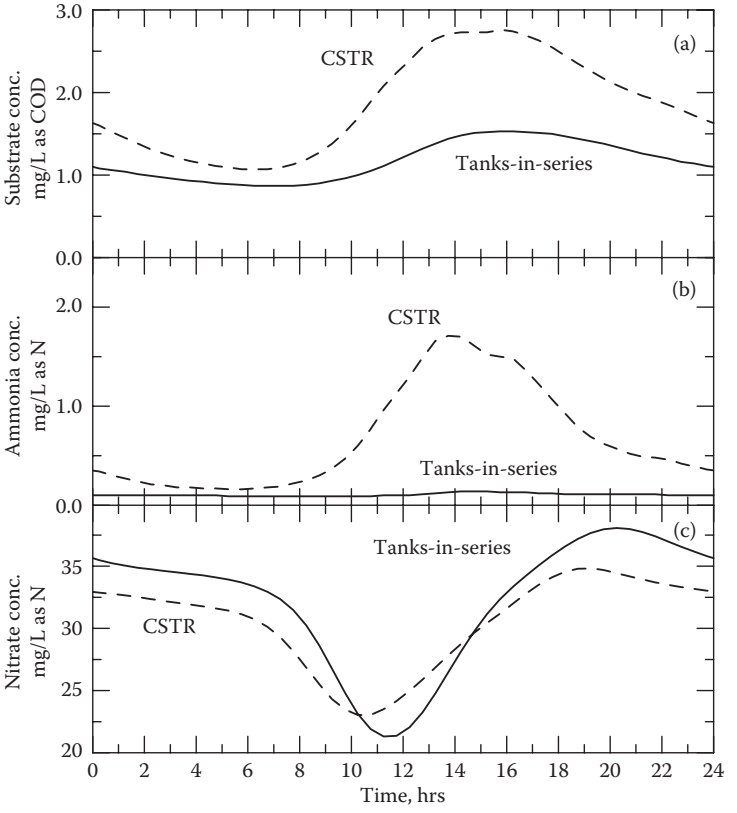

FIGURE 7.4 The time dependent response from the CAS or HPOAS system depicted in Figure 7.1 when subjected to the diurnal loading patterns shown in Figure 6.2. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Average influent flow = 1000 m^3/day, average influent concentrations are given in Table 6.6, biomass recycle flow = 500 m^3/day, volume of each reactor = 50 m^3, and SRT = 10 days. Parameters are listed in Table 6.3. The DO concentration was held constant at 2.0 mg/L.

transformation of the substrates. The effect is particularly significant for ammonia because of the small maximum specific growth rate of nitrifying bacteria.

An important assumption in all of the simulations was that the DO concentration was constant at 2.0 mg/L, a situation that can be achieved in practice through DO control. However, we saw in Figure 6.13 that imposition of the typical diurnal loading pattern causes considerable variation in the oxygen requirement within a single CSTR and that failure to meet that requirement causes deterioration of system performance. Thus it is important to consider how the diurnal loading influences the oxygen requirement in each bioreactor of the chain so that appropriate provisions can be made for delivering the needed amount. This is shown in Figure 7.5. In that figure, the vertical lines represent the range in the oxygen consumption rate experienced in each bioreactor as the loading varies over a 24-hour period. As expected, because the first bioreactor receives the variable waste load directly, it experiences the greatest variability, exhibiting a fourfold range in oxygen consumption rate. The second bioreactor receives only the substrate that is not removed by the first, and thus the magnitude of the oxygen requirement and the variability associated with it are lower, and so on down the chain. This means that any system that uses a tanks-in-series type reactor must be designed to handle much different oxygen requirements at different points in the system. The dots on the bars represent the steady-state oxygen requirement in each bioreactor caused by the entrance of a constant flow and concentration of wastewater to the system. They clearly show two things: first, a system designed only on the basis of the steady-state requirement would be inadequate during periods of peak loading; second, equalization of influent flow and concentration allows the use of a smaller oxygen transfer system. The figure also shows that the peak to average ratio changes from bioreactor to bioreactor, becoming particularly large in bioreactor 3. This is primarily due to nitrification, which occurs further downstream than organic substrate removal during periods of high loading. This is because nitrifying bacteria have a much smaller maximum specific growth rate than heterotrophic bacteria. As a consequence, the nitrifiers cannot increase their metabolism as much in response to the increased input rate of substrate, allowing a larger fraction of the ammonia-N to pass through to the downstream reactors during periods of high loading. For comparison purposes, the dashed lines in Figure 7.5 show the high, low, and steady-state oxygen requirements in a single CSTR. They show clearly that the first tanks in the chain require much higher oxygen input rates and experience a greater variation in requirement due to the lower degree of dampening associated with the small bioreactor. One reason for the use of single CSTRs is to take advantage of the smaller variability in oxygen requirement.

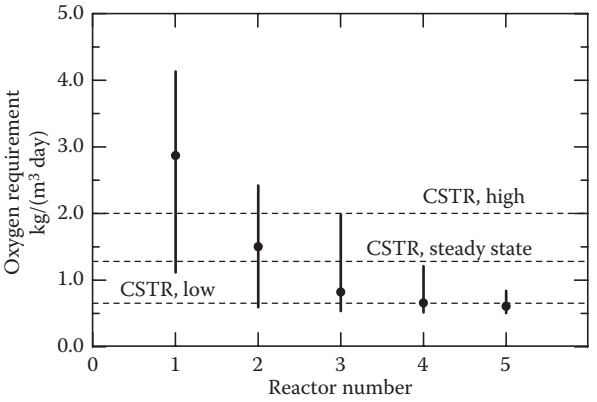

FIGURE 7.5 The variability of the oxygen requirement in each reactor for the CAS or HPOAS system described in Figure 7.4. The vertical lines represent the range observed over a 24-hr period and the dots represent the requirement when the influent flow and concentrations are constant at the flow-weighted average values (steady state). For comparison, the dashed lines show the extreme diurnal requirements, as well as the steady-state requirement, for a single CSTR with a volume of 250 m^3.

7.2.4 Variations within the System

The dots in Figure 7.5 demonstrated that each tank in a chain of CSTRs has a different reaction rate associated with it even when the system is operating at steady state. This suggests that further investigation of that aspect of system performance would be worthwhile. Thus, changes in the steady-state concentrations of three soluble constituents and the MLSS are shown in Figure 7.6 for three different SRTs.

The first thing to note in Figure 7.6 is that the MLSS concentration has little tank-to-tank variability. The same is true for the heterotrophic and autotrophic biomass concentrations, although they are not shown. This lack of change is because the SRT is large relative to the HRT, causing the biomass recycle to make the mixed liquor homogeneous, as discussed previously. A decline in the MLSS concentration as the solids move down the chain would only occur when the HRT approaches the SRT.

In contrast to the behavior of the particulate material, there are significant decreases in the concentrations of soluble organic substrate and ammonia-N as the fluid moves down the chain. All substrate is added to the first tank where it is mixed with biomass that has been returned from the settler. The presence there of high concentrations of soluble organic matter and ammonia-N allows both the heterotrophic and autotrophic biomass to attain relatively high specific growth and substrate removal rates, which, when combined with the high biomass concentration, gives rapid removal rates. This causes considerable reduction in soluble organic substrate and ammonia-N concentrations in the first bioreactor relative to the feed. Since the mass flow rates of soluble organic substrate and ammonia-N into the second bioreactor are much less than the rates into the first, the specific growth and substrate removal rates are reduced, but the high biomass concentration allows maintenance of an overall rate that is sufficiently large to cause more substrate to be removed. In other words, the process loading factor is very high in the first bioreactor but lower in the later ones. This means that the activity of the biomass varies as it moves from tank to tank.

In spite of the similarity in the mechanism associated with the removal of soluble organic matter and ammonia-N, comparison of parts a and b of Figure 7.6 shows that there are significant differences in their concentration profiles, particularly at low SRT. These differences result in part from the low maximum specific growth rate associated with autotrophic biomass. Because of it, the percentage of the influent ammonia-N removed in the first tank is smaller than the percentage of the soluble organic

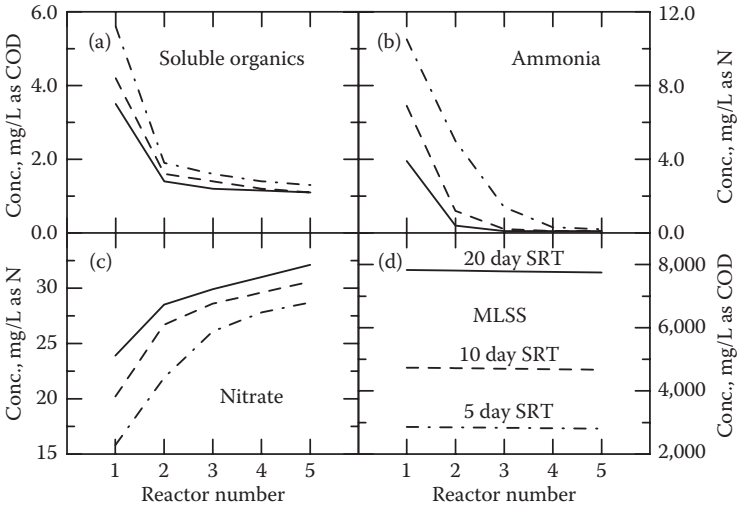

FIGURE 7.6 Effect of SRT on the steady-state concentrations of various constituents in each reactor of the CAS or HPOAS system described in Figure 7.2. To express the MLSS concentration in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

matter removed. This causes a greater percentage of the ammonia-N to move to the second tank, and so on, thereby causing nitrification to occur further down the bioreactor chain than organic substrate removal. In addition, because the half-saturation coefficient for autotrophs is lower than the coefficient for heterotrophs, the autotrophs tend to grow near their maximal rate in the first few bioreactors, particularly at lower SRTs where their mass in the system is small. Consequently, tank-to-tank removal of ammonia-N is more nearly linear than is the removal of organic substrate. In the final tanks the ammonia-N concentration approaches zero, making the specific removal rate quite low, but the presence of a high autotrophic biomass concentration keeps the overall rate high enough to scavenge any remaining ammonia from the system. The heterotrophic biomass respond in a similar manner, but the organic substrate concentration does not approach zero because substrate is continually being resupplied by biomass death and lysis. Thus, equilibrium is achieved in which the use of organic substrate just balances its release, causing a pseudosteady-state concentration to be maintained.

The tank-to-tank variability in the activity of the biomass is reflected best by the oxygen requirement, which is shown in Figure 7.7. There it can be seen that as in Figure 7.5, more oxygen is used in the first tanks because more substrate is removed in them. Furthermore, oxygen consumption in the later tanks reaches a relatively constant value reflective of endogenous metabolism and decay of the biomass.

The effect of SRT on the tank-to-tank performance is shown in Figures 7.6 and 7.7 by the three different curves. Because the system with the greatest SRT contains more biomass, organic substrate removal and nitrification occur more rapidly in it, causing lower concentrations to be attained in early tanks than in systems with lower SRT. In addition, at a lower SRT much less removal of ammonia-N occurs in the early tanks, for reasons discussed above. Comparison of parts b and c of Figure 7.6 shows that even though about the same ammonia-N concentration is ultimately achieved at all three SRTs, more nitrate-N is produced at higher SRTs. This is due to the greater death and lysis that occurs at longer SRTs, releasing more nitrogen, which acts as substrate for the nitrifying bacteria, producing more nitrate-N. Solids retention time has a fairly complicated effect on the pattern of oxygen consumption down the chain because it is influenced by organic substrate removal, nitrification, and biomass decay, all of which depend on the SRT. The longer the SRT, the greater the organic substrate removal in the first tank, and thus, the greater the oxygen requirement associated with that activity. Likewise, the longer the SRT, the greater the importance of biomass decay in the later tanks, also causing greater oxygen consumption. In intermediate tanks, however, differing degrees of nitrification occur at different SRTs, depending on the amount that has occurred in the early tanks. This makes it difficult to generalize about the effect of SRT on the amount of oxygen required in such tanks. Each situation will depend on the character of the wastewater. Consequently, any system employing this configuration must be built to ensure operational flexibility.

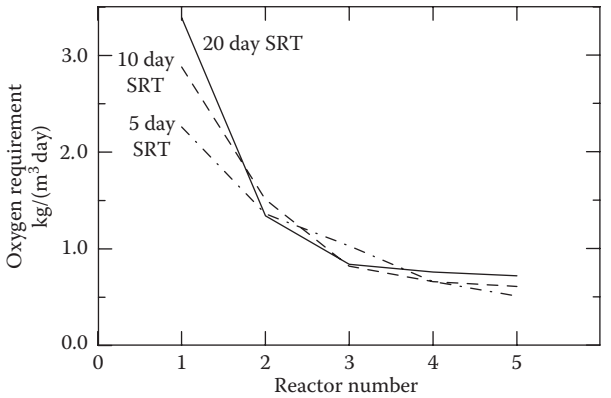

FIGURE 7.7 Effect of SRT on the steady-state oxygen requirement in each reactor of the CAS or HPOAS system described in Figure 7.2.

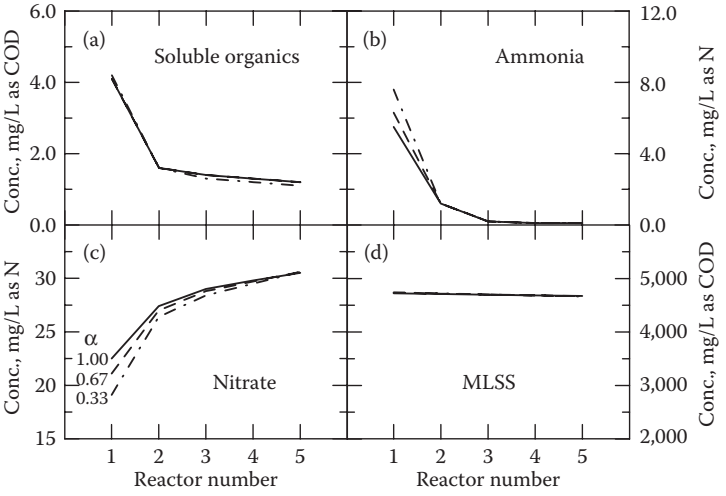

FIGURE 7.8 Effect of the biomass recycle ratio (α) on the steady-state concentrations of various constituents in each reactor of the CAS or HPOAS system described in Figure 7.2. SRT = 10 days. To express the MLSS concentration in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

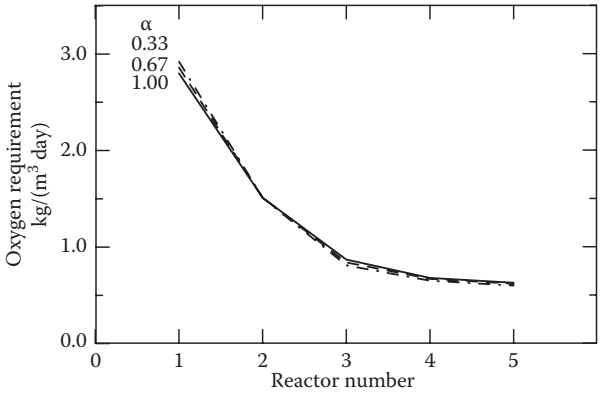

FIGURE 7.9 Effect of the biomass recycle ratio (α) on the steady-state oxygen requirement in each reactor of the CAS or HPOAS system described in Figure 7.2. SRT = 10 days.

Figures 7.8 and 7.9 show that as with a single CSTR, the rate of biomass recycle around a simple chain of CSTRs has relatively little effect on system performance. Theoretically, higher recycle flows around a chain of reactors act to make the system more homogeneous.[22] For example, in the extreme, an infinite recycle ratio would make the system behave like a CSTR. In practice, however, biomass recycle ratios in excess of 1.0 are seldom used, and thus the simulations presented in Figures 7.8 and 7.9 were limited to that value. Although the profiles do show slightly more homogeneity at higher recycle ratios, the effects are not significant, demonstrating that biomass recycle cannot be used as an operational tool for influencing substrate removal or oxygen uptake in a simple chain of CSTRs.

7.3 STEP FEED ACTIVATED SLUDGE

7.3.1 Description

Figure 7.10 presents the schematic diagram for the configuration used to simulate an SFAS system. As in Figure 7.1, five equal sized CSTRs in series were used, with all biomass recycle to the first

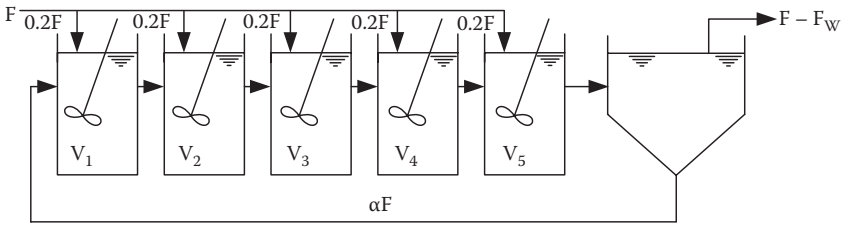

FIGURE 7.10 Schematic diagram of five CSTRs in series with all biomass recycle to the first reactor and the influent distributed evenly among the reactors. Although not shown, solids wastage is directly from all reactors. This configuration simulates step feed activated sludge (SFAS).

bioreactor, but in this case the feed was distributed evenly among the bioreactors. All other characteristics of the system, including the flow rates, feed concentrations, and so on, were the same as those used to simulate the performance of the tanks-in-series system.

7.3.2 Effect of SRT on Steady-State Performance

The effect of SRT on the steady-state performance of the system in Figure 7.10 is shown by the solid curves in Figure 7.11. For comparison, the performance of a single CSTR with volume equal to the total volume of the SFAS system is shown by the dashed curves. Examination of parts a, b, and c of the figure reveals that the effluent concentrations of soluble constituents from the SFAS system are slightly higher than those from the single CSTR. The differences are relatively small, however, and would probably be difficult to distinguish in practice. Further evidence for the similarity in performance between the SFAS system and a single CSTR is provided in Figure 7.12 where it may be seen that the oxygen requirements and solids wastage rates are almost identical in the two systems. This suggests that the mass of MLSS in the SFAS system is approximately the same as that in a single CSTR with the same SRT.

The failure of the SFAS system to perform as well as a simple CSTR (and by extension, as well as a simple chain) is a direct result of the feed distribution. If the effect of the wastage flow rate, F_w, is assumed to be small, a mass balance on MLSS about the final settler in a system with biomass recycle reveals that:

$$X_{M,T,r} \approx \frac{1+\alpha}{\alpha} X_{M,T}, \tag{7.7}$$

where $X_{M,T}$ is the MLSS concentration in total suspended solids (TSS) units entering the settler from the last bioreactor, $X_{M,T,r}$ is the concentration in the biomass recycle flow, and α is the recycle ratio. For the operational conditions described in the legend of Figure 7.11, α has a value 0.5, making the MLSS concentration in the biomass recycle stream three times the concentration in the last bioreactor of the chain. Because the influent flow is distributed evenly among the bioreactors, only one-fifth of it enters the first bioreactor, which provides little dilution of the biomass recycle flow and makes the MLSS concentration in the first bioreactor very high. Furthermore, as the MLSS moves from bioreactor to bioreactor, more influent flow enters, diluting it. This establishes a MLSS concentration gradient through the system. However, it will be recalled from Figure 7.12 that for a given SRT, the SFAS system has the same mass wastage rate of biomass as a single CSTR, suggesting that both systems contain the same mass of MLSS. Because of this and because the volume of the single CSTR is equal to the total volume in the SFAS chain, the MLSS concentration in the single CSTR must equal the average concentration in the SFAS system. Consequently, since the SFAS system contains a MLSS concentration gradient, the MLSS concentration in its last tank must be less than the concentration in the single CSTR. That this is true is shown in Figure 7.11. Now,

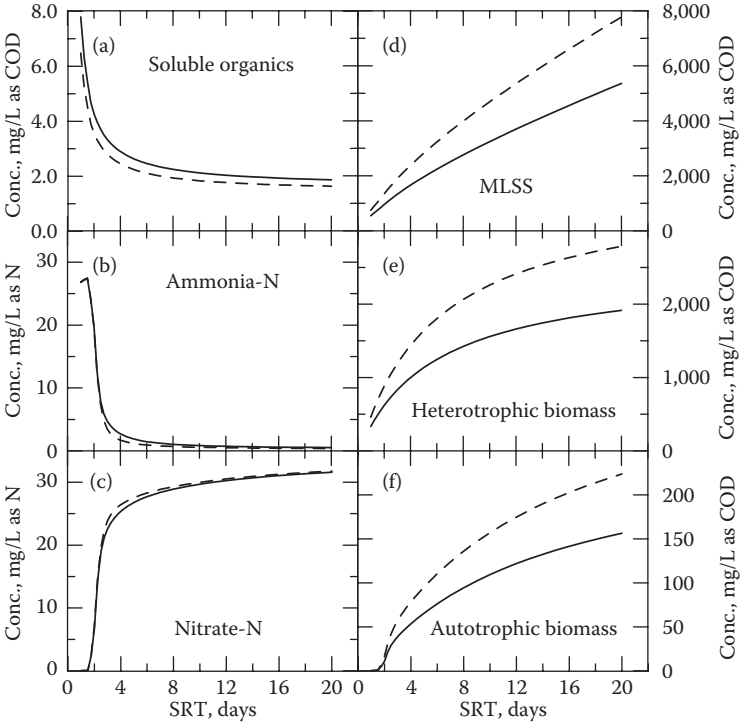

FIGURE 7.11 Effect of SRT on the steady-state concentrations of various constituents in the last reactor of the SFAS system depicted in Figure 7.10. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Influent flow = 1000 m^3/day, influent concentrations are given in Table 6.6, biomass recycle flow = 500 m^3/day, and volume of each reactor = 50 m^3. Parameters are listed in Table 6.3. The DO concentration was held constant at 2.0 mg/L. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

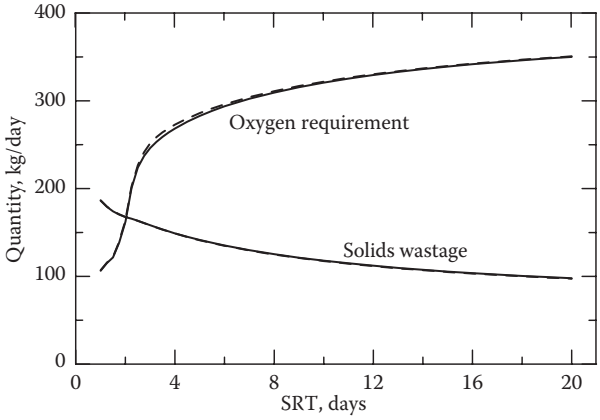

FIGURE 7.12 Effect of SRT on the total steady-state oxygen requirement and solids wastage rate for the SFAS system depicted in Figure 7.10 operating under the conditions listed in Figure 7.11. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. To express the solids wastage rate in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

since the influent flow rate into the last bioreactor of the chain is one-fifth the flow rate into the single CSTR, and since the volume of the last bioreactor is one-fifth the volume of the single CSTR, the mass flow rate of substrate per unit volume into the last bioreactor of the chain is the same as that into the single CSTR. However, since the biomass concentration is less, there is less biomass per unit of substrate added (i.e., the process loading factor is higher), allowing less substrate to be removed, as observed. Thus, the very configuration of the SFAS system prevents it from performing as well as a single CSTR or a simple chain. This raises the question as to why such a system would be used. One reason can be seen by examining the dynamic performance of the system.

7.3.3 Dynamic Performance

The dynamic performance of the SFAS system when subjected to the same diurnal loading pattern as the tanks-in-series system is shown in Figure 7.13. Again, the response of a single CSTR is shown for comparison. As would be expected from the entrance of feed directly into the last bioreactor with its low biomass concentration, the dynamic performance of the SFAS system is worse than that of the CSTR. Furthermore, although the difference is relatively small for soluble organic substrate, it is substantial for nitrification, with the maximum effluent ammonia-N concentration being almost twice as high as that from the CSTR. This is a direct result of the low maximum specific growth rate of nitrifying bacteria, which prevents them from responding rapidly enough to the changing input

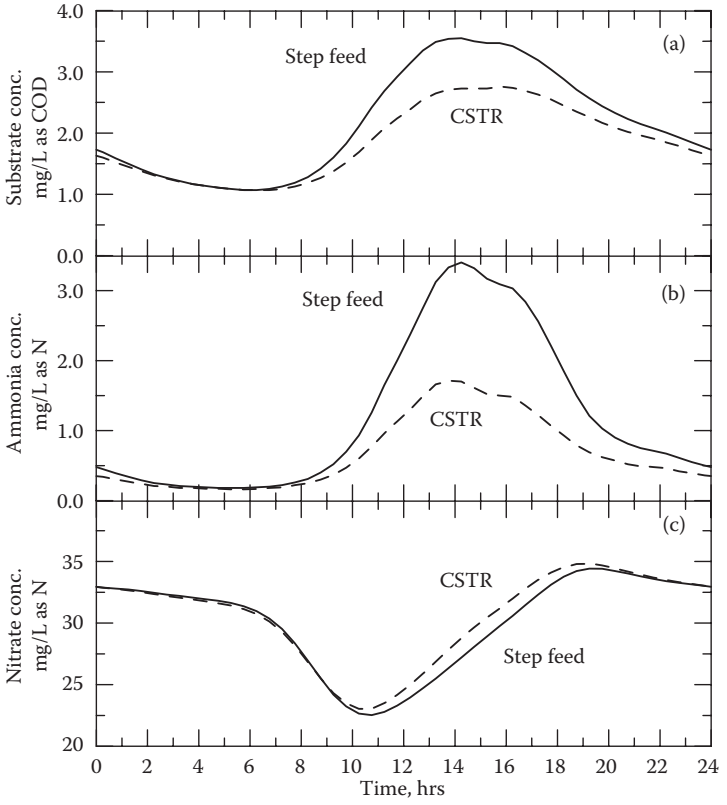

FIGURE 7.13 The time dependent response from the SFAS system depicted in Figure 7.10 when subjected to the diurnal loading patterns shown in Figure 6.2. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Average influent flow = 1000 m^3/day, average influent concentrations are given in Table 6.6, biomass recycle flow = 500 m^3/day, volume of each reactor = 50 m^3, and SRT = 10 days. Parameters are listed in Table 6.3. The DO concentration was held constant at 2.0 mg/L.

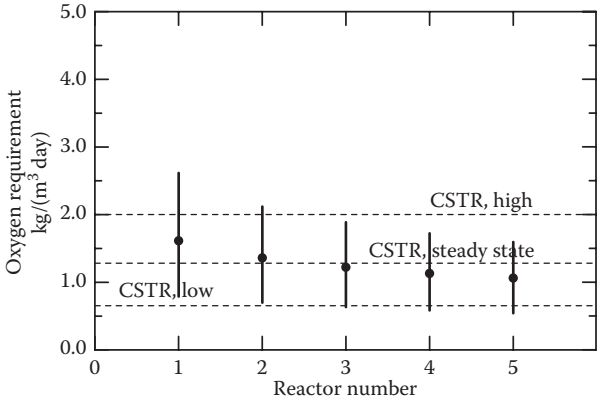

FIGURE 7.14 The variability of the oxygen requirement in each reactor for the SFAS system described in Figure 7.13. The vertical lines represent the range observed over a 24-hr period and the dots represent the requirement when the influent flow and concentrations are constant at the flow-weighted average values (steady state). For comparison, the dashed lines show the extreme diurnal requirements, as well as the steady-state requirement, for a single CSTR with a volume of 250 m^3.

rate of nitrogen into the last bioreactor. In addition, comparison of the effluent ammonia-N concentration to that from the simple chain in Figure 7.4 reveals that a SFAS system subjected to diurnal loading produces an effluent that is much worse than the chain. Thus, the dynamic performance of the SFAS system does not justify its use.

A major reason for employing a SFAS system, however, is revealed by Figure 7.14, where it can be seen that the distribution of the load along the length of the chain gives an oxygen consumption pattern that is much closer to that of a single CSTR than was the simple chain. This makes it much easier to provide the needed oxygen. Consequently, if a tanks-in-series system is not performing properly because of an inability to supply sufficient oxygen to the first tanks in the system, redistribution of a portion of the influent to tanks further down the chain may be able to alleviate the problem and produce an acceptable effluent, even though it may not be as good as that from a properly designed tanks-in-series system.

These simulations have only examined the case of distributing the feed evenly along the chain. Obviously, many possible distribution patterns exist and each will have a different impact on system performance. A study of the impact of alternative patterns will be left as an exercise for the reader.

7.3.4 Variations within the System

Tank-to-tank variations in concentration within an SFAS system are shown in Figure 7.15 for steady-state operation. Panel d of the figure shows the gradient in MLSS concentration discussed earlier, which is steeper at longer SRTs. This gradient is the other major reason for using the SFAS system.[18] Occasionally the settling characteristics of activated sludge will deteriorate so that the mass loading rate of solids entering the final settler must be reduced to prevent the settler from failing. Since the flow rate of wastewater being treated cannot be reduced, this requires reduction of the MLSS concentration. One way to do this is to reduce the SRT, but this may negatively impact other aspects of process performance, such as causing loss of nitrification. If the system were operating in the tanks-in-series mode, however, and contained the appropriate piping, switching to the SFAS mode would establish the MLSS concentration gradient, thereby reducing the concentration of MLSS entering the final settler while maintaining the same system SRT. Then, after the biomass settling characteristics had returned to normal, the operation could be returned to the tanks-in-series mode.

The concentration profiles of soluble organic substrate and ammonia-N are the opposite of the MLSS profiles. This follows from the fact that less biomass is available in the later bioreactors to

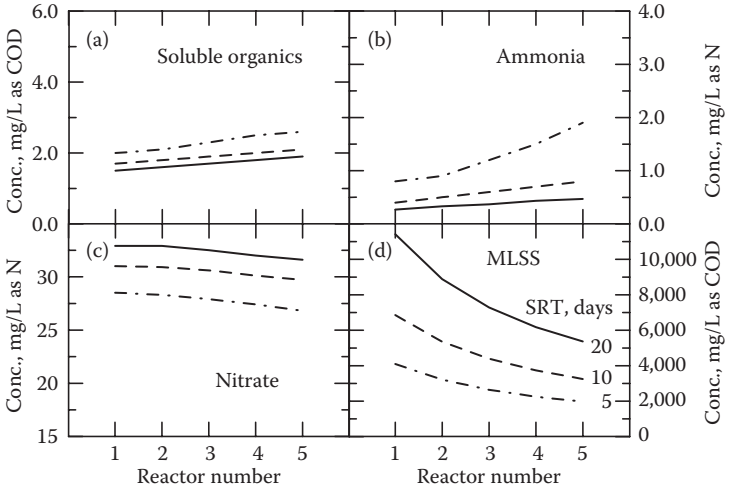

FIGURE 7.15 Effect of SRT on the steady-state concentrations of various constituents in each reactor of the SFAS system described in Figure 7.11. To express the MLSS concentration in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

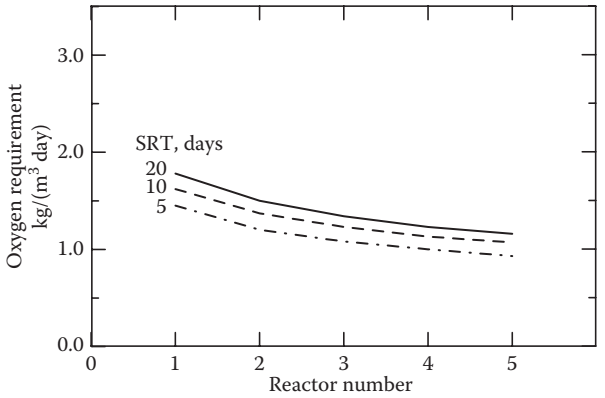

FIGURE 7.16 Effect of SRT on the steady-state oxygen requirement in each reactor of the SFAS system described in Figure 7.11.

act on the substrate and ammonia entering them, thereby causing the concentrations to rise. In other words, the process loading factor increases from bioreactor to bioreactor down the chain. The tank-to-tank changes in substrate and ammonia-N concentrations are less at longer SRTs, however, because the larger mass of biomass in the system at longer SRTs allows more complete reaction. This is particularly evident for nitrification, which is a slower reaction than organic substrate removal.

The steady-state profiles in oxygen requirement through the system are shown in Figure 7.16. They respond more systematically to changes in SRT than the profiles through the tanks-in-series system, shown in Figure 7.7, primarily because the soluble organic substrate and ammonia-N profiles are less severe. The major impact of an increase in SRT is to increase the oxygen requirement by increasing the importance of biomass decay.

The effect of the biomass recycle ratio on the steady-state concentration profiles is shown in Figure 7.17, while the effect on the oxygen requirement is illustrated in Figure 7.18. Examination of those figures shows that changes in the recycle ratio have a greater effect on the effluent quality of this system than on that of the simple chain (Figure 7.8) or the single CSTR (upon which the recycle ratio has no effect). This is due entirely to the hydraulics of the system. All of the systems

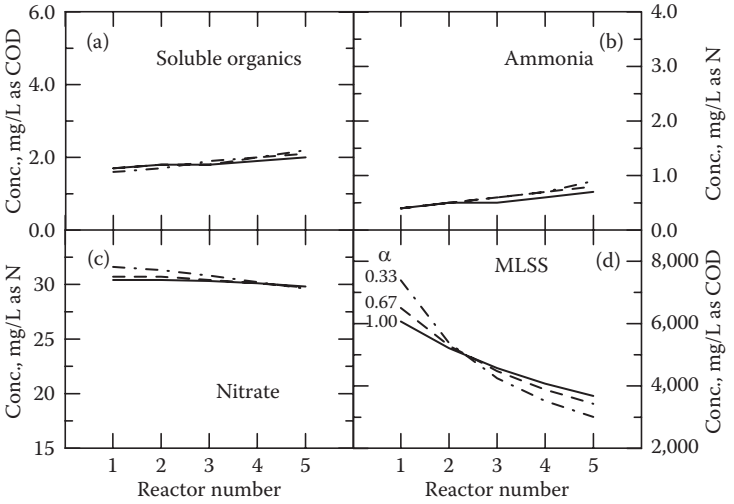

FIGURE 7.17 Effect of the biomass recycle ratio (α) on the steady-state concentrations of various constituents in each reactor of the SFAS system described in Figure 7.11. SRT = 10 days. To express the MLSS concentration in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

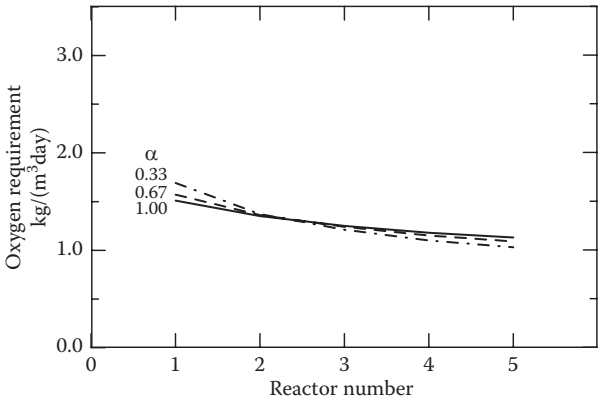

FIGURE 7.18 Effect of the biomass recycle ratio (α) on the steady-state oxygen requirement in each reactor of the SFAS system described in Figure 7.11. SRT = 10 days.

whose responses are shown in Figures 7.17 and 7.18 have the same SRT. Thus they all contain the same mass of MLSS. Furthermore, they all have the same reactor volumes, and thus they all have the same average MLSS concentrations. However, as shown by Equation 7.7, as the recycle ratio is increased, the difference between the concentration of MLSS entering the settler and the concentration leaving in the recycle flow decreases. This means that the MLSS concentration gradient along the bioreactor chain is reduced, as shown in Figure 7.17d. Furthermore, since the recycle ratio doesn't influence the average MLSS concentration, the MLSS concentration in the first bioreactors must decrease with increasing recycle while the MLSS concentration in the last ones must increase. This decreases the process loading factor in the last bioreactors, allowing more substrate and ammonia-N to be removed, thereby making their effluent concentrations lower. Finally, since an increase in the recycle ratio makes the system more homogeneous, it will make the utilization of oxygen more uniform, as shown in Figure 7.18.

Although the effects associated with changes in the recycle ratio are relatively small for this system configuration, they demonstrate clearly that the performance of suspended growth biological

reactors is not always independent of the recycle ratio. In fact, it has an even stronger effect on the next reactor configuration.

7.4 CONTACT STABILIZATION ACTIVATED SLUDGE

7.4.1 Description

As described previously, in the contact stabilization version of the activated sludge process, the biomass recycle flow enters a bioreactor (stabilization basin) where it undergoes aeration and reaction prior to its return to the bioreactor receiving the influent wastewater (contact tank). This system can be simulated as a two bioreactor chain, as shown in Figure 7.19, in which all biomass recycle enters the first bioreactor and all influent enters the second. To maintain consistency with previous simulations, a total bioreactor volume of 250 m^3 was assumed, split evenly between the two bioreactors. The influent flows and concentrations were the same as those used in previous simulations.

7.4.2 Effect of SRT on Steady-State Performance

The effect of the SRT on the steady-state performance of a CSAS system is shown by the solid curves in Figure 7.20, while the dashed curves present the performance of a single CSTR for comparison. Examination of the figure reveals that the effluent concentrations of soluble organic matter and ammonia-N are both higher than that of the CSTR, showing that the contact stabilization system does not perform as well. The reason becomes apparent when we consider the distribution of biomass within the system, as we did for the SFAS system. Examination of Figure 7.21 reveals that in spite of the slightly poorer performance, the CSAS system has almost the same excess biomass production as a single CSTR. As was argued for the SFAS system, this suggests that the CSAS system contains approximately the same mass of MLSS as the CSTR. However, the only flow entering the first bioreactor of the CSAS system is biomass recycle flow, which contains a much higher concentration of MLSS than the flow entering the settler. Since the CSAS system contains the same mass of MLSS as the CSTR, but the first bioreactor in it contains MLSS at a very high concentration, the MLSS concentration in the second bioreactor must be less than the concentration in the single CSTR, as shown in Figure 7.20. Furthermore, since the second bioreactor in the CSAS system receives the same influent wastewater flow rate as the single CSTR, but has only half the volume of the CSTR and contains a lower concentration of MLSS, the CSAS system cannot perform as well as the single CSTR. In other words, the process loading factor for the contact tank is much higher than it is for the CSTR, so less substrate will be removed. However, as we will see later, performance of the CSAS process can be changed by altering both the relative volumes of the two bioreactors and the recycle flow rate. This means that the degree of difference in performance between the two systems depends on the configuration chosen for the CSAS system. In addition, wastewater

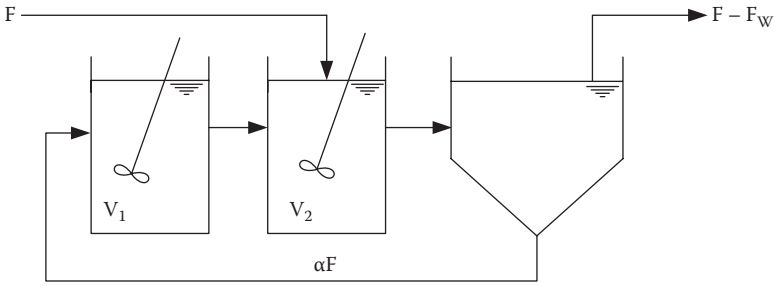

FIGURE 7.19 Schematic diagram of two CSTRs in series with all influent to the second (contact) reactor and all biomass recycle to the first (stabilization) reactor. Although not shown, solids wastage is directly from both reactors. This configuration simulates contact stabilization activated sludge (CSAS).

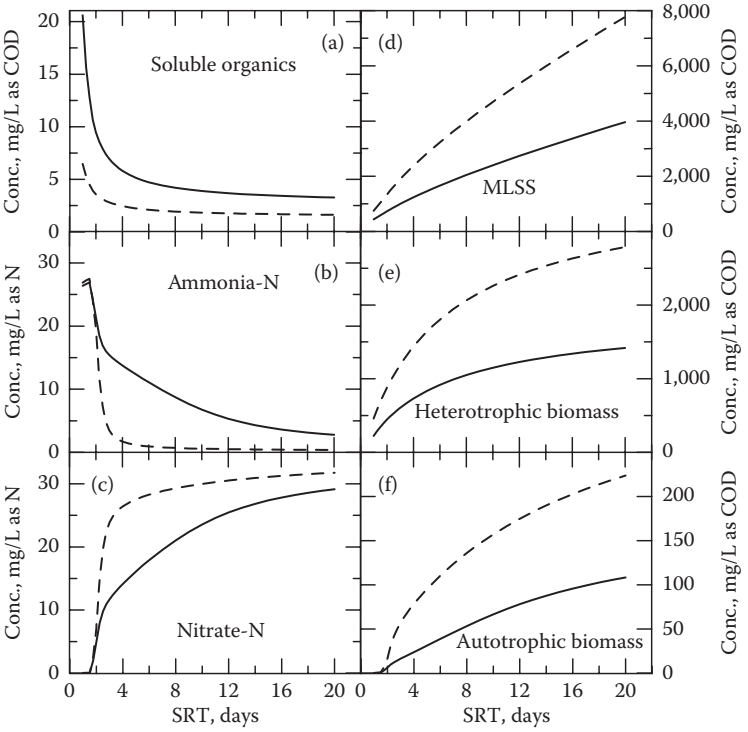

FIGURE 7.20 Effect of SRT on the steady-state concentrations of various constituents in the contact (second) reactor of the CSAS system depicted in Figure 7.19. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Influent flow = 1000 m^3/day, influent concentrations are given in Table 6.6, biomass recycle flow = 500 m^3/day, and volume of each reactor = 125 m^3. Parameters are listed in Table 6.3. The DO concentration was held constant at 2.0 mg/L. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

characteristics influence system performance. For example, colloidal and particulate organic matter are entrapped in the MLSS in the contact tank and undergo more complete biodegradation in the stabilization tank. This suggests that the CSAS system is better suited for wastewaters containing a higher fraction of their organic content in the colloidal form than in the soluble form. These and other factors influencing system performance will be discussed later when we consider the choice of system configuration for activated sludge systems. For now, however, we will concentrate on understanding system performance for the standard wastewater being used in all of these simulations.

First, consider Figure 7.20a where the removal of soluble organic matter is considered. The most striking thing about the performance of the CSAS system is how closely it parallels the single CSTR. The shapes of the curves are the same; only the magnitudes are different. This follows from the fact that the mass of heterotrophic biomass in the two systems is similar, which is caused by the nature of the organic substrate in the influent. Recall from Table 6.6 that the influent contains more particulate than soluble organic substrate. This particulate substrate is entrapped in the MLSS, making it available for microbial attack in both the contact and stabilization tanks. Because the SRT is the same in the CSAS system and the single CSTR, and because particulate substrate is attacked in both tanks, the opportunity for degradation of the particulate substrate is essentially the same in the two systems. That degradation results in the growth of heterotrophic biomass, which can then attack the soluble organic matter in the contact tank, resulting in even more biomass. Since more than half of the organic substrate is particulate and will be removed totally in the contact tank, and a substantial portion of the soluble organic substrate is removed at the SRTs studied, there is relatively little difference in the mass of heterotrophic biomass formed in the two systems, and thus

they perform in a similar manner. The concentration of soluble substrate in the effluent from the CSAS system is higher simply because it is removed only in the contact tank, which has a smaller volume and a lower biomass concentration than the single CSTR, giving the contact tank a higher process loading factor.

In contrast, the shape of the ammonia-N curve for the CSAS system is quite different from that of the single CSTR. This difference is also reflected by the shapes of the autotrophic biomass curves for the two systems and is primarily due to the fact that most of the influent nitrogen is present in the soluble form, but also in part to the low maximum specific growth rate of autotrophic nitrifying bacteria. Examination of Table 6.6 reveals that the wastewater contains some particulate organic nitrogen, which will be converted to ammonia-N as the particulate organic matter undergoes biodegradation. Because the particulate organic nitrogen is entrapped in the MLSS, it is present throughout the entire system and its biodegradation provides ammonia-N that is available to the nitrifying bacteria in both tanks. In addition, that portion of the mass inflow of soluble nitrogen that is recycled through the stabilization tank is also available to the nitrifiers in both bioreactors. For the recycle ratio used (0.5), this is about one-third of the influent soluble nitrogen. The remainder is only available to the nitrifiers in the contact tank. When the SRT is short, but long enough to prevent wash out of the nitrifiers, they will grow using the nitrogen that is available throughout the entire system and provide the basis for ammonia-N oxidation in the contact tank. However, the quantity of nitrifiers formed will be limited primarily by the amount of nitrogen that is available in the stabilization tank, because the residence time of the biomass is greater in it and there is greater opportunity for degradation of the particulate organic nitrogen, thereby making ammonia-N available to the nitrifiers. Nitrifiers grown in the stabilization tank will then pass to the contact tank where they can oxidize a portion of the ammonia-N entering from the feed. This limitation forms the break in the ammonia-N curve at short SRTs. As the SRT is increased, further reductions in the ammonia-N concentration will be due to its greater utilization in the contact tank. Because of the low half-saturation coefficient associated with autotrophic biomass growth, the autotrophs will be growing at their maximal rate in the contact tank and thus the mass of ammonia-N removed will be governed by the mass of autotrophs present. As long as excess ammonia-N is available, the mass of autotrophs will increase almost linearly with the SRT, which means that the ammonia-N concentration will decrease almost linearly, as shown. Only when the ammonia-N concentration drops sufficiently to cause the specific growth rate of the autotrophs in the contact tank to be governed by that concentration does the curve depart from linearity.

Figure 7.21 shows that significantly less oxygen is used in the CSAS system than in the single CSTR even though the solids wastage rates are very similar. This difference is due to the differences in the amount of nitrification. It will be recalled from the discussion in Section 6.3 that nitrification has a major impact on oxygen utilization but almost no impact on biomass production.

7.4.3 Dynamic Performance

Because the influent enters only the contact tank, because the concentration of biomass in that tank is low, and because its volume is half that of the single CSTR, we might expect the dynamic performance of this bioreactor system to be worse than any we have encountered so far and that is the case, as shown in Figure 7.22. The nitrification performance of the system is particularly poor. Reexamination of Figure 7.20 shows that at an SRT of 10 days, the value used for the dynamic simulation, steady-state nitrification is incomplete, with an effluent ammonia-N concentration of about 7 mg/L. This means that even at steady state, the nitrifying bacteria are growing near their maximal rate in the contact tank. Consequently, when the diurnal load is applied, no excess nitrification capacity exists to oxidize the additional ammonia-N that enters during peak loading periods, causing most of it to pass through to the effluent. Relatively complete nitrification only occurs when the influent mass flow rate of ammonia is sufficiently low for the mass of nitrifiers in the system to handle it.

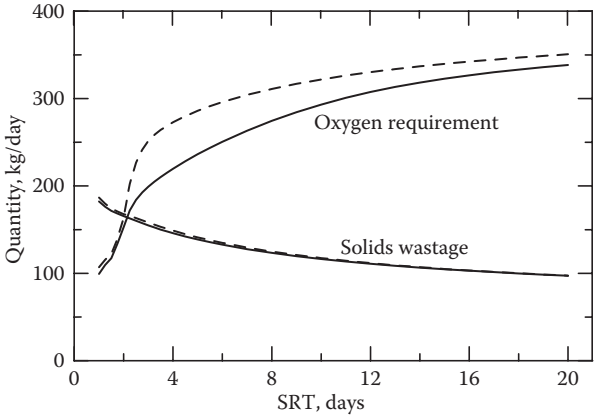

FIGURE 7.21 Effect of SRT on the total steady-state oxygen requirement and solids wastage rate for the CSAS system depicted in Figure 7.19 operating under the conditions listed in Figure 7.20. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. To express the solids wastage rate in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

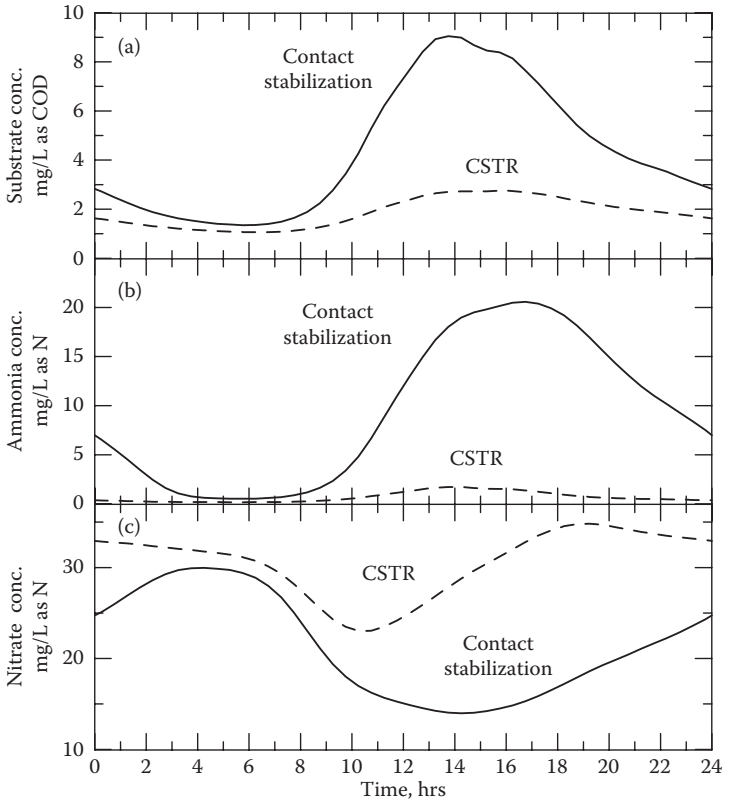

FIGURE 7.22 The time dependent response of the effluent from the CSAS system depicted in Figure 7.19 when subjected to the diurnal loading patterns shown in Figure 6.2. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Average influent flow = 1000 m^3/day, average influent concentrations are given in Table 6.6, biomass recycle flow = 500 m^3/day, volume of each reactor = 125 m^3, and SRT = 10 days. Parameters are listed in Table 6.3. The DO concentration was held constant at 2.0 mg/L.

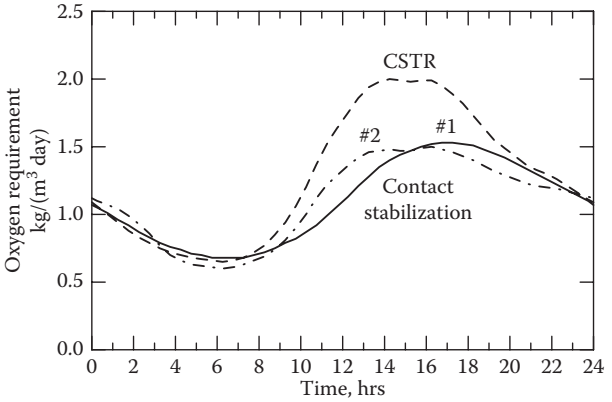

FIGURE 7.23 The time dependent variability in the oxygen requirement in each reactor of the CSAS system described in Figure 7.22. The solid curve represents the stabilization (first) reactor and the dashed-dot curve the contact (second) reactor. For comparison, the dashed curve shows the requirement in a single CSTR with a volume of 250 m^3.

The diurnal oxygen requirements in each of the bioreactors of the CSAS system are shown in Figure 7.23, along with the requirement in a single CSTR. The surprising thing about the curves is that the oxygen utilization rate in the stabilization tank is almost as dynamic as the utilization rate in the contact tank. Because the influent only flows through the contact tank and the stabilization tank receives a constant flow rate, one might expect the stabilization tank to show a less severe response. There are two reasons why it does not. One is that over half of the organic loading to the CSAS system is due to particulate organic matter, which is degraded in both bioreactors. Since its input varies in a diurnal manner, so does its degradation. The other is the transport of ammonia-N into the stabilization tank through the biomass recycle flow. Since the ammonia-N concentration gets quite high in the contact tank, an appreciable quantity enters the stabilization tank where the longer HRT and higher nitrifier mass allow its oxidation. The time lag associated with the transport of these materials causes a shift in the times at which the maximum and minimum uptake occur in the two tanks, however.

7.4.4 Effects of System Configuration

We saw earlier that the recycle ratio influences the performance of the SFAS system because of its effect on the distribution of biomass in the system. Thus, we would expect the recycle ratio to also affect the performance of the CSAS system, which it does, as shown in Figure 7.24. In this figure, which was generated for an SRT of 10 days, the dashed lines represent the concentrations in the contact (second) reactor, and thus represent the concentrations entering the settler, whereas the solid lines represent the concentrations in the stabilization (first) reactor.

Consider first the concentrations of heterotrophic biomass. The mass of heterotrophic biomass in the system is essentially independent of the recycle ratio between 0.1 and 1.0 because organic substrate removal is almost complete (relative to the influent) for all of those values. Thus, the differences in the concentrations in the two bioreactors shown in Figure 7.24e reflect primarily the effect of the recycle ratio on the concentrations of biomass entering the settler and leaving in the biomass recycle stream as given by Equation 7.7. Because the microbial mass is fixed, an increase in the recycle ratio simply shifts heterotrophic biomass from the stabilization tank to the contact tank. An increase in the mass of heterotrophic biomass in contact with the wastewater allows more soluble organic constituents to be removed, thereby improving system performance, as shown in Figure 7.24a. As might be expected, almost all soluble organic matter is gone from the stabilization tank regardless of the recycle ratio and the residual simply reflects a balance between its utilization and its production through biomass death and lysis.

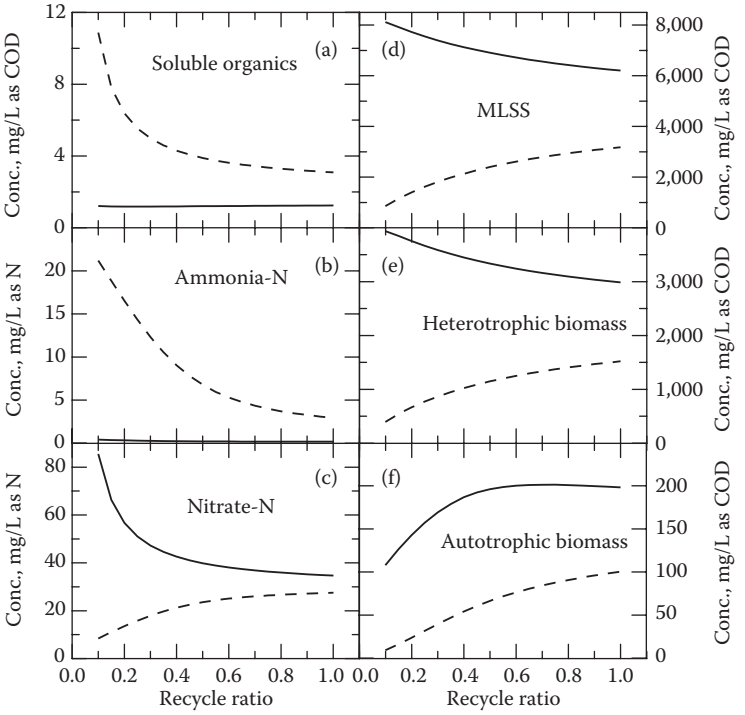

FIGURE 7.24 Effect of the recycle ratio (α) on the steady-state concentrations of various constituents in each reactor of the CSAS system described in Figure 7.20. SRT = 10 days. The solid curves represent the stabilization (first) reactor and the dashed curves the contact (second) reactor. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

The response of the autotrophic biomass is very different from that of the heterotrophic biomass and reflects the fact that the mass of autotrophic bacteria increases as the recycle ratio is increased. When the recycle ratio is small, only a small percentage of the biomass is in the contact tank. As a consequence, only a small fraction of the ammonia-N is oxidized. Furthermore, because the recycle flow rate is small, only a small portion of the ammonia-N in the contact tank effluent is transported to the stabilization tank for oxidation. Thus, only a small percentage of the total mass of nitrogen passing through the system is oxidized and only a small mass of autotrophic biomass is formed. An increase in the recycle ratio has two effects, however. First, it shifts more of the biomass from the stabilization tank to the contact tank, allowing more ammonia-N oxidation in the contact tank, thereby forming more autotrophic biomass. Second, it transports a greater fraction of the unreacted ammonia-N into the stabilization tank, allowing formation of even more autotrophic biomass. By the time the recycle ratio is around 0.8, the majority of the nitrogen flowing into the system is being oxidized, giving a relatively constant mass of nitrifiers, so that further changes in the recycle ratio simply act to redistribute them in the same manner as the heterotrophic biomass.

One other point about nitrification needs clarification and that concerns the concentration of nitrate-N in the stabilization tank. It can be seen in Figure 7.24c that the nitrate-N concentration is very high when the recycle ratio is very low. However, even though that concentration is high, it represents only a small fraction of the nitrogen flowing through the system. Furthermore, most of that nitrate comes from ammonia-N released as a result of biomass decay and particulate substrate degradation in the stabilization tank. Because the MLSS concentration is high there, the concentration of nitrogen released will be high. As stated above, however, it does not represent a large mass, as reflected by the low nitrate-N concentration in the final system effluent, which is the same as the concentration in the contact tank.

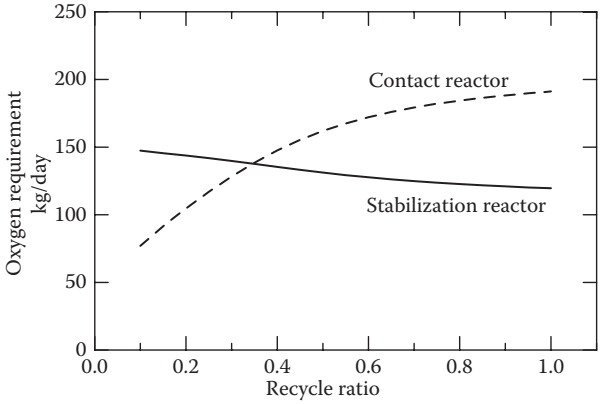

FIGURE 7.25 Effect of the recycle ratio (α) on the steady-state oxygen requirement in each reactor of the CSAS system described in Figure 7.20. SRT = 10 days. The solid curves represent the stabilization (first) reactor and the dashed curves the contact (second) reactor.

Figure 7.25 shows the effect of the recycle ratio on the oxygen utilization in the two tanks. Since an increase in the recycle ratio shifts biomass from the stabilization tank to the contact tank, it shifts the oxygen requirement in a like manner. In addition, however, an increase in the recycle ratio also allows more nitrification in both tanks, although more of the nitrification increase will occur in the contact tank. Thus, there is a greater increase in the oxygen requirement in the contact tank than there is a decrease in the stabilization tank.

Another way to shift the relative amounts of biomass in the two tanks is to change their sizes. Thus simulations were done in which the total bioreactor volume was held constant but the relative sizes of the two tanks were changed. The SRT was held constant at 10 days and all other conditions were the same as for the simulations shown in Figure 7.20. The results are shown in Figure 7.26 where they are plotted as a function of the fraction of the total system volume in the contact tank (bioreactor 2). When all of the volume is in that tank the system is just a single CSTR. Thus, the responses shown for that fraction are the same as those shown earlier for a single CSTR.

Consider first the concentrations of MLSS and heterotrophic biomass shown in parts d and e of Figure 7.26. For an SRT of 10 days, the mass of each in the system is reasonably constant over the range of relative tank sizes considered. This suggests that the changes in the concentrations of each shown in the figure are due primarily to their distributions within the system. Consequently, the concentration curves must satisfy two criteria. First, the sum of the masses in each tank (i.e., volume times concentration) must be relatively constant. Second, the concentration in the stabilization tank must be related to the concentration in the contact tank in approximately the same way that the concentration in the biomass recycle flow is related to the concentration entering the settler (i.e., through Equation 7.7). In other words, as more volume is shifted to the contact tank, a smaller percentage of the system biomass is at the concentration of the biomass recycle and thus the concentration in the system must increase in order to contain the same mass in the same total volume. Those changes in biomass concentration, in turn, impact on the concentrations of soluble constituents leaving the contact tank.

As might be expected from the previous discussion, the concentrations of soluble organic matter and ammonia-N leaving the system decrease as the fraction of the system volume in the contact tank increases. There are two reasons for this: the fraction of the biomass in contact with the wastewater increases, as discussed above, and the residence time of the wastewater in contact with the biomass increases. Both of these act to reduce the process loading factor in the contact tank, which acts to reduce the substrate concentration in it. However, it will be noted in Figure 7.26 that the responses of soluble organic substrate and ammonia-N are quite different. This follows from the fact that heterotrophic biomass arises from the destruction of both soluble and particulate organic

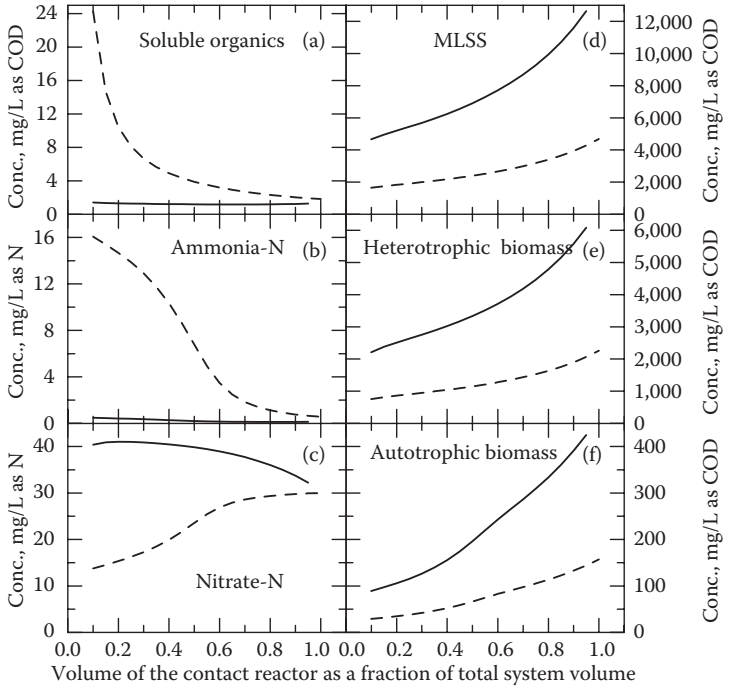

FIGURE 7.26 Effect of the relative volumes of the two reactors on the steady-state concentrations of various constituents in each reactor of the CSAS system described in Figure 7.20. The total system volume was constant at 250 m^3. SRT = 10 days. The solid curves represent the stabilization (first) reactor and the dashed curves the contact (second) reactor. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

matter, whereas autotrophic biomass comes primarily from destruction of soluble material, as discussed previously for changes in SRT. As a consequence, the fraction of autotrophs in the biomass increases as the fraction of the system volume in the contact tank increases, as reflected by the differences between the heterotrophic and autotrophic biomass curves.

Figure 7.27 shows the oxygen requirements in the two bioreactors as their relative sizes are changed. As would be anticipated from the previous discussion, oxygen demand is shifted from the stabilization tank to the contact tank as volume is shifted to the contact tank. The fact that the increase in the oxygen requirement in the contact tank exceeds the decrease in the requirement in the stabilization tank reflects the increased amount of nitrification that occurs as the relative volumes are changed.

The CSAS process is the most complex studied so far, since effluent quality can be affected by SRT, recycle ratio, and relative reactor volumes. This suggests that its design is more complex than that of simpler systems. Although the application of optimization techniques is the best way to arrive at a sound design, in Chapter 11 we will review some general rules that can be used by the engineer designing such a system.

7.5 MODIFIED LUDZACK–ETTINGER PROCESS

7.5.1 Description

All of the systems considered so far in this chapter have been totally aerobic. As a consequence, no significant denitrification occurs, and thus no nitrogen is removed; it is simply transformed from ammonia-N to nitrate-N. If one wished to add a denitrification reactor downstream of any of the

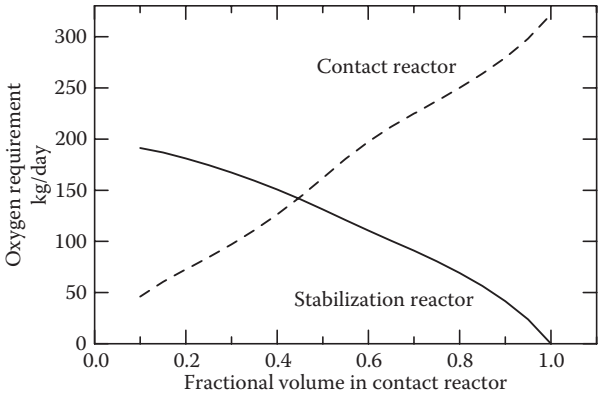

FIGURE 7.27 Effect of the relative volumes of the two reactors on the steady-state oxygen requirement in each reactor of the CSAS system described in Figure 7.20. The total system volume was constant at 250 m^3. SRT = 10 days. The solid curves represent the stabilization (first) reactor and the dashed curves the contact (second) reactor.

systems studied previously, it would be necessary to add an electron donor because when the SRT is made sufficiently long to allow oxidation of the ammonia-N to nitrate-N, all of the soluble organic substrate will be degraded. As a consequence, the only source of organic matter to serve as the electron donor for denitrification would be through hydrolysis of particulate substrate entrapped in the MLSS or released through biomass death and lysis. Hydrolysis is a slow reaction, however, and thus very large anoxic bioreactors would be needed to achieve even partial denitrification in this manner. Adding organic matter, such as methanol, to the bioreactor would overcome this problem but adds to the system's operating costs.

Ludzack and Ettinger[24] reasoned that it should be possible to use the readily biodegradable substrate in the wastewater itself as an electron donor to achieve partial denitrification if there were a way to bring the nitrate formed in the reactor system back to a point where the substrate was available. They achieved this by separating the bioreactor into two compartments, with the first receiving no aeration, and by pumping a stream of mixed liquor from the actively nitrifying aerobic zone back to the anoxic zone to carry nitrate to it. This system is referred to as the modified Ludzack-Ettinger (MLE) system to differentiate it from some of their earlier work. Although many other systems have since been developed within which both nitrification and denitrification occur with greater efficiency,[34,40] the MLE system represents one of the simplest within which both processes take place. Thus, it provides a convenient system with which to investigate through modeling the effects of various system variables on both processes.

The system chosen to simulate the MLE process is shown schematically in Figure 7.28. The bioreactors are each 125 m^3, as in the last system, but all influent flow and biomass recycle enter the first one, which is anoxic with a DO concentration of zero. The second bioreactor is maintained at a DO concentration of 2.0 mg/L, as has been done in all previous simulations. To provide nitrate for anoxic growth of the heterotrophic bacteria in the first bioreactor, mixed liquor from bioreactor 2 is pumped to bioreactor 1 at a rate of 2000 m^3/day or twice the influent flow rate. To distinguish this stream from the biomass recycle stream, which has a higher MLSS concentration and is at a lower flow rate, it will be called the mixed liquor recirculation (MLR) stream.

7.5.2 Effect of SRT on Steady-State Performance

The effects of SRT on the concentrations in the second bioreactor of the MLE system are shown by the solid curves in Figure 7.29. For comparison, the concentrations in a single aerobic CSTR with a volume of 250 m^3 are shown as the dashed curves. As expected, the most obvious difference

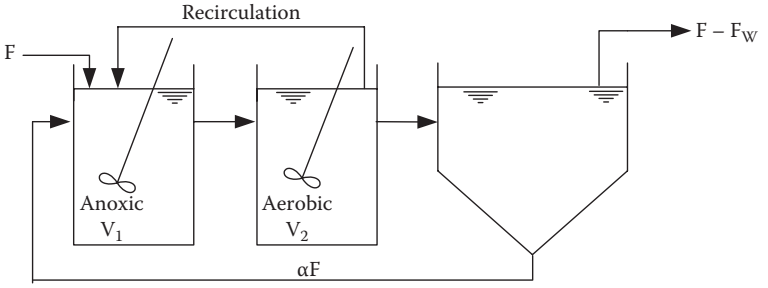

FIGURE 7.28 Schematic diagram of two CSTRs in series with all influent and all biomass recycle to the first reactor, in which the first reactor is anoxic and receives mixed liquor recirculation flow from the second, which is aerobic. Although not shown, solids wastage is directly from both reactors. This configuration simulates the MLE process.

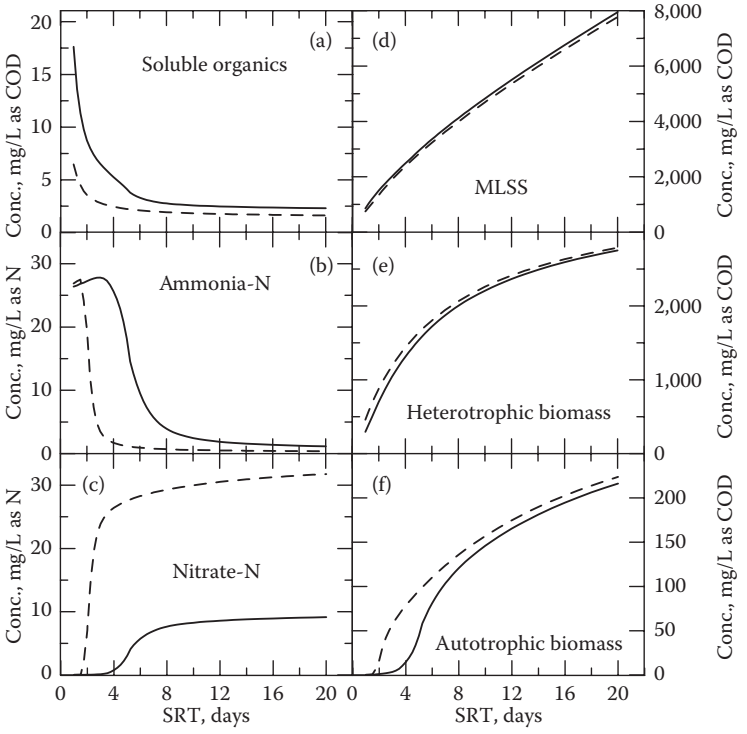

FIGURE 7.29 Effect of SRT on the steady-state concentrations of various constituents in the aerobic (last) reactor of the MLE system depicted in Figure 7.28. For comparison, the dashed curves represent the performance of a single aerobic CSTR with a volume of 250 m^3. Influent flow = 1000 m^3/day, influent concentrations are given in Table 6.6, biomass recycle flow = 500 m^3/day, mixed liquor recirculation flow = 2000 m^3/day, and volume of each reactor = 125 m^3. Parameters are listed in Table 6.3. The DO concentration is zero in the anoxic (first) reactor and 2.0 mg/L in the aerobic (second) reactor. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

between the two systems is the nitrate-N concentration, which is much lower in the MLE system due to the denitrification in the first bioreactor. There are some other differences that should be recognized, however.

An examination of the ammonia-N, nitrate-N, and autotrophic biomass curves shows that a longer system SRT is required for the onset of nitrification in the MLE system. This is because

nitrifiers can grow in only half of the system volume, the second half where DO is present. As discussed in Section 6.5.2, the aerobic SRT is recognized as the variable of importance in determining the growth of autotrophic bacteria in systems containing anoxic zones. Since the mass of biomass in the aerobic zone is half the total mass in this MLE system, the aerobic SRT is one-half of the total SRT. Examination of the curves shows that nitrification begins in the single CSTR when the SRT is about 1.5 days. Thus, we would expect nitrification to begin in the MLE system when the SRT is around 3 days, which it does. Furthermore, nitrification is well established in the CSTR when the SRT has reached 8 days and in the MLE system when it has reached 16 days. Thus, aerobic SRT is a valid concept for this system as well. Careful examination of the ammonia-N and autotrophic biomass curves reveals that, however, the curves for the single CSTR would not be coincident with the curves for the MLE system if both are plotted against aerobic SRT. This is primarily because of biomass decay in the systems, which responds to total SRT, not just aerobic SRT. For a given aerobic SRT, the net specific growth rates of the autotrophs in the two systems are different because more decay occurs in the MLE system due to its greater total SRT. Furthermore, since decay releases ammonia-N from the biomass, the input rate of ammonia-N into the MLE system is greater at a given aerobic SRT because of the larger amount of decay. These factors act together to cause the observed differences in the responses of the autotrophic biomass in the two systems.

The differences between the organic substrate curves are also related to the fact that only part of the system is aerobic. It will be recalled that a correction factor, η_g, is used in ASM No. 1 to express the fact that soluble substrate removal is slower under anoxic conditions. This retardation of the reaction rate in the anoxic zone acts to cause less soluble substrate removal in the MLE system. Furthermore, at shorter SRTs, incomplete nitrification means that insufficient electron acceptor is returned to the anoxic zone to allow biodegradation of all of the organic substrate. Thus, more substrate enters the aerobic zone, but because it is smaller than the single CSTR, less substrate can be removed.

One final difference that can be observed in Figure 7.29 is that the MLSS concentration is slightly higher in the MLE system even though the concentrations of heterotrophic and autotrophic biomass are lower. This reflects the presence of more slowly biodegradable substrate in the MLE system as a result of a lower rate of hydrolysis under anoxic conditions. This difference is also reflected in the solids wastage curves shown in Figure 7.30a.

Figure 7.30a also shows the oxygen requirements in the two systems. The most obvious difference between the two is that less oxygen is used in the MLE system. This is because of the use of nitrate-N as the terminal electron acceptor for the organic substrate that is oxidized in the first (anoxic) MLE reactor. If the nitrate utilization curve in Figure 7.30b is multiplied by 2.86, the oxygen equivalence factor for nitrate-N, and added to the oxygen requirement curve for the MLE system, the sum will be found to be similar to the oxygen requirement in the single aerobic CSTR at the longer SRTs where nitrification is fully established. The curves will not coincide at lower SRTs, however, because of differences in the extent of nitrification.

Another difference between the two systems, not shown in the figures, is that the MLE system will have a higher alkalinity. This is because denitrification increases alkalinity, as shown in Equation 3.22 and discussed in Section 6.4.1.

7.5.3 Effects of System Configuration

Two things can be altered in the MLE system, the amount of mixed liquor recirculation flow returning nitrate to the anoxic bioreactor and the relative sizes of the anoxic and aerobic zones. Like the single CSTR, however, the total system HRT has a negligible effect on system performance.[43] Thus, it would be instructive to examine each of the variables that has an effect. In so doing, the system SRT will be fixed at 10 days, which is adequate to allow relatively complete nitrification in the system with equal sized bioreactors.

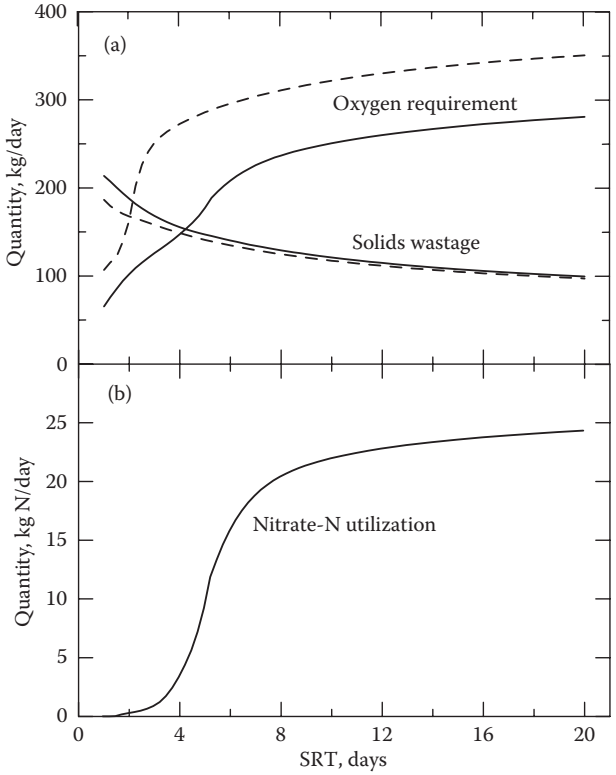

FIGURE 7.30 Effect of SRT on the total steady-state oxygen requirement, nitrate utilization rate, and solids wastage rate for the MLE system depicted in Figure 7.28 operating under the conditions listed in Figure 7.29. For comparison, the dashed curves represent the performance of a single aerobic CSTR with a volume of 250 m^3. To express the solids wastage rate in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

Figure 7.31 shows the effect of varying the mixed liquor recirculation rate, expressed as a ratio of the influent flow. Two types of curves are shown. The solid ones represent the concentrations in the first (anoxic) bioreactor and the dashed ones represent the concentrations in the second (aerobic). As might be expected, because the recycle and MLR flows distribute the MLSS around the system many times within one SRT, the MLSS and biomass behave in a completely mixed fashion and their concentrations are essentially the same in the two bioreactors. Furthermore, the concentrations of soluble organic matter and ammonia-N in the aerobic bioreactor are independent of the recirculation ratio because they are both controlled primarily by the aerobic SRT, which is constant. The only use of ammonia-N in the anoxic bioreactor is for biomass synthesis. Since that requirement is relatively small, the major factor influencing the ammonia-N concentration there is dilution by the recirculation flow, which has a uniformly low concentration. Consequently, the greater the recirculation ratio, the more dilution will occur, and the lower the concentration.

As far as the reactions in the system are concerned, the constituents most influenced by the recirculation are the soluble organic matter in the anoxic bioreactor and nitrate-N in both. When the recirculation flow is zero, nitrate can only enter the anoxic bioreactor through the biomass recycle flow (ratio = 0.5). Because the recycle flow rate is small, reaction in the anoxic bioreactor is limited by the availability of the electron acceptor and thus the anoxic bioreactor contains appreciable soluble organic matter and almost no nitrate-N. Furthermore, the minimal flow of nitrate-N to the first bioreactor limits the amount of denitrification in the system, thereby making the nitrate concentration in the effluent (i.e., the second bioreactor) high. As the recirculation flow is increased, more nitrate-N is returned to the first bioreactor, providing more of the electron acceptor and allowing

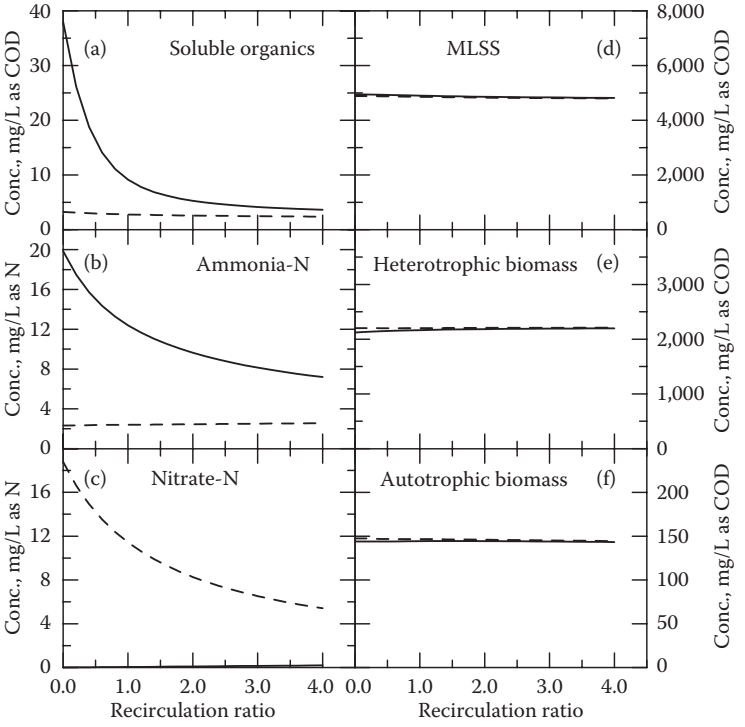

FIGURE 7.31 Effect of the mixed liquor recirculation ratio on the steady-state concentrations of various constituents in each reactor of the MLE system described in Figure 7.29. SRT = 10 days. The solid curves represent the anoxic (first) reactor and the dashed curves the aerobic (second) reactor. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

more anoxic heterotrophic growth, thereby reducing the concentration of soluble organic matter in that reactor. This increased denitrification also acts to diminish the amount of nitrate-N in the system effluent, as shown by the concentration in the second bioreactor. Up to a recirculation ratio of around 2.0, the availability of electron acceptor appears to limit the reactions in the anoxic bioreactor because appreciable soluble organic matter is present, but almost no nitrate-N. Above that point, however, the reactions are limited by both the electron donor and the electron acceptor because the concentrations of both are low. If the recirculation ratio were increased beyond 4.0, reaction in the anoxic bioreactor would become limited by the availability of an electron donor and the concentration of nitrate-N in it would increase. This, in turn, would cause the concentration of nitrate-N in the aerobic bioreactor to reach a limit below which further reduction could not be achieved. That final limit is determined by the relative amounts of organic matter (electron donor) and nitrogen (potential electron acceptor) in the influent. If the concentration of organic matter in the influent were higher than in this example, the lowest possible effluent nitrate-N concentration would be lower, whereas if the wastewater contained less organic matter relative to the amount of nitrogen present, the lowest attainable nitrate-N concentration would be higher. For this wastewater, which is similar to domestic sewage, it can be seen that recirculation ratios above 4.0 have little effect on the concentration of nitrate-N in the effluent. Thus, larger recirculation ratios are seldom used.

The mass utilization rates of nitrate and oxygen in the two bioreactors are shown in Figure 7.32. The nitrate utilization curve reflects the events just described, with the amount of nitrate used at zero recirculation being due to its return to the anoxic bioreactor by biomass recycle. As more recirculation is used, less oxygen is required because more of the electrons associated with the organic matter in the wastewater are being transferred to nitrate-N. As discussed previously, reduction of the amount of nitrate-N discharged is not the only reason a designer might decide to incorporate

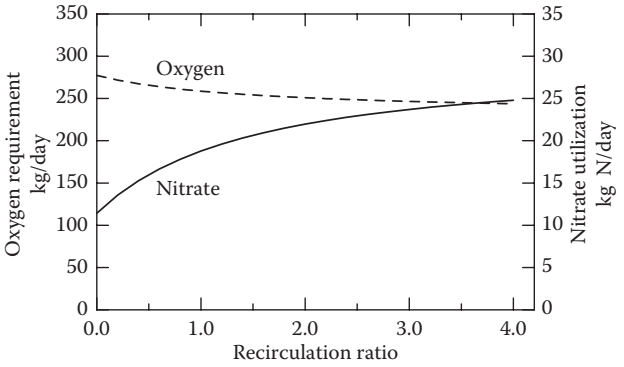

FIGURE 7.32 Effect of the mixed liquor recirculation ratio on the total steady-state oxygen requirement and nitrate utilization rate in the MLE system described in Figure 7.29. SRT = 10 days.

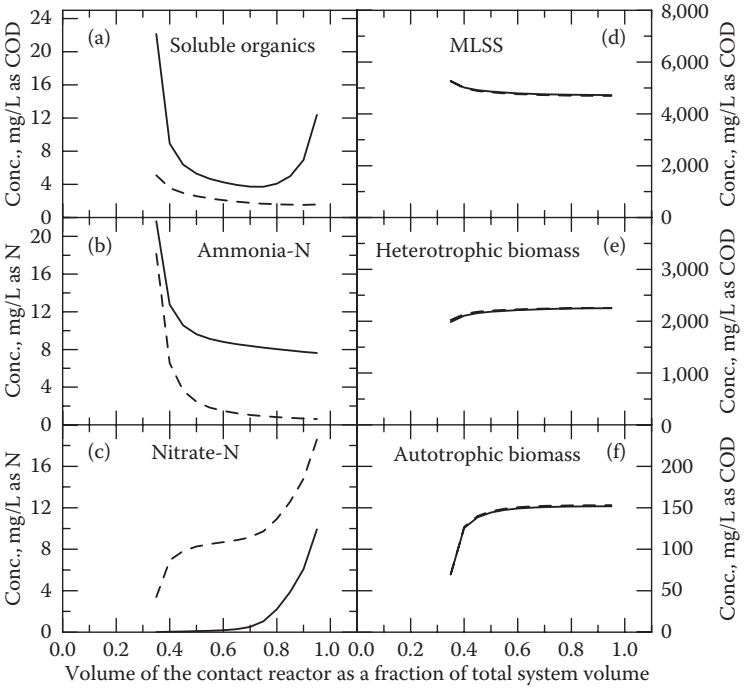

FIGURE 7.33 Effect of the relative volumes of the two reactors on the steady-state concentrations of various constituents in each reactor of the MLE system described in Figure 7.29. The total system volume was constant at 250 m^3. SRT = 10 days. The solid curves represent the anoxic (first) reactor and the dashed curves the aerobic (second) reactor. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

denitrification into a bioreactor system.[6] It may also be done to reduce power costs. In that case, the designer would choose the recirculation ratio that minimized total system power costs, including that for aeration and recirculation pumping.

Figure 7.33 shows the effects of varying the relative volumes of the two bioreactors, with the results plotted as a function of the volume in the aerobic bioreactor. An examination of Figure 7.33f shows that washout of the nitrifying bacteria began as the aerobic fraction of the system was reduced below 0.4. Since the system SRT is 10 days, this corresponds to an aerobic SRT of 4 days. Reexamination of Figure 7.29 reveals that aerobic SRTs in excess of 4 days are required to have

maximal nitrification with the kinetic parameters used in these simulations. Since stable nitrification is a prerequisite for successful operation of the MLE system (or any other system in which denitrification is to be achieved at the expense of nitrate formed in it), the minimum fraction of the bioreactor volume allocated to the aerobic zone is determined by the minimum allowable aerobic SRT. It will be recalled that ASM No. 1 was not developed for simulating prolonged anaerobic periods. However, once washout of the autotrophic biomass has occurred, no nitrate will be present, which means that bioreactor 1 will be anaerobic rather than anoxic. Because the model is not valid under those conditions and because it was apparent that failure of nitrification was occurring at an aerobic fraction of 0.35, lower aerobic fractions were not investigated.

The rise of ammonia-N in both bioreactors and the drop in nitrate-N in the aerobic bioreactor as the aerobic fraction is reduced reflects the washout of the autotrophic biomass discussed above. Furthermore, the rapid rise in the concentration of soluble organic matter as the aerobic fraction is reduced below 0.4 also reflects this wash out. When the amount of nitrate-N being returned to the first bioreactor is greatly reduced, the reactions in it became severely limited by the availability of an electron acceptor, preventing complete utilization of the organic matter. This limitation is also reflected in the heterotrophic biomass and MLSS concentrations. The former drops due to curtailed growth in the anoxic bioreactor whereas the latter rises due to the accumulation of particulate organic matter.

As the aerobic fraction is increased above 0.7 the nitrate-N and soluble organic matter concentrations in the anoxic bioreactor begin to rise because the HRT is insufficient to allow complete reaction. This effect becomes especially severe at aerobic fractions above 0.8, and both soluble organic matter and nitrate-N leave the anoxic bioreactor unreacted.

Aerobic fractions between 0.5 and 0.7 produce effluents with about the same total nitrogen concentrations (ammonia-N plus nitrate-N), although a greater fraction of that nitrogen is in the form of nitrate-N when the aerobic fraction is larger. This suggests that designers of MLE systems have some latitude in the distribution of the system volume between the anoxic and aerobic zones. Furthermore, that latitude increases as the system SRT is increased, suggesting that systems with longer SRTs have greater operational flexibility.

Figure 7.34 shows the effects of the aerobic fraction of the system volume on the nitrate and oxygen utilization rates. The curves are consistent with previous explanations. Nitrate utilization is low at small aerobic fractions because little is being produced and is low at large aerobic fractions because insufficient time is available in the anoxic bioreactor for its utilization. Likewise, oxygen utilization is low at small aerobic fractions because little nitrification is occurring, increases as more nitrification is achieved, and finally increases again as denitrification is curtailed, requiring

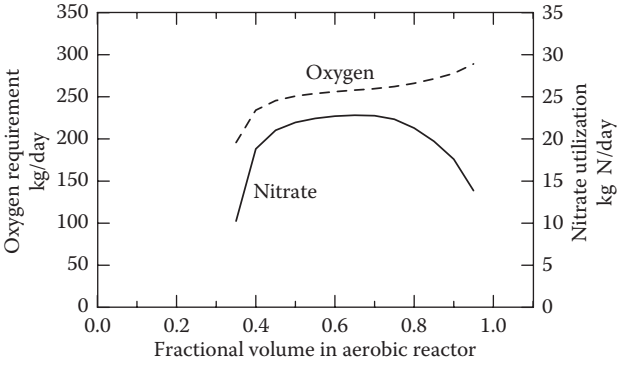

FIGURE 7.34 Effect of the relative volumes of the two reactors on the total steady-state oxygen requirement and nitrate utilization rate in the MLE system described in Figure 7.29. The total system volume was constant at 250 m^3. SRT = 10 days.

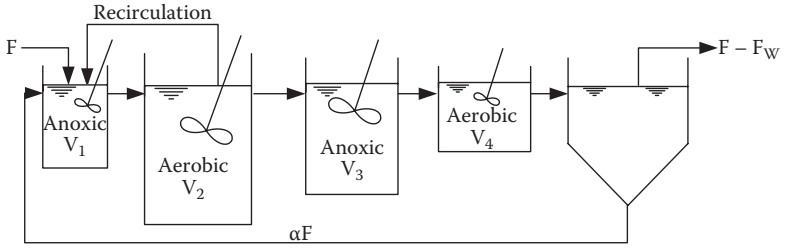

FIGURE 7.35 Schematic diagram of four CSTRs in series with all influent and all biomass recycle to the first reactor, in which the first and third reactors are anoxic and the second and fourth are aerobic. The first reactor receives mixed liquor recirculation flow from the second. Although not shown, solids wastage is directly from all reactors. This configuration simulates the four-stage Bardenpho process.

more of the electrons associated with organic matter in the influent to be transferred to oxygen as the terminal acceptor.

It was pointed out at the beginning of this section that system HRT has little impact on performance.[43] This point should be emphasized. The curves in Figures 7.33 and 7.34 change little when the system HRT is changed for a constant SRT. The important variable is the fraction of the HRT that is anoxic or aerobic, not the actual residence time in each zone.

Finally, it should be noted that ASM No. 1, the model used to perform these simulations, does not have the ability to predict the release of nitrous oxide, an important greenhouse gas, during denitrification. However, it has been noted experimentally that nitrous oxide can be released when the electron donor is limiting in an anoxic bioreactor.[17,20] Figures 7.31 and 7.33 show that this situation can develop when the recirculation ratio is too high or when the volume of the anoxic zone is too small. This suggests that such situations should be avoided. Although not used herein, models are now available that allow the production of nitrous oxide during denitrification to be considered.[15]

7.6 FOUR-STAGE BARDENPHO PROCESS

7.6.1 Description

As discussed in Section 7.1.1, a disadvantage of the MLE reactor configuration is that the effluent will always contain appreciable quantities of nitrate-N because nitrification occurs in the last bioreactor and the mixed liquor recirculation is from that bioreactor. The four-stage Bardenpho[3] process overcomes this by adding an anoxic bioreactor after the aerobic one in which denitrification can occur by biomass decay and the utilization of slowly biodegradable substrate. In addition, to prevent biomass settling problems associated with denitrification in the final settler, a small aerobic bioreactor is usually used as the final zone.[3] This reactor configuration is illustrated in Figure 7.35. Selection of the best combinations of bioreactor sizes is a complex question that requires the use of optimization techniques[43] and it will be discussed in Chapter 12. For the purpose of this discussion, however, the total bioreactor volume was kept at 250 m^3, the value used in all of the simulations in this chapter, and the aerobic fraction was kept at 50%, the value used in Figures 7.29–7.32. The sizes of bioreactors 1 through 4 were selected as 50, 100, 75, and 25 m^3, respectively. The mixed liquor recirculation ratio from bioreactor 2 to bioreactor 1 was maintained at 2.0, the value used in Figures 7.29, 7.30, 7.33, and 7.34.

7.6.2 Effect of SRT on Steady-State Performance

The effect of SRT on the performance of the four-stage Bardenpho system is shown by the solid curves in Figures 7.36 and 7.37. For comparison, the performance of the MLE system from

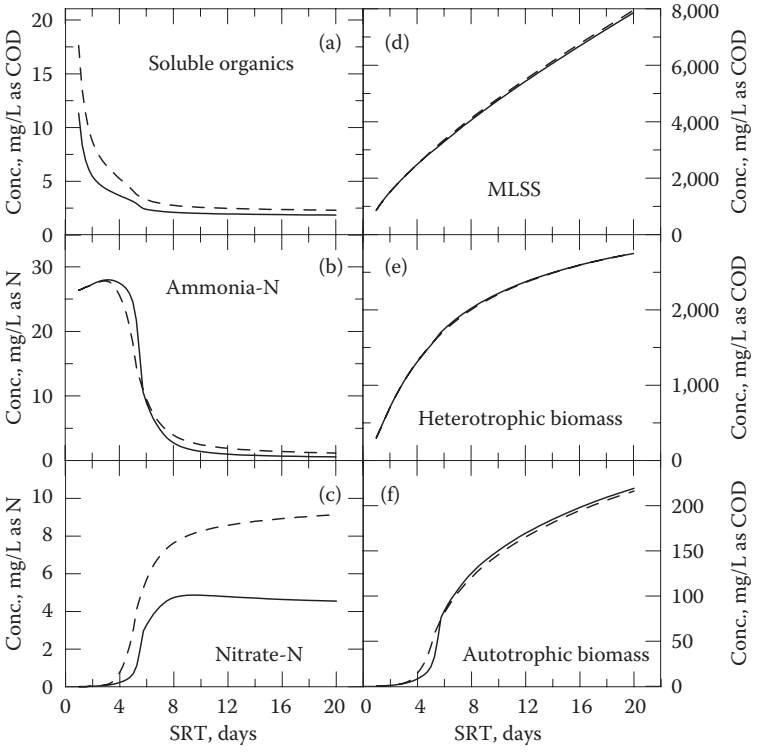

FIGURE 7.36 Effect of SRT on the steady-state concentrations of various constituents in the last reactor of the four-stage Bardenpho system depicted in Figure 7.35. For comparison, the dashed curves represent the performance of the MLE system described in Figure 7.29. Influent flow = 1000 m^3/day, influent concentrations are given in Table 6.6, biomass recycle flow = 500 m^3/day, mixed liquor recirculation flow = 2000 m^3/day, reactor volumes: $V_1 = 50$ m^3, $V_2 = 100$ m^3, $V_3 = 75$ m^3, and $V_4 = 25$ m^3. Parameters are listed in Table 6.3. The DO concentration was zero in the first and third (anoxic) reactors and 2.0 mg/L in the second and fourth (aerobic) reactors. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T} = 1.20$ mg COD/mg TSS.

Figures 7.29 and 7.30 is shown by the dashed curves. Comparison of the curves shows that both systems contain similar quantities of biomass, but that the Bardenpho system achieves an effluent with less ammonia-N and less nitrate-N than the MLE system by achieving more denitrification. This is done even though the aerobic and anoxic fractions are the same as in the MLE system. Even lower effluent concentrations could be attained by proper selection of the aerobic fraction, volume distribution among the various bioreactors, and recirculation ratio. However, it was not the intent here to minimize effluent nitrogen concentrations; rather, the point was to show the effect of adding additional anoxic and aerobic bioreactors.

A significant thing to note in Figure 7.36 is that the growth pattern of the autotrophic bacteria is different in the two bioreactor systems, even though the aerobic SRTs are the same. This emphasizes the point made earlier that although the concept of aerobic SRT is important to understanding the fate of nitrifying bacteria in systems containing anoxic zones, it is not the sole factor influencing their growth. System configuration is also important. This suggests that pilot scale studies coupled with system simulation are required to arrive at sound designs for such complex systems. In Chapter 12 we will see how the results from such studies have been combined with full-scale plant experience to allow development of design guidelines for biological nutrient removal systems.

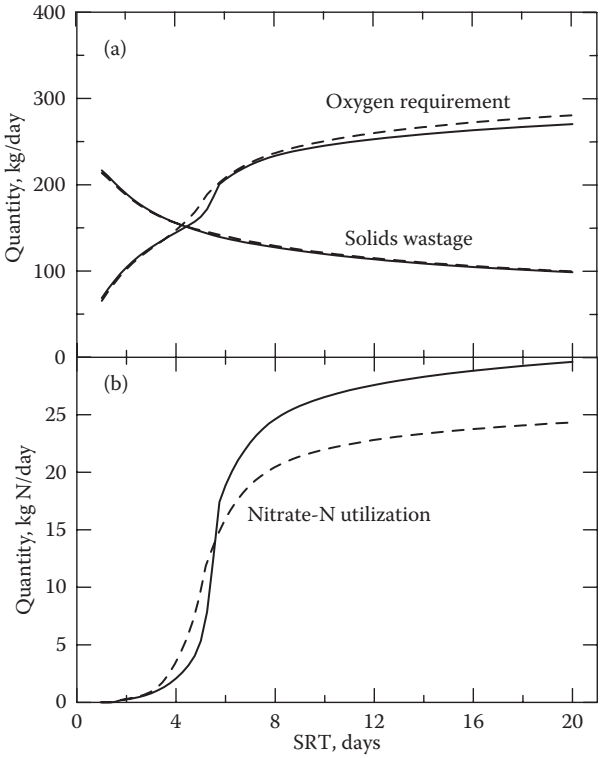

FIGURE 7.37 Effect of SRT on the total steady-state oxygen requirement, nitrate utilization rate, and solids wastage rate for the four-stage Bardenpho system depicted in Figure 7.35 operating under the conditions listed in Figure 7.36. For comparison, the dashed curves represent the performance of the MLE system described in Figure 7.29. To express the solids wastage rate in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

7.7 BIOLOGICAL PHOSPHORUS REMOVAL PROCESS

7.7.1 Description

As discussed in Section 2.4.6, certain bacteria, known collectively as PAOs or phosphate accumulating organisms, have the interesting characteristic of concentrating phosphate in poly-phosphate (Poly-P) granules when they are cycled between anaerobic and aerobic/anoxic conditions, thereby providing a means of biological phosphorus removal (BPR) from wastewater. The Poly-P acts as an energy reserve that allows the bacteria to rapidly take up acetate under anaerobic conditions, storing it as the polymer poly-β-hydroxyalkanoate (PHA). At the same time, glycogen is used to provide precursors and reducing power that are needed for PHA synthesis. Soluble phosphate is released in the process. The PHAs, in turn, provide energy for growth under aerobic/anoxic conditions. They also allow soluble phosphate to be taken up and stored as Poly-P and for the glycogen reserve to be replenished. The difference in energetics between aerobic/anoxic and anaerobic metabolism is such that more phosphate can be taken up than was released, providing a mechanism for concentrating phosphate within the biomass, allowing it to be removed via solids wastage.

The history of the development of biological phosphorus removal processes is one of the most fascinating ones in environmental engineering, beginning with observations of unexplained phosphorus removal in full-scale CAS systems, arguments over the reasons for that removal, simultaneous development of processes by various groups, conflicts over patent claims and infringements, and the use of molecular biology tools to characterize the ecology of PAOs, thereby improving our ability to predict the performance and design of BPR processes. Unfortunately, space does

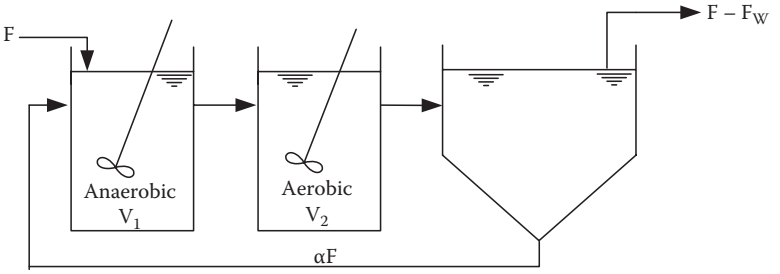

FIGURE 7.38 Schematic diagram of two CSTRs in series with all influent and all biomass recycle to the first reactor, which is anaerobic. Although not shown, solids wastage is from the second reactor, which is aerobic. The configuration simulates the Phoredox process.

not permit a review of that history here, but the reader is urged to consult Randall,[30] Stensel,[36] and Oehmen et al.[26] for part of the story.

The simplest process flow sheet for biological phosphorus removal incorporates two bioreactors in series, with the first being anaerobic and the second aerobic, as shown in Figure 7.38. This process was first presented in the open literature by Barnard[4] who termed it the Phoredox process to indicate that phosphorus removal will occur when a sufficiently low redox potential is achieved through use of an anaerobic zone. He reasoned that the anaerobic zone should be placed first in the process train to take advantage of the electrons available in the raw wastewater, just as is done in the MLE and four-stage Bardenpho processes of denitrification.[4] This same configuration was patented by Air Products and Chemicals, Inc. of Allentown, PA under the trademark Anaerobic/Oxic, or A/O process.[35] The major difference between the Phoredox and A/O processes is that in the latter the anaerobic and aerobic zones are divided into a number of equally sized completely mixed compartments.[36] For use in this section, we will consider only a single compartment for each. However, a large number of process flow sheets are available for biological phosphorus removal, both alone and in concert with nitrogen removal. They will be discussed in Chapter 12.

The simple process flow sheet shown in Figure 7.38 provides an opportunity to observe some very interesting interactions among the various types of bacteria in wastewater treatment systems. In the MLE and four-stage Bardenpho processes we observed interactions between heterotrophs and autotrophs. Introduction of the anaerobic zone allows the specialized PAOs to interact with both of those groups. (Because PAOs are heterotrophic, the term ordinary heterotrophic organisms [OHOs] will be used to distinguish the non-PAO heterotrophs from the PAOs.) To introduce those interactions and the effects that they have on design of BPR systems, simulations were performed in GPS-X (Table 6.4) using ASM No. 2d. As discussed in Section 6.1.4, ASM No. 2d is very complex because it seeks to incorporate a number of complicated processes that are not yet fully understood. Nevertheless, it is sufficiently conceptually accurate to illustrate several important points, and it is for that purpose that it is used herein. The limitations inherent in the assumptions of the model will be pointed out as the simulated performance of a Phoredox system is discussed.

The system chosen to represent the Phoredox process, shown in Figure 7.38, contains two bioreactors in series with the first being anaerobic and the second aerobic. For this simulation, the system has a total volume of 250 m^3, receives 1000 m^3/d of wastewater flow, and has a biomass recycle rate of 500 m^3/d from the clarifier to the first bioreactor. In this case, however, 20% of the total system volume is allocated to the first bioreactor. The first bioreactor is assumed to receive no dissolved oxygen whereas the second receives sufficient oxygen to maintain the DO concentration at 2.0 mg/L. The characteristics of the wastewater entering the system are given in Table 7.1. The components are mostly the same as those in Table 6.6, but others are included to be consistent with ASM No. 2d. For instance, the readily biodegradable substrate has been divided into two components, volatile fatty acids (primarily acetate) and readily fermentable substrate. Acetate is found in many wastewaters, particularly if the sewers are septic, and plays a major role in the metabolism of the PAOs, as

TABLE 7.1
Wastewater Characteristics Used to Simulate the Performance of the Phoredox Process

Component	Concentration
Inert particulate organic matter	35.0 mg/L as COD
Slowly biodegradable substrate	150.0 mg/L as COD
Readily (fermentable) biodegradable substrate	75.0 mg/L as COD
Volatile fatty acids (acetate)	40.0 mg/L as COD
Oxygen	0.0 mg/L as O_2
Soluble nitrate nitrogen	0.0 mg/L as N
Soluble ammonia nitrogen	25.0 mg/L as N
Soluble biodegradable organic nitrogen	6.5 mg/L as N
Particulate biodegradable organic nitrogen	8.5 mg/L as N
Soluble phosphate phosphorus	10.0 mg/L as P
Soluble biodegradable organic phosphorus	0.75 mg/L as P
Particulate biodegradable organic phosphorus	1.85 mg/L as P
Alkalinity	5.0 mmol/L

Source: Henze, M., Gujer, W., Takashi, M., Tomonori, M., Wentzel, M. C., Marais, G. v. R., and Van Loosdrecht, M. C. M., Activated Sludge Model No. 2d, *IWA Scientific and Technical Reports*, No. 9, International Water Association, London, 2000.

discussed above. Furthermore, phosphorus is included and is present in three different forms that parallel the forms used to represent nitrogen. The default parameter values for ASM No. 2d were used,[14] except as follows. The rate coefficient for PHA storage ($\hat{q}_{PHA}$) was set to 6 day^{-1};[2] the decay coefficients (lysis:regrowth approach) for PAOs ($b_{L,PAO}$), polyphosphate ($b_{L,PP}$), and PHA ($b_{L,PHA}$) were set to 0.1 day^{-1}.[41] Also, values for the autotrophic coefficients $\hat{\mu}_A$ and $b_{L,A}$ were used as given in Table 6.3, rather than using the default values in ASM No. 2d. Finally, abiotic phosphate precipitation was "turned off" to demonstrate the effect of biological phosphorus removal alone.

7.7.2 Effect of SRT on Steady-State Performance

The effects of SRT on the concentrations of various constituents in the second bioreactor (and the effluent) of the Phoredox system are shown by the solid curves in Figure 7.39. For comparison, the concentrations in a single aerobic CSTR with a volume of 250 m^3 receiving the same wastewater are shown as the dashed curves.

The responses of the soluble phosphate and the PAOs are shown in panels a and d, respectively, of Figure 7.39. No PAOs grow in the single aerobic CSTR because the proper environment is not provided, and thus the only phosphorus removal is that associated with its role as a macronutrient for biomass growth. Examination of Figure 7.39a reveals that the soluble phosphate concentration in the single aerobic CSTR increases as the SRT is increased. There are two reasons for this. First, the observed yield of biomass decreases as the SRT is increased, thereby decreasing the amount of nutrients required. Second, soluble phosphate is released from slowly biodegradable substrate as it is hydrolyzed. Hydrolysis is greater at longer SRTs, allowing more phosphate to be released. The net effect of these events is to make the amount of phosphate released exceed the amount incorporated into biomass. Thus, the concentration increases. Furthermore, at short SRTs, no phosphorus is removed in the Phoredox system because the SRT is below the minimum required for growth of PAOs. The minimum SRT relates only to the aerobic SRT because growth of PAOs occurs only under aerobic/anoxic conditions. The presence of the anaerobic zone is necessary for their growth,

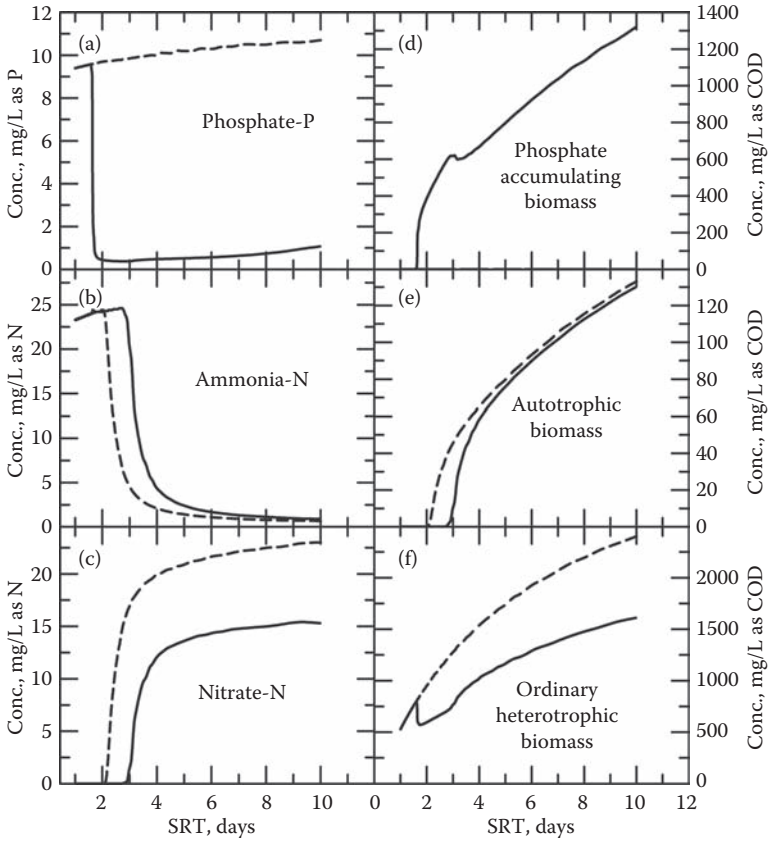

FIGURE 7.39 Effect of SRT on the steady-state concentrations of various constituents in the aerobic (last) reactor of the Phoredox system depicted in Figure 7.38. For comparison, the dashed curves represent the performance of a single aerobic CSTR with a volume of 250 m^3. Influent flow = 1000 m^3/day, influent concentrations are given in Table 7.1, biomass recycle flow = 500 m^3/day, volume of the anaerobic (first) reactor = 50 m^3, and volume of the aerobic (second) reactor = 200 m^3. The DO concentration is zero in the anaerobic reactor and 2.0 mg/L in the aerobic reactor. Parameters are from ASM No. 2d, with exceptions as described in the text. To express the biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

however, because they grow at the expense of stored PHAs, which are only formed under anaerobic conditions as acetate is taken up and stored at the expense of Poly-P. Once the minimum SRT is exceeded, a significant population of PAOs is established in the Phoredox system and phosphate uptake by them in the aerobic bioreactor is able to reduce the phosphate to very low levels.

An examination of Figure 7.39f reveals that there is a reduction in the amount of OHO biomass relative to the single aerobic CSTR over the SRT range where good phosphorus removal is occurring. This is because of competition between the PAOs and OHOs for substrate. In the anaerobic bioreactor of the Phoredox system, the PAOs are able to store acetate. The OHOs, on the other hand, cannot store or use acetate in the anaerobic zone; nor can they grow as long as oxygen or nitrate-N is unavailable as an electron acceptor. They can only produce acetate by fermentation of readily fermentable substrates. Consequently, because the PAOs in the anaerobic zone store the acetate formed by the OHOs, the amount of substrate entering the aerobic zone, where OHO growth occurs, is reduced, thereby reducing the quantity of OHOs that can be formed. However, because the PAOs contain the acetate that they stored in the anaerobic zone as PHA, they are able to grow.

Examination of Figure 7.39a reveals that excellent phosphorus removal occurs as long as the SRT is above 1.8 days. The minimum phosphate concentration occurs at 2.8 days and then slowly

increases as the SRT is increased. This is because nitrification starts at an SRT of around 2.8 days (Figure 7.39b and c). Beyond this point, autotrophic biomass can grow and convert ammonia-N to nitrate-N. The nitrate-N, in turn, is returned to the anaerobic zone via the biomass recycle, providing an electron acceptor and preventing the anaerobic zone from being truly anaerobic. The impact of the presence of nitrate in the anaerobic zone is manifested in two ways.

The first impact of the nitrate is straightforward. It provides an electron acceptor that can be used by the OHOs, allowing them to metabolize the acetate and readily fermentable substrate entering the anaerobic zone. In other words, the OHOs can compete effectively with the PAOs for substrate and if the rate of nitrate entry is sufficient to allow all acetate and readily fermentable substrate to be metabolized by the OHOs, the PAOs will lose out. On the other hand, if the rate of nitrate entry into the anaerobic tank is less than that required to metabolize all of the organic substrate via denitrification, the OHOs will carry out fermentation reactions on the balance, leading to the production of acetate, but in a lesser amount. Because the OHOs can't metabolize that acetate (because of insufficient electron acceptor) the PAOs will take it up and store it, thereby making it available for their growth in the aerobic zone. However, because less acetate is made available to the PAOs, there will be fewer of them and they will contain less PHA, thereby reducing the amount of phosphate that can be taken up in the aerobic zone. The severity of that effect will depend on the relative amounts of organic substrates, ammonia-N, and phosphorus in the wastewater.

The second impact of the presence of nitrate in the anaerobic zone is more complicated. We saw in Section 2.4.6 that some PAOs can use nitrate as an electron acceptor. Because the nitrate acts as an alternative to oxygen, if these denitrifying PAOs are cycled between anaerobic and anoxic conditions, they behave just like PAOs that are cycled between anaerobic and aerobic conditions, storing acetate in the anaerobic zone and metabolizing it in the anoxic zone, taking up phosphate in the process.[19,21] On the other hand, if they are cycled between an anoxic zone (with an abundant supply of nitrate) and an aerobic zone, they do not take up and store acetate; rather, they metabolize it in both zones, just like OHOs. Thus, although they would be present in the biomass, they would not remove any phosphate because they would not contain PHAs upon entering the aerobic zone. This is how the denitrifying PAOs behave when nitrate enters the anaerobic zone via biomass recycle from the clarifier when nitrification is occurring in the aerobic zone. Because they can use nitrate as their electron acceptor, they compete directly with the OHOs for substrate, thereby reducing the contribution of OHOs to the biomass. Although this allows the denitrifying PAOs to grow in the system, it diminishes their ability to remove phosphorus because they store less acetate as PHA. The impact of this effect is also a function of the amount of nitrate entering the anaerobic zone. More nitrate diminishes the PHA in the PAO community, thereby decreasing the phosphate removal.

Both of these impacts can be seen in Figure 7.39. Figure 7.39d illustrates that the rate of increase in PAOs with increasing SRT decreases relative to what it had been before nitrification started. Figure 7.39f illustrates an increase in the presence of OHOs once nitrification starts. Finally, Figure 7.39a illustrates that as more nitrate is generated through nitrification of decay products formed with increasing SRT, even more phosphate remains in the effluent. Again, the magnitude of the impact on phosphorus removal depends on the relative amounts of organic substrate, ammonia-N, and phosphorus in the wastewater.

Because the entry of nitrate into the anaerobic zone has a deleterious effect on phosphorus removal, one would expect that larger biomass recycle rates would cause a deterioration in phosphorus removal and this is exactly what happens, as shown in Figure 7.40. As can be seen, the phosphate concentrations leaving the system increase as the mass of nitrate-N entering the anaerobic zone via the recycle flow increases. In practice, recycle rates beyond 100% of the influent flow rate are seldom used because of negative impacts of high recycle rates on final clarifier design and operation. Nevertheless, even recycle ratios of 100% have severe impacts on phosphorus removal. At the other extreme, even though very low recycle rates benefit phosphorus removal when the system SRT is high enough to allow nitrification to occur, they are not practical because they, too, have negative

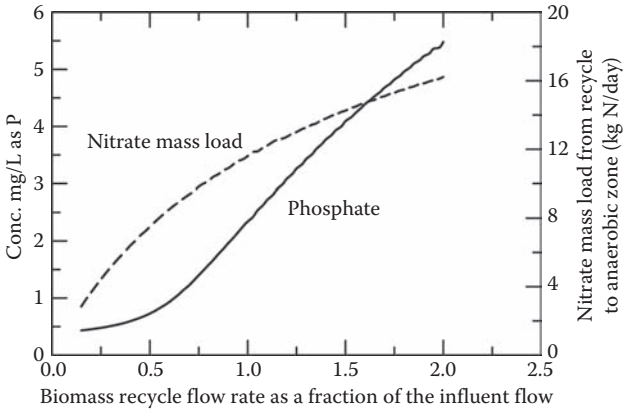

FIGURE 7.40 The impact of increasing the biomass recycle flow rate on the mass of nitrate returned to the first (anaerobic) reactor and the concentration of phosphorus in the second (aerobic) reactor of the Phoredox system depicted in Figure 7.38. Conditions are the same as those used in Figure 7.39 except that the SRT was held constant at eight days.

effects on final clarifiers. As a consequence of these effects, Phoredox and A/O processes tend to be operated with short SRTs and recycle flows around 50%.

Because it is often desirable to achieve carbon oxidation, nitrification, denitrification, and phosphorus removal all in a single system, several process flow sheets have been devised to overcome the effects associated with the entry of nitrate into the anaerobic zone. These systems minimize or eliminate the entry of nitrate into the anaerobic zone and can achieve both excellent phosphorus and nitrogen removal. They will be discussed in Chapter 12.

In summary, the important point to draw from Figures 7.39 and 7.40 is that there is a limited range of SRTs and recycle flow rates over which a simple Phoredox (or A/O) system will work properly. If the SRT is too short, the PAOs will wash out. If it is too long, nitrifying bacteria will grow, providing an inorganic electron acceptor to the anaerobic zone. This has two effects. First, some of the PAOs will use the nitrate as an electron acceptor, causing them to metabolize the acetate, rather than storing it as PHA. Second, the OHOs will compete effectively with the PAOs for substrate, thereby reducing PHA storage. Because less PHA is available to the PAOs in the aerobic bioreactor, less phosphate will be taken up. Furthermore, recycle flow rates must be kept low (but not too low) to avoid returning too much nitrate to the anaerobic zone.

7.7.3 Effects of System Configuration

The role of the anaerobic bioreactor in the Phoredox process is twofold. First and foremost, it provides the selective advantage that allows PAOs to grow in the system. It is there that they take up acetate, forming the PHAs that will serve as their energy source for growth in the aerobic bioreactor. Second, it provides a regime wherein fermentation may occur, providing acetate in excess of that available in the influent. As important as the anaerobic tank is, the entire system cannot be anaerobic because aerobic/anoxic conditions are required for growth of both PAOs and OHOs. Thus, there is an optimal balance between the sizes of the two tanks. In order to illustrate that balance, simulations were performed in which the total system volume was held constant while the relative sizes of the two tanks were varied. The system SRT was held constant at four days. The results of the simulations are shown in Figures 7.41 and 7.42. The former shows the responses of the soluble constituents while the latter presents the particulate ones. The solid curves represent the anaerobic (first) bioreactor and the dashed curves represent the aerobic (second) bioreactor. The dashed curves, therefore, also represent the system effluent with respect to the soluble constituents.

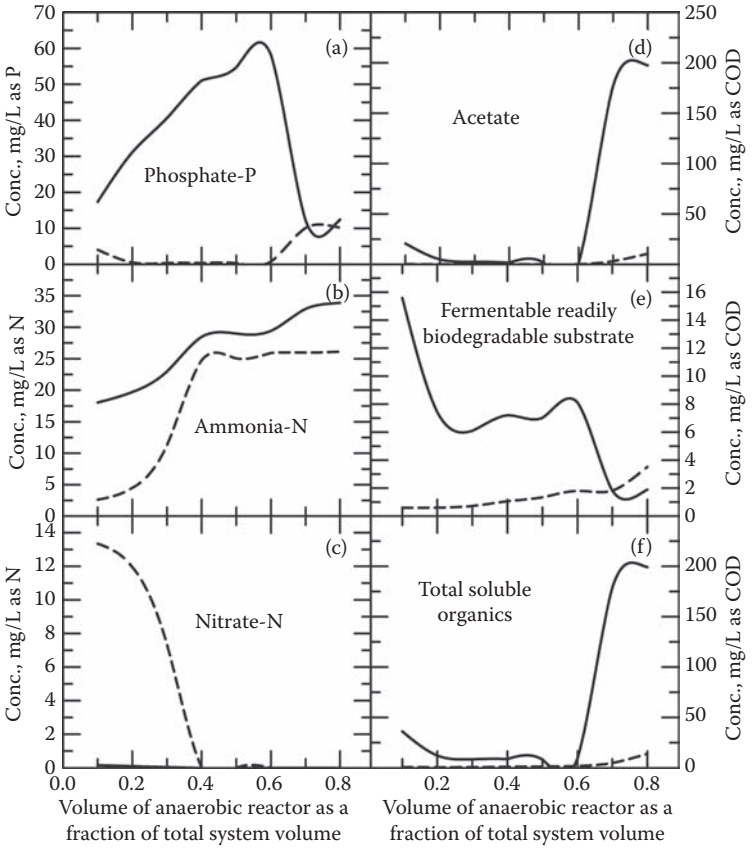

FIGURE 7.41 Effect of the relative volumes of the two reactors on the steady-state concentrations of various soluble constituents in each reactor of the Phoredox system depicted in Figure 7.38 and described in Figure 7.39. The total system volume was constant at 250 m^3. SRT = 4 days. The solid curves represent the anaerobic (first) reactor and the dashed curves the aerobic (second) reactor.

An examination of Figure 7.41a reveals that the best phosphorus removal occurs when the volume of the anaerobic tank is between 22 and 58% of the total system volume. It should be emphasized that this range is unique to this SRT, wastewater characteristics, and so on. It will be different for other situations. The important point is that there is indeed an optimal combination that maximizes the concentration of PAOs (Figure 7.42a), thereby allowing maximum phosphate-P removal. Panels a and d of Figure 7.41 also reveal that over that range, the concentration of phosphate-P reaches a maximum in the anaerobic tank while the concentration of acetate reaches a minimum. This is because of the mechanism of phosphate removal. Recall that in the anaerobic zone the PAOs hydrolyze stored Poly-P to gain energy for the uptake and storage of acetate as PHAs, releasing orthophosphate in the process (see Figure 7.41a and Section 2.4.6). Over the optimal range, maximum utilization has been made of the available acetate in the anaerobic tank, leading to the maximum release of soluble phosphate. That phosphate is then taken up again in the aerobic bioreactor as the PAOs grow by using the PHAs as substrate and store Poly-P.

The maximum acceptable anaerobic volume fraction is controlled by the minimum aerobic SRT for growth of the PAOs in the aerobic tank. Because the total volume is fixed, as the anaerobic volume is increased, the aerobic volume is decreased. Once the aerobic SRT is below the minimum required for growth of the PAOs, they wash out and the system fails from the perspective of phosphorus removal. This can be seen clearly in Figures 7.41a and 7.42a. The population of OHOs increases (Figure 7.42b) because they no longer have to compete with the PAOs for acetate.

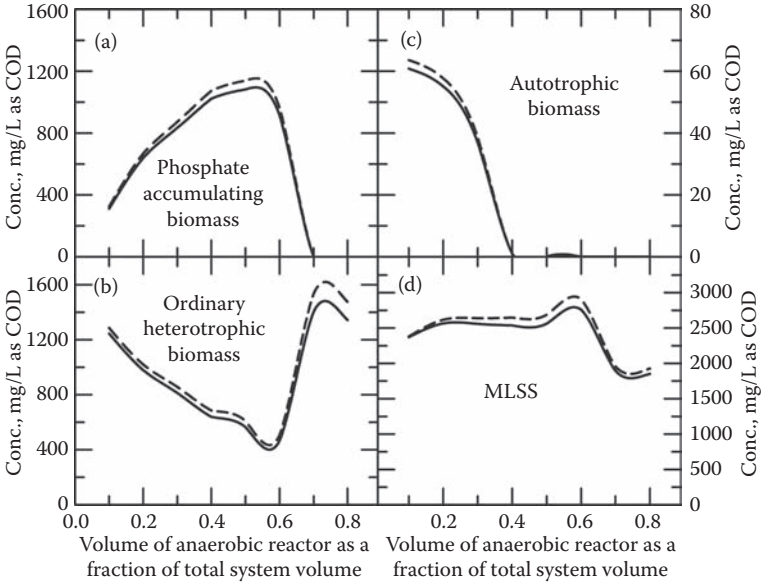

FIGURE 7.42 Effect of the relative volumes of the two reactors on the steady-state concentrations of various particulate constituents in each reactor of the Phoredox system depicted in Figure 7.38 and described in Figure 7.39. The total system volume was constant at 250 m^3. SRT = 4 days. The solid curves represent the anaerobic (first) reactor and the dashed curves the aerobic (second) reactor. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

Ultimately, as the anaerobic tank gets even larger, the entire system begins to act like a totally anaerobic system. The model was not intended for simulating such systems, so anaerobic fractions greater than 0.8 are not shown.

The explanation of the minimum acceptable anaerobic volume fraction is more complicated. If the anaerobic tank is very small, there is insufficient time for acetate production by fermentation and its uptake by the PAOs. Thus, the PAOs do not grow well in the system (Figure 7.42a). In addition, when the anaerobic tank is small, making the aerobic tank large, the aerobic SRT is sufficiently long to allow nitrifying bacteria to grow, thereby providing nitrate as an electron acceptor in the anaerobic zone through the biomass recycle. As a consequence, both denitrifying PAOs and OHOs metabolize the incoming substrate, thereby decreasing the formation of PHA by the PAOs, as described in Section 7.7.2. For the conditions used in the simulations, competition with OHOs and the coexistence of PAO groups exhibiting the two types of metabolism occur in anaerobic tanks that constitute less than 40% of the total system volume, thereby reducing the amount of fermentation that can occur, as well as the release of phosphate and storage of PHA. However, as the anaerobic tank is made larger, there is more opportunity for fermentation, with a gain in PAOs at the expense of OHOs (Figure 7.42a and b). Furthermore, as the size of the anaerobic tank is increased, the aerobic SRT is decreased, thereby reducing the population of autotrophic bacteria (Figure 7.42c) and the concentration of nitrate being returned to the first bioreactor (Figure 7.41c). This allows for better growth of the PAOs. Ultimately, the point is reached at which the aerobic SRT is smaller than the minimum SRT for the autotrophs and they wash out (Figure 7.42c), allowing the first tank to operate in a truly anaerobic mode, which maximizes phosphorus removal.

In summary, the important point to gain from these simulations is that there is an optimal split between the volumes of the anaerobic and aerobic bioreactors of the Phoredox (or A/O) system. Optimal phosphorus removal occurs when the aerobic SRT is small enough to exclude growth of autotrophic nitrifying bacteria, yet large enough to allow the PAOs to grow, and the time in the anaerobic tank is sufficiently large to allow efficient fermentation and uptake of the resulting acetate.

7.7.4 Factors Affecting the Competition between Phosphate Accumulating and Glycogen Accumulating Organisms

In Section 2.4.6, the metabolisms of PAOs and glycogen accumulating organisms (GAOs) were introduced, and their relative importance in BPR was discussed. Glycogen accumulating organisms can compete with PAOs for acetate in the anaerobic zone of a BPR system, thereby reducing the efficiency of PAOs in removing phosphate in the aerobic zone. As will be discussed in Chapter 12, very low effluent phosphate concentrations can be achieved only under phosphate-limiting conditions, which means that acetate must be provided in excess of that required to remove the phosphate. Glycogen accumulating organisms will grow on the excess acetate provided and, as a result, will always be present with PAOs in even the most effective BPR systems.[33] The goal in designing and operating a BPR system, therefore, is to minimize the degree to which GAOs proliferate in the biomass. This need to balance the competition between PAOs and GAOs has led to efforts to develop models that include GAOs. These models are early in their development and, therefore, are not presented here. However, it is important to discuss the current state of knowledge regarding PAO-GAO competition.

Four factors have an important influence on the growth competition between PAOs and GAOs: the type of carbon source, the phosphate to COD (P:COD) ratio in the influent, the pH, and the temperature. Significant progress has been made in determining how to manage these factors to bring operational stability to BPR systems. The influent P:COD ratio affects PAO-GAO competition by affecting polyphosphate stores in PAOs.[33] Under extended periods with low P:COD conditions (<0.02 g PO_4^-–P/g COD), PAOs cannot replenish their polyphosphate reserves, which adversely impacts their ability to take up acetate and other volatile fatty acids (VFAs). Under this condition, GAOs can establish a growth advantage with the unused VFAs and proliferate. Both acetic and propionic acids are prevalent among the VFAs formed during the fermentation of wastewater. Although we typically only consider acetate in modeling these systems, PAOs appear to use acetate and propionate equally well as a carbon and energy source. The situation for GAOs is more complex, with some preferring acetate and others using both acetate and propionate. Oehman et al.[26] provide a summary of the substrates upon which PAOs and GAOs can grow and the reader is referred there for details. Collectively, these results suggest that using substrate alone to control the growth of PAOs over GAOs is not sufficient to achieve stable BPR performance. Consequently, other factors need to be considered that can establish a preferential growth environment for PAOs over GAOs. One is pH. Phosphate accumulating organisms dominate in systems with a pH higher than about 7.[11,27,32,44] This is attributed to the larger energy investment required to take up acetate in the face of a larger pH differential between the inside and outside of the cell. Phosphate accumulating organisms meet the greater energy demand by degrading more polyphosphate, thereby releasing more phosphate in the process. Despite this, the rates of acetate uptake, glycogen hydrolysis, and PHA formation in PAOs are not affected over the pH range from 6.5 to 8.0.[10] In contrast, the acetate uptake rate decreases in GAOs with increasing pH,[9] possibly because they can only use glycogen to offset the higher energy demand for acetate transport. Finally, BPR performance tends to deteriorate during warmer periods and improve at lower temperatures,[7,42] suggesting that PAOs and GAOs have distinct growth-temperature curves that can be used to enhance selection of the former over the latter. Details about the factors influencing the competition between PAOs and GAOs are summarized by Lopez-Vazquez et al.[23] (see Figure 7.43), although other factors may also play a minor, but relevant, role.[26]

7.8 SEQUENCING BATCH REACTOR

7.8.1 Description

All of the process variations we have considered so far are continuous processes. As a consequence, the environments required to achieve a variety of objectives must be encountered spatially as the

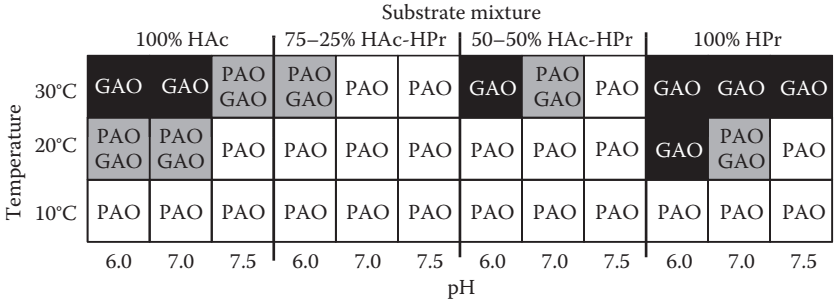

FIGURE 7.43 The population distributions of phosphate accumulating organisms (PAOs) and glycogen accumulating organisms (GAOs) as a function of the nature of the substrate (acetic acid [Hac] to propionic acid [HPr] ratio), pH, and temperature. Grey boxes represent coexistence; black and white boxes represent predominance of the organism listed. (After Lopez-Vazquez, C. M., Oehmen, A., Hooijmans, C. M., Brdjanovic, D., Gijzen, H. J., Yuan, Z., and van Loosdrecht, M. C. M., Modeling the PAO-GAO competition: Effects of carbon source, pH and temperature. *Water Research,* 43:450–62, 2009.)

wastewater and biomass move from tank to tank within the system. Because each tank has a fixed volume, the relative amount of process time spent under each environmental condition is fixed for a given influent flow rate. Alteration of those times requires alteration of the sizes of the various tanks, something that may or may not be achieved easily. It is possible, however, to accomplish the same results in a batch reactor by altering the environment temporally. In this situation, if the relative times devoted to the particular environments are not attaining the desired result, they can be changed easily by reprogramming the controllers that turn the pumps and blowers on and off. This flexibility is the major advantage associated with batch bioreactors.

The term sequencing batch reactor (SBR) stems from the sequence of steps that the reactor goes through as it receives wastewater, treats it, and discharges it, since all steps are accomplished in a single tank. A typical sequence is illustrated in Figure 7.44. The cycle starts with the fill period in which the wastewater enters the bioreactor. The length of the fill period is chosen by the designer and depends upon a variety of factors, including the nature of the facility and the treatment objectives. The main effect of the fill period, however, is to determine the hydraulic characteristics of the bioreactor. If the fill period is short, the process will be characterized by a high instantaneous process loading factor, thereby making the system analogous to a continuous system with a tanks-in-series configuration. In that case, the biomass will be exposed initially to high concentrations of organic matter and other wastewater constituents, but the concentrations will drop over time. Conversely, if the fill period is long, the instantaneous process loading factor will be small and the system will be similar to a completely mixed continuous flow system in its performance. This means that the biomass will experience only low and relatively constant concentrations of the wastewater constituents.

The fill period is followed by the react period in which the biomass is allowed to act upon the wastewater constituents. Actually, reactions (i.e., biomass growth and substrate utilization) also occur during the fill period, so the fill period should really be thought of as "fill plus react," with react continuing after fill has ended. Since a certain total react period will be required to achieve the process objectives, if the fill period is short, the separate react period will be long, whereas if the fill period is long, the separate react period will be short to nonexistent. The two periods are usually separately specified, however, because of the impact that each has on the performance of the system.

The environmental conditions established during the fill and react periods will determine which events occur. For example, if the fill and react periods are aerobic throughout, the events will be limited to carbon oxidation and nitrification. Consequently, performance of the SBR will lie somewhere between that of conventional and completely mixed activated sludge, depending on the length of the fill period. The elimination of aeration while maintaining mixing will allow denitrification to occur if

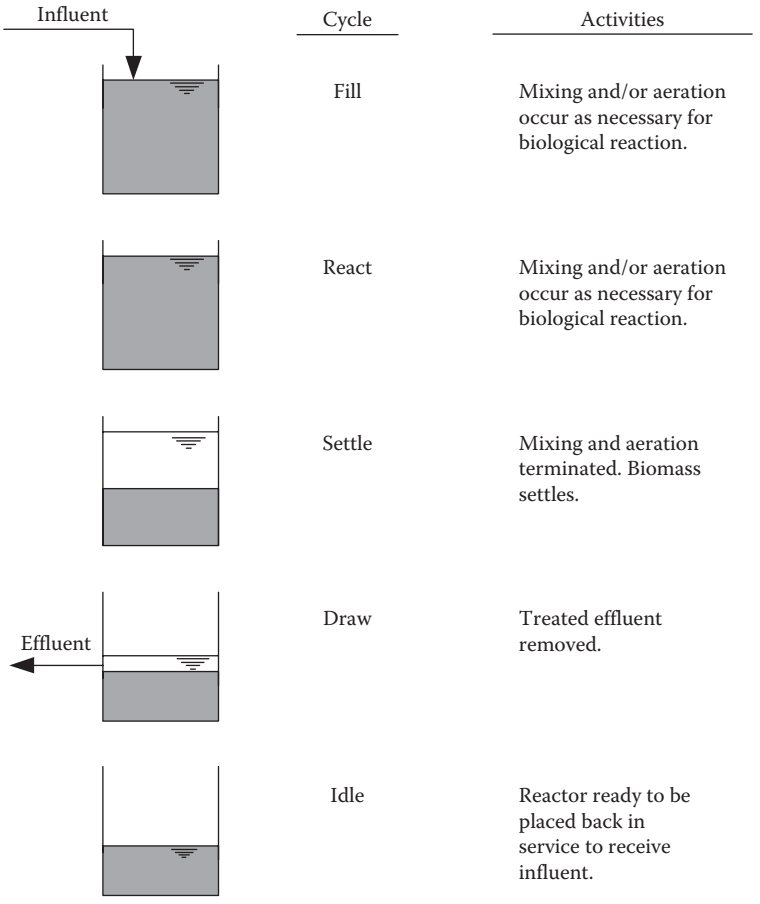

FIGURE 7.44 The sequence of events in a sequencing batch reactor (SBR).

nitrate is present. If the nitrate is generated by nitrification during the react period and is present in the liquid retained in the bioreactor at the end of the cycle, then imposition of a mixed but unaerated interval during the early portion of the fill and react periods will make the SBR behave like a continuous flow MLE system. Furthermore, inclusion of another mixed but unaerated interval later in the react period can make the SBR act like the Bardenpho process. On the other hand, if the SBR is operated with a short SRT so that no nitrate is produced, mixing without aeration during fill and react can lead to the selection of PAOs, allowing the SBR to behave like the Phoredox or A/O continuous systems. It is apparent from these few examples that SBRs can be designed and operated to mimic many different continuous flow processes. Readers wanting more information about this ability should consult the literature.[1,16]

After the react period has been completed, all mixing and aeration are stopped and the biomass is allowed to settle. Just as in continuous flow processes, this accomplishes two things. It produces a clear effluent suitable for discharge and it retains biomass for SRT control. Solids wastage may be accomplished at the end of the settle period, simulating the conventional wastage strategy for continuous flow systems, or it may be done at the end of the react period, simulating the Garrett[12] wastage strategy. Regardless of when wastage is done, however, after sufficient settling has occurred, treated effluent may be removed by decantation during the draw period. The amount of liquid and biomass retained in the bioreactor constitutes the biomass recycle for the next cycle. If a large volume is retained relative to the influent volume in order to provide nitrate for an initial denitrification period, then the retained volume is analogous to biomass recycle plus mixed liquor recirculation in a continuous process.

Finally, an idle period is generally allowed in each cycle to provide flexibility. This is particularly important for a system with several SBRs because it allows their operation to be synchronized for maximum effectiveness. Mixing and aeration may or may not be used during the idle period, depending on the overall process objectives, and the length of the idle period may vary from cycle to cycle as the system needs dictate. The beginning of the new fill period terminates the idle period and initiates a new cycle.

7.8.2 Analogy to Continuous Systems

In the previous section we saw that an SBR is capable of mimicking the operation of many different types of continuous systems, depending upon how the cycle is conducted. As a consequence, the performance of an SBR is similar to the performance of the analogous continuous system, provided that the two have the same SRT. Just as with continuous systems, the SRT determines the amount of biomass in the SBR, thereby determining its overall average performance. Furthermore, as with the other systems described in this chapter, it is not possible to derive analytical expressions describing the SBR performance. Because the biomass in an SBR is in a dynamic state, a set of dynamic differential equations must be written and solved to determine system performance. Oles and colleagues[28,29] developed a computer code that does this for ASM No. 1 and it was used here to examine the performance of SBRs accomplishing carbon oxidation, nitrification, and denitrification. Use of the program requires specification of the SBR operating characteristics. This may be accomplished by analogy with continuous systems.

The HRT and SRT for continuous flow systems were defined by Equations 4.15 and 5.1, respectively:

$$\tau = \frac{V}{F} \tag{4.15}$$

and

$$\Theta_c = \frac{V \cdot X}{F_w \cdot X_w}, \tag{5.1}$$

where X and X_w are the biomass concentrations in the bioreactor and wastage stream, respectively, regardless of the unit system in which they are measured, as long as consistent units are used. In the case of an SBR, the wastewater flow rate, F, and the solids wastage flow rate, F_w, must be defined in terms of the total volume processed and wasted per day, respectively. The retention times defined by the above equations must be considered to be nominal retention times when applied to an SBR because the reactor volume, V, serves two functions, reaction plus settling. If the SBR is properly designed, all reactions will have been completed by the end of the react period and little change will occur in the soluble constituents during the settle period. In other words, all biochemical events can be assumed to take place during only the fill and react periods, with nothing except liquid:solid separation occurring during the settle and decant periods. This is the same assumption that was made in writing the equations describing continuous reactor performance in that time spent in the settler was not considered. To make Equations 4.15 and 5.1 equivalently applicable to SBRs, we must define an effective HRT, τ_e, and an effective SRT, Θ_{ce}, by multiplying both equations by the fraction of a cycle devoted to fill plus react, ζ:

$$\tau_e = \frac{\zeta \cdot V}{F}, \tag{7.8}$$

$$\Theta_{ce} = \frac{\zeta \cdot V \cdot X}{F_w \cdot X_w}. \tag{7.9}$$

If wastage is accomplished at the end of the react period before settling is started, then X_w will be equal to X and Equation 7.9 will simplify to

$$\Theta_{ce} = \frac{\zeta \cdot V}{F_w}. \tag{7.10}$$

An SBR and a continuous system will have similar MLSS concentrations as long as the effective HRT and SRT of the SBR are equal to the HRT and SRT of the continuous system. As a consequence, they will have similar overall process performance if the hydraulic regime and other characteristics are similar. However, it should be stressed that this will only be true as long as the effective HRTs and SRTs for the two systems are comparable.

Having set the effective HRT and SRT for an SBR, the major task is to determine the number of cycles required per day, N_c, thereby fixing the volume of wastewater applied per cycle, F_c, since by definition:

$$F_c = \frac{F}{N_c}. \tag{7.11}$$

Fixing the number of cycles per day also fixes the cycle time because it is just the reciprocal of N_c. It should be emphasized that one cannot simply choose N_c freely. Rather it is tied to a number of other decision variables. One important decision variable is the volume retained in the SBR, V_{br}, following the draw period for the purpose of returning biomass for the next cycle. Let α represent that volume expressed as a fraction of the feed added per cycle:

$$\alpha = \frac{V_{br}}{F_c}. \tag{7.12}$$

It is analogous to the biomass recycle ratio in a continuous flow system and, hence, the same symbol is used. Another important decision variable is the volume retained in the SBR, V_{nr}, following the draw period for the purpose of returning nitrate for use as an electron acceptor at the beginning of the next cycle. Let β represent that volume expressed as a fraction of the feed added per cycle:

$$\beta = \frac{V_{nr}}{F_c}. \tag{7.13}$$

It is analogous to the mixed liquor recirculation ratio in a continuous flow system. Recognizing that the volume of the SBR, V, must accommodate the incoming flow per cycle plus the volume retained per cycle gives:

$$V = F_c(1 + \alpha + \beta). \tag{7.14}$$

Substitution of Equations 7.11 and 7.14 into Equation 7.8 relates the number of cycles per day to all of the important decision variables:

$$N_c = \frac{\zeta(1 + \alpha + \beta)}{\tau_e}. \tag{7.15}$$

Thus, it can be seen why N_c cannot be chosen freely. All of the decision variables are linked and the choice of one will affect the choice of another. Nevertheless, by proper choice of α, β, and τ_e it is possible to have an SBR that mimics closely the performance of a continuous flow system with an equal SRT.

7.8.3 Effects of Cycle Characteristics

To illustrate the similarity in performance of continuous flow and SBR systems, simulations were performed with SBRSIM[29] using the parameter values listed in Table 6.3 and the wastewater characteristics in Table 6.6. Because SBRSIM implements IWA ASM No. 1, the reactions included were the same as those in the simulations with SSSP[5] presented previously. The situation simulated was analogous to the MLE system depicted in Figure 7.28, with one important difference. A 10-minute fill period was used, which made the SBR system behave like a plug-flow continuous reactor. Thus, rather than having a CSTR for the anoxic and aerobic zones, each behaved in a plug-flow manner.

The first set of simulations was done to illustrate the effect of SRT on system performance, so the conditions employed were analogous to those described in the legend to Figure 7.29. This means that the effective HRT was six hours, α was 0.5, and β was 2.0. Furthermore, the first half of the combined fill plus react period was anoxic with no DO, whereas the second half was aerobic with a DO concentration of 2.0 mg/L. As stated above, the fill period lasted for 10 minutes and thus occurred under anoxic conditions.

Before examining the effects of SRT on the performance of the SBR, it would be instructive to examine the changes occurring within a cycle, thereby establishing the basis for system performance. Figure 7.45 illustrates the changes in the concentrations of readily biodegradable substrate (soluble organics), ammonia-N, and nitrate-N during a fill and react period of 101 minutes when the effective SRT was 10 days. The anoxic and aerobic periods each occupied 50% of the fill plus react time. As would be anticipated, the concentrations of soluble organics and ammonia-N rose during the fill period as wastewater was added to the volume of mixed liquor retained from the previous

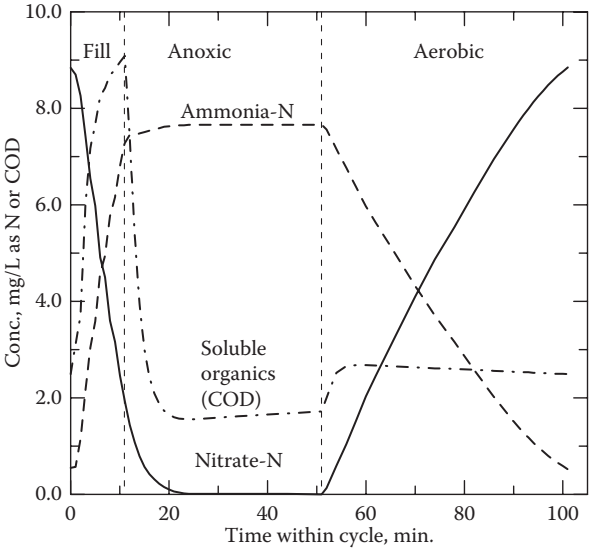

FIGURE 7.45 Performance of an SBR during a single cycle. The SBR was operated in a manner to mimic the performance of the MLE system shown in Figure 7.28. Effective HRT = 6 hr, effective SRT = 10 days, biomass recycle ratio = 0.5, mixed liquor recirculation ratio = 2.0, and aerobic fraction = 50%. Influent flow = 1000 m^3/day; influent concentrations are given in Table 6.6. Parameters are listed in Table 6.3. The DO concentration was zero during the anoxic period and 2.0 mg/L during the aerobic period.

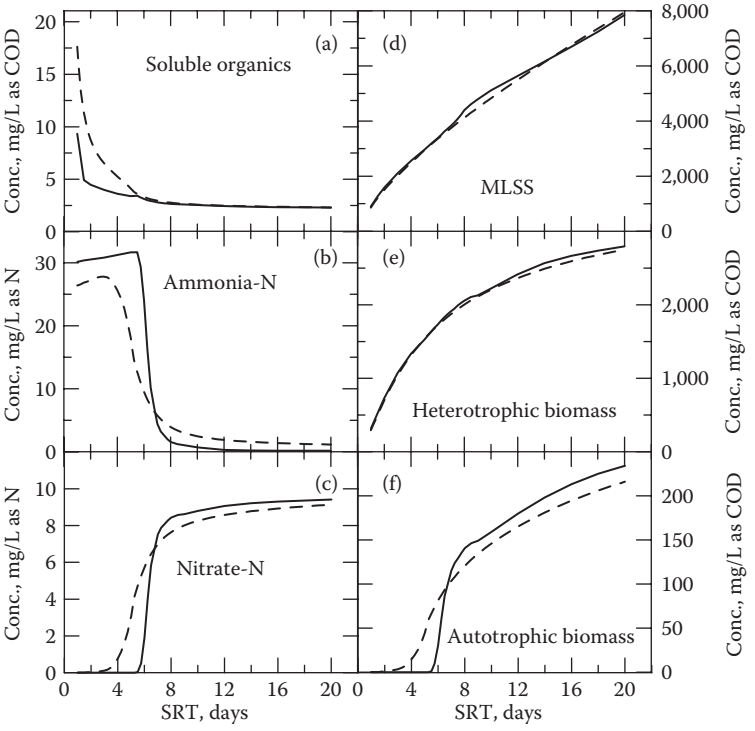

FIGURE 7.46 Effect of the effective SRT on the concentrations at the end of a cycle in an SBR operated in a manner to mimic the performance of the MLE system shown in Figure 7.28. For comparison, the dashed curves represent the concentrations in the aerobic (second) tank of that system. They are the same as the curves in Figure 7.29. With the exception of the SRT, the characteristics were the same as those listed in Figure 7.45. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

cycle. The concentration of nitrate-N, which came from the volume retained from the previous cycle, dropped as it served as an electron acceptor for biodegradation of the readily biodegradable substrate. It should be noted that carbon oxidation occurred throughout the fill period and that the reaction rate was sufficiently high to limit the buildup of soluble organics. Upon completion of the fill period, the remainder of the soluble organics and nitrate-N were rapidly depleted. In this case, the mass of nitrate-N retained from the previous cycle was well balanced with the mass of readily biodegradable COD added so that all nitrate-N was removed while little soluble organic matter remained. Nitrification was the main event during the aerobic react period and the reaction behaved in an almost zero-order manner over much of the time, as evidenced by the almost linear ammonia-N and nitrate-N curves. The soluble organics rose slightly because of their production by hydrolysis reactions. The model predicts that their concentration does not reach zero because they are continually being produced while they are being degraded. It is clear that the length of the anoxic period was excessive. Complete denitrification was achieved within 25 minutes, after which little occurred due to the lack of an electron acceptor. Furthermore, complete nitrification was not achieved during the aerobic period. Consequently, for this SRT, it would have been better to devote less time to anoxic conditions and more to aerobic conditions. The effects of the aerobic fraction of the fill plus react period will be addressed later.

The effect of SRT on the performance of this SBR system is shown by the solid lines in Figure 7.46. For comparison, the performance of an analogous MLE system is shown by the dashed lines. They are the same as the second bioreactor curves in Figure 7.29. The similarity in the performance of the two systems is apparent; the differences are due to the plug-flow nature of the SBR that allows it to achieve better removal of the soluble constituents. The particulate constituents are very similar,

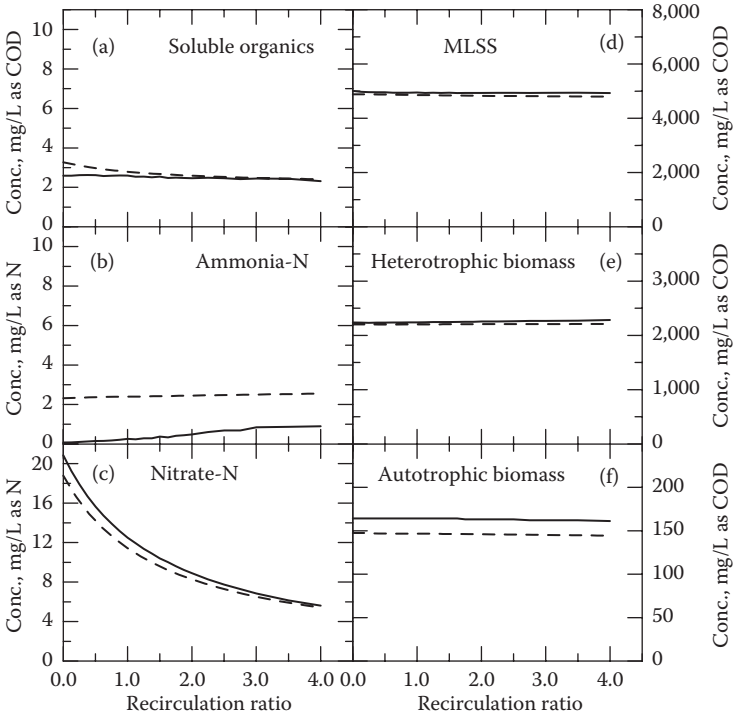

FIGURE 7.47 Effect of the mixed liquor recirculation ratio (β) on the concentrations at the end of a cycle in an SBR operated in a manner to mimic the performance of the MLE system shown in Figure 7.28. For comparison, the dashed curves represent the concentrations in the aerobic (second) tank of that system. They are the same as the curves in Figure 7.31. With the exception of the recirculation ratio, the characteristics were the same as those listed in Figure 7.45. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

however, for the same reason that the particulate constituents were the same in the CSTR and tanks-in-series systems. This emphasizes the point that continuous flow and batch systems with the same effective SRT will contain the same amount of biomass, thereby making the biomass concentrations the same when the effective HRTs are the same. Consequently, the simple analytical model for a CSTR in Chapter 5 can be used to help select the desired effective HRT for an SBR, just as it can be used for the other bioreactor systems discussed in this chapter.

It was stated earlier that the volume retained in an SBR at the end of each cycle is analogous to the combined effects of biomass recycle and mixed liquor recirculation in a continuous flow system. Consequently, changing that volume should have an effect on system performance analogous to the effects of changing the mixed liquor recirculation ratio on the MLE system. The solid lines in Figure 7.47 show the effects of changing the retained mixed liquor recirculation volume expressed as a ratio of the influent volume added per cycle (i.e., β). For comparison, the effects on the analogous MLE system are shown as dashed lines. They are the same as the second bioreactor curves in Figure 7.30. The similarity between the two systems is obvious. More complete nitrification occurs in the SBR because of its plug-flow nature (due to the short fill period), resulting in slightly more autotrophic biomass and slightly higher nitrate-N concentrations. Otherwise, the effects are essentially the same.

The third variable that was examined when we investigated the MLE system was the fraction of the total system volume that was aerobic. The analogous variable for the SBR is the fraction of the fill plus react period that is aerobic. Both can be expressed as the aerobic fraction. Figure 7.48 illustrates the effect of the aerobic fraction on the performance of both systems. Again, the similarity is striking, with the only differences being in the ammonia-N and autotrophic biomass concentrations, both of

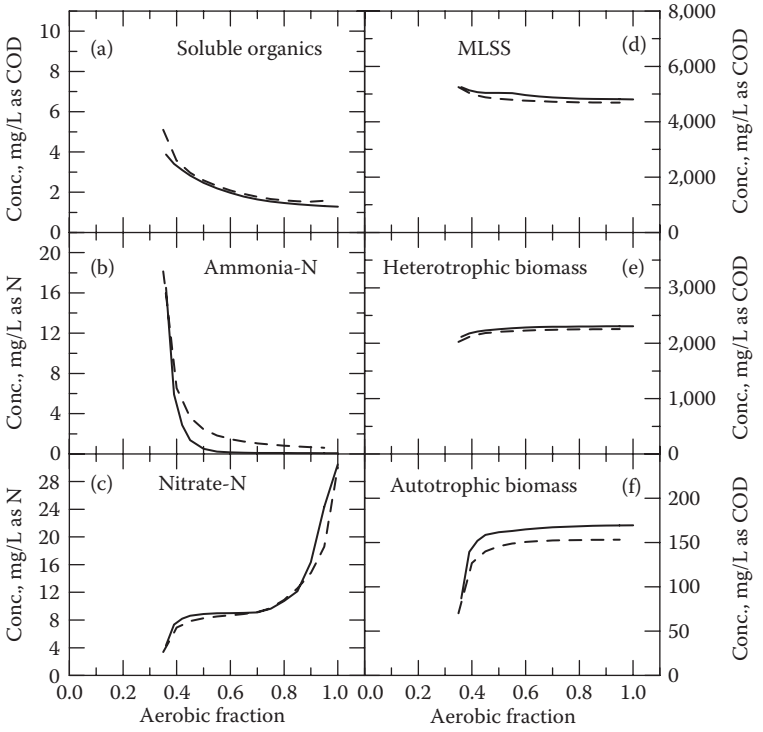

FIGURE 7.48 Effect of the fraction of the fill plus react period that is aerobic (aerobic fraction) on the concentrations at the end of a cycle in an SBR operated in a manner to mimic the performance of the MLE system shown in Figure 7.28. For comparison, the dashed curves represent the concentrations in the aerobic (second) tank of that system. They are the same as the curves in Figure 7.33. With the exception of the aerobic fraction, the characteristics were the same as those listed in Figure 7.45. To express the MLSS and biomass concentrations in TSS units, divide by $i_{O/XB,T}$ = 1.20 mg COD/mg TSS.

which are due to the plug-flow nature of this SBR. It will be recalled from Figure 7.45 that an anoxic period of around 30 minutes was adequate for complete denitrification at an SRT of 10 days, the value used in the simulations for Figure 7.48. This suggests that the best system performance would occur with an aerobic fraction of 60 to 70%, which is indeed the case, as shown in Figure 7.48.

In summary, SBR systems can be operated in ways that mimic the performance of the continuous flow systems described in this chapter. As long as the effective HRT and effective SRT are the same, the two systems will contain similar biomass concentrations and have similar performance. Furthermore, variation of the length of the fill period will allow the performance of the SBR to approach any degree of mixing between complete mixing and plug flow. As a consequence of their flexibility, SBRs are finding increased usage for wastewater treatment. Decisions related to their design will be addressed in Chapters 11 and 12 along with the continuous flow activated sludge and nutrient removal systems.

7.9 KEY POINTS

1. Several important variations of the activated sludge process can be simulated as systems containing continuous stirred tank reactors (CSTRs) in series with various feed points and recycle streams. They are conventional activated sludge (CAS), step feed activated sludge (SFAS), contact stabilization activated sludge (CSAS), modified Ludzack-Ettinger (MLE) process, four-stage Bardenpho process, and biological phosphorus removal (BPR) process.

2. Biomass in each bioreactor of a multiple reactor system has a unique specific growth rate in response to the concentration of the growth limiting nutrient(s) in the bioreactor, whereas the solids retention time (SRT) is defined with respect to the system as a whole. Thus, the SRT is not related to the specific growth rate in the same way that it is in a single bioreactor. Nevertheless, because the SRT is related to the net average specific growth rate of the entire system, it is still an important design and operational variable.
3. The process loading factor in each bioreactor of a chain of CSTRs is proportional to the specific growth rate of the bacteria in that bioreactor. Similarly, the instantaneous process loading factor achieved during the fill period for a sequencing batch reactor (SBR) is representative of the specific growth rate of the bacteria during that period.
4. At a given SRT, conventional or plug-flow activated sludge systems achieve slightly better removal or transformation of soluble constituents than a single CSTR, but exhibit almost identical biomass concentrations, oxygen utilization rates, and excess biomass production rates.
5. A CAS system exhibits less diurnal variability in its effluent organic substrate and ammonia-N concentrations than a CSTR operating at the same SRT, but experiences greater variability in its oxygen requirement.
6. Most soluble organic matter is removed in the early part of a CAS system, whereas nitrification occurs over a greater part of the bioreactor system. The longer the SRT, the nearer the influent end of the bioreactor the reactions are completed.
7. Within the normal range found in practice, changes in the recycle ratio of biomass around a CAS system have little impact on system performance.
8. At a given SRT and system hydraulic residence time (HRT), the mixed liquor suspended solids (MLSS) concentration entering the settler from the last tank of an SFAS system with influent distributed evenly along the system is less than the concentration from a CAS system or from a single CSTR. System performance, however, is only slightly worse than that of a single CSTR.
9. An SFAS system with influent distributed evenly along the system exhibits more diurnal variability in its effluent organic substrate and ammonia-N concentrations than a CSTR operating at the same SRT, but experiences about the same variability in its oxygen requirement.
10. The MLSS in an SFAS system has a concentration gradient from the influent to the effluent end, with the influent end of the system having higher concentrations. As a consequence, soluble organic substrate and ammonia-N exhibit the opposite concentration gradient.
11. Increasing the recycle ratio around an SFAS system decreases the concentration gradient through the system and gives better system performance.
12. Although the removal of soluble organic substrate by a CSAS system is not as complete as the removal in a single CSTR, its response to changes in SRT is qualitatively similar. The response of ammonia-N to such changes is both qualitatively and quantitatively different, however.
13. Because the contact tank in a CSAS system has a lower MLSS concentration and a smaller volume than a single CSTR with the same SRT and system volume, the variations in the effluent concentrations of soluble organic substrate and ammonia-N in response to typical diurnal variations in loading are greater than those of the CSTR.
14. The biomass recycle ratio has a large impact on the performance of a CSAS system. Up to a point, the larger it is, the lower the concentrations of soluble organic substrate and ammonia-N in the effluent.
15. For a fixed total system volume, the larger the contact tank in a CSAS system, the more the system performs like a single CSTR of equal SRT and total system HRT.

16. When operated at an SRT sufficient to achieve a high degree of nitrification, the MLE process discharges less nitrate-N and uses less oxygen than a single CSTR because denitrification occurs in the first (anoxic) bioreactor.
17. Although the aerobic SRT is an important determinant of the degree of nitrification achieved in systems with unaerated zones, it is not the only one.
18. If the rate of recirculation of MLSS from the aerobic zone to the anoxic zone of an MLE system is insufficient to return enough electron acceptor for the amount of electron donor entering in the influent, the removal of nitrate-N by the system will not be as high as it could be.
19. For a fixed system volume and SRT, there is a region of optimum aerobic volume fraction that results in the minimum discharge of nitrogen from an MLE system.
20. By providing a second anoxic zone that achieves denitrification through the use of slowly biodegradable substrate, the four-stage Bardenpho process can produce effluents with less total nitrogen than the MLE process can.
21. Selection of the sizes of the bioreactors in a four-stage Bardenpho system to achieve the minimum possible discharge of nitrogen requires the use of optimization techniques.
22. There is a limited range of aerobic SRT over which a simple Phoredox (or A/O) system will work properly. If the aerobic SRT is too short, the phosphate accumulating organisms (PAOs) will wash out. If it is too long, nitrifying bacteria will be able to grow, providing an inorganic electron acceptor to the anaerobic zone, which allows the ordinary heterotrophic organisms to out-compete the PAOs for substrate, reducing the amount of phosphate removed.
23. There is an optimal split between the volumes of the anaerobic and aerobic tanks in the Phoredox (or A/O) system. Optimal phosphorus removal occurs when the aerobic SRT is small enough to exclude growth of autotrophic nitrifying bacteria, yet large enough to allow the PAOs to grow, and the time in the anaerobic tank is sufficiently large to allow efficient fermentation and uptake of the resulting acetate.
24. Sequencing batch reactors (SBRs) are batch reactors that can be used to treat wastewaters on a semicontinuous basis by going through a repeated sequence of steps consisting of fill, react, settle, decant, draw, and idle. They are very flexible in operation and can be used to mimic many complex continuous reactor systems by altering the length of the fill and react periods and by employing different environmental conditions (aerobic, anaerobic, anoxic) during those periods.
25. Continuous flow processes and SBRs have similar performance as long as they have the same effective SRT and HRT, as well as similar environmental conditions.

7.10 STUDY QUESTIONS

1. State the main characteristics of each of the following biochemical operations and draw sketches showing how each may be simulated: conventional activated sludge (CAS), step feed activated sludge (SFAS), contact stabilization activated sludge (CSAS), modified Ludzack-Ettinger (MLE) process, four-stage Bardenpho process, and two-stage biological phosphorus removal (BPR).
2. Define the SRT for a system containing N bioreactors, each of volume V_i. What is the simplest way of controlling the SRT in such a system? Why?
3. Why does a CAS system remove or transform more soluble material than a CSTR with the same SRT, yet still have the same excess biomass production rate?
4. Draw a sketch comparing the variability in the oxygen requirement of a CAS system to that in a CSTR with the same SRT when both receive the same diurnal loading. What are the most important differences between the two systems? Why do they occur?

5. Draw a sketch depicting the tank-to-tank change in the concentrations of soluble organic matter and ammonia-N down a chain of bioreactors representative of CAS with an SRT sufficiently long to allow nitrification to occur. Why does the sketch look as it does? How will a change in the SRT influence the sketch? Why?
6. Why do changes in the recycle ratio of biomass around a CAS system have little impact on system performance?
7. Why is the MLSS concentration entering the settler from the last tank of an SFAS system less than the concentration from a CAS system or from a single CSTR with the same SRT and system HRT?
8. Why does an SFAS system exhibit more variability in its effluent organic substrate and ammonia-N concentrations in response to a diurnal load than does a CSTR with equal SRT and system HRT? How do the variations in oxygen requirement compare? Why?
9. Why do the concentrations of soluble organic substrate and ammonia-N increase as the flow moves from tank to tank in an SFAS system?
10. Why does an increase in the recycle ratio around an SFAS system cause slightly better system performance?
11. Using a computer code implementing one of the IWA activated sludge models, investigate the impact of distributing the influent to the SFAS system described in Figure 7.11 so that it is equally divided among the first four bioreactors, with none entering the fifth. How does the steady-state performance of this system over a range of SRTs compare to the system in Figure 7.11? How does it compare to the CSTR? How does the dynamic response of the system to a diurnal load compare to that of the other two systems? Discuss the reasons for the differences in response.
12. Why is the response of ammonia-N to changes in the SRT of a CSAS system qualitatively different from the response of soluble organic matter?
13. Even though the stabilization tank in the CSAS system receives biomass recycle flow at a constant rate, it still experiences variability in its oxygen requirement when the system receives typical diurnal variations in loading. Why?
14. Why does the biomass recycle ratio have a large impact on the performance of a CSAS system?
15. For a fixed total system volume, as the volume of the contact tank is increased in a CSAS system, the concentration of ammonia-N changes in a manner that is qualitatively different from the manner in which the soluble organic substrate concentration changes. Why?
16. Using a computer code implementing one of the IWA activated sludge models, investigate the effect of HRT on the performance of the CSAS system described in Figure 7.20. Maintain the SRT constant at a value of 12 days and vary the HRT over the range of 1.0 to 24 hours by changing the total system volume while keeping the two bioreactors of equal size. Explain your results.
17. Why does the MLE process fail when its operational conditions are such that nitrification cannot occur?
18. Although the aerobic SRT is an important determinant of the extent of nitrification achieved in a system with unaerated zones, equivalent degrees of nitrification will not be achieved in systems with and without unaerated zones, even though they may have the same aerobic SRTs. Why?
19. Draw a sketch showing the effect of mixed liquor recirculation ratio on the performance of the MLE process and explain why it has the effect that it does. Why does the impact of increased recirculation reach a point of diminishing return?
20. Why is there a region of optimum aerobic volume fraction that results in the minimum discharge of nitrogen from an MLE system of fixed total volume and SRT?

21. Using a computer code implementing one of the IWA activated sludge models, investigate the effect of HRT on the performance of the MLE system described in Figure 7.29. Maintain the SRT constant at a value of 12 days and vary the HRT by changing the total system volume while keeping the aerobic fraction constant at 50%. Explain your results.
22. Using a computer code implementing one of the IWA activated sludge models, explore the interactions between mixed liquor recirculation ratio and aerobic volume fraction for the MLE system described in Figure 7.29. Maintain the SRT constant at a value of 12 days and the total volume constant at 250 m^3. Vary the recirculation ratio over the range of 3 to 5 and the aerobic volume fraction over the range of 0.4 to 0.8. Plot total nitrogen concentration in the effluent as a function of aerobic volume fraction with recirculation ratio as the parameter and determine the best combination for the system. Then explain your results.
23. A four-stage Bardenpho system can achieve an effluent with less total nitrogen in it than an MLE system can, even though the two systems have the same SRT, total system volume, and aerobic volume fraction. Why?
24. Using a computer code implementing one of the IWA activated sludge models, investigate the effect on the effluent nitrogen concentration of changing the distribution of anoxic volume between bioreactors 1 and 3 in the Bardenpho system described in Figure 7.36. Use an SRT of 10 days. Why does the system respond as it does?
25. Explain why the occurrence of nitrification has a deleterious effect on biological phosphorus removal in the Phoredox process. Then suggest a process configuration that would achieve a high degree of biological phosphorus removal, complete nitrification, and partial denitrification and explain why it would work.
26. Using a computer code for IWA ASM No. 2d, demonstrate the effects of the key process variables on the performance of the process you suggested in Study Question 25. Use the wastewater characteristics listed in Table 7.1 and the kinetic parameters recommended in Section 7.7.2.
27. Describe how you would set up an SBR system to be analogous to the four-tank Bardenpho system described in Figure 7.36.
28. Using a computer code implementing one of the IWA activated sludge models, repeat Study Question 24 for the SBR system identified in Study Question 27.

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8 Stoichiometry, Kinetics, and Simulations of Anaerobic Biochemical Operations

Chapters 5, 6, and 7 considered aerobic/anoxic systems or systems in which an anaerobic reactor was provided in series with aerobic and/or anoxic reactors to provide an environment conducive to the selection of phosphate accumulating organisms. Strictly anaerobic bioreactors, on the other hand, have played an important role in wastewater treatment systems for the stabilization of particulate substrates, such as primary and secondary sludges, for the treatment of high strength soluble wastewaters, and for the production of acetate and other volatile fatty acids (VFAs) required in biological phosphorus removal (BPR) systems. Because distinct microbial communities are developed in strictly anaerobic systems and because of the important roles that they play in wastewater treatment we will consider them in this chapter. Fortunately, the recent development of the Anaerobic Digestion Model (ADM) No. 1 by a task group of the International Water Association (IWA)[4,5] allows us to do so quantitatively and to examine the performance of a simple anaerobic continuous stirred tank reactor (CSTR) through simulation.

As discussed in Section 2.3.3 and seen in Figure 2.4, in anaerobic operations three groups of bacteria are involved in acidogenesis and two in methanogenesis. Acidogenesis includes fermentation, anaerobic oxidation, and H_2 oxidizing acetogenesis. Fermentative bacteria convert amino acids and simple sugars to acetic acid, VFAs, and a minor amount of H_2. Bacteria performing anaerobic oxidation convert long chain fatty acids and VFAs to acetic acid and major amounts of H_2. Finally H_2 oxidizing acetogens form acetic acid from carbon dioxide and H_2, but they are considered to be of minor importance in anaerobic wastewater treatment operations and will not be considered here. The two groups of methanogens are aceticlastic methanogens, which split acetic acid into methane and carbon dioxide, and H_2 oxidizing methanogens, which reduce carbon dioxide to form methane.

To have a complete picture of the stoichiometry and kinetics of microbial growth and substrate utilization in anaerobic systems, the parameter values associated with these processes for all groups should be characterized. Unfortunately, because of the role of H_2 in regulating microbial activity, the close association between H_2 producing and H_2 consuming bacteria and archaea, and other complex interactions that exist in anaerobic systems,[16,26,46] this is not an easy task. Through development of ADM No. 1 by the IWA task group for mathematical modeling of anaerobic digestion processes,[4] many of the complex stoichiometric and kinetic interactions among the microbial groups in anaerobic systems have been organized in accordance with the performance of those groups. The stoichiometry and kinetics of the groups of anaerobic bacteria involved with anaerobic processes are discussed in detail in this chapter, and are used to demonstrate the performance of anaerobic digesters through the use of ADM No. 1.

8.1 STOICHIOMETRY OF ANAEROBIC BIOCHEMICAL OPERATIONS

In Section 2.3.3, we reviewed the multistep nature of anaerobic processes that result in methanogenesis, and showed the relationship between these steps in Figure 2.4. First, complex biodegradable particulate materials contained in the feed or formed during the process are broken down by nonbiological and biological processes through disintegration and hydrolysis to form monosaccharides, long chain

fatty acids (LCFAs), and amino acids, which are basic biochemical macromolecules (reaction 1 in Figure 2.4). These macromolecules serve as substrates for multiple acidogenic reactions that primarily produce acetate and H_2 (reactions 2 through 5 in Figure 2.4). Subsequently, the acetate and H_2 can each be metabolized through distinct pathways to form methane, which is where most of the chemical oxygen demand (COD) removed in the process ultimately ends up (reactions 6 and 7). The stoichiometry of these biochemical processes are such that very little new biomass is formed, and under ideal operation the soluble products of the various steps do not accumulate, thereby making the gases methane and carbon dioxide the main products. Here, we elaborate on the stoichiometry of these steps as well as significant physical and chemical processes that occur during anaerobic biochemical operations.

8.1.1 Solubilization of Particulate and High Molecular Weight Organic Matter

Anaerobic operations are used to treat wastes that contain both particulate and soluble constituents. Perhaps the most widely used application in wastewater treatment is the stabilization of primary sludge and waste biomass from aerobic/anoxic biological treatment processes. In this case, the waste undergoing anaerobic treatment is essentially all particulate and the treatment process is called anaerobic digestion. High strength wastewaters that contain little particulate matter, which are common among some food and chemical industries, are also frequently treated using anaerobic operations. When particulate substrates are being treated, the process must be designed in such a way that the particulate and high molecular weight organic matter can be solubilized because solubilization is a precondition for biodegradation. Although solubilization of particulate organic matter is thought of as a one-step hydrolytic process in aerobic/anoxic processes (Section 3.5), it is considered to be a two-step process in anaerobic systems. Those two steps are disintegration and hydrolysis.

Disintegration refers to the lysis of cells into organic particles consisting of the macromolecules that are part of living cells, and the further breakdown of those particles into their basic high molecular weight biochemical components: carbohydrates, nucleic acids, proteins, and lipids.[38] Disintegration also produces soluble and particulate inert material due to lysis of cellular material present in the particulate feed to the reactor. In ADM No. 1, all particles (cells and otherwise) that come into the digester and undergo disintegration are called composite particulate material, designated X_C. It contains particulate biochemical components plus inert material. Note that because X_C includes inert material, it is different from X_S that has been used previously in this book to denote biodegradable particulate organic matter. The particulate inert material that is generated during disintegration influences the fraction of the incoming particulate COD that is ultimately decomposed into biochemical component fractions, which are biodegradable. Therefore, it is important to estimate this fraction. If the incoming waste is biomass from an aerobic/anoxic bioreactor, Equation 5.33 can be used to estimate the inert particulate COD in it. We can estimate the particulate biochemical components of waste biomass using what we know about cellular composition. For example, rapidly growing cells contain approximately 56% protein, 24% nucleic acid, 12% lipid, and 8% carbohydrate on a dry mass basis.[33] Although slower growing cells, such as would be present in aerobic/anoxic biochemical operations, may have a different relative biochemical composition, it is expected that protein will still be the primary biochemical component.[37] Consequently, it should be noted that ADM No. 1 assumes that carbohydrates, proteins, and lipids contribute equal fractions (30% each) after X_C undergoes disintegration, with the remaining 10% being inert.[4] This is an assumption that probably reflects a desire to simplify the model. It has been shown that enhancing disintegration by external means (e.g., mechanical shearing or ultrasonic treatment) enhances biogas yield and sludge destruction, but causes poorer dewatering characteristics.[11,29,30,47] This demonstrates the important role that disintegration plays in anaerobic processes treating particulate organic matter.

Disintegration precedes hydrolysis, which is the continued degradation of the biochemical components into their soluble monomers: carbohydrates and nucleic acids into monosaccharides, proteins into amino acids, and lipids into LCFAs. Hydrolysis employs extracellular enzymes, which are either excreted into solution or directly onto particles to be hydrolyzed when the bacteria that

generate hydrolytic enzymes attach to particles.[49] Because hydrolysis is the conversion of a basic biochemical component from one form to another (e.g., particulate carbohydrate to monosaccharides), one mole of the monomer in a particle is assumed to generate one mole of soluble monomer upon hydrolysis. However, because most lipids in microbial cells are complex macromolecules, a small amount of carbohydrate is produced during lipid hydrolysis and it must be considered.[4]

8.1.2 Fermentation and Anaerobic Oxidation Reactions

The stoichiometries of fermentation and anaerobic oxidation reactions are vast and complex. They often rely upon syntrophic associations (when one species degrades a substrate only when the end product of that degradation is removed by another species) for reactions to occur. Here, we discuss key stoichiometric features of those reactions that are most relevant to anaerobic digestion. Therefore, we are focused on degradation of the amino acids, monosaccharides, and LCFAs that are produced through hydrolysis.

Acetate, propionate, and butyrate are the most commonly encountered stable fermentation products from monosaccharide acidogenesis.[4] Assuming six-carbon (hexose) monosaccharides as a starting point, the stoichiometries for the typical acidogenic (fermentation) reactions are given in Equation 8.1 for acetate formation, Equation 8.2 for butyrate formation, and Equation 8.3 for propionate plus acetate formation:

$$C_6H_{12}O_6 + 2H_2O \rightarrow 2CH_3COOH + 2CO_2 + 4H_2, \tag{8.1}$$

$$C_6H_{12}O_6 \rightarrow CH_3CH_2CH_2COOH + 2CO_2 + 2H_2, \tag{8.2}$$

and

$$3C_6H_{12}O_6 \rightarrow 4CH_3CH_2COOH + 2CH_3COOH + 2CO_2 + 2H_2O. \tag{8.3}$$

In some cases, such as during low pH fermentation, ethanol is a common product of fermentation and is increasingly important when two-stage (fermentation followed by methanogenesis) digestion is employed. Lactate is also a common product of hexose fermentation[25] but is not very stable in most anaerobic digesters.[4] Degradation of lactate follows the stoichiometry of glucose fermentation and, therefore, is not highlighted separately here or in ADM No. 1. Note that the fermentation of five-carbon monosaccharides, such as those produced by nucleic acid hydrolysis, can produce organic acids as well as alcohols as products. Energy is acquired from these reactions through substrate-level phosphorylation. The overall gain of adenosine triphosphate (ATP) is two to four moles of ATP per mole of glucose consumed, and is lower than for reactions fueled by electron transport phosphorylation. This helps to explain the moderate true growth yields for bacteria performing these reactions, which range from 0.10 to 0.17 g biomass COD/g carbohydrate COD consumed.

About half of all amino acid fermentation reactions occur primarily via Strickland reactions,[34,39] where some amino acids serve as an electron donor, some serve as an electron acceptor, and some play either role. Hydrogen gas is a relatively frequent by-product of amino acid fermentations (both Strickland and non-Strickland reactions), and the H_2 produced must be coupled to H_2 utilizing methanogenic bacteria in order to maintain a favorable thermodynamic environment for the fermentation to continue (this is discussed below under anaerobic oxidation). Because of the large number of amino acids a detailed listing of the stoichiometric relationships is beyond the scope of this book and the reader is referred elsewhere for details.[4,39] The true growth yield of bacteria performing amino acid acidogenic fermentation reactions is reported to be 0.06 g biomass COD/g amino acid COD consumed at near neutral pH, which is the condition typically found in single-stage anaerobic digestion. In two phase systems, which employ an upstream fermentation reactor where the pH can be maintained at values around 5.5, the true growth yield increases to 0.09–0.15 g biomass COD/g amino acid COD consumed.[4]

During anaerobic oxidation, the VFAs present in the reactor (propionate, butyrate, and valerate), as well as the LCFAs, are oxidized to acetic acid and H_2. Oxidation of LCFAs follows β-oxidation,[50] a cyclic process where one molecule of acetate is removed from the LCFA for each round of the cycle (in the case of LCFAs with an odd number of carbons, the last three carbons are released as propionate). The catabolic reactions for anaerobic oxidations of these compounds are shown in Table 8.1 (palmitate is used as the model molecule for LCFAs).[4] The oxidized product in these reactions is acetic acid and the electrons removed from the substrates are transferred to hydrogen ions, resulting in the formation of hydrogen gas. Although not shown in Table 8.1, the overall stoichiometry of anaerobic oxidation also leads to the formation of new biomass. However, the true growth yields associated with anaerobic oxidation are small: 0.04–0.05 g biomass COD/g substrate COD for propionate, 0.06 g biomass COD/g substrate COD for valerate and butyrate, and 0.04–0.045 g biomass COD/g substrate COD for LCFAs.[4,45] This means that the majority (about 70%) of the electrons present in the substrates that undergo anaerobic oxidation will end up in H_2. Furthermore, our inability to accurately predict the true growth yields for these metabolic processes does not significantly impact our ability to predict the overall performance of the system, since little biomass is formed.

Anaerobic processes operate close to thermodynamic equilibrium. This statement can be understood by inspecting Table 8.1, where the free energies for three different substrates undergoing anaerobic oxidation are shown for both standard conditions (ΔG^0) and for conditions typically found in anaerobic digesters ($\Delta G'$). The standard free energies of these three reactions are positive and, therefore, anaerobic oxidation is thermodynamically unfeasible under these conditions. However, under conditions found in anaerobic processes, methanogenic microorganisms use the products of anaerobic oxidation as substrates. This demand, coupled with the operating conditions experienced during anaerobic oxidation, leads to negative free energies ($\Delta G'$) and a thermodynamically feasible process. The free energy is most dependent on the dissolved H_2 concentration in the liquid phase and this concentration determines the feasibility of anaerobic oxidation. In anaerobic systems, H_2 oxidizing methanogens need to be in close physical proximity to the organisms carrying out anaerobic oxidation to maintain the H_2 concentration at low levels; hence, they grow syntrophically. This need for close physical proximity may be the reason why suspended growth anaerobic systems typically form dense and compact granules rather than dispersed growth. Later, in Section 8.2.2, when we discuss the kinetics of anaerobic oxidation, we will see how thermodynamic limitation of anaerobic oxidation can be included when modeling anaerobic systems.

TABLE 8.1
Reactions in Anaerobic Oxidation

Reaction	Stoichiometry	ΔG^0 (kJ/mol)[a]	$\Delta G'$ (kJ/mol)[b]
Anaerobic oxidation of propionate to acetate and hydrogen	$CH_3CH_2COOH + 2H_2O \rightarrow CH_3COOH + 3H_2 + CO_2$	76.2	−14.56
Anaerobic oxidation of butyrate to acetate and hydrogen	$CH_3CH_2CH_2COOH + 2H_2O \rightarrow 2CH_3COOH + 2H_2$	48	−25.6
Anaerobic oxidation of a long chain fatty acid to acetate and hydrogen[c]	$CH_3(CH_2)_{14}COOH + 14H_2O \rightarrow 8CH_3COOH + 14H_2$	404.8	−117.76

Source: Adapted from Batstone, D. J., Keller, J., Angelidaki, I., Kalyuzhnyi, S. V., Pavlosthathis, S. G., Rozzi, A., Sanders, W. T. M., Siegrist, H., and Vavilin, V. A., Anaerobic Digestion Model No. 1, *IWA Scientific and Technical Report No. 13,* IWA Publishing, London, 2002.

[a] Standard conditions are: temperature = 273 K, pressure = 0.1 MPa, and all concentrations = 1 M.

[b] The Gibbs free energies were calculated for the conditions in the reactor described in the ADM1 Technical Report: temperature = 298 K, pH 7, $\bar{p}_{H_2} = 10^{-5}$ bar, $CO_2 = HCO_3^- = 0.1M$, all organic acids (including LCFAs) = 0.001 M.

[c] Palmitate has been used as a model LCFA for this reaction.

8.1.3 Methanogenesis

As discussed in Section 2.3.3, the products of the acidogenic reactions, acetate and H_2, serve as the primary substrates for aceticlastic and H_2 utilizing methanogens, respectively. Although approximately two-thirds of methane is typically generated via aceticlastic methanogenesis, the actual distribution of methane production between these two types of methanogens is largely dictated by the nature of the substrate entering the digester and the distribution of products formed by acidogenesis. The H_2 utilizing methanogens must be coupled to the H_2 producing acetogens in order to keep the dissolved H_2 concentration within a narrow and specified range in order to prevent inhibition of the H_2 producing acetogens.

Both aceticlastic and H_2 utilizing methanogens have very low biomass yields, meaning they produce a very useful gaseous product but do not contribute significantly to the formation of new biomass in anaerobic processes. The aceticlastic methanogens have a biomass yield ranging from 0.03 to 0.07 g biomass COD/g acetate COD, while the H_2 utilizing methanogens have a biomass yield ranging from 0.01 to 0.06 g biomass COD/g H_2 COD.[4]

8.1.4 Physical and Chemical Processes in Anaerobic Systems

All biological reactions are influenced by physical and chemical processes. Under typical treatment conditions using aerobic/anoxic processes, the effects from physical and chemical processes are typically insignificant, as long as the alkalinity is sufficient to prevent large pH changes during treatment.[32] This is not the case for anaerobic processes, however, which are particularly influenced by a range of physical and chemical effects. Here, we discuss the most common physical and chemical processes that occur in anaerobic systems, including acid-base reactions, gas transfer, and precipitation.

8.1.4.1 Acid–Base Dissociations

A number of acid-base dissociations occur in anaerobic processes, and have a significant influence on pH, precipitation processes, and the composition of the gas phase. These reactions are quite rapid relative to the other biological, physical, and chemical processes occurring at the same time and, therefore, are often handled as a series of equilibrium reactions,[4] although they can also be explicitly modeled kinetically.[32] The most important dissociation reactions are related to the carbonate system, ammonium, and hydrogen sulfide. Chemicals associated with the carbonate system are generated during anaerobic processes through both fermentation (minor) and methanogenesis (major) because both produce CO_2 as a by-product. Ammonium is generated through cell lysis, hydrolysis, ammonification, and the metabolism of amino acids formed during hydrolysis. Hydrogen sulfide is a by-product of sulfate reducing bacteria, which we will not discuss in detail. However, it is important to recognize that they often exist in anaerobic processes, especially when the influent wastewater contains a large amount of sulfate, and that their by-products sometimes have significant impacts. In anaerobic digesters that receive waste biomass from a BPR system, the dissociation reactions associated with phosphate also become important. Furthermore, in fermentation reactors for VFA formation, acidic conditions occur and the acid-base dissociations of VFAs become important as well.

The carbonate system is a very complex, but well understood, acid-base process in anaerobic systems. Among the acid-base pairs included in the carbonate system, the $CO_{2(aq)}/HCO_3^-$ pair is the most relevant. It has a pK_a value of 6.35 at 25°C, which means that most of the total carbonate is in the HCO_3^- form for anaerobic processes maintained around neutral pH. Dissolved carbon dioxide ($CO_{2(aq)}$) is the predominant form over the acid form H_2CO_3 and, therefore, the state of the $CO_{2(aq)}/HCO_3^-$ dissociation pair is very much a function of $CO_{2(aq)}$ stripping. Similarly, the concentration of CO_3^{2-} is low in anaerobic processes operating near neutral pH, given the HCO_3^-/CO_3^{2-} pK_a of 10.3 (25°C). However, if enough divalent calcium (Ca^{2+}) is present, $CaCO_3$ precipitation can occur and influence the distribution of dissolved carbonate species. Anaerobic digestion model No. 1 only

considers the $CO_{2(aq)}/HCO_3^-$ acid-base pair, but cautions users of the influence that the other carbonate reactions may have on anaerobic processes. Overall, the carbonate system has a large impact on the pH of anaerobic processes, as can VFAs. The pK_a values for VFAs are typically around 4.8; therefore, VFA dissociation only becomes important in systems that are designed with a fermentation bioreactor. In those systems, the pH of the fermentation bioreactor is heavily influenced by the dissociation patterns of the VFAs present.

Ammonium/ammonia (NH_4^+/NH_3), H_2S/HS^-, and $H_2PO_4^-/HPO_4^{2-}$ are acid-base pairs that also influence anaerobic processes. The pK_a at 25°C for each is 9.25, 7.05, and 7.2, respectively. Ammonia stripping is not a significant factor at near neutral pH, given the high solubility of ammonium. However, the intentional precipitation of ammonium into the mineral struvite ($MgNH_4PO_4 \cdot 6H_2O$) is of increasing interest because it can cause costly scaling of reactors and piping systems. It can be managed in a way that allows for nutrient recovery to form a useful fertilizer.[13] Hydrogen sulfide (H_2S) is a different story, because it is readily stripped from the liquid phase to the gas phase and is highly corrosive. If enough cationic salts, such as iron, are present in the wastewater, HS^- can be precipitated and bound into the anaerobic biomass to prevent excessive H_2S emissions. Therefore, the state of the H_2S/HS^- acid-base balance is important in systems that receive enough sulfur to produce significant amounts of either form. The phosphate pair is important in systems that receive significant amounts of phosphorus, such as in the waste biomass from BPR systems or in urine treatment from source-separated waste management systems. Phosphate is also an important component in the precipitation of struvite, which will be discussed below.

8.1.4.2 Gas Transfer

Liquid-gas transfer in anaerobic processes refers primarily to the manner in which a constituent in the liquid phase is transferred to the gas phase. Equilibrium between the gas and liquid phases for constituents is estimated using Henry's Law that pertains to dilute solutions:

$$C_{G,i} = H_i \cdot S_i, \tag{8.4}$$

in which $C_{G,i}$ is the gas phase concentration of constituent i (atm), S_i is its liquid phase concentration (moles/m^3), and H_i is its Henry's Law coefficient (atmospheres of constituent i in the gas phase) (m^3 of liquid phase)/(moles of constituent i in the liquid phase). For volatile constituents of relevance to anaerobic processes, values of H_i indicate the relative tendency for the constituents to partition between the gas and liquid phases at equilibrium, with larger values indicating that a larger fraction of the constituent exists in the gas phase. The Henry's Law coefficients (atm·m^3/mole) at 25°C for biologically derived constituents of interest in anaerobic processes are:[24] $H_{H_2} = 1.28$; $H_{CH_4} = 0.747$; $H_{CO_2} = 0.00308$; $H_{H_2S} = 0.00102$; and $H_{NH_3} = 1.7 \times 10^{-5}$. Ammonia partitions mostly into the aqueous phase; however, ammonia emissions from some high strength wastes that undergo anaerobic treatment (e.g., animal wastes) can be significant.[12] Although H_2S is an important biologically derived gaseous component in anaerobic processes that receive higher than average sulfur concentrations, it is not included in ADM No. 1 because biological sulfate reduction was not considered. The transfer of CO_2 into the gas phase occurs routinely in anaerobic processes; however, excessive mixing or agitation can cause more CO_2 to strip than desired, which can cause the pH of the system to increase sharply, resulting in inadvertent precipitation of calcium and magnesium salts.[31] Finally, when anaerobic processes are used for wastewater treatment, the presence of CH_4 in the liquid phase must be considered. Given its high COD, dissolved CH_4 would need to be removed in an environmentally acceptable manner before the liquid effluent could be discharged.

8.1.4.3 Precipitation

Precipitation in anaerobic processes is complex; so much so that it was not included in ADM No. 1. Nevertheless, ignoring precipitation processes when modeling anaerobic processes can have an important impact on our ability to accurately estimate the pH and gaseous composition of a system.[4]

One of the main factors controlling precipitation is the concentration of the divalent cations present in the waste, especially Ca^{2+} and Mg^{2+}.[32] Most domestic wastewaters have low levels of these cations and ignoring precipitation for them is reasonable. On the other hand, systems that receive industrial discharges containing unusually high levels of these cations, or that use calcium- or magnesium-based salts in the treatment process may experience significant levels of precipitation. Undesirable precipitation in an anaerobic process can be extremely detrimental to the system, and the mineral struvite is among the most studied of the precipitation products. Struvite is highly insoluble (its solubility product ranges from $10^{-12.6}$ to $10^{-13.2}$)[31] and forms a very hard mineral that causes extensive damage to pipes, valves, and pumps.[8] Technologies have been developed to encourage the controlled formation of struvite so that the mineral can be recovered for use as a fertilizer,[13] and this is discussed in Chapter 23.

8.2 KINETICS OF ANAEROBIC BIOCHEMICAL OPERATIONS

As our understanding of the interactions in anaerobic processes has improved, engineers have sought to model anaerobic systems on a more fundamental level by including reaction steps for each important microbial group.[9,15] Many anaerobic processes are designed so that all the metabolic steps discussed here occur in a single reactor; therefore, the relative rates of the different metabolic processes become very important, especially as the products of some microorganisms (e.g., H_2 producing acetogens) are inhibitory to the activity of others (e.g., acetate forming fermenters). Although it is difficult to differentiate among the kinetics of multiple metabolisms in systems as complex as anaerobic bioreactors, an important first step is to develop the reaction rate expressions for the different metabolic steps because they provide information that is helpful in developing an appreciation for the kinetic characteristics of the various anaerobic microbes. Because a temperature of 35°C is commonly used for anaerobic operations, the following parameter values are for that temperature range.

8.2.1 Disintegration and Hydrolysis

It is widely agreed that disintegration follows a first-order expression as a function of the total composite particulate material concentration, X_C, and thus the rate of change of its concentration, r_{XC}, is given by

$$r_{XC} = -k_{dis} \cdot X_C, \tag{8.5}$$

where k_{dis} is the first-order disintegration rate coefficient.[4] Values for k_{dis} vary depending upon the temperature of digestion, but are between 0.4 and 0.5 d^{-1} for mesophilic systems, while disintegration in thermophilic systems is approximately twice as fast.[4] In anaerobic digesters receiving primary sludge and/or activated sludge, the rate of disintegration is believed to be the rate limiting step.[4]

Hydrolysis is more complex than disintegration. Because it is enzymatic, the actual linkage between the enzyme and the particle being hydrolyzed influences the rate at which the enzyme works. A number of models have been developed that consider the surface-mediated nature of enzyme catalyzed particle hydrolysis,[49] including forms similar to Equation 3.77. In cases where there is a high ratio of biomass to waste, as occurs with anaerobic digestion, hydrolysis simplifies to a first-order reaction rate,[48] and it is this form that is used in ADM No. 1. Because the enzymes responsible for the hydrolysis of the various biochemical components (i.e., carbohydrate, protein, or lipid) are different with unique reaction rates, rate expressions must be written for each. The generalized form for hydrolysis of a given particulate biochemical component, X_{comp}, is given by

$$r_{Xcomp} = -k_{h,comp} \cdot X_{comp}, \tag{8.6}$$

where $k_{h,comp}$ is the hydrolysis coefficient for that component. When incorporating hydrolysis using ADM No. 1, $k_{h,comp}$ is assumed to be at least an order of magnitude faster than k_{dis}, which represents the rate limiting step. This assumption minimizes the impact of hydrolysis on the overall rate at which biochemical monomers are generated for acidogenesis. However, most experimental estimates of the rate that particles breakdown into the soluble biochemical monomers reflect disintegration plus hydrolysis, even though the estimates are reported as hydrolysis rate coefficients. The values are typically in the range given for k_{dis} because disintegration is the rate limiting step in the process of forming soluble monomers. These coupled rates ($k_{h,overall}$) tell us something about the relative rates with which the different biochemical components break down. For example, carbohydrates and proteins tend to break down as much as an order of magnitude faster than lipids,[4] and carbohydrates break down a little faster than proteins.[48] Exact values are not given because estimates can vary widely and are a function of the source of waste and the hydraulic residence time over which hydrolysis occurs.[40] Estimates of $k_{h,overall}$ range from 0.15 to 0.25 d^{-1} for mesophilic digestion[37,40,45] and are around 0.4 d^{-1} for thermophilic digestion.[45]

8.2.2 Fermentation and Anaerobic Oxidation Reactions

Fermentative bacteria (group 2 in Figure 2.4) grow relatively rapidly on amino acids and simple sugars, and their kinetics can be represented by the Monod equation (Equation 3.36) with a $\hat{\mu}$ value on the order of 4 to 6 d^{-1} and a K_S value between 20 and 50 mg/L as COD.[15,45] Review of available data suggests that this reaction does not limit system performance.[15] The bacteria that oxidize long chain fatty acids (group 3 in Figure 2.4) and the bacteria that oxidize VFAs (group 4 in Figure 2.4) grow more slowly than the fermentative bacteria, and the extent to which they grow is largely dependent on the H_2 levels in the system. The values of $\hat{\mu}$ and K_S depend on the degree of saturation of the fatty acid serving as growth substrate, with saturated acids having lower $\hat{\mu}$ and K_S values than unsaturated ones.[15] Long chain fatty acid degraders are slow growers and have $\hat{\mu}$ values that range from 0.24 d^{-1} to 0.36 d^{-1}, and K_S values ranging from 400 to 500 mg/L as COD.[4,9] The same organisms are thought to be responsible for the oxidation of butyrate and valerate (reaction 4 in Figure 2.4) with $\hat{\mu}$ and K_S values of 1.2 d^{-1} and 200 mg/L as COD, respectively, thus making them the fastest growing organisms that perform anaerobic oxidation, although they are still slower than the organisms responsible for sugar and amino acid fermentation. Propionate, on the other hand, is degraded by more specialized bacteria that grow more slowly. Gujer and Zehnder[15] reported $\hat{\mu}$ and K_S values of 0.156 d^{-1} and 250 mg/L as COD, respectively, whereas Bryers[9] chose values of 0.079 d^{-1} and 800 mg/L as COD based on other studies. Although the two sets of values differ somewhat in magnitude, they both suggest that growth on propionate is much slower than growth on other fatty acids.

As discussed in Section 8.1.2, the thermodynamic feasibility of anaerobic oxidation is highly dependent on the H_2 concentration. A common way to include this dependence in the kinetics of this process is to use an expression that resembles the kinetics of noncompetitive inhibition of an enzymatic reaction:[4,6,36,45]

$$I_{H_2,kinetic} = \frac{1}{1 + \frac{S_{H_2}}{K_{I,H_2}}} \tag{8.7}$$

where $I_{H_2,kinetic}$ is a term (referred to herein as a H_2 switching function) that assumes a value of 1.0 when the H_2 concentration is equal to zero and becomes zero for H_2 concentrations much greater than the coefficient K_{I,H_2}. The maximum specific growth rate of a particular organism is multiplied by the H_2 switching function, thereby allowing it to change between $\hat{\mu}$ and zero, depending on the H_2 concentration. The values used for K_{I,H_2} should be such that as the H_2 concentration in the reactor increases to levels where anaerobic oxidation becomes thermodynamically unfeasible, then the

specific growth rate of the organisms performing anaerobic oxidation should approach zero, by making the H_2 switching function approach zero.

It is worth performing some simple calculations to find a relationship equivalent to Equation 8.7, but based solely on thermodynamic principles.[17,20,21] The equilibrium constant of a reaction is related to the free energy associated with that reaction:

$$K = \exp\left(-\frac{\Delta G^0}{RT}\right). \tag{8.8}$$

For any given reaction, not necessarily at equilibrium, we can calculate a reaction quotient (K_{RQ}). This quantity is calculated using the expression used to define the equilibrium constant, but with the actual concentrations of the species participating in the reaction. For example, consider the anaerobic oxidation of propionate to acetate (Table 8.1):

$$K_{RQ} = \frac{[\text{Acetate}][H_2]^3[CO_2]}{[\text{Propionate}]}. \tag{8.9}$$

In Table 8.1, the concentrations of the various species in an operating reactor are presented. Equation 8.9 can then be used to calculate K_{RQ} as a function of the H_2 concentration. The ratio K_{RQ}/K tells us how far away a given reaction is from its thermodynamic equilibrium. If the ratio is less than one, then, thermodynamically, the reaction can proceed further. If the ratio is equal to one, then the conditions within the reactor are exactly at thermodynamic equilibrium. A ratio larger than one is not thermodynamically possible. Equation 8.8 and the appropriate form of Equation 8.9 for each reaction described in Table 8.1, can be used to determine the ratio of K_{RQ}/K for all reactions as a function of the H_2 concentration in the reactor. This can be used to define a switch, $I_{H_2,thermo}$, which will range from zero to one, depending on the thermodynamic feasibility of the reaction under a certain set of conditions:

$$I_{H_2,thermo} = \begin{cases} 0 & \text{if} \quad \frac{K_{RQ}}{K} \geq 1 \\ \left(1 - \frac{K_{RQ}}{K}\right) & \text{if} \quad \frac{K_{RQ}}{K} < 1 \end{cases}. \tag{8.10}$$

In Figure 8.1, the behavior associated with the thermodynamic feasibility switch (Equation 8.10) is compared to the behavior of the kinetic switching function calculated with Equation 8.7 and the K_{I,H_2} values proposed in ADM No. 1: 5×10^{-3} mg/L as COD for LCFA oxidation, 3.5×10^{-3} mg/L as COD for propionate oxidation, and 1×10^{-2} mg/L as COD for butyrate oxidation. The main question we want to answer with Figure 8.1 is: Are the kinetic effects associated with Equation 8.7 consistent with the inherent thermodynamics of the system? To answer this question, first focus on the curves associated with Equation 8.10 or the thermodynamic approach. For the three substrates considered, the thermodynamic feasibility of the reactions is highly dependent on the H_2 concentration in the system, changing from feasible to infeasible over a very small range, with LCFA oxidation having the highest sensitivity. Additionally, as the H_2 concentration in the system increases, anaerobic oxidation of propionate is the first reaction to become thermodynamically limited, which explains why propionate tends to accumulate first in systems as the H_2 concentration increases. The next reaction to become infeasible is LCFA oxidation followed by butyrate oxidation.

When comparing the kinetic and thermodynamic approaches to modeling the effect of the H_2 concentration on anaerobic oxidation, several additional points are apparent in Figure 8.1. First, both switches start decreasing in value around the same H_2 concentration, which means that the

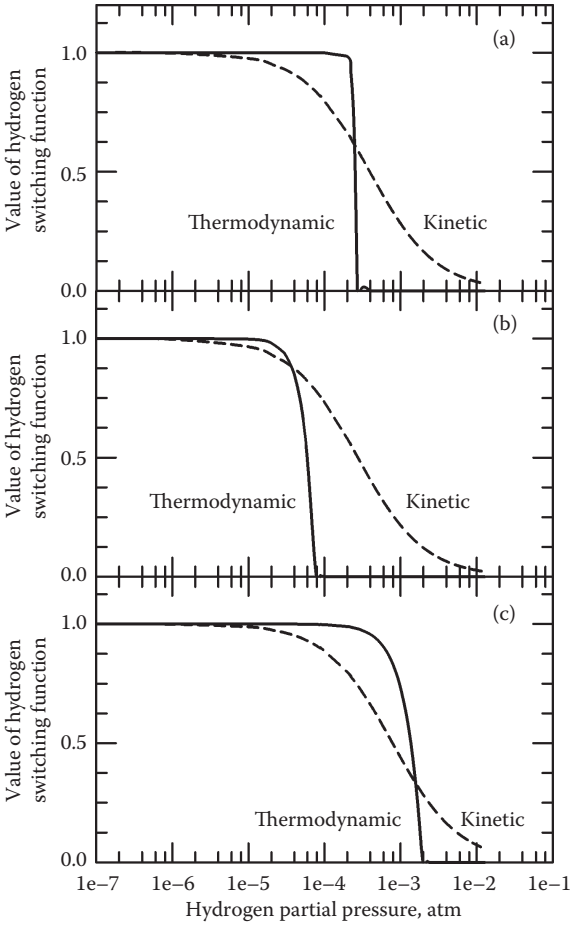

FIGURE 8.1 Effect of the H_2 partial pressure on the value of the hydrogen switch calculated using a kinetic approach or a thermodynamic approach. This switch modulates the maximum specific growth rate of the organisms performing anaerobic oxidation of (a) long chain fatty acids, (b) propionate, and (c) butyrate as the H_2 concentration varies. The conditions used were as described in Table 8.1, except for the H_2 partial pressure, which was varied as shown.

selection of the values of K_{I,H_2} was done in a way that reflects the fact that the system is approaching a state of thermodynamic limitation. This is a key aspect of anaerobic systems. Second, although the kinetic approach results in the correct on-set of the limitation, because of the nature of the kinetic switching function anaerobic oxidation is not turned off, even when the H_2 concentration attains values at which the reaction is thermodynamically infeasible. For steady-state conditions, this will likely not be a major concern, but it may be under dynamic conditions. It is possible that during dynamic simulations, processes that are thermodynamically infeasible remain active, which is not possible in reality. Nevertheless, the main points that should be taken from Figure 8.1 are that the maximum rate of anaerobic oxidation depends on the H_2 concentration and that this effect can be described using either a kinetic or thermodynamic approach.

Two other factors affect the maximum specific growth rates of the organisms performing anaerobic oxidation and these are pH and the free ammonia concentration. In Chapter 2 we discussed the importance of pH on the growth of microorganisms. This is particularly significant in anaerobic systems where the formation of large quantities of acids may cause large perturbations in the pH of the system and affect the delicate balance of the multiple organisms involved. The organisms

responsible for anaerobic oxidation are able to grow close to their maximum specific growth rate when the pH is higher than 5.5–6.5,[4,45] but they are far less sensitive than methanogens, especially aceticlastic methanogens.

8.2.3 Methanogenesis

Aceticlastic methanogenesis (reaction 6 in Figure 2.4) is a very important reaction in anaerobic operations because it produces about 70% of the methane. Two major types of aceticlastic methanogens can be present in anaerobic systems, but the one that predominates depends on the bioreactor conditions imposed because their growth kinetics are quite different. *Methanosarcina* can grow rapidly, but do not have a high affinity for acetate. Representative parameter values for them are 0.34 d^{-1} for $\hat{\mu}$ and 40 to 300 mg/L as COD of acetate for K_S.[4,44] *Methanosaeta* (formerly *Methanothrix*), on the other hand, grow more slowly, but have a higher affinity for acetate, as shown by a $\hat{\mu}$ value of 0.072 d^{-1} and a K_S value of 30 to 40 mg/L as COD of acetate.[44] Indeed, higher concentrations of acetate in anaerobic digesters tend to select for *Methanosarcina* over *Methanosaeta*.[25] Although parameters have been determined for these two types of aceticlastic methanogens, anaerobic digestion models like ADM No. 1 do not differentiate between them. Instead, typical values for the aceticlastic methanogens as a whole are reported to be 0.37 d^{-1} for $\hat{\mu}$ and 40 mg/L as COD of acetate for K_S.[45] Finally, the H_2 oxidizing methanogens produce methane from H_2, thereby keeping the H_2 concentration low and allowing the H_2 producing reaction to proceed as discussed in Section 2.3.2. The kinetic parameters for their growth have been reported to be $\hat{\mu} = 1.4$ to 2.0 d^{-1} and $K_S = 0.001$ to 0.6 mg/L as COD of dissolved H_2.[15,45,51]

Methanogens are vulnerable to inhibition caused by pH and free ammonia.[4] The aceticlastic methanogens become completely inhibited if the pH drops below 6, but can grow uninhibited at pH values greater than 7. The H_2 utilizing methanogens become completely inhibited if the pH drops below 5 but can grow uninhibited at pH values greater than 6. Furthermore, the aceticlastic methanogens are vulnerable to free ammonia inhibition. A noncompetitive model of the same form as shown in Equation 8.7 is used in ADM No. 1 to address this inhibition, with an inhibition coefficient estimated to be 30 mg/L.

8.2.4 Maintenance, Endogenous Metabolism, Decay, Lysis, and Death

Decay also occurs in anaerobic systems, but the values of the decay coefficients for such systems are lower than those for aerobic systems because the bacteria have much lower $\hat{\mu}$ values and the two parameters appear to be correlated. For example, Bryers[9] has reported decay coefficient values around 0.0096 d^{-1} for bacteria carrying out anaerobic oxidations and methanogenesis and values around 0.024 d^{-1} for fermentative bacteria. The IWA task group that developed ADM No. 1 did not differentiate among the different groups responsible for anaerobic digestion when estimating decay coefficients and used 0.02 d^{-1} for all of them.

8.2.5 Inhibition Factors in Anaerobic Biochemical Operations

The pH of an anaerobic system has a strong impact on $\hat{\mu}$, with an optimum around pH 7. Just as with nitrifying bacteria, this is in part because the nonionized form of the substrate (fatty acids in this case) serves as the actual substrate for growth and the amount of nonionized form will depend on the pH. As a consequence, relationships between $\hat{\mu}$ and pH are needed for the major groups of bacteria. On the other hand, some[1,14] have modeled acetic acid utilization with the Andrews equation using nonionized acetic acid as the substrate in combination with an expression for the ionization of acetic acid as a function of pH. It should be noted that this is an alternative approach to making $\hat{\mu}$ an explicit function of pH and the two should not both be used. In addition, the role of H_2 in regulating the utilization of propionic and butyric acids and the activity of the H_2 producing bacteria is

very important, but is not reflected in the parameter values reported above, which are all for low H_2 levels. Bryers[9] has argued that the H_2 effect is based on thermodynamics, and as such, does not translate directly into kinetic expressions, although we saw in Section 8.2.2 how the impacts of H_2 can be handled. Labib et al.,[23] on the other hand, have demonstrated inhibition of butyric acid utilization by H_2 separate from the thermodynamic effects. Thus, even though information on the kinetic impacts of H_2 is very limited, it is important, suggesting that additional studies are needed to allow development of appropriate rate expressions. In spite of these limitations, however, the kinetic parameters above provide a good sense of the relative capabilities of the microorganisms involved in anaerobic operations.

8.2.6 Effects of Temperature on Kinetic Parameters

Temperature is also known to play an important role in anaerobic operations. Most studies, however, have looked at overall system performance rather than at the impact on each of the groups of microorganisms discussed in Section 2.3.2. For example, Henze and Harremoës[16] combined data from seven studies to estimate the temperature coefficient for methanogenesis and found u to be 66.7 kJ/mole ($C = 0.10\ °C^{-1}$, $\theta = 1.105$) for a temperature range of 10 to 30°C. The methane production rate was constant from 30 to 40°C, and decreased for higher temperatures. Characklis and Gujer[10] used data from the literature to estimate that the value of u associated with K_S for acetic acid was −132.9 kJ/mole ($C = -0.199\ °C^{-1}$, $\theta = 0.819$), showing that K_S decreases as the temperature is increased for this process.

8.3 ANAEROBIC DIGESTION MODEL NO. 1

8.3.1 Components of Anaerobic Digestion Model No. 1

Anaerobic systems are very complex with multiple organisms working in concert and, in some cases, performing reactions that are close to thermodynamic equilibrium. Products formed within the reactor may cause growth inhibition and because various acids are produced in the reactor, large pH fluctuations are possible, which will also affect the growth rates of some microorganisms. The complexity of this system is best handled though the use of a mathematical model. The IWA Task Group for Mathematical Modeling of Anaerobic Digestion Processes developed such a model, ADM No. 1.[4] The model describes 19 biochemical processes, 3 processes for mass transfer between the liquid and gas phases, and 9 implicit algebraic variables. As with any mathematical model, ADM No. 1 is not perfect and several suggestions have been made to improve it.[3,21,41]

In its present form, ADM No. 1 has been successfully used to simulate a growing number of real systems[2,7,19] and the model captures the main events taking place in anaerobic systems, such as those described in Section 2.3.3 and the previous sections of this chapter. Because of the complexity of anaerobic systems, the main purpose of using a model here is to help us integrate all of the events taking place and to understand how they contribute to the overall performance of the system. We will consider the operation of an anaerobic digester treating a mixture of primary sludge and waste activated sludge.

8.3.2 Simulating the Anaerobic Digestion of Primary and Waste Activated Sludge

The anaerobic digester considered herein is the same as that used by the IWA Task Group on Benchmarking of Control Strategies for Wastewater Treatment Plants for the benchmark system model No.2,[18,35] also known as BSM No. 2. The liquid volume in the digester is 3400 m^3 and the headspace volume is 300 m^3. The digester is fed a mixture of primary and waste activated sludge at a flow rate of 170 m^3/day, resulting in a solids retention time (SRT) of 20 days for the base case. The feed composition is given in Table 8.2 and these values have been reported as the composition

TABLE 8.2
Feed Composition to the Anaerobic Digester

Component	Concentration
Monosaccharides	0.01 kg/m^3 as COD
Amino acids	0.001 kg/m^3 as COD
Total long chain fatty acids	0.001 kg/m^3 as COD
Total valerate	0.001 kg/m^3 as COD
Total butyrate	0.001 kg/m^3 as COD
Total propionate	0.001 kg/m^3 as COD
Total acetate	0.001 kg/m^3 as COD
Soluble hydrogen	10^{-8} kg/m^3 as COD
Soluble methane	10^{-5} kg/m^3 as COD
Inorganic carbon	0.04 kmole/m^3 as C
Inorganic nitrogen	0.01 kmole/m^3 as N
Soluble organic inerts	0.02 kg/m^3 as COD
Particulate composite organics	2.0 kg/m^3 as COD
Carbohydrates	5.0 kg/m^3 as COD
Proteins	20.0 kg/m^3 as COD
Lipids	5.0 kg/m^3 as COD
Sugar degraders (biomass)	0.0 kg/m^3 as COD
Amino acid degraders (biomass)	0.01 kg/m^3 as COD
Total long chain fatty acids degraders (biomass)	0.01 kg/m^3 as COD
Valerate and butyrate degraders (biomass)	0.01 kg/m^3 as COD
Propionate degraders (biomass)	0.01 kg/m^3 as COD
Acetate degraders (biomass)	0.01 kg/m^3 as COD
Hydrogen degraders (biomass)	0.01 kg/m^3 as COD
Particulate organic inert material	25.0 kg/m^3 as COD
Cations	0.04 kmole/m^3
Anions	0.02 kmole/m^3

Note: Data from Rosen, C. and Jeppsson, U., *Aspects on ADM1 Implementation within the BSM2 Framework,* University of Lund, Lund, Sweden, 2006. http://www.iea.lth.se/publications/Reports/LTH-IEA-7224.pdf.

of a mixture of primary and waste activated sludge generated using BSM No. 2.[43] The simulations were done using ADM No. 1,[4,5] with the default kinetic and stoichiometric parameters and an operating temperature of 35°C. The model was implemented[42] in MATLAB® and simulations where performed by forward integration of the equations in the model for 300 days, which ensured that a steady-state solution was found. Anaerobic digesters are usually operated as CSTRs, where the hydraulic residence time is equal to the SRT (Θ_c). To simulate the operation of the digester at various SRTs, the volume of the digester (V) was varied according to $V = F \cdot \Theta_c$, with the flow rate (F) being maintained constant at 170 m^3/day for all SRTs considered. Figure 8.2 shows the concentration of various constituents in the digester as a function of the SRT.

Consider first the behavior of the system when operated at SRTs below five days, where methanogenesis is not yet fully established. Disintegration and hydrolysis of composite particulate material, complex carbohydrates, lipids, and proteins occur when the system is operated at fairly low SRTs (around one day, Figure 8.2a). Monosaccharides and amino acids, either present in the feed

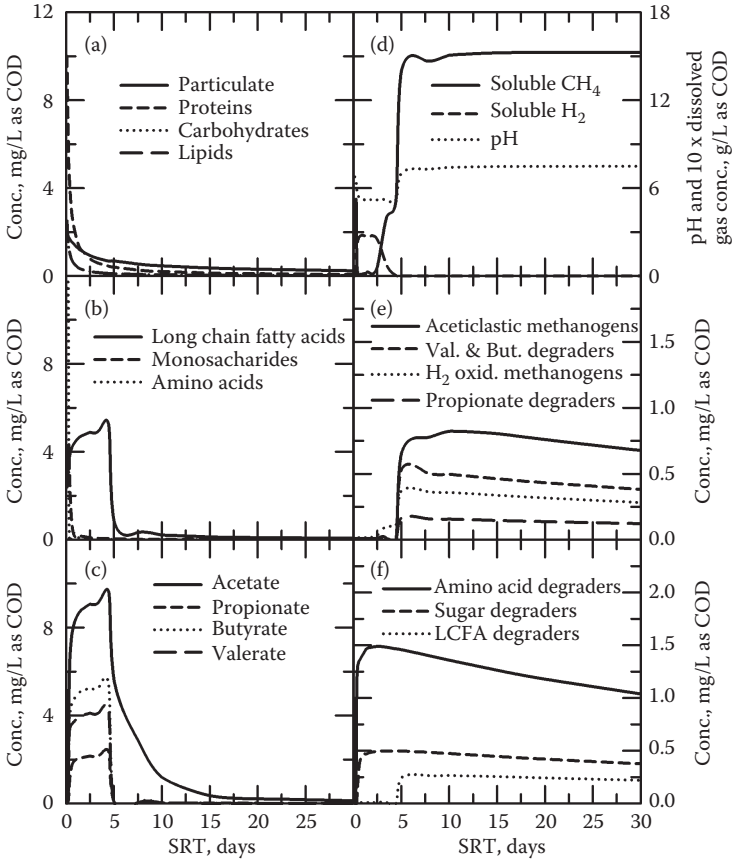

FIGURE 8.2 Effect of SRT on the steady-state concentration of various constituents in an anaerobic digester. The flow rate into the digester was constant (170 m^3/day) and the volume of the reactor was changed to achieve a given SRT ($V = F \cdot \Theta_c$). The influent concentrations are given in Table 8.2.

or formed though hydrolysis in the reactor, are rapidly fermented to the VFAs acetate, propionate, butyrate, and valerate (Figure 8.2b and c), which result in a decrease of the pH in the reactor (Figure 8.2d). It should be noted that in ADM No. 1, the stoichiometry describing the distribution of VFAs formed through fermentation is fixed, while in reality this distribution is dependent on the environmental conditions in the reactor.[28,41] The operation of an anaerobic reactor at these short SRTs is used in fermenters at some wastewater treatment plants where biological phosphorus removal is required, since it allows the conversion of complex organic materials into VFAs, the substrates used by phosphate accumulating organisms. An important aspect of fermentation is that the COD initially present in the feed is mainly maintained within the liquid phase, a desirable feature if maximizing VFA production is the main objective. During fermentation, a small amount of H_2 is generated and its concentration in the liquid phase starts increasing (Figure 8.2d). Lipids are also hydrolyzed, mostly (99%) to LCFAs and a small amount of monosaccharides. As shown in Figure 8.2b, LCFAs accumulate in the digester when the SRT is low (less than five days). The reason for this is that the bacteria responsible for oxidation of LCFAs to H_2 and acetate grow more slowly than fermenting bacteria, and the oxidation reaction is largely inhibited by H_2 (thermodynamically close to equilibrium) as well as by lower pH values.

At an SRT of about 2.5 days, H_2 oxidizing methanogens start growing in the system (Figure 8.2e). This results in a decrease in the soluble H_2 concentration and the accumulation of methane in the system (Figure 8.2d), but the H_2 concentration is still not low enough to allow anaerobic

oxidation to take place. As the operating SRT approaches five days, a variety of events occur in the reactor, shifting it from acidogenesis to methanogenesis and allowing anaerobic oxidation to take place. At that SRT, aceticlastic methanogens can grow in the system (Figure 8.2e), splitting acetate into methane and carbon dioxide, which causes the acetate concentration in the bioreactor to drop. In addition, there is also a large enough population of H_2 oxidizing methanogens to decrease the H_2 concentration sufficiently to allow anaerobic oxidation of the LCFAs and VFAs (Figure 8.2b and f) to acetate and H_2. This additional H_2 is used by the H_2 oxidizing methanogens, preventing the H_2 concentration from rising and blocking anaerobic oxidation. The pH of the system increases (Figure 8.2d) because the VFAs and LCFAs are oxidized to acetate, which is consumed by the aceticlastic methanogens.

It is interesting to see how the concentration of acetate changes as a function of SRT (Figure 8.2c). It increases in the reactor at low SRTs due to fermentation and then starts dropping at SRTs higher than five days, although this drop occurs in a different manner than that of the other VFAs in the system (Figure 8.2c). One reason for this is that acetate is not only a product of fermentation, but also a product of anaerobic oxidation, which is not the case for the other VFAs. In fact, the other VFAs are oxidized to acetate and H_2. The overall concentration of acetate in the reactor is a result of processes forming it (fermentation and anaerobic oxidation) and processes consuming it; namely, the growth of aceticlastic methanogens, which are slow growing organisms (as mentioned in Section 8.2.3 and as shown in Figure 8.2f).

Figures 8.2e and 8.2f are particularly useful for following the transformations taking place in the reactor as the SRT changes. Starting with low SRTs, the reactor goes through a fermentation phase, with the growth of amino acid and sugar degraders, followed by the appearance of H_2 oxidizing methanogens at an SRT of about 2.5 days, and moving finally to the growth of aceticlastic methanogens at an SRT of 5 days. Anaerobic oxidation also starts at an SRT around 5 days, as seen by the appearance of LCFA degraders, propionate degraders, and butyrate/valerate degraders.

Figure 8.3 shows the amount and composition of the gas produced in the reactor, which follows closely the events described in Figure 8.2. A small amount of H_2 is produced at low SRTs (less than one day) by fermentative reactions. Up to SRTs of 2.5 days, H_2 production occurs without H_2 oxidizing methanogens growing in the system. This clearly suggests a great potential role that anaerobic systems can play in the production of clean fuels. The accumulation of H_2 in the reactor, although desirable from the perspective of energy production, thermodynamically limits the extent of anaerobic oxidation. This results in only a small fraction of the electrons present in the feed ending up in H_2, with most of the electrons being associated with fairly reduced organic molecules (VFAs and

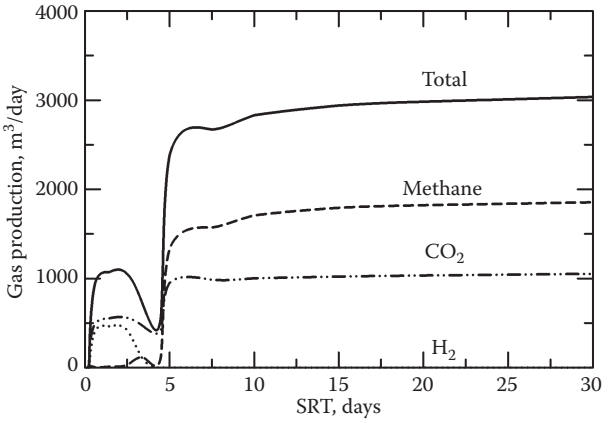

FIGURE 8.3 Effect of SRT on the volume and composition of gas produced in an anaerobic digester. The flow rate into the digester was constant (170 m^3/day) and the volume of the reactor was changed to achieve a given SRT ($V = F \cdot \Theta_c$). The influent concentrations are given in Table 8.2.

LCFAs). For maximum levels of energy production, H_2 should be removed from the system as fast as it is generated . Sparging the headspace of anaerobic reactors with inert gases has been successfully used to improve the yield of H_2 from anaerobic systems.[22,27] When the SRT approaches 2.5 days, the H_2 content of the gas starts dropping and methane starts being formed. As the SRT is increased further, aceticlastic methanogens start growing in the system, which, along with the occurrence of anaerobic oxidation, leads to a large degree of conversion of the COD in the influent to methane. These two processes result in a large increase in the volume of gas produced in the system, which contains virtually no H_2, but mainly methane and carbon dioxide.

An alternative way to look at the behavior of an anaerobic digester is to examine how the COD entering in the feed is converted and distributed between the gas and liquid phases (this includes soluble and particulate material as well as active biomass). From Figure 8.4a, we can see that when the SRT is lower than five days, most of the COD associated with the feed remains in the liquid phase, mainly in the form of VFAs and active biomass. It is necessary for aceticlastic methanogens to grow in the system to allow significant COD removal from the liquid phase through formation of methane.

The main objective of anaerobic digestion is to stabilize biosolids (primary and waste activated sludge), so it is worthwhile to examine how the volatile suspended solids (VSS) concentration changes in the digester as a function of SRT (Figure 8.4b). At very short SRTs (less than two days), the VSS in the reactor is substantially lower than in the feed, because disintegration and hydrolysis have relatively fast kinetics compared to the growth of aceticlastic methanogens. When the SRT approaches five days, several organisms are able to grow in the digester (as shown in Figure 8.2e and f) and, hence, a small increase in the VSS concentration occurs. With longer SRTs, the VSS concentration in the reactor approaches the concentration of the nonbiodegradable VSS in the feed, which provides an ultimate limit on the level of VSS destruction that can be achieved.

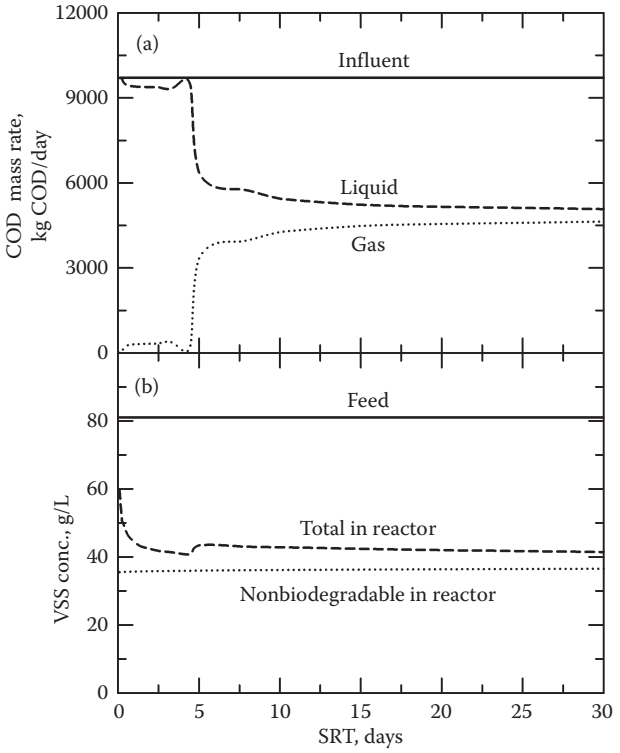

FIGURE 8.4 Effect of SRT on the fate of the COD in the feed and on the VSS destruction in an anaerobic digester. The flow rate into the digester was constant (170 m^3/day) and the volume of the reactor was changed to achieve a given SRT ($V = F \cdot \Theta_c$). The influent concentrations are given in Table 8.2.

As we will see in Chapter 15, the potential extent of VSS destruction that can be achieved in anaerobic digesters depends on the nature of the feed. Larger degrees of VSS destruction can be achieved in digesters treating only primary solids compared to digesters treating a mixture of primary solids and waste activated sludge. The reason for this is that because of decay of active biomass and accumulation of inert particulate organic material, waste activated sludge can contain a significant fraction of nonbiodegradable VSS. Furthermore, that fraction increases as the SRT of the activated sludge system is increased. Therefore, it follows that the maximum VSS destruction efficiency that can be achieved in an anaerobic digester receiving waste activated sludge is related to the SRT at which the activated sludge system is being operated, with lower levels of maximum potential destruction being observed in plants where the activated sludge system is operated at high SRT. An important point that results from inspection of Figure 8.4b is that the concentration of nonbiodegradable VSS in an anaerobic digester is virtually independent of the SRT at which the digester is operated. There are two reasons for this. The first is that true growth yield in anaerobic systems is very low, which will result in only a small amount of biomass being formed as organic material is removed. The second is that the decay rates in anaerobic systems are low, which leads to little generation of biomass debris.

As particulate material is disintegrated and hydrolyzed in an anaerobic digester, the nitrogen associated with the organic material (primarily protein) is released to the liquid phase. Because of the low growth yield under anaerobic conditions, little nitrogen is used for biomass synthesis and thus most ammonia remains in solution at very high concentrations, as shown in Figure 8.5. The presence of ammonia at such high levels is important for two reasons. The first is that free (nonionized) ammonia inhibits the growth of aceticlastic methanogens, which are essential for stable operation of anaerobic digesters. Luckily, as shown in Figure 8.5, the majority of the ammonia is present in the ionized form, because the pH is either slightly acidic or close to neutral. The second reason the presence of high ammonia concentrations is important concerns wastewater treatment systems that are required to produce nitrified effluents. After anaerobic digestion, dewatering is typically used to minimize the volume of solids removed to ultimate disposal. During dewatering, solids are concentrated by removing water, which in the case of anaerobically digested biosolids, have a very high concentration of ammonia. This ammonia-laden stream is usually recycled back to the liquid treatment portion of the plant, where it will be treated in an activated sludge bioreactor. If that bioreactor is operated at an SRT that allows nitrifiers to grow, the recycle stream can substantially increase its oxygen requirements because of the additional load of ammonia. Failure to account for

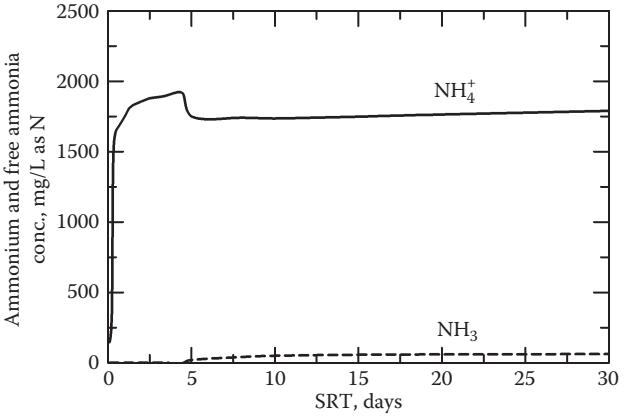

FIGURE 8.5 The effect of SRT on the ammonia (free and ionized) concentrations in an anaerobic digester. The flow rate into the digester was constant (170 m^3/day) and the volume of the reactor was changed to achieve a given SRT ($V = F \cdot \Theta_c$). The influent concentrations are given in Table 8.2.

this additional load when the aeration system is sized could result in a plant that is unable to meet its required effluent quality.

It should be noted that ADM No. 1 and the results shown in Figure 8.5 do not include the formation of precipitates, namely struvite, which can occur in real systems and will lead to a decrease in the concentration of soluble ammonia in the digester.

8.4 KEY POINTS

1. In anaerobic oxidation, volatile fatty acids (VFAs) and long chain fatty acids (LCFAs) are oxidized to acetic acid and H_2, the two substrates used in methanogenesis.
2. The microorganisms responsible for anaerobic oxidation have slower growth kinetics than the microorganisms responsible for fermentation.
3. Anaerobic oxidation occurs close to thermodynamic equilibrium and the thermodynamic feasibility of these reactions is highly dependent on the H_2 concentration.
4. As the H_2 concentration in anaerobic systems increases, anaerobic oxidation of propionate is the first process to become thermodynamically infeasible, followed by oxidation of LCFAs and finally oxidation of butyrate.
5. Mathematical models used to describe the behavior of anaerobic systems must consider the effects of H_2 on anaerobic oxidation. This can be done using a kinetic approach, but the parameters selected must be consistent with the thermodynamic basis of the reactions.
6. Anaerobic systems operated at solids retention times (SRTs) less than five days achieve extensive hydrolysis of particulate complex materials and fermentation of the hydrolysis products into VFAs and LCFAs. The majority if the chemical oxygen demand (COD) entering the reactor is maintained in the liquid phase.
7. A small amount of methane is produced in anaerobic reactors operated at SRTs lower than five days, due to the growth of H_2 oxidizing methanogens.
8. Long chain fatty acids accumulate in anaerobic reactors operated at SRTs lower than five days because H_2 accumulation inhibits anaerobic oxidation to H_2 and acetate.
9. The growth of aceticlastic methanogens causes the acetate concentration in the digester to drop, which leads to an increase in the pH. An increase in pH, along with a lower H_2 concentration in the reactor, allows anaerobic oxidation to take place.
10. Without anaerobic oxidation and fully established methanogenesis, little COD is removed in an anaerobic digester. These two processes start occurring when the SRT is larger than five days.
11. The composition of the feed to an anaerobic digester, more specifically the fraction of inert nonbiodegradable organic material, dictates the maximum fraction of volatile suspended solids (VSS) destruction that can be attained across the digester.
12. The ammonia concentration in anaerobic digesters is very high (several hundreds of mg/L as N). Nonionized free ammonia inhibits the growth of aceticlastic methanogens, which are essential for the stable operation of anaerobic digesters.

8.5 STUDY QUESTIONS

1. Explain the difference between disintegration and hydrolysis. In doing so, describe the status of the feed at various points in both processes by defining composite particulate material, biochemical components, and soluble monomers.
2. Describe the major groups of microorganisms participating in anaerobic operations and contrast their growth characteristics as described by their kinetic parameters.
3. Compare the yields of the different microorganisms involved in anaerobic operations and discuss the impact of their differences on the composition of the microbial community in an anaerobic digester.

4. Assume that the wasted biomass from Study Question 6d in Chapter 5 is sent to a mesophilic anaerobic digester. Estimate X_C entering the digester. Based on the typical biochemical composition of cells, what fraction of the COD entering the digester will be associated with monosaccharides, amino acids, long chain fatty acids, and inert material?
5. Explain the significance of the fact that the bacteria responsible for fermentative reactions are able to grow faster than the bacteria responsible for anaerobic oxidation. Focus your explanation in terms of the performance of anaerobic systems.
6. Explain the importance of the H_2 concentration in anaerobic oxidation.
7. In mathematical models for anaerobic systems, the effect of H_2 concentration on anaerobic oxidation is usually described using an expression that resembles the kinetics of noncompetitive inhibition. Explain how the selection of the inhibition coefficient is related to the thermodynamics of anaerobic oxidation.
8. Draw a sketch showing how the acetate concentration in an anaerobic digester changes as a function of the SRT and explain why the curve has the shape that it does.
9. Draw a sketch showing how the H_2 and methane concentrations in an anaerobic digester change as a function of the SRT and explain why the curves have the shapes that they do.
10. Draw a sketch showing how the LCFA concentration in an anaerobic digester changes as a function of the SRT and explain why the curve has the shape that it does.
11. Draw a sketch showing how the concentration of the following organisms present in an anaerobic digester change as a function of the SRT: sugar degraders, amino acid degraders, H_2 oxidizing methanogens, valerate and butyrate degraders, propionate degraders, LCFA degraders, and aceticlastic methanogens.
12. Draw a sketch and explain why the COD present in the feed to an anaerobic digester splits between the liquid and gas phases as a function of the SRT.
13. Explain why the limit on VSS destruction that can be ultimately attained in an anaerobic digester in a wastewater treatment plant depends on the SRT at which the activated sludge system in the plant is operated.
14. Explain the significance of the existence of high levels of ammonia in anaerobic digesters.

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9 Techniques for Evaluating Kinetic and Stoichiometric Parameters

In the preceding chapters of Part II, we have examined models for characterizing the performance of ideal suspended growth bioreactors. Before those models can be used for design and evaluation of wastewater treatment systems, however, values must be available for the kinetic and stoichiometric parameters in them. Some of those values may be obtained from the literature or from experience with the particular wastewater to be treated. Generally, however, parameters must be evaluated experimentally during treatability studies. In this chapter we will review procedures for evaluating the parameters in the models presented in Chapters 5–7.

9.1 TREATABILITY STUDIES

Treatability studies must be carefully designed. That is because biological reactors are subject to their own version of the Heisenberg uncertainty principle. Biological wastewater treatment systems all use mixed microbial communities, which involve complex interactions among their members. In other words, bioreactors contain their own ecosystems. As a consequence, the physical characteristics of the treatment system and the manner in which it is operated will influence the composition of the community that develops within it, as well as the physiological state of the microorganisms in that community. The nature of the community and the state of the microorganisms in it, in turn, will define the kinetics exhibited by it, which will determine its performance. So, when we set up a bench-scale bioreactor in the laboratory, seed it with biomass from a suitable source, and begin to feed it with a wastewater, we are, to a large degree, predetermining the outcome of the study, although we may have no idea beforehand just what that outcome will be. This is not to suggest that treatability studies are a waste of time. Quite to the contrary. They are essential to successful design, particularly for unique wastewaters. This follows from the fact that the nature of the wastewater has the primary impact on community composition, with bioreactor configuration playing a secondary, but important, role.

Because of the impact of the nature of the bioreactor system on the composition of the microbial community that develops, it is necessary to approach the study of complex systems in a staged manner. Generally, initial studies are done with the simplest possible system. If the focus is only on oxidation of carbon and/or nitrogen, then the system will usually contain a single continuous stirred tank reactor (CSTR). However, if denitrification is to be considered along with carbon and nitrogen oxidation, then a modified Ludzak-Ettinger (MLE) system consisting of two CSTRs in series must be used. Finally, the consideration of biological phosphorus removal requires two to three CSTRs in series, depending upon whether nitrification will also occur. Regardless of the bioreactor configuration, if gravity sedimentation or membrane separation is to be used for biomass recycle in the full-scale system, then it should be employed during the treatability study because its presence will influence the nature of the microbial community present. Both gravity sedimentation and membrane separation require the presence of floc-forming bacteria for proper performance, and thus it is essential that such bacteria be present in the microbial community used to develop the kinetic and stoichiometric data from which the system will be designed. In addition, the growth of biomass in

floc particles can influence the apparent values of some parameters because of mass transfer limitations within the floc.[9]

The simplest system should be run at a number of solids retention times (SRTs) while appropriate performance data are collected. Generally, it is desirable to operate at a minimum of four different SRT values. Because of the importance of SRT in determining both the nature of the microorganisms present and their physiological state, the bioreactors should be run for an extended period of time before data collection begins. Usually, the time is expressed in terms of SRT, although what constitutes a sufficient number of SRTs is subject to debate. Because mixed microbial communities are dynamic,[3,29] the goal is to achieve functional stability rather than community stability; that is, we want to ensure that measured variables such as the mixed liquor suspended solids (MLSS) concentration, effluent quality, and so on are stable and exhibit no trends over time. It is generally accepted that three SRTs is the minimum time required to achieve functional stability, and so most treatability studies are operated for three SRTs before sampling is begun. Strategies have been proposed for assessing stability once sampling has begun.[50] Sampling should continue long enough to determine whether the system is functionally stable. This normally requires several SRTs, with many studies using two as a practical lower limit, although four would be preferable. When a treatability study involves assays for the determination of biodegradation kinetics for individual organic substrates, sufficient assays should be performed during the sampling period to allow the variability in the parameter values to be assessed and incorporated into the design uncertainty.[3] Because of the necessity to operate at several SRTs for extended periods of time, most studies run several systems simultaneously.

During operation, it is important that tight control be maintained over SRT. This means that computation of the SRT, Θ_c (or the wastage rate, F_w, required to maintain a desired SRT), must consider the biomass lost in the effluent from the settler, $X_{M,T,e}$:

$$\Theta_c = \frac{V \cdot X_{M,T}}{F_w X_{M,T,w} + (F - F_w) X_{M,T,e}}, \tag{9.1}$$

where $X_{M,T}$ and $X_{M,T,w}$ are the MLSS concentrations in the bioreactor and wastage stream, respectively, V is the bioreactor volume, and F is the influent flow rate. For simplicity, most bench-scale bioreactors are operated with the Garrett flow scheme with wastage directly from the bioreactor. It is generally not feasible to waste continually from a bench-scale bioreactor because of the small flows involved. However, wastage should be done a sufficient number of times per day to limit the amount of biomass removed each time to no more than 5% of the total biomass in the system.

After the data have been analyzed and initial estimates have been obtained for the parameter values, simulation studies should be run with an appropriate model to allow investigation of alternative bioreactor configurations. From those studies, the systems considered to be the most likely to meet the effluent criteria can be chosen and subjected to preliminary engineering analysis. One or more of those systems should then be chosen for further bench- or pilot-scale testing to verify the selection and tune the model parameters for the biomass that develops in the chosen system.

Throughout the entire staged approach to treatability testing, care should be taken to ensure that conditions are optimal for microbial growth. Bioreactors cannot be expected to perform satisfactorily if environmental conditions are inadequate. For example, no inorganic nutrients should be limiting. Guidelines are given in Section 3.8.2 to help in the determination of the proper amounts. In addition, sufficient oxygen should be provided to aerobic bioreactors so that it is not rate limiting and the pH should be maintained near neutrality. Finally, the temperature should be maintained as constant as possible because the kinetic and stoichiometric parameters will be unique to that temperature. If large seasonal temperature fluctuations are expected in the final facility, then studies should be conducted at the extreme values to allow correction of the parameter values for temperature effects using one of the approaches presented in Section 3.9.1.

9.2 SIMPLE SOLUBLE SUBSTRATE MODEL WITH TRADITIONAL DECAY AS PRESENTED IN CHAPTER 5

This model should be applied only to wastewaters without significant quantities of particulate organic matter for which the primary focus is on oxidation of organic matter, measured as chemical oxygen demand (COD). It can also be applied to situations in which nitrification is an objective. The focus here will be on carbon oxidation, but the reader can extend the principles presented to nitrification, in which case the substrate would be ammonia-N rather than soluble, biodegradable organic matter. Generally, particulate organic matter is operationally defined as the material that will be retained on a 0.45 μm pore size filter. Many colloidal sized particles will pass such a filter and thus, in a strict sense, "soluble" organic matter may not all be truly soluble. Nevertheless, for purposes of parameter estimation it is generally acceptable to apply the model of Chapter 5 to any wastewater in which the organic matter will all pass such a filter.

9.2.1 Data to Be Collected

For this application the test bioreactors should be simple CSTRs with biomass recycle. They should be operated at a number of SRTs and the following data should be collected during the steady-state period following stabilization. (It should be noted that the presence of a subscript O in a symbol denotes its concentration in the influent.)

S_{CO}	=	Soluble COD in the influent (mg/L)
S_C	=	Soluble COD in the bioreactor (mg/L)
$X_{total,T}$	=	Total biomass in the bioreactor in total suspended solids (TSS) units (mg/L)
$X_{total,Tw}$	=	Total biomass in the waste solids in TSS units (mg/L) (This will be the same as $X_{total,T}$ if the Garrett flow scheme is used.)
$X_{total,Te}$	=	Total biomass in the final effluent in TSS units (mg/L)
f_A	=	Active fraction of biomass
V	=	Reactor volume under aeration (L)
F	=	Influent flow rate (L/h)
F_w	=	Waste solids flow rate (L/h)

Several points should be noted about the data to be collected. Because the biomass data are in TSS units, the yield, $Y_{H,T}$, will have units of TSS formed per unit of substrate COD removed. Biomass data could also be collected in either volatile suspended solids (VSS) or COD units, in which case the yield values obtained would have similar units. When the yield is expressed in either TSS or VSS units, then knowledge is also needed of the COD/TSS or COD/VSS conversion factor to allow COD balances to be made. The values in Equations 5.12 and 5.13 can be assumed or COD tests can be run on biomass samples to experimentally determine the values. The active fraction of the biomass, f_A, is the most difficult data to collect during treatability studies. As a consequence, most studies do not try to measure it. In Section 9.3 we will examine how to estimate parameters in the absence of such data. A number of techniques have been proposed for measuring the active fraction, but all are tedious and subject to error. Consequently, they are used mostly in a research setting. The most direct method is the slide culture technique of Postgate,[46] which involves plating bacteria on microscope slides and observing the fraction that divide. Its main weakness, when used with mixed microbial communities, is that not all of the bacteria will necessarily be able to grow under the conditions on the slide. Fluorescent stains have been used to differentiate between actively growing and less active bacteria, but questions have been raised about the accuracy of the technique.[5] An indirect method involves quantifying the amount of adenosine triphosphate (ATP) present per unit of

biomass. It has been used successfully because the amount of ATP per viable cell is relatively independent of SRT and ATP is quickly lost from nonviable cells.[44,52] Another indirect method involves the measurement of the amount of deoxyribonucleic acid (DNA) present per unit of biomass. Like ATP, it is relatively independent of the SRT[39] and is quickly degraded when cells die and lyse.[49]

The data collected during the treatability study will be used to estimate the values of $\hat{\mu}_H$, K_S, $Y_{H,T}$, b_H, and f_D. In the process of doing this we will also have to estimate the soluble inert COD, S_I. Because many of the equations describing the performance of a CSTR can be reduced to linear form, graphical procedures have commonly been used to estimate the parameters. Linear transformations usually change the structure of the error in a data set, and thus nonlinear parameter estimation techniques are preferable whenever possible. However, an explanation of them is beyond the scope of this book. Because the linear techniques are commonly used, they will be described. Some of the parameters appear in more than one equation, making it necessary to determine them in a sequential manner when the linear techniques are employed. Regardless of the estimation technique employed, however, it is important to recognize that all parameters estimated from a data set are interrelated. Consequently, an error in the estimation of one will influence the estimated values of the others. This means that more emphasis should be placed on the parameter set as a whole than on any individual values within it.

9.2.2 Determination of $Y_{H,T}$ and b_H

The first parameters to be estimated are the biomass yield, $Y_{H,T}$, and the traditional decay coefficient, b_H. As presented in Chapter 5, $Y_{H,T}$ has units of mg biomass TSS formed per mg of substrate COD used whereas b_H has units of hr^{-1}. Both can be obtained from a rearranged form of Equation 5.29:

$$X_{B,H,T} \cdot \tau = \frac{\Theta_c \cdot Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}. \tag{5.29}$$

Examination of Equation 5.29 reveals that the units of the true growth yield must be consistent with the units of the active biomass concentration; that is, the biomass concentration must be measured in TSS units to give a true growth yield in TSS units. As discussed in the preceding section, it is sometimes desirable to measure the biomass concentration in COD or VSS units. In that case, the true growth yield will have similar units. It makes no difference which unit system is used provided that it is used consistently and that the appropriate COD conversion factor, $i_{O/XB,T}$ or $i_{O/XB,V}$, is used when COD balances are performed. During the experimental studies, measurements were made of the total biomass concentration, $X_{total,T}$, not the active biomass, $X_{B,H,T}$. Thus, use must be made of the active fraction to get $X_{B,H,T}$:

$$X_{B,H,T} = f_A \cdot X_{total,T}. \tag{9.2}$$

Measurements were also made of the soluble COD in the feed, S_{CO}, and in the bioreactor, S_C, not the biodegradable COD, which is the substrate. The concentrations of biodegradable COD can be obtained from the measured soluble COD values by subtracting the inert soluble COD, S_I, which passes through the bioreactor:

$$S_{SO} = S_{CO} - S_I \tag{9.3}$$

$$S_S = S_C - S_I, \tag{9.4}$$

where S_{SO} and S_S are the biodegradable COD concentrations in the feed and bioreactor, respectively. However, for use in Equation 5.29, knowledge of the inert soluble COD is not required because it cancels out:

$$S_{SO} - S_S = S_{CO} - S_C. \tag{9.5}$$

Substitution of Equations 9.2 and 9.5 in Equation 5.29 yields:

$$f_A \cdot X_{total,T} \cdot \tau = \frac{\Theta_c \cdot Y_{H,T}(S_{CO} - S_C)}{1 + b_H \cdot \Theta_c}. \tag{9.6}$$

Data are collected on the effect of SRT on f_A, $X_{total,T}$, and S_C, with the hydraulic residence time (HRT), τ, and S_{CO} as controlled input values. Consequently, sufficient information is available for an estimation of $Y_{H,T}$ and b_H. The most suitable method for doing this depends on the structure of the errors in the data.[2] If the errors have constant variance, then a nonlinear least squares technique applied directly to Equation 9.6 without transformation is the most appropriate method.[2] Such techniques are available in spreadsheet and graphics programs. If nonlinear least squares estimation is inappropriate, then a linear least squares technique must be used. Linear least squares procedures are available on engineering calculators as well as in spreadsheet and graphics programs. Equation 9.6 can be linearized to give:

$$\frac{S_{CO} - S_C}{f_A \cdot X_{total,T} \cdot \tau} = \frac{b_H}{Y_{H,T}} + \frac{1}{Y_{H,T}} \cdot \frac{1}{\Theta_c}. \tag{9.7}$$

A plot of $(S_{CO} - S_C)/(f_A \cdot X_{total,T} \cdot \tau)$ versus $1/\Theta_c$ will give a straight line with a slope of $1/Y_{H,T}$ and an ordinate intercept of $b_H/Y_{H,T}$. This is illustrated in Figure 9.1.

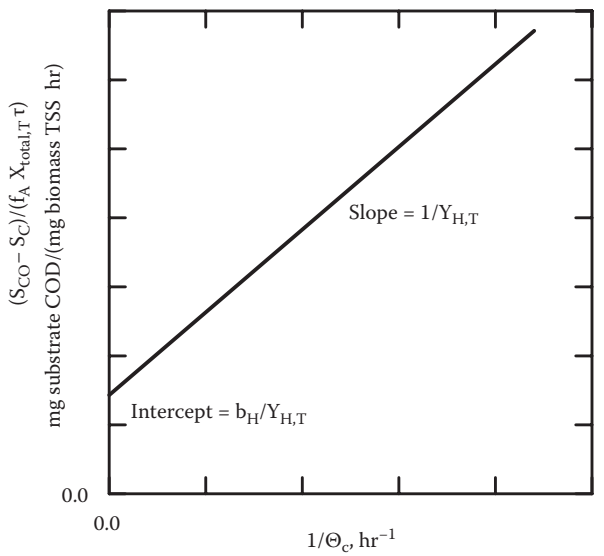

FIGURE 9.1 Plot of Equation 9.7 for the determination of $Y_{H,T}$ and b_H.

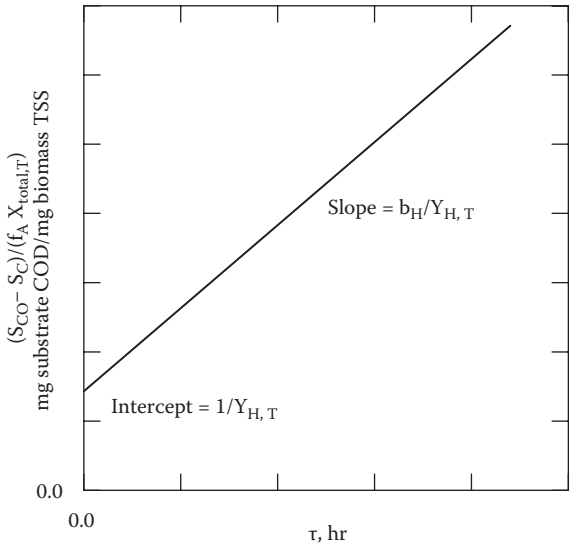

FIGURE 9.2 Plot of Equation 9.8 for the determination of $Y_{H,T}$ and b_H.

If the study is conducted in CSTRs without biomass separators, then the SRT (Θ_c) and the HRT (τ) will be the same, and Equation 5.31 must be used:

$$X_{B,H,T} = \frac{Y_{H,T}(S_{SO} - S_S)}{1 + b_H \cdot \tau}. \tag{5.31}$$

Making the same substitutions and linearizing gives:

$$\frac{S_{CO} - S_C}{f_A \cdot X_{total,T}} = \frac{1}{Y_{H,T}} + \frac{b_H}{Y_{H,T}} \cdot \tau. \tag{9.8}$$

A plot of $(S_{CO} - S_C)/(f_A \cdot X_{total,T})$ versus τ will give a straight line with a slope of $b_H/Y_{H,T}$ and an ordinate intercept of $1/Y_{H,T}$. This is illustrated in Figure 9.2.

The linearization technique giving Equations 9.7 and 9.8 is better than alternative linearizations when the error is either normally or log-normally distributed and the coefficient of variation is less than 11% or when the error is uniformly distributed, regardless of the coefficient of variation.[40]

9.2.3 Determination of f_D

The value of f_D can be determined from Equation 5.36 using nonlinear least squares analysis with b_H as a fixed value obtained from the preceding analysis:

$$f_A = \frac{1}{(1 + f_D \cdot b_H \cdot \Theta_c)}. \tag{5.36}$$

Alternatively, rearrangement gives:

$$\frac{1}{f_A} = 1 + f_D \cdot b_H \cdot \Theta_c. \tag{9.9}$$

A plot of $1/f_A$ versus Θ_c will give a straight line with a slope of $f_D \cdot b_H$. The ordinate intercept should pass through 1.0. Because b_H is known, f_D can be calculated.

9.2.4 Estimation of Inert Soluble COD, S_I

Before the kinetic parameters describing microbial growth and substrate utilization can be estimated, data must be available on the soluble substrate concentration, S_S. This requires knowledge of the inert soluble COD, S_I, as shown in Equation 9.4. One way to determine S_I is to remove an aliquot of the mixed liquor from one of the bioreactors operating at an SRT of 10 days or more, place it in a batch reactor, and aerate it. The soluble COD should be measured over time and when it reaches a stable residual value that value can be considered to be equivalent to the concentration of inert soluble COD in the feed.[25]

Alternatively, S_I can be estimated as the truly soluble COD remaining in the effluent from a bioreactor operated with a long SRT.[35] The rationale for this technique is that at longer SRTs, the amount of soluble, readily biodegradable COD remaining in the effluent will be negligibly small. Consequently, essentially all soluble COD remaining will be nonbiodegradable. The truly soluble COD is obtained by flocculating an effluent sample from the bioreactor with the longest SRT with $ZnSO_4$ at pH 10.5 (forming $Zn(OH)_2$ floc) prior to filtration through a 0.45 μm membrane filter. The flocculation step effectively removes colloidal organic matter that might pass through the filter, leaving only S_I.

The Dutch Foundation for Applied Water Research has developed guidelines for wastewater characterization in which S_I is determined from the COD in an aliquot of effluent from a bioreactor after filtration through a 0.1 μm membrane filter, which they consider adequate for retention of all colloidal organic matter without the flocculation step.[48] For municipal wastewater, S_I is taken as 90% of the filtered COD when the bioreactor is operated at a longer SRT value.

All of the above techniques are approximations. As discussed in Section 2.4.3, bacteria produce soluble microbial products as they degrade organic matter. Consequently, part of the inert organic matter remaining will actually be of microbial origin. However, because the models employed herein do not explicitly account for soluble microbial products, it is acceptable to consider it as part of the inert soluble COD as long as a constant influent soluble COD, S_{CO}, is used in the treatability study. Germirli et al.[18] have proposed a simple technique whereby the residual COD from the test described above may be partitioned into inert soluble COD from the influent and from microbial activity. It may be used in situations in which it is necessary to explicitly account for soluble microbial products.

9.2.5 Estimation of Monod Parameters, $\hat{\mu}_H$ and K_S

Once S_I is know, S_S can be calculated with Equation 9.4, thereby allowing $\hat{\mu}_H$ and K_S to be determined from Equation 5.22 using the value of b_H determined previously:

$$S_S = \frac{K_S(1/\Theta_c + b_H)}{\hat{\mu}_H - (1/\Theta_c + b_H)}. \tag{5.22}$$

As with $Y_{H,T}$ and b_H, the best way to determine $\hat{\mu}_H$ and K_S is through the use of a nonlinear least squares analysis. The S_S is treated as the dependent variable, with $1/\Theta_c + b_H$ as the independent variable. If a nonlinear least squares routine is unavailable, or if the error structure is inappropriate for nonlinear techniques, Equation 5.22 can be linearized to allow linear least squares analysis to be used. Three techniques are available.[12]

9.2.5.1 Hanes Linearization

If the plot is to be drawn by eye without the use of a least squares analysis, the Hanes linearization is preferable:

$$\frac{S_S}{1/\Theta_c + b_H} = \frac{K_S}{\hat{\mu}_H} + \frac{S_S}{\hat{\mu}_H}. \tag{9.10}$$

As illustrated in Figure 9.3, a plot of $S_S/(1/\Theta_c + b_H)$ versus S_S will yield a straight line with a slope of $1/\hat{\mu}_H$ and an ordinate intercept equal to $K_S/\hat{\mu}_H$. The scales can be chosen to give good estimates of $\hat{\mu}_H$ and K_S, but the linear least squares technique cannot be used to find the line of best fit because both axes contain terms that are subject to error (i.e., S_S).

9.2.5.2 Hofstee Linearization

If it is desired to use the linear least squares technique to determine the line of best fit, the Hofstee linearization should be used:

$$\frac{1/\Theta_c + b_H}{S_S} = \frac{\hat{\mu}_H}{K_S} - \frac{1}{K_S}(1/\Theta_c + b_H). \tag{9.11}$$

This is illustrated in Figure 9.4. The use of the reciprocal of S_S may amplify the error in S_S and make it difficult to fit the line by eye. Furthermore, because the independent variable $(1/\Theta_c + b_H)$ appears in both axes there will be some degree of inevitable correlation.

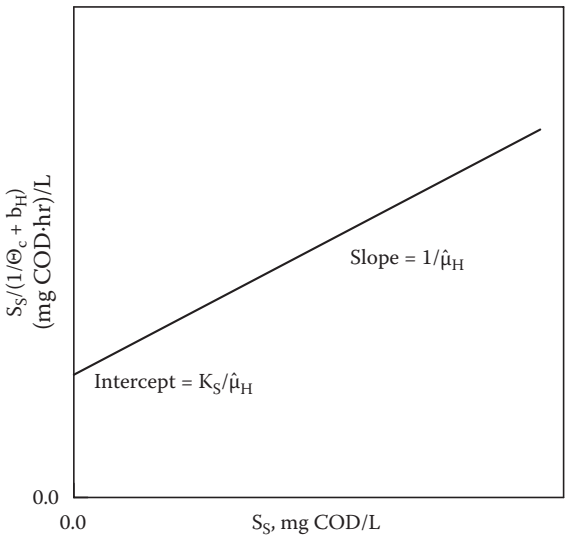

FIGURE 9.3 Hanes plot (Equation 9.10) for the determination of $\hat{\mu}_H$ and K_S.

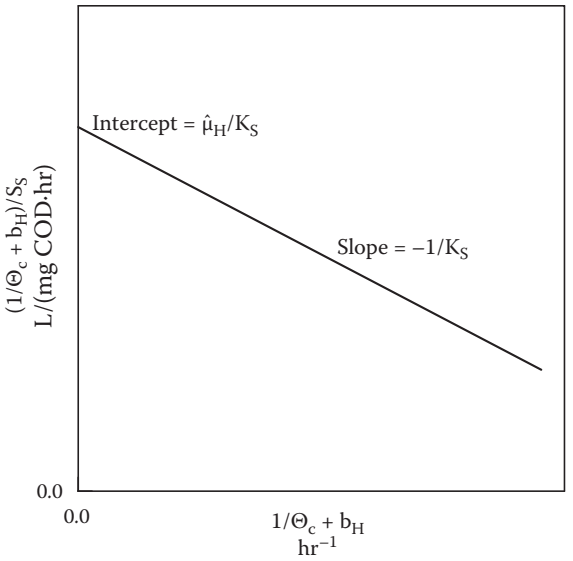

FIGURE 9.4 Hofstee plot (Equation 9.11) for determination of $\hat{\mu}_H$ and K_S.

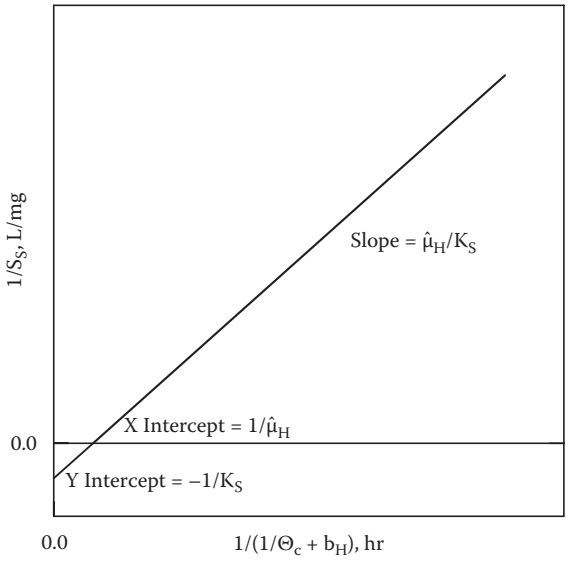

FIGURE 9.5 Lineweaver-Burk plot (Equation 9.12) for the determination of $\hat{\mu}_H$ and K_S.

9.2.5.3 Lineweaver–Burk Linearization

To eliminate the correlation in Equation 9.11, a slightly rearranged form of the equation may be used:[40]

$$\frac{1}{S_S} = \frac{\hat{\mu}_H}{K_S}\left(\frac{1}{1/\Theta_c + b_H}\right) - \frac{1}{K_S}. \tag{9.12}$$

This is illustrated in Figure 9.5. The Lineweaver-Burk method has been found to give a deceptively good fit, even with unreliable data points, and thus some authors have recommended against its

use.[12] Ong,[40] however, has found that Equation 9.12 works well for bioreactor data when the error is either normally or log-normally distributed.

9.2.6 Estimation of $k_{e,T}$

In situations in which the organic compounds in the wastewater are relatively easy to degrade, the soluble biodegradable COD in the effluent from the CSTRs is likely to be small relative to the half-saturation coefficient. This makes it difficult to obtain independent estimates of $\hat{\mu}_H$ and K_S. In that situation it is better to use the first-order approximation of the Monod equation to express the biodegradation kinetics of the wastewater constituents. It was given in Equation 5.53:

$$S_S = \frac{1}{Y_{H,T} \cdot k_{e,T}}(1/\Theta_c + b_H). \tag{5.53}$$

In that case, a plot of S_S versus $(1/\Theta_c + b_H)$ will yield a straight line that passes through the origin. Its slope will be $1/(Y_{H,T} \cdot k_{e,T})$ and can be determined from linear least squares analysis. Because the value of $Y_{H,T}$ is already known, the slope can be used to determine the value of $k_{e,T}$, the mean reaction rate coefficient.

Example 9.2.1

A treatability study was performed on an industrial wastewater using CSTRs with biomass recycle. The wastewater contained a complex mixture of soluble organic compounds with a total soluble COD of 350 mg/L. The wastewater contained no particulate matter. The studies were run in lab-scale bioreactors that had a volume of 6.0 L. The flow rate was maintained at a constant rate of 1.0 L/hr and the SRT was maintained at the desired values by wasting excess biomass directly from the reactors. The wastage rate was corrected for the loss of biomass in the effluent. Batch studies with biomass removed from one of the reactors revealed that the concentration of inert soluble COD was 35 mg/L. The COD/TSS conversion factor, $i_{O/XB,T}$, was determined to have a value of 1.20 mg COD/mg biomass TSS. Using the data provided in Table E9.1, estimate the values of the parameters $\hat{\mu}_H$, K_S, $Y_{H,T}$, b_H, and f_D. Use the linearized forms of the equations to do this.

a. The first task is to determine $Y_{H,T}$ and b_H. This is done by plotting $(S_{CO} - S_C)/(f_A \cdot X_{total,T} \cdot \tau)$ versus $1/\Theta_c$ as indicated in Equation 9.7. The values of S_C; the soluble COD, f_A; the active fraction; and $X_{total,T}$, the total biomass concentration are given in Table E9.1. Because the bioreactors all have a volume of 6 L and receive a flow rate of 1.0 L/hr, the HRT, τ, has a value of 6.0 hr. The resulting plot is shown in Figure E9.1. An application of linear least squares

TABLE E9.1
Data Collected during a Treatability Study with a Soluble Industrial Wastewater

SRT Hrs	Soluble COD mg COD/L	Total Biomass mg TSS/L	Active Fraction
48	44.2	833	0.95
96	39.9	1450	0.91
144	38.7	1930	0.87
192	38.2	2330	0.83
288	37.6	3000	0.76
384	37.4	3570	0.71

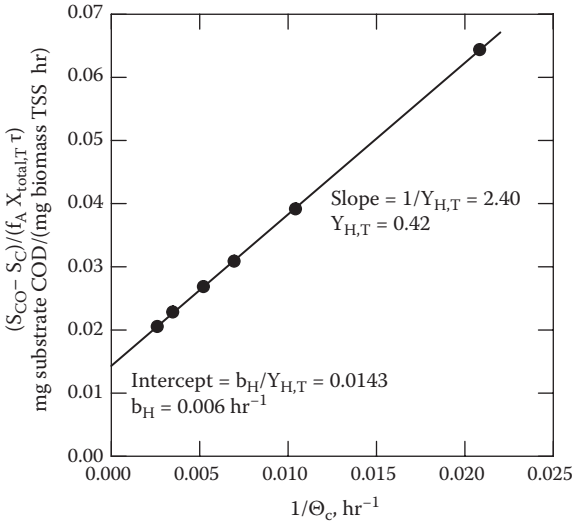

FIGURE E9.1 Determination of $Y_{H,T}$ and b_H in Example 9.2.1.

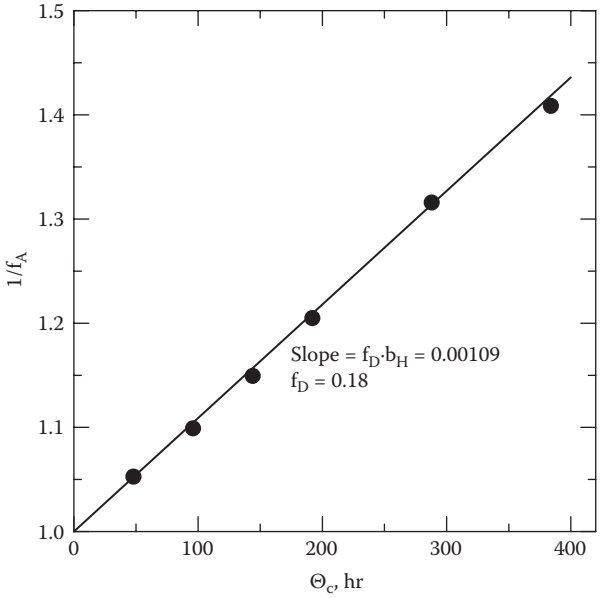

FIGURE E9.2 Determination of f_D in Example 9.2.1.

analysis to the data reveals that the slope is 2.40 and the ordinate intercept is 0.0143. The slope is equal to $1/Y_{H,T}$; consequently, $Y_{H,T}$ has a value of 0.42. Because the biomass concentration was measured as TSS, the units of $Y_{H,T}$ are mg biomass TSS formed/mg substrate COD removed. The intercept is equal to $b_H/Y_{H,T}$; consequently, b_H has a value of 0.006 hr^{-1}.

b. The second task is to estimate the value of f_D. This is done by plotting $1/f_A$ versus Θ_c as indicated by Equation 9.9. The resulting plot is shown in Figure E9.2. Application of linear least squares analysis to the data reveals that the slope of the line is 0.00109 hr^{-1}. The slope is equal to $f_D \cdot b_H$. Since b_H has a value of 0.006 hr^{-1}, f_D has a value of 0.18. It is dimensionless.

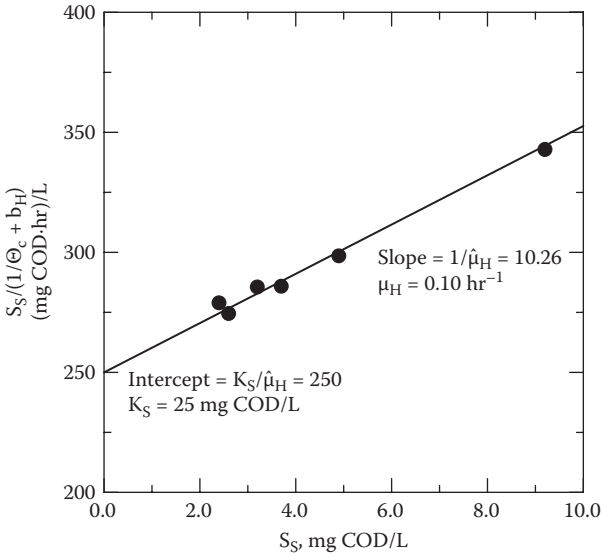

FIGURE E9.3 Determination of $\hat{\mu}_H$ and K_S in Example 9.2.1.

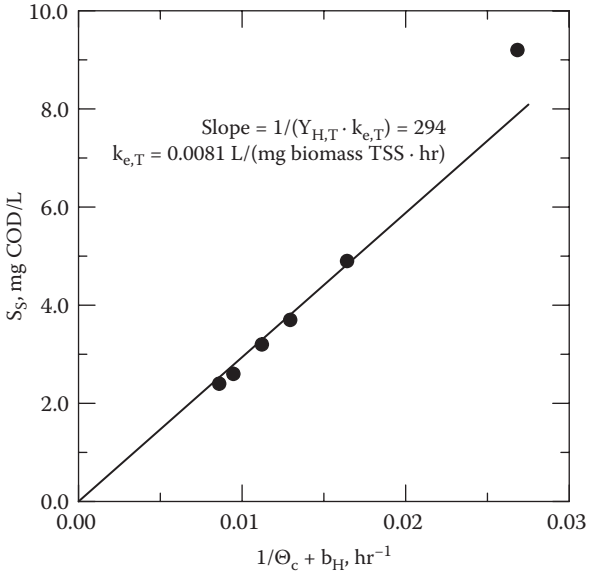

FIGURE E9.4 Determination of $k_{e,T}$ in Example 9.2.1.

c. The final task is to determine the values of $\hat{\mu}_H$ and K_S. Before this can be done, it is necessary to calculate the biodegradable COD concentration, S_S. This is done by subtracting the inert soluble COD, S_I from the soluble COD values given in Table E9.1, as indicated by Equation 9.4. The Hanes linearization will be used because it does not require taking the reciprocal of S_S. Because the values of S_S are generally small, taking their reciprocal can greatly amplify the error in them, making it difficult to see the trends. The Hanes linearization requires plotting $S_S/(1/\Theta_c + b_H)$ versus S_S as indicated in Equation 9.10. The plot is shown in Figure E9.3. It suggests that $\hat{\mu}_H$ has a value of 0.10 hr^{-1} and K_S has a value of 25 mg COD/L. One thing that is evident from the plot is that small errors in the estimated values of S_S have a large effect on the

estimates of the slope and the intercept. This is because the substrate concentration was low at all SRTs studied. While this suggests that it would be good to collect data at shorter SRTs, where the values of S_S would be larger, this would be difficult to do because bioflocculation would be poor (see Section 2.3.1). If the scatter in the data is so great that no confidence can be placed in the resulting parameter values, then it would be better to determine $k_{e,T}$, the mean reaction rate coefficient, which combines $\hat{\mu}_H$, K_S, and $Y_{H,T}$ into a single, first-order coefficient. This is done by plotting S_S versus $(1/\Theta_c + b_H)$ as indicated by Equation 5.53. Figure E9.4 shows the result of doing that in this case. Examination of the plot shows that only the lowest five points can be approximated as forming a straight line. For S_S values above 5 mg/L, S_S is not negligible with respect to K_S and thus the first-order approximation of the Monod equation does not apply. However, for S_S values below 5 mg/L, the linear approximation is acceptable. As can be seen in the plot, $k_{e,T}$ has a value of 0.0081 L/(mg biomass TSS · hr). This is slightly lower than the theoretical value 0.0095 L/(mg biomass TSS · hr) that comes from its definition and the previously determined values of $\hat{\mu}_H$, K_S, and $Y_{H,T}$ (see Sections 3.2.8 and 5.1.8). The lower measured value is a direct result of the approximate nature of first-order kinetics and the mean reaction rate coefficient.

9.3 SIMPLE SOLUBLE SUBSTRATE MODEL WITH TRADITIONAL DECAY IN THE ABSENCE OF DATA ON THE ACTIVE FRACTION

As discussed in Section 9.2.1, the active fraction of the biomass, f_A, is difficult to assess and thus data will generally not be collected on it during routine treatability studies for wastewater treatment. As a result, it will generally be impossible to determine b_H and $Y_{H,T}$ by the techniques outlined in Section 9.2.2. Rather, an additional experiment must be conducted that will allow independent determination of b_H. Once b_H is known, the value of $Y_{H,T}$ can be determined by using an assumed value for f_D of 0.20. The justification for assuming a value for f_D is that it is an inherent characteristic of biomass and, as such, should not change greatly from system to system.[25]

9.3.1 Data to Be Collected

With the exception of f_A, data will be collected from CSTRs exactly as described in Section 9.2.1. In addition, however, an experiment will be conducted for the direct determination of the decay coefficient for heterotrophic biomass, b_H. Biomass should be removed from one of the CSTRs and placed into a batch reactor where it will be aerated for several days. Generally, the longer the experiment is run, the more accurate the assessment of b_H will be. On a regular basis, biomass should be removed from the batch reactor and placed into a respirometer where the oxygen uptake rate (OUR) can be measured. Data on the decline in OUR over time will be used to assess b_H.[36] Historically, many investigators have used data on the change in TSS or VSS concentration over time for the determination of b_H, but the OUR technique has been shown to give more reproducible results provided proper precautions are taken.[36] The control of pH is one of those precautions. Since the bioreactor used for wastewater treatment will be maintained at neutral pH, it is important that the batch reactor used to provide the data for determining b_H also be maintained at neutral pH. The other major precaution concerns nitrification. If nitrification is well established in the CSTR from which the biomass is obtained, then continued nitrification will cause no problems in the experiment, provided that the pH is controlled and adequate oxygen is provided. This is because the ammonia released by biomass decay will be oxidized as it is released and no accumulation will occur. Consequently, the additional oxygen uptake associated with nitrification will be proportional to the oxygen uptake associated with heterotrophic decay and will not alter the change in the OUR over time.[36] It will only alter the magnitude of the OUR. If nitrification is not well established in the CSTR from which the biomass is obtained for the batch reactor, then the onset of nitrification in the batch reactor will distort the OUR data, making it unusable for the intended purpose. In that situation, nitrification should be inhibited during the OUR measurements by the addition of 20 mg/L of thiourea.[41]

9.3.2 Determination of b_H

The value of b_H can be obtained by two techniques from the data on the change in OUR over time. The rationale is as follows. For the traditional decay approach, the rate of oxygen utilization associated with decay was given by Equation 3.58:

$$OUR = -r_{SO} = (1 - f_D) b_H \cdot X_{B,H,T} \cdot i_{O/XB,T} \quad (O_2 \text{ units}). \tag{3.58}$$

In a batch bioreactor that receives only biomass there will be no soluble substrate, so the only reaction will be decay. Performing a mass balance on active biomass for the batch reactor and substituting the appropriate terms from the matrix in Table 5.2 reveals (the term $i_{O/XB,T}$ cancels out):

$$\frac{dX_{B,H,T}}{dt} = -b_H \cdot X_{B,H,T}, \tag{9.13}$$

which when integrated over time t tells us:

$$X_{B,H,T}\big|_t = X_{B,H,TO} \cdot e^{-b_H t}, \tag{9.14}$$

where $X_{B,H,TO}$ is the initial biomass concentration in the batch reactor. Substitution of Equation 9.14 into Equation 3.58 reveals that the oxygen uptake rate in the batch bioreactor at any time t is given by

$$OUR\big|_t = (1 - f_D) b_H \cdot X_{B,H,TO} \cdot i_{O/XB,T} \cdot e^{-b_H t}. \tag{9.15}$$

Equation 9.15 can be used directly to estimate b_H through the application of nonlinear least squares techniques. Examination of the equation shows that there are three unknowns, b_H, $X_{B,H,TO}$, and f_D, but that f_D and $X_{B,H,TO}$ cannot be determined independently of each other. As discussed above, however, a value of 0.20 can generally be assumed for f_D, allowing $X_{B,H,TO}$ and b_H to be determined. It should be noted that $X_{B,H,TO}$ is specific to the biomass used in the batch test and has no utility as a parameter.

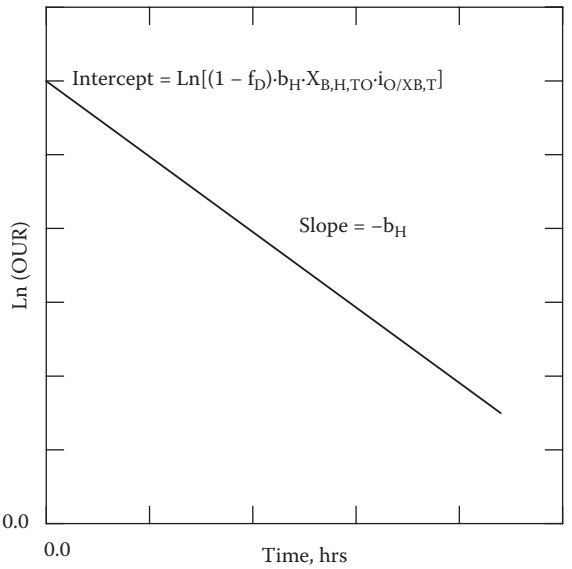

FIGURE 9.6 Plot of Equation 9.16 to determine b_H.

In the absence of nonlinear techniques, Equation 9.15 may be transformed, allowing linear least squares techniques to be used. Taking the natural log of both sides reveals:

$$\ln(\text{OUR})|_t = \ln[(1-f_D)b_H \cdot X_{B,H,TO} \cdot i_{O/XB,T}] - b_H \cdot t. \tag{9.16}$$

Consequently, an alternative technique is to plot ln(OUR) versus time and use linear least squares to obtain the slope, which is $-b_H$. This is illustrated in Figure 9.6. The ordinate intercept can be used to estimate $X_{B,H,TO}$ from assumed values for f_D and $i_{O/XB,T}$, but this is seldom done because $X_{B,H,TO}$ has limited utility, as noted above.

9.3.3 Determination of $Y_{H,T}$

Once the value b_H has been estimated and a value has been assumed for f_D, it is possible to estimate $Y_{H,T}$ from a rearranged form of Equation 5.34:

$$\frac{X_{total,T}}{S_{SO}-S_S} = Y_{H,T}\left[\left(\frac{\Theta_c}{\tau}\right)\frac{(1+f_D \cdot b_H \cdot \Theta_c)}{(1+b_H \cdot \Theta_c)}\right]. \tag{9.17}$$

Recall from Equation 9.5 that $S_{SO} - S_S$ is equal to $S_{CO} - S_C$. Because data have been collected from the CSTRs on the effect of SRT on $X_{total,T}$ and S_C, a plot may be made of $X_{total,T}/(S_{CO} - S_C)$ versus the bracketed term on the right side. It will have a slope equal to $Y_{H,T}$.

If the treatability study is conducted in CSTRs without biomass separators, then the SRT and the HRT will be the same, requiring a rearranged form of Equation 5.35 to be used:

$$\frac{X_{total,T}}{S_{SO}-S_S} = Y_{H,T}\left[\frac{(1+f_D \cdot b_H \cdot \tau)}{1+b_H \cdot \tau}\right]. \tag{9.18}$$

Again, $S_{CO} - S_C$ may be used in place of $S_{SO} - S_S$.

9.3.4 Determination of S_I, $\hat{\mu}_H$, K_S, and $k_{\varepsilon,T}$

Once f_D, b_H, and $Y_{H,T}$ are known, the estimation of the other parameters may proceed as described in Sections 9.2.4, 9.2.5, and 9.2.6.

Example 9.3.1

A treatability study was performed on an industrial wastewater using CSTRs with biomass recycle. The wastewater contained a complex mixture of soluble organic compounds with a total soluble COD of 400 mg/L. The wastewater contained no particulate matter. The studies were run in lab-scale bioreactors that had a volume of 6.0 L. The flow was maintained at a constant rate of 1.0 L/hr and the SRT was maintained at the desired values by wasting excess biomass directly from the reactors. The wastage rate was corrected for the loss of biomass in the effluent. Data collected from the bioreactors are shown in Table E9.2. In addition, biomass was removed from the bioreactor with an SRT of 96 hr, doubled in concentration by settling, and placed in a batch reactor to allow measurement of the OUR over time for estimation of b_H. The data from this test are shown in Table E9.3. Using the information provided, estimate the values of the parameters b_H and $Y_{H,T}$. Assume that f_D has a value of 0.20 and $i_{O/XB,T}$ has a value of 1.20 mg COD/mg biomass TSS.

a. The first task is to determine b_H from the batch data given in Table E9.3. This is done by plotting the natural log of the OUR versus time. The resulting plot is shown in Figure E9.5. The value of b_H is equivalent to the slope of the line, which is 0.007 hr^{-1}.

TABLE E9.2
Data Collected during a Treatability Study with a Soluble Industrial Wastewater

SRT Hrs	Soluble COD mg COD/L	Total Biomass mg TSS/L
48	29.6	1180
96	27.6	2020
144	27.0	2680
192	26.8	3230
288	26.5	4170
384	26.4	4980

TABLE E9.3
Data Collected during a Batch Aeration Test for the Determination of b_h

Time Hrs	OUR mg O_2/(L·hr)
0	24.0
12	22.0
24	20.2
36	18.6
48	17.1
72	14.5
96	12.2
120	10.3
144	8.7
192	6.2
240	4.5

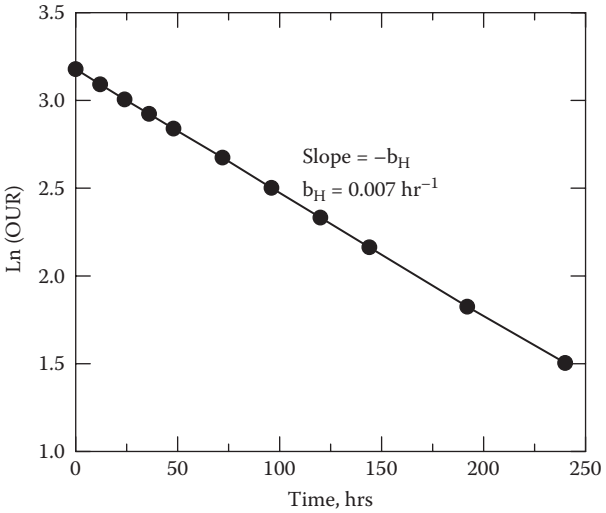

FIGURE E9.5 Determination of b_H in Example 9.3.1.

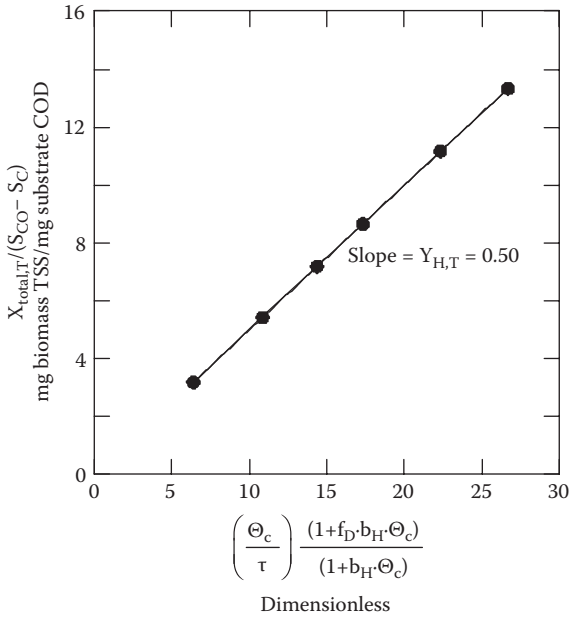

FIGURE E9.6 Determination of $Y_{H,T}$ in Example 9.3.1.

b. Once b_H has been evaluated it is possible to estimate $Y_{H,T}$ from Equation 9.17, provided a value is available for f_D. No data are available on the active fraction of the biomass, so it is necessary to assume a value for f_D. The risk in doing this is small because f_D is not thought to vary greatly from system to system. Figure E9.6 shows the plot of Equation 9.17 when f_D is assumed to be 0.20. The slope is equal to the value of $Y_{H,T}$, which is 0.50 mg biomass TSS/mg substrate COD.

9.4 USE OF BATCH REACTORS TO DETERMINE MONOD KINETIC PARAMETERS FOR SINGLE SUBSTRATES

The values of the Monod kinetic parameters $\hat{\mu}_H$ and K_S determined by the techniques discussed in Section 9.2.5 are for general organic matter as measured by COD. Frequently, however, it is desirable to know the kinetic parameters associated with the biodegradation of a single organic compound in a wastewater. Because of the relatively low specific growth rates at which suspended growth bioreactors operate, it is permissible to use parameter values determined during metabolism of the compound as the sole carbon and energy source without regard for metabolism of the other organic compounds present in the wastewater,[17] provided that none of the other compounds have strongly inhibitory effects. Such parameter values may be determined during batch experiments, but care must be exercised in the way in which those experiments are performed. This follows directly from the effect that growth conditions have on the physiological state of the biomass.

9.4.1 Intrinsic versus Extant Kinetics

The physiological state of a microbial culture is determined by the manner in which it is grown. This follows from the fact that the physiological state is a measure of the macromolecular composition of the cells in the culture. If the culture is grown at a very high rate, the cells' enzyme synthesizing system will be fully developed for rapid growth and the cells will contain high levels

of all enzymes. If the culture is grown at a slow rate, the cells will adjust the level of their enzyme synthesizing system to conserve resources. This means that they will contain lower enzyme levels. Consideration must be given to the physiological state of a culture and how it may change when batch experiments are designed for measuring kinetic parameter values. This is necessary because the parameter values measured in the experiment will depend upon the physiological state of the biomass.

Two extreme conditions represent the limits that the physiological state of a culture may attain during a kinetic experiment.[23] At one extreme, if the physiological state is not allowed to change, the resulting kinetic parameter values are reflective of the conditions of the biomass in the bioreactor from which the microorganisms used in the kinetic test were obtained. Because those parameters reflect the conditions "currently existing" in the parent bioreactor, they have been called "extant." At the other extreme, if the physiological state of the culture is allowed to change during the test to the point that the cells' protein synthesizing system is fully developed and the bacteria have an enzyme system that allows them to grow at the fastest rate possible on the test substrate at the given temperature and pH, the resulting kinetic parameters are said to be "intrinsic." The name follows from the fact that the parameter values are dependent only on the nature of the substrate and the types of bacteria in the culture. They are independent of the history of the culture (i.e., they are intrinsic). Parameter values obtained from experiments that allow the physiological state to be between these two extremes may be either "defined" or "undefined," depending upon the care with which the experiment is run and reported.

Extant kinetic parameters are better indicators of the removal of individual organic substrates in activated sludge bioreactors operated under steady-state conditions than are intrinsic parameters.[33] In addition, extant parameters can also be used to predict the extent of nitrification in bioreactors.[8] Unfortunately, neither extant nor intrinsic parameters adequately characterize the removal of individual organic substrates in highly dynamic activated sludge systems such as sequencing batch reactors.[34] Thus, additional research is required to resolve that issue. Nevertheless, at this time extant parameter values appear to have greater utility to wastewater treatment engineers. The main utility of intrinsic parameters appears to be for comparing the biodegradability of organic compounds.

9.4.2 Intrinsic Kinetics

The key to determining intrinsic kinetic parameter values during a batch test is to provide sufficient substrate to allow the bacteria performing the biodegradation to fully develop their protein synthesizing and enzyme systems. This can usually be accomplished when the initial substrate to biomass ratio ($S_{SO}/X_{B,HO}$) is at least 20 when both concentrations are expressed as COD.[6] In addition, the initial substrate concentration should be greater than the expected value of K_S. Since intrinsic K_S values for individual substrates tend to be less than 10 mg/L as COD,[22] experiments in which S_{SO} is 20 mg/L as COD have proven to be successful.[6] Consequently, the initial biomass concentration should be on the order of 1 mg/L as COD. It should be noted that the initial biomass concentration applies only to that portion of the biomass that is active in biodegradation of the test compound (i.e., the competent biomass). This will generally be only a fraction of the total biomass in a suspended growth system like activated sludge.

The substrate and biomass are placed into a batch reactor and the course of biodegradation is followed over time. Three types of data may be collected and used: substrate disappearance, biomass growth, or oxygen consumption. This follows directly from the proportionality of the three rates during balanced growth as depicted in Equation 3.34. The equivalency of the three data sets has also been shown experimentally.[11] Mass balance equations for substrate and biomass in a batch reactor must be written using the simplified model of Chapter 5 and solved simultaneously by appropriate numerical methods using assumed values for the parameters. If oxygen

consumption data are to be used for parameter estimation, the resulting theoretical substrate and biomass curves can be converted into the equivalent oxygen consumption curve using a COD balance as depicted by Equation 3.32. The theoretical data curve is then compared to the measured data curve and the parameter values are adjusted until the best agreement is achieved between the theoretical and measured curves. The parameter values associated with the best-fit curve are considered to be the best estimates of them. Because of the nature of the Monod (and Andrews) equation, a robust fitting routine is required to find the true best estimate of the parameters.[47] Details of a test procedure employing the collection and analysis of oxygen consumption data may be found in Brown et al.[6]

9.4.3 Extant Kinetics

Typically, samples of biomass are removed from operating activated sludge bioreactors (either lab-scale or full-scale) and used in batch tests to determine extant kinetic parameters. The requirements for determining extant values of the Monod (or Andrews) kinetic parameters are just the opposite of the requirements for the determination of intrinsic values. This follows from the fact that extant parameter values reflect the conditions of the biomass in the bioreactor from which they were obtained. Consequently, the test conditions need to be such that few changes occur in the physiological state of the biomass or the composition of the microbial community during the test. This can be achieved by keeping the value of $S_{SO}/X_{B,HO}$ (both expressed as COD) small during the batch tests, with 0.02 being a typical value. Again, it should be emphasized that $X_{B,HO}$ should reflect only that portion of the biomass that is capable of degrading the test compound (i.e., the competent biomass). Because the amount of substrate added is very small relative to the amount of biomass present, biomass growth may be neglected in any given test, allowing the use of only the mass balance equation for substrate. This simplifies the computations but means that data can only be collected on substrate loss or oxygen consumption.

The most commonly used technique for determining extant kinetic parameters relies on a single substrate injection into a batch respirometer with determination of the kinetic parameters by fitting the theoretical oxygen consumption curve to the observed oxygen consumption curve.[16] However, as pointed out in the preceding paragraph, because the quantity of substrate is small relative to the amount of biomass present, biomass growth may be ignored. This makes the fitting routine simpler because only the substrate mass balance equation must be solved. It also allows a closed-form solution, eliminating the need for a numerical solution of nonlinear differential equations.[21] As in the intrinsic test, S_{SO} should also be equal to or greater than the expected K_S value.

Determination of the maximum specific growth rate during extant tests requires knowledge of the competent biomass concentration in the respirometer. Wastewaters contain many organic compounds and most are degraded by only a portion of the microbial community in the bioreactor treating them. This makes it difficult to assess the competent fraction in that community. Because the proportion of enzymes in the biomass responsible for the biodegradation of a given compound should be related to the energy provided to the biomass by that compound, the competent fraction in a bioreactor has been approximated by the fraction of the influent biodegradable COD provided by the compound.[32] This approximation has tended to underestimate the competent fraction,[4] most likely because most of the bacteria in a mixed microbial community are involved in the biodegradation of more than one organic compound. Consequently, a calibration technique has been developed by which the competent fraction can be estimated.[15]

The extant respirometric technique may be used for determining both the biodegradation kinetics of individual organic compounds[16] and the kinetics of nitrification.[8] Details regarding the technique are too involved to provide herein, but the reader is encouraged to consult the cited references.

9.5 COMPLEX SUBSTRATE MODEL WITH LYSIS:REGROWTH APPROACH TO DECAY AS PRESENTED IN CHAPTER 6 (INTERNATIONAL WATER ASSOCIATION ACTIVATED SLUDGE MODEL NO. 1)

We saw in Chapter 6 that the International Water Association (IWA) has developed a number of activated sludge models (ASMs). All are complex, with many parameters. Before one of the models can be used to simulate the performance of a system treating a wastewater, values must be assigned to those parameters. In addition, the characteristics of the wastewater must be quantified. In this section we will briefly examine the procedures for obtaining information for ASM No. 1. Details can be found elsewhere,[25,37,41] as can a number of examples.[19,20,30,37] Space does not permit the presentation of similar information for the other IWA models.

Not all parameters in ASM No. 1 need to be evaluated for every wastewater, either because they do not change greatly or because the model output is not very sensitive to their values. This allows fixed values to be assumed for them.[25] These are listed in Table 9.1 and the values given in Table 6.3 are satisfactory for most purposes. However, it is important to recognize that many of the parameters are interrelated and that the value obtained for one will affect the value obtained for another. Thus, assumptions should be made with care.

9.5.1 Data to Be Collected

Because ASM No. 1 contains more elements than the simple soluble substrate model, evaluation of the parameters involved is more complicated. This requires more bioreactors to be operated.

Just as with the simple model, several CSTRs should be operated with steady-state feed over a range of SRTs. At least one should be operated with an SRT short enough to preclude nitrification. Others should have SRTs in excess of five days, providing mixed liquor that is low in undegraded particulate substrate and effluent that is low in readily biodegradable substrate. These bioreactors will be important to the evaluation of the parameters related to biomass production and electron acceptor requirement. Consequently, the data collected from them should be similar to that listed in Section 9.2.1, with a few exceptions. One is the active fraction, f_A, which is seldom measured. Another is that because of the need to make COD balances, it will be expedient to measure MLSS concentrations as COD. This means that the heterotrophic yield, Y_H, will be expressed as biomass COD formed per unit of substrate COD formed. If periodic measurements are also made of TSS and VSS, then the COD conversion factors $i_{O/XM,T}$ and $i_{O/XM,V}$ can be estimated, allowing modeling results and yield values to be expressed as TSS or VSS if desired. In addition, the total COD in the influent to the bioreactors should be measured.

Finally, since information is needed about denitrification rates, a reactor system containing two CSTRs in series should be operated as an MLE system. The system SRT, the mixed liquor recirculation flow rate, and the relative volumes of the aerobic and anoxic bioreactors should be selected to give stable nitrification and denitrification. Biomass from this system will be used to provide data on the nitrate utilization rate as outlined later.

9.5.2 Characterization of Wastewater and Estimation of Stoichiometric Coefficients

Activated sludge model No. 1 requires that the organic matter in a wastewater be partitioned into several components, as indicated in Tables 6.1 and 6.2: X_I, inert particulate organic matter; S_I, inert soluble organic matter; X_S, slowly biodegradable substrate; and S_S, readily biodegradable substrate. It is important to recognize that the distinction between the two types of biodegradable substrate is operationally defined and does not necessarily correspond to readily distinguishable physical characteristics, such as soluble and particulate, in spite of the symbols used in the model. Thus, characterization of the wastewater must be accomplished experimentally. One of the procedures for wastewater characterization, the determination of the concentration of inert soluble organic

TABLE 9.1
Parameters That May Be Assumed

Symbol	Description
Y_A	Yield for autotrophic biomass
f'_D	Fraction of biomass leading to debris
$i_{N/XB}$	Mass of nitrogen per mass of COD in biomass
$i_{N/XD}$	Mass of nitrogen per mass of COD in biomass debris
$K_{O,H}$	Oxygen half-saturation coefficient for heterotrophic biomass
K_{NO}	Nitrate half-saturation coefficient for denitrifying heterotrophic biomass
$K_{O,A}$	Oxygen half-saturation coefficient for autotrophic biomass

TABLE 9.2
Parameters and Characteristics That Must Be Evaluated and Information Needed

Symbol	Name	Prior Information Needed
S_{NOO}	Soluble nitrate-N concentration in wastewater	
S_{NHO}	Soluble ammonia-N concentration in wastewater	
S_{IO}	Soluble inert COD concentration in wastewater	
S_{NIO}	Soluble inert organic-N concentration in wastewater	
S_{NSO}	Soluble biodegradable organic-N concentration in wastewater	S_{NIO}
Y_H	Yield for heterotrophic biomass	
S_{SO}	Concentration of readily biodegradable COD in wastewater	Y_H
$b_{L,A}$	Decay coefficient for autotrophic biomass	
$\hat{\mu}_A$	Maximum specific growth rate for autotrophic biomass	$b_{L,A}$
K_{NH}	Ammonia-N half-saturation coefficient for autotrophic biomass	
$b_{L,H}$	Decay coefficient for heterotrophic biomass	Y_H, f'_D
X_{IO}	Inert suspended organic matter concentration in wastewater	f'_D, $b_{L,H}$, S_{SO}, S_{IO}
X_{SO}	Slowly biodegradable organic matter concentration in wastewater	
X_{NSO}	Slowly biodegradable organic-N concentration in wastewater	S_{SO}, X_{SO}, S_{NSO}
η_g	Correction factor for $\hat{\mu}_H$ under anoxic conditions	
η_h	Correction factor for hydrolysis under anoxic conditions	
$\hat{\mu}_A$	Maximum specific growth rate for heterotrophic biomass	Y_H, X_{SO}, X_{IO}, S_{SO}, f'_D
K_S	Half-saturation coefficient for heterotrophic biomass	Y_H, X_{SO}, X_{IO}, S_{SO}, f'_D
k_h	Maximum specific hydrolysis rate	All other parameters
K_X	Half-saturation coefficient for hydrolysis of slowly biodegradable substrate	All other parameters
k_a	Ammonification rate	

matter, S_I, is the same as the procedure presented in Section 9.2.4 and thus will not be repeated here. However, it should be noted that because inert soluble organic matter passes through a bioreactor unchanged, the concentration in the bioreactor, S_I, is the same as the concentration in the feed, S_{IO}. We will focus in this section on the quantification of the other three fractions. Before that can be done, however, Y_H must be known.

9.5.2.1 Determination of Y_H

The methods employed in Sections 9.2.2 and 9.3.3 for determining Y_H cannot be used here because of the presence of particulate organic matter in the MLSS. Thus, the approach that has been recommended is to observe Y_H directly as biomass is grown on only the soluble component of the wastewater.[25] An aliquot of wastewater should be filtered to remove the particulate COD, placed into a batch bioreactor, and seeded with a small amount of biomass from one of the CSTRs. Care should be exercised to obtain a sample with as high a soluble COD concentration as possible and to make the concentration of seed biomass very small relative to the initial soluble COD (<1%). This is because Y_H is defined as the amount of biomass grown in the absence of decay and the only way this can be directly observed is to allow a small amount of biomass to grow rapidly on a large amount of substrate. Samples should then be taken over time so that the total and soluble COD can be measured as biomass growth proceeds. The biomass COD can be calculated as the difference between the total COD and soluble COD, and the yield can be determined from its definition:

$$Y_H = \frac{\Delta \text{ biomass COD}}{\Delta \text{ soluble COD}}. \tag{9.19}$$

Generally, the best way to determine Y_H is to plot the biomass COD as a function of the soluble COD removed and take the slope of the resulting line. Alternatively, the biomass can be measured directly by suspended solids measurements, rather than as COD, in which case the yield is expressed in terms of TSS or VSS, depending upon the measurement used. Every effort should be made to get an accurate estimate of Y_H because errors in its determination will influence the estimates of the various fractions of the wastewater organic matter.

9.5.2.2 Determination of Influent Readily Biodegradable COD (S_{SO})

One procedure originally recommended for determination of the readily biodegradable COD concentration is a bioassay requiring the use of a bioreactor receiving a square wave input of feed.[14,25] Because of the complexity of that assay, a number of alternatives have been proposed.[41,53] Two will be presented here, a batch bioassay[14] and a rapid physical assay[35] that is very simple.

In the bioassay procedure, data are collected from a batch reactor containing MLSS taken from one of the CSTRs. Wastewater is added to the reactor and OUR measurements are made frequently to define the change in OUR over time as the substrates are utilized. Figure 9.7 illustrates the idealized OUR response when the MLSS used in the test is from a bioreactor that is fully nitrifying. The areas correspond to the mass of oxygen per unit reactor volume associated with the events occurring in the reactor in order of their relative reaction rates. Area 1 is associated with the utilization of readily biodegradable substrate, area 2 with nitrification, area 3 with the utilization of slowly biodegradable substrate, and area 4 with lysis and decay (endogenous metabolism). If the biomass is taken from a CSTR that is not nitrifying, or if an inhibitor of nitrification is added, then no nitrification will occur and the resulting OUR curve will be simpler and easier to interpret (area 2 will be eliminated). Care should be exercised to ensure that an appropriate substrate to biomass (F/M) ratio is used in the test. Figure 9.8 illustrates why this is important. As can be seen there, if the F/M is either too high or too low the identification and quantification of the individual areas will be difficult. Once area 1 has been quantified, the readily biodegradable substrate concentration in the wastewater, S_{SO}, can be calculated with Equation 9.20:[37]

$$S_{SO} = \left(\frac{M_{O_2}}{1 - Y_H}\right)\left(\frac{V_{MLSS} + V_{WW}}{V_{WW}}\right), \tag{9.20}$$

where M_{O_2} is the mass of oxygen consumed per liter of batch reactor volume during utilization of the readily biodegradable substrate (area 1 in Figure 9.7), V_{MLSS} is the volume of MLSS used in the

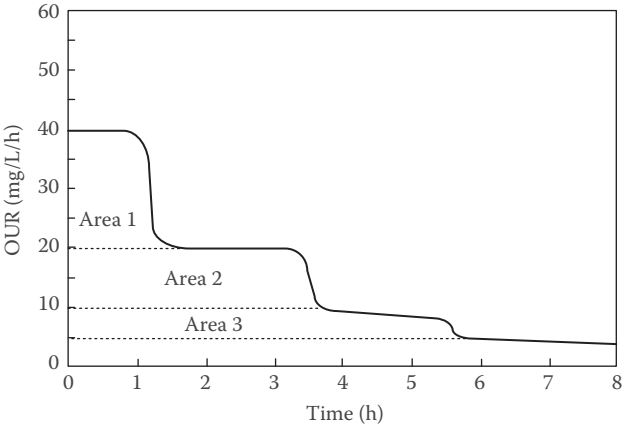

FIGURE 9.7 Idealized OUR response observed in an aerobic batch test using MLSS from a CSTR in which nitrification is well established. (Reprinted from Melcer, H., Dold, P. L., Jones, R. M., Bye, C. M., Takacs, I., Stensel, H. D., Wilson, A. W., Sun, P., and Bury, S., *Methods for Wastewater Characterization in Activated Sludge Modeling*, Report 99-WWF-3, Water Environment Research Foundation, Water Environment Federation, Alexandria, Virginia, 2003. Copyright © Water Environment Research Foundation, Alexandria, Virginia. With permission.)

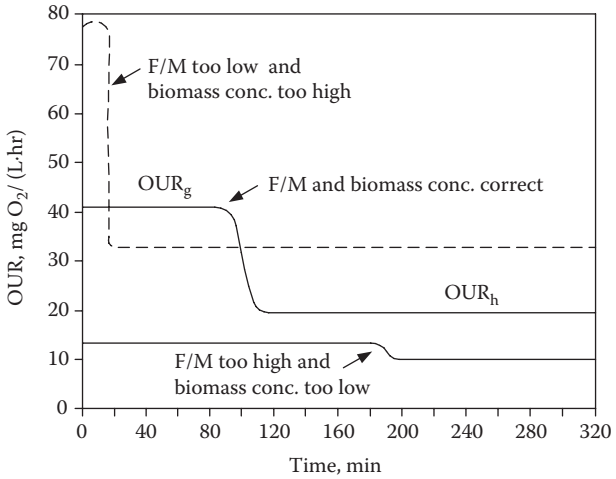

FIGURE 9.8 Effect of changing the substrate to biomass ratio (F/M) on the OUR in a batch reactor. The faster rate corresponds to the use of readily biodegradable substrate (OUR_g), whereas the slower rate corresponds to the use of slowly biodegradable substrate (OUR_h). (Reprinted from Ekama, G. A., Dold, P. L., and Marais, G. v. R., Procedures for determining influent COD fractions and the maximum specific growth rate of heterotrophs in activated sludge systems. *Water Science and Technology,* 18 (6): 91–114, 1986. Copyright © IWA Publishing. With permission.)

test, and V_{WW} is the volume of wastewater used in the test. It is important to recognize that Y_H must be in COD units for use in Equation 9.20.

The physical assay for S_{SO} developed by Mamais et al.[35] is much quicker than the bioassay and gives results that correlate well with the original bioassay for domestic wastewater. In it, samples of raw wastewater are flocculated by adding $ZnSO_4$, mixing vigorously for one minute, and adjusting the pH to 10.5 with NaOH. They are then allowed to settle quiescently before a sample of clear supernatant is withdrawn and filtered through a 0.45 μm membrane filter. The flocculation step removes colloidal organic matter that otherwise would pass through the filter and be measured as "soluble" material. The COD of the filtrate is the total soluble COD of the wastewater. Subtraction

of the inert soluble COD, S_I, provides the value of the readily biodegradable COD, S_{SO}. A major advantage of this technique is that more samples can be analyzed, giving a better measure of the long-term average concentration of the readily biodegradable COD in the wastewater than can be obtained with the bioassay. If the wastewater is from an industrial facility or contains significant industrial components, then the physical assay may overestimate S_{SO} because some of the soluble organic compounds may be biodegraded slowly.

9.5.2.3 Determination of Influent Inert Particulate COD (X_{IO})

The total COD in a wastewater is given by

$$\text{Total influent COD} = COD_{TO} = X_{SO} + X_{IO} + S_{SO} + S_{IO}. \tag{9.21}$$

The total influent COD (COD_{TO}) can be measured and S_{SO} and S_{IO} can be estimated by the procedures given previously. If either the influent slowly biodegradable COD, X_{SO}, or X_{IO} is determined, the other can be calculated from Equation 9.21. It is best to estimate X_{IO} as a parameter for fitting the model to data showing the effect of SRT on the concentration of MLSS in the CSTRs. The value of X_{SO} can then be calculated from the total influent COD using Equation 9.21. Ignoring autotrophic biomass, which is usually negligible, the MLSS in the CSTRs comes from four major sources: growth of heterotrophs, production of microbial debris, accumulation of inert suspended organic matter from the feed, and accumulation of undegraded slowly biodegradable substrate. If the SRT is greater than five days, the values of X_S and S_S in a bioreactor will be negligibly small relative to X_{SO} and S_{SO}. This means that the MLSS concentration in a CSTR in COD units (X_M) can be approximated by the simple soluble substrate model that includes inert particulate COD (X_{IO}), provided that the measurement of the influent biodegradable COD includes the biodegradable particulate substrate. In other words, even though we are estimating parameters for ASM No. 1, Equation 5.58 can be rewritten in COD units and used to estimate X_{IO}:

$$X_M = \left(\frac{\Theta_c}{\tau}\right)\left[X_{IO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c) Y_H (COD_{TO} - S_{IO} - X_{IO})}{1 + b_H \cdot \Theta_c}\right]. \tag{9.22}$$

Everything in Equation 9.22 except X_{IO} is either known or can be assumed. The value of b_H can be evaluated independently by the technique in Section 9.3.2. It is important to recognize that the traditional decay coefficient, b_H, is used at this point. That is because Equation 9.22 is from the simple model. The value of Y_H has already been evaluated. The value of f_D can be assumed to be 0.20, just as before with the simple model. Again, the value of f_D for the simple model is used in this computation because Equation 9.22 is from that model. The values of COD_{TO} and S_{IO} are measured values. The SRT (Θ_c) and the HRT (τ) are experimentally controlled variables. Generally, it is best to maintain a fixed HRT during the operation of the CSTRs with different SRTs, thereby making the MLSS concentration a function of only the SRT. The value of X_{IO} can then be estimated by using a one-dimensional search routine that chooses X_{IO} to minimize the error sum of squares when the predicted MLSS concentration is compared to the values measured in the CSTRs operated over a range of SRTs. If it is necessary to change the HRT as well as the SRT during the studies, then τ should be moved to the left side of Equation 9.22 and predicted values of $X_M \cdot \tau$ should be compared to the measured values during the estimation of X_{IO}.

9.5.2.4 Characterization of Nitrogen-Containing Material

Activated sludge model No. 1 includes terms for nitrogen as well as COD. Thus, it is also necessary to characterize them during treatability studies. All soluble forms can be determined by appropriate chemical analyses on the wastewater. The concentration of inert soluble organic nitrogen in the wastewater can be determined by performing Kjeldahl nitrogen tests on aliquots of the samples used

to determine the inert soluble COD (see Section 9.2.4). Subtraction of the inert soluble organic nitrogen from the soluble organic nitrogen in the influent gives the concentration of readily biodegradable organic nitrogen in the feed, S_{NSO}. Readily and slowly (X_{NSO}) biodegradable organic nitrogen in the wastewater are then proportioned in the same way as readily and slowly biodegradable COD:

$$\frac{S_{NSO}}{X_{NSO} + S_{NSO}} = \frac{S_{SO}}{X_{SO} + S_{SO}}. \tag{9.23}$$

Because everything in Equation 9.23 except X_{NSO} is known, its value can be calculated. The sum of the biodegradable organic nitrogen species and the inert soluble organic nitrogen in the wastewater should be less than the total organic nitrogen, which can be measured. The difference between the two is the inert particulate organic nitrogen. Although the latter is not used in the model, its value should be calculated as a check on the procedures. A negative value is evidence of an error.

9.5.3 Estimation of Kinetic Parameters

9.5.3.1 Aerobic Growth of Heterotrophs

Because of all of the terms in ASM No. 1 and the generation of soluble substrate from slowly biodegradable substrate, it is not possible to use the approach described in Section 9.2.5 to obtain $\hat{\mu}_H$ and K_S. Consequently, an alternative approach must be used. The extant batch technique described in Section 9.4.3 may be applied using the soluble fraction of the wastewater. Because the main function of $\hat{\mu}_H$ and K_S in ASM No. 1 is to allow the maximum oxygen uptake rate of an operating bioreactor to be calculated, the extant parameter values should be measured using biomass taken from a bioreactor with a short SRT. The biomass concentration used in the estimation of the specific rates should correspond to that of the active heterotrophic biomass, $X_{B,H}$. Sufficient information in the form of stoichiometric and kinetic parameters is available at this point to allow its estimation for the bioreactor from which the biomass in the test is obtained, thereby allowing its estimation for the batch test reactor. Substrate injections should be made with coagulated and filtered wastewater, thereby providing only readily biodegradable substrate during the test.

9.5.3.2 Decay of Autotrophs

Because the technique for determining the maximum specific growth rate of the autotrophic nitrifying bacteria ($\hat{\mu}_A$; described below) requires knowledge of the decay coefficient for nitrifiers, it must be quantified first. Historically, most investigators assumed a value for the autotrophic decay coefficient because it was thought to be small and of negligible importance. Recent studies, however, have indicated that it is larger than had previously been thought.[37] Therefore, assumption of a small value can introduce significant error into the estimate of $\hat{\mu}_A$.[37] The reason that the autotrophic decay coefficient had been underestimated lies in the test method typically employed to measure it. The rationale and test method are very similar to the batch method for measuring the heterotrophic decay coefficient described in Section 9.3.2. In this case, however, the nitrate production rate is used instead of the OUR as the indicator of the relevant (nitrifier) biomass present. The problem with the test is that the biomass used, which typically comes from one of the CSTRs with a sufficient SRT to have fully developed nitrification, contains heterotrophs as well as autotrophs. Furthermore, as we saw in Section 6.3.2, the mass of heterotrophs present will be much greater than the mass of autotrophs. Because the heterotrophs are also decaying, nitrogen will be released from them, which provides substrate for the autotrophs. Thus, even though no external substrate is provided to the reactor, autotrophic substrate is generated internally, allowing autotrophic growth to occur even though decay is the predominant reaction. As a consequence, the measured autotrophic decay rate is less than the actual autotrophic decay rate.

Nonlinear curve-fitting procedures have been developed to allow the true autotrophic decay rate to be determined from the batch test by accounting for the nitrogen released as a result of heterotrophic decay.[37] If one is estimating parameters for a wastewater with a large industrial component then use of that procedure is recommended. However, if one is dealing with a predominantly municipal wastewater, it would probably be satisfactory to assume a value of 0.0071 hr^{-1} for the autotrophic decay coefficient with a temperature correction factor, θ, of 1.029.[37] This issue will doubtless receive additional study in the future and thus the accuracy of using an assumed value will become clearer.

Because of the way the autotrophic decay coefficient is measured, it is a traditional decay coefficient, b_A. However, it will be recalled from Section 3.3.2 that for autotrophic bacteria the traditional and lysis:regrowth decay coefficients are equal; thus, $b_A = b_{L,A}$.

9.5.3.3 Aerobic Growth of Autotrophs

As far as the design and control of nitrifying bioreactor systems are concerned, the maximum specific growth rate coefficient for autotrophic biomass, $\hat{\mu}_A$, is the most critical parameter value. There are two reasons for this. First, it determines the SRT at which nitrifying biomass will be eliminated from the system, thereby fixing the minimum acceptable SRT at which the system can be operated. Second, it can be affected strongly by chemicals in the wastewater; more strongly, in fact, than the half-saturation coefficient. Thus, it is important that an accurate assessment of $\hat{\mu}_A$ be obtained in an environment that represents the wastewater undergoing treatment. Several procedures have been proposed for measuring $\hat{\mu}_A$,[13,24,37,41] but the simplest involves a batch experiment started with a small amount of biomass. Effluent should be collected from one of the continuous CSTRs, preferably one with a short SRT so that little nitrification will have occurred, making the initial concentrations of nitrate- and nitrite-N small. If necessary, ammonia-N should be added to the bioreactor to bring the concentration to approximately 40 mg/L. The bioreactor should then be seeded with biomass from a bioreactor with an active nitrifying population, but the initial nitrifying biomass concentration should be less than 1.0 mg/L. Because the production of nitrate-N and nitrite-N is proportional to the amount of nitrifying bacteria formed, the change in the concentration of oxidized nitrogen (NO_3^--N + NO_2^--N) can be used to estimate $\hat{\mu}_A$.[1,24] The concentration of oxidized nitrogen in the bioreactor should be measured over time as it increases through growth of the autotrophs and the natural log of (NO_3^--N + NO_2^--N) should be plotted versus time. The plot should give a straight line, with a slope of $\hat{\mu}_A - b_{L,A}$. If the wastewater in question is primarily municipal in origin, then a value 0.0071 hr^{-1} at 20°C can be assumed for $b_{L,A}$.[37] However, if the wastewater contains a significant industrial component, it would be safer to measure the autotrophic decay coefficient experimentally. Information for doing this can be found in Melcer et al.[37] Once the value of $b_{L,A}$ has been assumed or measured, $\hat{\mu}_A$ can be calculated. Examples of this procedure can be found in the literature.[1,20,37]

Although the half-saturation coefficient for nitrifying bacteria, K_{NH}, is not affected as strongly by organic contaminants as the maximum specific growth rate, it may be influenced and thus should be determined. This may be done with the extant kinetic parameter technique of Lamb et al.[31] and Chudoba and colleagues.[7,10] Biomass should be removed from a CSTR that is fully nitrifying and placed into a respirometer without dilution. Small quantities of ammonia-N should be injected and the net respiration rate measured in response to the injections. This provides data on OUR as a function of injected ammonia-N concentration, which is equivalent to data on the nitrification rate as a function of ammonia-N concentration. This can be analyzed by the Hanes technique (Figure 9.3) to estimate K_{NH} using OUR in place of the specific growth rate of the autotrophs, μ_A. Because the autotrophic biomass concentration is unknown, the maximum OUR cannot be related to $\hat{\mu}_A$ but the value of K_{NH} will be valid for use with the value of $\hat{\mu}_A$ determined from the batch test described in the preceding paragraph.

9.5.3.4 Decay of Heterotrophs

The heterotrophic decay coefficient, $b_{L,H}$, is very important to predictions of biomass production and oxygen requirements, so it must be determined for the particular wastewater under study. Therefore, biomass should be removed from one of the CSTRs and used in the batch procedure of Section 9.3.2 to determine the traditional decay coefficient, b_H, which can be used to determine X_{IO} as discussed in Section 9.5.2. Biomass can be removed from any of the CSTRs, but correction for the effects of nitrification will be easier if the biomass is fully nitrifying as discussed in Section 9.3.1. Once the value of b_H has been obtained it can be used to calculate $b_{L,H}$ with Equation 3.69:

$$b_{L,H} = \frac{b_H}{[1 - Y_H(1 - f_D')]}. \tag{3.69}$$

The value of f_D' should be assumed to be 0.08.

9.5.3.5 Correction Factors for Anoxic Conditions, η_g and η_h

Two important parameters in ASM No. 1 are η_g and η_h because they correct the rates of growth and hydrolysis reactions when they occur under anoxic conditions. As discussed in Section 6.1.2, this is required because only a portion of the biomass is capable of functioning under anoxic conditions and the maximum specific growth rate coefficient and yield are different with the two electron acceptors. Thus, the model needs some way to reflect these facts. Tests to measure η_g and η_h are performed at the same time by evaluating oxygen and nitrate consumption rates in two batch bioreactors that are equivalent in every respect except for the terminal electron acceptor.[25] The biomass for the tests should come from the MLE bioreactor run as part of the parameter estimation study because it will contain biomass capable of functioning under both aerobic and anoxic conditions. The rationale for the test is as follows. Immediately after biomass is brought into contact with wastewater in a batch bioreactor, the activity in the bioreactor will be dominated by growth of the heterotrophs on the readily biodegradable substrate. However, as soon as the readily biodegradable substrate is exhausted, the activity will be predominantly due to the use of substrate arising from hydrolysis of the slowly biodegradable substrate. Therefore, by comparing the activity of a biomass sample in both of these regions under both aerobic and anoxic conditions it is possible to estimate η_g and η_h.

Conceptually, the experiment is very simple. Biomass is removed from the MLE system and placed into two batch bioreactors, one of which is maintained under aerobic conditions with oxygen as the terminal electron acceptor and the other of which is kept under anoxic conditions with nitrate as the terminal electron acceptor. The latter bioreactor should be constructed to minimize oxygen transfer to the liquid.[45] Wastewater is then added to both bioreactors and the OUR and nitrate utilization rate (NUR) are measured in the appropriate bioreactors as the substrate is depleted. The OUR is normally measured by placing a dissolved oxygen (DO) probe in the aerobic bioreactor. By providing a mechanical mixer to keep the biomass in suspension, it is possible to turn off the air supply periodically and measure the OUR directly by the rate of decrease in the DO concentration over a short time period. The NUR is normally measured by manually removing samples from the bioreactor over time, stopping the reaction, removing the biomass, and measuring the nitrate-N concentration. The NUR in the two reaction regions is then determined from the slopes of the plot of nitrate-N over time as illustrated in Figure 9.9. Care should be exercised in the measurement of the NUR to ensure that nitrite is not accumulating in the bioreactor. If it is, then the results should be expressed in terms of the net amount of electron acceptor used expressed as the equivalent amount of oxygen used. Care should also be taken to ensure that an appropriate substrate to biomass (F/M) ratio is used in the tests. Figure 9.8 illustrates why this is important in terms of the OUR. If the F/M ratio is too low, the time required for readily biodegradable substrate removal will be too short to get a good measure of the rate. Conversely, if the ratio is too large, the difference in rate between

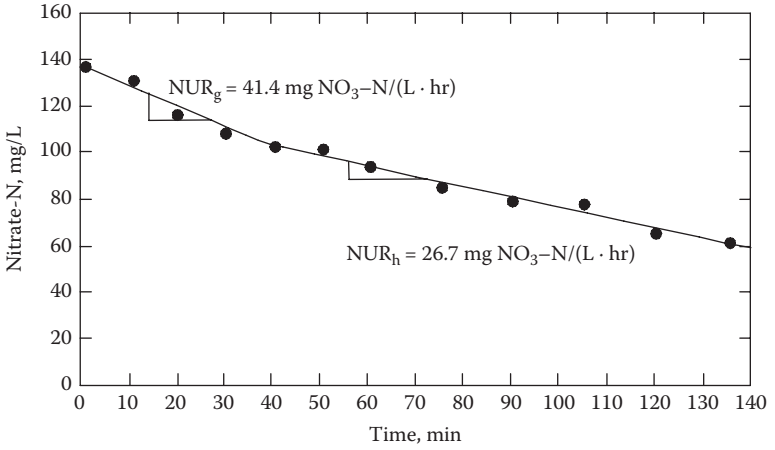

FIGURE 9.9 Nitrate utilization in a batch reactor. The faster rate corresponds to the use of readily biodegradable substrate (NUR_g), whereas the slower rate corresponds to the use of slowly biodegradable substrate (NUR_h). (Reprinted from Givens, S. W. Brown, E. V., Gelman, S. R., Grady Jr., C. P. L., and Skedsvold, D. A., Biological process design and pilot testing for a carbon oxidation, nitrification and denitrification system. *Environmental Progress,* 10:133–46, 1991. Copyright © American Institute of Chemical Engineers. With permission.)

the two zones will not be sufficiently large to clearly distinguish between them. The value of η_g can be calculated once data are available for OUR_g and NUR_g:

$$\eta_g = \frac{2.86 \times NUR_g}{OUR_g}. \tag{9.24}$$

Likewise, the value of η_h can be calculated from OUR_h and NUR_h:

$$\eta_h = \frac{2.86 \times NUR_h}{OUR_h}. \tag{9.25}$$

9.5.3.6 Hydrolysis and Ammonification

Three parameters remain to be evaluated. Two are the parameters characterizing hydrolysis of slowly biodegradable substrate, k_h and K_X. The third is the parameter describing ammonification, k_a. Estimation of these parameters is often accomplished by measuring the OUR in a batch reactor containing biomass from one of the continuous reactors.[27,28,42] As with determination of the readily and slowly biodegradable substrate concentrations, proper selection of the F/M ratio for the test is important to an estimation of the parameters. In this case emphasis is on modeling of the progression of OUR over time in area 3 of Figure 9.7 and the F/M ratio should be selected to provide a well-defined curve for fitting. Use of a nitrification inhibitor allows elimination of area 2, which helps to delineate the metabolism of slowly biodegradable substrate in area 3. It also allows the model to be simplified, as does the fact that the batch test is conducted under fully aerobic conditions. The hydrolysis parameters, k_h and K_X, are determined by nonlinear curve fitting of ASM No. 1 to the OUR data. All other parameters in the model are known or assumed, requiring only those two to be determined. Procedures are available whereby other parameters can be determined simultaneously, but they require the use of sophisticated fitting routines to overcome issues of parameter identifiability.[27,28] If nitrification is excluded from the bioreactor, ammonia will build up as ammonification occurs. Consequently, an estimation of k_a can be based on the release of ammonia during the test.[25]

9.5.4 Order of Determination

Evaluation of the parameters and the wastewater characteristics must proceed in a particular order because the values of some are needed before others can be obtained. Table 9.2 summarizes the order of their determination.

9.6 USING TRADITIONAL MEASUREMENTS TO APPROXIMATE WASTEWATER CHARACTERISTICS FOR MODELING

As seen in the preceding section, characterization of a complex wastewater in a manner suitable for use with ASM No. 1 is quite involved and represents a significant investment of time and money. Consequently, such characterizations are not ordinarily done as part of the routine measurements made at wastewater treatment plants. Rather, in the United States, wastewaters are normally characterized in terms of the concentrations of total suspended solids (TSS), volatile suspended solids (VSS), five-day biochemical oxygen demand (BOD_5), ammonia-N, total Kjeldahl nitrogen (TKN), and alkalinity. Total COD is also commonly measured, but the frequency is usually less than that of the other characteristics, although it is increasing. Furthermore, distinction is seldom made between soluble and particulate phases during measurements of BOD_5, COD, and TKN. Because there are circumstances in which it would be advantageous to conduct preliminary modeling studies prior to conducting detailed treatability studies, it would be very useful to be able to translate the traditional data available in the records of wastewater treatment plants into a form that can be used with the models presented herein. Luckily, with a few simplifying assumptions, this can be done for domestic wastewaters. Such translations cannot be made for industrial wastewaters, however, because each is unique.

As indicated by Equation 9.21, the total COD in a wastewater (COD_{TO}) is made up of four components: particulate biodegradable COD (X_{SO}), soluble biodegradable COD (S_{SO}), particulate inert COD (X_{IO}), and soluble inert COD (S_{IO}). As a consequence, data on the total COD in the wastewater is essential to the determination of the other constituents. If no COD data are available, the total COD of domestic wastewater can be approximated as[38,51]

$$COD_{TO} \approx (2.1)(BOD_5). \tag{9.26}$$

The wastewater biodegradable COD (COD_{BO}) can be estimated from the ultimate biochemical oxygen demand (BOD_u), which, in turn, can be estimated from the BOD_5:

$$BOD_u = \frac{BOD_5}{1 - 10^{-5k}}, \tag{9.27}$$

where k is the biochemical oxygen demand (BOD) rate coefficient with units of day^{-1}. For domestic wastewater, the relationship between the ultimate BOD and the five-day BOD can be approximated as[38]

$$BOD_u \approx (1.5)(BOD_5). \tag{9.28}$$

The biodegradable COD is greater than the ultimate BOD because the latter does not account for the electrons retained in the biomass debris formed during the BOD test. Consequently,

$$COD_{BO} = \frac{BOD_u}{(1 - f_D \cdot Y_H)}, \tag{9.29}$$

where Y_H is in COD/COD units and f_D can be assumed to have a value of 0.20 mg debris COD/mg biomass COD.[25] Substituting Equation 9.27 for the ultimate BOD into Equation 9.29 allows the biodegradable COD to be estimated from the five-day BOD:

$$COD_{BO} = \frac{BOD_5}{(1 - f_D \cdot Y_H)(1 - 10^{-5k})}. \tag{9.30}$$

For wastewaters with a significant industrial component, both k and Y_H would have to be measured to allow the conversion, but for domestic wastewater, Y_H can be assumed to be 0.60,[25] and Equation 9.28 can be used as the relationship between the two types of BOD, giving:

$$COD_{BO} \approx (1.14)(BOD_u) \approx (1.71)(BOD_5). \tag{9.31}$$

The division of biodegradable COD into slowly and readily biodegradable fractions requires specific knowledge of the wastewater in question. This is necessary because slowly biodegradable substrate is not necessarily the same as particulate biodegradable substrate, even though it is considered to be particulate in the models. Rather, some of the slowly biodegradable substrate may pass through the filters used to determine VSS concentrations. As discussed earlier, the physical assay of Mamais et al.[35] provides a simple means of determining the readily biodegradable COD in a wastewater. If no other basis is available for making the division, its use is encouraged, even during preliminary studies. If that cannot be done, a "best guess" division must be made based on experience.

The nonbiodegradable, or inert, COD (COD_{IO}) is the difference between the total COD and the biodegradable COD:

$$COD_{IO} = COD_{TO} - COD_{BO}. \tag{9.32}$$

It must be partitioned into soluble (S_{IO}) and particulate (X_{IO}) forms. Experience suggests that 35 to 40% of the particulate organic matter in domestic wastewater is nonbiodegradable.[25,26] Particulate organic matter is represented by the VSS. If one assumes that the elemental composition of the inert particulate organic matter is similar to that of protein, which has a COD equivalent of 1.5 g COD/g protein (Table 3.1), and that protein is totally volatile in a volatile suspended solids test, then

$$X_{IO} \approx (0.375)(1.50)(VSS) = 0.56(VSS). \tag{9.33}$$

The soluble inert COD can be calculated by difference:

$$S_{IO} = COD_{IO} - X_{IO}. \tag{9.34}$$

If one were going to use BOD as a measure of biodegradable organic matter, it would be better to measure the ultimate carbonaceous BOD (BOD_u) than to measure BOD_5, because the relationship expressed by Equation 9.29 is subject to less error and variability than the relationship between biodegradable COD and BOD_5 expressed by Equation 9.30, for which an assumed value of k is required. Although Equation 9.31 can be used as a rough approximation for domestic wastewater, the relationship between biodegradable COD and BOD_5 depends on the rate of oxygen consumption in the BOD test, as indicated by Equation 9.30. Since that rate will be influenced by the nature of the organic chemicals present in the sample being tested, the presence of industrial discharges to a municipal wastewater treatment plant may well change the relationship from that associated with a

strictly domestic wastewater. Thus, there is a high degree of uncertainty associated with Equation 9.31 under that circumstance and it is better to use Equation 9.30 with a measured value of the BOD rate coefficient, k. The best course of action, however, is to use COD as the measure of organic substrates, accounting for the nonbiodegradable material in the ways outlined earlier.

Some of the wastewater characteristics used in modeling are routinely measured directly: ammonia-N, nitrate-N, and alkalinity. The ammonia-N and nitrate-N concentrations can be used without conversion. Most domestic wastewaters contain no nitrate-N, although industrial wastewaters might. Alkalinity is typically measured as $CaCO_3$, but is expressed as mM/L in ASM No. 1 and No. 2. Since the molecular weight of $CaCO_3$ is 100, the conversion is simple.

Without treatability studies, the other nitrogen forms used in modeling must be deduced from the routine measurements made at wastewater treatment plants. The total organic nitrogen concentration (ON_{TO}) is the difference between the TKN and ammonia-N (S_{NHO}) concentrations in the influent:

$$ON_{TO} = TKN - S_{NHO}. \tag{9.35}$$

Furthermore, the wastewater total organic nitrogen can be divided into soluble, particulate, biodegradable, and inert fractions:

$$ON_{TO} = S_{NSO} + S_{NIO} + X_{NSO} + X_{NIO}. \tag{9.36}$$

The concentration of soluble, inert organic nitrogen (S_{NIO}) in domestic wastewaters typically ranges from 1 to 2 mg/L as N,[43] suggesting that a value of 1.5 mg/L as N can be assumed for preliminary modeling without fear of gross error. The particulate inert organic nitrogen is associated with the particulate inert organic matter. The nitrogen content of this material can be assumed to be equal to $i_{N/XD}$, the nitrogen content of biomass debris. Consequently, the concentration of particulate inert organic nitrogen (X_{NIO}) can be approximated as

$$X_{NIO} \approx i_{N/XD} \cdot X_{IO}. \tag{9.37}$$

When possible, distribution of biodegradable organic nitrogen between the soluble and particulate phases should be based on data. In the absence of specific data, the biodegradable organic nitrogen is often distributed into particulate and soluble fractions in the same proportions as the slowly and readily biodegradable fractions of the COD, as indicated in Equation 9.23.

The major difficulty in moving between COD and TSS units for suspended solids is the presence of fixed suspended solids (FSS) in wastewaters:

$$FSS = TSS - VSS. \tag{9.38}$$

Fixed suspended solids are inorganic and thus have no COD. As a consequence, they are not considered when particulate concentrations are expressed on either a COD or a VSS basis. However, they must be considered when the concentrations of particulate materials, such as MLSS, are expressed in TSS units. Fixed suspended solids undergo no reactions in biochemical operations. Rather, they behave like inert organic solids, which are discussed in Section 5.2.2. Consequently, when the MLSS concentration is being calculated in TSS units, the influent FSS concentration, given the symbol X_{FO}, should be used in addition to the influent inert organic solids, X_{IO}, in the appropriate equations. It should be handled in exactly the same manner as influent inert organic solids in all computations.

Example 9.6.1

Conventional characterization of a domestic wastewater following primary clarification is given in the upper portion of Table E9.4. Translate that information into a form that can be used in ASM No. 1.

a. The first task is to estimate the concentration of biodegradable COD. This is done by using the BOD_5 and Equation 9.31:

$$COD_{BO} = (1.71)(155) = 265 \text{ mg COD/L}.$$

b. The inert COD concentration can then be calculated from the total COD using Equation 9.32:

$$COD_{IO} = 325 - 265 = 60 \text{ mg COD/L}.$$

c. The concentration of particulate inert COD can be estimated from the VSS concentration using Equation 9.33:

$$X_{IO} = (0.56)(61.5) = 35 \text{ mg COD/L}.$$

d. This, in turn allows the soluble inert COD concentration to be calculated from Equation 9.34:

$$S_{IO} = 60 - 35 = 25 \text{ mg COD/L}.$$

TABLE E9.4
Translation of Traditional Wastewater Characteristics into a Form Suitable for Modeling

Component	Concentration
Conventional Wastewater Characterization	
TSS	82 mg/L
VSS	61.5 mg/L
BOD_5	155 mg/L
Total COD	325 mg/L as COD
Ammonia-N	25 mg/L as N
Total Kjeldahl Nitrogen (TKN)	43.5 mg/L as N
Nitrate-N	0.0 mg/l as N
Alkalinity	200 mg/L as $CaCO_3$
Characterization as Required for Use in ASM No. 1	
Particulate inert organic matter	35 mg/L as COD
Soluble inert organic matter	25 mg/L as COD
Slowly biodegradable substrate	150 mg/L as COD
Readily biodegradable substrate	115 mg/L as COD
Oxygen	0 mg/L as O_2
Soluble nitrate nitrogen	0 mg/L as N
Soluble ammonia nitrogen	25 mg/L as N
Soluble biodegradable organic nitrogen	6.5 mg/L as N
Particulate biodegradable organic nitrogen	8.5 mg/L as N
Alkalinity	2 mM/L

Note: The wastewater is considered to be typical of domestic wastewater that has undergone primary sedimentation.

e. Partitioning of the biodegradable COD into slowly and readily biodegradable fractions requires some knowledge of the nature of the wastewater. Additional information suggests that 43% of the biodegradable COD is readily biodegradable. Consequently,

$$S_{SO} = (0.43)(265) = 115 \text{ mg COD/L}$$

and

$$X_{SO} = 265 - 115 = 150 \text{ mg COD/L.}$$

f. The concentration of organic nitrogen in the wastewater can be obtained as the difference between the TKN and the ammonia-N concentrations as expressed in Equation 9.35:

$$ON_{TO} = 43.5 - 25 = 18.5 \text{ mg N/L.}$$

g. The biodegradable organic nitrogen concentration must be obtained by subtracting the concentrations of the soluble and particulate inert organic nitrogen. The concentration of soluble inert organic nitrogen can be assumed to be 1.5 mg N/L, as discussed previously. The concentration of particulate inert organic nitrogen can be estimated with Equation 9.37 by assuming a value for $i_{N/XD}$. A reasonable value is 0.06 mg N/mg COD, as indicated in Table 6.3. Consequently:

$$X_{NIO} = (0.06)(35) = 2 \text{ mg N/L.}$$

Use of Equation 9.36 gives the biodegradable organic nitrogen concentration:

$$S_{NSO} + X_{NSO} = 18.5 - 1.5 - 2.0 = 15 \text{ mg N/L.}$$

h. Partitioning of the biodegradable organic nitrogen in accordance with Equation 9.23 gives:

$$S_{NSO} = (15.0)\left(\frac{115}{150+115}\right) = 6.5 \text{ mg N/L.}$$

Consequently,

$$X_{NSO} = 15.0 - 6.5 = 8.5 \text{ mg N/L.}$$

The estimated characteristics of the wastewater are listed in the lower portion of Table E9.4. Comparison of them to the values in Table 6.6 shows that with the exception of alkalinity they are the same. The alkalinity is a very site-specific characteristic, being dependent on the nature of the carriage water.

9.7 KEY POINTS

1. Care should be exercised in the design of a treatability study because the manner in which bioreactors are configured and operated has a strong effect on the microbial community that develops, thereby influencing the values of the kinetic and stoichiometric parameters obtained from the study.
2. During treatability studies to evaluate the parameters in the simple model of Chapter 5, data related to effluent quality and biomass concentrations should be collected from continuous stirred tank reactors (CSTRs) operated over a range of solids retention times (SRTs). Biomass concentrations may be expressed as chemical oxygen demand (COD), total suspended solids (TSS), or volatile suspended solids (VSS) as long as appropriate conversion factors are used to allow COD balances to be made.

3. Because of the interrelationships among the parameters in the simple model of Chapter 5, their values must be estimated in a certain order. The values of $Y_{H,T}$ and b_H are evaluated together, but b_H must be known before f_D and the Monod parameters ($\hat{\mu}_H$ and K_S) can be evaluated. The concentration of inert soluble COD, S_I, must also be known before the Monod parameters can be estimated.
4. Nonlinear parameter estimation techniques are preferable, but several methods of linearizing the Monod equation are in common use for determining the values of the kinetic parameters in it. The efficacies of those linearizations depend on the nature of the error in the data set.
5. The active fraction of the biomass is difficult to measure and thus is not routinely measured during treatability studies. Luckily, f_D does not vary greatly, allowing a value of 0.20 to be assumed with little risk of error.
6. When data are not available concerning the active fraction of the biomass in the CSTRs used for treatability studies, a separate batch experiment must be performed to determine b_H. The experiment measures the change in oxygen uptake rate (OUR) of the biomass over time. Once the value of b_H is known, it may be used with an assumed f_D value to determine $Y_{H,T}$. The Monod kinetic parameters can then be evaluated in the same manner as previously described.
7. Consideration must be given to the physiological state of a culture and how it may change during batch experiments for determining biodegradation kinetics for single organic compounds. If the physiological state is not allowed to change, the resulting parameter values are called "extant" parameters because they reflect the conditions of the culture in the bioreactor from which they were removed prior to testing. If the test conditions allow the enzyme system of the culture to develop to the point that the microorganisms can grow at the fastest rate possible on the test substrate at the given temperature and pH, the kinetic parameters are called "intrinsic."
8. Both intrinsic and extant biodegradation kinetic parameter values can be determined during batch experiments, although the experimental conditions required are quite different. To obtain intrinsic parameter estimates the initial substrate to relevant biomass ratio should be at least 20 when both are measured as COD. In contrast, to obtain extant parameter estimates the ratio should be less than 0.02. For both estimates, the initial substrate concentration should exceed the expected K_S value.
9. Because of the large number of parameters required by activated sludge model (ASM) No. 1, extensive treatability testing is required to fully evaluate them. Several CSTRs with biomass recycle should be run over a range of SRTs in excess of five days to provide data for evaluating biomass production and electron acceptor requirement. In addition, two CSTRs in series should be operated as a modified Ludzak-Ettinger (MLE) system to provide biomass capable of denitrification.
10. Characterization of a wastewater and estimation of the stoichiometric coefficients requires data from CSTRs operated over a range of SRTs. The estimate of Y_H must be obtained from an experiment in which a small amount of biomass is used to seed a batch bioreactor containing only the soluble portion of the wastewater. The concentration of inert soluble COD is estimated in the same way as for the simple model. The concentration of readily biodegradable COD can either be obtained from OUR data from a batch reactor or by an assay involving coagulation and filtration. The concentration of inert particulate organic matter is estimated by fitting a simplified model to data relating the mixed liquor suspended solids (MLSS) concentration in the CSTRs to their SRTs. The concentration of slowly biodegradable COD can be calculated from the total influent COD and the previously measured COD values.
11. The most important kinetic parameter in ASM No. 1 is $\hat{\mu}_A$, the maximum specific growth rate coefficient for nitrifying bacteria. Thus, it is important that it be determined in the

wastewater matrix. This can be done by measuring the increase in nitrate-N and nitrite-N concentrations over time in a batch growth experiment and plotting $\ln(NO_3^-\text{-N} + NO_2^-\text{-N})$ versus time. The slope of the resulting plot is $\hat{\mu}_A - b_A$, thereby allowing the value of $\hat{\mu}_A$ to be calculated. Care must be exercised in the selection of b_A because it has a direct impact on the estimated value of $\hat{\mu}_A$.

12. Tests to measure η_g, the correction factor for heterotrophic growth under anoxic conditions and η_h, the correction factor for hydrolysis of slowly biodegradable substrate under anoxic conditions, are performed at the same time by evaluating oxygen and nitrate consumption rates in two batch bioreactors that are equivalent in every respect except for the terminal electron acceptor.
13. The parameters characterizing hydrolysis of slowly biodegradable substrate, k_h and K_X, can be evaluated with OUR data collected during a batch experiment with biomass from one of the CSTRs. Because all other parameters are known, k_h and K_X can be estimated by the nonlinear curve fitting of ASM No. 1 to the OUR data.
14. By using appropriate approximations, it is possible to use traditional wastewater characteristics (TSS; VSS; five-day biochemical oxygen demand, BOD_5; ammonia-N; and total Kjeldahl nitrogen, TKN) to estimate the concentrations of the constituents required to use ASM No. 1. However, specific information on the particular wastewater in question is required to make the split between readily and slowly biodegradable substrate concentrations.

9.8 STUDY QUESTIONS

1. Explain why treatability studies are often run in stages, especially for design of more complex systems.
2. Describe how you would modify the procedures presented in Section 9.2 to provide the information required to design a system to achieve nitrification in addition to carbon oxidation. In your description, tell how you would estimate the kinetic parameters describing nitrification.
3. Discuss the efficacy of the various techniques for estimating the Monod parameters ($\hat{\mu}_H$ and K_S) from experimental data relating the effluent soluble biodegradable COD to the SRT of the bioreactors.
4. A treatability study was performed on a wastewater using a CSTR with cell recycle. The wastewater was totally soluble and had a COD of 474 mg/L. The studies were run in a lab-scale bioreactor that had a volume of 6.0 L. The flow rate was maintained at a constant value of 1.0 L/hr and the SRT was maintained at the desired values by wasting excess biomass directly from the bioreactors. A batch experiment with mixed liquor removed from the CSTR revealed that the concentration of inert soluble COD in the wastewater is 24 mg COD/L. Using the data provided in Table SQ9.1, determine the parameters $\hat{\mu}_H$, K_S, $Y_{H,T}$, b_H, and f_D. Use all three of the linearization techniques for estimating $\hat{\mu}_H$ and K_S and compare their effectiveness. Could the first-order approximation (i.e., $k_{e,T}$) be used to characterize this wastewater over the SRT range studied? Why?
5. Explain why nitrification does not interfere with the determination of b_H from OUR measurements in a batch bioreactor when nitrification is well established in the CSTR from which the biomass was obtained, but does interfere when only partial nitrification occurs in the CSTR.
6. A treatability study was performed on a wastewater using CSTRs with cell recycle. The wastewater was totally soluble and had a COD of 286 mg/L. The studies were run in lab-scale bioreactors with a volume of 8.0 L. The flow rate was maintained at a constant value of 2.0 L/hr and the SRT was maintained at the desired values by wasting excess biomass directly from the bioreactors. Data collected from the CSTRs are shown in Table SQ9.2.

TABLE SQ9.1
Data Collected from CSTRs with Cell Recycle during a Treatability Study of a Soluble Wastewater

SRT Hrs	Soluble COD mg COD/L	Biomass mg TSS/L	Active Fraction
48	62.0	1150	0.95
96	45.8	1980	0.90
144	41.1	2590	0.85
192	38.8	3100	0.81
288	36.7	3920	0.74
384	35.6	4610	0.68

TABLE SQ9.2
Data Collected from CSTRs with Cell Recycle during a Treatability Study of a Soluble Wastewater

SRT Hrs	Soluble COD mg COD/L	Biomass mg TSS/L
48	41.7	1280
96	39.5	2150
144	38.8	2820
192	38.5	3390
288	38.2	4370
384	38.0	5230

TABLE SQ9.3
Data Collected during a Batch Aeration Test for the Determination of b_H

Time Hrs	OUR mg O_2/(L · hr)
0	37.0
12	33.4
24	30.2
36	27.2
48	24.6
72	20.1
96	16.4
120	13.3
144	10.9
192	7.2
240	4.8

TABLE SQ9.4
Characteristics of a Domestic Wastewater in Traditional Terms

Component	Concentration
TSS	125 mg/L
VSS	100 mg/L
BOD_5	225 mg/L
Total COD	475 mg/L as COD
Ammonia-N	35 mg/L as N
Total Kjeldahl Nitrogen (TKN)	60 mg/L as N
Nitrate-N	0.0 mg/l as N
Alkalinity	200 mg/L as $CaCO_3$

No data on the active fraction of the biomass were taken during the study. A batch experiment with mixed liquor removed from one CSTR revealed that the concentration of inert soluble COD in the wastewater is 36 mg COD/L. Another batch experiment with biomass taken from the CSTR with an SRT of 144 hr was performed for the determination of b_H. The results are shown in Table SQ9.3. Using the available data, determine the values of b_H and $Y_{H,T}$ describing the biomass.

7. Explain what is meant by the term "physiological state" and why it will influence the values of the kinetic parameters describing biodegradation of an organic compound when those parameters are measured in batch experiments.
8. Explain why oxygen consumption measurements can be used as a surrogate for measurements of biomass growth or substrate utilization during batch tests for determining intrinsic or extant kinetic parameters.
9. Describe the types of bioreactors that should be operated and the type of data that should be collected during treatability studies to evaluate the parameters in ASM No. 1.
10. Describe how the various fractions of the wastewater COD (inert soluble, inert particulate, readily biodegradable, and slowly biodegradable) may be estimated during treatability studies for use of ASM No. 1.
11. Describe and contrast the procedures used to evaluate the kinetic parameters for aerobic growth of heterotrophs and aerobic growth of autotrophs for use in ASM No. 1.
12. Describe the procedure for estimating the correction factors for growth and hydrolysis under anoxic conditions, η_g and η_h, for use in ASM No. 1.
13. Describe the procedure for estimating the parameters describing hydrolysis of slowly biodegradable substrate under aerobic conditions.
14. A domestic wastewater has the characteristics listed in Table SQ9.4. Use those characteristics to estimate the concentrations of the various constituents required for using ASM No. 1. A listing of those constituents is provided in the lower portion of Table E9.4. Assume that the readily biodegradable substrate is 35% of the total biodegradable COD.

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Part III

Applications: Suspended Growth Reactors

Part I presents the fundamental principles upon which the design and evaluation of the biochemical operations used in wastewater treatment systems are based. These principles are then applied in Part II to the modeling of ideal suspended growth bioreactors. In Part III we apply these principles to the practical design and operation of suspended growth biological wastewater treatment systems. Chapter 10 provides an overview of the design and evaluation of these systems, whereas the remaining chapters address specific suspended growth applications. Chapter 11 describes the design of activated sludge systems for the removal of biodegradable organic matter, the stabilization of particulate organic matter, and the oxidation of ammonia-N. The use of aerobic selectors to control the growth of certain types of filamentous bacteria is also considered. Chapter 12 addresses the design and operation of suspended growth biological nutrient removal systems. Single-sludge nitrogen removal, phosphorus removal, and combined nitrogen and phosphorus removal systems are considered, along with separate stage denitrification systems. The use of anoxic and anaerobic zones to control solids settleability is also addressed. The use of aerobic digestion to stabilize the waste solids (both primary and secondary) produced in the liquid process train of a wastewater treatment plant is the topic of Chapter 13. Conventional aerobic digestion systems are considered, along with anoxic/aerobic digestion systems and autothermal thermophilic aerobic digesters. Chapter 14 addresses the use of anaerobic processes for the treatment of high strength wastewaters and sludges. Both suspended growth and combined suspended and attached growth processes are considered. Finally, pond and lagoon systems are considered in Chapter 15. The environments in these systems are complex and deviate more than the environments in the other named biochemical operations from the ideal reactors considered in Part II. In spite of that, the fundamental principles developed in Parts I and II of this book can be applied to their design and operation.

10 Design and Evaluation of Suspended Growth Processes

Most of this book addresses the technical aspects of the design and evaluation of the biochemical operations used in wastewater treatment systems. This chapter, in contrast, addresses the process of designing and evaluating such operations, with particular emphasis on those that use suspended growth bioreactors. It has several purposes. First, it provides a transition between the fundamental principles presented in Parts I and II of this book and the detailed application of those principles to specific named biochemical operations in Part III. Second, it illustrates that the design and evaluation of biochemical operations is iterative and provides a perspective on the typical steps involved. Third, it addresses the basic decisions that must be made to select among the various suspended growth biochemical operations, as well as those that are common to all of them. Finally, it contrasts the various levels of design and evaluation, from preliminary to simulation-based. Even though this material is presented in the context of suspended growth bioreactors, much of it is also applicable to the design and evaluation of attached growth bioreactors. Consequently, the reader should also refer to this material while reading Part V.

The term "design and evaluation" is used here to reflect the range of tasks that biological process engineers must perform. In some instances, new facilities, or significant expansions of existing facilities, are needed to provide sufficient treatment capacity and/or capability. The term "design" refers to the process of determining the size and configuration of such new facilities. In other instances, a facility may already exist, but its treatment capacity and/or capability may not be known precisely. The term "evaluation" refers to the process of rationally determining that capacity and/or capability. The same process engineering principles are utilized in both situations.

10.1 GUIDING PRINCIPLES

Before investigating the design and evaluation of biochemical operations, it would be helpful to summarize the basic fundamental principles that arose in Parts I and II. These few essential principles, which provide the basis for all design and evaluation, are summarized in Table 10.1.

First, the biochemical environment imposed upon a bioreactor determines the nature of the microbial community that develops and the character of the biological reactions that they perform. If the engineer ensures that a high concentration of dissolved oxygen is maintained at all times, then organic substrates can be oxidized to carbon dioxide and water, providing energy for heterotrophic biomass growth and removing chemical oxygen demand (COD) from solution. In addition, ammonia-N can be oxidized to nitrate-N by autotrophic nitrifying bacteria, providing for their growth as well. The introduction of an anoxic zone in an otherwise aerobic bioreactor allows the nitrate-N formed by the autotrophic bacteria to be used as a terminal electron acceptor by facultative heterotrophic bacteria, converting it to nitrogen gas, thereby removing nitrogen from the wastewater. Furthermore, a properly positioned anaerobic zone in an otherwise aerobic bioreactor will allow phosphate accumulating organisms (PAOs) to effectively compete with ordinary heterotrophic bacteria for substrate, leading to a biomass enriched in phosphate. Wastage of that biomass removes phosphorus from the system. Finally, if the engineer provides a totally anaerobic environment in which neither oxygen nor nitrate-N is ever present, then an entirely different microbial community will develop in which methane is an important end product. The key point is that the engineer has control over the type of

TABLE 10.1
Guiding Principles for the Design and Evaluation of Suspended Growth Biochemical Operations

1. The biochemical environment determines the nature of the microbial community that develops in a bioreactor and the character of the reactions they perform.
2. The SRT is the most important design and control parameter available to the engineer.
3. A COD balance across a bioreactor provides valuable information about the amount of electron acceptor required and the amount of excess biomass produced.
4. The excess biomass production rate is essentially the same for all suspended growth systems with a given SRT and biochemical environment, regardless of the bioreactor configuration.
   a. The total mass of biomass in such systems will be the same, regardless of bioreactor configuration.
   b. The total mass of electron acceptor required for the removal of organic matter will be same in such systems, regardless of bioreactor configuration, although the distribution of need will be different.
5. Only the mass of biomass in a bioreactor system is specified by the descriptive analytical expressions, not the concentration. The concentration is only specified after the bioreactor volume, or HRT, has been specified.

microbial community that may be present through the decisions that are made about the biochemical environment.

Second, the solids retention time (SRT) is the most important design and control parameter available to the engineer. This follows directly from its relationship to the specific growth rate of the biomass in the bioreactor, as reflected by Equation 5.21 for a simple continuous stirred tank reactor (CSTR). Thus, while the biochemical environment provides the potential for growth of a given microbial population, the SRT, in concert with the bioreactor configuration, determines whether that potential will be realized. Furthermore, the SRT and the bioreactor configuration determine the extent of reaction in the system, thereby influencing the effluent substrate concentration, the excess biomass production rate, the rate at which electron acceptor must be provided, and the overall process performance.

Third, a COD balance across a bioreactor provides valuable information about the amount of electron acceptor required and the amount of excess biomass produced. As stated in Equation 3.94, the COD removed in a bioreactor must equal the oxygen equivalents of the terminal electron acceptor used plus the COD of biomass formed. Furthermore, at steady state, the amount of biomass formed is equal to the observed yield ($Y_{Hobs,T}$) times the amount of substrate used, as reflected by Equation 5.41 for a simple CSTR. This means that the oxygen equivalent of the terminal electron acceptor required is just equal to ($1 - Y_{Hobs,T} \cdot i_{O/XB,T}$) times the amount of substrate used, as stated by Equation 5.45. The parameter $i_{O/XB,T}$ is the unit COD of biomass. Thus, it can be seen that this simple balance is very useful for making initial estimates of excess biomass production and electron acceptor requirements.

Fourth, the excess biomass production rate is essentially the same for all suspended growth systems with a given SRT and biochemical environment, regardless of the bioreactor configuration.

This was seen in the simulations of the various systems in Chapter 7 and has two important implications. First, the total mass of biomass in those systems will be the same, regardless of the system configuration. This means that expressions derived for a single CSTR, which can be solved analytically, can be used to estimate the total mass of biomass or mixed liquor suspended solids (MLSS) in a complex bioreactor system for which analytical solutions are not possible. Second, the total amount of electron acceptor required for removal of organic matter will be independent of the bioreactor configuration, although the distribution of the electron acceptor will not be. The same statement cannot be made about the amount of oxygen required for nitrification, however, because the extent of nitrification is much more dependent on the bioreactor configuration than is the extent of COD removal, as seen in Chapter 7. Nevertheless, these are very powerful tools, especially for preliminary designs.

Finally, Equation 5.30 demonstrated that only the mass of biomass in a bioreactor system is specified by the descriptive analytical equations, not the concentration. Rather, the concentration is only specified after the bioreactor volume, or the hydraulic residence time (HRT), has been specified. The same is true for the MLSS as well. From a design and analysis perspective, this means that a designer has one free design variable that may be freely chosen, within reasonable limits. A large part of the design process is concerned with that choice because it strongly influences the potential interactions among the various unit operations that must be considered during system design.

10.2 ITERATIVE NATURE OF PROCESS DESIGN AND EVALUATION

The design of biological wastewater treatment systems is typically an iterative process. There are two reasons for this: (1) the definition of the problem to be solved evolves throughout the design, and (2) the database upon which the design is based improves as additional investigations are completed. Nevertheless, a "freeze" point must be reached during any project where the solution to the problem is fixed and then implemented. The steps that lead to this point are evolutionary in nature, with the problem statement continuously being redefined and potential solutions being evaluated and discarded until the "best" solution is selected. Most designs begin with an initial concept that is rather general in nature. For example, during early discussions for a design, a decision may be made to treat a particular wastewater in an activated sludge system, even though the size of the system, its specific configuration, and the nature of any special features are unknown. Those questions, and many more, are addressed as the design proceeds, resulting in an ever more refined estimate of the required facilities. The increasing database that develops during the design process also allows refinement of the design. Typically the designer's understanding of the strength and nature of the wastewater, the characteristics of the treatment system, the effluent discharge standards, and the needs and desires of the treatment system owner and operator evolve as the project progresses.

Figure 10.1 illustrates this iterative nature of process design. The first step is to define the project objectives and requirements, allowing identification of the most reasonable potential solutions based on the current state of knowledge. The costs and scope of those potential solutions are then estimated using rough calculations of bioreactor size, oxygen requirements, solids wastage, and so on. Next, the potential advantages and disadvantages of the alternative solutions are considered and a decision is made whether to more fully evaluate each one. When the potential advantages of a particular alternative solution are not sufficient to warrant further consideration, it is dropped and more study is devoted to those remaining. Each iteration around the loop results in more refined information, which allows better estimates to be made of the size and cost of the alternatives. Consequently, the use of more refined techniques is called for. This same logic can be applied to the refinement of a selected alternative. Additional studies to refine a given alternative are conducted only as long as the benefits derived from them outweigh their costs.

The iterative nature of process design and evaluation makes it clear that several levels of refinement are required. To begin the design process, a preliminary assessment must be made based on

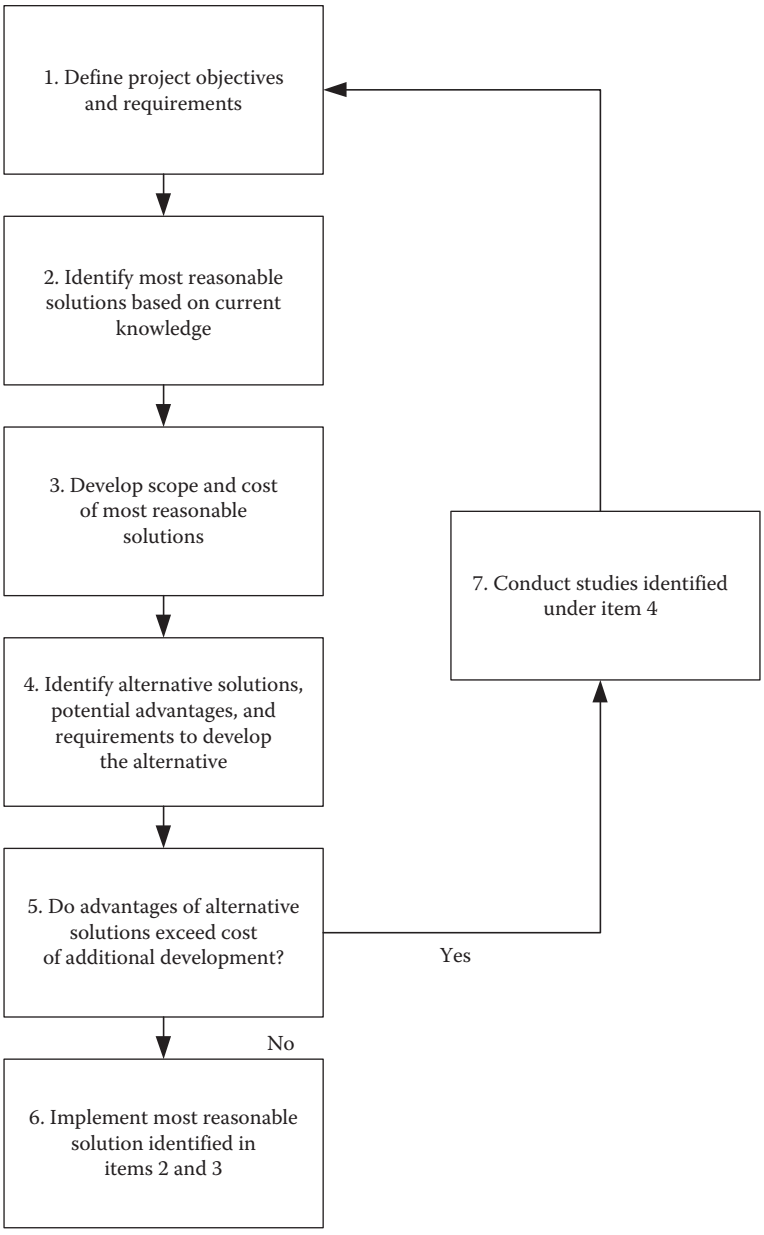

FIGURE 10.1 The iterative nature of process design and evaluation.

limited data. In spite of its preliminary nature, this assessment must be conceptually sound because important decisions will be based on it. In some instances, the preliminary assessment may be sufficiently precise to allow the project to proceed directly to implementation. This will usually occur for smaller projects where the cost of a conservative design is small compared to the cost of refining the estimates of facility requirements. It also occurs frequently for applications where significant experience already exists upon which to base the selection of the preliminary process design parameters. In other instances, little experience may exist with the subject wastewater or proposed treatment system, making the initial preliminary assessment quite uncertain. In that case treatability studies as outlined in Chapter 9 must be performed, leading to parameters that may be used in models. In some cases, particularly for the activated sludge process, it may be possible

to base design decisions on the simple stoichiometric model of Chapter 5 by incorporating a few broadening assumptions, such as grouping together slowly and readily biodegradable substrate. In other cases, such as large nutrient removal projects, even a small amount of uncertainty can result in significant over-expenditures. In those situations, additional testing to quantify the parameters in Activated Sludge Model (ASM) No. 1, 2, or 2d may be merited, allowing alternative designs to be considered by simulation, thereby reducing uncertainty. In the next section we will consider the decisions that must be made in any design situation, regardless of its level. Then, in Section 10.4 we will consider the approaches used in the various levels of design.

10.3 BASIC DECISIONS DURING DESIGN AND EVALUATION

Design of a biochemical operation requires that decisions be made that are consistent with the guiding principles summarized in Section 10.1. Some of those decisions establish the nature of the facility, whereas others determine its size. Since the biochemical environment has a profound effect, its choice is one of the earliest decisions that must be made. Then the SRT is chosen, determining the various factors that follow from it, such as the mass of biomass, electron acceptor requirement, and so on. Finally, the interrelationships between the bioreactor and the other unit operations in the system must be considered.

10.3.1 Biochemical Environment

One of the fundamental decisions faced by a designer is whether to use an aerobic/anoxic or an anaerobic operation. As discussed in Section 2.3.2, in aerobic/anoxic operations, heterotrophic bacteria use oxygen or nitrate-N as their terminal electron acceptor while using biodegradable organic matter as an energy and carbon source for growth. Furthermore, the presence of dissolved oxygen in such systems allows for the growth of autotrophic nitrifiers, which use ammonia-N as an electron donor, producing nitrate-N. In contrast, as discussed in Section 2.3.3, when both dissolved oxygen and nitrate-N are absent, alternative electron acceptors must be used. In fermentative systems, the biodegradable organic matter itself serves as the terminal electron acceptor, yielding soluble fermentation products, whereas in methanogenic systems carbon dioxide is the major acceptor, yielding methane.

Table 10.2 compares the features of aerobic/anoxic and anaerobic wastewater treatment systems. Both systems are capable of achieving high organic removal efficiencies. However, the effluent quality from an aerobic/anoxic system will generally be excellent while that from an anaerobic

TABLE 10.2
Comparison of Aerobic/Anoxic and Anaerobic Systems

	System	
Feature	**Aerobic/Anoxic**	**Anaerobic**
Organic removal efficiency	High	High
Effluent quality	Excellent	Moderate
Sludge production	High	Low
Nutrient requirements	High	Low
Energy requirements	High	Low to moderate
Temperature sensitivity	Low	High
Methane production	No	Yes
Nutrient removal	Possible	Negligible

system will be moderate. Aerobic and anoxic conditions allow extensive removal of biodegradable organic matter, particularly soluble material. In addition, the biomass in aerobic/anoxic systems is generally well flocculated, resulting in low effluent suspended solids concentrations. In contrast, although a high percentage of the biodegradable organic matter is converted to methane and carbon dioxide in anaerobic systems, the resulting concentrations of soluble biodegradable organic matter can still be relatively high and the produced solids may be poorly flocculated. As a result, the quality of the effluent from an anaerobic system does not generally equal that from an aerobic system.

Waste solids production is high in aerobic/anoxic systems due to the large amount of energy made available for the synthesis of new biomass, resulting in relatively high yield values. Consequently, nutrient requirements are also high. In contrast, the biomass production and associated nutrient requirements for anaerobic systems are low because the relatively small amount of available energy makes the yield low. Power requirements for aerobic systems are high because oxygen must be transferred to serve as the electron acceptor, although this need will be reduced when anoxic zones are present. In contrast, the power requirements for anaerobic systems are low to moderate and generally represent the energy required to heat and mix the bioreactor. Heating requirements can be significant, but energy for heating is typically provided by the methane produced. Temperature control is critical in many (but not all) anaerobic systems because the methanogens are quite sensitive to changes in temperature. If the temperature is uniformly low, adequate performance can be achieved by designing and operating at a very long SRT to compensate for the low maximum specific growth rate associated with the low temperature. The performance of aerobic systems, on the other hand, is less sensitive to changes in temperature and the SRTs needed to achieve necessary performance at low temperatures are not as extreme as required for anaerobic systems. Finally, removal of nitrogen and phosphorus is possible in aerobic/anoxic systems, whereas nutrient removal is negligible in anaerobic systems.

These features combine to provide advantages to aerobic systems for the treatment of low strength wastewaters and to anaerobic systems for the treatment of high strength wastewaters. Figure 10.2 presents the wastewater concentration ranges over which aerobic/anoxic and anaerobic bioreactors

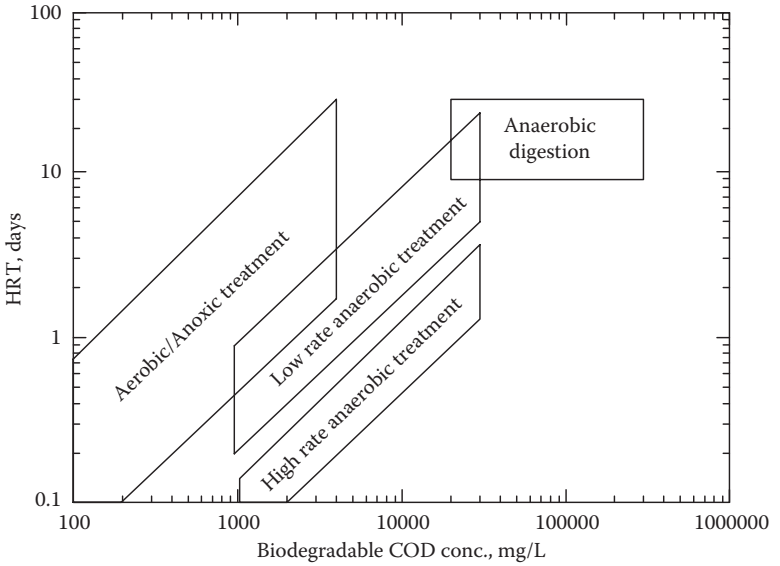

FIGURE 10.2 Typical operating ranges for aerobic/anoxic and anaerobic suspended growth biochemical operations. (Adapted from Hall, E. R., Anaerobic treatment of wastewaters in suspended growth and fixed film processes. *Design of Anaerobic Processes for the Treatment of Industrial and Municipal Wastes*, 41–118, eds. J. F. Malina Jr. and F. G. Pohland, Technomic Publishing Co., Inc., Lancaster, PA, 1992.)

are typically applied and the ranges of HRT typically required. Both ranges are approximate and are provided only as general descriptors. The HRT range reflects both the range of SRTs required and the degree of separation between the SRT and the HRT achieved with each technology. Due to their ability to produce high quality effluents, aerobic/anoxic systems are typically used for wastewaters with biodegradable COD concentrations less than 1000 mg/L. Although anaerobic systems can be applied to treat wastewaters in this concentration range, the effluent quality will generally not meet discharge standards, thereby requiring aerobic polishing. However, the combination of an anaerobic system followed by an aerobic system is usually not economical compared to a fully aerobic system for these wastewaters. In addition, low strength wastewaters typically result in insufficient methane production to heat the wastewater to the optimum temperature. Both aerobic/anoxic and anaerobic systems are used to treat wastewaters with biodegradable COD concentrations between 1000 and 4000 mg/L. Again, aerobic polishing of the anaerobic process effluent will be required if high quality is needed. Finally, in many instances the advantages of anaerobic systems outweigh the advantages of aerobic/anoxic systems for the treatment of wastewaters with biodegradable COD concentrations over 4000 mg/L. The typical operating ranges for various anaerobic treatment systems are also presented in Figure 10.2. In general, low-rate and high-rate anaerobic systems employ biomass recycle to increase the SRT relative to the HRT, whereas anaerobic digestion systems do not, making the SRT and HRT identical.

Additional factors will also influence the relative economics of aerobic/anoxic versus anaerobic systems. Consequently, investigations must be conducted to distinguish the relative advantages and disadvantages of each biochemical environment for wastewaters with biodegradable COD concentrations near the overlap region in Figure 10.2. These include the wastewater temperature, flow rate, and composition and they will be discussed in greater detail in Chapters 11, 12, and 14. Nevertheless, Figure 10.2 can be used for preliminary screening of biological wastewater treatment options.

10.3.2 Solids Retention Time

As illustrated in Part II of this book, the SRT exerts a dominant effect on the capabilities and performance of a biochemical operation. For example, it affects the types of microorganisms that can grow in a bioreactor, as well as their activity, thereby determining effluent quality. Because of the multiple effects associated with SRT, many factors must be considered during its selection. In fact, it is seldom possible to select the SRT based on a single criterion, such as effluent substrate concentration. The range of typical SRT values is already known for many applications, and often an appropriate SRT can be selected based on experience. In this section we will consider such situations.

Before addressing appropriate SRT values, it should be emphasized that the selected SRT must always exceed the minimum SRT associated with the microorganisms responsible for a particular required biochemical transformation. As discussed in Section 5.1.5, the minimum SRT is the value below which a particular group of microorganisms is unable to grow in a suspended growth bioreactor. As expressed in Equation 5.25, it is a function of the influent concentration of the limiting substrate for the microorganisms of interest and the kinetic parameters describing their growth on that substrate. The kinetic parameter that exerts the most pronounced effect is the maximum specific growth rate coefficient, $\hat{\mu}$. Since the $\hat{\mu}$ value for heterotrophs growing on readily biodegradable substrate is high, the minimum SRT for them is very low. In contrast, because the $\hat{\mu}$ value for autotrophic nitrifying bacteria is very low, the minimum SRT associated with them may be quite high. The same may be true for heterotrophs growing on xenobiotic chemicals. If the SRT is maintained at a value less than the minimum SRT for the subject bacteria, they will be wasted from the bioreactor faster than they grow and a stable population will not develop. In other words, washout occurs, as discussed in Section 5.1.5. Conversely, if the operating SRT exceeds the minimum SRT, then the subject bacteria will be able to grow in the process and the reaction will occur. However,

as seen in Chapters 5 and 7, the degree of conversion will depend on the operating SRT and the bioreactor configuration, and both must be chosen to meet effluent quality goals. This generally requires the operating SRT to be well above the minimum SRT. The ratio of the operating SRT to the minimum SRT is called the safety factor (or alternately the design or operating factor, depending upon which activity is underway). To ensure that their washout does not occur, the safety factor for the most slowly growing microorganisms required in a bioreactor should always exceed 1.5, although larger values may be required in some circumstances. Furthermore, larger values may result when other factors control the choice of the SRT. Factors affecting the choice of the SRT for various named biochemical operations will be discussed in subsequent chapters. Only a brief overview is given here.

10.3.2.1 Aerobic/Anoxic Systems

Figure 10.3 illustrates the ranges of operating SRTs over which various events will occur in aerobic/anoxic systems. Because the ranges represent operating SRTs, the lower limits reflect the application of typical safety factors to the minimum SRTs associated with the microorganisms responsible for a particular event. The upper limits reflect SRT values above which little additional reaction occurs in a CSTR. Because ordinary heterotrophs and PAOs can grow under both aerobic and anoxic conditions, the SRT values in Figure 10.3 can be thought of as any combination of aerobic and anoxic SRT values. On the other hand, nitrifying bacteria can grow only under aerobic conditions. Consequently, the SRT values in Figure 10.3 should be thought of as aerobic SRTs when considering nitrification.

The first thing to be noted in Figure 10.3 is that removal of biogenic soluble organic matter occurs at low SRTs, typically over a range of about 0.5 to 1.5 days. For SRT values in excess of this range, the degradation of soluble organic matter will be essentially complete. This follows from the fact that the $\hat{\mu}$ values for heterotrophic bacteria growing under aerobic/anoxic conditions on such substrates are relatively high, as discussed in Section 3.2.10. Furthermore, in municipal wastewater treatment systems, bacterial growth and substrate removal are assisted by the presence of microorganisms in the influent wastewater, which prevents washout, as discussed in Section 5.2.3. The solubilization and metabolism of particulate organic matter typically occurs over a SRT range of 2 to 4 days, with degradation being essentially complete at longer SRT values. Stabilization of biomass through decay and similar reactions will occur over a broad range of SRTs, but is generally thought to be insignificant for SRT values less than about 10 days, as illustrated in Figure 5.10. Increasing

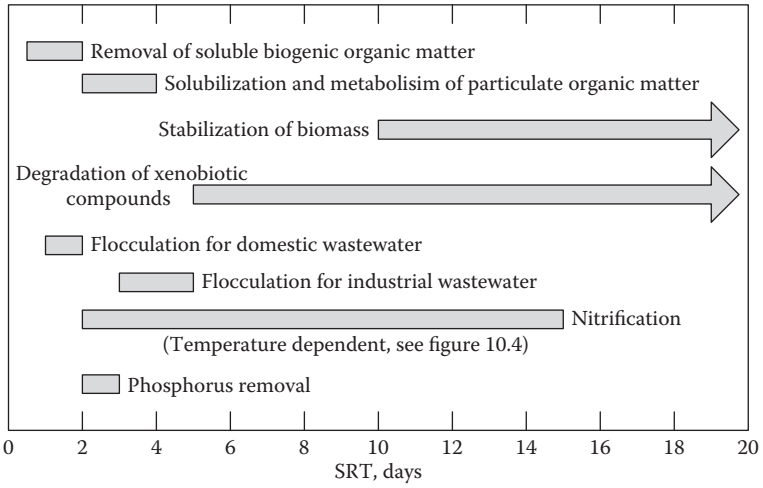

FIGURE 10.3 Typical SRT ranges for various biochemical conversions in aerobic/anoxic bioreactor systems at 20°C.

stabilization is obtained as the SRT is increased beyond 10 days. Relatively long SRT values are often required to biodegrade xenobiotic compounds. As a general rule, the SRT should be at least 5 days to biodegrade some of these materials, but often an SRT in excess of 10 days is necessary for complete biodegradation.

The SRT must also be sufficiently long to allow flocculent growth of heterotrophic bacteria because biomass separation and recycle in suspended growth systems requires such a condition. A wide range of values has been reported in the literature, but practical experience indicates that when domestic wastewater is being treated, flocculation can be obtained at SRT values as short as one day. For industrial wastewaters, however, longer values may be required, typically ranging from three to five days. This difference may be due to differences in the nature of the substrates in the two types of wastewater, or to the higher concentration of bacteria typically present in domestic wastewater. A complicating factor is the growth of filamentous bacteria, which may be exacerbated by the use of low SRT values. Because of their importance, the topics of bioflocculation and filamentous organism growth were discussed in Sections 2.3.1 and 2.3.2, respectively. They will be discussed further in Section 11.2.1. For the present time, however, it is only necessary to recognize that a minimum SRT value exists below which bioflocculation will not occur and that the minimum value is a function of wastewater type and other factors.

As discussed in Section 3.2.10, the autotrophic nitrifying bacteria have lower $\hat{\mu}$ values than most heterotrophic bacteria, and thus they require a longer SRT to survive in an aerobic bioreactor, as reflected in Figure 10.3. In addition, their maximum specific growth rate coefficient is more sensitive to changes in temperature than that of heterotrophs. Consequently, a broad range of SRTs has been shown for nitrification. Denitrification is accomplished by heterotrophic bacteria, which have relatively high $\hat{\mu}$ values, allowing them to grow at relatively low SRTs if nitrate-N is present in the influent wastewater. However, since most wastewaters contain ammonia-N rather than nitrate-N, a relatively long SRT is required for nitrification/denitrification systems to allow production of nitrate-N by the relatively slow growing nitrifiers.

The range of SRT values required for growth of the PAOs utilized in biological phosphorus removal and anaerobic selector systems is also presented in Figure 10.3. The $\hat{\mu}$ values for the PAOs are lower than those of ordinary heterotrophic bacteria, which are not capable of accumulating phosphorus. Consequently, the lower limit on the SRT for phosphorus removal is generally higher than that for soluble substrate removal.[16] On the other hand, the lower limit on the SRT for phosphorus removal is similar to the lower limit for nitrification, suggesting that it may be difficult to operate a bioreactor for phosphorus removal without experiencing the problems caused by the presence of nitrate, as discussed in Section 7.7.2.

Additional insight into the growth characteristics of nitrifying bacteria and PAOs can be gained by examining Figure 10.4, where the effects of temperature on the minimum SRT for each type of bacteria are shown for aerobic conditions. The values in the figure are computed values using typical kinetic parameter values and temperature correction coefficients.[11,15] It should be emphasized that the curves are for the minimum SRTs at which washout will occur, not operating SRTs. No safety factors are included. An examination of Figure 10.4 reveals that at temperatures below 20°C the minimum SRT for PAOs is sufficiently smaller than the minimum SRT for nitrifiers to allow operation of a phosphorus removal system without nitrification occurring, but that at temperatures above 24°C this would be extremely difficult to do.

Two final points should be made about Figure 10.3. First, the design SRT must reflect the limiting event that is required in the system. For example, even though an SRT as low as 0.5 days would be sufficient to remove soluble organic matter, it could not be used as the design SRT for a system using sedimentation to remove the biomass because it would not be sufficient to allow bioflocculation to occur. Rather, the SRT would have to be above one or two days to allow flocculation during treatment of domestic wastewater and above three to five days for an industrial wastewater. Second, any event that can occur at the chosen SRT will, provided the environmental conditions are adequate. For example, if an SRT of 15 days were used to achieve stabilization of biomass, provision should be made for

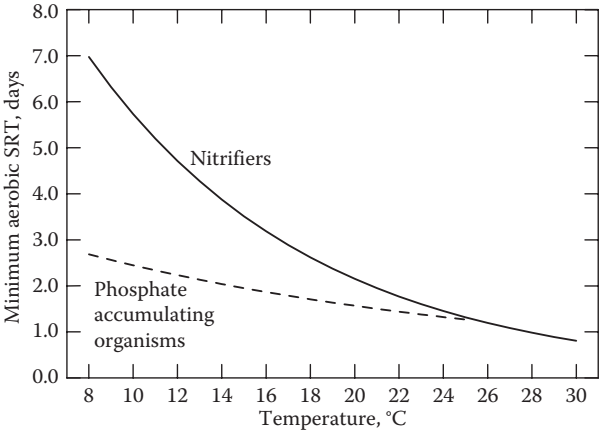

FIGURE 10.4 Effect of temperature on the minimum aerobic SRT required to grow nitrifiers and phosphate accumulating organisms (PAOs). The nitrifier curve was adapted from Sedlak, R. I., ed., *Phosphorus and Nitrogen Removal from Municipal Wastewater: Principles and Practice*, 2nd ed., Lewis Publishers, Ann Arbor, MI, 1991, and the PAO curve was developed from data presented by Mamais, D. and Jenkins, D., The effects of MCRT and temperature on enhanced biological phosphorus removal. *Water Science and Technology*, 26 (5/6): 955–65, 1992.

nitrification since the SRT is long enough for it to occur. Figure 10.3 can be very helpful as a reminder of what is likely to occur at various SRTs, thereby helping a designer to consider all possible events.

10.3.2.2 Anaerobic Systems

A similar approach can be used for anaerobic systems. Figure 2.4 depicts the biochemical conversions that occur in such systems and Figure 10.5 indicates the SRT ranges over which they occur at 35°C. Longer SRT values will generally be required for lower temperatures.

Hydrolysis of particulate carbohydrates and proteins to produce simple sugars and amino acids is a relatively rapid reaction, which is essentially complete for SRT values in excess of about three days. In contrast, the hydrolysis of lipids to form long chain fatty acids and other soluble reaction products is a much slower reaction that does not generally occur for SRT values less than about six days. A significant difference also exists for the various acidogenic reactions that convert the hydrolysis products into acetic acid and hydrogen. Fermentation of amino acids and simple sugars occurs very rapidly and, generally, will not be rate limiting. In contrast, the anaerobic oxidation of fatty acids to acetic acid and hydrogen is much slower. The oxidation of propionic acid is particularly slow in comparison to the other anaerobic oxidations.

Various methanogenic reactions are also possible, depending on the SRT. Hydrogen oxidizing methanogens can grow quite rapidly; a complete population of such organisms will generally develop for SRT values in excess of about 1.5 days. In contrast, the aceticlastic methanogens grow much more slowly, and there are significant differences in the maximum specific growth rate coefficients of the two major types. *Methanosarcina* grows relatively rapidly and a complete population will be available at SRTs in excess of about 5 days. In contrast, *Methanosaeta* grows relatively slowly and will not generally be present unless the SRT is in excess of about 12 days.

Analysis of the relative growth characteristics of the various anaerobic microorganisms results in some important observations. They are summarized in the three paragraphs that follow.

1. A relatively low SRT must be maintained if an anaerobic process is to achieve acidogenesis without significant methanogenesis. Hydrolysis of carbohydrates and proteins will generally be complete at SRTs of about three days, and the simple sugars and amino acids produced will be converted into acetic acid, other volatile acids, and hydrogen. Hydrogen

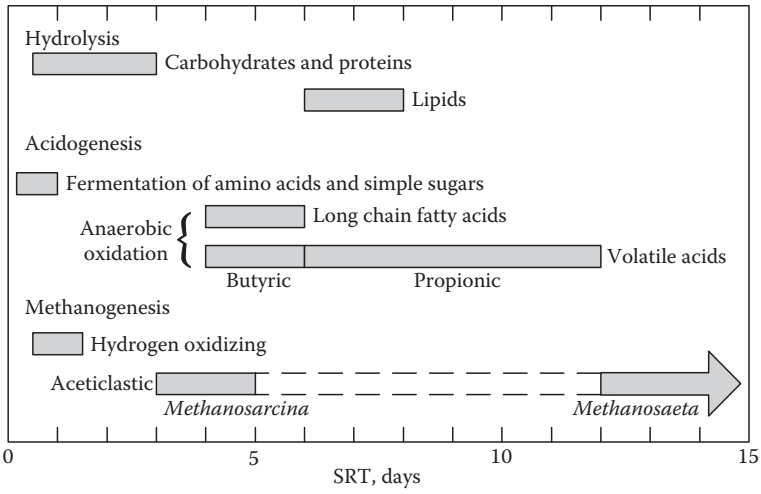

FIGURE 10.5 Typical SRT ranges for various biochemical conversions in anaerobic bioreactor systems at 35°C.

oxidizing methanogens are capable of growing at these SRTs, so much of the hydrogen produced will be converted into methane. The quantity will be small, however, since only limited amounts of hydrogen are produced through fermentation reactions. The volatile acids will accumulate under these conditions because longer SRTs are required for growth of the acetogenic bacteria that convert them to acetic acid and hydrogen by anaerobic oxidation. An SRT of less than three days must be maintained to prevent the growth of *Methanosarcina*, which would convert acetic acid to methane and carbon dioxide. Such a short SRT will not be sufficient for hydrolysis of lipids, so they will remain unreacted.

2. Anaerobic treatment of a wastewater containing carbohydrate and protein with production of methane can be accomplished at SRT values of about eight days. At this SRT, the carbohydrates and proteins will be hydrolyzed; the hydrolysis products will be converted by fermentation and anaerobic oxidation into acetic acid, carbon dioxide, and hydrogen; and the acetic acid and hydrogen will be utilized for methane production. In fact, significant methane formation will occur at SRT values as low as five to six days, but significant quantities of propionic acid will accumulate because this SRT is too short to allow the growth of bacteria that anaerobically oxidize propionic acid to acetic acid and hydrogen.
3. Solids retention time values in excess of eight days will be required to stabilize wastewaters containing significant quantities of lipids, such as primary sludges from domestic treatment systems. Generally, a minimum SRT of about 10 days is specified to ensure complete and reliable degradation of lipids in anaerobic bioreactors.

The performance of anaerobic bioreactors is affected by many factors in addition to the SRT, such as temperature, pH, and the presence of toxic materials. In addition, although anaerobic systems have demonstrated the ability to degrade xenobiotic materials, relatively long SRTs are often required. Nevertheless, the information presented in Figure 10.5 illustrates the relative effects of SRT on the growth of the various types of microorganisms found in anaerobic bioreactors and the resulting impact on the types of biochemical conversions that will occur.

10.3.3 Items from Process Stoichiometry

As seen from the guiding principles summarized in Section 10.1, once the biochemical environment and the SRT have been selected, a number of important items follow directly from the

stoichiometry of biomass growth and substrate utilization. First, for a given wastewater flow rate and concentration, the mass of biomass in the system is fixed. Consequently, the mass of MLSS is also fixed. Since only the mass of MLSS is specified, the designer may freely choose either the MLSS concentration or the bioreactor volume, thereby fixing the other. The values chosen will depend on the nature of the biochemical operation, the bioreactor configuration, and constraints that consider the interactions between the various unit operations in the system. Consideration of those factors is an important component of system design and will be discussed in detail in the chapters to follow.

Another item that is determined by the process stoichiometry is the mass rate at which solids must be wasted from the bioreactor system to maintain the desired SRT. It is very significant because it will be used to size the solids processing system for the facility.

The process stoichiometry also determines the quantity of electron acceptor required. This follows directly from the mass of biodegradable COD entering the bioreactor system and the mass of solids wasted from the system, as expressed by the COD balance that has been stressed throughout this book. If the biochemical environment is aerobic, the electron acceptor will be oxygen, and the calculated oxygen requirement will be central to the sizing of the oxygen transfer system. On the other hand, if the system is aerobic/anoxic, the computation will fix the sum of the oxygen and nitrate requirements, but additional information will be needed to determine the relative amounts of the two electron acceptors required. Finally, if the system is anaerobic, the electron acceptor requirement can be translated directly into the methane production rate, which can be used to estimate the quantity of energy available for use within the facility.

Finally, stoichiometry can be used to estimate the nutrient requirements for the bioreactor, as discussed in Sections 3.8.2 and 5.1.6. While it will seldom be necessary to add nutrients to bioreactors treating domestic wastewaters, many industrial wastewaters lack sufficient quantities of one or more macro- or micronutrients, and thus they must be added for successful operation of a bioreactor. Appropriate planning for such additions is an important component of design.

The quantitative information listed above can be developed with various degrees of precision, depending on the nature of the information available and the use to which it will be put. In Section 10.4 we will examine the various levels of design and evaluation and their appropriate use.

10.3.4 Interactions among Decisions

We have seen above that the designer can freely choose either the MLSS concentration or the bioreactor volume once the mass of MLSS in the system has been determined. While that choice is a "free" one in the sense that no equation specifies it, it cannot be made in isolation. Rather, it must be made with consideration of its impact on other unit operations in the system. Two such interactions are particularly important: those with the downstream biomass separation device, typically a gravity clarifier but also possibly a membrane filter, and those with the mixing and/or aeration system.

For a given application, a wide range of MLSS concentrations can be used. Thus, it would appear that, from an economic standpoint, the most cost-effective design would use the highest possible MLSS concentration, resulting in the smallest possible bioreactor. However, this overlooks the fact that higher MLSS concentrations impose an increased load on the downstream biomass separation device. The result is that larger clarifiers or membrane filters are required to separate the applied solids and concentrate them for return to the upstream bioreactor. Clarifiers must be sized to avoid thickening failure, which is a function of the settleability of the solids, the applied MLSS concentration, and the recycle flow rate from the clarifier to the bioreactor. Likewise, the amount of fluid that may be applied per unit surface area for a membrane filter (referred to as the flux) must be reduced when the applied MLSS concentration exceeds about 10,000 mg/L. Thus, a trade-off exists between the use of a small bioreactor with a large secondary clarifier or membrane filter, and vice versa. When clarifiers are used for biomass separation, the trade-off between bioreactor and

clarifier size can be analyzed if the settleability of the solids is known by formulating a variety of feasible options and determining the most cost-effective one.[6] During preliminary design, however, the settling characteristics of the solids are seldom known. Fortunately, correlations have been developed between various solids settleability indices and the settling characteristics of activated sludge,[3,10] and these can be used to quantify expected settling characteristics. These relationships are the basis for secondary clarifier operating diagrams that can be used to select appropriate clarifier areas and recycle flow rates for preliminary design purposes.[3] A complete explanation of those diagrams and their use requires knowledge of solids flux theory,[10] which is beyond the scope of this chapter. Nevertheless, the reader should be aware of the existence of the diagrams and the fact that they can be used for both design and evaluation of final clarifiers associated with bioreactors. Likewise, experience is evolving about the effects of MLSS concentration on allowable membrane flux rates.[2,14]

If the proposed bioreactor is aerobic, provisions must be made to transfer the needed oxygen to the liquid phase for use by the biomass. In addition, regardless of the nature of the biochemical environment, provision must be made for keeping the MLSS in suspension without subjecting it to so much shear that it will not flocculate and settle properly. Generally, for economic reasons, in aerobic bioreactors the same equipment is used to transfer oxygen and to keep the MLSS in suspension. This results in certain constraints that must be considered during design and evaluation. Figure 10.6 illustrates those constraints.

The volumetric power input to a bioreactor is the power applied per unit volume for mixing and/or oxygen transfer, regardless of whether it comes from a mechanical mixer or from the movement of air discharged into the bioreactor. A minimum volumetric power input is required to provide the turbulence needed to keep solids in suspension. Its value depends on the type of biochemical operation being used, and appropriate values will be provided in subsequent chapters. Nevertheless, it is

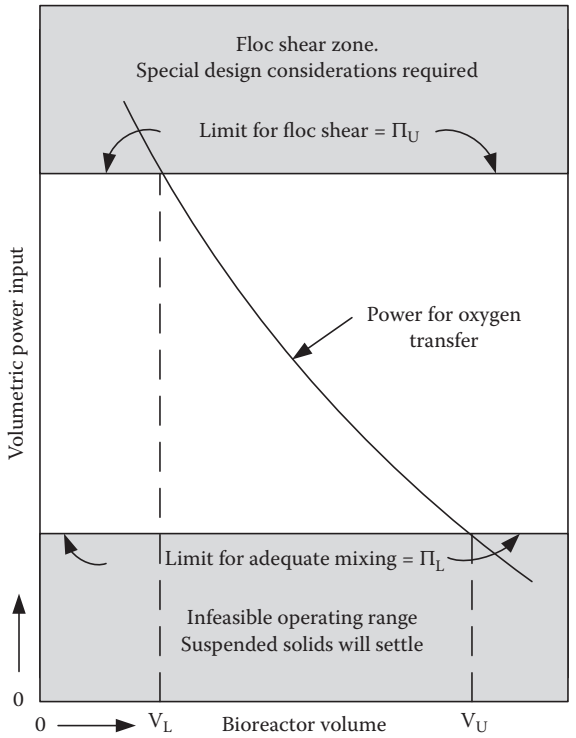

FIGURE 10.6 Effect of bioreactor volume on the volumetric power input required for oxygen transfer and its relationship to solids suspension and floc shear.

shown schematically in the figure as the lower horizontal line labeled "limit for adequate mixing." Conversely, a maximum volumetric power input exists above which excessive floc shear will occur, making it very difficult to achieve adequate clarification of the treated wastewater. As with the mixing constraint, its value depends on the type of biochemical operation under consideration, but for illustrative purposes it is shown in the figure as the upper horizontal line labeled "limit for floc shear."

The amount of oxygen required by biomass treating a given wastewater is fixed once the SRT of the system is fixed. For a given type of aeration device, transfer of that quantity of oxygen will require a total power input that is essentially independent of the volume of the bioreactor. Consequently, the larger the bioreactor, the smaller the volumetric power input for oxygen transfer, as illustrated by the curve in Figure 10.6. The intersections of that curve with the two horizontal limit lines define the upper and lower feasible bioreactor volumes, V_U and V_L, respectively. If the bioreactor volume were greater than V_U, the total power input to the bioreactor, as determined by the product of the volumetric power input limit for mixing times the bioreactor volume, would be greater than the power input needed for oxygen transfer. This is inefficient and wastes power. On the other hand, if the bioreactor volume were less than V_L, the biomass would be subjected to excessive shear forces, which will disrupt flocculation and clarification. Thus, when decisions are being made about the MLSS concentration and associated bioreactor volume, consideration should be given to the impact of the bioreactor volume on the volumetric power input. The quantitative aspects of this consideration will be presented in the chapters to follow.

10.4 LEVELS OF DESIGN AND EVALUATION

The iterative nature of process design and evaluation was discussed in Section 10.2. One consequence of it is that the level of detail required will change as the design progresses. We have described elsewhere how a series of process models can be applied at various stages in a project to provide a consistent and increasingly precise estimate of biological process requirements.[5] In this section we will present the essential elements required for each level of design.

10.4.1 Preliminary Design and Evaluation Based on Guiding Principles

In many instances, aerobic/anoxic and anaerobic systems can be sized on an approximate basis using the guiding principles set forth in Section 10.1. The resulting preliminary process design and evaluation provides an initial assessment of the capacity and capability of an existing biological treatment facility and of the changes required to expand it. It can also provide initial information about the nature of a proposed new facility. These preliminary estimates allow development of a preliminary scope and cost estimate for the project of concern that can subsequently be refined as further information is gathered.

After a decision has been made about the biochemical environment, information such as that provided in Figures 10.3 and 10.5 can be used to select the SRT. Both of these decisions require a degree of experience with the type of wastewater under consideration, but the factors discussed in Sections 10.3.1 and 10.3.2 can provide guidance during the decision-making process. In addition, the information in Figure 10.4 can be used in conjunction with Figure 10.3 to provide additional insight into the circumstances under which nitrification is likely to occur. Such an application is illustrated in the following example.

Example 10.4.1.1

Consider the design of an activated sludge system to treat a readily biodegradable industrial wastewater. The discharge standards require removal of biodegradable organic matter, but no limits have been set on effluent ammonia-N, total nitrogen, or phosphorus.

a. What SRT would be selected to provide reliable treatment and a high quality effluent while also minimizing the size of the bioreactor?
   From Figure 10.3, it can be seen that a SRT in excess of two days is required to provide essentially complete removal of soluble, biodegradable organic matter. However, because the facility is to treat an industrial wastewater, a SRT of three to five days should be used to achieve good flocculation. Consequently, select a SRT of five days as a conservative measure for preliminary process sizing.
b. Will the system nitrify at the selected SRT?
   Figure 10.3 suggests that nitrification will occur at a SRT of five days, provided that nothing in the wastewater inhibits the nitrifying bacteria. However, to gain more insight, the effects of temperature should be considered by using Figure 10.4. Nitrification will occur as long as the bioreactor SRT is above the minimum SRT for the nitrifying bacteria. Entering Figure 10.4 with the design SRT of five days reveals that nitrification will occur down to a temperature of about 12°C. At that temperature, however, the nitrifying bacteria will be on the verge of washing out. As a result, nitrification will not be stable. However, the warmer the temperature, the greater the design SRT will be relative the minimum SRT and the more stable nitrification will be. This information can be combined with information about the annual variations in temperature to make a decision about when nitrification is likely to occur.
c. Could enhanced biological phosphorus removal capabilities be incorporated into the design of this facility without increasing the design SRT, if the need were to arise in the future?
   Figure 10.3 indicates that the growth of PAOs requires an SRT of at least two to three days. Since the design SRT of five days exceeds this value, growth of phosphorus removing bacteria is possible. Furthermore, an examination of Figure 10.4 reveals that it should be possible to sustain biological phosphorus removal down to quite low temperatures.

After selection of the SRT, estimates must be made of the various items related to process stoichiometry, as discussed in Section 10.3.3. This, too, can be done on the basis of the guiding principles. An important principle articulated in Section 10.1 is that for a given SRT, Θ_c, the mass of biomass in a bioreactor, $X_{total,T} \cdot V$, is fixed for a given treatment situation (i.e., flow, F, and concentration of wastewater, S_{SO}). This follows directly from combining Equations 5.35 and 5.38 and invoking the definition of the HRT:

$$X_{total,T} \cdot V = \Theta_c Y_{Hobs,T} F(S_{SO} - S_S), \tag{10.1}$$

where $Y_{Hobs,T}$ is the observed yield and S_S is the effluent substrate concentration. The concept embodied in Equation 10.1 can be carried another step, thereby giving a simple equation with general utility for preliminary process design. For most domestic wastewaters, both inert and biodegradable organic matter will be present in the influent, requiring us to be concerned about the MLSS concentration rather than just the biomass concentration. We saw in Section 5.2.2 that the simple model of Chapter 5 could be extended to include inert particulate organic matter, X_{IO}, and in Section 9.5.2 that it could even incorporate slowly biodegradable substrate, X_{SO}, provided that the SRT was sufficiently long. Because of that, and recognizing that for the SRTs used in practice $S_S << S_{SO}$, we can write an approximate equation for the mass of MLSS that follows directly from Equation 10.1 and the fundamental principle it embodies:

$$X_{M,T} \cdot V = \Theta_c Y_{n,T} F(S_{SO} + X_{SO}). \tag{10.2}$$

Examination of Equation 10.2 reveals that it differs from Equation 10.1 in three important ways. First, it expresses the MLSS concentration, $X_{M,T}$, and includes both readily and slowly biodegradable substrate, as discussed above. The most important difference is in the nature of the yield coefficient,

however. The coefficient employed, $Y_{n,T}$, is called the net process yield, which differs from the observed yield. In Equation 5.38 we saw that the observed yield accounts for the effect of biomass decay on the net amount of heterotrophic biomass formed. As such it is a fundamental parameter that decreases in value as the SRT is increased. The net process yield accounts for both that effect and the impact of inert solids and slowly biodegradable substrate on the MLSS in the bioreactor. As such, it is an empirical parameter that depends on the nature of the wastewater under treatment. However, if both X_{IO} and X_{SO} are zero, $Y_{n,T}$ becomes equivalent to $Y_{Hobs,T}$.

In spite of its empirical nature, for many wastewaters, typical values of net process yield are known, or can be calculated using data from operating systems. An example of one relationship between net process yield and SRT is shown in Figure 10.7.[18] Several items should be noted. First,

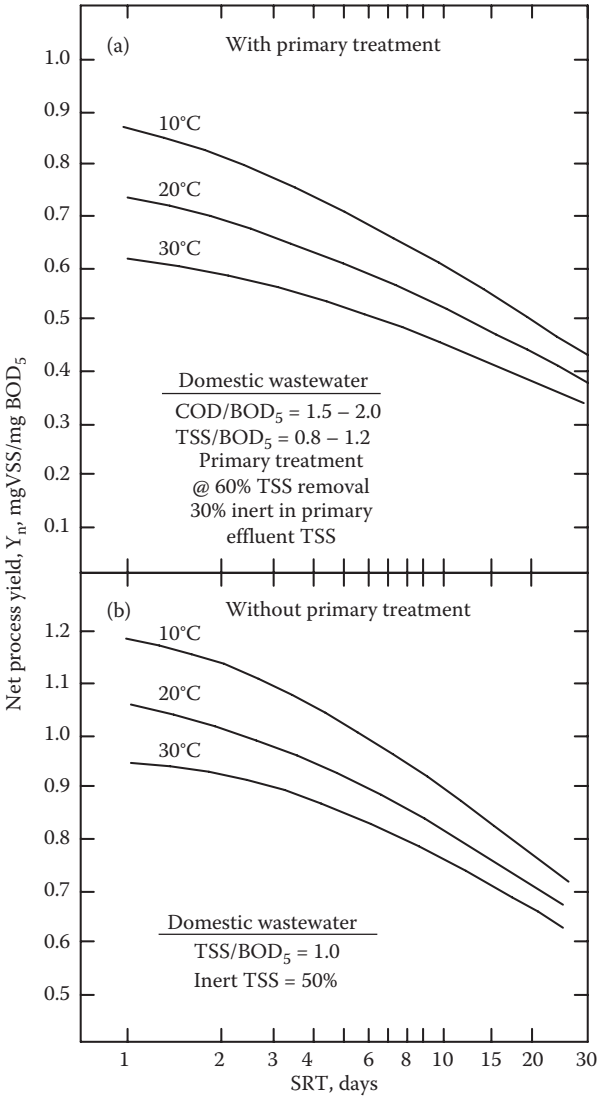

FIGURE 10.7 Net yield as a function of SRT and temperature for domestic wastewater (a) with primary treatment and (b) without primary treatment. (From *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, 4th ed., Copyright © Water Environment Federation, Alexandria, Virginia, 1998. Reprinted with permission from Water Environment Federation.)

the net process yield is affected by SRT in much the same way as the observed yield, as would be expected (compare Figure 10.7 to Figure 5.4). Second, the net process yield values are larger when the wastewater has not been subjected to primary treatment (i.e., settling) because the values of X_{IO} and X_{SO} are larger. Finally, in this figure the net process yield has units of mg VSS/mg five-day biochemical oxygen demand (BOD_5), making its symbol $Y_{n,V}$. Actually, any unit system could be used to report net process yield, as long as it is consistent with the methods used to measure the MLSS concentration and the substrate. We have elected to use total suspended solids (TSS) units as the measure of MLSS in this book. Activated sludges typically have a volatile solids content of approximately 75%, so the MLSS values determined with net process yield values from Figure 10.7 can be converted to a TSS basis by dividing by 0.75.

The utility of Equation 10.2 is that it can be used to estimate the mass of MLSS that will be present in the bioreactor, regardless of its configuration. The distribution of those solids will depend on the system configuration, as illustrated in Chapter 7 through simulation.

The net yield can also be used to calculate other information needed for a preliminary process design, such as the excess solids wastage rate, which is needed for design of the solids handling and disposal system. In Chapter 5 we derived Equation 5.41 that expressed the excess biomass wastage rate in terms of $Y_{Hobs,T}$. By analogy to the development of Equation 10.2, an expression can be developed from which the MLSS mass wastage rate, $W_{M,T}$, can be calculated during preliminary design:

$$W_{M,T} = FY_{n,T}(S_{SO} + X_{SO}). \tag{10.3}$$

This equation has particular significance because it has another important use. If an existing treatment facility is to be upgraded, then information will be available in the plant records on the amount of solids wasted daily, as well as the quantity and strength of the wastewater. Such information allows the net process yield to be calculated from the historical records with Equation 10.3. It can then be used in Equation 10.2 to estimate $X_{M,T} \cdot V$ for use in the design.

As discussed in Section 10.3.3, another factor that must be considered during design is the electron acceptor requirement. The basic COD balance states that the amount of electron acceptor required is equal to the biodegradable COD entering a bioreactor minus the COD of the solids wasted from the bioreactor. This is what led to Equation 5.45, which gives the oxygen requirement in a simple CSTR receiving a soluble substrate. By analogy to the development of the previous preliminary design equations, an equation can be developed that relates the oxygen requirement, RO, to the flow and waste load in a manner similar to Equation 10.3:

$$RO = FY_{O_2}(S_{SO} + X_{SO}). \tag{10.4}$$

In this case, a new empirical coefficient, Y_{O_2}, the process oxygen stoichiometric coefficient, has been defined, the units of which must be consistent with the units on S_{SO} and X_{SO}. Since BOD_5 is a commonly used measure of biodegradable substrate concentration in practice, Y_{O_2} is commonly expressed as mg O_2/mg BOD_5. For domestic wastewater, the value of Y_{O_2} is known as a function of the SRT, as illustrated in Figure 10.8.[17] Two things are of note about this figure. First, the values of Y_{O2} exceed 1.0. This follows directly from the fact that BOD_5 is not a measure of all of the electrons available in a substrate, as discussed in Section 9.6. Second, Y_{O_2} increases as the SRT is increased in a manner consistent with Figure 5.6, illustrating that the empirical relationship depicted by Equation 10.4 conforms to the fundamental principles developed earlier. For an existing facility, Equation 10.4 can be rearranged to calculate the process oxygen stoichiometric coefficient associated with a given waste load. Procedures for measuring the process oxygen requirement of such a facility are described elsewhere.[4,17]

The concept embodied in the COD mass balance and reflected in Equation 5.45 can be generalized to any electron acceptor by expanding the substrate term to include slowly biodegradable

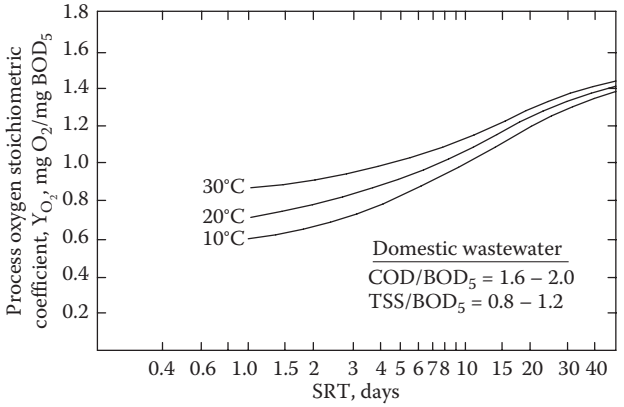

FIGURE 10.8 Process oxygen stoichiometric coefficient as a function of SRT and temperature for domestic wastewater. (From U.S. Environmental Protection Agency, *Design Manual—Fine Pore Aeration Systems*, EPA/625/1-89/023, Cincinnati, OH, 1989.)

substrate, as was done with Equation 10.4, and incorporating an appropriate conversion coefficient, $i_{O/EA}$, reflecting the COD mass equivalent of the particular acceptor as listed in Table 3.1:

$$\text{REA} = \frac{F(S_{SO} + X_{SO})(1 - Y_{Hobs,T} \cdot i_{O/XB,T})}{-i_{O/EA}}, \tag{10.5}$$

where REA is the mass per time of electron acceptor required and the substrate concentrations are in COD units. If oxygen is the electron acceptor, $i_{O/EA}$ has a value of −1.0 g COD/g O_2, whereas if nitrate-N is the acceptor, $i_{O/EA}$ has a value of −2.86 g COD/g N. If carbon dioxide serves as the electron acceptor, leading to methane formation, the value of $i_{O/EA}$ is −5.33 g COD/g C. In that case, however, it is more common to be interested in the amount of methane produced, rather than in the amount of carbon reduced. That quantity can be calculated by recognizing that all electrons removed end up in methane and that at standard temperature and pressure, the oxidation of 1.0 kg COD results in the formation of 0.35 m^3 of methane, as discussed in Section 2.3.3:

$$CH_4 \text{ production} = [F(S_{SO} + X_{SO})(1 - Y_{Hobs,T} \cdot i_{O/XB,T})]\left(0.35 \frac{m^3\ CH_4}{kg\ COD}\right). \tag{10.6}$$

As in Equation 10.5, the substrate concentrations are in COD units for Equation 10.6.

Finally, for wastewaters that have a significant industrial component, the requirement for nutrients should be checked to ensure that adequate amounts are provided. A conservative estimate of the nutrient requirement can be obtained by assuming that all waste solids, as calculated from Equation 10.3, are biomass and that the nutrients contained in those solids must be provided in the influent or by supplementation. Thus, the mass per day of nutrients required may be estimated by multiplying the solids wastage rate by the appropriate factor from Table 10.3, which was derived from Table 3.3. The supplementation requirement will be the difference between the amount required and the amount available in the influent.

All of the equations in this section are based on the guiding principles articulated in Section 10.1. Although they are approximate and incorporate several assumptions, they are sufficiently accurate for use during preliminary design when decisions are being made about the feasibility of various process alternatives. Thus, they have considerable utility in practice. Their use is illustrated in the following example.

TABLE 10.3
Approximate Nutrient Requirements

	Approximate Requirement*	
Nutrient	g/kg of VSS Wasted	g/kg of TSS Wasted
Nitrogen	125	104
Phosphorus	25	21
Potassium	14	12
Calcium	14	12
Magnesium	10	8
Sulfur	8.5	7
Sodium	4.3	3.6
Chloride	4.3	3.6
Iron	2.8	2.4
Zinc	0.3	0.2
Magnesium	0.1	0.2

*Based on Table 3.3

Example 10.4.1.2

The preliminary design for a domestic wastewater treatment plant is being developed. The wastewater flow rate is 10,000 m^3/day, and the BOD_5 concentration is 200 mg/L (= 200 g/m^3). The plant discharge standards are 30 mg/L BOD_5 and 30 mg/L TSS. Experience suggests that stable and reliable compliance with this discharge standard can be achieved using an SRT of five days. No primary clarifiers are to be provided. For the purpose of this assessment, the wastewater temperature can be assumed to be 20°C. Make a preliminary assessment of the size of bioreactor required if the MLSS concentration is maintained at 2500 mg/L as TSS. Assume that 75% of the MLSS is volatile. Also estimate the kg/day of solids that must be disposed of and the kg/day of oxygen that must be supplied. Because the wastewater is totally domestic, it can be assumed to contain adequate nutrients.

a. What would be an appropriate net process yield value for this application?
   The value of $Y_{n,V}$ can be estimated from Figure 10.7. Since primary treatment will not be used, part b of the figure is the appropriate part to use. For an SRT of five days, $Y_{n,V}$ has a value of 0.92 mg VSS/mg BOD_5 at 20°C. Assuming that the MLSS will be 75% volatile, the value of $Y_{n,T}$ is 1.2 mg TSS/mg BOD_5.
b. What is the $X_{M,T} \cdot V$ product for this application? From Equation 10.2:

$$X_{M,T} \cdot V = (5)(1.2)(10{,}000)(200) = 12{,}000{,}000 \text{ g TSS}$$
$$= 12{,}000 \text{ kg TSS}.$$

c. If the allowable MLSS concentration is 2500 mg/L (= 2500 g/m^3), what size would the bioreactor be?
   Since $X_{M,T} \cdot V$ = 12,000,000 g TSS,

$$V = \frac{12{,}000{,}000}{2500} = 4800 \text{ m}^3.$$

d. What would the solids wastage rate be?
The solids wastage rate can be estimated with Equation 10.3:

$$W_{M,T} = (10{,}000)(1.2)(200) = 2{,}400{,}000 \text{ g TSS/day}$$
$$= 2400 \text{ kg TSS/day}.$$

e. What would the process oxygen requirement be?
From Figure 10.8, at a five-day SRT and a temperature of 20°C, Y_{O_2} has a value of 0.90 mg O_2/mg BOD_5. Using this, the oxygen requirement can be estimated from Equation 10.4:

$$RO = (10{,}000)(0.9)(200) = 1{,}800{,}000 \text{ g } O_2\text{/day}$$
$$= 1800 \text{ kg } O_2\text{/day}.$$

The information provided in the example above is adequate for a ballpark estimate of the costs associated with a new treatment facility. However, it would not be adequate for more detailed estimates because such estimates require consideration of temperature variations, peak loadings, and so on. While information like that in Figures 10.7 and 10.8 could be used when considering those effects, the analysis becomes much more detailed and is beyond the scope of this chapter. Such considerations will be covered in the chapters that follow.

In summary, even when little other information is available, it is often possible to develop a preliminary process design and evaluation for a wastewater treatment facility using the guiding principles developed in Parts I and II of this book. The general procedure is as follows:

1. Select the design SRT based on the treatment objectives and anticipated process operating conditions. An initially conservative approach is recommended to ensure that the bioreactor is adequately sized.
2. Estimate the net process yield ($Y_{n,T}$) and the process oxygen stoichiometric coefficient (Y_{O_2}) based on the best information available. The basis may be actual operating data from an existing facility, values from the literature, such as those in Figures 10.7 and 10.8 or an estimate based on knowledge of and experience with similar wastewaters.
3. Calculate the $X_{M,T}$·V product using Equation 10.2.
4. Calculate the solids wastage rate using Equation 10.3 and the process oxygen requirement using Equation 10.4.
5. Use the value of the $X_{M,T} \cdot V$ product determined in item 3 to consider the interactions between the selected size of the bioreactor, the final clarifier, and the oxygen transfer/mixing system in order to arrive at a selection that meets all requirements.

For an existing facility, this will provide a preliminary estimate of existing capacity. For a new facility or a facility expansion, a preliminary estimate of capital and operating costs can be prepared for use in ongoing project evaluation and planning. The selection of a conservative SRT value results in conservative estimates of each. It is important to note that no estimates of process effluent quality have been developed yet. Rather, the fact that effluent quality is largely determined by the process SRT has been exploited to allow selection of an SRT that will result in the desired degree of conversion. The design to achieve a specified effluent quality requires more refined techniques.

10.4.2 Stoichiometric-Based Design and Evaluation

As noted above, preliminary design and evaluation based on the guiding principles does not allow precise estimates of process effluent quality. That requires values of the kinetic parameters for

biodegradation of the constituent of interest. For some situations, approximate or typical values of those parameters are known, allowing more precise estimates of process effluent quality to be made without the necessity for running treatability studies. In other cases, appropriate values of the kinetic parameters are unknown and they must be measured directly for the subject wastewater. Likewise, the mass of biomass required in the bioreactor, the solids wastage rate, and the process oxygen requirement are all functions of the characteristics of the wastewater to be treated and the microbial populations that develop in the treatment system. Consequently, more precise estimates of the bioreactor size, and so on require values for the kinetic and stoichiometric parameters that are specific to a particular wastewater. Again, for some situations, appropriate values of the wastewater constituent parameters can be selected based on experience, but in other instances the values of those parameters must be measured directly for the subject wastewater and the particular type of bioreactor under consideration.

Because the International Water Association (IWA) activated sludge models have been developed fairly recently, most of our experience in modeling bioreactors for design and evaluation has come from the simple model of Chapter 5. Although it was presented as a soluble substrate model, much experience exists with it when the organic matter is present in both a soluble and particulate state. In that case, influent concentrations are expressed as BOD_5 or as biodegradable COD without regard to their physical state, whereas effluent concentrations are commonly expressed as soluble material alone. Such a practice assumes that bioflocculation will entrap the bulk of the undegraded particulate organic matter, making soluble material the main contributor of organic matter in the effluent. Treatability studies for quantification of the parameters in the simple model of Chapter 5 were outlined in Sections 9.2 and 9.3, while procedures for translating between the various measurement techniques were discussed in Section 9.6. By using such procedures it is possible to use the Chapter 5 model to design systems involving organic substrate removal and nitrification. Because most of the computations with that model are based on the stoichiometry of microbial growth, it has been called a stoichiometric-based model.[5] It has severe restrictions for use with biological nutrient removal systems. Thus, one should strive to use the IWA models for them.

When a design or evaluation is sufficiently advanced to warrant the use of modeling, many factors are considered that were not considered during preliminary design or evaluation. Thus, the approach is more involved, although the same basic strategy is used. Consequently, a full discussion of this subject will be presented in the remaining chapters of Part III, with each chapter being devoted to a particular type of named biochemical operation. The material in this section will be limited to a few major points.

When modeling is being used, the choice of the SRT should consider the value required to meet a desired effluent quality. If the bioreactor is to be completely mixed, then a rearranged form of Equation 5.22 may be used to select the minimum allowable design SRT because lower SRTs would not produce the required effluent quality:

$$\Theta_c = \frac{K_S + S_S}{S_S(\hat{\mu}_H - b_H) - K_S b_H}, \tag{10.7}$$

where $\hat{\mu}_H$ and K_S are the Monod parameters and b_H is the traditional decay coefficient. This computation may be made on the basis of total soluble biodegradable organic matter, using either biodegradable COD or BOD_5 as the measure of S_S, or on the basis of specific organic chemicals, in which case S_S would represent them. No matter what the basis for S_S, however, the same basis must be used for $\hat{\mu}_H$ and K_S. Alternatively, if the system must meet an ammonia-N criterion, Equation 10.7 could be applied to the autotrophic bacteria by replacing the heterotrophic parameter values with autotrophic ones. In that case, however, appropriate safety factors may also be applied to the design. Once the minimum allowable design SRT has been computed with Equation 10.7, it must

be compared to the various constraints discussed in the context of Figure 10.3. In other words, even though modeling will be used in the design, the choice of SRT is still limited by the same constraints discussed previously, and they may control the choice. This subject will be discussed more in subsequent chapters.

Once the SRT has been selected, the design or evaluation of a system based on the stoichiometric model of Chapter 5 proceeds in exactly the same way as the preliminary design presented in Section 10.4.1. The differences are in the equations used and the additional factors that are considered. If the concentrations of the waste organic constituents are expressed in COD units and the MLSS concentration is expressed in TSS units, the equations in Chapter 5 may be used directly, with one minor modification: the influent concentration is the sum of the easily and slowly biodegradable constituents entering the bioreactor. In other words, S_{SO} is replaced by the sum of S_{SO} and X_{SO}. Consequently, the mass of solids in the bioreactor can be calculated with a modified form of Equation 5.58:

$$X_{M,T} \cdot V = \Theta_c F\left[X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c) Y_{H,T} (S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \Theta_c} \right]. \tag{10.8}$$

The solids wastage rate can be calculated from a modified form of Equation 5.61:

$$W_{M,T} = F\left[X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c) Y_{H,T} (S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \Theta_c} \right]. \tag{10.9}$$

The heterotrophic oxygen requirement can be calculated from a modified form of Equation 5.43:

$$RO_H = F(S_{SO} + X_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c) Y_{H,T} \cdot i_{O/XM,T}}{1 + b_H \cdot \Theta_c} \right]. \tag{10.10}$$

It should be recognized that because of the presence of inert organic solids and fixed solids in the MLSS and waste solids, their COD equivalent, $i_{O/XM,T}$, will be different from the COD equivalent of the biomass, $i_{O/XB,T}$, and will depend on the magnitude of the influent inert solids concentration, $X_{IO,T}$, as well as its COD equivalent. As a result, a measured value of $i_{O/XM,T}$ will normally be required to calculate the oxygen requirement from a COD balance across the bioreactor system. The nutrient requirement can still be calculated from the solids wastage rate as discussed above.

The equations above can be used regardless of the bioreactor configuration, as discussed in Section 10.1. However, by the time a design or evaluation has progressed to the point that a treatability study has been run to provide the parameter values for use in the model, it will be necessary to consider the distribution of MLSS and oxygen within the bioreactor system. This requires the use of heuristic rules based on experience. These will be discussed in subsequent chapters.

10.4.3 Simulation-Based Design and Evaluation

When a complex biochemical operation, such as a biological nutrient removal system, is being designed or evaluated, even a small degree of uncertainty can result in large consequences in terms of effluent quality or system cost. Thus, use of the simple stoichiometric-based model is usually not adequate, except as a starting point for more detailed studies. Because of the cost of treatability studies, particularly at the pilot-plant scale, the potential exists for significant cost savings through

the use of simulation. Computer codes implementing the various IWA activated sludge models, listed in Table 6.4, provide excellent means for doing this.

Starting with the default parameters provided in the IWA reports,[8,9] one can use the simple model or the guiding principles of Section 10.1 to choose a starting bioreactor system. The performance of that system can then be evaluated by simulation and systematic changes in system configuration and operating conditions can be investigated, just as was done in Chapter 7. Through this approach, the most feasible bioreactor systems and the most sensitive parameter values can be identified for experimental study in the lab using the procedures presented in Section 9.5. The results of the experimental study will provide new parameter estimates that can then be used in another round of simulations, leading to the one or two most feasible configurations, which can be investigated at pilot scale. In other words, the iterative process of design and evaluation illustrated to Figure 10.1 is applied to both the simulation and the testing phases of the project. The resulting system will have much less uncertainty associated with it.

10.4.4 Effluent Goals versus Discharge Requirements

Collection of the kinetic and wastewater characterization data described above allows relatively precise estimates of process effluent quality to be developed. However, the inherent variability in the performance of such processes must be recognized. Precise data allow accurate prediction of the mean performance of a treatment process. In contrast, facility discharge requirements are typically expressed in "not to exceed" terms so that variability in process performance must be considered when establishing a process design. Numerous procedures exist for translating discharge requirements into process design requirements. One which is frequently used is to establish an effluent goal that is more stringent than the specified discharge requirement and incorporates an allowance for the expected variations in treatment performance.

Several procedures are available to select effluent goals to allow reliable compliance with specified discharge requirements. In some instances effluent goals are selected based on experience with a particular wastewater and application. Other procedures are based on the statistical characteristics of a process effluent, and their use requires prior knowledge of those characteristics.[12] Monte Carlo simulation methods can be used when appropriate kinetic and waste load parameters and their statistical distributions are known.[1] Still other procedures are general in nature and can be applied to a wide range of applications. The procedure of Roper et al.[13] illustrates this latter approach and is presented here for illustrative purposes. Roper et al.[13] collected effluent quality data from a wide variety of full-scale wastewater treatment facilities and plotted extreme values versus the mean process performance. They found that the two were well correlated and that the relationship was generally linear. Figure 10.9 presents several of the relationships they developed and illustrates the association between the extreme values typically considered in discharge requirements and the mean performance that can be predicted using process kinetics and mean process loadings. Relationships such as those presented in Figure 10.9 can be used to select treatment goals to allow reliable compliance with facility discharge standards.

10.4.5 Optimization

Further steps in the refinement of the design and evaluation of a biological process fall under the general heading of optimization. A wide variety of approaches can be considered here, including pilot plant studies, further wastewater characterization and analysis, detailed analyses of the performance of existing facilities treating similar wastewaters or using similar processes, computer simulation of process performance to characterize the dynamic performance of a proposed process design, and detailed engineering studies to evaluate cost trade-offs for the subject process. Detailed descriptions of these steps are beyond the scope of this book. However, the reader should recognize

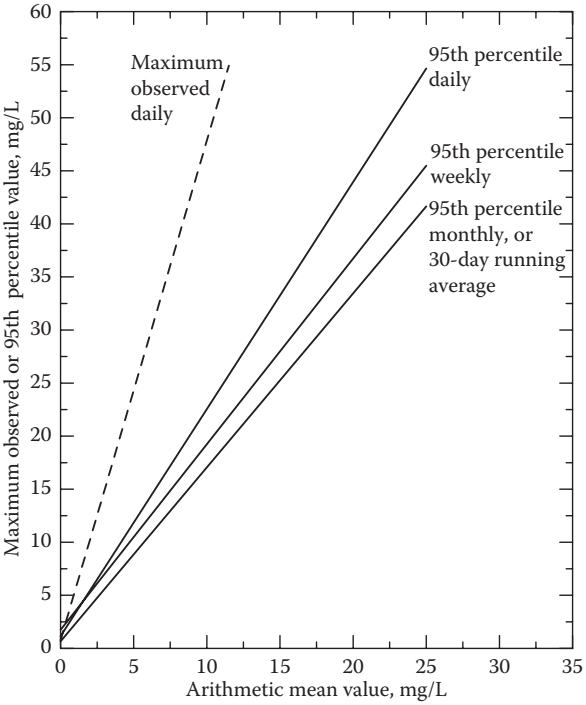

FIGURE 10.9 Relationship between arithmetic average and 95th percentile or maximum observed effluent BOD_5 and TSS concentrations for full-scale wastewater treatment plants. (Adapted from Roper, R. E., Jr., Dickey, R. O., Marman, S., Kim, S. W., and Yandt, R. W., Design effluent quality. *Journal of the Environmental Engineering Division, ASCE,* 105:309–21, 1979.)

that many tools are available that can be applied to further refine the design and evaluation of a biological wastewater treatment process.

10.5 KEY POINTS

1. Five guiding principles provide the basis for the design and evaluation of all suspended growth bioreactors:
   - The biochemical environment imposed upon a bioreactor determines the nature of the microbial community that develops and the character of the biological reactions that they perform.
   - The solids retention time (SRT) is the most important design and control parameter available to the engineer.
   - A chemical oxygen demand (COD) balance across a bioreactor provides valuable information about the amount of electron acceptor required and the amount of excess biomass produced.
   - The excess biomass production rate is essentially the same for all suspended growth systems with a given SRT and biochemical environment, regardless of the bioreactor configuration. As a consequence, the total mass of biomass and the total amount of electron acceptor required for removal of organic matter in those systems will be the same, regardless of the system configuration.
   - Only the mass of biomass in a bioreactor system is specified by the descriptive analytical equations, not the concentration. The concentration is only specified after the bioreactor volume has been specified.

2. The design process is iterative in nature and contains feedback loops. This is due to two factors: (1) improved problem definition and (2) additional data, both of which become available as the design and evaluation proceeds. Consequently, the design is refined as more information becomes available. The iterative nature of design demands the use of a variety of process design procedures, which must be consistent and based on the guiding principles. The benefits derived from using a more sophisticated design procedure must justify its additional costs.
3. The operational and performance characteristics of aerobic/anoxic and anaerobic bioreactors differ significantly. Aerobic/anoxic systems are generally used to treat wastewaters with biodegradable COD concentrations less than 1000 mg/L, while anaerobic systems are usually used to treat those with concentrations greater than 4000 mg/L. Either can be used for wastewaters with concentrations between those limits, depending on the specific application. Effluents from an anaerobic bioreactor must generally be polished in an aerobic system if a high quality is needed.
4. The SRT selected for a bioreactor must always exceed the minimum SRT associated with the microorganisms responsible for a particular desired biochemical transformation.
5. Biochemical conversions that are controlled by the SRT in aerobic/anoxic bioreactor systems include metabolism of biodegradable soluble and particulate organic matter, stabilization of removed organic matter, biodegradation of xenobiotic compounds, bioflocculation, growth of nitrifying bacteria, and growth of phosphate accumulating organisms (PAOs).
6. Biochemical conversions that are controlled by the SRT in anaerobic bioreactor systems include hydrolysis of carbohydrates, proteins, and lipids; fermentation of simple sugars and amino acids; anaerobic oxidation of fatty acids; and methanogenesis.
7. Several important characteristics of bioreactors can be determined from process stoichiometry once the SRT has been fixed, regardless of the bioreactor configuration. These are the mass of biomass in the system, the mass rate of solids wastage, the quantity of electron acceptor that must be supplied, and the amount of nutrients needed.
8. Because only the mass of biomass in a bioreactor is fixed by process stoichiometry, the designer may freely choose either the mixed liquor suspended solids (MLSS) concentration or the bioreactor volume, thereby fixing the other. That decision must be made, however, while considering its impact on the size of the downstream biomass separation unit (either a gravity clarifier or a membrane filter) and the volumetric power input required for oxygen transfer and/or mixing.
9. During preliminary design and evaluation of a bioreactor system, $Y_{n,T}$, the net process yield, may be used to calculate the mass of MLSS and the solids wastage rate for the system. A similar coefficient, Y_{O_2}, the process oxygen stoichiometric coefficient, can be used to estimate the oxygen requirement. They are based on experience at full-scale treatment facilities and incorporate the effects of SRT and wastewater characteristics.
10. More precise estimates of the bioreactor size, and so on require values for the kinetic and stoichiometric parameters that are specific to a particular wastewater. These may be obtained by treatability studies and used in stoichiometric models that are based on the simple model of Chapter 5. When a design or evaluation is sufficiently advanced to warrant the use of modeling, many factors may be considered that were not considered during preliminary design or evaluation. Thus, the approach is more involved, although the same basic strategy is used.
11. When a complex biochemical operation, such as a biological nutrient removal system, is being designed or evaluated, even a small degree of uncertainty can result in large consequences in terms of effluent quality or system cost. Thus, the potential exists for significant cost savings through the use of simulation with the International Water Association activated sludge models.

10.6 STUDY QUESTIONS

1. One of the guiding principles summarized in Section 10.1 is that the excess biomass production rate is essentially the same for all suspended growth systems with a given SRT and biochemical environment, regardless of the bioreactor configuration. Discuss the implications and significance of this fact to the design and evaluation of bioreactor systems.
2. Explain why the design and analysis of bioreactor systems is an iterative process and why the approach used must change as the process proceeds.
3. Define the terms "minimum SRT" and "safety factor" as applied to bioreactor design and evaluation, discuss their significance, and explain their use.
4. An aerobic/anoxic bioreactor is to be designed, and the first step is the selection of the design SRT. Discuss the differences in the SRT required to achieve removal of biodegradable organic matter (both readily and slowly biodegradable), flocculation, biological phosphorus removal, nitrification, and stabilization of the produced microorganisms. What factors determine the SRT range for each treatment objective?
5. An anaerobic bioreactor is to be designed, and the first step is the selection of the design SRT. Discuss the differences in the SRT required to achieve fermentation of influent particulate matter into volatile fatty acids, stabilization of carbohydrates and proteins, and stabilization of lipids. What factors determine the SRT range for each treatment objective?
6. An industrial wastewater treatment plant treats a flow of 5000 m^3/day with a COD concentration of 3500 mg/L. The process operates at a 25-day SRT and produces an effluent with a soluble COD concentration of 125 mg/L. Solids are wasted from the system at a rate of 7000 kg TSS/day. What is the net process yield ($Y_{n,T}$) for this facility, expressed in units of mg TSS/mg COD?
7. A bioreactor system treats 60,000 m^3/day of wastewater with a BOD_5 concentration of 200 mg/L and a TSS concentration of 210 mg/L, which are typical of unclarified domestic wastewater. The average temperature is 20°C. The discharge standards are 30 mg/L BOD_5 and 30 mg/L TSS on a maximum monthly basis (i.e., typical secondary treatment). Calculate the mass of MLSS in the bioreactor (in TSS units), the process solids wastage rate (in TSS units), and the process oxygen requirements for this application. Neglect process oxygen requirements for nitrification. Justify your choice of SRT.
8. The addition of primary treatment to the domestic wastewater treatment plant described in Study Question 7 is being evaluated. Pilot studies demonstrate that primary clarification will remove 30% of the BOD_5 and 60% of the TSS contained in the influent wastewater. The discharge standards will remain the same at 30 mg/L BOD_5 and 30 mg/L TSS on a maximum monthly basis, and thus the SRT will remain the same. Calculate the mass of MLSS (in TSS units), the process solids wastage rate from the bioreactor (in TSS units), and the process oxygen requirements for this application. Neglect process oxygen requirements for nitrification.
9. Biological phosphorus removal is to be added to the domestic wastewater treatment plant described in Study Question 7. How does this affect the mass of biomass (in TSS units), the process solids wastage rate (in TSS units), and process oxygen requirements for this application? Neglect process oxygen requirements for nitrification.
10. The domestic wastewater treatment plant described in Study Question 7 is to be upgraded to provide nitrification. How does this affect the mass of biomass and the process solids wastage rate (both in TSS units)?

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11 Activated Sludge

Activated sludge is the workhorse suspended growth biological wastewater treatment process used to remove a wide range of pollutants. This chapter addresses the use of activated sludge to remove soluble and particulate organic matter and to convert ammonia- and organic-N to nitrate-N. Chapter 12 addresses the use of activated sludge processes to remove the nutrients nitrogen and phosphorus—a group of processes typically referred to as biological nutrient removal processes.

11.1 PROCESS DESCRIPTION

Since its inception by Arden and Lockett in 1914,[5] the activated sludge process has grown in popularity until today it is the most widely used biological wastewater treatment process. Much experimentation has occurred since its initial development and many variations are known. Most of the variations can be grouped into the nine named operations listed under activated sludge in Table 1.2. Theoretical simulations of the performance of many of those variations were presented in Chapters 5, 6, and 7. The purpose here is to describe and compare them as they appear in practice.

11.1.1 General Description and Facilities

Figure 11.1 provides a schematic of a general activated sludge process. Aeration basins are typically open tanks containing equipment to transfer oxygen into solution and to provide mixing energy to keep the mixed liquor suspended solids (MLSS) in suspension. The depth is determined largely by the characteristics of the oxygen transfer/mixing system and typically ranges from 3 to 7.5 m. In special cases, depths as low as 2 m or as great as 20 m may be used. The bioreactor may be constructed of concrete, steel, or as an earthen basin lined with clay or an impermeable membrane. Vertical sidewalls are typically used with concrete and steel structures, while sloping sidewalls are used with earthen structures. A wide variety of bioreactor configurations (i.e., length-to-width ratios) can be used. Bioreactor configuration, the characteristics of the oxygen transfer/mixing equipment, and the distribution of the influent and return activated sludge (RAS) flows will affect the mixed liquor flow pattern within the bioreactor, which will affect process performance, as demonstrated in Chapter 7.

As discussed in Chapter 10, a single device is used both to transfer oxygen and to keep the MLSS in suspension. Typical devices include diffused air (both coarse and fine bubble), floating or fixed mechanical surface aerators (both high speed and low speed), jet aerators, and submerged turbine aerators. Auxiliary mechanical mixers are used when the aeration device does not provide sufficient mixing energy to keep the MLSS in suspension.

Most activated sludge processes use separate clarifiers; exceptions include membrane bioreactor activated sludge (MBRAS), where membranes are used for liquid-solids separation, and sequencing batch reactor activated sludge (SBRAS), where clarification occurs in the same vessel as the biological reactions. In all cases the secondary clarifier or its equivalent provides two functions. One (the clarification function) is removal of the MLSS to produce a clarified effluent that meets the effluent suspended solids goal. The other (the thickening function) is concentration of the settled solids for return to the bioreactor. The return of solids to the bioreactor allows the residence time of the solids in the system (solids retention time, SRT) to exceed the residence time of the liquid (hydraulic residence time, HRT), thereby "intensifying" the reaction rate. A wide range of secondary clarifier configurations can be used, although circular and rectangular are the most common.

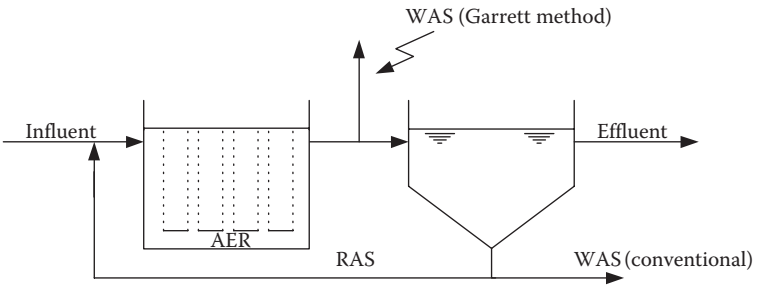

FIGURE 11.1 Typical activated sludge process.

Clarifier configuration has less impact on process performance than bioreactor configuration, as long as the clarifier is sized properly. Clarifiers contain an effluent collection device (typically an overflow weir and effluent collection launder) and a settled solids collection device. Most clarifiers are also provided with equipment to collect solids that float to the surface. The secondary clarifier can provide temporary storage of MLSS transferred from the bioreactor during periods of high flow. This occurs when suspended solids are transferred to the clarifier at a rate greater than they can be removed in the RAS and waste activated sludge (WAS) streams. The solids storage capacity of a secondary clarifier is largely a function of its depth and the settling characteristics of the MLSS, but routine use of the secondary clarifier for solids storage is generally not recommended.

Membrane filters (either microfilters or ultrafilters) are used with MBRAS in place of secondary clarifiers, as described in Chapter 1. Micro- and ultrafilters are particle separation filters that retain a wide range of particulate and colloidal matter. While it is possible for dispersed microorganisms to be retained by membrane filters, the attainment of acceptable filtration rates and the avoidance of rapid membrane fouling require that good bioflocculation occur within the bioreactor. Thus, the biomass in MBRAS has characteristics similar to the biomass produced in other activated sludge processes.[53] Essentially any of the various types of micro- or ultrafiltration membranes can be used with this process. As described in Section 1.3.1 and illustrated in Figure 1.9, the membranes may either be external to the biological reactor or submerged in a portion of it (referred to as submerged or immersed MBRAS). Due to their greater economy, submerged MBRAS systems are most commonly used today. Because of the use of aeration to transport accumulating biomass away from the membrane elements, the RAS flow in MBRAS often contains elevated dissolved oxygen (DO) concentrations, in comparison to other activated sludge options where the DO in the RAS is essentially zero. This difference becomes especially important when MBRAS is used for biological nutrient removal, as described in Chapter 12.

This book focuses on biological process design rather than on facility design. Readers interested in further information about the design of oxygen transfer devices, mixing systems, membranes, and final clarifiers are referred elsewhere.[41,59,67,74,77,79]

11.1.2 Process Options and Comparison

Table 11.1 summarizes the defining characteristics, benefits, and drawbacks of the nine activated sludge process options described in Chapter 1. Conventional activated sludge (CAS) and completely mixed activated sludge (CMAS) require similarly sized bioreactors and clarifiers, resulting in similar capital and operating costs. However, sludge settleability is generally better for CAS than for CMAS, due to greater growth of filamentous microorganisms in CMAS systems. Completely mixed activated sludge offers the benefit of greater resistance to inhibitory shock loads.

In comparison to CAS and CMAS, step feed activated sludge (SFAS) and contact stabilization activated sludge (CSAS) offer the advantage of reduced bioreactor volumes and correspondingly lower capital cost. As discussed in Sections 7.3 and 7.4, this occurs because the feeding pattern

TABLE 11.1
Comparison of Activated Sludge Process Options

Process	Defining Characteristics	Benefits	Drawbacks
Conventional activated sludge (CAS)	• Rectangular bioreactor with tapered aeration • SRT typically 3–8 days, generally less than 15 days • HRT generally 4 to 8 hours for municipal wastewater with primary clarifiers	• Performance well characterized and predictable • Process and facility design well known • Operational parameters well characterized • Useful in a wide range of applications	• Moderate capital and operating costs • Moderate sludge settleability
Step feed activated sludge (SFAS)	• Multiple feed points to each bioreactor • SRT similar to CAS • MLSS gradient leading to lower MLSS concentration entering settler than CAS with same SRT and volume	• Reduced bioreactor requirements compared to CAS • Able to process high peak flows in comparison to CAS • Reduced capital cost compared to CAS	• More complex operation
Contact stabilization activated sludge (CSAS)	• Separate zone containing only RAS • Other zone contains MLSS • SRT similar to CAS	• Reduced bioreactor requirements compared to CAS • Reduced capital cost compared to CAS	• More complex operation • Reduced nitrification efficiency
Completely mixed activated sludge (CMAS)	• Uniform conditions throughout bioreactor • SRT similar to CAS	• Simple design and operation • More resistant to toxic shock loads than others • Useful in a wide range of applications	• Moderate capital and operating costs • Susceptible to the growth of filamentous organisms
Extended aeration activated sludge (EAAS)	• SRT generally 20 to 30 days • Special bioreactor configuration (such as closed loop) and oxygen transfer device often used to reduce power requirements to suspend solids	• Simple design and operation • Reduced sludge production • Organically stable waste sludge • High quality, well-nitrified effluent	• Large bioreactor volumes required • Poor sludge settleability • Higher oxygen requirement
High purity oxygen activated sludge (HPOAS)	• Staged bioreactor with cocurrent flow of liquid and oxygen • SRT of 1.5 to 4 days typical • High volumetric loading—exceeding that achievable with atmospheric oxygen	• Small reactor volume • More resistant to organic shock loads than others • Minimal air emissions	• Mechanically complex • Incompatible with low process loading factors

(*Continued*)

TABLE 11.1 (Continued)
Comparison of Activated Sludge Process Options

Process	Defining Characteristics	Benefits	Drawbacks
Membrane bioreactor activated sludge (MBRAS)	• SRT generally 10 to 25 days • Membrane filter allows high MLSS concentration (7 to 20 g/L)	• Compact system • High quality effluent • Performance independent of sludge settleability	• Peak flow limited by membrane capacity • Membrane fouling • Complex equipment • High capital and operating costs
Selector activated sludge (SAS)	• Small initial bioreactor, often staged • Selector aerated or mixed	• Excellent sludge settleability • Compatible with most activated sludge process options • Proven process	• Increased mechanical complexity • Increased sensitivity to toxic organic compounds
Sequencing batch reactor activated sludge (SBRAS)	• Microbial activity and clarification in same basin	• Simple design and operation • High quality effluent • Process operational characteristics adjustable by adjusting operational cycles	• Discontinuous discharge • Relatively large reactor volumes

produces a distribution of MLSS that results in a higher average MLSS concentration in the aeration basin than in the effluent leaving it and entering the secondary clarifier. Since the clarifier influent MLSS concentration is the limiting factor, the higher average MLSS concentration allows SFAS and CSAS aeration basins to be smaller than those in comparable CAS or CMAS systems. However, the reduced volume may or may not result in reduced cost, depending on actual basin configurations. Contact stabilization activated sludge and SFAS systems are generally more complex. The ability of SFAS and CSAS systems to maintain stable but elevated effluent ammonia-N concentrations (i.e., to partially nitrify) can be a benefit when only partial ammonia-N oxidation is needed.[31] Finally, the ability to adjust the feeding pattern in response to varying hydraulic and organic loading rates can be a significant advantage, especially for SFAS.

Extended aeration activated sludge (EAAS) systems are simple to design and operate, produce a high quality effluent, and produce reduced quantities of more stabilized solids than other, comparable activated sludge systems. This is achieved at the expense of larger and more expensive bioreactors, increased oxygen requirements, and, in some instances, poor solids settling characteristics. As noted in Section 1.3.1 and Table 11.1, unique aeration basin configurations, such as the closed loop bioreactor, are used to reduce the energy input required to maintain solids in suspension and to better balance energy inputs required for mixing and oxygen transfer.

High purity oxygen activated sludge (HPOAS) offers the benefits of small bioreactor volumes, the ability to meet high oxygen requirements caused by shock loads of biodegradable organic matter, and minimal air emissions. In contrast, they are mechanically complex and incompatible with situations requiring low process loading factors because under those circumstances high purity oxygen is not necessary to meet oxygen requirements. Historically, the use of high purity oxygen in activated sludge systems was claimed to result in an altered biological process with reduced sludge production rates, reduced oxygen requirements, and improved sludge settleability.[40] However, more thorough analysis demonstrated that the biological characteristics of HPOAS systems are the same as other activated sludge systems.[13,45]

The replacement of the secondary clarifier by a membrane filtration unit in an MBRAS system significantly changes the characteristics of the system and the quality of the effluent produced. The use of membrane filtration allows higher MLSS concentrations to be maintained, thereby allowing relatively long SRTs within modestly sized aeration basins. The long SRT, coupled with excellent removal of particulate and colloidal matter by the membrane filters, results in a very high quality effluent. The systems are also quite compact since smaller aeration basins and no secondary clarifiers are needed. However, membrane systems are relatively expensive and complex, and the control of membrane fouling is a significant issue. Energy costs are generally higher because of the energy required to operate the membrane system and the reduced process oxygen transfer efficiency resulting from the elevated MLSS concentrations often used.[14,53,59,77]

Selector activated sludge (SAS) systems were developed to improve activated sludge settleability, and they have proven capable of doing so.[17,36,62,72] Selectors can be incorporated into most of the activated sludge options, although they may make the biomass more susceptible to inhibitory shock loads because of the high process loading factors in them.

The sequencing batch reactor activated sludge (SBRAS) process offers the benefits of simple design and operation, the production of a high quality effluent, and the capability to adjust process operational characteristics by adjusting operational cycles.[34] However, effluent is removed on a discontinuous basis, which can negatively impact the operation of downstream treatment processes unless the effluent flow is equalized. Relatively large bioreactors are required to accommodate the addition of influent wastewater and the removal of treated effluent.

11.1.3 Typical Applications

Activated sludge is the most widely utilized biochemical operation, with many thousands of operating examples. It has been used to treat nearly every type of wastewater that contains biodegradable

organic matter and has been applied to flows ranging up to 5,000,000 m^3/day. Extensive documentation of such applications exist in the literature and in standard references.[23,41,74,75]

Activated sludge is applicable to wastewaters with a wide range of concentrations. As illustrated in Figure 10.2, aerobic suspended growth bioreactors are typically used to treat wastewaters with biodegradable chemical oxygen demand (COD) concentrations up to about 4000 mg/L. Above that, anaerobic bioreactors are often more cost-effective. Technically, activated sludge systems with clarifiers can be used to treat wastewaters with biodegradable COD concentrations ranging from about 50 to 10,000 mg/L. The technical constraints are derived from sludge settleability and the ability to maintain a suitable SRT. For dilute wastewaters, the rate of generation of new biomass may be so low that it is less than the rate at which biomass is lost in the effluent from a clarifier. Under such circumstances, it is impossible to accumulate biomass in the bioreactor and the necessary SRT cannot be achieved. This constraint is removed with MBRAS. Attached growth bioreactors, which employ alternate mechanisms for biomass retention, provide other options for dilute wastewaters.

For high concentration wastewaters, the constraint relates to the ability of a clarifier to consolidate the MLSS for recycle to the bioreactor. If a wastewater contains a high concentration of biodegradable COD, the concentration of MLSS produced could exceed the thickening capacity of a clarifier. For example, treatment of a wastewater with a biodegradable COD concentration of 20,000 mg/L could result in the production of 8000 mg/L of total suspended solids (TSS). This exceeds the typical MLSS concentrations achievable in activated sludge systems, as discussed in Section 10.3.4. Thus, it would not be possible to settle this mixed liquor in a clarifier and produce a more concentrated RAS stream for recycle to the bioreactor, thereby making it impossible for the SRT to exceed the HRT. This constraint is relaxed with MBRAS systems, which can sustain MLSS concentrations up to 20,000 mg/L. When the wastewater concentration is too high for activated sludge systems, either a lagoon within which the HRT and the SRT are the same or an anaerobic process that has a lower yield would have to be used.

Plant size is one of the factors that affects the type of activated sludge process used, particularly for municipal applications. Extended aeration activated sludge and SBRAS are most often used in smaller municipal wastewater treatment plants, those with design flows up to about 20,000 to 60,000 m^3/day. Their simple operation and reliable performance more than offset the increased cost associated with the larger bioreactors required. Primary clarifiers are also typically not provided to simplify the solids treatment and disposal system. Conventional activated sludge and CMAS are most often used in larger plants, typically with primary clarifiers. The cost savings associated with the smaller bioreactors needed for these processes are sufficient to justify the more intensive operation required. The use of primary clarifiers is an economic—not a process—issue. For example, primary clarifiers can be used successfully in association with EAAS processes, whereas CAS or CMAS systems can be successfully designed and operated without primary clarifiers. The smaller bioreactors and reduced off-gas volumes associated with HPOAS make this process useful for larger plants located in urban areas where compact designs are needed.

Extended aeration activated sludge is also often used to treat industrial wastewaters that contain slowly biodegradable, xenobiotic, and/or inhibitory materials. The long SRT used with this process is often necessary to treat these difficult to degrade materials. Furthermore, SBRAS is being used increasingly in industrial applications due to its simplicity and the periodic nature of some industrial discharges. The long SRT often used with the SBRAS process is another factor in its selection for these applications. Likewise, MBRAS is being increasingly used in these applications due to the long SRT that can be maintained and the increased capability of these systems to retain biomass and produce an effluent with low concentrations of particulate matter.

As discussed in Section 10.3.2, a long SRT is typically needed when nitrification is to be maintained to control effluent ammonia-N concentrations. Extended aeration activated sludge, MBRAS, and SBRAS systems typically nitrify under all conditions due to the long SRTs associated with them. Conventional activated sludge, CMAS, SFAS, and CSAS processes can also be designed

with an SRT sufficiently long to achieve reliable nitrification. Furthermore, SFAS and CSAS can be designed and operated to achieve partial oxidation of ammonia-N, as illustrated in Sections 7.3 and 7.4. On the other hand, HPOAS is not generally used for nitrification applications due to its low pH and short SRT.

Membrane bioreactor activated sludge is being used increasingly because of its compact nature, ease of automation, reliable performance, and high quality effluent. These features make MBRAS ideal for remotely located water reclamation plants and for industrial applications.[60] While MBRAS has historically been used only in small wastewater treatment plants, subsequent developments are allowing it to be used in larger plants.[14] Ready for application in both developed and developing countries, MBRAS may play an important role in a wide range of water reclamation systems.[18,22]

In some instances, a wastewater treatment plant may be designed so that it can be operated in more than one activated sludge mode, depending on the nature of the influent wastewater. Many plants experience significant increases in influent flow during wet weather, either because of the infiltration of rainwater into the sanitary sewer system or because the collection system is a combined sewer system. As a consequence, CAS may be used during dry weather to provide the maximum treatment efficiency and the greatest ease of operation, whereas either SFAS or CSAS may be used during wet weather to allow higher hydraulic loadings to be processed while maintaining adequate performance.[55] In fact, due to the reduced pollutant concentrations that often exist during high flow periods, effluent quality may be no different than that achieved during dry weather using the CAS process.

11.2 FACTORS AFFECTING PERFORMANCE

Many factors affect the performance of activated sludge systems and the more important ones will be discussed in this section. They also affect design and operation, and their impacts will be discussed in those contexts in Sections 11.3 and 11.4.

11.2.1 Floc Formation and Filamentous Growth

As discussed in Section 2.3.1, successful operation of activated sludge systems requires development of a flocculent biomass that settles rapidly and compacts properly in the clarifier; failure to do so can lead to process failure. Individual bacteria are colloidal in size. They will not settle in conventional clarifiers or in SBRAS, and will foul the membranes used in MBRAS at an unacceptable rate. Consequently, they must be aggregated (i.e., flocculated). An ideal activated sludge floc is strong and compact so that it settles or filters rapidly, producing a dense sludge for recycle to the bioreactor and a clear, high quality supernatant for discharge as treated effluent. Achieving this objective requires the proper proportion of floc-forming and filamentous bacteria.

As noted in Section 2.3.1, SRT has an important impact on bioflocculation; often a value of at least three days is recommended for achieving good bioflocculation in activated sludge systems. Nevertheless, many municipal wastewater treatment plants have been successfully designed and operated at SRTs as low as one day.[23,27,55] Apparently, the presence of microorganisms in municipal wastewaters, which alters the relationship between SRT and specific growth rate as discussed in Section 5.2.3, and the presence of extracellular polymeric substances (EPSs), result in a reduction in the SRT required to achieve good bioflocculation. This reduction is reflected in the typical SRT operating ranges presented in Figure 10.3. The relative concentrations of monovalent and divalent cations can also affect bioflocculation and sludge settling characteristics. In general the ratio of monovalent to divalent cations (on a milli-equivalent basis) should be less than 0.5 to produce a well-flocculated biomass with good settling characteristics.[33,36]

Good bioflocculation allows for the retention of floc-forming bacteria in activated sludge systems. In addition to the production of EPS, the filament population must be managed to produce the

strong flocs needed for optimum performance while avoiding the excessive filament quantities that produce a sludge that settles slowly and compacts poorly. About 30 morphologically different types of filamentous bacteria routinely exist in activated sludge systems. However, they can be organized into four groups, as summarized in Table 11.2. Group I, Low DO Aerobic Zone Growers, are encouraged by a low DO concentration. In fact, in some instances the bioreactor DO concentration can be manipulated to control the relative proportion of floc-forming and filamentous bacteria. Group II, Mixotrophic Aerobic Zone Growers, oxidize sulfide to sulfur granules. As indicated in Table 2.5, the filamentous bacteria *Thiothrix*, *Beggiatoa*, and 021N can obtain energy from the oxidation of hydrogen sulfide, which provides an advantage for them when it is present. Group III, Other Aerobic Zone Growers, and Group IV, Aerobic, Anoxic, and Anaerobic Zone Growers, are present due to the growth rate environment created by the configuration of the bioreactor (as described below) and due to a longer SRT. Low pH will encourage the growth of filamentous fungi. Table 11.2 summarizes control methods for these groups of filamentous organisms. It is generally desirable to use permanent filament control methods such as those listed in Table 11.2 and discussed below. However, in

TABLE 11.2
Proposed Filamentous Organism Groups

	Group I—Low DO Aerobic Zone Growers
Features	• readily metabolizable substrates • low DO • wide SRT range
Organisms	*S. natans,* Type 1701, *H. hydrossis*
Control	• aerobic, anoxic, or anaerobic selectors • increase SRT • increase aeration basin DO concentration
	Group II—Mixotrophic Aerobic Zone Growers
Features	• readily metabolizable substrates, especially low molecular weight organic acids • moderate to high SRT • sulfide oxidized to stored sulfur granules • rapid nutrient uptake rates under nutrient deficiency
Organisms	Type 021N, *Thiothrix spp.*
Control	• aerobic, anoxic, or anaerobic selectors • nutrient addition • eliminate sulfide and/or high organic acid concentrations (eliminate septicity)
	Group III—Other Aerobic Zone Growers
Features	• readily metabolizable substrates • moderate to high SRT
Organisms	Type 1851, *N. limicola spp.*
Control	• aerobic, anoxic, or anaerobic selectors • reduce SRT
	Group IV—Aerobic, Anoxic, Anaerobic Zone Growers
Features	• grow in aerobic, anoxic, and anaerobic zones • high SRT • possible growth on hydrolysis products of particulates
Organisms	Type 0041, Type 0675, Type 0092, *M. parvicella*
Control	Largely unknown but: • maintain uniformly adequate DO in aerobic zone and stage the aerobic zone

Source: Adapted from Jenkins, D., Richards, M. G., and Daigger, G. T., *The Causes and Control of Activated Sludge Bulking, Foaming, and Other Solids Separation Problems,* 3rd ed., Lewis Publishers, Ann Arbor, MI, 2004.

certain instances it may be more economical to use nonspecific toxicants such as chlorine or hydrogen peroxide to control filament growth. The use of such techniques is discussed in Section 11.4.3.

Classically the presence of filaments in an activated sludge system is explained based on the relative growth kinetics of filamentous and floc-forming bacteria, as illustrated in Figure 11.2. In general, for a particular substrate, floc-forming bacteria have higher Monod parameter ($\hat{\mu}$ and K_S) values than filamentous bacteria. In other words, the floc formers can grow faster when the substrate concentration is high, but the filamentous bacteria have a higher affinity for the substrate and can grow faster when its concentration is low. For example, if the substrate concentration is S_1 in Figure 11.2, the specific growth rate of the floc-forming bacteria is higher than that of the filaments and the floc formers will out-compete the filaments. If, on the other hand, the substrate concentration is S_2, the specific growth rate of the filaments is higher than that of the floc formers and the filaments will out-compete the floc formers. This illustrates the characteristics of an environment that favors the growth of filamentous bacteria: the substrate must be supplied continuously in a manner that results in a low concentration. Continuous substrate supply is required so that biomass growth can occur. The residual concentration must be low to provide a competitive advantage for the filamentous organism. Said simply, filamentous organisms are good scavengers; they consume substrates more efficiently than floc-forming bacteria. Consequently, filaments proliferate under conditions that favor scavenging organisms. Such conditions typically occur in CMAS and sometimes occur in CAS, CSAS, and SFAS.

The general objective of the activated sludge process is to remove biodegradable organic matter. This is achieved by creating conditions in which it is the limiting substance. Consequently, the presence of filamentous bacteria with a high affinity for nitrogen, phosphorus, or DO (Groups I and II) indicates that these nutrients may be limiting bacterial growth. The solution to problems caused by excessive growths of these filamentous bacteria is the addition of the limiting nutrient. For nitrogen and phosphorus, residual concentrations of approximately 1 mg/L are desired. For DO, the required residual concentration is a function of the process loading factor or respiration rate, as illustrated in Figure 11.3.[44] This relationship exists because DO concentrations are measured in the bulk solution while bacterial growth occurs within the floc particle. As the process loading factor is increased, the biomass uses oxygen at a faster rate and a higher bulk DO concentration is required to ensure the penetration of DO throughout the floc particle.

Other filaments are present because of the configuration of the bioreactor. These filaments, principally of Groups III and IV, are more competitive than floc formers when readily biodegradable organic matter is consistently supplied, but maintained at uniformly low concentrations. Variations

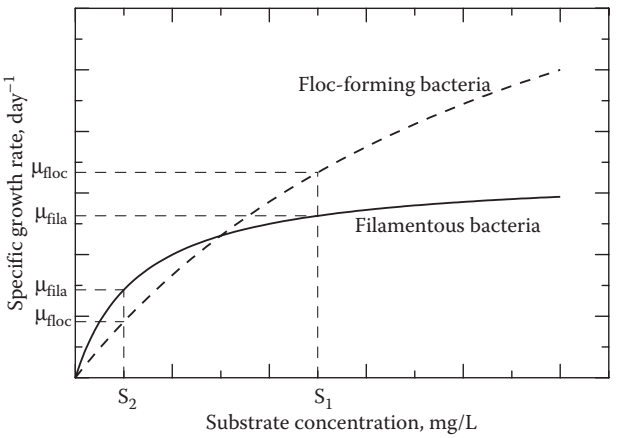

FIGURE 11.2 Comparison of the growth kinetics of floc forming and filamentous bacteria.

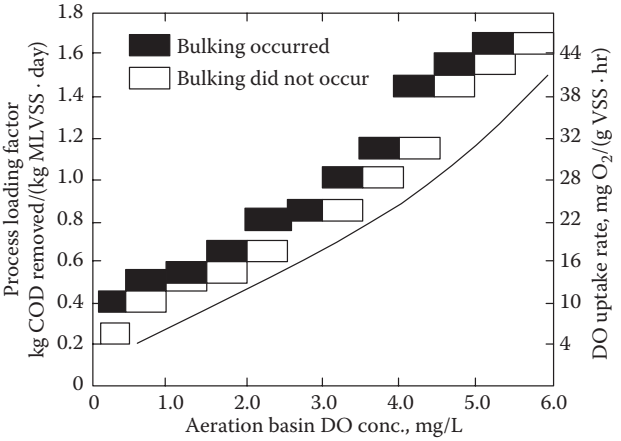

FIGURE 11.3 Combinations of process loading factors and aeration basin dissolved oxygen concentrations where bulking and nonbulking sludges occur in CMAS system. (Reprinted from Palm, J. C., Jenkins, D., and Parker, D. S., Relationship between organic loading, dissolved oxygen concentration and sludge settleability in the completely-mixed activated sludge process. *Journal, Water Pollution Control Federation,* 52:2484–506, 1980. With permission. Copyright © Water Environment Federation Alexandria, Virginia.)

in readily biodegradable organic matter concentrations through activated sludge flocs can also contribute to this effect.[38] One approach for controlling the growth of filamentous bacteria with these characteristics is to configure the bioreactor to create a substrate concentration gradient through it. This is known as kinetic selection because it is based on the relative kinetics of floc-forming and filamentous bacteria. The goal is to produce a substrate concentration at the inlet to the bioreactor that favors the growth of floc-forming bacteria at the expense of the filamentous bacteria, as illustrated in Figure 11.2 by the concentration S_1. Likewise, the high concentration of readily biodegradable organic matter allows it to penetrate throughout the entire activated sludge floc particle, which encourages the development of compact sludge flocs that settle rapidly and compact well. In some cases extremely compact and readily settleable flocs called granular floc particles can develop.[21] In either case these conditions can be created by providing highly plug-flow conditions within the bioreactor, which results in the rapid removal of readily biodegradable organic matter. Finally, the uptake of readily biodegradable organic matter in the initial portion of the bioreactor minimizes its availability in the remaining portion of the bioreactor, thereby contributing to the control of Group I and II filaments.

Because of the effects of substrate concentration gradients on the competition between filamentous and floc-forming bacteria, the sludge volume index (SVI) of activated sludge is influenced by the residence time distribution in the bioreactor as characterized by the equivalent number of tanks in series. This is illustrated in Figure 11.4.[66] In a process with a low equivalent number of tanks in series, the residual readily biodegradable substrate concentration in the first tank will be relatively low, similar to S_2 in Figure 11.2. This low concentration favors the growth of filamentous bacteria, resulting in a high SVI. As the number of equivalent tanks in series is increased, the readily biodegradable substrate concentration in the first tank increases until it approaches S_1, which favors the growth of floc-forming bacteria. Consequently, SVIs are generally low for bioreactors with a flow pattern characterized as five tanks in series or more.

The desired conditions in the initial equivalent bioreactor have been identified by calculating the process loading factor in that tank and correlating the activated sludge SVI with it. The process loading factor for the initial tank is calculated by using the mass flow rate of biodegradable organic matter in the process influent and the mass of biomass in the equivalent initial tank. As discussed in Section 5.1.7, the process loading factor is linearly related to specific growth rate. Thus, such correlations identify the specific growth rate in the initial equivalent tank required to produce a residual

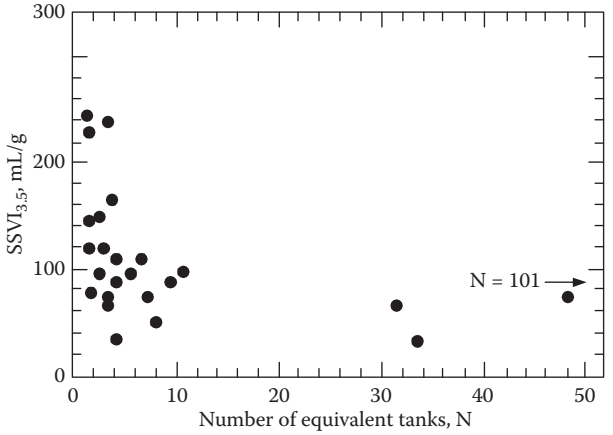

FIGURE 11.4 Relationship between activated sludge settling characteristics as expressed by the SVI and the bioreactor hydraulic characteristics as expressed by the equivalent number of tanks in series. (Reprinted from Tomlinson, E. J., and Chambers, B., *The Effect of Longitudinal Mixing on the Settleability of Activated Sludge*. Technical Report TR 122, Water Research Centre, Stevenage, 1978. With permission from Water Research Centre.)

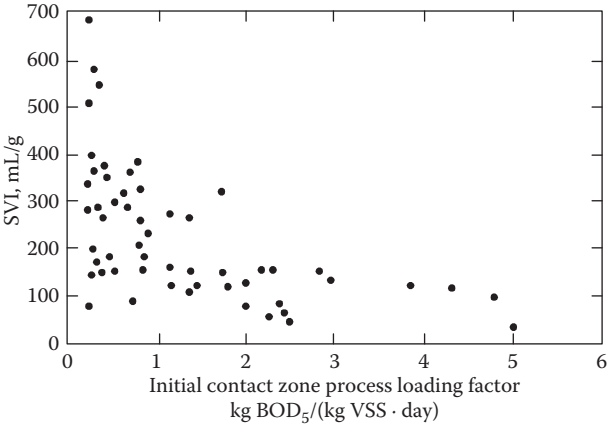

FIGURE 11.5 Effect of initial contact zone process loading factor on SVI. (Reprinted from Tomlinson, E. J., *Bulking—A Survey of Activated Sludge Plants*. Technical Report TR35, Water Research Centre, Stevenage, 1976. With permission from Water Research Centre.)

biodegradable substrate concentration comparable to S_1 in Figure 11.2. One such correlation is presented in Figure 11.5[65] for the full-scale wastewater treatment plants considered in Figure 11.4. The results indicate that, for these plants, an initial process loading factor of 2 kg of five-day biochemical oxygen demand (BOD_5)/(kg volatile suspended solids, VSS,·day) produces a sufficiently high specific growth rate to encourage the growth of floc-forming bacteria over filamentous bacteria. The use of correlations such as these to design SAS systems will be discussed in Section 11.3.4.

Another approach for controlling the growth of filaments with a high affinity for readily biodegradable organic matter is by metabolic selection. Metabolic selection is accomplished by eliminating DO as a terminal electron acceptor in the selector and either providing nitrate-N to create an anoxic selector or excluding both DO and nitrate-N to create an anaerobic selector. Some strains of floc-forming bacteria are able to take up readily biodegradable organic matter under either anoxic

or anaerobic conditions, whereas many filamentous bacteria cannot. Thus, control of the terminal electron acceptor provides a powerful selective pressure against filamentous bacteria. Metabolic selectors are discussed in Chapter 12.

Group IV filamentous bacteria are typically observed in processes with long SRTs, especially those with completely mixed bioreactors.[9,26,36,62] Some evidence suggests that the growth of at least some of these organisms is encouraged by cyclic low DO concentrations, which can occur with oxygen transfer systems such as mechanical surface aerators.[36,47,62] Their growth is controlled by maintaining plug-flow conditions and uniform DO concentrations throughout the bioreactor. The occurrence of these organisms in biological nutrient removal systems is discussed further in Chapter 12.

A recent study of 44 full-scale activated sludge systems confirms these observations.[30] Aerobic, anaerobic, and anoxic selectors effectively improved sludge settleability for short SRT activated sludge systems. A gradient in the process loading factor was necessary for effective selection for plants using aerobic selectors whereas the key to effective performance of anoxic and anaerobic selectors was the complete removal of the readily biodegradable organic matter. A gradient in the process loading factor was not necessary for them, but staging helped to minimize short-circuiting of readily biodegradable organic matter into the main aerobic zone. Selectors by themselves did not effectively control sludge settleability in long-SRT activated sludge plants, indicating the importance of aerobic zone configuration. Bench-scale results demonstrated that selectors remove readily biodegradable organic matter, but only limited amounts of slowly biodegradable organic matter.[81] The presence of slowly biodegradable organic matter in a completely mixed aerobic zone contributes to poorer sludge settleability. Staging of anoxic selectors can improve the removal of slowly biodegradable organic matter and improve sludge settleability.

11.2.2 Solids Retention Time

It should be clear by now that SRT is a primary factor determining the performance of activated sludge systems. The theoretical impacts of SRT on the concentration of soluble constituents for a variety of activated sludge processes were discussed in Chapters 5 through 7. The role of SRT in achieving bioflocculation was discussed in Section 2.3.1. Finally, the general factors that must be considered in the selection of the SRT for all biochemical operations were presented in Section 10.3.2. This section emphasizes certain common observations concerning the effect of SRT on activated sludge process performance.

Figure 10.3 illustrated that the design and operating SRT for activated sludge systems treating biogenic organic matter is generally controlled by bioflocculation, not the removal of soluble substrate. To demonstrate this point, Figure 11.6 shows the effect of SRT on the effluent soluble COD from CMAS bioreactors that received a feed with a soluble COD of 375 mg/L.[8] Soluble COD decreased rapidly as the SRT was increased from 0.25 to 1 day. At an SRT of two days, the COD reached a minimum value, which was maintained until the SRT exceeded eight days. Beyond eight days the COD increased again, probably as a result of the production of soluble microbial products.[16,49] Since those products are resistant to biodegradation, the biodegradable COD was essentially constant for SRTs in excess of two days. However, Figure 2.3 revealed that the SRT for the same bioreactors had to exceed two days to obtain effective bioflocculation. Thus once the SRT was long enough for effective bioflocculation to occur, further increases had only minor effects on soluble substrate removal. Consequently, for easily degradable substrates like those in domestic wastewaters, selection of the SRT is almost always controlled by factors other than soluble substrate removal.

The preceding paragraph helps to explain another common observation; that is, bioreactor configuration often has no observable impact on soluble effluent quality for many applications.[64] Although we saw in Section 7.2.2 that CAS systems theoretically have lower effluent soluble

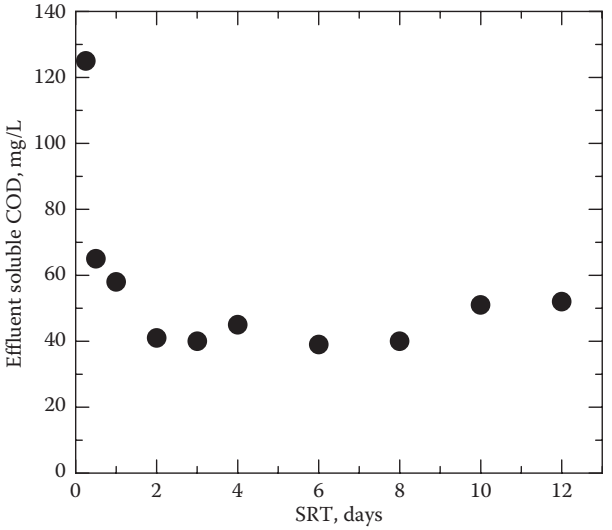

FIGURE 11.6 Effect of SRT on effluent soluble COD from a CMAS system. (Data from Bisogni, J. J., and Lawrence, A. W., Relationships between biological solids retention time and settling characteristics of activated sludge. *Water Research,* 5:753–63, 1971.)

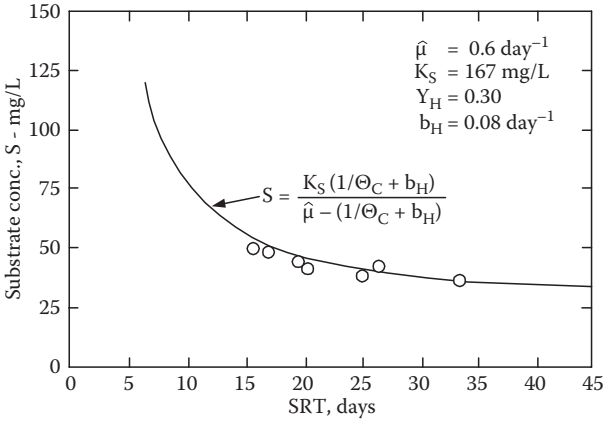

FIGURE 11.7 Effect of SRT on effluent COD from a full-scale CMAS reactor. (Adapted from Campbell, H. J., and Rocheleau, R. F., Waste treatment at a complex plastics manufacturing plant. *Journal, Water Pollution Control Federation,* 48:256–73, 1976.)

substrate concentrations than CMAS systems, little difference is observed in practice. Consequently, for treatment of easily biodegradable substrate, the choice of the activated sludge process variation is usually driven by factors other than soluble effluent quality.

Longer SRT values may be required for the treatment of industrial wastewaters containing more difficult to degrade materials, which may also be inhibitory to biological growth. This is illustrated in Figure 11.7, which presents data from a CMAS process treating a plastics manufacturing wastewater.[11] However, note that SRT had relatively little effect on effluent quality over the operating range for the facility. This is typical in that most activated sludge systems are designed and operated over a range of SRTs where little difference in the concentration of soluble organic matter in the effluent is observed. When the kinetic parameters for a particular application have been determined, Equation 10.7 can be used to calculate the SRT required to achieve a specified target effluent quality from a CMAS process. Because the substrate is slowly biodegradable, however,

bioreactor configuration is likely to have more of an effect than it did with domestic wastewater. Consequently, consideration of other activated sludge process options will require pilot studies or simulations to arrive at a design SRT value or the application of safety factors, much as is required for nitrification.

While operation at relatively short SRTs is possible for many wastewaters, it can result in high excess solids production rates because short SRTs do not permit hydrolysis of particulate organic matter,[42] as discussed in Section 6.2. This will result in a corresponding reduction in process oxygen requirements, consistent with the COD mass balance described by Equation 3.94 and articulated in the Guiding Principle No. 3 of Section 10.1. The economic consequences of increased solids production and reduced oxygen requirement must be evaluated for each application.

Activated sludge systems are often designed to operate at long SRTs to achieve stabilization of entrapped particulate organic matter and heterotrophic biomass or to biodegrade xenobiotic and other slowly biodegradable organic compounds. This can lead to limited growth of filamentous bacteria resulting in pinpoint floc, and thus measures should be taken to minimize its formation. One technique is to use completely mixed reactors, which will encourage the growth of filaments. Another is careful control of the DO concentration to encourage the growth of a few low DO filaments. Both techniques require care on the part of the designer and operator to prevent excessive growth of filaments, with their attendant problems. Nevertheless, it is important to recognize that pinpoint floc is not an inevitable consequence of a long SRT operation. On the other hand, inadequate bioflocculation occurs with some industrial wastewaters and is exacerbated by operation at a long SRT. Thus, experience with particular wastewaters should be used to determine whether long SRT operation is possible and to establish the measures necessary for achieving both adequate bioflocculation and sufficient filaments to build a stable floc.

Although nitrification may not be required at a particular facility, it will occur any time the SRT exceeds the minimum SRT for the nitrifying bacteria. As illustrated in Figure 10.4, nitrification can occur at SRTs as low as one to two days when the mixed liquor temperature exceeds 20°C. To avoid the operational difficulties resulting from an inadequate oxygen supply, the oxygen transfer system should be designed for the oxidation of both carbon and nitrogen if the SRT could possibly be long enough to allow nitrification.

Both organic substrate removal and nitrification may be required in some situations, and an SRT must be selected that will allow both effluent quality goals to be met. As discussed in Section 10.3.2, this requires the design SRT to exceed the minimum SRT of the most slowly growing microorganisms by a sufficient degree to have stable performance. Nitrifying bacteria are usually the most slowly growing bacteria in activated sludge systems and thus they generally control the design in this situation. Selection of the design SRT for nitrification is considered in Section 11.3.2.

Membrane bioreactor activated sludge is generally operated at SRTs sufficiently long to achieve reliable nitrification. This occurs for two reasons. First, research and experience indicate that longer SRTs are needed to produce a mixed liquor that can be effectively filtered in the membrane unit. Second, the minimum size of the bioreactor is limited by oxygen transfer (Section 10.3.4), making sufficient bioreactor volume available to easily achieve longer SRTs at the high MLSS concentrations possible in MBRAS.

The nature and capacity of the excess solids processing system is influenced strongly by the SRT of an activated sludge process. Low SRTs, on the order of one to three days, require a solids processing system with the capability to continuously receive and process large and variable quantities of WAS. At such SRTs a substantial portion of the MLSS inventory is wasted each day. Consequently, little capacity exists within the activated sludge process to absorb variations in inventory caused by variations in process loadings or by interruptions in waste solids processing. Instead, the capacity to absorb such variations must be built into the solids processing system. In contrast, an activated sludge process operating at an SRT of 5 to 10 days generally has sufficiently large bioreactors to temporarily accumulate excess solids. Consequently, the solids processing system serving it need

not have a large capacity to absorb variations. Finally, activated sludge systems with very long SRTs, such as EAAS, often practice periodic solids wasting, and the excess solids processing system should be sized accordingly. In order to foster process stability, any given incident of solids wasting should not result in a decrease in the bioreactor MLSS concentration of more than 10%. For a process operating at an SRT of 20 days, solids need to be wasted only every other day to meet this criterion.

11.2.3 Mixed Liquor Suspended Solids Concentration

Activated sludge processes can be successfully operated over a wide range of MLSS concentrations. In fact, the MLSS concentration itself does not affect the performance of the process; rather, performance is controlled by the mass of MLSS present. Furthermore, as expressed by the Guiding Principle No. 5 in Section 10.1, once the SRT for a biochemical operation has been selected, the mass of biomass in it becomes fixed. As discussed in Section 10.3.4, selection of the MLSS concentration (and consequently, the bioreactor volume) requires consideration of the interactions of the bioreactor with the liquid-solids separation system and the mixing/aeration system. Although a wide range of MLSS concentrations is possible, practical designs typically limit the MLSS concentration between 2000 and 5000 mg/L for clarifier-based systems such as CAS, EAAS, and CMAS.

Higher MLSS concentrations can be sustained in MBRAS systems because membranes can successfully process such concentrations. Mixed liquor suspended solids concentrations as high as 20,000 mg/L or more have been successfully maintained in MBRAS, although such high concentrations begin to adversely impact membrane hydraulic capacity.[14,59,77] For current membrane systems, performance is generally optimized when the MLSS concentration in the membrane zone is limited to about 10,000 mg/L.[77] This is generally accomplished by limiting the MLSS concentration in the flow entering the membrane section to no more than about 8000 mg/L and maintaining a recirculation flow from the membrane section back to the bioreactor of four-to-one. Elevated MLSS concentrations—in excess of 10,000 mg/L—can also result in reduced oxygen transfer performance as the mixed liquor becomes viscous.[53,77] Thus, current best practice for MBRAS is to limit MLSS concentrations to less than 10,000 mg/L.

A minimum MLSS concentration is necessary to allow the development of a flocculent biomass. If the process is operated at MLSS concentrations below this value, bioflocculation will be poor, entrapment of particulate organic matter will be inadequate, and a good settling activated sludge floc will not develop. The result will be a turbid, poor quality effluent. Although actual experience must define the minimum MLSS concentration for a particular process, they typically fall between 500 and 1000 mg/L. Thus, the maximum range in MLSS concentration typically used in activated sludge systems is between 500 and 5000 mg/L for systems with clarifiers and between 1000 and 10,000 mg/L for MBRAS systems.

11.2.4 Dissolved Oxygen

The primary effect of the DO concentration on activated sludge performance is on the growth of filamentous bacteria. Guidance on the selection of the appropriate DO concentration to control low DO filamentous bulking is presented in Figure 11.3. Many books recommend the maintenance of a minimum DO concentration of 2 mg/L in activated sludge processes. As indicated by Figure 11.3, however, 2 mg/L may not be sufficient in some cases and may be excessive in others. Rather, the required DO concentration depends on the process loading factor and specific oxygen uptake rate (SOUR) in a given bioreactor. Although it is considered prudent to design oxygen transfer systems to achieve a DO concentration of at least 2 mg/L, many activated sludge processes operate quite satisfactorily at lower DO concentrations, thereby achieving significant power cost savings.[79] It all depends on the process loading factor imposed.

11.2.5 Oxygen Transfer and Mixing

For economic reasons, the equipment used to transfer oxygen in activated sludge systems also provides the turbulence necessary to maintain solids in suspension. This results in constraints on process design and operation. One concern is the volumetric power input, Π, which is the power supplied per unit volume, either directly by mechanical aerators or indirectly by compression of the air for diffused aeration systems. As discussed in Section 10.3.4 and illustrated in Figure 10.6, the volumetric power input must be sufficiently large to keep solids in suspension, but not so large as to cause excessive floc shear. Another concern is the maximum volumetric oxygen transfer rate (i.e., the mass of oxygen transferred per unit time per unit volume) that can be attained economically with the equipment available.

As discussed in Sections 5.1.6 and 10.4, once the SRT has been selected for an activated sludge process, the oxygen requirement is fixed. This, in turn, fixes the amount of power that must be expended to supply the oxygen. Although this book does not address the design of oxygen transfer systems, it is necessary to approximate the power required to ensure that the volumetric power input is in the feasible region as indicated by Figure 10.6. For mechanical aeration systems, the required power input can be approximated from

$$P = \frac{RO}{\eta_P}, \tag{11.1}$$

where P is the power input in kW, RO is the oxygen requirement in kg/hr, and η_P is the in-process energy efficiency for the mechanical aeration system in kg O_2/(kW·hr). The value of η_P typically ranges from 0.7 to 1.2 kg O_2/(kW · hr).[79] For diffused air systems, the process air requirement can be calculated from the following dimensional expression:

$$Q = \frac{6.0\ RO}{\eta_Q}, \tag{11.2}$$

where Q is the air flow rate in m^3/min, RO is the oxygen requirement in kg/hr, and η_Q is the field oxygen transfer efficiency expressed as the percentage of the oxygen in the air actually transferred to the liquid. The value of η_Q depends on the nature of the diffuser and the depth at which the air is released.[79] It typically lies in the range of 6 to 15%, but with values as low as 4% and as high as 80% observed under unusual circumstances. The volumetric power input required to meet the oxygen requirement is obtained by dividing P or Q by the bioreactor volume.

As illustrated in Figure 10.6, the lower limit on the volumetric power input, Π_L, is determined by the need to maintain solids in suspension. For mechanical aeration systems, this input, $\Pi_{L,P}$, is around 14 kW/1000 m^3. The manufacturer of a particular aeration device should be consulted for a more exact value. For spiral roll diffused aeration systems, a minimum air input rate, $\Pi_{L,Q}$, of 20 m^3/(min·1000 m^3) is generally required. For full floor coverage aeration systems, the corresponding value is 37 m^3/(min·1000 m^2 of bioreactor floor area).[41,67,74,79] If these requirements exceed the volumetric power input required to meet the oxygen requirements, the power input required for solids suspension must be provided. This results in increased power usage and higher DO concentrations than required purely for process reasons. Consequently, the minimum volumetric power input can be used with the power required to meet the minimum oxygen requirement to establish the upper feasible bioreactor volume, V_U, in m^3:

$$V_U = \frac{1000\ P}{\Pi_{L,P}} \text{ or } \frac{1000\ Q}{\Pi_{L,Q}}. \tag{11.3}$$

The volumetric power input must also be less than the value that causes excessive shear of the activated sludge floc, Π_U. Excessive shear will disperse a portion of the solids into a poorly settleable form that will not be removed in a clarifier and will, therefore, pass into the effluent. For mechanical aeration systems, the typical maximum volumetric power input, $\Pi_{U,P}$, is 60 kW/1000 m^3, while for diffused aeration systems $\Pi_{U,Q}$ is approximately 90 m^3/(min·1000 m^3).[41,67,74,79] To avoid excessive floc shear, the reactor volume should be no smaller than the lower feasible bioreactor volume associated with Π_U, which has been designated $V_{L,FS}$ to emphasize that it is determined by the floc shear criterion. Its value in m^3 is given by

$$V_{L,FS} = \frac{1000\ P}{\Pi_{U,P}} \quad \text{or} \quad \frac{1000\ Q}{\Pi_{U,Q}}. \tag{11.4}$$

The necessity to indicate that the minimum feasible volume calculated with Equation 11.4 comes from the floc shear criterion stems from the fact that there is a maximum volumetric rate at which oxygen can be transferred in activated sludge systems, and it also imposes a lower limit on the bioreactor volume. This maximum rate is device specific, and the manufacturer of the particular equipment of interest must be contacted to determine the appropriate maximum value for a given application. For example, for a floor coverage diffused air system, the limitation may be caused by the maximum number of diffusers that can be placed in the bioreactor per unit of floor area. Nevertheless, for the types of oxygen transfer systems typically used today, the maximum volumetric oxygen transfer rate that can be achieved economically on a sustainable basis is around 100 g O_2/(m^3·hr), which is equivalent to 0.10 kg O_2/(m^3·hr). During short-term transients this rate can sometimes be pushed to 150 g O_2/(m^3·hr), but typical mechanical aeration equipment should not be counted on to deliver oxygen at such a high rate on a sustainable basis because of excessive wear. Thus, if such high transfer rates are needed, specialized high efficiency transfer systems must be used. As a result of this constraint, the lower limit on bioreactor volume based on oxygen transfer, $V_{L,OT}$, should also be calculated:

$$V_{L,OT} = \frac{RO}{0.10}, \tag{11.5}$$

where RO is expressed as kg O_2/hr and the $V_{L,OT}$ is in m^3. The smallest allowable reactor volume is given by the larger of $V_{L,FS}$ and $V_{L,OT}$.

Although floc shear has been correlated with the volumetric power input, a more fundamental parameter describing flocculation in general is the root-mean-square velocity gradient, G, and thus it is often used when examining flocculation in activated sludge systems. For diffused air systems, G (sec^{-1}) can be calculated as

$$G = \left(\frac{Q \cdot \gamma \cdot h}{60 \cdot V \cdot \mu_w} \right)^{0.5}, \tag{11.6}$$

where Q is the airflow rate in m^3/min, γ is the liquid specific weight in N/m^3, h is the liquid depth above the diffuser in m, V is the bioreactor volume in m^3, μ_w is the absolute viscosity in N·sec/m^2, and 60 is the conversion from minutes to seconds. Note the direct relationship between the volumetric air flow rate and G. For mechanical aerators, G can be calculated as

$$G = \left(\frac{1000\ P}{V \cdot \mu_w} \right)^{0.5}, \tag{11.7}$$

where P is the aerator power input in kW ([W] = N·m/sec) and V and μ_w have the same units as above. Again, note the direct relationship between the volumetric power input and G.

For diffused air activated sludge systems utilizing final clarifiers, effluent suspended solids concentrations have been correlated with G, with G values in excess of 125 sec^{-1} causing values to rise.[20] A G value of 125 sec^{-1} corresponds closely to a volumetric air input rate of around 20 $m^3/(min \cdot 1000\ m^3)$ and a volumetric power input of around 14 kW/1000 m^3, which are the minimum considered necessary to keep MLSS in suspension (i.e., Π_L). At the other extreme, the G values associated with Π_U are on the order of 270 sec^{-1}, and above that value, excessive floc destruction occurs. Between those extremes, there is a continual rise in clarifier effluent suspended solids concentration with increasing G,[20] and thus a designer can use the calculated G value to get an idea about likely effluent suspended solids concentrations. A prudent designer will anticipate having clarifier effluent suspended solids concentrations above the minimum attainable unless provisions are made for reflocculation prior to clarification. In contrast, floc shear is not an issue for diffused air systems utilizing membranes. In MBRAS the bioreactor volume is generally limited by the volumetric oxygen transfer rate. Consequently, G values are typically sufficiently high to cause significant floc shear, making the floc size in MBRAS smaller than in systems utilizing final clarifiers.[59] This is acceptable since the smaller floc in MBRAS is still easily filterable in the membrane system.

For mechanically aerated facilities, another factor that must be considered is the location of the aerator relative to the discharge to a final clarifier.[20] This is because mechanical aerators have very high localized velocity gradients. Consequently, in such systems the type and layout of the aerators has a stronger effect on clarifier effluent suspended solids concentrations than does the average G based on the overall volumetric power input.

Sheared floc can be reflocculated.[70] Thus, if the value of G exceeds 125 sec^{-1}, which will be true for most facilities, clarifier effluent quality can be improved by passing the activated sludge through a reflocculation zone prior to a final clarifier. A reflocculation time of 20 minutes at a G value of about 15 sec^{-1} may be appropriate.[70,76] The same thing can be accomplished in CAS systems by using low mixing energy in the latter stages where the oxygen requirement is low, but care must be exercised to keep all solids in suspension.

11.2.6 Nutrients

As discussed in Section 3.8.2, adequate nutrients are required to allow balanced growth of biomass in biochemical operations. Failure to provide them can have several consequences. For example, low nutrient concentrations can favor the growth of filamentous bacteria over floc formers, as discussed in Section 11.2.1, resulting in a poor settling activated sludge. More severe nutrient deficiencies can result in unbalanced growth of all bacteria, leading to the production of exocellular slime. In severe cases, the slime gives the activated sludge a jelly-like consistency, resulting in a sludge that settles slowly and compacts poorly.[36,72] Virtually no liquid-solids separation will occur in such cases.

Procedures to calculate nutrient requirements are described in Sections 5.1.6 and 10.4.1, and Tables 3.3 and 10.3 provide guidance as to the quantities needed. Experience suggests that such calculations should consider only the inorganic nitrogen and phosphorus available in the influent wastewater.[36] Organic nitrogen and phosphorus will be released into solution and become available to the biomass as organic matter is biodegraded. However, the rate of biodegradation of some of these materials can be relatively slow, making the associated nutrients unavailable to heterotrophic bacteria metabolizing readily biodegradable organic matter. Thus, limiting nutrient concentrations can occur within the process, even though the total mass of nutrients may be adequate. Consistent maintenance of residual inorganic nitrogen and phosphorus concentrations throughout the process of approximately 1 mg/L should be adequate.

11.2.7 Temperature

Temperature affects the performance of activated sludge systems as a result of its impact on the rates of biological reactions. Procedures for estimating the magnitudes of its effects are presented in Section 3.9. Two additional factors must be considered: the maximum acceptable operating temperature and the factors that affect heat loss and gain by the process.

The maximum acceptable operating temperature for typical activated sludge systems is limited to about 35 to 40°C, which corresponds to the maximum temperature for the growth of mesophilic organisms. Even short-term temperature variations above this range must be avoided since thermal inactivation of mesophilic bacteria occurs quickly. Successful operation can also be obtained if temperatures are reliably maintained above about 45 to 50°C, since a thermophilic population will develop, provided that thermophilic bacteria exist with the capability to degrade the wastewater constituents. Unacceptable performance will result for temperatures between about 40 and 45°C due to the limited number of microorganisms that can grow within this range. These considerations are particularly important for the treatment of high temperature industrial wastewaters.

One factor that affects heat gains in biological processes is the production of heat as a result of biological oxidation. As discussed in Section 2.4.1, the growth of bacteria requires that a portion of the electron donor be oxidized to provide the energy needed for biomass synthesis. Energy is also needed for cell maintenance. This oxidation and subsequent use of the energy results in the conversion of that energy into heat. Although this may seem surprising at first, it is directly analogous to the release of energy that occurs when material is burned; the only difference is the oxidation mechanism. The amount of heat released in the biooxidation of carbonaceous and nitrogenous material is directly related to the oxygen utilized by the process. For each gram of oxygen used to oxidize carbonaceous material, 3.5 kcal of energy are released.[37,39] The value for oxidation of ammonia is less well defined, but, based on thermodynamic considerations, is about 1.25 kcal of energy released for every gram of oxygen used to oxidize ammonia.[48] Since 1 kcal is sufficient energy to raise the temperature of one liter of water 1°C, the impact of this heat release depends on the wastewater strength. For example, a typical domestic wastewater requires only one gram of oxygen for each 10 liters treated, and thus the temperature rise would be only 0.35°C, a negligible amount. On the other hand, it is not unusual for an industrial wastewater to require one gram of oxygen for each liter treated, in which case the temperature rise would be 3.5°C. This could be quite significant, particularly if the wastewater itself is warm.

Other heat gains and losses occur in biological systems. Heat inputs to the system include the heat of the influent wastewater, solar inputs, and mechanical inputs from the oxygen transfer and mixing equipment. Heat outputs include conduction and convection, evaporation, and atmospheric radiation. Models for accurately calculating heat balances across suspended growth bioreactors have been developed.[6,54,61] They are discussed in Section 15.2.5.

If experience or a heat balance suggests the likelihood of unsatisfactorily high or unstable temperatures, the bioreactor should be configured to maximize heat losses. Measures to accomplish this include the use of relatively large basins to increase the HRT, shallow sidewater depths to increase basin surface area, aboveground construction to maximize conductive and convective heat losses, and the selection of an oxygen transfer device, such as mechanical surface aeration, which maximizes heat loss. Another solution is to provide mechanical cooling of the process influent or the bioreactor contents. Designs such as HPOAS or facilities using deep bioreactors with diffused aeration will have minimal heat loss and should be avoided in this situation. In fact, they may require mechanical cooling even when large heat inputs are not expected.

Although heat gain is not generally a concern during treatment of municipal wastewaters, heat loss can be, depending on the type of oxygen transfer system used and the bioreactor HRT.[10] For

example, submerged oxygen transfer systems, such as diffused aeration, have low heat losses, whereas mechanical surface aerators have high losses. This difference may influence the geographic region in which a particular type of oxygen transfer device can be used. When needed, heat loss can be minimized through proper facility design.

11.3 PROCESS DESIGN

11.3.1 Overview

The basic approach to the design of suspended growth biochemical operations is presented in Chapter 10. In this chapter we focus that approach on activated sludge systems and examine the types of decisions that are required in their design. As discussed in Chapter 10, design is an iterative procedure and can take place at several levels of sophistication, depending on the information available to the designer. At the simplest level, in which little information is available, a preliminary design can be accomplished by applying the guiding principles articulated in Table 10.1. This approach is illustrated in Section 10.4.1, and the steps involved are summarized at the end of that section.

The next level, stoichiometric-based design, uses the simple model of Chapter 5, as extended in Section 10.4.2, and requires that specific information be available about the nature of the wastewater and the parameter values describing its biodegradation. That information can sometimes be obtained from historical records at a facility that is to be expanded, or from facilities treating similar wastewaters when a new system is being designed. In this case, it will usually be necessary to convert the information from traditional measurements, such as BOD_5, into the more descriptive measurements, such as biodegradable COD, in use today. The procedures for doing this are presented in Section 9.6. In other cases, treatability studies will be required to provide the necessary information. The procedures for conducting them are presented in Sections 9.2 and 9.3. Stoichiometric-based design provides quantitative information about the mass of MLSS to be contained in the activated sludge process, the steady-state oxygen requirement, and the mass of excess solids to be disposed of daily. The equations are for a single completely mixed bioreactor, such as in CMAS, but as the Guiding Principle No. 4 in Table 10.1 states, the calculated values are applicable to any of the activated sludge variants. Thus, they can be used as the basis for decisions about bioreactor configuration and the distribution of MLSS and oxygen supply within the bioreactors. These decisions require heuristic approaches, which are presented in the material that follows. However, it should be recognized that all of the calculated values are based on the daily average flow and substrate concentration entering the facility, even though wastewater treatment facilities are subject to diurnally variable inputs, as illustrated in Figure 6.2. Thus, unless the activated sludge process is to be preceded by equalization, the impact of those variations on the design must be considered. This also requires the application of heuristically derived approaches.

As discussed in Section 10.4.3, the third level of design is simulation-based design. It is the most precise way to consider the impact of dynamic loadings on activated sludge systems and requires the use of a suitable dynamic model, such as one of the International Water Association (IWA) Activated Sludge Models (ASMs) and a computer code that implements it. Several such codes are listed in Table 6.4; all are simulation programs, not design programs. This means that the designer must choose a particular activated sludge process and provide the sizes of the component bioreactors as input to the programs. Simulations are then run to examine the performance of the process and the output is used to assess its acceptability. Should the design be unacceptable, the process must be modified and another simulation run. This procedure must be repeated in a logical manner until an acceptable design is arrived at, with the output providing needed information about the distribution of oxygen, MLSS, and so on. Thus, the designer must have already accomplished a basic process design before beginning the simulations. This can be done with either of the first two approaches.

However, characterization of the wastewater constituents and the parameters describing treatment is much more complex than that for the other design levels. The techniques for performing that characterization are described in Section 9.5.

The primary focus of this section is on stoichiometric-based design. There are several reasons for doing this. First, preliminary design based on the guiding principles is discussed in sufficient detail in Section 10.4.1 to allow its application. It need not be expanded upon here. Second, as discussed above, simulation-based design requires the designer to provide a basic process flow diagram as input to the simulation program. The most effective way of doing this is by stoichiometric-based design. Third, execution of a stoichiometric-based design requires the designer to understand the most important aspects of activated sludge design. Thus, it provides an excellent framework within which to present them.

As discussed in Section 9.6, many types of measurements have been used to express the concentrations of wastewater and activated sludge constituents. However, to allow us to focus on the decisions to be made during design and not distract the reader with multiple unit conversions, we use only biodegradable COD as the measure of organic substrate and TSS as the measure of MLSS. We have chosen the former because of its fundamental importance as a measure of available electrons, and the latter because of its widespread use in the profession. Section 9.6 provides information needed to convert between unit systems. In addition, to provide continuity in the examples, we use a standard wastewater throughout that is typical of domestic wastewater after primary treatment. It is the one in Table E9.4, and the translation between the conventional characterization in the top of the table and the more complete characterization in the bottom is explained in Example 9.6.1. Finally, we must emphasize that the calculations presented in this book are meant only to illustrate the procedures and decisions the process designer must make. The numerical results should not be considered to be typical of the application of biochemical operations to any particular real wastewater.

11.3.2 Factors to be Considered during Design

11.3.2.1 Selection of the Appropriate Process Option

The selection of a particular activated sludge process is based on many considerations, including the wastewater characteristics, effluent quality goals, facility capital and operating costs, facility operational objectives, other processes at the facility, and the desires of the owner. Consequently, a full discussion of the selection of the process option is beyond the scope of this book. Nevertheless, a few generalizations are possible. Conventional activated sludge is popular for treatment of domestic wastewater because of its proven reliability and ability to achieve high effluent quality, including complete nitrification. However, for treatment of industrial wastewaters containing inhibitory organic compounds, CMAS has advantages, although special consideration must be given to the settling properties of the resulting biomass. If a wastewater contains a high percentage of readily biodegradable organic matter and no inhibitory materials, SAS may be required to control filamentous sludge bulking. On the other hand, if the wastewater contains a high percentage of colloidal organic matter that can be removed by entrapment in the biofloc, and the readily biodegradable substrate can be removed at an SRT shorter than that associated with good bioflocculation, then CSAS and SFAS have distinct advantages relative to system volume, although full nitrification may be difficult to achieve. For small communities wishing to minimize the number of operational personnel and types of unit operations on site, EAAS is popular. When little space is available and the emission of volatile organic compounds must be minimized, situations commonly associated with industrial facilities, HPOAS is often used. Membrane bioreactor activated sludge is often used when a high quality effluent (such as for water reclamation) is needed, the plant is to be remotely located, and/or a compact facility is needed. For additional information on the selection of the process option the reader should consult design manuals, such as Manual of Practice (MOP) No. 8.[74]

11.3.2.2 Selection of the Solids Retention Time

The effects of SRT on activated sludge performance are discussed in Section 11.2.2 while the factors that must be considered during its selection are covered in Section 10.3.2. Consideration of the information in those sections makes it clear that selection of the SRT is a multifaceted decision requiring input from a number of sources. An important consideration, of course, is the SRT needed to meet the required effluent quality. If a single completely mixed bioreactor is to be used, such as in CMAS or EAAS, the SRT required to produce a particular effluent COD is given by Equation 10.7:

$$\Theta_c = \frac{K_S + S_S}{S_S(\hat{\mu}_H - b_H) - K_S b_H}. \tag{10.7}$$

As discussed in Section 10.4.2, the parameters associated with autotrophs ($\hat{\mu}_A$, K_{NH}, and b_A) can be substituted for the heterotrophic parameters ($\hat{\mu}_H$, K_S, and b_H) and the ammonia-N concentration, S_{NH}, can be substituted for the organic substrate concentration, S_S, to determine the SRT required to achieve a required ammonia-N concentration through nitrification. In either case, the SRT calculated from Equation 10.7 would not necessarily produce the required effluent quality. There are several reasons for this. One is uncertainty in the kinetic parameters, influent characteristics, natural variability in the microbial community, and other factors. Such factors cause statistical variability in the effluent quality as illustrated in Figure 10.9.[52] Thus, the calculated SRT must be multiplied by an appropriate safety factor, ς_U, to account for that uncertainty, or the value of S_S (or S_{NH}) must be chosen with the uncertainty in mind.

Because Equation 10.7 represents only steady-state performance, another factor that must be considered is the impact of loading variations on process performance. Loading variations take two forms, day-to-day variations caused by seasonal and other events, and diurnal loading variations within a day, such as those illustrated in Figure 6.2. Discussion of the factors that go into decisions about design loadings and the use of equalization to dampen them is beyond the scope of this book. However, it is important to recognize that seasonal loading variations must be considered by the designer in selecting the SRT. With regard to typical diurnal loading variations, examination of the curves labeled CSTR (continuous stirred tank reactor) in Figure 7.4 reveals that COD removal is much less subject to their effects than is nitrification. Consequently, the safety factor for uncertainty is often sufficient to guard against unsatisfactory effluent organic substrate concentrations as a result of diurnal loading variations, but not against high effluent ammonia-N concentrations. Rather, an additional safety factor is required to account for the effects of ammonia-N loading variations on nitrification. It is the peak load safety factor, ς_{PL}:

$$\varsigma_{PL} = \frac{(F \cdot S_{NHO})_{Peak}}{(F \cdot S_{NHO})_{Avg}}, \tag{11.8}$$

where F is the influent flow rate and S_{NHO} is the influent ammonia-N concentration. The choice of the period over which $(F \cdot S_{NHO})_{Peak}$ is defined (i.e., peak diurnal within the average day, maximum month, maximum week, etc.) depends on the nature of the discharge standard that must be met, the plant process flow diagram, and so on, and as such, is also beyond the scope of this book. Consequently, the reader should consult other sources, such as the U.S. EPA Nitrogen Control Manual[68] for more information.

Finally, nitrification is very sensitive to DO concentration, as discussed in Section 6.3.1 and illustrated in Figure 6.7. That sensitivity can be expressed by a double Monod expression, such as Equation 3.46, which is used in ASM No. 1 (Table 6.1). Since DO was not included in Equation 10.7, it can be included by defining another safety factor, ς_{DO}, which is the reciprocal of the DO term in

Equation 3.46 in which S_O is the DO concentration and $K_{O,A}$ is the DO half-saturation coefficient for autotrophs:

$$\varsigma_{DO} = \frac{K_{O,A} + S_O}{S_O}. \tag{11.9}$$

A similar safety factor does not need to be applied to organic substrate removal because the half-saturation coefficient for DO for heterotrophic bacteria is sufficiently small to make Equation 11.9 approach a value of 1.0 at typical bioreactor DO levels.

Considering all of these safety factors, the required value of the SRT, $\Theta_{c,r}$, can be determined from the computed SRT, Θ_c, as

$$\Theta_{c,r} = \Theta_c \cdot \varsigma_U \cdot \varsigma_{PL} \cdot \varsigma_{DO}. \tag{11.10}$$

For organic substrate removal, ς_{PL} and ς_{DO} are normally set at 1.0, as discussed above, simplifying the expression. However, the full expression is typically used for nitrification. The application of Equation 11.10 results in a very conservative design for nitrification because the safety factors are multiplicative. Consequently, the application of simulation-based design techniques, which can explicitly account for dynamic conditions, often allows reductions in the SRT required to achieve adequate nitrification performance from a CMAS system.

The application of Equation 11.10 to nitrification in a CAS system, which behaves like a plug-flow or tanks-in-series system, would result in the selection of a longer SRT than is actually necessary. This can be seen in Figure 11.8 where the effluent ammonia-N concentration is plotted as a function of SRT for activated sludge systems with various numbers of CSTRs in series. The curves in the figure were developed by simulation with ASM No. 1, using the parameter values in Table 6.3. The DO concentration was set at 4.0 mg/L so that it would not be a factor. To illustrate the impact of bioreactor configuration, consider a situation in which the desired ammonia-N concentration is 1.0 mg/L as N. An examination of the figure reveals that the calculated SRT for a CMAS system (N = 1) would be 4.4 days, whereas the SRT required for a CAS system (N = 9) would be only 2.36 days. Thus, if the SRT calculated with Equation 10.7 (which is for a CMAS system) were

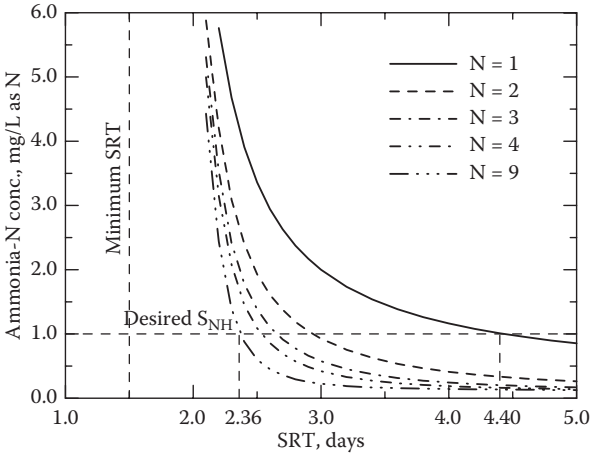

FIGURE 11.8 Effect of SRT and the hydraulic characteristics of an activated sludge bioreactor, as expressed by the equivalent number of tanks in series, N, on the effluent ammonia-N concentration as simulated with ASM No. 1. The parameter values used are listed in Table 6.3.

substituted into Equation 11.10 to determine the required SRT for a CAS system, it is clear that the CAS system would be badly overdesigned. Furthermore, the overdesign becomes worse the lower the desired ammonia-N concentration. Consequently, a different approach is used for CAS and other systems that approach plug-flow.

The approach to selection of the required SRT for nitrification in CAS systems is based on another observation from Figure 11.8; when a system behaves in a plug-flow manner, the SRT needed to achieve a low effluent ammonia-N concentration is only slightly larger than the minimum SRT. This suggests a convenient way of determining the required SRT for a CAS system. Since the SRT associated with a given effluent ammonia-N concentration from a CAS system cannot be calculated directly for substitution into Equation 11.10, designers use the minimum SRT instead, recognizing that an application of the multiple safety factors will make the required SRT sufficiently long to achieve the desired effluent quality in a reliable manner. Consequently, for nitrification in CAS and similar systems:

$$\Theta_{c,r} = \Theta_{c\min} \cdot \varsigma_U \cdot \varsigma_{PL} \cdot \varsigma_{DO}. \tag{11.11}$$

The minimum SRT for nitrification can be estimated with Figure 10.4, or it can be calculated with Equation 5.25 if the appropriate kinetic parameters are known.

$$\Theta_{c\min} = \frac{K_{NH} + S_{NHO}}{S_{NHO}(\hat{\mu}_A - b_A) - K_{NH} \cdot b_A}. \tag{5.25a}$$

The equation has been modified to include the autotrophic parameters ($\hat{\mu}_A$, K_{NH}, and b_A) and thus is renumbered as Equation 5.25a. The influent ammonia-N concentration, S_{NHO}, is the appropriate nitrogen concentration to use in Equation 5.25a because organic nitrogen is unlikely to have been metabolized by the heterotrophic bacteria and made available at short SRTs. Regardless of the technique used to determine the minimum SRT, the value should be based on the coldest wastewater temperature at which nitrification is required. The application of Equation 11.11 provides a much more realistic required SRT than Equation 11.10 for CAS and other plug-flow systems.

Example 11.3.2.1

A CAS system is to be designed to fully nitrify (i.e., produce an effluent ammonia-N concentration less than 1.0 mg/L) at a temperature of 15°C when the DO concentration in the bioreactor is 2.0 mg/L. The wastewater has been sufficiently characterized to allow the safety factor for uncertainty to be set to 1.0. The characterization also revealed that the peak to average ammonia-N loading for the wastewater is 1.75 and that the half-saturation coefficient for DO for nitrifiers is 0.75 mg/L. Determine the required SRT for the design.

Since a CAS system is to be used, Equation 11.11 is the appropriate expression with which to calculate the required SRT. The value of the minimum SRT can be obtained from Figure 10.4, which gives a value of 3.5 days. The peak load safety factor, ς_{PL}, has a value of 1.75. The safety factor for uncertainty, ς_U, has a value of 1.0. The DO safety factor can be calculated with Equation 11.9:

$$\varsigma_{DO} = \frac{0.75 + 2.0}{2.0} = 1.375.$$

Thus,

$$\Theta_{c,r} = (3.5)(1.0)(1.75)(1.375) = 8.4 \text{ days}.$$

The application of Equations 10.7 and 11.10 to a selection of the SRT for a CMAS system producing an effluent ammonia-N concentration of 1.0 mg/L as N results in a value of 17.3 days. Thus, it can be seen that the use of CAS results in a smaller SRT and thus a smaller system size.

11.3.2.3 Consideration of the Effects of Temperature

Because temperature affects the kinetic parameters describing biological reactions, as described in Section 3.9, it has important impacts on process design. First, effluent quality is worse at the coldest temperature. Thus, the parameter values used in the selection of the SRT should always reflect the lowest sustainable temperature anticipated in the bioreactor. This is particularly important when nitrification is required because of the extreme sensitivity to temperature exhibited by the maximum specific growth rate coefficient for nitrifying bacteria. Second, at SRTs encountered in practice, excess biomass production will be greatest at the lowest temperature. This has two important impacts; it implies that excess solids handling and processing systems should be designed for winter operation, and it suggests that the mass of MLSS in the system, $X_{M,T} \cdot V$, will be greatest then as well. Consequently, the value of $X_{M,T} \cdot V$ for use in bioreactor sizing should be calculated for the lowest sustainable temperature anticipated in the bioreactor. Finally, the oxygen requirement, RO, will be greatest at the highest operating temperature. Consequently, the oxygen transfer system must be designed for summer operation. With regard to this last point, it is important to recognize that the occurrence of nitrification should be checked for summer temperatures, even if it will not occur in winter. The minimum SRT for nitrification should be calculated with Equation 5.25a using temperature corrected parameter values, or it should be estimated with Figure 10.4. If the design SRT exceeds that value, a provision must be made for supplying the additional needed oxygen since it may equal that required for organic substrate removal, as seen in Section 6.3.2. On the other hand, for domestic wastewater, the nitrifiers generally will not make a significant contribution to the mass of MLSS in the system and need not be considered. However, such a generalization cannot be made about industrial wastewaters; each of them is unique.

The most commonly used temperature adjustment technique for the kinetic and stoichiometric parameters characterizing activated sludge is Equation 3.99:

$$k_1 = k_2 \cdot \theta^{(T_1 - T_2)}, \tag{3.99}$$

where k represents any parameter and θ is its temperature coefficient. Generally the reference temperature, T_2, is 20°C and that is the case with the parameters used herein. Typical values of θ are discussed in Section 3.9.2. Values selected for use in the examples of this chapter are given in Table E11.1.

TABLE E11.1
Temperature Correction Factors

Parameter	θ
$\hat{\mu}_H$	1.08
b_H and $b_{L,H}$	1.04
K_S and Y_H	1.00
$\hat{\mu}_A$	1.11
b_A and $b_{L,A}$	1.04
K_{NH}	1.14
Y_A	1.00
K_h	1.08
K_a	1.08

11.3.2.4 Consideration of the Effects of Transient Loadings

Figure 6.3 illustrates two important points about the effect of typical diurnal loadings on an activated sludge process; the oxygen requirement is influenced quite strongly, whereas there is little impact on the mass of biomass present. This means that bioreactor sizing, which is based on the value of $X_{M,T}$·V, can be done on the basis of average loads. Design of the oxygen transfer system, on the other hand, and the impact of that system on the mixing energy input, must be based on the expected peak loading.

Additional insight into the sizing of the oxygen transfer system comes from further examination of Figure 6.3. The figure shows the oxygen consumption associated only with carbon oxidation, since the maximum specific growth rate coefficient for autotrophs was set to zero during the simulation. As discussed in Section 6.2.2, the use of soluble substrate causes a more severe transient response than the use of particulate substrate because the latter must be hydrolyzed, which is a slow reaction. Nevertheless, the hydrolysis reactions are rapid enough to cause some of the particulate substrate applied during peak loading periods to be used. Thus, the transients in both readily and slowly biodegradable substrate must be considered when estimating peak oxygen requirements. Because of the complexities of the reactions involved, dynamic simulation is the only truly accurate way to assess the transient oxygen requirement in an activated sludge process. Nevertheless, it would be advantageous to have a way to approximate the peak oxygen requirement for a single tank system like CMAS.

During a short-term transient loading, biodegradable organic matter will be oxidized to synthesize new cell mass, but little additional decay will occur because the decay rate is proportional to the active biomass concentration, which changes little during the transient. Consequently, the additional oxygen requirement associated with the short-term increase in loading will be proportional to $(1 - Y_{H,T} \cdot i_{O/XB,T})$. Thus, the fractional transient increase in oxygen requirement will be less than the fractional increase in the biodegradable organic matter loading.

Simulations conducted using wastewater characteristics and parameters similar to those in Tables 6.6 and 6.3, respectively, demonstrated that transient peak oxygen requirements correspond to oxidation of all of the additional readily biodegradable organic matter applied, but only a portion of the additional slowly biodegradable organic matter applied.[2] This was true for a broad range of SRTs and load peaking factors. The load peaking factor is the peak mass loading divided by the average mass loading. Furthermore, the fraction of the additional slowly biodegradable organic matter oxidized, $f_{XS,H}$, decreased as the load peaking factor increased. Based on these considerations, the transient state oxygen requirement for the growth of heterotrophic bacteria ($RO_{H,TS}$) may be estimated as follows:

$$RO_{H,TS} = (1 - Y_{H,T} i_{O/XB,T})[\Delta(F \cdot S_{SO}) + f_{XS,H} \Delta(F \cdot X_{SO})], \tag{11.12}$$

where $\Delta(F \cdot S_{S0})$ is the transient increase in the loading of readily biodegradable organic matter above the average loading, and $\Delta(F \cdot X_{S0})$ is the transient increase in the loading of slowly biodegradable organic matter above the average. The value of $f_{XS,H}$ will generally range from 0.5 to 1.0, with smaller values being associated with larger transient loading increases.[2]

The peak oxygen requirement due to heterotrophic activity is the sum of the steady-state oxygen requirement, as given by Equation 10.10, and the transient-state oxygen requirement as given by Equation 11.12. The oxygen transfer rate to the system must be capable of meeting both requirements, in addition to any oxygen utilization by the autotrophic bacteria.

The transient increase in the oxygen requirement due to nitrification is more complicated for a number of reasons. The first is that SRT has a much stronger effect than it does on the heterotrophic oxygen requirement.[2] At SRTs that are above the minimum SRT for the autotrophic bacteria but below that required for full nitrification at steady state, transient loadings will have no effect on the rate of nitrification, and hence on the oxygen consumption associated with it, because

nitrification will already be occurring at close to its maximum rate. At long SRTs where full nitrification can occur even during the transient, the increase in oxygen consumption rate will be proportional to the increase in loading, just as it is for heterotrophic bacteria, although the proportion oxidized will be different. At SRTs that are just sufficient to give full nitrification at steady state, the ammonia nitrogen concentration may rise sufficiently during the transient to allow the rate of nitrification to reach its maximum value, thus causing the oxygen consumption rate to rise, but by a smaller amount than the loading increase. A second complicating factor is that not all nitrogen in the influent is in a form that is available to the autotrophic bacteria. Some will be in the form of biodegradable organic nitrogen. This nitrogen will become available only as the organic matter containing it undergoes decomposition. Based on the arguments in the preceding paragraph, we would expect all of the nitrogen associated with the readily biodegradable substrate to be made available as ammonia-N, but only the fraction $f_{XS,H}$ of that associated with the slowly biodegradable substrate. Finally, some of the ammonia-N entering during the transient will be incorporated into the extra biomass formed during the transient as the additional organic matter is removed. This, too, must be accounted for.

For the situation in which the SRT is sufficiently long to allow full nitrification during the peak loading period, the transient state oxygen requirement associated with the autotrophic bacteria, $RO_{A,TS}$ can be calculated with an equation analogous to Equation 11.12:

$$RO_{A,TS} = (4.57 - Y_{A,T} i_{O/XB,T})[\Delta(F \cdot S_N)_{a,TS}], \tag{11.13}$$

in which $\Delta(F \cdot S_N)_{a,TS}$ is the transient increase in ammonia-N available to the autotrophic bacteria. It is given by

$$\begin{aligned}\Delta(F \cdot S_N)_{a,TS} = {} & \Delta(F \cdot S_{NHO}) + \Delta(F \cdot S_{NSO}) + f_{XS,H}\Delta(F \cdot X_{NSO}) \\ & -0.087 Y_{H,T} i_{O/XB,T}[\Delta(F \cdot S_{SO}) + f_{XS,H}\Delta(F \cdot X_{SO})],\end{aligned} \tag{11.14}$$

where $\Delta(F \cdot S_{NHO})$, $\Delta(F \cdot S_{NSO})$, and $\Delta(F \cdot X_{NSO})$ are the transient increases in the loadings of ammonia-N, soluble biodegradable organic nitrogen, and particulate biodegradable organic nitrogen above the average. The negative term in Equation 11.14 accounts for the additional use of nitrogen associated with synthesis of the heterotrophic bacteria during the transient organic loading. The result from Equation 11.14 must be added to the steady-state autotrophic oxygen requirement to determine the peak autotrophic requirement for this situation.

For the situation in which nitrification is not complete during the transient, causing the ammonia-N concentration to rise high enough to allow the autotrophic bacteria to grow at their maximal rate, the maximum autotrophic oxygen utilization rate, $RO_{A,max}$, can be calculated from a modified form of Equation 5.42:

$$\begin{aligned}RO_{A,max} = {} & \left[\left(\frac{4.57 - Y_{A,T} i_{O/XB,T}}{Y_{A,T} i_{O/XB,T}}\right)\hat{\mu}_A + (1 - f_D) b_A\right] \cdot \\ & (X_{B,A,T} \cdot V) i_{O/XB,T}\left(\frac{S_O}{K_{A,O} + S_O}\right).\end{aligned} \tag{11.15}$$

The mass of autotrophic bacteria in the system, $X_{B,A,T} \cdot V$, should be that associated with the average loading on the system. The last term is included because of the sensitivity of the autotrophic nitrifying bacteria to the DO concentration and the likelihood of that concentration falling during the transient. Consequently, the DO concentration used should be the concentration expected during the

transient. In situations where this condition occurs, the peak autotrophic oxygen requirement will just be $RO_{A,max}$, because that value cannot be exceeded. Consequently, the determination of which situation controls is made by comparing the two potential peak requirements; the smaller of the two controls.

There will be circumstances, particularly during preliminary design, where insufficient information is available to allow the procedures above to be used. In that case it may be satisfactory to multiply the heterotrophic and autotrophic steady-state oxygen requirements by an oxygen peaking factor to arrive at the transient state oxygen requirements. Figure 11.9 provides oxygen peaking factors as a function of the load peaking factor. It was generated from simulations conducted using wastewater characteristics and parameters similar to those used in Tables 6.6 and 6.3, respectively.[2] Because the oxygen peaking factor for the removal of organic matter was not influenced strongly by SRT, the curve for carbon oxidation should be safe for a broad range of SRTs. The curve for nitrification, on the other hand, is only valid for SRTs above 10 days.

A final point about transient loadings concerns their impact on mixing energy input and floc shear. It will be recalled from Section 11.2.5 that there is both a lower and an upper limit on the volumetric power input to an activated sludge bioreactor, with the lower limit being the minimum energy required to keep the MLSS in suspension, and the upper limit being set to prevent floc shear. The ratio of the upper to the lower limit is around 4.5. It is not unusual, however, for the ratio of the maximum loading to the minimum loading within a day at a wastewater treatment plant to be greater than 4.5, particularly for small plants.[2] This suggests that the ratio of the maximum to minimum oxygen requirements associated with diurnal loadings can be greater than 4.5. In that situation, since the volumetric power input required for oxygen transfer is directly proportional to the oxygen requirement, it would be impossible to meet both the upper and lower limits on power input. Consequently, most designers use the upper limit and the peak oxygen requirement during sizing of the bioreactor and then limit the turn down on the aeration system to meet the lower limit during low loading, recognizing that the DO concentration in the bioreactor will be higher than needed then. The other alternative is to include flow equalization in the process flow diagram.

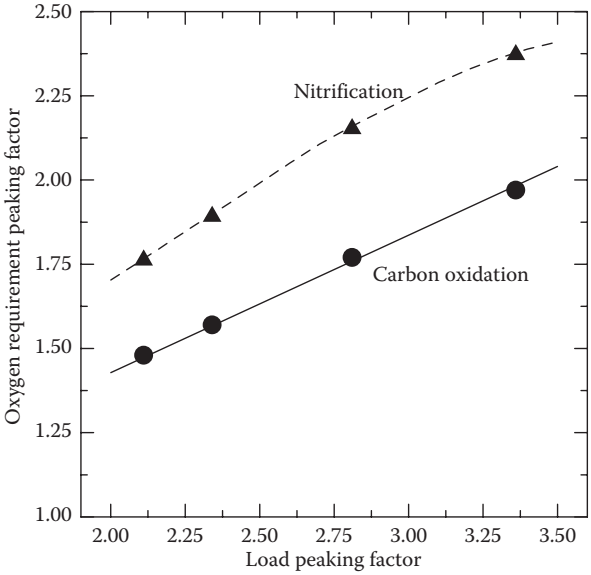

FIGURE 11.9 Effect of the load peaking factor on the oxygen peaking factor for a CMAS system receiving a diurnally varying input. (Data from Amalan, S., Analysis of Factors Affecting Peaking Phenomena in Activated Sludge Oxygen Requirements Due to Diurnal Load Variations, MS Thesis, Clemson University, South Carolina, 1992.)

11.3.2.5 Distribution of Volume, Mixed Liquor Suspended Solids, and Oxygen in Nonuniform Systems

As indicated by the Guiding Principle No. 4 in Table 10.1, the total mass of biomass and the total oxygen requirement in the various alternative activated sludge systems will all be essentially the same, provided they all have the same SRT. Thus, they can be calculated by the simple model of Chapter 5 as modified in Section 10.4.2. However, when the design involves the distribution of flows or volumes into reactors in series, the biomass and oxygen requirement must also be distributed appropriately. This can be done for the steady-state case by the application of mass balances and appropriate heuristics. Distribution of the transient-state oxygen requirement is more difficult. Because the procedures involved are unique to each activated sludge variation, they will be considered individually in the sections that follow.

Table 11.3 summarizes the general steps required to complete the process design of an activated sludge system. The following examples illustrate specific aspects of the relevant calculations.

TABLE 11.3
Summary of Activated Sludge Process Design Procedure

1. Summarize process design and loading conditions including maximum, minimum, and average sustained temperature; maximum, minimum, average, and design process influent flows and pollutant loadings; and desired effluent quality.
2. Where necessary, convert process loadings, expressed in conventional parameters (such as BOD_5) into the units used in the process design (such as biodegradable COD).
3. Select the process configuration.
4. Select the design SRT.
5. Calculate the steady-state oxygen requirement for maximum, minimum, and average sustained temperature conditions based on the CMAS configuration.
6. Calculate the diurnal maximum and minimum oxygen requirements for the conditions above.
7. Determine the range of allowable bioreactor volumes based on the CMAS model. The minimum volume can be limited either by the maximum achievable volumetric oxygen transfer rate or by floc shear. The maximum bioreactor volume will be limited by mixing. In some cases it may be necessary to compromise as not every condition can be accommodated.
8. Using the maximum and minimum volumes determined in Step 7, calculate the maximum and minimum MLSS concentrations using the CMAS model.
9. Considering cost trade-offs between the bioreactor and the final clarifier (or membrane system), choose the MLSS concentration within the allowable range and calculate the associated bioreactor volume.
10. Calculate the waste sludge production rate using the CMAS model.
11. For processes with MLSS concentrations that vary through the bioreactor, such as CSAS and SFAS, calculate the MLSS distribution.
12. For processes with variations in oxygen requirements, calculate the distribution of oxygen requirements through the bioreactor.
13. Based on the above, make any necessary adjustments in the process design and summarize the results in tabular form.

11.3.3 Design of a Completely Mixed Activated Sludge System—The General Case

Because it is the simplest, the basic design process will be outlined for a CMAS system. All examples will be developed for the wastewater characteristics given in Table E9.4 and the kinetic and stoichiometric parameters given in Table E11.2. The values for a temperature of 20°C are the same as those in Table 6.3 after conversion to COD/TSS units. As indicated earlier, all organic substrate concentrations will be expressed as biodegradable COD and all MLSS concentrations will be expressed as TSS. Two situations will be considered. First, to illustrate basic principles, the case of

full equalization (i.e., the steady-state case) will be considered. Then we will consider the impacts of diurnal variations in loading (i.e., the case without equalization).

11.3.3.1 Basic Process Design for the Steady-State Case

The first task in a process design is to establish the maximum and minimum sustained temperatures likely to be encountered in the activated sludge system. The stoichiometric and kinetic parameters are then adjusted to those temperatures using Equation 3.99, as discussed in Section 11.3.2. The temperature adjusted parameters are used in selection of the design SRT. Because we have already discussed the selection of the SRT, it will not be considered further here. Rather, we will assume that the decision has already been made.

Example 11.3.3.1

A CMAS system is to be designed to remove organic matter from a wastewater with the characteristics given in Table E9.4. Removal of ammonia-N is not required, so the system does not have to nitrify. Consequently, an SRT of three days has been chosen for the design. The average design wastewater flow rate is 40,000 m^3/day and full equalization will maintain the loading at the average value throughout the day. The oxygen transfer system will be sized to maintain the DO concentration above 1.5 mg/L under all conditions. The parameter values characterizing the wastewater at 20°C are given in Table E11.2. However, the lowest sustained temperature anticipated is 15°C and the highest is 25°C. Prepare a table of temperature adjusted parameter values by using the temperature coefficients in Table E11.1.

All temperature adjustments are made with Equation 3.99, in which k_2 is the value of the parameter at reference temperature T_2. Using b_H as an example,

$$b_{H,15} = (0.18\ \text{day}^{-1})\ 1.04^{(15-20)} = 0.15\ \text{day}^{-1}$$

$$b_{H,25} = (0.18\ \text{day}^{-1})\ 1.04^{(25-20)} = 0.22\ \text{day}^{-1}.$$

The values of the other parameters are given in Table E11.2.

The next step in the process design is to calculate the oxygen requirement for the system. As mentioned in Section 11.3.2, that should be done for the highest expected sustained temperature because that is when the highest oxygen requirement will occur. The information has two uses.

TABLE E11.2
Stoichiometric and Kinetic Parameter Values from Table 6.3 after Conversion to COD/TSS Units and Adjustment for Temperature

Parameter	Units	Value at 20°C	Value at 15°C	Value at 25°C
$\hat{\mu}_H$	day^{-1}	6.0	4.1	8.8
K_S	mg/L as COD	20	20	20
$Y_{H,T}$	mg TSS/mg COD	0.50	0.50	0.50
b_H	day^{-1}	0.18	0.15	0.22
f_D	mg TSS/mg TSS	0.20	0.20	0.20
$i_{O/XB,T}$	mg COD/mg TSS	1.2	1.2	1.2
$\hat{\mu}_A$	day^{-1}	0.77	0.46	1.3
K_{NH}	mg/L as N	1.0	0.52	1.9
$K_{O,A}$	mg/L as O_2	0.75	0.75	0.75
$Y_{A,T}$	mg TSS/mg N	0.20	0.20	0.20
b_A	day^{-1}	0.10	0.08	0.12

First, it provides the base requirement for design of the oxygen transfer system. That aspect of design will not be covered in this book, so the reader is referred to other sources for it.[41,67,74,79] Second, the maximum oxygen requirement will be used with the upper limit on the volumetric power input, Π_U, to select the lower feasible reactor volume based on floc shear, $V_{L,FS}$, as given by Equation 11.4. The maximum oxygen requirement will also be used in Equation 11.5 to calculate the lower limit on bioreactor volume based on oxygen transfer, $V_{L,OT}$. The minimum oxygen requirement, which will occur at the lowest sustained operating temperature, must also be calculated. It will be used with the lower limit on the volumetric power input, Π_L, to select the upper feasible bioreactor volume, V_U, as given by Equation 11.3. Those volume limits will then be used to make the final selection of the bioreactor volume and the associated MLSS concentration.

The oxygen requirement for removal of organic matter by the heterotrophs, RO_H, can be calculated with Equation 10.10:

$$RO_H = F(S_{SO} + X_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T}i_{O/XB,T}}{1 + b_H \cdot \Theta_c}\right]. \tag{10.10}$$

If nitrification will occur, then the oxygen requirement associated with it can be calculated with a slightly modified version of Equation 6.2, shown below as Equation 11.16:

$$RO_A = F(S_{N,a} - S_{NH})\left[4.57 - \frac{(1 + f_D \cdot b_A \cdot \Theta_c)Y_{A,T}i_{O/XB,T}}{1 + b_A \cdot \Theta_c}\right]. \tag{11.16}$$

In this expression the influent nitrogen concentration is designated as $S_{N,a}$, representing the nitrogen available to the nitrifiers, rather than the influent ammonia-N concentration as given in Equation 6.2. There are two reasons for this. First, at SRTs long enough to allow nitrification, essentially all biodegradable COD will be used, releasing all organic nitrogen as ammonia-N. Thus, $S_{N,a}$ must include all biodegradable organic nitrogen in addition to the ammonia-N. Second, the heterotrophic bacteria will use some of the nitrogen in the synthesis of their biomass, making it unavailable to the nitrifiers. Thus, the available nitrogen concentration is given by

$$S_{N,a} = S_{NHO} + S_{NSO} + X_{NSO} - NR(S_{SO} + X_{SO} - S_S), \tag{11.17}$$

where NR is the heterotrophic nitrogen requirement given by Equation 5.46 after substitution of Equation 5.38 for the observed yield, $Y_{Hobs,T}$:

$$NR = 0.087\left[\frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T}i_{O/XB,T}}{1 + b_H \cdot \Theta_c}\right]. \tag{11.18}$$

The effluent concentration of ammonia-N can be calculated with Equation 5.22 by substituting the kinetic parameters for the autotrophic bacteria in place of the heterotrophic parameters. Because nitrification is so sensitive to the DO concentration, however, it would be wise to consider its effect on the effluent ammonia-N concentration. This can be done by using the double Monod equation, Equation 3.46, in place of the Monod equation when deriving Equation 5.22. The resulting equation can be simplified, however, by recognizing that the term $S_O/(K_{O,A} + S_O)$ is just the reciprocal of the DO safety factor, as given by Equation 11.9. Substituting for it gives:

$$S_{NH} = \frac{K_{NH}(1/\Theta_c + b_A)}{(\hat{\mu}_A/\varsigma_{DO}) - (1/\Theta_c + b_A)}. \tag{11.19}$$

If nitrification is not a design objective during selection of the SRT, then the minimum SRT for nitrification should be checked to determine whether it is likely to occur, because if it does, it will have a large impact on the oxygen requirement. This can be done either with Figure 10.4 or with Equation 5.25a. The effect of DO concentration on the SRT required for nitrification can be considered by multiplying the minimum SRT by the DO safety factor, ς_{DO}, as was done in Equation 11.11. However, neither of the other two safety factors should be used in this application.

The following example illustrates the technique for determining the oxygen requirement and the upper and lower limits on the bioreactor volume.

Example 11.3.3.2

Continue with the CMAS design started in Example 11.3.3.1 and determine the maximum and minimum steady-state oxygen requirements, and the lower and upper limits on the feasible bioreactor volumes. The design SRT is three days. Assume that bubble aeration will be used, with oxygen transfer efficiency, η_Q, of 10%.

a. What is the maximum steady-state heterotrophic oxygen requirement?
This can be calculated with Equation 10.10 using the parameter values in Table E11.2 for 25°C, since the oxygen requirement is maximum at the warmest temperature. We saw earlier that almost all readily biodegradable COD will be removed at SRTs in excess of one day, so the value of S_S can safely be assumed to be negligible. As given in Example 11.3.3.1, the flow rate is 40,000 m³/day. Thus, for consistency in units, concentrations should be expressed with m³ as the measure of volume. From Table E9.4, $S_{SO} = 115$ mg/L = 115 g/m³ and $X_{SO} = 150$ mg/L = 150 g/m³. Furthermore, all time dependent coefficients in Table E11.2 are expressed with days as the unit of time, so the SRT should be expressed in days for consistency:

$$RO_H = (40{,}000)(115+150)\left\{1-\frac{[1+(0.2)(0.22)(3.0)](0.50)(1.20)}{1+(0.22)(3.0)}\right\}$$

$$= 6{,}260{,}000 \text{ g } O_2/\text{day} = 6260 \text{ kg } O_2/\text{day}.$$

b. Will the system nitrify in the summer when the temperature is 25°C?
This can be determined either with Figure 10.4 or with Equation 5.25a. Because the needed parameter values are available, we will use Equation 5.25a with a value of S_{NHO} of 25 mg/L from Table E9.4. Again, the values of the parameters are for 25°C from Table E11.2:

$$\Theta_{cmin} = \frac{(1.9+25)}{(25)(1.30-0.12)-(1.9)(0.12)} = 0.92 \text{ day}.$$

This value is for high DO concentrations. Because the lowest expected DO concentration is 1.5 mg/L, the DO safety factor, as given in Equation 11.9 should be applied, using the value of $K_{O,A}$ from Table E11.2:

$$\varsigma_{DO} = \frac{(0.75+1.5)}{1.5} = 1.50.$$

Therefore, the minimum SRT is

$$\Theta_{cmin} = (0.92)(1.5) = 1.38 \text{ days}.$$

Since the design SRT is three days, nitrification will occur in the summer.

c. What is the concentration of nitrogen available to the autotrophic bacteria for nitrification?
This can be calculated with Equation 11.17. From Table E9.4, $S_{NHO} = 25$ mg/L as N = 25 g/m^3, $S_{NSO} = 6.5$ mg/L as N = 6.5 g/m^3, $X_{NSO} = 8.5$ mg/L as N = 8.5 g/m^3, $S_{SO} = 115$ mg/L as COD = 115 g/m^3, and $X_{SO} = 150$ mg/L as COD = 150 g/m^3. As in part a above, S_S may be neglected. All that is needed is the value of NR, the nitrogen requirement of the heterotrophs. This can be calculated with Equation 11.18 using parameter values from Table E11.2 for 25°C:

$$NR = 0.087\left\{\frac{[1+(0.20)(0.22)(3.0)](0.5)(1.20)}{1+(0.22)(3)}\right\}$$

$$= 0.036 \text{ mg N used/mg COD removed.}$$

Substituting this into Equation 11.17 gives:

$$S_{N,a} = 25 + 6.5 + 8.5 - 0.036(115 + 150) = 30.5 \text{ g/m}^3 \text{ as N.}$$

d. What is the steady-state autotrophic oxygen requirement?
This can be calculated with Equation 11.16, which requires knowledge of the effluent ammonia-N concentration. Since nitrification is not likely to be complete at an SRT of three days, the effluent ammonia-N concentration must be calculated with Equation 11.19. Using the parameter values from Table E11.2 for 25°C and the value of the DO safety factor calculated in part b above

$$S_{NH} = \frac{1.9(1/3.0+0.12)}{(1.30/1.5)-(1/3.0+0.12)} = 2.1 \text{ mg/L as N} = 2.1 \text{ g/m}^3 \text{ as N.}$$

Substitution of this value into Equation 11.16 gives the autotrophic oxygen requirement:

$$RO_A = (40{,}000)(30.5-2.1)\left\{4.57-\frac{[1+(0.2)(0.12)(3.0)](0.20)(1.20)}{1+(0.12)(3.0)}\right\}$$

$$= 4{,}980{,}000 \text{ g } O_2\text{/day} = 4980 \text{ kg } O_2\text{/day.}$$

e. What is the maximum steady-state oxygen requirement?
The maximum steady-state oxygen requirement is the sum of the heterotrophic and autotrophic oxygen requirements:

$$RO = 6260 + 4980 = 11{,}240 \text{ kg } O_2\text{/day} = 468 \text{ kg/hr.}$$

The oxygen transfer system must be designed to transfer this amount, plus an appropriate factor of safety. Note that the autotrophic oxygen requirement is almost as much as the heterotrophic requirement. This points out why it is so important to determine whether it is likely that nitrification will occur, even when the system is not being designed with nitrification as an objective.

f. What is the lower limit on the bioreactor volume based on the mixing energy constraint to avoid floc shear?
This lower limit on the bioreactor volume can be calculated with Equation 11.4, after estimation of the required airflow rate, Q, with Equation 11.2. The oxygen requirement in

Equation 11.2 has units of kg/hr, giving Q in m^3/min. The same units are used for Q in Equation 11.4:

$$Q = \frac{(6.0)(468)}{10} = 281\ m^3/min.$$

The lower limit on the CMAS bioreactor volume based on floc shear comes from application of Equation 11.4 using 90 $m^3/(min \cdot 1000\ m^3)$ as an appropriate value for $\Pi_{U,Q}$:

$$V_{L,FS} = \frac{(1000)(281)}{90} = 3{,}120\ m^3.$$

g. What is the lower limit on the bioreactor volume based on the maximum sustainable volumetric oxygen transfer rate?
This lower limit on the bioreactor volume can be calculated with Equation 11.5 using the total oxygen requirement of 468 kg/hr:

$$V_{L,OT} = \frac{468}{0.10} = 4{,}680\ m^3.$$

This value is larger than the volume associated with the constraint on floc shear, and thus it controls.

h. What is the minimum steady-state heterotrophic oxygen requirement?
This must be calculated for winter conditions when the temperature is 15°C. The procedure is exactly the same as in part a above, except that the parameter values for 15°C from Table E11.2 are used. The result of that computation is

$$RO_H = 5{,}820{,}000\ g/day = 5820\ kg/day.$$

i. Will the system nitrify in the winter when the temperature is 15°C and the DO concentration is 1.5 mg/L?
Using the same procedure as in part b above, but with the parameter values for 15°C from Table E11.2, the minimum SRT for nitrification is found to be 4.0 days. Thus, nitrification will not occur in the winter.

j. What is the minimum steady-state oxygen requirement?
Since nitrification will not occur, the minimum steady-state oxygen requirement is just the minimum heterotrophic oxygen requirement:

$$RO = 5820\ kg/day = 242.5\ kg/hr.$$

This is only slightly more than half of the summer requirement. Since this is all that will be required during winter operation, the oxygen transfer system must be designed with sufficient turn down capacity to allow this amount to be delivered in an economic manner.

k. What is the upper limit on the bioreactor volume based on the mixing energy constraint to keep all biomass in suspension?
The upper limit on the bioreactor volume can be calculated with Equation 11.3, after estimation of the required airflow rate, Q, with Equation 11.2. The oxygen requirement in Equation 11.2 has units of kg/hr, giving Q in m^3/min. The same units are used for Q in Equation 11.3:

$$Q = \frac{(6.0)(242.5)}{10} = 146\ m^3/min.$$

The upper limit on the CMAS bioreactor volume comes from application of Equation 11.4 using 20 $m^3/(min \cdot 1000\ m^3)$ as an appropriate value for $\Pi_{L,Q}$:

$$V_U = \frac{(1000)(146)}{20} = 7300\ m^3.$$

l. Any bioreactor volume between 4680 and 7300 m^3 can be used as long as it results in an MLSS concentration that is acceptable.

After the range of feasible bioreactor volumes has been calculated, the next step in the design of a CMAS system is to determine the MLSS concentration associated with each extreme reactor volume. This will provide a range of feasible MLSS concentrations from which a design value can be chosen after consideration of the size of the final settler or membrane system, as discussed in Section 10.3.4. Use Equation 10.8 to calculate the mass of MLSS, $X_{M,T} \cdot V$:

$$X_{M,T} \cdot V = \Theta_c F\left[X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c) Y_{H,T} (S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \Theta_c} \right]. \tag{10.8}$$

This computation is made for the coldest anticipated sustained wastewater temperature because the decay coefficient will be smallest then, resulting in the highest quantity of biomass, as discussed in Section 11.3.2. The value of the effluent soluble biodegradable COD, S_S, is generally small enough to be neglected in the computation, even for cold conditions. The equation does not include the contribution of autotrophic biomass to the mass of MLSS in the bioreactor. This is because for domestic wastewater, their contribution will be negligible, as discussed in Section 6.3.2. This may not be true for wastewaters containing a high nitrogen content, and in those circumstances another term should be added to account for the contribution of the nitrifying bacteria. It is similar to the right term within the brackets, except that the parameters and the influent concentrations would represent nitrification:

$$X_{B,A,T} \cdot V = \Theta_c F\left[\frac{(1 + f_D \cdot b_A \cdot \Theta_c) Y_{A,T} (S_{N,a} - S_{NH})}{1 + b_A \cdot \Theta_c} \right]. \tag{11.20}$$

Example 11.3.3.3

Continuing with the CMAS design begun in Example 11.3.3.1, determine the range of feasible MLSS concentrations.

a. What value of $X_{IO,T}$ should be used in the calculation?
The $X_{IO,T}$ includes both the fixed suspended solids (FSS) and the nonbiodegradable VSS. In Section 9.6 it was stated that 35 to 40% of the particulate organic matter in domestic wastewater is nonbiodegradable. As a result, a nonbiodegradable fraction of 0.375 was used in Example 9.6.1, where Table E9.4 was developed. From that table, the VSS concentration was 61.5 mg/L. This suggests that the nonbiodegradable VSS concentration is 0.375×61.5 or 23 mg/L. The FSS concentration is the difference between the TSS and VSS concentration, or 82 – 61.5 = 20.5 mg/L. Therefore, the value of $X_{IO,T}$ is

$$X_{IO,T} = 23 + 20.5 = 43.5\ mg/L = 43.5\ g/m^3.$$

b. What is the mass of MLSS present in the CMAS system at 15°C?
This can be calculated with Equation 10.8. As given in Example 11.3.3.1, the flow rate is 40,000 m^3/day. Thus, for consistency in units, concentrations should be expressed with m^3 as the measure of volume. From Table E9.4, $S_{SO} = 115$ mg/L = 115 g/m^3 and $X_{SO} = 150$ mg/L = 150 g/m^3. Furthermore, all time dependent coefficients in Table E11.2 are expressed with days as the unit of time, so the SRT should be expressed in days for consistency. The mass of MLSS is calculated for winter conditions using the parameters from Table E11.2:

$$X_{M,T} \cdot V = (3.0)(40{,}000)\left\{43.5 + \frac{[1+(0.20)(0.15)(3.0)](0.5)(115+150)}{1+(0.15)(3.0)}\right\}$$

$$= 17{,}170{,}000 \text{ g MLSS.}$$

c. What are the upper and lower limits on the MLSS concentration?
The highest feasible MLSS concentration, $X_{M,T,U}$, is associated with the smallest feasible bioreactor volume, 4680 m^3:

$$X_{M,T,U} = \frac{17{,}170{,}000}{4680} = 3670 \text{ g/m}^3 = 3670 \text{ mg/L.}$$

The smallest feasible MLSS concentration, $X_{M,T,L}$, is associated with the largest feasible bioreactor volume, 7300 m^3:

$$X_{M,T,L} = \frac{17{,}170{,}000}{7300} = 2350 \text{ g/m}^3 = 2350 \text{ mg/L.}$$

d. The choice of MLSS concentration between those limits must be made by considering the cost trade-offs between the bioreactor and the final settler or membrane system, as discussed in Section 10.3.4. It is important to note that the constraints on mixing energy input and oxygen transfer limit the range of values that need to be considered.

The final step in the design of the activated sludge process is the calculation of the solids wastage rate, $W_{M,T}$. It is very straightforward and is simply an extension of the procedure used to calculate $X_{M,T} \cdot V$. For the stoichiometric approach with the simplified model of Chapter 5, the appropriate equation is Equation 10.9:

$$W_{M,T} = F\left[X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c) Y_{H,T} (S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right]. \tag{10.9}$$

This computation is generally made for winter conditions since that is when the most solids must be disposed of. As with the computation of $X_{M,T} \cdot V$, no term is included for the contribution of the autotrophic bacteria. If their contribution to the waste solids is likely to be significant, then an appropriate term should be added to the equation. The actual flow rate of waste solids will depend on whether solids are wasted from the settler underflow or directly from the bioreactor. Both the flow rate and the daily mass of waste solids are used to size the solids handling system.

Example 11.3.3.4

Continue the design begun in Example 11.3.3.1 by calculating the solids wastage rate from the process. This is done for the low temperature condition using Equation 10.9:

$$W_{M,T} = (40{,}000)\left\{43.5 + \frac{[1+(0.20)(0.15)(3.0)](0.5)(115+150)}{1+(0.15)(3.0)}\right\}$$

$$= 5{,}720{,}000 \text{ g TSS/day} = 5720 \text{ kg TSS/day}.$$

11.3.3.2 Consideration of the Effects of Transient Loadings

The basic design given above considered a system that used equalization to dampen transient loadings. Quite frequently, however, facilities are designed without equalization, and thus consideration must be given to the impacts of typical diurnal loadings or other transients on the system. As discussed in Section 11.3.2, diurnal loadings have little impact on the mass of MLSS in the system because of the dampening effect of the SRT relative to the HRT. Consequently, the main consideration during design is on the oxygen requirement and the feasible bioreactor volumes while meeting the mixing energy and oxygen transfer constraints. Generally, the peak loading is most important because of its impact on the size of the oxygen transfer system and the potential for floc shear. Little consideration is given to the minimum daily loading because of its short duration. If aeration rates are turned down to the lower limit on mixing energy during that period and that provides more oxygen than is needed, the penalty in power costs will be small.

The basic procedure for calculating the additional oxygen requirement as a result of a transient load was discussed in Section 11.3.2, where Equations 11.12 through 11.15 were presented. The peak oxygen requirement for the process, RO_P, is then the sum of the average and transient-state oxygen requirements for both the heterotrophic and the autotrophic biomass. However, as discussed in Section 11.3.2, consideration must be given to whether the autotrophic bacteria have reached their maximum possible growth and oxygen consumption rates. Thus, the peak oxygen requirement is given by the smaller of the two expressions:

$$RO_P = RO_H + RO_{H,TS} + RO_A + RO_{A,TS} \tag{11.21}$$

or

$$RO_P = RO_H + RO_{H,TS} + RO_{A,max}. \tag{11.22}$$

In the example that follows, we will examine the impact of transient loads on the design of the CMAS system considered in Examples 11.3.3.1 through 11.3.3.4.

Example 11.3.3.5

Continue with the design of the CMAS system begun in Example 11.3.3.1, which has an average daily flow rate of 40,000 m^3/day. In this case, however, no equalization will be employed so that the system routinely experiences a peak loading 2.5 times the average daily loading. What will the peak oxygen requirement be? What oxygen peaking factor does the peak oxygen requirement represent?

a. What is the peak transient oxygen requirement for the heterotrophic bacteria?
   This is calculated with Equation 11.12 in which $\Delta(F \cdot S_{SO})$ and $\Delta(F \cdot X_{SO})$ represent the transient increases in the readily and slowly biodegradable organic matter, respectively. Since the peaking factor is 2.5, which is the ratio of the peak to the average loading, the transient

increase of the biodegradable materials above the average loading is 1.5 times that average. Thus:

$$\Delta(F \cdot S_{SO}) = 1.5(40{,}000)(115) = 6{,}900{,}000 \text{ g COD/day}$$
$$= 6900 \text{ kg COD/day}$$
$$\Delta(F \cdot X_{SO}) = 1.5(40{,}000)(150) = 9{,}000{,}000 \text{ g COD/day}$$
$$= 9000 \text{ kg COD/day}.$$

A value must be assumed for $f_{XS,H}$, the fraction of additional slowly biodegradable substrate oxidized during the transient. We stated earlier that values typically lie between 0.5 and 1.0, with smaller values being associated with larger transient increases. Since the magnitude of the transient load is neither extremely high nor low, we will assume that $f_{XS,H}$ has a value of 0.75. The peak oxygen requirement will occur at the warmest temperature. Consequently, substituting the parameter values for 25°C from Table E11.2 into Equation 11.12, along with the transient increases calculated above, gives:

$$RO_{H,TS} = [1-(1.20)(0.5)][6900+(0.75)(9000)] = 5460 \text{ kg } O_2/\text{day}.$$

b. What is the peak transient oxygen requirement for the autotrophic bacteria assuming that they are able to oxidize the additional ammonia load?
We use the same procedure as above to calculate the transient nitrogen loadings:

$$\Delta(F \cdot S_{NHO}) = 1.5(40{,}000)(25) = 1{,}500{,}000 \text{ g N/day} = 1{,}500 \text{ kg N/day}$$
$$\Delta(F \cdot S_{NSO}) = 1.5(40{,}000)(6.5) = 390{,}000 \text{ g N/day} = 390 \text{ kg N/day}.$$
$$\Delta(F \cdot X_{NSO}) = 1.5(40{,}000)(8.5) = 510{,}000 \text{ g N/day} = 510 \text{ kg N/day}$$

Use Equation 11.14 and $f_{XS,H} = 0.75$ to calculate the transient increase in the ammonia-N concentration available to the autotrophic bacteria:

$$\Delta(F \cdot S_N)_{a,TS} = 1500 + 390 + (0.75)(510) - (0.087)(120)(0.50)$$
$$\cdot[6900+(0.75)(9000)] = 1560 \text{ kg N/day}.$$

Using Equation 11.13 and the appropriate parameters for 25°C from Table E11.2:

$$RO_{A,TS} = [4.57-(1.2)(0.20)](1560) = 6750 \text{ kg } O_2/\text{day}.$$

c. What is the maximum potential autotrophic oxygen requirement assuming that the autotrophic bacteria are not able to oxidize all of the transient nitrogen input, causing the ammonia-N concentration to rise sufficiently for them to reach their maximal growth rate? As in the other examples, assume that a DO concentration of at least 1.5 mg/L is maintained.

This is calculated using Equation 11.15 with parameter values for the maximum temperature of 25°C. The use of this equation requires knowledge of the mass of autotrophic bacteria in the system based on average loading conditions, $X_{B,A,T} \cdot V$. This can be calculated for average conditions with Equation 11.20 using the autotrophic parameter values for 25°C from Table E11.2 and the values of $S_{N,a}$ and S_{NH} from Example 11.3.3.2:

$$X_{B,A,T} \cdot V = (3.0)(40{,}000)\left\{\frac{[1+(0.20)(0.12)(3.0)](0.20)(30.5-2.1)}{1+(0.12)(3.0)}\right\}$$

$$= 537{,}000 \text{ g autotrophic MLSS.}$$

Substitution of this mass into Equation 11.15 gives:

$$RO_{A,max} = \left\{\left[\frac{4.57-(1.2)(0.20)}{(1.2)(0.20)}\right](1.3)+(1-0.20)(0.12)\right\}\cdot (1.2)(537{,}000)\left(\frac{1.5}{0.75+1.5}\right)$$

$$= 10{,}100{,}000 \text{ g } O_2/\text{day} = 10{,}100 \text{ kg } O_2/\text{day}.$$

d. What is the peak autotrophic oxygen requirement?
The peak autotrophic oxygen requirement is the smaller of $RO_A + RO_{A,TS}$ and $RO_{A,max}$. The value of RO_A, determined in Example 11.3.3.2, is 4980 kg O_2/day. Thus, $RO_A + RO_{A,TS} = 4980 + 6750 = 11{,}730$ kg O_2/day. However, as seen above, $RO_{A,max} = 10{,}100$ kg O_2/day. Thus, ammonia-N breakthrough occurs and the peak autotrophic oxygen requirement is 10,100 kg O_2/day.

e. What is the peak oxygen requirement, RO_P?
Since $RO_{A,max} < (RO_A + RO_{A,TS})$, Equation 11.22 gives the peak oxygen requirement. Recalling the value of RO_H from Example 11.3.3.2:

$$RO_P = 6260 + 5460 + 10{,}100 = 21{,}820 \text{ kg } O_2/\text{day} = 909 \text{ kg } O_2/\text{hr}.$$

f. What is the oxygen peaking factor?
The peaking factor for the oxygen requirement is the peak requirement divided by the average from Example 11.3.3.2, which was 11,240 kg O_2/day:

$$\text{oxygen peaking factor} = \frac{21{,}820}{11{,}240} = 1.94.$$

This occurred for a transient peak in the organic loading on the process of 2.5. This value is slightly higher than the value obtained from Figure 11.9. The difference is due entirely to the difference in the peaking factor for the heterotrophic activity and is caused primarily by the approximations associated with Equation 11.12.

After the peak oxygen requirement has been estimated, it may be used to refine the range of feasible bioreactor volumes and MLSS concentrations. For the steady-state case, the lower limit on the bioreactor volume, V_L, was calculated for the average summer oxygen requirement. When transient loadings occur, that volume must be calculated on the basis of the peak summer oxygen requirement, thereby raising the value of V_L. The impact of the peak oxygen requirement will be examined in the example below for the design we have been considering.

Example 11.3.3.6

What is the range of allowable bioreactor volumes and MLSS concentrations for the CMAS system considered in Examples 11.3.3.2, 11.3.3.3, and 11.3.3.5?

a. What is the lower limit on the bioreactor volume based on the mixing energy constraint to avoid floc shear?
   This lower limit on the bioreactor volume can be calculated with Equation 11.4, after estimation of the required airflow rate, Q, with Equation 11.2. The oxygen requirement in Equation 11.2 should be the peak value with units of kg/hr, giving Q in m^3/min. The same units are used for Q in Equation 11.4. Using a peak oxygen requirement of 909 kg/hr from Example 11.3.3.5, Q is

$$Q = \frac{(6.0)(909)}{10} = 545 \text{ m}^3/\text{min}.$$

   The lower limit on the CMAS bioreactor volume based on floc shear comes from the application of Equation 11.4 using 90 $m^3/(min \cdot 1000\ m^3)$ as an appropriate value for the upper limit on the volumetric power input, $\Pi_{U,Q}$:

$$V_{L,FS} = \frac{(1000)(545)}{90} = 6060 \text{ m}^3.$$

   This value is almost twice the steady-state value because the peak oxygen requirement was almost twice the steady state.

b. What is the lower limit on the bioreactor volume based on the maximum oxygen transfer rate?
   The lower limit on the CMAS bioreactor volume based on the maximum oxygen transfer rate comes from application of Equation 11.5:

$$V_{L,OT} = \frac{909}{0.10} = 9090 \text{ m}^3.$$

   This value is also much larger than the steady-state value. Because it is the larger of the two lower limits, it will control.

c. What is the upper limit on bioreactor volume based on the mixing energy constraint to keep all biomass in suspension?
   The upper limit does not change for the reasons given earlier. It remains 7300 m^3 as calculated in Example 11.3.3.2.

d. What range of bioreactor volumes can be used?
   An examination of the upper and lower limits on the bioreactor volume reveals that they both cannot be met since the lower limit is greater than the upper limit. Thus, the designer must consider the consequences of violating one or both of them. The lower limit of 9090 m^3 was chosen to keep the volumetric oxygen transfer rate no greater than 100 g $O_2/(m^3 \cdot hr)$ at all times, including the diurnal peak, which does not last long. Thus, this volume could be considered to be conservative. Another option would be to use the lower limit associated with floc shear, which is 6060 m^3. This requires a peak oxygen transfer rate of 150 g $O_2/(m^3 \cdot hr)$, which is achievable but would require special attention to the design of the oxygen transfer system. Alternatively, the bioreactor volume could be set equal to the upper limit based on the mixing energy required to keep biomass in suspension during the period when the average oxygen requirement is minimum, which is 7300 m^3. If that volume is chosen, the peak oxygen transfer rate will be 125 g $O_2/(m^3 \cdot hr)$, which will be easier to attain than 150. Thus, while any volume between 6060 and 7300 m^3 could be chosen, it would be more prudent to choose a value near the upper end of the range.

e. What is the new range of MLSS concentrations?
   The imposition of the transient load on the system doesn't change the mass of MLSS in the system, $X_{M,T} \cdot V$. It is the same as calculated in Example 11.3.3.3, 17,170,000 g MLSS. Thus, the lower limit remains 2350 mg/L. The new upper limit is

$$X_{M,T,U} = \frac{17{,}170{,}000}{6060} = 2800 \text{ g/m}^3 = 2800 \text{ mg/L}.$$

f. The transfer of oxygen and the avoidance of floc shear during peak loading conditions greatly limit the designer's options while balancing the costs of the bioreactor and the final settler or membrane system. This is necessary, however, because of the nature of activated sludge oxygen transfer devices and because the effluent from a CMAS bioreactor passes directly to the settler. More latitude, with respect to floc shear, can be gained in the design by adding a flocculation chamber prior to the settler, as discussed in Section 11.2.5 or by adding equalization prior to the CMAS system, as discussed above.

This section has focused on CMAS design as a means for presenting the basic factors that must be considered in the design of any activated sludge system. Like CMAS, EAAS and MBRAS usually have both the biomass and the oxygen requirement distributed uniformly throughout the bioreactor. Consequently, the design procedure for an EAAS or MBRAS system is essentially the same. For EAAS, the only difference is that, due to the long SRTs typically used, floc shear is not usually a factor limiting the bioreactor size. Rather, the major issue is to maintain MLSS in suspension in an economical manner. In contrast, due to the higher MLSS concentrations that can be maintained in MBRAS the bioreactor size is typically limited by the volumetric oxygen transfer capacity.

11.3.4 Conventional, High Purity Oxygen, and Selector Activated Sludge—Systems with Uniform Mixed Liquor Suspended Solids Concentrations, but Variations in Oxygen Requirements

The basic approach to the design of all other activated sludge systems is the same as that presented in Section 11.3.3, although CMAS is the only variation for which that approach can be used directly. However, because of the Guiding Principle No. 4 in Table 10.1, much of the information can be used for the other variations, with appropriate modifications. The activated sludge variations for which the least modification is required are those that have uniform MLSS concentrations throughout. These are conventional activated sludge (CAS), high purity oxygen activated sludge (HPOAS), and selector activated sludge (SAS). As discussed in Sections 7.2 and 11.1.2, CAS and HPOAS can both be considered to behave as a number of completely mixed tanks in series. In HPOAS the separate tanks are real, since the bioreactor is staged by partitioning it, whereas in CAS the tanks are imaginary, representing the residence time distribution of the bioreactor. For purposes of design and analysis, however, both can be considered to be made of equal size tanks in series. Examination of the simulations presented in Figure 7.6 shows that the variation in the MLSS concentration from tank to tank is insignificant, justifying the assumption of uniform concentration throughout the system. An SAS system generally is designed with a series of small completely mixed tanks preceding the main bioreactor, as shown in Figure 1.7. As indicated in Section 11.1.2, the main bioreactor may be one large completely mixed basin or it may be like CAS, in which case it can be considered to behave as a number of tanks in series. In either case, however, the SAS system can be modeled for design and analysis as several unequal tanks in series but with uniform MLSS concentration throughout.

For all of these systems, decisions about the design SRT are made in the same way as previously discussed. Moreover, the mass of MLSS in the system, $(X_{M,T} \cdot V)_{System}$, can be calculated directly with Equation 10.8 and the total steady-state oxygen requirement can be estimated by summing Equations 10.10 and 11.16. The total transient-state oxygen requirement can be estimated with the techniques discussed in Section 11.3.2 and illustrated in Example 11.3.3.5 for a CMAS system. The added level of complexity arises from the need to distribute that oxygen

requirement appropriately throughout the reactor system and to size the oxygen transfer system in a corresponding manner.

The need to spatially distribute the oxygen requirement in CAS and HPOAS systems can be seen by examining Figure 7.5, which shows both the steady state and range of diurnal oxygen requirements in such systems. There it can be seen that the variation from the first to the last tank is large, particularly when diurnal loading variations are imposed on the system. Similar variations will occur in SAS systems, with peak requirements occurring in the selector. Failure to properly account for such spatial variations will lead to poor performance and/or an uneconomic system. Simulation is the most accurate way to predict the variation in oxygen requirement and its use is encouraged, but sufficient information for simulation is often unavailable, requiring approximations. Thus, before we consider the design approach for these systems, we need to consider how to approximate the required spatial oxygen distribution.

11.3.4.1 Approximate Technique for Spatially Distributing Oxygen Requirements

The spatial distribution of the oxygen requirement requires consideration of the different events contributing to oxygen utilization and the rates at which they occur. First consider a system operating at steady state with a uniform loading. Utilization of readily biodegradable substrate is very rapid and will usually be complete even in systems with very short SRTs. It is the major contributor to the high utilization rate in tank one of Figure 7.5. Utilization of slowly biodegradable substrate, on the other hand, is slower because it is limited by the rate of hydrolysis reactions. However, we saw in Figure 10.3 that it is often complete in systems with SRTs as short as two days. This suggests that most slowly biodegradable substrate will have been used in the first third to one-half of a tanks-in-series system, depending on the system SRT. Its use contributed to a substantial portion of the oxygen consumption in the first two tanks in Figure 7.5. Biomass decay, on the other hand, is a very slow reaction that occurs at a constant rate throughout the entire activated sludge process because its rate is driven solely by the biomass concentration, which can be considered to be uniform, as discussed earlier. It contributed the base rate seen in tanks four and five of Figure 7.5, and the same contribution also occurred in all of the preceding tanks. Finally, nitrification is also a slow process but faster than decay. Furthermore, its rate is driven by the ammonia-N concentration, which is soluble and attains a maximum at moderate ammonia-N concentrations. Consequently, nitrification will occur at its maximal rate in the first few tanks in the system, but often will be completed before the last tank is reached, depending on the system SRT. In Figure 7.5, nitrification contributed to oxygen consumption primarily in the first three tanks.

Use can be made of the generalizations in the preceding paragraph for partitioning the total oxygen requirement into its component parts. Consider first the heterotrophic oxygen requirement. Rearrangement of Equation 10.10, after neglecting S_S, allows the oxygen requirement to be divided into two component parts that associated with biomass synthesis and that caused by decay:

$$\begin{aligned} RO_H = {} & F(S_{SO} + X_{SO})(1 - Y_{H,T} i_{O/XB,T}) \\ & + F(S_{SO} + X_{SO})(Y_{H,T} i_{O/XB,T}) \left[\frac{(1 - f_D)(b_H \cdot \Theta_c)}{1 + b_H \cdot \Theta_c} \right]. \end{aligned} \tag{11.23}$$

The first term on the right side of the equation is the oxygen used for synthesis of new biomass, whereas the second term is the oxygen utilization for biomass decay. The synthesis term can be further subdivided into oxygen utilization for biomass synthesis from readily biodegradable substrate, $RO_{H,SS}$:

$$RO_{H,SS} = F \cdot S_{SO}(1 - Y_{H,T} i_{O/XB,T}), \tag{11.24}$$

and oxygen utilization for biomass synthesis from slowly biodegradable substrate, $RO_{H,XS}$:

$$RO_{H,XS} = F \cdot X_{SO}(1 - Y_{H,T} i_{O/XB,T}). \tag{11.25}$$

This subdivision is desirable because of the differences in the rates of utilization of the two substrate types, as discussed above. The oxygen utilization due to decay of biomass, $RO_{H,D}$, may be given by one term because the type of substrate from which biomass was grown has no effect on its decay rate:

$$RO_{H,D} = F(S_{SO} + X_{SO})(Y_{H,T} i_{O/XB,T})\left[\frac{(1 - f_D)(b_H \cdot \Theta_c)}{1 + b_H \cdot \Theta_c}\right]. \tag{11.26}$$

These component oxygen utilization terms can be used to determine the profile of heterotrophic oxygen utilization through a CAS, HPOAS, or SAS system. The distribution of oxygen utilization for decay is the easiest. Because biomass is distributed evenly throughout these systems and because the rate of decay is proportional to the biomass concentration, the rate of oxygen utilization due to decay is the same throughout the system. As a result, the mass of oxygen required for decay in any tank, i, of a multitank system is just:

$$RO_{H,Di} = RO_{H,D}\left(\frac{V_i}{V_T}\right), \tag{11.27}$$

where V_i is the volume of tank i and V_T is the total system volume.

The distribution of the oxygen requirement for the utilization of slowly biodegradable substrate is less exact and is dependent on an approximation. Let $\Theta_{c/XS}$ be the SRT at which slowly biodegradable substrate utilization would be essentially complete in a CMAS system. If we then recognize that biomass is uniformly distributed in a tanks-in-series system, and neglect any kinetic benefits to hydrolysis of having a tanks-in-series configuration, then we can approximate the fraction of the system volume within which biomass synthesis occurs on slowly biodegradable substrate, $f_{V,XS}$, as

$$f_{V,XS} = \frac{\Theta_{c/XS}}{\Theta_c}. \tag{11.28}$$

We saw earlier that slowly biodegradable substrate can be fully utilized at SRTs as short as two days. If we considered that figure to be applicable in a system with an SRT of four days, we might expect oxygen utilization for synthesis of biomass from slowly biodegradable substrate to occur in the first half of the system. Furthermore, if the system could be characterized as being equivalent to five tanks in series, we would expect 40% of that oxygen utilization to occur in each of the first two tanks and 20% in the third. None would occur in the last two tanks. Alternatively, if the system behaved like three tanks in series, we might expect 67% of the utilization to occur in the first tank, 33% in the second, and none in the third. While this technique is crude, it at least provides a means to approximate where oxygen utilization is likely to occur.

Distribution of the oxygen requirement for biomass synthesis from readily biodegradable substrate requires computation of the volume of a fictitious completely mixed bioreactor, V_F, which receives the influent stream and the RAS flow and reduces the substrate concentration to a desired level. The oxygen requirement would then be apportioned to the initial fraction of the activated sludge bioreactor that contained an equivalent volume. If we let S_{SF} be the desired readily biodegradable substrate concentration in the fictitious completely mixed bioreactor, then it follows from

Monod kinetics (Equation 3.36) that the specific growth rate of the biomass in that reactor, $\mu_{H,F}$, must be

$$\mu_{H,F} = \hat{\mu}_H \frac{S_{SF}}{K_S + S_{SF}}. \tag{11.29}$$

A mass balance on readily biodegradable substrate in that reactor, neglecting the contribution of hydrolysis of slowly biodegradable substrate, gives:

$$\mu_{H,F} = \frac{Y_{H,T}[F(S_{SO} - S_{SF}) - F_r(S_{SF} - S_S)]}{f_A \cdot V_F \cdot X_{M,T,F}}, \tag{11.30}$$

where S_S is the readily biodegradable substrate concentration in the effluent from the entire activated sludge process, F_r is the RAS flow rate, and $X_{M,T,F}$ is the MLSS concentration in the fictitious bioreactor. Because the composition of the MLSS in any activated sludge system is the same throughout, the active fraction in the MLSS of the fictitious bioreactor will be the same as the active fraction in the activated sludge system under consideration, which is governed by the system SRT. Making use of this fact, and assuming that $F_r(S_{SF} - S_S) \ll F(S_{SO} - S_{SF})$, that $S_{SF} \ll S_{SO}$, and that $S_S \ll (S_{SO} + X_{SO})$, it can be shown that:

$$f_{XM,F} = \frac{V_F \cdot X_{M,T,F}}{(V \cdot X_{M,T})_{System}} = \left(\frac{1/\Theta_c + b_H}{\mu_{H,F}}\right)\left(\frac{S_{SO}}{S_{SO} + X_{SO}}\right). \tag{11.31}$$

In other words, the fraction of the system MLSS in the fictitious bioreactor, $f_{XM,F}$, is proportional to the ratio of the average net specific growth rate in the process $(1/\Theta_c + b_H)$ relative to the required specific growth rate in the fictitious bioreactor, $\mu_{H,F}$, as given by Equation 11.29. For the case under consideration here, the MLSS concentration in the fictitious bioreactor will be the same as the concentration throughout the process. Therefore, the fraction of the total system volume in which the readily biodegradable substrate is removed, $f_{V,SS}$, is given by

$$f_{V,SS} = \frac{V_F}{V_T} = \left(\frac{1/\Theta_c + b_H}{\mu_{H,F}}\right)\left(\frac{S_{SO}}{S_{SO} + X_{SO}}\right), \tag{11.32}$$

where V_T is the total system volume.

The fraction given by Equation 11.32 represents the smallest possible fraction of the system volume within which the readily biodegradable substrate concentration could be reduced to S_{SF}. The possibility exists that the fraction will be larger because hydrolysis of slowly biodegradable substrate will be occurring, a contribution that was not considered in the derivation of Equation 11.30. The exact contribution of hydrolysis is difficult to estimate without using a model like ASM No. 1, which would defeat the purpose for which these equations are given. However, the largest possible fraction of the system within which the readily biodegradable substrate might be removed can be calculated by assuming that all of the slowly biodegradable substrate is hydrolyzed and contributes to substrate removal in the fictitious bioreactor. When that assumption is made, the right parenthetical term, $S_{SO}/(S_{SO} + X_{SO})$, goes to one. Thus, the largest possible fraction can be calculated with Equation 11.32 with that term set equal to 1.0. The two fractions give the designer bounds within which to apportion the oxygen consumption associated with biomass synthesis from readily biodegradable substrate.

In using Equation 11.29 to calculate $\mu_{H,F}$ for substitution into Equation 11.32, the choice of S_{SF} is very important. It must be large enough to represent the rapid rate of removal of readily biodegradable substrate, but low enough for the assumptions to be valid. Generally, a value of 10% of S_{SO} should be adequate.

The total oxygen requirement for nitrification can be calculated with Equation 11.16. It can be partitioned into two components, synthesis and decay, just as the heterotrophic oxygen requirement was partitioned. However, as shown in Figure 11.8, the effluent ammonia-N concentration from a tanks-in-series system will be less than the concentration from a CMAS system. This will make the oxygen requirement for nitrification slightly larger. Because the effluent ammonia-N concentration cannot be easily predicted without simulation, the effluent ammonia-N concentration should be assumed to be zero. This will provide a slightly conservative estimate of the oxygen requirement. The oxygen requirement associated with synthesis of the autotrophic biomass, $RO_{A,SN}$, can be calculated from

$$RO_{A,SN} = F \cdot S_{N,a}\left(4.57 - Y_{A,T} i_{O/XB,T}\right). \tag{11.33}$$

The oxygen requirement associated with decay of the autotrophic biomass, $RO_{A,D}$, can be calculated with an equation like Equation 11.26:

$$RO_{A,D} = F \cdot S_{N,a}\left(Y_{A,T} i_{O/XB,T}\right)\left[\frac{(1-f_D)(b_A \cdot \Theta_c)}{1 + b_A \cdot \Theta_c}\right]. \tag{11.34}$$

If the wastewater undergoing treatment is a domestic wastewater, the contribution of autotrophic decay to the total oxygen requirement will typically be negligible. However, this may not be the case for industrial wastewaters, and thus both autotrophic oxygen requirements should be calculated explicitly.

The distribution of the oxygen requirement associated with synthesis of the nitrifiers within a tanks-in-series system is very straightforward since nitrification can be assumed to behave as a zero-order reaction in the first part of such a system. The maximum mass nitrification rate, $RN_{A,max}$, can be estimated from a modified form of Equation 5.7 that is analogous to Equation 11.15 for the maximum autotrophic oxygen requirement:

$$RN_{A,max} = \left(\frac{\hat{\mu}_A}{Y_{A,T}}\right)\left(X_{B,A,T} \cdot V\right)\left(\frac{S_O}{K_{A,O} + S_O}\right), \tag{11.35}$$

in which $X_{B,A,T} \cdot V$ is calculated with Equation 11.20. Because nitrification behaves as a zero-order reaction over most of the system, the fraction of the system volume over which autotrophic nitrification will occur, $f_{V,A}$, is just given by

$$f_{V,A} = \frac{F \cdot S_{N,a}}{RN_{A,max}}. \tag{11.36}$$

If the value of $f_{V,A}$ is greater than 1.0, then nitrification will not be complete, which for the tanks-in-series configuration suggests that the SRT is too short and washout will occur (recall Figure 11.8). Actually, this never should occur since the value of $X_{B,A,T} \cdot V$ is required to get $RN_{A,max}$. If the SRT were too low, that value could not be calculated. Thus, a value of $f_{V,A}$ greater than 1.0 suggests that an error has been made somewhere in the computations. Once the value of $f_{V,A}$ is known, the oxygen requirement for nitrification, RO_A, is apportioned proportionally over the fraction of the system

within which nitrification occurs, just as the oxygen requirement for biomass synthesis from slowly biodegradable substrate was apportioned.

As was done for the oxygen requirement associated with decay of heterotrophs, that associated with decay of the autotrophs should be distributed evenly throughout the entire system.

Example 11.3.4.1

Consider the activated sludge system that was the subject of all of the examples in Section 11.3.3. Instead of a CMAS system, however, the system is to be configured as a CAS system with hydraulic characteristics equivalent to four tanks-in-series. Distribute the steady-state oxygen requirement associated with the summer equalized loading.

a. How should the oxygen requirement associated with biomass synthesis from readily biodegradable substrate be distributed?
Calculate the oxygen requirement with Equation 11.24. The readily biodegradable substrate concentration is 115 mg/L (= 115 g/m^3) and the flow rate is 40,000 m^3/day.

$$RO_{H,SS} = (40{,}000)(115)[1-(0.50)(1.20)] = 1{,}840{,}000 \text{ g } O_2/\text{day}$$
$$= 1840 \text{ kg } O_2/\text{day}.$$

Calculate the specific growth rate in a fictitious bioreactor capable of removing 90% of the readily biodegradable substrate. In that case, $S_{SF} = 11.5$ mg/L. Utilization of Equation 11.29 gives:

$$\mu_{H,F} = (8.8)\left(\frac{11.5}{20+11.5}\right) = 3.21 \text{ days}^{-1}.$$

Use it in Equation 11.32 to calculate the smallest possible fraction of the total system volume within which the readily biodegradable substrate is removed. The SRT of the CAS system is three days.

$$f_{V,SS} = \left(\frac{1/3.0+0.22}{3.21}\right)\left(\frac{115}{115+150}\right) = 0.075.$$

The largest possible fraction within which the readily biodegradable substrate is removed can be calculated by setting the right parenthetical term equal to 1.0, giving a value of 0.17. Thus, the readily biodegradable substrate will be removed in 7.5 to 17% of the system volume. Since the CAS system behaves as four equal tanks-in-series, 25% of the system volume is in each tank. Since the readily biodegradable substrate will be removed in less than 25% of the volume, all of the oxygen requirement associated with it should be apportioned to tank No. 1.

b. How should the oxygen requirement associated with biomass synthesis from slowly biodegradable substrate be distributed?
Calculate the oxygen requirement with Equation 11.25. The slowly biodegradable substrate concentration is 150 mg/L (= 150 g/m^3) and the flow rate is 40,000 m^3/day:

$$RO_{H,XS} = (40{,}000)(150)[1-(0.50)(1.20)] = 2{,}400{,}000 \text{ g } O_2/\text{day}$$
$$= 2400 \text{ kg } O_2/\text{day}.$$

Use Equation 11.28 to estimate the fraction of the system volume within which biomass growth on slowly biodegradable substrate occurs. Since the total system SRT is three days, all of the substrate must be fairly easy to degrade. Therefore, it seems reasonable to assume that $\Theta_{c,XS}$ has a value of two days:

$$f_{V,XS} = \frac{2.0}{3.0} = 0.67.$$

Each tank in the system contains 25% of the system volume. Since 67% of the system volume has 100% of the oxygen requirement, 25% of the volume will have 37% of the oxygen requirement (i.e., 0.25/0.67 = 0.37). Therefore, tanks No. 1 and No. 2 will both receive 37% of the oxygen requirement, and tank No. 3 will receive the remainder or 26%. Therefore:

$$RO_{H,XS,1} = (0.37)(2400) = 888 \text{ kg } O_2\text{/day},$$

$$RO_{H,XS,2} = (0.37)(2400) = 888 \text{ kg } O_2\text{/day},$$

and

$$RO_{H,XS,3} = (0.26)(2400) = 624 \text{ kg } O_2\text{/day}.$$

Tank No. 4 will have no oxygen requirement associated with the use of slowly biodegradable substrate.

c. How should the oxygen requirement associated with heterotrophic biomass decay be distributed?
The total oxygen requirement for heterotrophic decay can be calculated with Equation 11.26:

$$RO_{H,D} = (40{,}000)(115+150)(0.50)(1.20)\left[\frac{(1-0.20)(0.22)(3.0)}{1+(0.22)(3.0)}\right]$$

$$= 2{,}020{,}000 \text{ g } O_2\text{/day} = 2020 \text{ kg } O_2\text{/day}.$$

The oxygen requirement associated with decay should be distributed equally to all reactors. Since each reactor contains 25% of the system volume, each will have 25% of the oxygen requirement. Thus, each tank will require 505 kg O_2/day for biomass decay.

d. How should the oxygen requirement associated with synthesis of autotrophic biomass be distributed?
Calculate the oxygen requirement for nitrification with Equation 11.33. The value of $S_{N,a}$ is obtained from Equation 11.17 and can be assumed to be the same as for the CMAS system or 30.5 mg/L as N:

$$RO_{A,SN} = (40{,}000)(30.5)[4.57-(0.2)(1.20)] = 5{,}280{,}000 \text{ g } O_2\text{/day}$$

$$= 5280 \text{ kg } O_2\text{/day}.$$

The fraction of the system volume over which nitrification will occur can be calculated with Equation 11.36, which requires knowledge of the maximum mass nitrification rate from Equation 11.35. That equation, however, requires the mass of autotrophic biomass in the system, which can be calculated with Equation 11.20. Just as the oxygen requirement was

slightly higher in the CAS system because of the lower effluent ammonia-N concentration, so will the mass of autotrophic biomass be higher. In this case it will be

$$X_{B,A,T} \cdot V = (3.0)(40{,}000)\left\{\frac{[1+(0.20)(0.12)(3.0)](0.20)(30.5)}{1+(0.12)(3.0)}\right\}$$

$$= 577{,}000 \text{ g autotrophic MLSS.}$$

It can now be substituted into Equation 11.35. Since the minimum DO concentration in the system is to be 1.5 mg/L, this gives:

$$RN_{A,max} = \left(\frac{1.3}{0.20}\right)(577{,}000)\left(\frac{1.5}{0.75+1.5}\right) = 2{,}500{,}000 \text{ g N/day}$$

$$= 2500 \text{ kg N/day.}$$

This can now be used with Equation 11.36 to find the fraction of the system volume over which nitrification occurs:

$$f_{V,A} = \frac{(40{,}000)(30.5)}{2{,}500{,}000} = 0.49.$$

Each tank in the system contains 25% of the system volume. Since 49% of the system volume receives 100% of the oxygen requirement, 25% of the volume will receive 51% of the oxygen requirement (i.e., 0.25/0.49 = 0.51). Therefore, tank No. 1 will receive 51% of the oxygen requirement and tank No. 2 will receive the remainder or 49%. Therefore:

$$RO_{A,SN,1} = (0.51)(5280) = 2690 \text{ kg } O_2\text{/day}$$

and

$$RO_{A,SN,2} = (0.49)(5280) = 2590 \text{ kg } O_2\text{/day.}$$

e. How should the oxygen requirement associated with autotrophic biomass decay be distributed?
The oxygen requirement associated with decay of the autotrophic biomass can be calculated with Equation 11.34:

$$RO_{A,D} = (40{,}000)(30.5)(0.20)(1.20)\left[\frac{(1-0.20)(0.12)(3.0)}{1+(0.12)(3.0)}\right]$$

$$= 62{,}000 \text{ g } O_2\text{/day} = 62 \text{ kg } O_2\text{/day.}$$

This will be distributed equally to all bioreactors, so each has an oxygen requirement of 15.5 kg O_2/day, which is negligible, compared to the other oxygen requirements in the system.

f. The total oxygen requirement in each equivalent tank of the CAS system is given in Table E11.3, where the component oxygen requirements are summarized.
An examination of Table E11.3 reveals several things. First, the total heterotrophic oxygen requirement is 6260 kg O_2/day, which is the same as in the CMAS system, as calculated in Example 11.3.3.2. This is to be expected since the equations for distributing the oxygen requirement were derived by partitioning the equations for a CMAS system. Nevertheless, the agreement between the values serves as a convenient check on the calculations. Second, the total autotrophic oxygen requirement is 5340 kg O_2/day, which is slightly higher than the

TABLE E11.3
Distribution of Steady-State Oxygen Requirement in CAS System of Example 11.3.4.1

	Oxygen Requirement, kg O_2/day			
Component	**Tank 1**	**Tank 2**	**Tank 3**	**Tank 4**
Readily biodegradable organic matter	1840	0	0	0
Slowly biodegradable organic matter	888	888	624	0
Heterotrophic decay	505	505	505	505
Nitrification	2690	2590	0	0
Autotrophic decay	16	16	16	16
Total	5939	3999	1145	521

4980 kg O_2/day calculated for the CMAS system in Example 11.3.3.2. The higher value is due to the greater extent of nitrification that will occur in a CAS-type system. Third, the majority of the oxygen requirement occurs in the first two equivalent tanks of the CAS-type system, which is in agreement with the simulations in Section 7.2.2 and experience in the field. This means that the volumetric oxygen transfer rate in the first equivalent tanks of a CAS-type system will be greater than the rate in a CMAS system if the total bioreactor volumes are the same. Fourth, the oxygen requirements in the last equivalent tanks of a CAS-type system are small, suggesting that the required volumetric oxygen transfer rate is smaller than that in an equivalent CMAS system.

g. How accurate is the approximate technique for distributing the oxygen requirement?
Simulations conducted with ASM No.1 gave smaller oxygen requirements in the first two equivalent tanks and larger requirements in the last two, than the values indicated in Table E11.3. The values from the simulation are more accurate because they come from a true dynamic model that has been shown to mimic real-world activated sludge bioreactors well.[7] The reasons for this difference are the assumptions made about biodegradation of slowly biodegradable substrate and nitrification in the approximate technique. Thus, when the approximate technique of this section is used to distribute the steady-state oxygen requirement in a CAS, HPOAS, or SAS system, its tendency to overpredict the oxygen requirement at the start of the system and to underpredict the requirement at the end of the system should be kept in mind.

The distribution of the transient-state oxygen requirement is even more difficult to approximate than the distribution of the steady-state requirement. Consequently, simulation with a dynamic model like one of the IWA activated sludge models is the only way to accurately determine how the oxygen supply should be distributed to a CAS, HPOAS, or SAS system receiving diurnal loads. The reason for this is the uncertainty associated with the biodegradation of the slowly biodegradable organic matter and with nitrification. If it is necessary to make such a distribution without simulation, then it should be done by applying the same principles used to distribute the steady-state requirement. Since that distribution is based on dividing the oxygen requirement into its component parts, it is directly applicable to the transient-state case, making it unnecessary to use the approach used for a CMAS system. In applying the principles presented above, one need only consider the impact of the transient loading on the synthesis oxygen requirement. The oxygen requirements associated with decay will not change because the mass of MLSS in the system changes little in response to the transient loading, as discussed earlier.

11.3.4.2 Design of Conventional Activated Sludge Systems

The design of a CAS system is an iterative process because of the need to distribute the oxygen requirement throughout the bioreactor. This requires knowledge of its hydraulic characteristics, expressed as the number of equivalent tanks-in-series, which are determined by its size and

configuration. As a consequence, the designer must select a bioreactor volume and shape, determine the number of equivalent tanks-in-series, and distribute the oxygen requirement among those equivalent tanks. Only then is it possible to check the mixing intensities in the equivalent tanks. If they are not adequate, then another bioreactor volume and shape must be selected and the process repeated until an acceptable combination is obtained. Luckily, the selection process is simplified somewhat by a relaxation of the mixing constraints as a result of the tanks-in-series behavior of a CAS bioreactor. The mixing intensity in the first equivalent tank may exceed the floc shear limitation, Π_U, as long as the mixing intensity in subsequent equivalent tanks is sufficiently small to allow for reflocculation of the biomass.[70] However, consideration must still be given to the maximum practical oxygen transfer rate. At the other end, where the oxygen requirement is small, the mixing intensity in the last equivalent tank can be set equal to the limit for adequate mixing, Π_L, without significant sacrifice in power costs because only a small percentage of the total system volume will be involved.

To start the process, Equation 10.8 may be used to calculate the mass of MLSS in the system $(X_{M,T} \cdot V)_{System}$. Using the procedures in Section 11.3.3, the total steady-state oxygen requirement can then be used to select the upper and lower limits on bioreactor volume that would exist if the system were configured as a CMAS system. The design volume for the CAS system should be intermediate between these values because the oxygen requirement in the first equivalent tank will be greater than the rate in the CMAS system and the rate in the last equivalent tank will be smaller, as illustrated in Example 11.3.4.1. Generally, however, the selected volume should be closer to the upper limit for the CMAS system because of limitations on the maximum rate of oxygen transfer in the first equivalent tank. Since the choice of this volume fixes the target MLSS concentration (through the calculated $[X_{M,T} \cdot V]_{System}$ value), it should be based on consideration of the anticipated solids settling characteristics and the proposed final settler design, including the biomass recycle ratio, α. The bioreactor configuration (i.e., its length, L; width, W; and depth, H) is then selected by considering the site characteristics. The next task is to estimate the number of equivalent tanks-in-series, N, for the chosen dimensions. This may be done with an empirical relationship developed at the Water Research Centre in England:[67]

$$N = 7.4\left[\frac{F(1+\alpha)L}{W \cdot H}\right], \tag{11.37}$$

in which the units of the flow rate, F, are m^3/sec, and L, W, and H are all expressed in meters. N must be an integer value and thus should be rounded appropriately. Once N is known, then the volume of each equivalent tank, V_i, may be obtained by dividing the chosen system volume, V, by N. The volumetric oxygen transfer rate in any tank, $r_{SO,i}$, can then be obtained from

$$r_{SO,i} = \frac{RO_i}{V_i}, \tag{11.38}$$

where

$$RO_i = RO_{H,SS,i} + RO_{H,XS,i} + RO_{H,D,i} + RO_{A,SN,i} + RO_{A,D,i}. \tag{11.39}$$

For average load conditions, the highest practical sustained volumetric oxygen transfer rate is approximately 100 g $O_2/(m^3 \cdot hr)$ for conventional oxygen transfer systems, as discussed previously. Values as high as 150 g $O_2/(m^3 \cdot hr)$ can be obtained under peak load conditions, but values greater than 100 should not be used for prolonged operation of mechanical aeration systems because of

rapid wear of the components. Thus, when a conventional oxygen transfer system is to be used, if the calculated volumetric oxygen transfer rate exceeds 100 g O_2/(m^3·hr) a larger total tank volume must be chosen or special attention should be given to the design of the oxygen transfer system. Use of a larger total tank volume will require alteration of the tank dimensions, which will require recalculation of the number of equivalent tanks-in-series, N. Because the use of Equation 11.37 requires rounding to arrive at an integer value for N, judgment must be exercised in deciding whether a change in N is justified. If N is changed, the oxygen requirement must be redistributed and the procedure repeated. After a system size has been determined based on oxygen transfer in the first equivalent tank, the mixing energy input into the last few equivalent tanks must be calculated using RO_i for them and the techniques presented for a CMAS system. If the mixing energy input is less than Π_L, additional mixing must be provided to maintain the MLSS in suspension. The objective is to arrive at a system size that will allow the high oxygen requirements in the front of the system to be met while minimizing the amount of mixing energy in excess of Π_L required in the back of the system. Consideration must be given to both the maximum oxygen requirement for summer operation and the minimum for winter operation. Consequently, this exercise will require judgment and compromise.

Example 11.3.4.2

Continue with the problem of Example 11.3.4.1. A tentative decision has been made to use a reactor volume of 7500 m^3 for the CAS system, configured in such a way that the hydraulic characteristics will be equivalent to four tanks-in-series. Is this an acceptable choice?

a. What is the required volumetric oxygen transfer rate in the first equivalent tank?
   From Table E11.3 of Example 11.3.4.1, the required oxygen transfer rate to the first tank is 5939 kg O_2/day, which is 247 kg O_2/hr. The volume of the first equivalent tank is one-fourth of the total system volume or 1875 m^3. The volumetric oxygen transfer rate can be calculated with Equation 11.38, giving a value of

$$r_{SO,1} = \frac{247}{1875} = 0.132 \text{ kg } O_2/(m^3 \cdot hr) = 132 \text{ g } O_2/(m^3 \cdot hr).$$

   This value is higher than can be achieved on a sustained basis with conventional mechanical oxygen transfer equipment, but can be attained with diffused aeration by giving careful consideration to the design of the oxygen transfer system. Thus, the selected volume is acceptable from the standpoint of the maximum oxygen transfer rate.
b. What air flow rate will be required in the last equivalent tank if the oxygen transfer efficiency is 10%?
   From Table E11.3, the required oxygen transfer rate in the last equivalent tank is 521 kg O_2/day (22 kg O_2/hr). The air flow rate required to achieve this can be calculated with Equation 11.2 in which RO must be expressed as kg O_2/hr:

$$Q = \frac{(6.0)(22)}{10} = 13.2 \text{ m}^3/\text{min}.$$

   The air flow rate required to keep the MLSS in suspension can be calculated from Equation 11.3 using the equivalent tank volume of 1875 m^3:

$$Q = \frac{(20)(1875)}{1000} = 37.5 \text{ m}^3/\text{min}.$$

The air flow rate required to keep the biomass in suspension is larger than that required to deliver the needed oxygen and thus it controls. The tank volume cannot be decreased to reduce this amount because then it would be impossible to deliver the needed oxygen to the first equivalent tank. This commonly occurs in CAS systems. The design can be considered to be acceptable.

11.3.4.3 Design of High Purity Oxygen Activated Sludge Systems

Design of an HPOAS system is essentially the same as the design of a CAS system, with two significant differences. First, the system is actually staged, so that the number of tanks-in-series is chosen by the designer. Thus, there is no need to iterate when distributing the oxygen requirement from tank to tank. Second, because high purity oxygen is introduced into the first stage and the gas moves sequentially from stage to stage from the inlet to the outlet, the partial pressure of oxygen in the gas phase decreases as it moves through the system. As a result, the gas phase is most heavily enriched in oxygen in that part of the system where the required oxygen transfer rate is highest, allowing high oxygen transfer rates to be achieved with lower power inputs than would be required if the gas phase were air. This makes the power inputs more uniform throughout the system and alleviates some of the problems associated with balancing the power inputs into CAS systems. Of course, mixing must still be checked to ensure that sufficient energy is expended to maintain the MLSS in suspension.

11.3.4.4 Design of Selector Activated Sludge Systems

Selector activated sludge systems are used when either the wastewater or the bioreactor configuration of the chosen activated sludge variation have a tendency to favor the growth of filamentous bacteria. A selector counteracts that tendency. As described in Section 11.1.2, a selector is a highly loaded section at the inlet end of an activated sludge bioreactor. The high loading condition creates an environment that favors the growth of floc-forming rather than filamentous bacteria, as discussed in Section 11.2.1. This results in improved solids settling characteristics, allowing the use of higher MLSS concentrations and/or higher secondary clarifier solids loading rates.

The selector must be properly sized if kinetic selection of floc-forming bacteria is to occur. First, it must be large enough to remove the majority of the readily biodegradable organic matter applied. If it is not, some of that organic matter will pass through the selector into the remaining portion of the bioreactor where environmental conditions favor the growth of filamentous bacteria. Second, it must be small enough to maintain the readily biodegradable substrate concentration on the right side of the crossover point on a plot of specific growth rate versus substrate concentration for the two bacterial types, as shown in Figure 11.2. If it isn't, environmental conditions favoring the growth of floc-forming bacteria will not be established in the selector, resulting in the preferential growth of filamentous bacteria in it.

Several factors must be considered when choosing the size and configuration of a selector. Because it is difficult to generalize about the kinetic parameters describing growth of floc-forming and filamentous bacteria, it is impossible to identify the readily biodegradable substrate concentration associated with the crossover point on the μ:S_S curve. However, empirical evidence has suggested the minimum initial process loading factor that will lead to good settling biomass, as shown in Figure 11.5 and discussed in Section 11.2.1. Since the process loading factor is proportional to the specific growth rate, specification of the process loading factor for the selector is analogous to specifying the crossover point on the μ:S_S curve. Consequently, a selector design is typically based on specification of a process loading factor for it. However, because influent loadings vary, it is difficult to specify a single selector volume that is large enough to remove all of the readily biodegradable organic matter while also providing a sufficiently high process loading factor under all loading conditions. This problem can be solved by using a staged selector, as illustrated in Figure 1.7. During periods of low loading, the process loading factor will be sufficiently high in the first stage to favor the floc-forming bacteria, yet the bulk of the readily

biodegradable substrate will be removed. During periods of high loading, on the other hand, the additional selector stages ensure that the readily biodegradable organic matter is removed prior to the main bioreactor.

Our understanding of the organism selection and organic substrate removal mechanisms occurring within SAS systems continues to evolve and, consequently, so are the approaches used to design selector systems. Based on current knowledge, however, the following general approach has been successful:[36]

1. A minimum of three stages should be used. The first two stages should each contain 25% of the total selector volume, while the third stage should contain the remaining 50%.
2. The total selector volume is chosen to give an overall process loading factor of 3 kg total biodegradable COD/(kg MLSS · day), which is analogous to about 1.75 kg BOD_5/(kg MLSS · day). Note that even though the purpose of the selector is to remove readily biodegradable organic matter, the process loading factor is based on the total biodegradable COD. This is because the empirical evidence upon which it is based (i.e., Figure 11.5) did not distinguish between the readily and slowly biodegradable organic matter in the wastewater.

This relatively low overall process loading factor results in good removal of readily biodegradable organic matter, even during periods when the organic loading is temporarily higher. Furthermore, the use of the selected overall process loading factor will result in a process loading factor in the initial selector stage of 12 kg COD/(kg MLSS · day). A reference to Figure 11.5 indicates that this high initial stage process loading factor provides a factor of safety to ensure that sufficiently high process loading factors are achieved regardless of the influent load.

Based on the above considerations, the selector volume is chosen from a rearrangement of the definition of the process loading factor as given by Equation 5.48:

$$V_s = \frac{F(S_{SO} + X_{SO})}{U_s \cdot X_{M,T}}, \tag{11.40}$$

where the influent COD includes both readily and slowly biodegradable substrate, and V_s and U_s are the volume and process loading factor for the selector, respectively. The MLSS concentration is the same as in the rest of the activated sludge system, and is determined from $(X_{M,T} \cdot V)_{System}$ as calculated with Equation 10.8, where V is the selected total system volume. The volume of the rest of the system is just $V - V_s$.

Determination of the required oxygen transfer rate within a selector is not straightforward. Experience with some selector installations indicates that respiration rates in the first stage may be as high as 40 to 60 g O_2/(kg MLSS·hr).[19,36] These values are high, but are less than would be expected if the heterotrophic bacteria were growing at their maximum specific growth rate. It appears that, in some instances, the specific growth rates in the selector are sufficiently high to trigger a substrate storage response called the "selector effect."[36] In those cases, selector oxygen requirements corresponding to the oxidation of only about 20% of the COD removed have been observed.[36] It has been hypothesized that the remainder of the removed substrate is stored as intracellular carbon storage polymers such as glycogen and/or poly-β-hydroxyalkanoate. Because COD is conserved when storage polymer formation occurs, it is not necessary to supply sufficient oxygen in the selector to oxidize all of the removed substrate. Rather, the stored substrate is oxidized in the main bioreactor, and the oxygen requirement associated with it must be met there. Further research is needed to refine our understanding of the conditions under which this phenomenon occurs. In the meantime, it is prudent to design aerobic selector systems with significant flexibility and capacity in terms of the oxygen that can be transferred.

The determination of the oxygen requirement in the selector can be made by the same techniques used to spatially distribute the oxygen requirement in multitank systems, as discussed earlier in this section. From a conservative perspective, the oxygen requirement in the selector can be calculated by assuming that all of the readily biodegradable organic matter is removed and used for biomass synthesis in the selector. In other words, the selector effect is assumed to not occur. The oxygen requirement associated with the removal of slowly biodegradable organic matter can be distributed in proportion to the selector volume as a fraction of the system volume in which biomass synthesis occurs on slowly biodegradable substrate as determined with Equation 11.28. The oxygen requirement for decay is given by Equation 11.27 in which V_i is the selector volume, V_s. Finally, nitrification will be occurring at its maximal rate in the selector, as given by Equation 11.15. Because the selector is highly loaded, the volumetric oxygen transfer rate is likely to be quite high, particularly during periods of peak loading. As a consequence, special care should be given to the design of the oxygen transfer system in it to ensure that the needed transfer rate can be achieved.

The oxygen transfer system for the main bioreactor should be designed as if the selector were not present. There are two reasons for doing this. First, it will be necessary to bypass the selector periodically for maintenance and other purposes, placing the entire oxygen requirement into the main bioreactor. Second, the extent to which the selector effect will occur is usually unknown. However, the larger it is, the more the oxygen requirement is shifted to the main reactor. Thus, designing the main bioreactor to handle the entire oxygen requirement ensures that any situation can be handled.

Example 11.3.4.3

Consider the CMAS system that was the subject of the examples in Section 11.3.3. In Example 11.3.3.6 the range of acceptable bioreactor volumes and associated MLSS concentrations was determined for the unequalized case and was found to be small. It was also found that it would be preferable to design the system for the maximum feasible volume. Thus, assume that a volume of 7300 m^3 was chosen, giving an MLSS concentration of 2350 mg/L. Size a three-compartment selector for the system and determine the maximum potential oxygen requirement under average loading conditions.

a. What is the size of the selector?
   As seen above, aerobic selectors are typically sized with an overall process loading factor of 3.0 kg total biodegradable COD/(kg MLSS·day). Adopting that value and recognizing that the MLSS concentration in the selector will be the same as the MLSS concentration in the main bioreactor (i.e., 2350 mg/L = 2.35 kg/m^3), the selector volume can be calculated with Equation 11.40. Recalling that the average flow rate is 40,000 m^3/day and that the average readily and slowly biodegradable substrate concentrations are 115 mg COD/L (0.115 kg/m^3) and 150 mg COD/L (0.150 kg/m^3), respectively, gives:

$$V_s = \frac{(40{,}000)(0.115+0.150)}{(3.0)(2.35)} = 1500 \text{ m}^3.$$

b. How is the selector configured?
   It should be configured as three tanks-in-series, with the first two each being 25% of the total volume and the third 50%. Thus, the first two selectors each have a volume of 375 m^3 and the third a volume of 750 m^3.
c. What is the volume of the main CMAS bioreactor?
   The total system volume is not changed by the addition of the selector, so the main reactor volume is reduced by the volume of the selector. Therefore:

$$V_{CMAS} = 7300 - 1500 = 5800 \text{ m}^3.$$

d. What is the oxygen requirement in the selector under average load conditions in the summer?
The oxygen requirement associated with biomass synthesis from readily biodegradable substrate was determined in Example 11.3.4.1 to be 1840 kg O_2/day by using Equation 11.24. All of it will occur in the selector.

The oxygen requirement for biomass synthesis from slowly biodegradable substrate was determined in Example 11.3.4.1 to be 2400 kg O_2/day by using Equation 11.25. Because the rest of the system is not compartmentalized, the utilization of slowly biodegradable substrate will occur uniformly throughout it. Therefore, the utilization of slowly biodegradable substrate in the selector can just be assumed to be in proportion to its volume as a fraction of the total system volume, as can the oxygen requirement associated with it, $RO_{H,XS,s}$. (If the remainder of the system was like a CAS system, rather than being a CMAS system, the oxygen requirement would have to be distributed in the same way as in Example 11.3.4.1.) Therefore:

$$RO_{H,XS,s} = \left(\frac{1500}{7300}\right)2400 = 490 \text{ kg } O_2/\text{day}.$$

The oxygen requirement associated with decay of heterotrophic biomass was determined in Example 11.3.4.1 to be 2020 kg O_2/day by using Equation 11.26. The amount of oxygen required in the selector for heterotrophic decay, $RO_{H,D,s}$, will be in proportion to its volume:

$$RO_{H,D,s} = \left(\frac{1500}{7300}\right)2020 = 415 \text{ kg } O_2/\text{day}.$$

The oxygen requirement associated with autotrophic biomass synthesis was determined in Example 11.3.4.1 to be 5280 kg O_2/day by using Equation 11.33. If the DO concentration in the selector is maintained at 1.5 mg/L, then nitrification will occur at the maximum rate, as calculated in that example. Using the logic used in Example 11.3.4.1 to determine the fraction of the nitrification oxygen requirement occurring in each tank of a CAS system, the fraction of the nitrification oxygen requirement occurring in the selector is (1500/7300) ÷ 0.49 = 0.42. Therefore, oxygen requirement associated with autotrophic biomass synthesis in the selector, $RO_{A,SN,s}$, is

$$RO_{A,SN,s} = (0.42)(5280) = 2210 \text{ kg } O_2/\text{day}.$$

Finally, the oxygen requirement associated with decay of autotrophic biomass was determined in Example 11.3.4.1 to be 62 kg O_2/day by using Equation 11.35. The amount of oxygen required in the selector for autotrophic decay, $RO_{A,D,s}$, will be in proportion to its volume:

$$RO_{A,D,s} = \left(\frac{1500}{7300}\right)62 = 13 \text{ kg } O_2/\text{day}.$$

The total oxygen requirement in the selector, RO_s, is the sum of all the above components:

$$RO_s = 1840 + 490 + 415 + 2210 + 13 = 4968 \text{ kg } O_2/\text{day} = 207 \text{ kg } O_2/\text{hr}.$$

e. What is the required volumetric oxygen transfer rate?
The required volumetric oxygen transfer rate is determined by dividing RO_s by V_s, giving a value of 138 g $O_2/(m^3 \cdot hr)$. This can be achieved by giving special attention to the design of the oxygen transfer system.

Techniques for designing aerobic selectors will continue to evolve as our understanding of microbial competition and metabolic selection increase. Consequently, the approach presented here can be expected to be modified or replaced in the future.

11.3.5 Step Feed and Contact Stabilization Activated Sludge—Systems with Nonuniform Mixed Liquor Suspended Solids Concentrations

Examination of Figures 7.15 and 7.26 reveals that SFAS and CSAS have two characteristics in common; the MLSS concentration is not uniform throughout the system and the concentration in the tank that discharges to the final settler or membrane system is the lowest of all of the tanks in the system. These are a direct result of the influent and recycle flow distributions as discussed in Sections 7.3.4 and 7.4.4. Furthermore, they provide the justification for the choice of these activated sludge variations for a particular installation. Guiding Principle No. 4a in Table 10.1 states that all activated sludge variants with the same SRT contain about the same mass of biomass. As a consequence, when an SFAS or CSAS process is designed to have the same MLSS concentration entering the final settler or membrane system as one of the activated sludge processes with uniform biomass concentration, the SFAS or CSAS process will always have a smaller total volume. Furthermore, the savings in system volume will be greater the longer the system SRT and the easier the organic substrate is to remove. Thus, SFAS and CSAS are often chosen for situations where space is limited. Another reason for designing a system so that it can be operated as SFAS was discussed in Section 7.3.4 and the same reason also applies to CSAS. If a CAS or CMAS system receives extremely high flows periodically, the high flow rate may cause the solids loading on a final settler to exceed allowable values, leading to loss of MLSS, poor effluent quality, and process failure. However, if the operating configuration can be switched to SFAS or CSAS, the MLSS concentration entering a final settler will be reduced, thereby keeping the solids loading on the clarifier within an acceptable range.[63] Thus, another reason wastewater treatment plants are designed with the ability to operate in the SFAS or CSAS mode is for operational flexibility.

One consequence of the hydraulic characteristics of a CSAS system is that its performance is much more dependent on the recycle ratio (and thus on the RAS flow rate) than other activated sludge variations, as illustrated in Figures 7.24 and 7.25. This is because of the effect that the recycle ratio has on the MLSS concentration gradient through the system, with higher recycle ratios diminishing the gradient. For the same reason, the performance of an SFAS system is also somewhat sensitive to the recycle ratio, but much less so, as illustrated in Figures 7.17 and 7.18. Nevertheless, for both systems, more consideration must be given to the impacts of the RAS flow rate during system design, particularly when its impact on final settler size is taken into consideration. As far as effluent quality is concerned, it is often acceptable to consider only the minimum anticipated recycle ratio because it is the critical one, causing the maximum MLSS gradient and minimizing the amount of biomass in the last bioreactor. If the system produces an acceptable effluent soluble substrate concentration at that recycle ratio, it will also do so at all higher ones, as shown in Figures 7.17 and 7.24. On the other hand, the oxygen requirement in the contact tank of a CSAS system can increase significantly as the recycle ratio is increased, particularly when nitrification is occurring, as shown in Figure 7.25. Therefore, the distribution of oxygen requirements should be examined at both the upper and lower anticipated recycle ratios.

For an existing system with uniform MLSS concentration, the impact of switching to a SFAS or CSAS operational configuration can be calculated by using mass balance techniques based on a selected RAS flow rate, the anticipated distribution of influent and RAS flows to the various tanks, and the mass of MLSS in the system as calculated with Equation 10.8. For the design of a new system in either SFAS or CSAS mode, the situation is more complicated because of the large number of choices involved, particularly for CSAS as illustrated in Section 7.4.4. Nevertheless, in all situations the over-riding criterion is that the specific growth rate of the biomass in the tank discharging to the final settler must be low enough to allow effluent quality objectives to be met. Furthermore, all of

the criteria discussed earlier about oxygen transfer and mixing energy input must also be met. This requires distribution of the oxygen requirement among the various tanks. Because it is the simpler of the two, we will first investigate SFAS. Then we will consider CSAS.

11.3.5.1 Design of Step Feed Activated Sludge Systems

As shown in Figure 1.5 and illustrated schematically in Figure 7.10, SFAS systems are usually configured so that they behave as equal-sized tanks-in-series with flow distributed equally to each tank. Other configurations can, and often are, used. For example, the equivalent tanks may not all be of equal size or influent may not be distributed to all, particularly the last. Furthermore, in some cases it may be advantageous to distribute the influent in unequal portions to the various tanks. Because it would be impossible to quantitatively describe all possible configurations, this presentation is limited to the case of equally sized tanks with equal distribution of flows to all of them. The concepts presented can be extended to other configurations by the reader should the need arise. The design of a SFAS system follows the same general approach as the design of the other alternatives presented earlier, with some additional steps. The emphasis here will be on the additional steps.

The first task in the design is to choose the bioreactor configuration, fixing the number of equivalent tanks in series, N, and the SRT, choices that are made by considering the factors presented earlier. Once those choices have been made, the mass of biomass in the system, $(X_{M,T} \cdot V)_{System}$, can be calculated by using Equation 10.8 for a CMAS system.

The next task is to ensure that the selected configuration and SRT will result in the desired effluent quality. For that to occur, the specific growth rate of the heterotrophs in the last tank, $\mu_{H,N}$, must be equal to or smaller than the specific growth rate as calculated by the Monod equation, Equation 3.36. The approach used to calculate that specific growth rate is similar to the approach for determining the specific growth rate in the fictitious bioreactor used in the distribution of the oxygen requirement in CAS, HPOAS, and SAS systems, as discussed in Section 11.3.4. Thus, Equation 11.29 may be used in which S_{SF} is set equal to the desired effluent substrate concentration. The chosen effluent concentration can be either the readily biodegradable substrate concentration or the ammonia-N concentration, depending on the type of standard that must be met. A lower specific growth rate will be required to meet an ammonia-N standard than a soluble biodegradable COD standard. Nevertheless, for the development that follows, organic substrate removal will be assumed to be the objective, making the heterotrophic specific growth rate controlling. The logic is similar to that used to derive Equations 11.30 and 11.31, except that in this case the production of soluble substrate by hydrolysis of slowly biodegradable substrate will make a significant contribution to biomass growth in the last tank. If we assume that utilization of both readily and slowly biodegradable substrate is important, it can be shown that the specific growth rate in the last tank is given by

$$\mu_{H,N} = \frac{(X_{M,T} \cdot V)_{System}(1/\Theta_c + b_H)}{N(X_{M,T,N} \cdot V_N)}, \tag{11.41}$$

in which V_N is the volume of the last tank (which is the same as V_T/N since all tanks have the same volume) and $X_{M,T,N}$ is the MLSS concentration in the last tank. The derivation of this equation is dependent on the assumption that the mass flow rate of substrate into the last tank from the preceding tank is much less than the mass flow rate from the influent, which will generally be true. Since the value of $\mu_{H,N}$ as calculated with Equation 11.41 must be less than or equal to the value of $\mu_{H,N}$ associated with the desired effluent substrate concentration as calculated with Equation 11.29, then

$$\frac{X_{M,T,N} \cdot V_N}{(X_{M,T} \cdot V)_{System}} = f_{XM,N} \geq \frac{1/\Theta_c + b_H}{N \cdot \mu_{H,N}}, \tag{11.42}$$

where $\mu_{H,N}$ has been calculated with Equation 11.29. In other words, as long as the fraction of MLSS in the N^{th} tank, $f_{XM,N}$, is greater than or equal to the right side of Equation 11.42, the effluent quality will be acceptable.

The fraction of MLSS in any given tank of a SFAS system, $f_{XM,i}$, is determined totally by the hydraulics of the system. If both growth and wastage are neglected, then the fraction of MLSS in any tank i of an SFAS system in which the influent flow is split equally among N equal size tanks is given by

$$\frac{X_{M,T,i} \cdot V_i}{(X_{M,T} \cdot V)_{System}} = f_{XM,i} = \frac{\dfrac{1.0+\alpha}{(i/N)+\alpha}}{\displaystyle\sum_{i=1}^{N} \frac{1.0+\alpha}{(i/N)+\alpha}}, \tag{11.43}$$

where α is the biomass recycle ratio, which is the RAS flow rate, F_r, divided by the influent flow rate to the system, F. Furthermore, since the last tank is tank N, the fraction of MLSS in the last tank is

$$\frac{X_{M,T,N} \cdot V_N}{(X_{M,T} \cdot V)_{System}} = f_{XM,N} = \frac{1.0}{\displaystyle\sum_{i=1}^{N} \frac{1.0+\alpha}{(i/N)+\alpha}}. \tag{11.44}$$

Figure 11.10 is a plot of Equation 11.44 over the range of recycle ratios and number of equivalent tanks likely to be encountered in practice. To determine whether a proposed SFAS system will work, the fraction of MLSS in the last tank should be determined with Equation 11.44 or Figure 11.10 for the smallest anticipated recycle ratio and that value should be used to determine if Equation 11.42 is satisfied. As long as that fraction is greater than the right side of Equation 11.42, the desired effluent quality will be met and a SFAS system can be used.

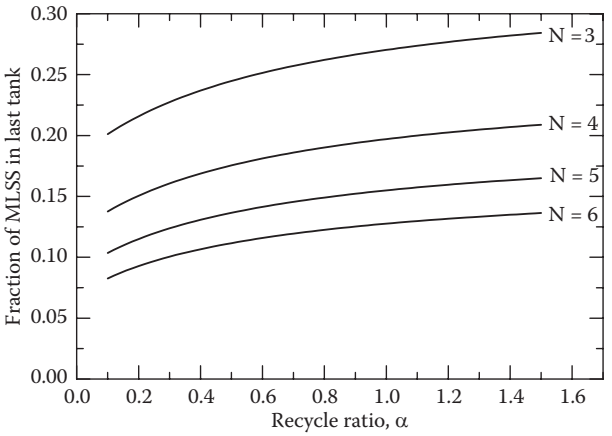

FIGURE 11.10 Effect of the biomass recycle ratio on the fraction of MLSS in the last tank of an SFAS system, as calculated with Equation 11.44.

Example 11.3.5.1

Consider the wastewater that was the subject of the examples in Section 11.3.3 and 11.3.4. Consideration is being given to using a SFAS system with an SRT of three days that is equivalent to four tanks in series with equal distribution of the influent to all tanks. The effluent quality objective is 10 mg/L as COD of readily biodegradable organic matter. Can that objective be met if the recycle ratio is 0.5?

a. What specific growth rate is required in the last tank of the SFAS system?
   Calculate the specific growth rate in the last tank, $\mu_{H,N}$, using Equation 11.29 and the kinetic parameters for winter conditions, since they will control. The desired substrate concentration is 10 mg/L. Using the kinetic parameters from Table E11.2 gives:

$$\mu_{H,N} = (4.1)\left(\frac{10}{20+10}\right) = 1.37 \text{ days}^{-1}.$$

b. What is the smallest fraction of the MLSS that can be in the last tank?
   Use of $\mu_{H,N}$ in Equation 11.42 gives the smallest fraction of the MLSS that can be in the last tank:

$$f_{XM,N} \geq \frac{1/3.0+0.15}{(4)(1.37)} = 0.088.$$

   Thus, as long as more than 8.8% of the MLSS is in the last tank, the effluent quality goal can be met.
c. Will the proposed system be capable of meeting the effluent quality goal?
   The actual fraction of MLSS in the last tank can be calculated with Equation 11.44, or read from Figure 11.10. For a system equivalent to four tanks in series with a recycle ratio of 0.5, the fraction of MLSS in the last tank can be seen from the figure to be 0.175. Thus, the SFAS system is capable of meeting the effluent quality goal. In fact, examination of the figure reveals that the goal can be met no matter what the recycle ratio is.

The next task in the design is to distribute the steady-state oxygen requirement to each of the equivalent tanks. This can be done using the techniques described for spatially distributing the oxygen requirement in CAS systems, but in a simpler manner. The oxygen requirements for heterotrophic biomass synthesis from slowly and readily biodegradable substrate can be calculated with Equations 11.24 and 11.25, respectively. Likewise, the requirement for synthesis of autotrophic biomass can be calculated with Equation 11.33. As seen in Figure 7.11, the extent of nitrification in SFAS is essentially the same as in a CMAS system. However, the ammonia-N concentration in the early tanks will be lower, as shown in Figure 7.15. Thus, to be conservative, Equation 11.33, which assumed that the residual ammonia-N concentration was negligible, should be used without modification. Because the influent to the type of SFAS system considered here is distributed evenly among the tanks, all of the synthesis oxygen requirements should be apportioned evenly as well. If some other influent flow distribution were used, the synthesis oxygen requirements should be apportioned in the same manner as the flow. The oxygen requirement due to heterotrophic decay can be calculated with Equation 11.26, while the requirement for autotrophic decay can be calculated with Equation 11.34. Both should be apportioned to each tank in accordance with the fraction of the MLSS in each as given by Equation 11.43. Although the recycle ratio will influence that distribution, the effect will be relatively small so the distribution can be made on the basis of the most commonly used ratio. Because the MLSS concentration decreases from tank to tank down the chain, so will this oxygen requirement. The transient-state oxygen requirement can be distributed in a similar

manner by using the approaches developed for CMAS systems and considering the distribution of influent to all tanks.

For equal influent flow distribution to all tanks, the first tank will have the highest oxygen requirement, and thus it is used to determine the minimum acceptable bioreactor volume based on oxygen transfer. This can be calculated with Equation 11.5. Even though the last tank will have a somewhat lower oxygen requirement than the first, it should be used to calculate the lower limit based on floc shear because its effluent goes to the final settler or membrane system. That limit can be calculated with Equation 11.4. Both calculations should be made on the basis of the summer time oxygen requirement, with either the steady- or transient-state requirement being used, depending upon the nature of the influent flow. The last tank will have the lowest oxygen requirement, and thus it should be used to determine the maximum acceptable bioreactor volume. The upper limit on the size of the last tank can be calculated with Equation 11.3 by using the minimum steady-state or transient-state oxygen requirement (as needed) as calculated for winter operating conditions.

Once the limits on tank volume are known, they can be used with the mass of MLSS in the last tank to establish an acceptable range of MLSS concentrations for that tank. The mass of MLSS in the last tank is just the fraction of biomass in that tank multiplied by $(X_{M,T} \cdot V)_{System}$ as calculated with Equation 10.8 for winter conditions. Since the MLSS concentration in the last tank is the same as the concentration entering the final settler, various concentrations within the allowable range can be investigated for their effects on final settler size and operation. Once one is selected, it, in turn, establishes the total system volume, just as in the other designs.

Sometimes it may be desirable to alter the configuration of an existing activated sludge system to that of SFAS. In that situation, the planned change must be evaluated to be sure that it can be made while still accomplishing the treatment objectives. The steps for doing that are similar to those for designing a SFAS system, except that the system volume and number of equivalent tanks-in-series are known, thereby fixing the volume of each tank. Thus, the major questions are whether the desired effluent quality can be met with the existing system SRT and whether the needed amount of oxygen can be transferred to each tank while meeting the constraints on mixing energy input and oxygen transfer. The steps outlined above can be followed to answer those questions.

11.3.5.2 Design of Contact Stabilization Activated Sludge Systems

A schematic diagram of a CSAS system is shown in Figure 7.19. There it can be seen that the influent enters the contact tank, from which it flows directly to a final settler or membrane system. The biomass recycle from the settler flows to the stabilization basin where additional reactions can occur before the biomass is returned to the contact tank. As far as organic substrates are concerned, removal from the wastewater undergoing treatment must occur in the contact tank, where the majority of the soluble substrate is metabolized. Particulate organic matter is entrapped in the mixed liquor, with its degradation occurring in both the contact and stabilization tanks. As discussed in Section 7.4, one characteristic of CSAS is that partial nitrification can occur, with the extent depending on both the system SRT and the fraction of the biomass in the contact tank. These systems are seldom designed specifically to achieve partial nitrification, and thus the design approach presented here is based on biodegradable COD removal. However, because some nitrification will occur if the proper environmental conditions are achieved, its occurrence must be considered or the estimation of the amount of oxygen required in the system will be incorrect. While it is possible to derive solvable analytical equations for organic substrate removal with a few simplifying assumptions that are unlikely to be violated, the same is not true for nitrification. Only approximations can be achieved. This means that the only truly accurate way to estimate the degree of nitrification and the distribution of the associated oxygen requirement is through simulation with a model like one of the IWA activated sludge models. However, because hand calculations are very useful during preliminary design we will present an approach for using them to evaluate the degree of nitrification likely to occur. Their approximate nature should be recognized, however, and appropriate caution should be exercised in their use.

The first task in the design of a CSAS process is the selection of the system SRT, which requires the consideration of many factors as discussed previously. After that, the relative amount of biomass in the two tanks must be selected to ensure that the desired effluent quality is attained. This, in turn, requires selection of the biomass recycle ratio and the fraction of the system volume allocated to each tank. Because system performance depends on both of those factors, as illustrated in Figures 7.24–7.27, CSAS systems should not be built with separate vessels for the contact and stabilization tanks because such a design fixes their relative volumes. Rather, both should be in the same vessel, with a nonload-bearing, curtain wall between them. This will allow their relative sizes to be changed as circumstances require.

Selection of the fraction of biomass in the contact tank requires the following. The heterotrophic specific growth rate in the contact tank, $\mu_{H,C}$, must be consistent with the desired effluent readily biodegradable substrate concentration. Thus, as in SFAS design, Equation 11.29 can be used to calculate $\mu_{H,C}$ by setting S_{SF} equal to the desired effluent substrate concentration. The kinetic parameters should be those for the coldest expected operating condition since it will control. Following the same logic used to derive Equations 11.30 and 11.31, it can be shown that the specific growth rate in the contact tank is given by

$$\mu_{H,C} = \left[\frac{(X_{M,T} \cdot V)_{System}(1/\Theta_c + b_H)}{(X_{M,T,C} \cdot V_C)}\right]\left(\frac{S_{SO}}{S_{SO} + X_{SO}}\right), \tag{11.45}$$

when it is assumed that no utilization of slowly biodegradable substrate occurs. In Equation 11.45, V_C is the volume of the contact tank and $X_{M,T,C}$ is the MLSS concentration in it. Since the value of $\mu_{H,C}$ as calculated with Equation 11.45 must be less than or equal to the value of $\mu_{H,C}$ associated with the desired effluent substrate concentration as calculated with Equation 11.29, then

$$\frac{X_{M,T,C} \cdot V_C}{(X_{M,T} \cdot V)_{System}} = f_{XM,C} \geq \left(\frac{1/\Theta_c + b_H}{\mu_{H,C}}\right)\left(\frac{S_{SO}}{S_{SO} + X_{SO}}\right), \tag{11.46}$$

where $\mu_{H,C}$ has been calculated with Equation 11.29. In other words, as long as the fraction of MLSS in the contact tank, $f_{XM,C}$, is greater than or equal to the right side of Equation 11.46, the effluent quality will be acceptable. Since the system contains only two tanks, the remainder of the biomass is in the stabilization basin. The derivation of Equation 11.45 is based on the assumption that no slowly biodegradable substrate is used in the contact tank. Some hydrolysis will occur, however, allowing a portion of the slowly biodegradable substrate to be used. The effect of slowly biodegradable substrate utilization in the contact tank is to increase the required fraction of biomass in that tank, with the right parenthetical term in Equation 11.46 approaching one as all of the slowly biodegradable substrate is used there, an event that is unlikely to occur. Because of the uncertainty associated with the fraction of the slowly biodegradable substrate that will be used in the contact tank, a designer should calculate the required minimum value of $f_{XM,C}$ with Equation 11.46 twice, once as written and once with the right parenthetical term set to one. A design value for $f_{XM,C}$ between those two extremes should then be chosen.

The fraction of the biomass in the contact tank is determined solely by the system hydraulics. If both growth and wastage are neglected, as they were for the SFAS system, then

$$f_{XM,C} = \frac{\alpha \cdot \upsilon}{1 + \alpha - \upsilon}, \tag{11.47}$$

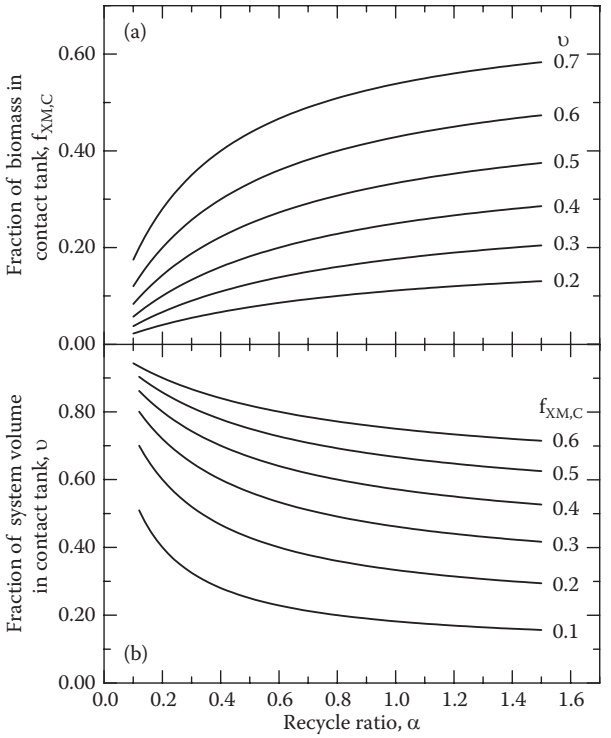

FIGURE 11.11 Effect of the biomass recycle ratio on: (a) the fraction of biomass in the contact tank and (b) the fraction of the system volume in the contact tank for a CSAS system, as calculated with Equation 11.47.

where α is the biomass recycle ratio and υ is the fraction of the system volume in the contact tank:

$$\upsilon = \frac{V_C}{V_C + V_S} = \frac{V_C}{V_T}, \tag{11.48}$$

where V_S is the volume of the stabilization basin and V_T is the total system volume. Many combinations of α and υ can result in the same value of $f_{XM,C}$, as shown in Figure 11.11, where Equation 11.47 is plotted in two ways. Figure 11.11a shows that for any value of υ, the fraction of MLSS in the contact tank will increase as the recycle ratio is increased. Thus, the selection of υ should be made for the smallest anticipated recycle ratio. Because of the criterion given in Equation 11.46, the needed effluent quality will be obtained for any larger value of the recycle ratio.

Example 11.3.5.2

Consider the wastewater that has been the subject of all of the examples in this chapter. Consideration is being given to using a CSAS process with an SRT of three days. The effluent quality objective is 10 mg/L as COD of readily biodegradable organic matter. What fraction of the biomass and what fraction of the system volume should be in the contact tank if the smallest recycle ratio that will be used is 0.30?

a. What specific growth rate will be required in the contact tank?
   Calculate the specific growth rate in the contact tank, $\mu_{H,C}$, using Equation 11.29 and the kinetic parameters for winter conditions, since they will control. The desired substrate concentration is 10 mg/L. Using the kinetic parameters from Table E11.2 gives:

$$\mu_{H,C} = (4.1)\left(\frac{10}{20+10}\right) = 1.37 \text{ days}^{-1}.$$

b. What is the smallest fraction of the MLSS that can be in the contact tank?
Use of $\mu_{H,C} = 1.37$ in Equation 11.46 gives the smallest fraction of the MLSS that can be in the contact tank.

$$f_{XM,C} \geq \left(\frac{1/3.0+0.15}{1.37}\right)\left(\frac{115}{115+150}\right) = 0.15.$$

Thus, assuming that no utilization of slowly biodegradable substrate occurs in the contact tank, as long as 15% or more of the MLSS is in that tank, the effluent quality goal can be met. Recalculation of the fraction with the right parenthetical term set to one shows that at least 35% of the MLSS would have to be in the contact tank if all of the slowly biodegradable substrate was used there. This is unlikely to occur, but to provide a factor of safety, choose a fraction of 25% for design.

c. What fraction of the system tank volume should be in the contact tank with a recycle ratio of 30%?
The required fraction of the system volume in the contact tank, υ, can be calculated with a rearranged form of Equation 11.47:

$$\upsilon = \frac{1+0.30}{1+0.30/0.25} = 0.59.$$

This value could also have been read from Figure 11.11b. Thus, 59% of the system volume must be in the contact tank to ensure that 25% of the system biomass is there. This large fraction is due to the short SRT of the system. Generally, the larger the SRT, the smaller the contact tank as a fraction of the total system.

After the fraction of biomass and the fraction of total system volume in the contact tank have been selected, the oxygen requirement must be distributed between the two reactors. This must be done for summer and winter conditions, and for the maximum and minimum anticipated recycle ratios. The minimum oxygen requirement will occur in the winter, and it will be used to determine the upper limit on the total system volume, just as with the other systems. Likewise, the summer conditions will give the maximum oxygen requirement, which determines the lower limit on the system volume, as discussed previously. Increasing the recycle ratio shifts some of the oxygen requirement from the stabilization basin to the contact tank, and also increases the oxygen requirement due to nitrification, as shown in Figures 7.24 and 7.25. Consequently, the extreme conditions must be examined so that the controlling situation can be identified.

The heterotrophic oxygen requirement can be distributed by a technique similar to that used to distribute the oxygen requirement in a CAS system, with special consideration of the unique characteristics of the CSAS system. The oxygen requirement for synthesis of biomass from readily biodegradable substrate can be calculated with Equation 11.24. All of it will occur in the contact tank. The oxygen requirement for synthesis of biomass from slowly biodegradable substrate can be calculated with Equation 11.25. Because slowly biodegradable substrate is entrapped in the MLSS where it undergoes hydrolysis, which is a slow reaction, it is likely to occur in both tanks. Although the relative amount of slowly biodegradable substrate utilization occurring in each tank can only be determined by simulation with a model like one of the IWA activated sludge models, in the absence of such simulations it seems reasonable to distribute the oxygen requirement to the two tanks in proportion to the mass of MLSS in them. The oxygen requirement due to heterotrophic biomass decay can be calculated with Equation 11.26. It, too, should be distributed in proportion

to the mass of MLSS in each tank. Because of the effects of the recycle ratio on the distribution of MLSS in the system, its extreme values should be considered. The maximum value will maximize the heterotrophic oxygen requirement in the contact tank and minimize it in the stabilization basin, whereas the minimum recycle ratio will have just the opposite effect.

Distribution of the oxygen requirement associated with nitrification is more difficult and cannot be done precisely with analytical equations. However, it is possible to approximate it for those situations in which substantial nitrification will occur; that is, when the system SRT is well in excess of the minimum SRT for nitrification. The rationale used to derive the equations is similar to that used to derive Equations 11.45 and 11.46, as well as 11.30 and 11.31, with additional criteria. First, nitrification will occur in the stabilization basin, making the concentration of ammonia-N entering the contact tank in the RAS flow approach zero. Second, when stable partial nitrification is occurring in the contact tank, it cannot be assumed that the mass of nitrogen oxidized in the stabilization tank is much less than the mass oxidized in the contact tank. In fact, it may be more. Furthermore, neither can it be assumed that the change in ammonia-N concentration across the contact tank is approximately equal to the change in concentration across an equivalent CMAS system, since more nitrification will occur in the latter. Rederiving Equation 11.31 with these changes, and with the assumption that all forms of nitrogen are equally available in both tanks, gives the following approximate expression for the fraction of autotrophs in the contact tank, $f_{XA,C}$:

$$f_{XA,C} \approx \frac{1/\Theta_c + b_A}{\mu_{A,C}}\left[1-\alpha\left(\frac{S_{NHC}}{S_{N,a}-S_{NHC}}\right)\right], \tag{11.49}$$

where $\mu_{A,C}$ is the specific growth rate of the autotrophs in the contact tank as calculated with Equation 11.29 using S_{NHC}, which is the ammonia-N concentration in that tank. The fraction of autotrophs in the contact tank is the same as the fraction of the MLSS in the contact tank, $f_{XM,C}$, because the mixed liquor is homogeneous throughout the system. Substitution of Equation 11.29 into Equation 11.49 and rearrangement gives a quadratic equation allowing the effluent ammonia-N concentration, S_{NHC}, to be calculated for any given biomass recycle ratio, α, with its associated value of $f_{XM,C}$, as determined by Equation 11.47:

$$\begin{aligned}&\left(1+\alpha-\frac{f_{XM,C}\cdot\hat{\mu}_A}{1/\Theta_c+b_A}\right)S_{NHC}^2\\&+\left[K_{NH}(1+\alpha)-S_{N,a}\left(1-\frac{f_{XM,C}\cdot\hat{\mu}_A}{1/\Theta_c+b_A}\right)\right]S_{NHC}-K_{NH}\cdot S_{N,a}=0,\end{aligned} \tag{11.50}$$

where $S_{N,a}$ is the available nitrogen concentration, as given by Equation 11.17. The value obtained from Equation 11.50 is only approximate because of the approximate nature of Equation 11.49. Use of the highest anticipated value of α in the solution of Equation 11.50 will minimize the ammonia-N concentration in the contact tank, which will maximize the oxygen requirement associated with nitrification in that tank, while minimizing the oxygen requirement in the stabilization basin. Use of the lowest anticipated recycle ratio will have just the opposite effect.

Once the ammonia-N concentration in the contact tank has been estimated, the oxygen requirement due to nitrification can be apportioned to the two vessels. Because particulate organic nitrogen will have a different fate than soluble nitrogen, a distinction must be made between the two, even though that was not done in the derivation of Equation 11.49. Let $S'_{N,a}$ be the available soluble nitrogen:

$$S'_{N,a} = S_{N,a} - X_{SNO}. \tag{11.51}$$

Then the oxygen requirement due to synthesis of autotrophic biomass from soluble nitrogen in the contact tank, $RO_{A,SN,C}$, can be calculated from a modified form of Equation 11.33:

$$RO_{A,SN,C} = F(S'_{N,a} - S_{NHC})(4.57 - Y_{A,T}i_{O/XB,T}). \quad (11.52)$$

The oxygen requirement in the stabilization tank associated with synthesis of autotrophic bacteria from soluble nitrogen, $RO_{A,SN,S}$, is then

$$RO_{A,SN,S} = \alpha \cdot F \cdot S_{NHC}(4.57 - Y_{A,T}i_{O/XB,T}). \quad (11.53)$$

The value of α used in Equation 11.53 should be consistent with the value used to solve Equation 11.50. The total oxygen requirement associated with synthesis of autotrophic biomass from particulate organic nitrogen, $RO_{A,XNS}$, can be estimated as

$$RO_{A,XNS} = F \cdot X_{NSO}(4.57 - Y_{A,T}i_{O/XB,T}). \quad (11.54)$$

It is apportioned to each of the vessels in proportion to the fraction of biomass in each. That apportionment will depend on the recycle ratio, as discussed previously. The oxygen requirement due to decay of autotrophs can be ignored because its contribution is less than the uncertainty associated with the other determinations.

Transient-state oxygen requirements can be estimated by combining the techniques presented in Section 11.3.2 with those presented above. As with the other multitank activated sludge systems, there will be considerable uncertainty associated with the estimates because of the need to distribute the transient oxygen requirement spatially. Consequently, simulation with a model like ASM No. 1 is a much more certain way of estimating such requirements.

Example 11.3.5.3

Continue considering the CSAS system whose analysis was begun in Example 11.3.5.2. What would the summer time steady-state oxygen requirement be if 59% of the system volume was in the contact tank and the system was being operated with a recycle ratio of 0.30? As determined in Example 11.3.5.2, the contact tank of such a system would contain 25% of the MLSS.

a. What is the heterotrophic oxygen requirement in each tank?
The oxygen requirement for heterotrophic synthesis from readily biodegradable substrate can be determined with Equation 11.24. Since the SRT is the same as in all previous examples, this computation is the same as the one made in part a of Example 11.3.4.1. Thus, $RO_{H,SS}$ = 1840 kg O_2/day. All of this will occur in the contact tank.

The oxygen requirement for heterotrophic synthesis from slowly biodegradable substrate can be determined with Equation 11.25. This computation is the same as the one made in part b of Example 11.3.4.1. Thus, $RO_{H,XS}$ = 2400 kg O_2/day. Since the contact tank contains 25% of the biomass, 25% of this oxygen requirement will occur in it and 75% in the stabilization tank. Thus:

$$RO_{H,XS,C} = (0.25)(2400) = 600 \text{ kg } O_2/\text{day}$$

and

$$RO_{H,XS,S} = (0.75)(2400) = 1800 \text{ kg } O_2/\text{day}.$$

The oxygen requirement associated with heterotrophic decay can be estimated with Equation 11.26. This was done in part c of Example 11.3.4.1. Thus, $RO_{H,D} = 2020$ kg O_2/day. It should also be distributed to the two vessels in proportion to the fraction of the MLSS that each contains. Thus:

$$RO_{H,D,C} = (0.25)(2020) = 505 \text{ kg } O_2/\text{day}$$

and

$$RO_{H,D,S} = (0.75)(2020) = 1515 \text{ kg } O_2/\text{day}.$$

Summing these values gives the heterotrophic oxygen requirement in each tank:

$$RO_{H,C} = 1840 + 600 + 505 = 2945 \text{ kg } O_2/\text{day}$$

and

$$RO_{H,S} = 1800 + 1515 = 3315 \text{ kg } O_2/\text{day}.$$

b. What is the autotrophic oxygen requirement in each tank?
The first task is to estimate the concentration of ammonia-N leaving the contact tank, S_{NHC}, using Equation 11.50. In that equation, $\alpha = 0.30$, $f_{XM,C} = 0.25$, $\hat{\mu}_A = 1.3$ day^{-1}, $\Theta_c = 3$ days, $b_A = 0.12$ day^{-1}, $K_{NH} = 1.9$ mg N/L, and $S_{N,a} = 30.5$ mg N/L. Setting up the quadratic equation gives:

$$0.583\, S_{NHC}^2 - 6.162\, S_{NHC} - 57.950 = 0.$$

The solution of this equation reveals that S_{NHC} is 16.6 mg /L as N. This compares to a concentration of 2.1 mg/L from the CMAS system (see Example 11.3.3.2). This difference is consistent with the differences between the two systems shown in Figure 7.20.

The oxygen requirement from oxidation of soluble nitrogen in the contact tank, $RO_{A,SN,C}$, can be calculated with Equation 11.52 after $S'_{N,a}$ has been estimated with Equation 11.51 using the value of $S_{N,a}$ of 30.5 calculated in Example 11.3.3.2:

$$S'_{N,a} = 30.5 - 8.5 = 22.0 \text{mg/L as N}$$

and

$$RO_{A,SN,C} = (40{,}000)(22.0 - 16.6)[4.57 - (0.20)(1.20)]$$
$$= 935{,}000 \text{ g } O_2/\text{day} = 935 \text{ kg } O_2/\text{day}.$$

The oxygen requirement associated with oxidation of soluble nitrogen in the stabilization basin, $RO_{A,SN,S}$, can be calculated with Equation 11.53 using a value of 16.6 mg/L as the ammonia-N concentration entering the basin through the RAS flow, which is 30% of the system influent flow:

$$RO_{A,SN,S} = (0.30)(40{,}000)(16.6)[4.57 - (0.2)(1.20)]$$
$$= 862{,}000 \text{ g } O_2/\text{day} = 862 \text{ kg } O_2/\text{day}.$$

The oxygen requirement from oxidation of particulate organic nitrogen, $RO_{A,XNS}$, can be calculated with Equation 11.54:

$$RO_{A,XNS} = (40{,}000)(8.5)[4.57 - (0.2)(1.20)]$$
$$= 1{,}470{,}000 \text{ g } O_2/\text{day} = 1470 \text{ kg } O_2/\text{day}.$$

Since 25% of the biomass is in the contact tank, 25% of this oxygen requirement will be exerted there and 75% in the stabilization basin. Consequently, the oxygen requirement from oxidation of particulate organic nitrogen in the two vessels is

$$RO_{A,XNS,C} = (0.25)(1470) = 368 \text{ kg } O_2/\text{day}$$

and

$$RO_{A,XNS,S} = (0.75)(1470) = 1102 \text{ kg } O_2/\text{day}.$$

Summing these values gives the autotrophic oxygen requirement in each tank:

$$RO_{A,C} = 935 + 368 = 1303 \text{ kg } O_2/\text{day}$$

and

$$RO_{A,S} = 862 + 1102 = 1964 \text{ kg } O_2/\text{day}.$$

c. What is the total oxygen requirement in each tank?
The total oxygen requirement in each tank is obtained by summing the heterotrophic and autotrophic requirements:

$$RO_C = 2945 + 1303 = 4248 \text{ kg } O_2/\text{day}$$

and

$$RO_S = 3315 + 1964 = 5279 \text{ kg } O_2/\text{day}.$$

The total oxygen requirement in the whole system is 9530 kg O_2/day, which is less than that required by the CMAS system as calculated in Example 11.3.3.2. The difference is due to the lower degree of nitrification in the CSAS system.

The contact tank comprises 59% of the system volume and experiences 45% of the oxygen requirement, whereas the stabilization basin contains 41% of the system volume and experiences 55% of the oxygen requirement. These percentages are specific to the operating conditions imposed and depend on the SRT, the recycle ratio, and the fraction of system volume in each tank, as illustrated in Figures 7.25 and 7.27.

After the oxygen requirements have been estimated for winter and summer and for both the maximum and minimum planned recycle ratios, the next task in the design of a CSAS system is to determine the total bioreactor volume, thereby fixing the volume of each tank, the MLSS concentration in each, and the final settler size. This is done in the same way as for the other activated sludge designs. An extra level of complexity is involved, however, because it is not apparent before the exercise whether the contact tank or the stabilization basin will control the design. For instance, in the example above, the stabilization basin had the greater oxygen requirement per unit volume, with 55% of the oxygen requirement in 41% of the volume. Thus, for that operational situation (summer with minimum recycle flow), the stabilization basin will control the minimum system volume. The engineer must consider all of the possible scenarios to determine which actually control the upper and lower limits on the system volume. Once those volumes have been determined, the range of permissible MLSS concentrations is fixed by the mass of MLSS in

the system as calculated with Equation 10.8. Finally, that information is used with the anticipated recycle ratios and projections of solids settling properties to arrive at an economic combination of bioreactor and settler sizes.

11.3.6 Batch Reactors—Sequencing Batch Reactor Activated Sludge

As discussed in Section 7.8, batch processes offer opportunities for flexibility that can be advantageous in some circumstances. For example, when flows are highly variable or when the character of the contaminants changes on a regular and periodic basis, batch processes allow those special needs to be met. Furthermore, by changing the length of the fill period within a cycle, batch processes can be made to behave like continuous flow processes with hydraulic characteristics anywhere between a perfect CSTR and a perfect plug-flow reactor (PFR). Consequently, sequencing batch reactor activated sludge (SBRAS) systems have found increasing popularity in recent years.

Although SBRAS systems are often treated as if special design techniques are required, they actually operate according to the same principles as other activated sludge processes, as discussed in Section 7.8.2, and can be designed according to them. Consequently, the mass of MLSS in the system is defined by Equation 10.8, the overall daily solids wastage rate is given by Equation 10.9, the total daily heterotrophic oxygen requirement is given by Equation 10.10, and the total daily autotrophic oxygen requirement by Equation 11.16, provided that the effective SRT is used in their computation. The three primary differences between SBRAS and other activated sludge systems are that the oxygen requirement must be distributed to each of the discrete operating cycles and then apportioned within each cycle in a manner consistent with the length of the fill period, the fraction of each operating cycle that is not devoted to biological reaction must be accounted for in the design, and the interaction between the bioreactor and the secondary clarifier must be analyzed differently because biological reaction and sedimentation occur in the same vessel (although at different times in the cycle). The first of these can be accomplished by the same techniques used to apportion steady- and transient-state oxygen requirements in the various continuous-flow activated sludge systems. The second can be accounted for through use of the effective HRT, τ_e, and the effective SRT, Θ_{ce}, when using the CMAS equations listed above. The third can be considered through simple procedures to be presented below.

The effective HRT and effective SRT account for the portion of the SBRAS cycle that is not utilized for biological reaction (i.e., that portion devoted to settling and decanting, draw), as illustrated in Figure 1.4. In Section 7.8.2, ζ is defined as the fraction of the total cycle devoted to fill plus react. Using those definitions, the effective HRT is defined as

$$\tau_e = \frac{\zeta \cdot V}{F}, \tag{7.8}$$

and the effective SRT is defined as

$$\Theta_{ce} = \frac{\zeta \cdot V \cdot X_M}{F_w \cdot X_{M,w}}. \tag{7.9}$$

Selection of the effective SRT is governed by the same considerations as the selection of the SRT for any other activated sludge system. Relatively low values can be used if the removal of biodegradable organic matter is the primary objective. However, SBRAS systems are often used in small wastewater treatment plants where sludge stabilization is also an important consideration. Thus, effective SRTs of 10 days or more are often used.

Selection of the system volume, and hence the effective HRT, must consider not only mixing and oxygen transfer, but also the anticipated settling properties of the MLSS. That is necessary because

the bioreactor volume must be sufficient to contain not only the flow added per cycle, F_c, but the volume retained after effluent is decanted to contain the recycled biomass, V_{br}, where

$$F_c = \frac{F}{N_c} \tag{7.11}$$

and

$$V_{br} = \alpha \cdot F_c. \tag{7.12}$$

As discussed in Section 7.8.2, N_c is the number of cycles per day and α is analogous to the recycle ratio in a continuous flow system. Consequently, when an SBRAS system is being designed as a simple activated sludge system with no nutrient removal capabilities:

$$V = F_c(1+\alpha) \tag{11.55}$$

or

$$V = F_c + V_{br}. \tag{11.56}$$

The primary factors considered in the selection of both N_c and ζ are adding influent wastewater and removing treated effluent, allowing a sufficient time for settling and decanting in the operating cycle, and allowing for peak hydraulic flows through the bioreactor. These factors, in turn, are affected by the total number of bioreactors selected and the desired operational cycle. Values of N_c often range from 4 to 6 cycles per day. Since the time required for settling and decanting is relatively constant, whereas the length of a cycle gets shorter as the number of cycles per day is increased, the value of ζ decreases as N_c increases. Nevertheless, values of ζ often range from 0.5 to 0.7. Considerable flexibility exists in the selection of both parameters.

The minimum bioreactor volume associated with a given number of cycles per day will result when V_{br} is just large enough to contain the MLSS retained for use in the next cycle after the effluent has been decanted. The mass of MLSS in the bioreactor after solids settling and effluent decanting is the same as the mass of MLSS at the end of the react period, which is given by Equation 10.8. Therefore:

$$X_{M,T,r} \cdot V_{br} = (X_{M,T} \cdot V)_{System}, \tag{11.57}$$

where $X_{M,T,r}$ is the settled (retained) MLSS concentration. Thus, it can be seen that when the settled MLSS concentration is as large as possible, the retained volume, V_{br}, will be as small as possible. Although experience with treatment of a particular wastewater is the best way to select the maximum achievable value of $X_{M,T,r}$, it can also be estimated from the definition of the sludge volume index, SVI. As discussed in Section 2.3.1, the SVI is defined as the volume in mL occupied by a gram of solids after 30 minutes of quiescent settling. If we take the SVI as being indicative of the highest concentration to which the MLSS can be settled, then the maximum attainable settled solids concentration, $X_{M,T,r,max}$ (in mg/L), will be given by

$$X_{M,T,r,max} \approx \frac{10^6}{SVI}. \tag{11.58}$$

It follows from Equation 11.57 that the smallest retained volume, $V_{br,min}$, is given by

$$V_{br,min} = \frac{(X_{M,T} \cdot V)S_{System}}{X_{M,T,r,max}}. \tag{11.59}$$

Consequently, the lower limit on the SBR volume, V_L, is obtained by substituting Equation 11.59 into Equation 11.56:

$$V_L = F_c + \frac{(X_{M,T} \cdot V)_{System}}{X_{M,T,r,max}}. \tag{11.60}$$

Although care should be exercised in the selection of $X_{M,T,r,max}$, SBRAS systems offer some flexibility in the control of the MLSS settling properties. Recall that SBRs can achieve any hydraulic characteristic between a CSTR and a PFR simply by changing the length of the fill period. This means that the instantaneous process loading factor can be changed simply by changing the length of the fill period. Since the process loading factor influences the competition between filamentous and floc-forming bacteria, as discussed in Section 11.2.1, changing the length of the fill period allows control of the settling properties, provided that the oxygen transfer system is capable of meeting the imposed oxygen requirement. Thus, steps can be taken during operation to ensure that the selected SVI is achieved, as long as a realistic value was selected during design.

With the above information in mind, we can now set forth the steps in the design of an SBR system. After selection of the effective SRT, the mass of MLSS in the system is calculated with Equation 10.8. The number of cycles per day is then selected, taking into account the length of time in each cycle to be devoted to settling and decanting. Since the flow to be treated is known, this fixes both F_c and ζ. The estimation of an attainable SVI value allows computation of $X_{M,T,r,max}$ with Equation 11.58, which allows calculation of the minimum possible bioreactor volume with Equation 11.60. Selection of the design bioreactor volume requires consideration of the oxygen requirement and the constraints on the rate of oxygen transfer and floc shear, just as with all of the other activated sludge systems. Equations 10.10 and 11.16 can be used to calculate the total daily oxygen requirement. While the mass of oxygen required per cycle will just be the total daily requirement divided by the number of cycles per day, the oxygen transfer rate will depend on the length of the fill period during that cycle. If the fill period lasts throughout the entire react cycle, then the system will behave like a CSTR and the oxygen transfer rate will be equivalent to that in a CMAS system, allowing the techniques for that system to be used. On the other hand, the shorter the fill period relative to the react period, the more the system will behave like a PFR, requiring the techniques used during CAS design to be used. No matter which technique is employed, however, the design reactor volume must be selected to ensure proper mixing and oxygen transfer. Floc shear is seldom a problem with an SBRAS process because flocculation will generally occur later in the react period when aeration rates are lower. Suspension of solids is not a problem in facilities that provide separate mixing equipment. Nevertheless, the designer should verify that these constraints will be satisfied either by calculation or logical analysis. After the volume is known, the anticipated MLSS concentration can be calculated by dividing $(X_{M,T} \cdot V)_{System}$, as calculated with Equation 10.8, by the selected bioreactor volume. The effective HRT can then be determined with Equation 7.8 and the flow rate of waste solids can be calculated with Equation 7.9 after it has been determined when in the cycle solids will be wasted, thereby determining $X_{M,T,w}$. The required solids mass wastage rate is also given by Equation 10.9, so a system check can be made. The following example illustrates the unique aspects of SBRAS design.

Example 11.3.6.1

An SBRAS facility is to be constructed to treat the wastewater considered throughout Section 11.3. A partial degree of sludge stabilization is important, so an effective SRT of 10 days is to be maintained. Four cycles per day are to be utilized at average flow to allow more cycles to be used to treat peak flows. A value of ζ of 0.5 will be used for design purposes, again to allow operational flexibility during peak flows. Experience indicates that the SVI in the process will generally not exceed 120 mL/g. What is the minimum possible bioreactor volume that could be used and what MLSS concentration would be associated with its use? What effective HRT would the SBRAS have at average flow?

a. What value of $(X_{M,T} \cdot V)_{System}$ should be used for design purposes?
The $(X_{M,T} \cdot V)_{System}$ is calculated using Equation 10.8 at the winter operating temperature of 15°C. Values of all of the parameters in Equation 10.8 are given in Table E11.2. Using these values and $\Theta_c = 10$ days:

$$(X_{M,T} \cdot V)_{System} = (10.0)(40{,}000) \cdot \left\{43.5 + \frac{[1+(0.20)(0.15)(10.0)](0.5)(115+150)}{1+(0.15)(10.0)}\right\}$$
$$= 45{,}000{,}000 \text{ g MLSS.}$$

b. What value of $X_{M,T,r,max}$ is appropriate for this design?
As noted above, the SVI for this application is expected to be less than 120 mL/g. Using Equation 11.58:

$$X_{M,T,r,max} = \frac{10^6}{SVI} = \frac{10^6}{120} = 8300 \text{ mg/L} = 8300 \text{ g/m}^3.$$

c. What is the minimum possible bioreactor volume?
Since there will be four cycles per day, and the average daily flow rate is 40,000 m^3/day, the flow per cycle, F_c, is 10,000 m^3. Substitution of this value into Equation 11.60 gives:

$$V_L = 10{,}000 + \frac{45{,}000{,}000}{8300} = 15{,}400 \text{ m}^3.$$

d. What is the bioreactor MLSS concentration?
This is calculated by dividing the mass of MLSS in the system, as calculated in part a, by the bioreactor volume:

$$X_{M,T} = \frac{45{,}000{,}000}{15{,}400} = 2900 \text{ g/m}^3 = 2900 \text{ mg/L}.$$

e. What is the effective HRT of the bioreactor.
From Equation 7.8:

$$\tau_e = \frac{\zeta V}{F} = \frac{(0.5)(15{,}400)}{40{,}000} = 0.19 \text{ day}.$$

11.3.7 Process Optimization Using Dynamic Models

As discussed in Section 10.4.3, dynamic simulation is an important tool that allows an engineer to refine the design of an activated sludge system or to evaluate alternative operating strategies for an existing facility. The necessity for using this tool increases as the system to be designed is further removed from the assumptions inherent in the simple stoichiometric models that are the basis for the analytical expressions in this chapter. For example, if a CMAS system were to be used to treat a wastewater flowing at a constant rate and containing a constant concentration of soluble pollutants, the approach presented in this chapter would be very accurate because the assumptions in the model are consistent with the nature of the problem. In fact, the equations would even do a very reasonable job under dynamic loading conditions. If part of the pollutants were particulate, on the other hand, the equations would still be accurate for a CMAS system operated under constant loading conditions, but they would be more approximate as dynamic loads were applied because they contain no rate expressions for hydrolysis of particulate substrates. However, by using approximations based on experience, it is still possible for the approach to give adequate information about transient-state oxygen requirements in a single tank system. The approaches presented in this chapter are weakest for design of multitank systems treating wastewaters containing both soluble and particulate pollutants, regardless of the nature of the influent flow. This is because the apportionment of the oxygen requirement among the various tanks requires assessment of the rates of degradation of both soluble and particulate constituents, information that is not incorporated into the simple model upon which the approaches are based. However, experience can help the engineer make the decisions required, although the information will always be approximate. Consequently, as the system configuration becomes more complex and as the influent conditions become more dynamic, the engineer must exercise more caution in the application of the approaches presented and should rely more and more on experience to make decisions.

Given the situation described in the preceding paragraph, what should be done when the experience base is small, either for an individual engineer or for the profession? The answer to that question is to rely more on dynamic simulation. As discussed in Part II of this text, the IWA activated sludge models adequately represent a number of suspended growth biological treatment systems, and are particularly effective for activated sludge systems of the type presented here. Furthermore, software packages implementing them are readily available, as indicated in Table 6.4. Using the techniques described in Chapter 9, the parameters in ASM No. 1 can be assessed with sufficient accuracy to allow it to adequately mimic the performance of real systems.[7] Therefore, an engineer can use the approaches presented in this chapter to decide on tentative sizes for the various bioreactors in an activated sludge system, and then use simulation to investigate the oxygen requirement in each vessel in the system under a variety of anticipated dynamic loading scenarios. That output can then be used to evaluate the ability to transfer the needed oxygen while meeting the constraints on floc shear and mixing, allowing modification of the design as needed. Additional rounds of simulations can then be used to further refine the design. Because the models are known to reflect reality,[7] the engineer can have more confidence in the proposed design than he/she could have in a design based on the approximate approaches presented here. This allows smaller factors of safety to be used, resulting in more economic designs, and so on.

In addition to the benefits presented above, simulation allows a neophyte engineer to build an experience base in a relatively short time. By investigating a variety of activated sludge variations by simulation, and comparing the output to the approximations obtained through the analytical equations, an engineer can quickly learn how much uncertainty is associated with the approximate analytical approach for a given configuration. That information then becomes part of the engineer's experience base, which allows better use to be made of the approximate techniques when they are suitable. Returning to the concepts presented in Chapter 10, a good engineer will use the tool that is appropriate for the job at hand. Simulation is one of those tools, but it should not be viewed as

something that is only justified for large and complex systems. Rather, the ease with which it can be done suggests strongly that it should be used to learn more about systems, just as it was used in Chapters 6 and 7, thereby increasing the engineer's experience base. Thus, it should be thought of as an important extension of this book.

11.4 PROCESS OPERATION

Control of activated sludge systems is accomplished by the application of the principles described throughout this book. While common sense and a good understanding of the process are keys to successful activated sludge operation, some specialized techniques have been developed. These specialized techniques are discussed in this section.

11.4.1 Solids Retention Time Control

Throughout this book, we have emphasized the importance of SRT in determining the performance and operating characteristics of suspended growth bioreactors. Consequently, control of the SRT is the key to achieving reliable, consistent performance from activated sludge systems.[35] However, it is important to recognize that control of the SRT is a long-term control strategy, allowing the system to respond to long-term changes in process loadings and to seasonal differences in required operating conditions. This is because a time equivalent to 2 to 3 SRTs will be required before a performance change is observed as a result of a change in the SRT.

11.4.1.1 Determination of Solids Wastage Rate

The equation defining SRT (Equation 5.1) was based on the assumption that the biomass separator (i.e., the final clarifier or membrane system) was perfect, so that no biomass was lost in the final effluent. While this assumption can be considered to be true for membrane systems, all pilot- and full-scale clarifiers lose some suspended solids to the effluent, and this must be accounted for in the computation of the SRT for systems employing clarification for biomass separation. In that case, Equation 9.1 must be used to calculate the SRT:

$$\Theta_c = \frac{V \cdot X_{M,T}}{F_w \cdot X_{M,T,w} + (F - F_w) X_{M,T,e}}. \tag{9.1}$$

Examination of Equation 9.1 reveals that suspended solids are lost from the process in two ways: (1) in the WAS ($F_w \cdot X_{M,T,w}$) and (2) in the process effluent (($F–F_w$)$X_{M,T,e}$). Loss of solids in the WAS is intentional wastage, while loss in the process effluent is unintentional wastage. Under certain circumstances, unintentional wastage can represent a sizable proportion of the total solids wastage from the system, and exclusion of it from the SRT calculation can result in a significant error.[50] For the more general situation in which the MLSS concentration is not uniform throughout the activated sludge system, the following expression applies when wastage is from a single point:

$$\Theta_c = \frac{\sum_{i=1}^{n} V_i \cdot X_{M,T,i}}{F_w \cdot X_{M,T,w} + (F - F_w) X_{M,T,e}}. \tag{11.61}$$

In this case, the volume V_i with its associated MLSS concentration, $X_{M,T,i}$, may represent a distinct bioreactor, such as in CSAS, or it may represent a zone within a bioreactor, such as in SFAS.

Rearranging Equation 11.61 provides a general expression from which required sludge wastage rate can be calculated:

$$F_w = \frac{\dfrac{\left(\sum_{i=1}^{n} V_i \cdot X_{M,T,i}\right)}{\Theta_c} - (F - F_w)X_{M,T,e}}{X_{M,T,w}}. \tag{11.62}$$

Because the SRT is expressed in days, F_w is the volume of sludge that must be wasted every day, expressed as an average daily flow rate. Solids may be wasted continuously at rate F_w, or they may be wasted discontinuously so that the total volume wasted in a day equals F_w. Continuous wastage maintains a more constant solids inventory.

Two different values of the SRT can be calculated, depending on whether the activated sludge in the clarifier is included in the calculation. Unless denitrification occurs, resulting in sludge flotation and the uncontrolled loss of suspended solids in the clarifier effluent, the solids in the clarifier do not influence the performance of an activated sludge system.[58] That is why they were excluded from the definition of SRT used in Part II of this book. However, in practice, inclusion of the solids within the clarifier is necessary to achieve proper control over the solids inventory within the activated sludge system. For those systems in which the mass of solids in the clarifier is typically a small portion of the total system inventory, exclusion of the clarifier from the SRT calculation is acceptable. On the other hand, for those systems in which the solids in the clarifier are a significant and/or variable portion of the total system inventory, the clarifier must be included in the SRT calculation. In this case, the SRT based on the solids within the bioreactor alone must also be calculated and monitored to ensure that sufficient biomass is in contact with the incoming wastewater to achieve the necessary treatment.

Control of the SRT provides the required mass of biomass in an activated sludge system. For systems in which the MLSS concentration is not uniform throughout, such as CSAS and SFAS, it is important that the solids be distributed appropriately to achieve the desired performance. This requires adjustment of the solids recycle ratio, or the relative volumes allocated to the various zones, to achieve the desired specific growth rate at the point of effluent discharge to the clarifier, as discussed in Chapter 7 and Section 11.3.

Because maintenance of the SRT is a long-term control strategy, it is typically done by using running averages, in which the desired SRT itself is the time period over which the averages are calculated.[69] Two techniques are typically used. In one, values of the SRT are calculated on a daily basis and used to calculate the running average. In the other, running averages are calculated for the various solids concentrations and flow rates, which are measured on a daily basis, and those running averages are used to calculate the SRT. Examples of both approaches can be found in activated sludge operations manuals.[73,75] The fact that the SRT provides long-term control of the activated sludge process is reflected in the typical rule-of-thumb that no operating parameter should be changed by more than 20% on any given day. This suggestion is particularly appropriate for the solids wastage rate.

Although continuous wastage of activated sludge maintains the most constant solids inventory, it is not practiced at all plants. Rather, wastage is done discontinuously, for a variety of reasons. In that case, the solids wasting frequency should be selected relative to the SRT so that it does not result in more than a 5 to 10% variation in the MLSS concentration. Thus, if a plant is being operated at a short SRT, solids wastage should occur several times a day, such as once a shift. On the other hand, if the SRT is long, it may be possible to waste solids only once a day while still keeping the variation in MLSS concentration less than 10%.

When solids are not wasted continuously, care must be exercised to ensure that the measured waste solids concentration is truly representative. Collection of a composite sample of the waste activated sludge is critical because the solids concentration during the period when wastage is occurring may be quite different from the daily average solids concentration at the wastage location. This can be the case when solids are wasted from the RAS stream and RAS flow rates are not proportional to the influent flow rate. It is especially true if wasting is done from a solids hopper within the clarifier. Due to the critical nature of the waste solids concentration in determining the SRT, waste activated sludge sampling and analysis procedures should be verified for each system.

11.4.1.2 Solids Retention Time Control Based on Direct Analysis of Mixed Liquor Suspended Solids Concentration

Using this approach, the mass of suspended solids in the activated sludge system and the mass wasted intentionally and unintentionally are measured directly. The SRT is then calculated from the measured values. Flow rates are measured on a continuous basis, and daily totalized values are determined. Those values give the average daily flow rate for each relevant stream. Composite samples are collected from each stream, and the suspended solids concentration is measured for each. The MLSS concentrations are typically expressed on a TSS basis, but they can also be measured and expressed on a COD or VSS basis. The basis of measurement has no impact on the calculation of the SRT as long as only one basis is used. This can be seen by examining Equations 11.61 and 11.62.

A major problem associated with SRT control based on MLSS analysis is that information about the waste solids concentration is not available until after the solids have been wasted, making it necessary to base the wastage calculation on an estimate of that concentration. There are two reasons for this. One is the need to use a composite sample of the waste solids stream, as discussed above. The other is the lag time required for a suspended solids analysis. Time is required to take a sample, filter it, dry it, and weigh it. As a consequence, estimates of the suspended solids concentrations required in Equations 9.1, 11.61, and 11.62 are historical, not current, which is one reason running averages are usually used in the computation. This is another reason for never changing the solids wastage rate by more than 20% in any given day.

11.4.1.3 Solids Retention Time Control Based on Centrifuge Analysis of Mixed Liquor Suspended Solids Concentration

Centrifuge analysis is often used to reduce the lag time associated with MLSS analysis.[73,75] A sample is placed in a graduated centrifuge tube and centrifuged for a standard time at a standard speed, after which the volume of the resulting solids pellet is determined. Suspended solids concentrations are measured on parallel samples, and a relationship between the solids pellet volume in the centrifuge tube and the MLSS concentration of the sample is developed. This relationship can then be used as a calibration curve for estimation of the MLSS concentration when the centrifuge analysis is applied to other samples. The benefit of this approach is that it can be accomplished very quickly, thereby reducing the lag time between sample collection and analysis. It also reduces the number of time consuming MLSS analyses that must be conducted, although periodic analyses are required to verify the calibration.

11.4.1.4 Hydraulic Control of Solids Retention Time

Hydraulic control of the SRT utilizes relationships between the suspended solids concentrations in the bioreactor and the wastage to eliminate suspended solids concentrations from the SRT calculation.[28,71] As a consequence, only the bioreactor volume and process flow rates are required to calculate the SRT for activated sludge process variations with uniform MLSS concentrations. If the

mass flow rate of suspended solids in the effluent is negligible compared to the mass flow rate of suspended solids in the wastage, then Equation 9.1 may be simplified to

$$\Theta_c \approx \frac{V \cdot X_{M,T}}{F_w \cdot X_{M,T,w}}. \tag{11.63}$$

This expression is equivalent to Equation 5.1, except that an approximately equal to sign has been used to emphasize that the expression is approximate and depends on the validity of the assumption concerning the effluent suspended solids. For the Garrett wastage approach, in which solids are wasted directly from the bioreactor, $X_{M,T,w}$ is equal to $X_{M,T}$. Substituting this into Equation 11.63 gives an expression equivalent to Equation 5.2:

$$\Theta_c \approx \frac{V}{F_w}. \tag{11.64}$$

For the conventional wastage approach, in which solids are wasted from the clarifier underflow, $X_{M,T,w}$ is related to $X_{M,T}$ through the RAS flow rate, F_r. Substituting the appropriate relationship into Equation 11.63 gives an expression equivalent to Equation 5.82:

$$\Theta_c \approx \frac{V}{F_w}\left(\frac{F_w + F_r}{F + F_r}\right). \tag{11.65}$$

Since only bioreactor volume and process flow rates are included in Equations 11.64 and 11.65, suspended solids concentrations need not be measured to allow the SRT to be calculated on an approximate basis. The term hydraulic control of the SRT is based on the fact that only flow rates must be set to control the SRT. Thus, for the Garrett wastage approach:

$$F_w \approx \frac{V}{\Theta_c}, \tag{11.66}$$

and for the conventional wastage approach:

$$F_w \approx \frac{V \cdot F_r}{(F + F_r)\Theta_c - V}, \tag{11.67}$$

which is equivalent to Equation 5.83, except for the approximately equal to sign.

This approach offers significant potential for simplifying activated sludge operation and the concepts have been applied to systems with uniform MLSS concentrations.[51] Use is made of a calibration chart to account for suspended solids in the process effluent and consideration is given to the constraints imposed by the thickening limitations of the clarifiers. The concepts can also be used with activated sludge systems with nonuniform MLSS concentrations.

11.4.2 Qualitative Observations

Qualitative observations provide valuable information on the actual operating conditions in activated sludge systems, thereby helping operators to make decisions about alterations in process control parameters.[73,75,80] They are a necessary adjunct to long-term SRT control and application of the process fundamentals described above.

11.4.2.1 Bioreactor

Important information can be gained from observing the color and appearance of the biomass in an activated sludge bioreactor. The mixed liquor in a well-operating bioreactor will be brown in color, and a small amount of fresh, crisp, white foam will be present on the liquid surface. An earthy, musty aroma will be present that represents normal biological degradation products. A black color, or the presence of the "rotten egg" odor of hydrogen sulfide, indicates that inadequate aeration is being provided. The presence of voluminous, billowing white foam indicates that inadequate treatment is occurring. The foam is a result of incomplete degradation of surfactants contained in the influent wastewater and/or the production of surface active agents by rapidly growing bacteria. Such a condition indicates either that the SRT is too short or that a toxicant has entered the plant and inactivated a portion of the biomass. This condition can be remedied by increasing the SRT and/or by locating and eliminating the source of the toxicant.

Under some conditions, thick, viscous brown foam may accumulate on the surface of the bioreactor. This is typically caused by the growth of actinomycetes such as *Gordonia*, although other filamentous bacteria such as *Microthrix parvicella* can also cause it.[36,62] In extreme cases, the foam can accumulate sufficiently to overflow the bioreactor, resulting in unsightly and unsafe conditions. Accumulations within the clarifier can result in carryover of solids to the effluent, degrading it. The factors that cause such foams are complex, and our understanding of these factors continues to evolve. The causative organisms are relatively slow growing and can, in some instances, be controlled by lowering the SRT. Their maximum specific growth rate is about the same as the phosphate accumulating organisms, which, as illustrated in Figure 10.4, corresponds to lower minimum aerobic SRTs than for the nitrifiers. The use of SRT to control the growth of foam-causing organisms is incompatible with nitrifying applications.

Growth and accumulation of foam-causing organisms in activated sludge systems is facilitated by physical designs that trap accumulated foam in the bioreactor. Consequently, good design practice requires foam to flow easily out of the bioreactor to the clarifier, where it is collected and removed from the system.[36,62] Once collected, the foam should be directed to the solids handling system in a way that minimizes recycle; direction to the plant headworks or to a gravity thickener from which the overflow is recycled should be avoided. Selectors can provide limited control over the growth of the causative bacteria, but they are ineffective if foam is recycled as described above. Another approach is a classifying selector, which involves preferentially wasting floating material from the system at a point where it tends to collect. This practice results in enhanced removal of the foam-causing organisms because the removed foam is selectively enriched with them.[12,43,62]

If all else fails, a high concentration chlorine spray can be applied directly to the foam on the bioreactor.[1,36] A fine mist is used so that the chlorine will contact the foam without entering the mixed liquor itself. The chlorine oxidizes the foam and the microorganisms in it, thereby destroying them. Since the foam-causing bacteria will be enriched in the foam, this practice selectively removes them from the system. Another approach that has proven effective is the addition of polymer to either the mixed liquor or the RAS to incorporate the foam-causing organisms into the mixed liquor and allow them to be wasted effectively through the WAS.[36,56]

11.4.2.2 Clarifier

Visual observations of the clarifier indicate its operational status and also provide clues about the settling characteristics of the activated sludge. Denitrification in the clarifier solids blanket causes large clumps of solids to rise to the liquid surface, break, and spread over it. This phenomena, known as clumping, is caused when nitrate-N is present, DO is depleted in the solids blanket, and the residence time of the solids in the blanket is long enough to allow the generation of sufficient nitrogen gas bubbles to cause the solids to float. It can be controlled by altering the SRT of the activated sludge and/or by increasing the RAS flow rate to reduce the size and residence time of the solids blanket. The SRT may be decreased to reduce the degree of nitrification, or it may be

increased to reduce the respiration rate of the activated sludge, thereby reducing the production rate of nitrogen gas. Alternatively, an anoxic zone may be incorporated into the bioreactor to remove nitrate-N, thereby reducing its concentration in the clarifier. The design and operation of activated sludge systems with anoxic zones are discussed in Chapter 12.

Two problems are often associated with long SRTs and/or excessive DO concentrations. They are ashing and pin floc. Both result from an inadequate filament backbone in the activated sludge floc. Growth at long SRTs and high DO concentrations prevents the growth of filamentous microorganisms, which are necessary to produce a strong floc that is resistant to mechanical turbulence. Ashing is the term used when small dark brown to gray particles rise in the clarifier. Pin floc refers to dense, granular solids that flocculate poorly and result in particles that settle rapidly but leave behind a turbid supernatant. Such a situation can be alleviated by reducing the SRT and/or the DO concentration to encourage the growth of a moderate amount of filamentous bacteria.

Straggler floc and dispersed growth are often observed at short SRTs. Straggler floc are relatively large floc particles (0.25 to 0.5 mm) that appear to be light, fluffy, and almost buoyant within the clarifier. They are caused by excessive quantities of filamentous microorganisms, which expand the activated sludge floc resulting in slowly settling floc particles that can be carried out of the clarifier by hydraulic currents. As discussed in Section 2.3.1, dispersed growth is caused by inadequate flocculation and can often be corrected by increasing the SRT.

Observation of the clarifier solids blanket thickness (i.e., the vertical distance between the bottom of the clarifier and the top of the blanket) can provide an early indication of settleability problems. However, care must be exercised to properly interpret such observations because a blanket that is increasing in thickness may simply indicate improper system operating conditions. For example, an inadequate RAS flow rate will cause the blanket to increase in thickness as more solids are applied to it than can be removed by the RAS. An unintended increase in the activated sludge SRT can also cause the blanket to rise by increasing the MLSS concentration beyond that which can be handled by the clarifier. On the other hand, if an increase in the solids blanket thickness is coupled with an increase in the SVI, then deterioration in solids settling and compaction characteristics is probably the cause. The term "bulking" should only be applied to that situation and not to the accumulation of solids within the clarifier. Bulking is generally caused by the growth of excessive quantities of filamentous bacteria, and its causes and cures are discussed in Section 11.2.1. If the corrective measures discussed there cannot be successfully applied or if an immediate remedy is required to prevent permit violations, then either oxidants such as chlorine or hydrogen peroxide may be applied to the activated sludge to reduce the population of filamentous bacteria, or polymers or other coagulants may be added to improve sludge settleability. The addition of oxidants to control the relative populations of filamentous and floc-forming bacteria is discussed in Section 11.4.3.

11.4.2.3 During Sludge Volume Index Measurement

Changes in activated sludge settling characteristics, referred to as "sludge quality" in the operations manuals,[73,75] can often be observed in the vessel used to measure the SVI (settleometer) before they have an impact on the clarifier solids blanket thickness. Phenomena such as ashing, pin floc, straggler floc, and dispersed floc can be observed more precisely in the settleometer than in the clarifier. Denitrification will also often be observed first in the settleometer because its small hydraulic head allows the development of sufficient gas to float the solids before it occurs in the full-scale clarifier.[32] An increase in the SVI value indicates an increase in the population of filamentous bacteria.

Some operators measure the settling velocity of the activated sludge in the settleometer during the SVI test and use it as a control parameter.[73,75] They then adjust the DO concentration and the SRT in the activated sludge system (sometimes called "oxidation pressure") until the relative populations of low DO filaments and floc-forming bacteria are balanced to attain the desired sludge settling velocity. The settling characteristics of the sludge can be adjusted relatively quickly using this technique. Longer times are required, however, to recover acceptable settleability for sludge with a high SVI because of the need to waste a large inventory of filamentous bacteria from the

system. This emphasizes the need to carefully monitor solids settling characteristics and maintain them within acceptable ranges.

The SVI analysis is not appropriate for MBRAS because performance is not affected by the settleability of the mixed liquor. Rather, it is affected by the filterability of the mixed liquor, which is characterized using the capillary suction time (CST) test.[77] In this analysis, a sludge sample is placed on filter paper in a standard test apparatus and the time for water to flow out of the sludge sample and through the filter paper a fixed distance (generally 1 cm) is measured.[3,73] Trends in CST are monitored to detect changes in MBRAS sludge filterability. These trends can be used to identify factors affecting filterability and to adjust operational conditions. Information on the factors affecting sludge filterability is accumulating as more MBRAS installations come into operation.

11.4.2.4 Microscopic Examination

Regular microscopic examination of an activated sludge can provide insight into the factors affecting system performance and greatly assist with process optimization.[78] One type of microscopic examination uses a relatively inexpensive microscope to determine the types of Eucarya present and to characterize the overall structure of the activated sludge floc. Such an examination can supplement observations of the bioreactor, the clarifier, and the characteristics of the sludge in the settleometer. The types of Eucarya present tend to correlate with the operating SRT and can provide rapid, visual confirmation that the proper value is being achieved. A rapid change in the types of Eucarya present can be an indication of an inadequate DO concentration or the presence of a toxicant. A second type of microscopic examination is used to characterize the types of filamentous bacteria present using the technique of Eikelboom[24,25] as modified by Jenkins et al.[36] This examination requires a research grade microscope and a high degree of specialized training. The information obtained can be used to identify corrective actions required to alleviate filamentous bulking problems, as discussed in Section 11.2.1 and indicated in Table 11.2. Facilities that do not generally experience filamentous sludge bulking problems need not conduct this type of examination on a regular basis.

11.4.3 Activated Sludge Oxidation to Control Settleability

The application of oxidizing agents to activated sludge can be used to control the growth of filamentous bacteria. Because such chemicals oxidize filamentous bacteria faster than floc-forming ones, they reduce the relative population size of the filaments in the activated sludge, thereby influencing its settling characteristics.[36] Due to chlorine's low cost and ready availability, it is the oxidant used most often for this purpose, although others, such as hydrogen peroxide, can be used with equal effect. Three factors are important in the use of chemical oxidation to control activated sludge settling characteristics: proper control of the oxidant dose, selection of an appropriate dose point, and mixing at the dose point.

Because the purpose of oxidant addition is to destroy part of the activated sludge, and because the oxidant should be added continuously, the dosing rate should be expressed as the mass of oxidant added per day per mass of activated sludge in the system. Typical units are g of oxidant/(kg MLVSS · day), with mixed liquor volatile suspended solids (MLVSS) being used to represent the activated sludge because only the organic fraction of the sludge will react with the oxidant. The dose required will depend on the severity of the filamentous bulking and the speed with which it is desirable to reduce the SVI, with dosing rates typically ranging from a low of about 2 g Cl_2/(kg MLVSS · day) to a high of about 10. Prior to the initiation of chlorination, a target SVI is selected that will give the desired overall system performance, and a target SVI range is also chosen, often ± 20 mL/g. The dosing rate is then selected, chlorination is initiated, and the response of the system is monitored in terms of changes in the SVI and the abundance of filaments, as revealed by periodic microscopic examination. The dose is adjusted in accordance with the response until the SVI is within the target range. The dose is reduced if the SVI is within the target range but decreasing,

while it is increased if the SVI is increasing. Chlorination is terminated when the SVI falls below the target range. If time is available to reduce the SVI slowly, then chlorination can begin at a relatively low dose of about 2 g Cl_2/(kg MLVSS · day) and be slowly increased until a downward trend in the SVI is established. If the SVI must be reduced rapidly, then a more aggressive dose, on the order of 6 g Cl_2/(kg MLVSS · day), should be selected, with the dose being reduced when a downward SVI trend is established.

The dose point should be selected to avoid contact with the influent wastewater while achieving a desired dosing frequency. Convenience and adequate mixing should also be considered. Direct contact with the influent wastewater must be avoided because the oxidant will react with any organic matter present at the dose point, thereby reducing its effectiveness against the activated sludge. Dosing frequency is important because biomass circulates through activated sludge systems and the entire inventory must be exposed to the oxidant about three times per day. If the dosing frequency is less than this, significant filament growth can occur in the fraction of the biomass that is not being dosed. For some systems, such as CAS, CMAS, CSAS, and SFAS, the RAS stream is an excellent location to add the oxidant. Little influent organic matter is present in the RAS stream and the bioreactor HRT is short enough that biomass will be circulated through the clarifier several times per day, thereby ensuring the needed dosing frequency. For other processes, such as EAAS, the bioreactor HRT is much longer and biomass circulates to the clarifier less than once per day. In such cases the dosing frequency will be insufficient and the oxidant may not be fully effective in controlling filament growth. In this case, the oxidant should be added directly to the bioreactor to obtain the needed dosing frequency.

Mixing at the dose point is important because the reaction between oxidant and biomass is very fast. Thus, rapid mixing must be provided to avoid over-oxidation of a portion of the biomass and under-oxidation of the remainder. A diffuser is generally required to ensure good contact between the added oxidant and the process flow. In addition, mechanical or other mixing may also be needed to ensure adequate dispersion of the oxidant. When the oxidant is added directly to the bioreactor, adequate mixing is often provided by the aeration equipment itself. The reader is referred to the book by Jenkins et al.[36] for further discussion of the use of oxidants to control activated sludge settling characteristics.

11.4.4 Dynamic Process Control

Regulation of the SRT provides long-term control of an activated sludge system. Furthermore, visual observations provide feedback on the success of the SRT control strategy and allow fine-tuning of certain operating parameters, such as the DO concentration. While this approach will permit operational adjustments in response to seasonal variations in process operating conditions and long-term changes in process loadings, most activated sludge systems are also subject to short-term loading variations. Loading variations occur on a diurnal and a day-to-day basis and process operation must be adjusted accordingly. Facilities may also be subject to shock organic loadings due to industrial discharges and shock hydraulic loadings due to the inflow and infiltration of precipitation into the wastewater collection system. Process operation must be adjusted in response to these variations as well. Activated sludge systems possess some ability to respond to short-term loading variations, particularly if they occur regularly and over a period of several hours. However, this capacity is limited by the capability of bacteria to rapidly adjust their enzyme levels and by the time required to grow significant quantities of additional biomass. Thus, some deterioration in performance should be expected in response to dynamic loadings. Dynamic simulation, as illustrated in Chapters 6 and 7, is one technique that can be used to define the deterioration in effluent quality that will result from variations in process loadings. It can be used to determine the treatment limits for a particular process configuration and the benefits to be gained from the use of equalization to dampen load variations. Dynamic simulation can also be used to identify alternative process design and operating conditions that will improve the dynamic response of the system.

Several factors must be considered to ensure that an activated sludge system has adequate dynamic response capability. One is selection of an appropriate SRT.[29,57] The procedure used to select the design SRT for a nitrifying activated sludge system illustrates the principle; a safety factor for the design is selected with consideration of process loading variations. Equalization can be applied to smooth the organic and hydraulic loading variations. It will generally be a necessity for industrial wastewaters in which significant short-term organic loading variations occur as a result of the operation of the production facility that generates the wastewater. On the other hand, experience indicates that the activated sludge process is generally quite resilient and can accept the dry weather variations in organic and hydraulic loadings typical of domestic wastewaters, while still providing acceptable performance. Even in that situation, however, flow equalization can improve the performance of the entire treatment train. Improved response to organic and hydraulic variations can also be achieved by providing alternative activated sludge process operating modes. For example, step feeding capability can be provided to allow peak hydraulic loadings to be processed without causing clarifier thickening failure.

While not yet in routine use, much research is being directed at the development and application of automatic control technology.[4,46] Process instrumentation is available to monitor suspended solids concentrations, DO concentrations, oxygen uptake rates, and residual nutrient concentrations (ammonia, nitrate, and phosphate). The data collected by such sensors, supplemented with laboratory and flow rate data, can be interfaced with dynamic process models to allow investigation of alternative operating strategies. Such approaches have been used off-line to analyze the performance of existing activated sludge systems,[15] and are becoming available for online monitoring and process optimization. Expert systems are also being developed to assist with activated sludge process operation. All in all, these developments offer significant potential to optimize activated sludge operation, thereby improving performance and reliability and reducing costs.

11.5 KEY POINTS

1. Activated sludge bioreactors are typically open basins containing mechanical equipment to transfer oxygen and maintain the flocculent biomass in suspension. Several devices are used to do this, including coarse and fine bubble diffused air, mechanical surface aerators, jet aerators, and submerged turbines.
2. The clarifiers or membrane systems used in activated sludge systems provide two functions: separation of the flocculent biomass to produce a clarified effluent (clarification), and concentration of the biomass for recycle to the upstream bioreactor (thickening).
3. Nine major activated sludge process options exist: conventional (CAS), step feed (SFAS), contact stabilization (CSAS), completely mixed (CMAS), extended aeration (EAAS), high purity oxygen (HPOAS), membrane bioreactor (MBRAS), selector (SAS), and sequencing batch reactor (SBRAS).
4. CAS and CMAS contain uniform mixed liquor suspended solids (MLSS) concentrations and typically have hydraulic residence times (HRTs) ranging from four to eight hours and solids retention times (SRTs) ranging from 3 to 15 days. CAS uses a plug-flow bioreactor, while CMAS uses a single, completely mixed bioreactor. Sludge settleability is usually better for CAS facilities due to better control of the growth of filamentous bacteria. CMAS offers greater resistance to upset by inhibitory organic chemicals.
5. The return activated sludge (RAS) is sequentially diluted as it flows through SFAS and CSAS systems, resulting in an outlet MLSS concentration that is lower than the average concentration in the bioreactor. As a consequence, SFAS and CSAS bioreactors can be smaller than CAS or CMAS bioreactors with the same SRT. SFAS and CSAS bioreactors can be more complex than CAS and CMAS bioreactors, and process operation can be more complex.

6. EAAS systems use long SRTs (20 to 30 days) to partially stabilize the biomass produced. This requires large bioreactors (HRTs of 24 hours or more), but results in excellent process stability and the production of a high quality effluent.
7. In HPOAS systems, biomass and oxygen enriched gas flow cocurrently through a staged, covered bioreactor. This allows use of high volumetric organic loading rates and short HRTs (generally two to four hours). Short SRTs are also generally used (one to five days).
8. The SAS system uses a highly loaded section at the inlet end of the bioreactor (the selector) to create conditions favorable to the growth of floc-forming bacteria relative to filamentous bacteria. Selectors can be incorporated into the other activated sludge options.
9. MBRAS replaces the conventional gravity-based secondary clarifier with a membrane liquid-solids separation system for biomass retention. The result is a system that can reliably retain higher MLSS concentrations and produce an effluent with lower concentrations of particulate and colloidal matter.
10. SBRAS incorporates biological reaction and sedimentation into a single vessel. Microprocessors are used to automatically control the influent flow, aeration, mixing, and effluent decanting functions. They are used most often in smaller wastewater treatment plants.
11. For many activated sludge processes, the formation of good settling activated sludge floc that will produce a clear effluent low in effluent total suspended solids (TSS) requires a balance between floc-forming and filamentous bacteria.
12. Bioflocculation, which occurs as a result of exocellular polymers produced by the biomass, typically requires a minimum SRT (one day for domestic wastewaters and three days for industrial wastewaters).
13. Excessive quantities of filamentous bacteria cause activated sludge bulking. Although about 30 types of filamentous bacteria can be found in activated sludge systems, they comprise four groups with different characteristics. Appropriate dissolved oxygen (DO) concentrations (for the process loading factor/respiration rate), adequate nutrients, control of H_2S loading, control of SRT, and installation of selectors can be used to control filament growth to maintain sufficient but not excessive levels.
14. The SRT required for many activated sludge systems is determined by the need for bioflocculation. In other cases, longer SRTs are needed to nitrify, to treat certain industrial wastes containing less biodegradable organic matter, and to stabilize the biomass produced. The reliability and capacity of the solids processing system must also be considered in the selection of the SRT.
15. Two factors limit the bioreactor MLSS concentration. One is solids thickening, which limits the maximum economical MLSS concentration to about 5000 mg/L as TSS for systems using clarifiers. MBRAS systems can employ higher MLSS concentrations. The other is bioflocculation, which typically requires a minimum MLSS concentration of 500 to 1000 mg/L as TSS.
16. The primary effect of the DO concentration is on the growth of filamentous bacteria, although it will also affect the occurrence of nitrification. A DO concentration of 2 mg/L is a reasonable benchmark, but in some situations successful treatment can be obtained with lower values, whereas in others, higher values will be required.
17. Use of the oxygen transfer equipment both to transfer oxygen and to maintain solids in suspension places constraints on the size of the bioreactor. If it is too small, then the volumetric power input required to transfer the needed oxygen will cause floc shear. If it is too large, then mixing will control the volumetric power input, resulting in increased power requirements.
18. Adequate nutrients are required to allow balanced growth of biomass in activated sludge systems. Nutrient limitations can result in the growth of undesirable quantities of filamentous

bacteria and/or the production of exocellular slime. Both interfere with activated sludge settling and compaction.

19. The temperature of an activated sludge system must be maintained in either the mesophilic (35 to 40°C) or the thermophilic (45 to 60°C) range; it should not fluctuate between the two. The oxidation of organic matter and ammonia-N results in the liberation of heat, whereas the physical configuration of the bioreactor and the nature of the oxygen transfer device influence the loss of heat.
20. The design of an activated sludge process generally consists of the following six steps:
    a. Select the activated sludge option and SRT based on wastewater characteristics, effluent quality goals, facility capital and operating costs, and operational objectives.
    b. Calculate the mass of MLSS in the system $(X_{M,T} \cdot V)_{System}$, the quantity of waste sludge, $W_{M,T}$, and the oxygen requirement, RO, by using the modified stoichiometric model of Chapter 5.
    c. Distribute the oxygen requirement as required by the system configuration.
    d. Calculate the upper and lower limits on the bioreactor volume based on mixing, floc shear, and oxygen transfer.
    e. Using $(X_{M,T} \cdot V)_{System}$ and the output from Step d, calculate the upper and lower limits on the MLSS concentration and choose an MLSS concentration within those limits based on consideration of final settler design. Calculate the bioreactor volume associated with the chosen MLSS concentration.
    f. Optimize the system using tools such as activated sludge model (ASM) No. 1.

    Further detail is provided in Table 11.3.
21. The mass of MLSS and the quantity of waste sludge for any activated sludge system can be calculated once the SRT is established. They are calculated for cold weather conditions because they produce the largest values, thereby determining the required size of the bioreactor.
22. The system oxygen requirement can also be calculated once the SRT is fixed. It will be greatest at the highest temperature, and thus summer conditions are used in design. This requirement must be apportioned among the different vessels in a multitank system. Transient requirements must also be considered when sizing the oxygen transfer system.
23. The size of the bioreactor is determined from $(X_{M,T} \cdot V)_{System}$ based on the design MLSS concentration. The choices of the MLSS concentration and the bioreactor volume must be consistent with the constraints given in Key Points 15 and 17.
24. Aerobic selectors prevent filamentous sludge bulking by allowing removal of the readily biodegradable organic matter in an environment with a sufficiently high specific growth rate for kinetic selection of floc-forming bacteria. To ensure that this occurs under a variety of loading conditions, selectors are staged and are sized to give a desired average process loading factor.
25. The RAS flow rate and the distribution of the influent wastewater must be considered when sizing the bioreactor for systems with nonuniform MLSS concentrations, such as SFAS and CSAS, because both affect the distribution of MLSS within the system.
26. In the absence of simulation, the spatial distribution of the oxygen requirement in multitank systems can be approximated by partitioning it into its component parts and assigning each to the appropriate portion of the bioreactor system. Convenient divisions are the oxidation of readily biodegradable substrate, the hydrolysis and subsequent oxidation of slowly biodegradable substrate, nitrification, and decay of heterotrophs.
27. Because of the analogy between continuous flow and batch activated sludge systems, SBRAS systems can be designed with the same basic procedures as the other activated sludge options. The primary difference is that the smallest allowable bioreactor size is governed by its role as the final settler.

28. Dynamic models such as ASM No. 1 can be used to refine designs developed using the procedures described above. They allow estimation of the impact of short-term loading variations on effluent quality, and provide better estimates of the spatial and temporal variations in oxygen requirements.
29. Three procedures are routinely used to determine the waste activated sludge (WAS) flow rate required to achieve a given SRT: MLSS analysis, centrifuge analysis, and hydraulic control. The WAS flow rate should not be changed by more than 20% each day, but solids wasting should be frequent enough so that MLSS concentrations do not change by more than 10%.
30. Observations of the color and appearance of the biomass in an activated sludge bioreactor provide information about the system operating conditions.
31. Phenomena such as denitrification in the final clarifier and an imbalance between floc-forming and filamentous bacteria can be detected by visual observations of the clarifier.
32. Thick, brown, viscous foam resulting from excessive quantities of microorganisms such as *Nocardia* and *Microthrix pavicella* can cause severe operating problems. Their growth can be controlled by low SRT operation, selective foam wasting, spraying a chlorine mist on the foam, and polymer addition to the mixed liquor or RAS.
33. Settleability of activated sludge systems using clarifiers is monitored using the Sludge Volume Index (SVI) test. The filterability of activated sludge systems using membranes is monitored using the capillary suction time (CST) test. Trends in these indicators are used to make adjustments in sludge settleability and filterability, respectively.
34. Microscopic examination of the activated sludge biomass should be performed routinely. Observation of the Eucarya present provides visual confirmation of the SRT value. Rapid changes in the Eucarya present indicate an inadequate DO concentration or the presence of toxic materials. Identification of the types of filamentous bacteria present can be used to determine the conditions causing excessive filament growth and the associated sludge bulking problems.
35. Oxidants such as chlorine and hydrogen peroxide can be used to oxidize excessive quantities of filamentous bacteria and control biomass settling characteristics. Oxidant doses are expressed as g oxidant/(kg MLVSS·day). The dose point should be selected to avoid contact with the influent wastewater while achieving a desired dosing frequency; convenience and adequate mixing should also be considered.
36. The dynamic response of an activated sludge system is constrained by the mass of biomass present in the system and by the ability of the microorganisms to synthesize additional enzymes. Longer SRTs generally provide greater capability to metabolize added organic matter. In some cases, equalization is necessary to limit the variations in activated sludge process loadings.

11.6 STUDY QUESTIONS

1. Prepare a table comparing the following factors for the nine major types of activated sludge: SRT, HRT, MLSS concentration, recycle ratio, bioreactor configuration, and design approach.
2. List the main advantages and disadvantages of an activated sludge system in comparison to anaerobic wastewater treatment processes. When would an activated sludge process be used in comparison to an anaerobic process?
3. List the benefits and drawbacks of each of the nine major activated sludge variations. When is each typically applied?
4. Explain why the SRT is generally maintained between 3 and 15 days to obtain an activated sludge that settles well.

5. Three benefits have been attributed to HPOAS: an increased rate of treatment, increased density of biomass with an associated increased settling rate, and reduced rate of excess biomass production. Discuss and evaluate these claims.
6. Discuss the role of both floc-forming and filamentous bacteria in the formation of a good settling activated sludge, explaining why an optimum balance exists.
7. Explain the kinetic selection mechanism as applied to SAS. Describe how the selector should be configured to take advantage of this mechanism and explain why.
8. Discuss the types of filamentous microorganisms that often occur in activated sludge systems and relate the filament types to the environmental conditions favoring them.
9. Discuss the impacts of DO concentration on the performance of activated sludge systems. Identify all potential effects and their typical importance.
10. Discuss the relationship between filament growth and activated sludge floc structure.
11. Discuss the benefits of plug-flow conditions within an activated sludge system. Describe how the process loading factor for the initial contact zone should be selected to optimize sludge settleability.
12. Why should the MLSS concentration in an activated sludge system employing a final clarifier normally lie between 500 and 5000 mg/L as TSS? What factors affect the choice of the value? Why are higher concentrations allowable in systems using membranes for liquid-solids separation?
13. Why must both an upper and a lower limit be placed on the mixing energy supplied per unit volume to an activated sludge bioreactor? List appropriate limit values for diffused air and mechanical surface aeration systems.
14. Why are heat losses from diffused air activated sludge systems less than those from systems using mechanical surface aerators? Does the HRT affect heat loss, and if so, how?
15. Using the wastewater characteristics in Table E9.4, the stoichiometric and kinetic parameters in Table E11.2, and the temperature correction factors in Table E11.1, design a CMAS system to treat an average wastewater flow rate of 30,000 m^3/day using an SRT of seven days. Assume a constant loading. Use mechanical surface aeration and justify all assumptions and decisions. The lowest sustained winter temperature is 13°C and the highest sustained summer temperature is 24°C.
16. The diurnal peak loading for the CMAS system considered in Study Question 15 is twice the average loading. What is the peak oxygen requirement during both low and high temperature operating conditions? Does consideration of the peak oxygen requirement affect the selection of the bioreactor volume? What bioreactor volume and MLSS concentration would you recommend under this condition? Why?
17. Design a CAS system for the situation considered in Study Question 15. Use a diffused air oxygen transfer system and assume that the in-process oxygen transfer efficiency is 12%. Also assume that the hydraulic characteristics of the CAS bioreactor are equivalent to three tanks in series.
18. Use a computer code implementing one of the IWA activated sludge models to evaluate the steady-state distribution of oxygen requirements in the CAS system sized in Study Question 17.
19. Assume that the influent flow and pollutant concentrations for the problem considered in Study Question 17 vary as indicated in Figure 6.2. Use a computer code implementing one of the IWA activated sludge models to determine the effect of these diurnal variations on the system effluent quality and the oxygen requirement in each equivalent tank of the CAS system designed in Study Question 17. Do this for the summer conditions only. How does the flow-weighted average effluent quality compare with the steady-state value obtained in Study Question 17? Would it be possible to deliver oxygen rapidly enough to meet the peak oxygen requirement in the first tank of the system sized in Study Question 17? Why?

If the system is incapable of meeting the peak oxygen requirement, what modifications to the design would be required? What other impacts would such changes have?

20. Repeat Study Question 15 for an SAS system. Assume that the main bioreactor is a single completely mixed vessel, and determine the required sizes for all tanks in the system. Assume that diffused aeration with 12% oxygen transfer efficiency is to be used.
21. An EAAS system is to be designed to treat the wastewater described in Study Question 15. Select and justify the SRT that will be used for the design. Determine the mass of MLSS in the system, the waste solids quantity, and the oxygen requirement under both summer and winter conditions. Assume that the bioreactor will be configured as an oxidation ditch with vertical mechanical surface aerators. The in-process oxygen transfer capacity of the aerators is 1.2 kg O_2/(kW · hr), and the minimum volumetric power input required to maintain solids in suspension is 7 kW/1000 m^3. Determine the allowable range in bioreactor volumes.
22. A four pass SFAS system with an SRT of seven days is to be used to treat the wastewater considered in Study Question 15. The total bioreactor volume is 5625 m^3 and the RAS flow rate is 15,000 m^3/day. Influent wastewater is to be distributed uniformly to each of the four passes. Determine the MLSS concentration in each of the four passes, as well as the oxygen requirement. Use a diffused air oxygen transfer system and assume that the in-process oxygen transfer efficiency is 12%. Is the bioreactor volume acceptable as far as the constraints on mixing energy and oxygen transfer are concerned? Why? How does the winter effluent soluble substrate concentration compare to that from the CMAS system in Study Question 15?
23. For the problem considered in Study Question 15, design a CSAS system to produce an effluent with a concentration of readily biodegradable substrate of less than 5 mg/L as COD, while maintaining an SRT of seven days. For this design assume that the contact tank MLSS concentration will be 2500 mg/L as TSS and that the stabilization basin MLSS concentration will be 8000 mg/L as TSS. Use a diffused air oxygen transfer system and assume that the in-process oxygen transfer efficiency is 12%. Size the contact and stabilization basins and check the oxygen transfer, mixing, and floc shear requirements.
24. Use a computer code implementing one of the IWA activated sludge models to determine the distribution of the steady-state oxygen requirement between the contact and stabilization basins for Study Question 23 and compare the requirements to the estimates made in that study question.
25. An SBRAS process is to be designed to treat the wastewater defined in Study Question 15. An effective SRT of 15 days is to be used, and the SVI is expected to be less than or equal to 120 mL/g. Develop a plot that demonstrates the effect on the total reactor volume of the number of operating cycles per day and the fraction of the total bioreactor cycle devoted to fill plus react.
26. Discuss the factors that must be considered when selecting a wasting frequency for an activated sludge system.
27. Define each of the following terms and tell what each indicates about the operation of an activated sludge system: clumping, ashing, pin floc, straggler floc, and dispersed growth.
28. Discuss the factors that must be considered during selection of the dosing rate and dose point for application of chlorine to control filamentous sludge bulking.
29. Why is SRT control considered to be a long-term operational strategy?

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12 Biological Nutrient Removal

Biological nutrient removal (BNR) processes are activated sludge processes that incorporate anoxic and/or anaerobic zones to provide nitrogen and/or phosphorus removal. Many BNR variants have been developed, representing a wide range of nutrient removal capabilities. This chapter presents the basic design and operational principles for several of them. It builds upon the theoretical concepts presented in Chapters 2, 3, 6, and 7 and the practical concepts presented in Chapters 10 and 11.

12.1 PROCESS DESCRIPTION

12.1.1 General Description

Figure 12.1 illustrates the general configuration of a BNR process for both nitrogen and phosphorus removal and the general location of the anaerobic (ANA), anoxic (ANX), and aerobic (AER) zones. Table 12.1 summarizes the biochemical transformations occurring in each of those zones. It also presents the functions that those zones provide, as well as those required to remove each nutrient. A key point is that an aerobic zone is required in essentially all BNR systems. It is required for nitrogen removal because nitrifying bacteria are obligate aerobes. Either an aerobic or an anoxic zone following an anaerobic zone is required for phosphorus removal because the stored and exogenous organic matter must be oxidized by the phosphate accumulating organisms (PAOs) in an aerobic/anoxic environment to generate the energy required for growth. In most cases, however, an aerobic zone serves that purpose. Ordinary heterotrophic organisms (OHOs) grow in both anoxic and aerobic zones. Table 12.1 may be used to understand the relative roles and the interactions between the zones in BNR processes.

Because BNR processes are variations of the activated sludge process, they are constructed using the same materials and equipment components. The major differences are: division of the bioreactor into anaerobic, anoxic, and aerobic zones; provision of mixed liquor recirculation (MLR) pumping facilities; and provision of mixing equipment in the anaerobic and anoxic zones to maintain solids in suspension while minimizing oxygen transfer. Figure 12.2 illustrates two types of mixers often used. Further discussion of the physical facilities is provided elsewhere.[6–8,51,55,61,66–68]

12.1.2 Process Options and Comparison

Biological nutrient removal process options have been referred to as an "alphabet soup" due to the large number defined, often by acronyms. The large number makes it impossible to describe here all of the available processes, let alone their unique features and relative advantages. Instead, the focus will be on those used most often and those with characteristics that differentiate them most significantly from the others. Table 12.2 summarizes the defining characteristics, benefits, and drawbacks of the BNR process options described in Chapter 1, along with a few others that will be introduced here. Some additional process options will also be mentioned in the text.

The modified Ludzak-Ettinger (MLE) process (Figure 1.13) offers good nitrogen removal, moderate bioreactor volume requirements, alkalinity recovery, good sludge settleability, reduced oxygen requirements compared to traditional activated sludge systems, and simple control. However, a high level of nitrogen removal cannot generally be achieved, as discussed in Section 7.5.3. Practical MLR flow rates limit nitrate-N removal to between 60 and 85%. As illustrated in Figure 7.36, this limitation does not exist for the four-stage Bardenpho process (Figure 1.14), which includes a

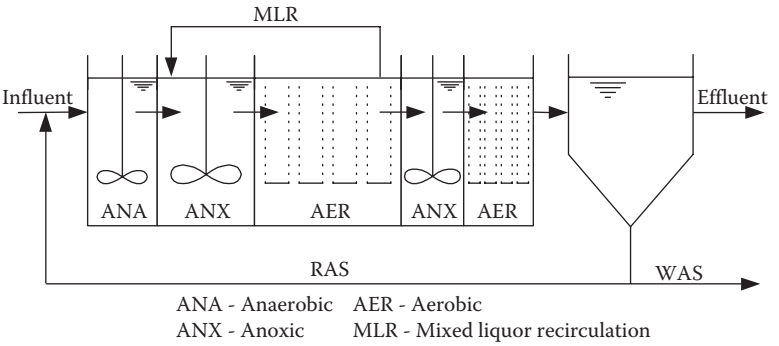

FIGURE 12.1 Single sludge biological nutrient removal process.

TABLE 12.1
Summary of Biological Nutrient Removal Process Zones

Zone	Biochemical Transformations	Functions	Zone Required For
Anaerobic	• Uptake and storage of VFAs by PAOs with associated phosphorus release • Fermentation of readily biodegradable organic matter by heterotrophic bacteria	• Selection of PAOs	• Phosphorus removal
Anoxic	• Denitrification • Alkalinity production • Metabolism of exogenous substrate by denitrifying OHOs • Metabolism of stored and exogenous substrate by denitrifying PAOs • Phosphorus uptake by denitrifying PAOs	• Conversion of NO_3-N to N_2 • Selection of denitrifying bacteria • Growth of denitrifying OHOs • Formation of polyphosphate • Growth of denitrifying PAOs	• Nitrogen removal
Aerobic	• Nitrification and associated alkalinity consumption • Metabolism of exogenous substrate by OHOs • Metabolism of stored and exogenous substrate by PAOs • Phosphorus uptake by PAOs	• Conversion of NH_3-N to NO_3-N • Nitrogen removal through gas stripping • Formation of polyphosphate • Growth of nitrifiers • Growth of OHOs • Growth of PAOs	• Nitrogen removal • Phosphorus removal

second anoxic zone. Performance data from full-scale wastewater treatment plants demonstrate this difference.[55] Processes with one anoxic zone typically produce effluents with total nitrogen concentrations ranging between 5 and 10 mg/L as N, while processes with two anoxic zones typically produce effluents with concentrations ranging between 1.5 and 4 mg/L as N.[55] However, this improved performance is at the expense of a larger bioreactor volume. Another benefit of the MLE and four-stage Bardenpho processes is alkalinity production by denitrification in the initial anoxic zone, which offsets some of the alkalinity consumed by nitrification in the aerobic zone. Denitrification

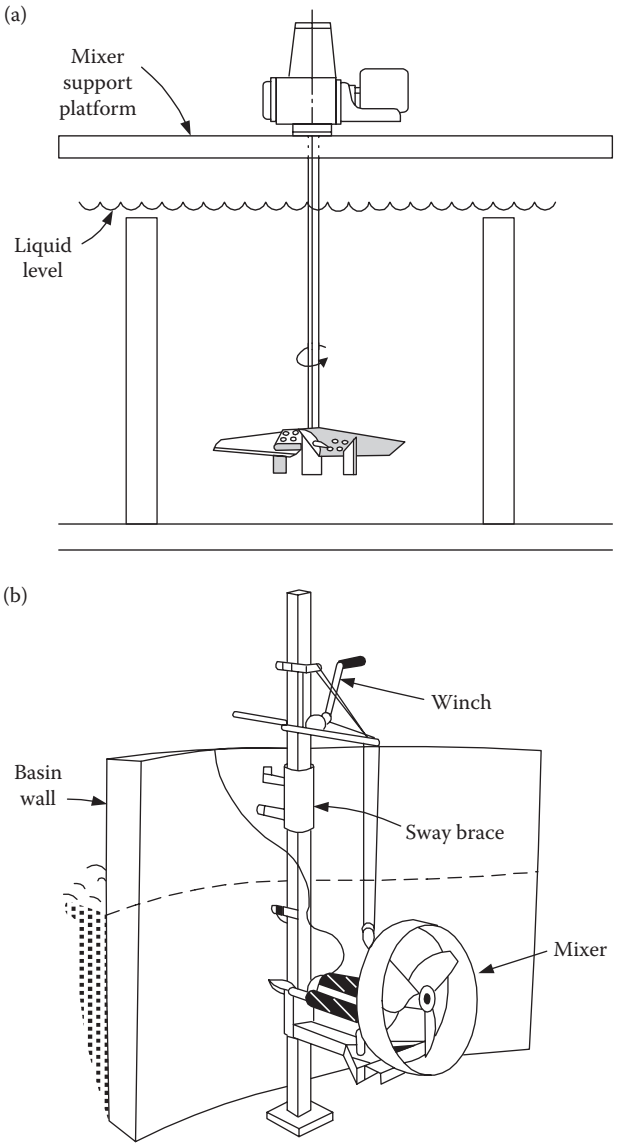

FIGURE 12.2 Typical vertical (a) and horizontal (b) mixers for anaerobic and anoxic zones.

also reduces the oxygen requirement in the aerobic zone because nitrate-N serves as the electron acceptor during oxidation of some of the biodegradable organic matter, thereby removing the need for oxygen to do so. These effects are discussed in Sections 6.3, 6.4, 7.5, and 7.6, and illustrated in Figure 7.30. The reduced power requirement for oxygen transfer in the aerobic zone offsets some or all of the energy required to mix the anoxic zone and to pump the mixed liquor recirculation. Good sludge settleability can be obtained with both the MLE and four-stage Bardenpho processes because the initial anoxic zone acts as a selector to control the growth of filamentous bacteria. This occurs because many filamentous bacteria are not capable of taking up substrate rapidly under anoxic conditions and thus the presence of an anoxic zone results in metabolic selection, as discussed in Section 11.2.1. The incorporation of an anoxic zone can also minimize denitrification problems in the secondary clarifier by reducing nitrate-N concentrations, making it impossible to generate sufficient nitrogen gas to cause sludge flotation.

TABLE 12.2
Biological Nutrient Removal Process Comparison

Process	Defining Characteristics	Benefits	Drawbacks
		Nitrogen Removal	
MLE	• ANX and AER in series • Recirculation from AER to ANX	• Good nitrogen removal • Moderate reactor volume • Alkalinity recovery • Good solids settleability • Reduced oxygen requirement • Simple control	• High level of nitrogen removal not generally possible
Four-stage Bardenpho	• ANX and AER like MLE • Downstream ANX for denitrification • Downstream AER for nitrogen gas stripping	• Excellent nitrogen removal • Alkalinity recovery • Good solids settleability • Reduced oxygen requirement • Simple control	• Large reactor volume
Denitrification in aerobic reactor	• SRT significantly longer than minimum SRT to nitrify • Low DO zones so that denitrification can occur	• Alkalinity recovery • Reduced energy requirement • Easily applied to some existing facilities	• Large reactor volume • Complex control • May result in poor sludge settleability
Separate stage suspended growth denitrification	• Receives nitrified effluent • ANX and AER in series • No mixed liquor recirculation	• Excellent nitrogen removal • Minimum reactor volume	• Requires upstream nitrification • Supplemental electron donor required • High energy requirement
		Phosphorus Removal	
A/O	• ANA and AER in series • No mixed liquor recirculation • Low aerobic SRT to avoid nitrification	• Minimum reactor volume • Good phosphorus removal • Good solids settleability • Simple operation	• Phosphorus removal adversely impacted if nitrification occurs
		Nitrogen and Phosphorus Removal	
A2/O	• ANA, ANX, and AER in series • RAS directed to ANA • Recirculation from AER to ANX	• Good nitrogen removal • Moderate reactor volume • Alkalinity recovery • Good solids settleability • Reduced oxygen requirement • Simple control	• High level of nitrogen removal not generally possible • Moderate phosphorus removal
VIP and UCT	• Like A^2/O except RAS directed to ANX and mixed liquor recirculated from ANX to ANA • ANA, ANX, and AER staged for VIP	• Good nitrogen removal • Good phosphorus removal • Moderate reactor volume • Alkalinity recovery • Good solids settleability • Reduced oxygen requirement • Simple control	• High level of nitrogen removal not generally possible • An additional MLR step is required

TABLE 12.2 (CONTINUED)
Biological Nutrient Removal Process Comparison

Process	Defining Characteristics	Benefits	Drawbacks
JHB	• Like A²/O except RAS directed to separate ANX preceding ANA • RAS flows from initial ANX to ANA, which receives influent	• Good nitrogen removal • Good phosphorus removal • Moderate reactor volume • Alkalinity recovery • Good solids settleability • Reduced oxygen requirement • Simple control	• High level of nitrogen removal not generally possible • An additional ANX required
Five-stage Bardenpho	• Four-stage Bardenpho with initial ANA added • RAS to ANA	• Excellent nitrogen removal • Alkalinity recovery • Good solids settleability • Reduced oxygen requirement • Simple control	• Large reactor volumes • Moderate to poor phosphorus removal

Significant nitrogen removal can also occur in sequencing batch reactor activated sludge (SBRAS) systems.[1,51,55,57] The performance of SBRAS systems and their analogy to continuous flow activated sludge systems, particularly the MLE process, are discussed in Section 7.8. The design and operation of SBRAS systems are discussed in Chapter 11. Many operating SBRAS facilities use an anoxic fill period or anoxic/aerobic cycling as an anoxic selector to control the growth of filamentous bacteria, also resulting in a significant degree of nitrogen removal. Step feed activated sludge (SFAS) processes can also be configured to accomplish biological nitrogen removal by providing an anoxic zone at some or all of the feed addition points.[9,28,36,37] Significant nitrogen removal can be accomplished with this approach. Even more complete nitrogen removal can be accomplished by adding another anoxic zone downstream of the feed points, analogous to the Bardenpho process. It would not receive feed directly.

Significant denitrification is observed in some nitrifying activated sludge systems even though separate anoxic zones are not purposefully provided. This occurs when the operating solids retention time (SRT) is significantly greater than that required to achieve nitrification and regions of low dissolved oxygen (DO) concentration develop that allow denitrification to occur.[17] If the SRT sufficiently exceeds that required for nitrification, the low DO regions will not negatively impact the bioreactor's ability to nitrify. Low DO concentrations can occur in specific regions as a result of the bioreactor and aerator configurations, as illustrated in Figure 12.3, or they can occur inside the activated sludge floc particles because of the requirement to transport oxygen into the floc by diffusion.[21,39] Extensive denitrification has been observed in some bioreactors oxygenated with point source aerators, such as mechanical surface aerators.[51,62] Relatively complete denitrification, ranging up to 70 to 90%, has been observed in oxidation ditch activated sludge systems,[10,17,24,51,55] which incorporate the features of point source aeration and mixed liquor recirculation on an even larger scale (see Figure 1.3). Biological phosphorus removal (BPR) has also been observed in these systems, again as a result of the development of low DO zones.[14,17,41,42] The aeration input to continuous flow systems has also been cycled to create periods of high and low DO to encourage denitrification, as discussed in Section 6.5.2. Systems that encourage denitrification in an aerobic bioreactor provide the benefits of alkalinity recovery and reduction in oxygen requirement associated with the MLE and four-stage Bardenpho processes, but with smaller total energy requirements since mixing and MLR facilities are generally not required. Some existing activated sludge facilities can easily be retrofitted. However, relatively large bioreactor volumes may be required since the microbial

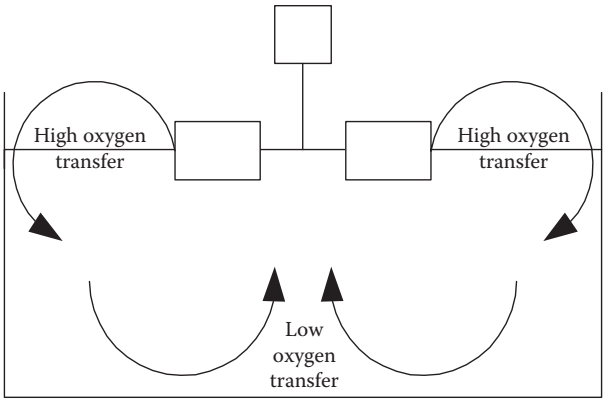

FIGURE 12.3 Variation in oxygen transfer rate in a biological reactor with a point source aerator.

environment is not optimized and control can be more complex because of the need to restrict oxygen input to allow the anoxic regions to develop. Poor sludge settleability may result due to the development of poorly defined aerobic and anoxic zones, which encourage the growth of Group IV filamentous bacteria (see Table 11.2).[35,59]

Separate stage denitrification systems (Figure 1.12) offer the benefits of excellent nitrogen removal and minimum bioreactor volumes. However, extensive upstream treatment is needed to convert ammonia-N to nitrate-N, a supplemental electron donor such as methanol is typically required, and energy is required to mix and aerate the bioreactor. Furthermore, the benefits of reduced oxygen requirements and alkalinity recovery associated with the other processes are not available since denitrification occurs after biodegradable organic matter has been removed from the wastewater and ammonia-N has been converted to nitrate-N.

When process objectives are limited to carbon oxidation and phosphorus removal, the anaerobic/oxic (A/O) process (Figure 1.11) offers the benefits of minimum bioreactor volume, good phosphorus removal, good solids settleability, and simple operation. Good solids settleability results from the anaerobic zone functioning as a selector to control the growth of filamentous bacteria. However, as discussed in Section 7.7, the primary drawback is that phosphorus removal is adversely impacted if nitrification occurs because of the recycle of nitrate-N to the anaerobic zone via the return activated sludge (RAS).

The anaerobic/anoxic/oxic (A^2/O) process (Figure 1.15) offers the benefits and drawbacks of the MLE process, along with the capability to remove a significant degree of phosphorus. Like the A/O process, however, its phosphorus removal capability is adversely impacted by the presence of nitrate-N in the RAS. This limitation is eliminated in the University of Cape Town (UCT) process (Figure 1.16) by directing the RAS to an anoxic zone. However, this added capability is obtained at the expense of another mixed liquor recirculation pumping step, the anoxic recirculation (AR). If the anaerobic and anoxic zones are each staged, the aerobic zone possesses some plug-flow characteristics, and a relatively short SRT is used (sufficient to nitrify but not for sludge stabilization), then the process is known as the Virginia Initiative Plant (VIP) process rather than the UCT process.[23] Several studies have demonstrated the superior phosphorus removal capabilities of the VIP and UCT processes in comparison to the A/O and A^2/O processes in nitrifying applications.[13,22,23,69] The Johannesburg (JHB) process (Figure 12.4) provides another option for eliminating the adverse impact of nitrate in the RAS by denitrifying it before it flows into the anaerobic zone.[46] Although the denitrification rate in the RAS anoxic zone is slow due to the absence of an exogenous carbon source, the high biomass concentration (undiluted RAS) allows acceptable volumetric rates. The JHB process is one of the more frequently used biological phosphorus and nitrogen removal processes due to its mechanical simplicity. The five-stage Bardenpho process (Figure 12.1—referred

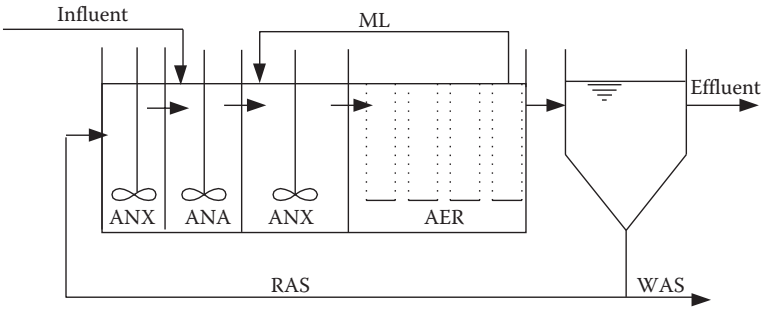

FIGURE 12.4 Johannesburg process.

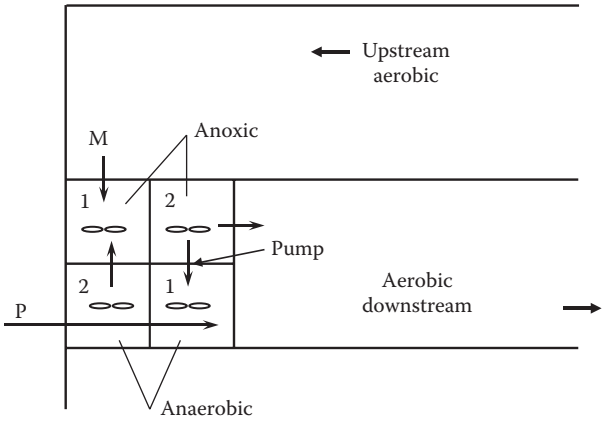

FIGURE 12.5 Configuration of SFAS feed point to accommodate biological phosphorus removal. Mixed liquor (M) from the upstream aerobic zone passes through the two-stage anoxic zone before entering the downstream aerobic zone. Denitrified mixed liquor from stage 2 of the anoxic zone is pumped to the first stage of the anaerobic zone where it mixes with influent wastewater (P). The flow then passes through stage 2 of the anaerobic zone before entering stage 1 of the anoxic zone.

to as the single sludge BNR process) provides the advantages of the four-stage Bardenpho process, along with the capability to remove phosphorus. Although nitrate-N recycle to the anaerobic zone is minimized in this process, experience indicates that its phosphorus removal capability may be only moderate to poor.[55] This results from the relatively long SRTs often used, which lead to relatively low waste sludge production rates, relatively low phosphorus uptake, and secondary release of phosphate in the second anoxic zone.

Biological phosphorus removal can also be incorporated into nitrifying and denitrifying SFAS processes by configuring some or all of the feed points with anaerobic zones receiving influent wastewater and recirculation from an adjacent anoxic zone,[47] as illustrated in Figure 12.5. The main mixed liquor flow path is from the upstream aerobic zone, through the staged anoxic zone, to the downstream aerobic zone. The anaerobic zone is created by recirculating denitrified mixed liquor to the staged anaerobic zone where it is contacted with influent wastewater. The influent wastewater and recirculated denitrified mixed liquor flow through the staged anaerobic zone in the opposite direction of the mixed liquor flow through the anoxic zone and enter stage 1 of the anoxic zone where any unmetabolized biodegradable organic matter provides substrate for denitrification.

Membrane bioreactor activated sludge (MBRAS) can also be configured to achieve biological nitrogen and phosphorus removal by incorporating anoxic and anaerobic zones into the bioreactor. Two challenges exist in accomplishing this: (1) the elevated DO concentrations in the

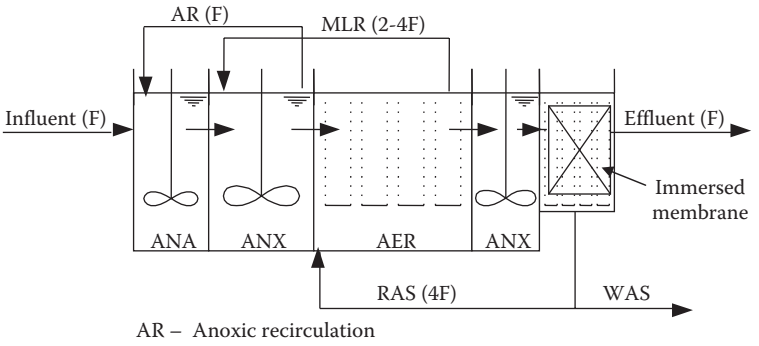

FIGURE 12.6 MBRAS process incorporating biological nitrogen and phosphorus removal. (Adapted from Fleischer, E. J., Broderick, T. A., Daigger, G. T., Fonseca, A. D., Holbrook, R. D., and Murthy, S. N., Evaluation of membrane bioreactor process capabilities to meet stringent effluent nutrient discharge requirements. *Water Environment Research*, 77:162–78, 2005.)

recirculation from the membrane system caused by membrane aeration (which is a necessary component of submerged membrane systems) and (2) the high recirculation rates from the membrane section (generally in the range of 4:1).[16] These effects can be addressed by directing the return flow from the membrane zone to the aerobic zone (where it contributes to satisfying the oxygen demand) and providing separate recirculation streams to upstream anoxic and anaerobic zones. Figure 12.6 illustrates this approach, including a downstream anoxic zone for further denitrification.[29]

Many other biological nitrogen and phosphorus removal processes have been developed and have received some full-scale use. Oxidation ditch processes, such as Orbal[17] and Biodenitro,[11] are widely used for nutrient removal. Full-scale facilities have been modified by turning off aerators and/or by other simple modifications to create the zones necessary to achieve nitrogen and phosphorus removal.[14,17,41,42,51,62] The potential to enhance the removal of nutrients by similar modifications of activated sludge systems appears to be limited only by the imagination and understanding of the process fundamentals by plant designers and operators.

Fermentation of primary sludge to generate an influent stream high in volatile fatty acids for use in systems that remove both nitrogen and phosphorus is a well-established process.[5–7] It offers significant potential for enhancing the performance and improving the reliability of BNR systems.[48,49,58] The impact of fermentation on the performance of BNR facilities is described in Section 12.2.3, while the basic principles of sludge fermentation and the design of fermenters are discussed in Chapter 14.

Several BNR processes are emerging to treat the ammonia-rich recycle streams produced in some solids processing systems.[63,76] As discussed in Chapters 13 and 14, the destruction of organic matter during digestion results in the release of the nitrogen contained in the organic matter as ammonia-N. This can be partially nitrified, and also potentially denitrified, in aerobic digestion processes, but it is conserved in anaerobic digestion processes. If the digested sludge is subsequently dewatered, most of the released ammonia is contained in the liquid separated from the solids. That ammonia-rich stream is typically returned to the main liquid treatment process for treatment. Although the flow rate of this stream is generally small compared to the influent wastewater flow rate, the ammonia mass flow rate can be substantial. For anaerobic digestion processes, the nitrogen mass flow rate in the sludge dewatering stream can increase the nitrogen mass loading on a mainstream BNR process by between 25 and 40%. Moreover, because this stream generally contains insufficient electron donor to allow the nitrogen to be denitrified, it can adversely affect the performance of the mainstream BNR process. One way to address these impacts is to store the sludge dewatering liquid and return it to the mainstream BNR process at times when it is more capable of receiving the nitrogen mass loading. In addition, a number of processes are being developed to address these

impacts. Adding to the "alphabet soup" of BNR process options, they can generally be grouped into four categories as follows:

1. Processes that return the ammonia-rich stream to specific locations within the mainstream BNR process train. For example, if one is using SFAS configured with anoxic zones at the feed points, the ammonia-rich process stream can be added to the initial pass.[20]
2. Processes that use sidestream bioreactors to grow nitrifiers to seed into the main BNR process. These allow a reduced aerobic SRT to be used in the BNR process, reducing its size. This approach is especially applicable for solids process trains using anaerobic digestion due to the uniform and elevated temperature (generally about 35°C) maintained in anaerobic digesters. Since most of this heat is retained in the sludge dewatering stream, the elevated temperature allows nitrification to be accomplished at a reduced SRT in the sidestream bioreactor.
3. Sidestream nitrification processes that convert ammonia only to nitrite rather than nitrate. Oxidation of ammonia to nitrite rather than nitrate reduces the oxygen requirement for ammonia oxidation by 25% and the electron donor requirement for denitrification by 40%.
4. Sidestream processes incorporating the autotrophic anammox denitrification process. In these processes about half of the ammonia is converted to nitrite and the other half remains as ammonia. The nitrite subsequently serves as the electron acceptor and the remaining ammonia as the electron donor, thereby converting both into N_2.

Both knowledge and experience with these process options are evolving rapidly and are expected to result in consensus concerning their practical applicability.

12.1.3 Typical Applications

Single sludge nitrogen removal systems, such as the MLE and four-stage Bardenpho processes, are the preferred alternatives for nitrogen removal during wastewater treatment. Because they can achieve performance comparable to those processes when configured properly, step feed systems with anoxic zones are also popular for larger plants (to minimize the tankage requirements), for plants with significant site constraints, and for retrofits in which the use of step feeding can avoid the construction of the additional tankage that would be required to implement MLE or four-stage Bardenpho technologies. Physical-chemical technologies, such as ammonia stripping, ion exchange, and breakpoint chlorination, have been found to be expensive options that are difficult to operate and maintain.[55,61] Although separate stage biological nitrogen removal processes are technically feasible, they are often not cost-effective relative to single sludge systems. However, they may be the best alternative when nitrogen removal must be added to an existing wastewater treatment plant, particularly if little space is available to construct additional facilities. In such instances, attached growth bioreactors (Chapter 21), such as packed bed downflow denitrifying bioreactors, moving bed bioreactors, or fluidized bed bioreactors, may be better than suspended growth denitrification systems. Consequently, although separate stage suspended growth denitrification systems are capable of providing a high level of performance,[77] they have received little full-scale use.[55,61] Many examples of full-scale suspended growth single sludge biological nitrogen removal facilities exist and are documented in the literature.[6,7,55,61,67]

Biological phosphorus removal is seeing increased use compared to chemical phosphorus removal and offers the advantages of reducing or eliminating chemical usage and decreasing the quantity of waste sludge that must be processed. These benefits must be evaluated relative to the capital and operating costs associated with the construction of the necessary facilities to accomplish biological phosphorus removal. In most instances BPR facilities also incorporate nitrogen removal, either because nitrogen removal is required or because nitrification is required and the incorporation of denitrification is needed to mitigate the adverse impacts of nitrate recirculation to the anaerobic zone on biological phosphorus removal. Thus, biological nitrogen and phosphorus

removal is an accepted and widely used approach. Solids fermentation is being used frequently to improve the performance and reliability of BPR facilities. Sedlak[55] presents several examples of full-scale BPR facilities.

Biological nutrient removal capabilities have also been incorporated into the design of some activated sludge facilities due to the process benefits they provide. When properly designed, initial anoxic and anaerobic zones function as selectors to enhance solids settleability by controlling the growth of filamentous bacteria. The inclusion of denitrifying capabilities in nitrifying activated sludge systems can also reduce alkalinity and energy consumption, and can reduce solids flotation problems in secondary clarifiers by decreasing the nitrate-N concentration in them. In several instances, these benefits have proven to be cost-effective and have improved the performance and reliability of the activated sludge systems into which they were incorporated.[18,35,44] Thus, BNR systems are a standard and widely applied technology, even when nutrient removal is not required.

12.2 FACTORS AFFECTING PERFORMANCE

Many of the factors that affect the performance of BNR processes are similar to those that affect activated sludge systems. This section will discuss those factors that are of particular concern in BNR systems.

12.2.1 Solids Retention Time

Solids retention time plays the same role in BNR processes as in activated sludge systems. In this case, however, the SRTs in the separate zones are of more interest than the total SRT (Θ_c), since they control what occurs in those zones,[40] as discussed in Chapter 6 and 7. In Section 6.5.2, the aerobic SRT ($\Theta_{c,AER}$) was defined as the portion of the total system SRT that is aerobic. Because the SRT is defined in terms of the mass of mixed liquor suspended solids (MLSS, $X_{M,T}$) in a system (Equations 5.1 and 7.1), the fraction of the system SRT that is aerobic is equivalent to the fraction of the MLSS in the system that is maintained under aerobic conditions, $f_{XM,AER}$. Consequently:

$$\frac{\sum (X_{M,T} \cdot V)_{AER}}{(X_{M,T} \cdot V)_{System}} = f_{XM,AER} = \frac{\Theta_{c,AER}}{\Theta_c}. \tag{12.1}$$

Likewise, for the anoxic SRT ($\Theta_{c,ANX}$):

$$\frac{\sum (X_{M,T} \cdot V)_{ANX}}{(X_{M,T} \cdot V)_{System}} = f_{XM,ANX} = \frac{\Theta_{c,ANX}}{\Theta_c}, \tag{12.2}$$

and the anaerobic SRT ($\Theta_{c,ANA}$):

$$\frac{\sum (X_{M,T} \cdot V)_{ANA}}{(X_{M,T} \cdot V)_{System}} = f_{XM,ANA} = \frac{\Theta_{c,ANA}}{\Theta_c}. \tag{12.3}$$

The MLSS concentration in any given tank will depend on the positions of influent flows, solids recycle streams, and MLR streams. Like conventional activated sludge (CAS) systems, MLE, four-stage Bardenpho, A/O, and A^2/O systems can be considered to have uniform MLSS concentrations throughout. As a consequence, the fraction of the biomass held under any given environmental condition is just equal to the fraction of the system volume maintained under that condition. Because of the anoxic mixed liquor recirculation (AR) in the UCT and VIP processes, however, the MLSS

concentration in their anaerobic zones will be smaller than the MLSS concentration in the rest of the system. Consequently, in those processes, the fraction of the system volume that is anaerobic is larger than the fraction of biomass held under anaerobic conditions and the fraction of the SRT that is anaerobic. In contrast, in the JHB process the fraction of the system volume that is anoxic is smaller than the fraction of biomass held under anoxic conditions and the fraction of the SRT that is anoxic. Likewise, MLSS concentrations can vary through the various zones of MBRAS configured for biological nitrogen and phosphorus removal when the membrane zone recirculation flow is added to the aerobic zone and mixed liquor is recirculated to upstream anoxic and/or anaerobic zones as illustrated in Figure 12.6.

One way in which the aerobic SRT can dramatically affect a BNR system is through organism selection. If the system is to be designed and operated to achieve phosphorus removal, the aerobic SRT must be sufficiently long to allow PAOs to grow. Figure 10.4 provides guidance for selection of the required aerobic SRT as a function of temperature. Although the values are relatively low, full-scale activated sludge systems can be operated at aerobic SRTs below that required to grow PAOs.[56] Thus, care must be exercised to ensure that the aerobic SRT is sufficiently long. Just how long that SRT must be is determined in part by whether nitrification is required. If nitrification is an objective, it will control the choice of the aerobic SRT because, as Figure 10.4 illustrates, the minimum aerobic SRT for nitrification is longer than the minimum for growth of PAOs. If nitrification is not an objective, on the other hand, making the aerobic SRT short enough to preclude nitrification allows phosphorus removal to be maximized. This is particularly important for a system like the A/O process, in which nitrate-N will be recycled to the anaerobic zone should nitrification occur. Interestingly, Figure 10.4 suggests that at temperatures above about 25°C, it may be very difficult to operate at an aerobic SRT sufficiently high to allow PAOs to grow while also excluding nitrifying bacteria. Consequently, under those circumstances, systems like the VIP, UCT, and JHB processes have distinct advantages. Appropriate safety factors must be added to the values presented in Figure 10.4 to derive design aerobic SRTs, as illustrated in Section 11.3.2.

Experience also suggests appropriate ranges for the SRT in the anaerobic and anoxic zones. If the removal of readily biodegradable substrate is the primary objective in these zones, then an SRT of about 1 day for temperatures above 20°C is appropriate, while it must be increased to about 1.5 days for colder temperatures.[44] These criteria would also be used if either anaerobic or anoxic zones were being incorporated into a system as selectors to control filament growth. As discussed in Section 11.2.1, readily biodegradable substrate must be removed upstream of the main, aerobic section of the bioreactor if the growth of filamentous bacteria is to be controlled. Since this generally represents a baseline process objective, these values represent the minimum ones that should be used for the combination of the anaerobic and anoxic zones. The anaerobic plus anoxic SRT must be increased if significant quantities of slowly biodegradable substrate are to be metabolized in the zones. Anaerobic plus anoxic SRTs in the two to three day range have been used successfully in some pilot- and full-scale BNR systems.[23,53] In such circumstances, the portion of the total SRT allocated to the anaerobic zone may be as short as 0.5 days if the concentrations of volatile fatty acids (VFAs) and readily biodegradable organic matter in the process influent are sufficiently high.[40] Increases in the anaerobic SRT will allow increased fermentation of biodegradable organic matter in the anaerobic zone, resulting in increased production of VFAs and increased biological phosphorus removal. Increased anoxic zone SRT will result in increased metabolism of slowly biodegradable substrate there and increased nitrogen removal capacity. Kinetic expressions such as those incorporated into International Water Association (IWA) Activated Sludge Model (ASM) No. 2d must be used to specifically determine how much the anaerobic and anoxic SRTs must be increased to achieve a particular set of objectives.

Longer SRTs in BNR systems allow more complete metabolism of organic matter, which can increase nitrogen removal but adversely impact phosphorus removal. Denitrification requires an electron donor. Readily biodegradable organic matter can be used rapidly as an electron donor, but slowly biodegradable organic matter must be hydrolyzed before it can be used. Hydrolysis is

a relatively slow reaction under anoxic conditions and, consequently, long anoxic SRTs may be required when slowly biodegradable substrate must be used as the electron donor.[23,45] On the other hand, phosphorus removal can be adversely affected by the use of relatively long anoxic or aerobic SRTs.[40] This may occur for at least three reasons: (1) long SRTs result in reduced solids production so that less phosphorus is removed from the process in the waste activated sludge (WAS); (2) long aerobic SRTs result in relatively complete oxidation of organic storage products and a reduced rate of phosphorus uptake in the aerobic zone; and (3) decay reactions cause secondary release of phosphorus (i.e., the release of phosphorus without a corresponding uptake and storage of biodegradable organic matter). Thus, SRTs beyond that just required to meet treatment objectives should be avoided for BPR systems.

In summary, for BPR processes an anaerobic SRT of 1 to 1.5 days should be used, depending on temperature. The aerobic SRT for such systems should be selected from Figure 10.4, with an appropriate safety factor applied. A safety factor of about 1.5 is probably sufficient for most applications. For biological nitrogen removal processes, the initial anoxic SRT must be at least 1 to 1.5 days (depending on temperature), but a larger value may be used if significant hydrolysis of slowly biodegradable substrate is required. The aerobic SRT should be selected to achieve reliable nitrification, as discussed in Sections 11.2.2 and 11.3.2. For processes that remove both nitrogen and phosphorus, an anaerobic plus anoxic SRT of at least two to three days should generally be used, with higher values being used if significant hydrolysis of slowly biodegradable organic matter is necessary in the anaerobic and anoxic zones. The anaerobic SRT should be at least 0.5 days. The aerobic SRT should be selected in the same fashion as for nitrogen removal processes, although care should be exercised to use the minimum value that provides the necessary system performance.

12.2.2 Ratios of Wastewater Organic Matter to Nutrient

The concentration of biodegradable organic matter relative to the nutrient concentrations in an influent wastewater can dramatically affect the performance of a BNR system. This is because of the key role biodegradable organic matter plays in nutrient removal. Nitrogen removal is accomplished when biodegradable substrate is used as the electron donor by denitrifying bacteria under anoxic conditions. Phosphorus removal is accomplished when VFAs, which are either a part of the influent readily biodegradable substrate or are formed from it, are taken up and stored by PAOs in the anaerobic zone, thereby allowing them to increase the phosphorus content of the MLSS in the aerobic zone. Since only some PAOs can denitrify and their efficiency at forming polyphosphate under those conditions is less,[4,26,75] it would seem that the organic matter requirements for nitrogen and phosphorus removal could be additive when both nutrients are to be removed. As discussed in Section 12.3.3, however, evidence suggests that this is not the case and only the greater of the two requirements (i.e., for nitrogen or phosphorus removal) need be supplied.

Consider first the amount of organic matter required to remove nitrogen. As shown in Table 3.1, the chemical oxygen demand (COD) mass equivalent of nitrate-N as an electron acceptor is –2.86 g COD/g N. However, this does not mean that an influent biodegradable COD to total Kjeldahl nitrogen (TKN) ratio of 2.86 will allow complete denitrification. There are several reasons for this. First, some of the biodegradable organic matter in the influent is incorporated into biomass, which is why the $\Delta S/\Delta N$ ratio for denitrification, as calculated with Equation 6.4, is always greater than 2.86. Second, some of the influent nitrogen is incorporated into the biomass that grows in the process, which is why the nitrogen available for nitrification, $S_{N,a}$, as calculated with Equation 11.17, is always less than the influent TKN concentration. Third, some of the biodegradable organic matter is slowly biodegradable and may not be metabolized rapidly enough to be available in the anoxic zones. Thus, rather than being used for denitrification, it will be oxidized aerobically. Finally, differences in system configuration result in different utilization efficiencies for the biodegradable organic matter. Table 12.3 provides general guidance concerning the amenability of various wastewaters (characterized in terms of the amount of organic matter relative to the amount of nitrogen)

TABLE 12.3
Biological Nitrogen Removal Efficiency Expected at Various Influent Organic Matter to Nitrogen Ratios

Influent Organic Matter to Nitrogen Ratio			Nitrogen Removal
COD/TKN	BOD_5/NH_3-N	BOD_5/TKN	Efficiency
<5.0	<4.0	<2.5	Poor
5.0–7.0	4.0–6.0	2.5–3.5	Moderate
7.0–9.0	6.0–8.0	3.5–5.0	Good
>9.0	>8.0	>5.0	Excellent

to biological nitrogen removal. The values given can be used to screen candidate wastewaters to determine how difficult it may be to achieve good nitrogen removal. Alternatively, procedures such as those developed in South Africa are available to estimate the appropriate COD/TKN ratio for various nitrogen removal processes.[25,69] Experience has indicated that, in many instances, these procedures provide conservative estimates of the required ratio.[51] That is, adequate performance can be obtained at values of the COD/TKN ratio lower than the recommended value. They may be useful, however, in circumstances where "worst case" values need to be estimated.

Recent research has revealed an impact of COD/NO_3-N ratio in the feed to denitrifying bioreactors on the nature of the gases emitted.[33,34,38,54] When the COD/NO_3-N ratio is too low to allow complete denitrification, the residual dissolved nitrogen will not necessarily be in the form of nitrate-N. Rather, some or all will be present as the denitrification intermediates nitrite, nitric oxide, and nitrous oxide. Because nitric oxide and nitrous oxide are gases of low solubility, they will escape to the atmosphere, which is undesirable because nitric oxide is an important precursor of ground-level ozone and nitrous oxide is a strong greenhouse gas (approximately 300 times as strong as carbon dioxide). Unfortunately, the quantities of gases emitted are influenced by a number of factors in addition to the COD/NO_3-N ratio, including the type of nitrogen removal process employed and the nature of the wastewater, making it difficult to predict their emission rates. Consequently, when COD/TKN ratios are low, consideration should be given during pilot testing to the potential for undesirable gaseous emissions and the investigation of ways to eliminate them.

Next consider the amount of organic matter required to remove phosphorus. The ratios of COD and five-day biochemical oxygen demand (BOD_5) to total phosphorus (TP) and soluble BOD_5 to soluble phosphorus (SP) are often used to judge the phosphorus removal potential of a wastewater.[51,55,60,66] Figure 12.7[60] illustrates the impacts of the BOD_5/TP ratio and the soluble BOD_5/SP ratio on the performance of some full-scale BPR processes. Results such as these demonstrate that a relationship exists, but that it varies with process type and operating conditions. The concept of a minimum organic matter requirement for biological phosphorus removal has resulted in a distinction between carbon limited and phosphorus limited wastewaters. A carbon limited wastewater is one in which insufficient organic matter is available to remove all of the phosphorus. As a consequence, phosphorus will be present in the process effluent at a concentration determined by the relative concentrations of phosphorus and organic matter in the influent. A phosphorus limited wastewater is one in which more than sufficient organic matter is available to remove the phosphorus. Consequently, the effluent phosphorus concentration will generally be low when it is treated in a BPR process. Thus, a phosphorus limited wastewater is desirable when a good quality effluent must be produced.

Recognition of the concept of carbon limited wastewaters has resulted in the development of benchmark ratios expressing the amount of organic matter required to remove a unit of phosphorus by various BPR processes. Such ratios have been determined from pilot- and full-scale BPR systems operating under carbon limited conditions and have been used to characterize the capabilities

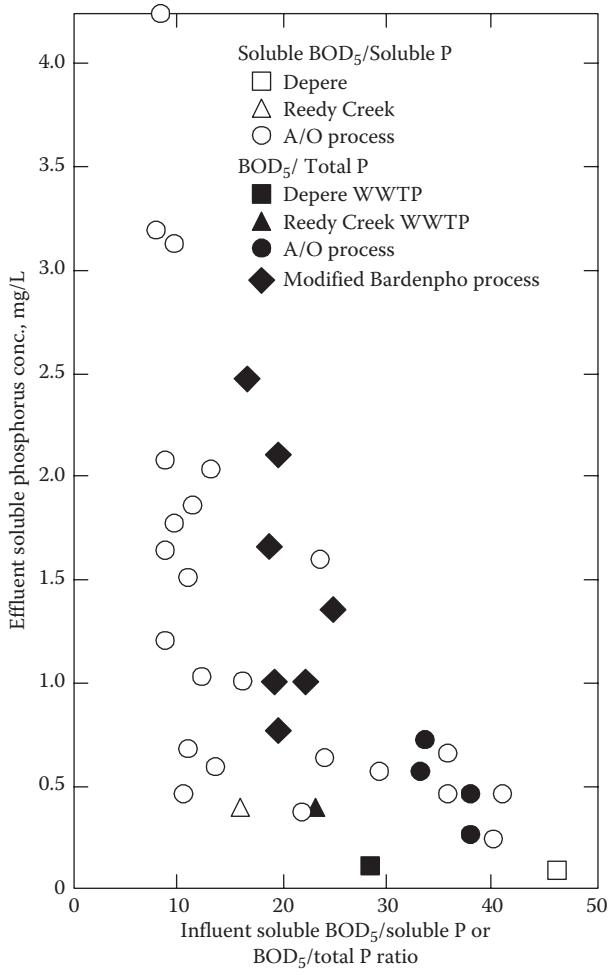

FIGURE 12.7 Example relationship between the influent BOD_5/P ratio and the effluent soluble phosphorus concentration. (From Tetreault, M. J., Benedict, A. H., Kaempfer, C., and Barth, E. D., Biological phosphorus removal: A technology evaluation. *Journal, Water Pollution Control Federation*, 58:823–37, 1986. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

of those systems.[55] A commonly used ratio is the BOD_5 to phosphorus removal ratio ($BOD_5/\Delta P$), which is calculated as

$$\frac{BOD_5}{\Delta P} = \frac{BOD_5 \text{ in biological process influent}}{TP \text{ in biological process influent} - SP \text{ in biological process effluent}}. \quad (12.4)$$

Phosphorus removal (ΔP) is quantified as the total phosphorus in the biological process influent minus the soluble phosphorus in the biological process effluent. Influent total phosphorus is used because both soluble and particulate phosphorus are acted upon. Soluble phosphorus is in the form of, or rapidly converted to, inorganic phosphate, which is the form taken up and stored by the PAOs. Particulate phosphorus is either hydrolyzed and released as soluble phosphorus or it is entrapped in the MLSS and removed in the waste solids. In either case, the particulate phosphorus affects the overall phosphorus removal by the process. Effluent soluble phosphorus is used because effluent particulate phosphorus is generally associated with suspended solids that have escaped the clarifier and are a function of the efficiency of the clarifier, not the biological process.

TABLE 12.4
BOD_5 and COD to Phosphorus Removal Ratios Associated with Various Types of Biological Phosphorus Removal Processes

Type of BPR Process	$BOD_5/\Delta P$ Ratio (mg BOD_5/mg-P)	COD/ΔP Ratio (mg COD/mg-P)
High efficiency (e.g., A/O without nitrification, VIP, UCT)	15–20	26–34
Moderate efficiency (e.g., A/O and A^2/O with nitrification)	20–25	34–43
Low efficiency (e.g., Bardenpho)	>25	>43

Table 12.4 provides typical ranges for $BOD_5/\Delta P$ associated with a variety of BPR processes. For consistency, values are also presented for COD/ΔP that were calculated from the BOD_5 values using Equation 9.31. Biological phosphorous removal processes are categorized as high, moderate, or low efficiency, depending upon the amount of organic matter required to remove a unit of phosphorus in them. Highly efficient BPR processes, such as the VIP or JHB process or the A/O process operating under nonnitrifying conditions, require only 15–20 mg BOD_5 (26–34 mg COD) to remove a mg of phosphorus. In these processes, essentially no nitrate-N is recycled to the anaerobic zone, either because it is not generated (for the nonnitrifying A/O process) or it is removed (for the VIP and JHB processes), and hence electron donor is not required to denitrify, with the result that all of the organic matter is available for phosphorus removal. They also both have low SRTs, which maximizes phosphorus uptake and waste solids production. A moderately efficient process, such as a nitrifying A/O or A^2/O process, will require 20–25 mg of BOD_5 (34–43 mg COD) to remove a mg of phosphorus. More organic matter is required for these systems because some will be consumed by OHOs in the anaerobic zone due to denitrification of the nitrate-N recycled there in the RAS. The ratio will be even higher for a low efficiency process, such as a five-stage Bardenpho process operating at long SRT, which require more than 25 mg of BOD_5 (43 mg COD) to remove a mg of phosphorus because of the extensive denitrification occurring.

The information in Table 12.4 can assist design engineers in choosing the appropriate type of BPR process for a particular wastewater. If the amount of COD in the wastewater is high relative to the amount of phosphorus (COD/P), then almost any type of BPR process is applicable and the engineer has broad latitude in selecting the one to be applied. Conversely, the lower the COD/P ratio, the less latitude the designer has. For example, if only 30 mg COD is available for each mg of phosphorus in the wastewater, only high efficiency BPR processes can be used. If the process selected has a COD/ΔP value that is lower than the COD/P value of the wastewater, the process will operate under phosphorus limited conditions, which will allow good removal of phosphorus. It must be emphasized that the organic matter to nutrient ratios discussed above are for the influent to the biological treatment system, not the ratios for the influent to the entire wastewater treatment plant. This is because the ratio can be significantly altered by treatment upstream of the biological process and by recycle streams from the solids handling system.

The potential for nitrogen inputs to biological processes from solids handling systems was discussed in Section 12.1.1. Solids handling systems can also result in solubilization of removed phosphorus and its recycle back to the liquid treatment process train, particularly if the solids are held under anaerobic conditions, which cause PAOs to release phosphorus. This includes, for example, wet wells and gravity sludge thickeners, as well as anaerobic digesters. Interestingly, full-scale experience indicates that of some of the phosphorus released during anaerobic digestion of BPR waste solids can precipitate and be retained with the solids, especially in anaerobic digesters.[51,55] Precipitates include struvite ($MgNH_4PO_4$), brushite ($CaHPO_4 \cdot 2H_2O$), and vivianite [$Fe_2(PO_4)_3 \cdot H_2O$]. In some cases, these precipitates can be purposefully formed as a means of recovering phosphorus,

as discussed in Section 23.4.1. Nevertheless, the process designer must be aware of the potential impacts of phosphorus recycle and balance the requirements of the liquid and solids processing trains to obtain an optimum treatment system.

12.2.3 Composition of Organic Matter in Wastewater

The composition of the organic matter present in a wastewater, particularly its biodegradability, also affects the performance of BNR processes. In the anaerobic zone, PAOs transport short chain VFAs into the cell and store them as poly-β-hydroxyalkanoates (PHAs).[72,75] There are two sources of VFAs for the PAOs; they are either present in the influent wastewater or they are produced by fermentation of other readily biodegradable substrate by facultative heterotrophs. The uptake of VFAs is a rapid process, while fermentation is a slower process.[23,70,72] Ideally a wastewater that is to be treated in a BPR system will contain a high proportion of VFAs; this will result in their rapid uptake by the PAOs and a relatively small anaerobic SRT can be used. At a minimum, a sufficiently high concentration of fermentable organic matter must be present to generate VFAs for uptake by the PAOs. Between 7 and 10 mg of VFAs must be available in the anaerobic zone (either in the influent wastewater or produced through fermentation) for each mg of phosphorus as P to be removed by the PAOs. Thus, the readily biodegradable substrate concentration in the influent wastewater, particularly the VFA concentration, will significantly affect the performance of a BPR system.

The readily biodegradable substrate concentration of the influent wastewater will also affect the denitrification rate in an initial anoxic zone. Denitrification is rapid when readily biodegradable substrate is available, but is much slower when only slowly biodegradable substrate is present. This is because the use of slowly biodegradable substrate is controlled by the rate of hydrolysis, which is relatively slow under anoxic conditions. Consequently, if the amount of readily biodegradable substrate entering an initial anoxic zone is insufficient to remove the nitrate-N added, the anoxic zone must be large enough to provide time for the hydrolysis of slowly biodegradable substrate. If it is not, the potential exists for the release of nitric and nitrous oxide to the atmosphere.[33,34,38,54] Hydrolysis and fermentation of slowly biodegradable substrate in an upstream anaerobic zone can produce readily biodegradable organic matter that can pass into an anoxic zone and produce a high rate of denitrification there.[15]

Significant fermentation will occur in some wastewater collection systems, resulting in a wastewater that contains sufficient quantities of readily biodegradable substrate (particularly VFAs) to allow efficient biological phosphorus removal and denitrification. Ideal conditions for fermentation include: warm temperatures; low velocities, which minimize reaeration; and force main systems, which maintain the wastewater under anaerobic conditions and in contact with the fermentative bacteria that grow as slimes on the sewer walls. When fermentation does not occur in the wastewater collection system, the influent wastewater can be treated to convert slowly biodegradable organic matter into a more readily biodegradable form. As discussed in Section 12.1.2, fermentation is an established technology that can be used to accomplish this conversion. Either the raw wastewater itself can be fermented, or primary solids can be separated and fermented. Solids fermentation is discussed in Section 14.3.3.

12.2.4 Effluent Total Suspended Solids

The quality of the effluent from a biological wastewater treatment system is determined by the concentrations of soluble and particulate matter in it. This is particularly significant with BPR systems because of the elevated phosphorus content of the MLSS and their effect on the particulate phosphorus in the effluent. The phosphorus content of the MLSS in a BPR system will typically average about 3.5 to 6%, with values as high as 8 to 12% achievable in some cases. In contrast, conventional activated sludge will typically range from 1.5 to 2% units of phosphorus (P) per unit of volatile suspended solids (VSS). Figure 12.8 illustrates the effect that increasing the phosphorus content of the

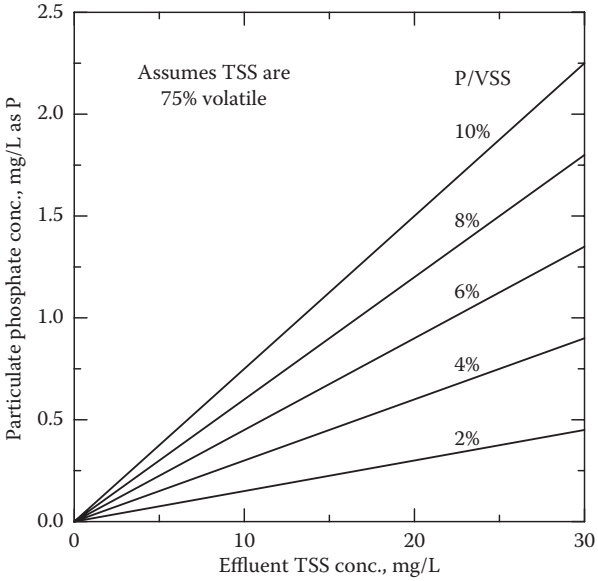

FIGURE 12.8 Effect of the effluent TSS concentration and the mixed liquor P/VSS ratio on the effluent particulate phosphorus concentration.

MLSS can have on the particulate phosphorus concentration in the effluent from a BPR system. It indicates that significant quantities of phosphate can be contributed if effluent total suspended solids (TSS) concentrations exceed about 10 mg/L. Biological nutrient removal process mixed liquor also contains organic-N, with typical values in the 10 to 12% range on an N/VSS basis. This can amount to 1 to 2 mg/L of nitrogen for effluent TSS concentrations in the 10 to 30 mg/L range. These concentrations are significant in applications where an effluent low in total nitrogen must be produced.

Fortunately, BNR systems generally produce a well-flocculated sludge that settles well in the clarifier and produces a clear effluent that is relatively low in suspended solids.[45] Nevertheless, the impact of effluent suspended solids on effluent nitrogen and phosphorus concentrations must be considered carefully when BNR systems are used.

12.2.5 Environmental and Other Factors

A number of environmental factors affect the performance of BNR systems. These factors also affect activated sludge systems, but their impacts can be more significant for BNR systems. The primary impact of temperature is on the kinetics of the various biochemical conversions, and its effect can be predicted quite well using the temperature correction factors described in Section 3.9.1. In general, temperature has the greatest impact on the nitrifying bacteria, as illustrated in Figure 10.4, but PAOs are also significantly affected.[43] Decreasing temperature will also reduce denitrification rates, resulting in the need for larger anoxic zones and/or in reduced nitrogen removal. Decreasing temperature in the collection system can reduce the rate of fermentation of organic matter and alter the composition of the wastewater entering the biological treatment system.[51,52] This impact is difficult to predict, but provides one of the major reasons that wastewater characterization or pilot plant studies should be conducted over an extended period of time. Both laboratory results and full-scale experience demonstrate that PAOs are adversely impacted by temperatures greater than about 35°C.[43,50] Thus, temperatures greater than this should be avoided if biological phosphorus removal is to be achieved. Alternately, chemical phosphorus removal should be provided as a backup system if temperatures will occasionally exceed this value.

Available data indicate that the activity of nitrifying bacteria is significantly reduced as the pH drops below 7.0, as illustrated in Figure 3.4, and that the impact of decreasing pH is much greater for them than for the PAOs or the denitrifying bacteria. Much of those data were collected in batch reactors using unacclimated cultures, but full-scale experience and some laboratory studies using acclimated cultures suggest that nitrifiers can acclimate to lower pHs, in the 6.5 to 7.0 range, with little decrease in activity.[51,61] Many nitrifying activated sludge systems operate quite successfully at pH values in this range, with efficiency dropping off only as the pH drops below 6.5. This potential should be considered in the design and operation of BNR systems, and site-specific data should be collected using acclimated cultures if this becomes a significant cost or operational issue. The pH can also adversely impact biological phosphorus removal as depressed pH (below about 7.0) favors the growth of glycogen accumulating organisms, which compete with PAOs for VFAs in the anaerobic zone.[27] Consequently, depressed pH should be avoided whenever biological phosphorus removal is to be provided.

Dissolved oxygen concentrations affect the rates of nitrification, denitrification, and phosphorus removal in a variety of ways. Enough DO must be present in the aerobic zone to allow growth of nitrifiers and PAOs at adequate rates. In general, the DO requirements for nitrifiers are controlling due to their high half-saturation coefficient for oxygen. Although a DO concentration of 2 mg/L is often specified to obtain efficient nitrification, effluent quality goals can be obtained at lower DO concentrations if aerobic SRTs are sufficiently long. This effect is illustrated in Figure 3.3 where it can be seen that many combinations of DO and ammonia-N concentrations can exist for a specified nitrifier specific growth rate (i.e., for a given system aerobic SRT). At a fixed aerobic SRT, a reduction in the DO concentration will result in an increase in the ammonia-N concentration, but the increase may be small if the aerobic SRT is sufficiently long. Consequently, DO concentrations must be evaluated on a relative basis and adjusted in accordance with system performance requirements. Operation at aerobic zone DO concentrations below 2 mg/L may result in adequate nitrification and phosphorus uptake, while also encouraging additional denitrification. Low DO operation, on the other hand, may encourage the growth of Group IV filamentous bacteria, as discussed previously.

The addition of DO to anoxic and anaerobic zones should be minimized because it is used preferentially as a terminal electron acceptor, thereby reducing the amount of readily biodegradable substrate available for denitrification or for uptake by the PAOs. If the quantity of readily biodegradable organic matter in the influent wastewater is high, larger oxygen inputs can be tolerated while still producing acceptable process performance because sufficient quantities of organic matter will still be available for the other needs. Care should be taken not to introduce too much oxygen, however, because DO concentrations will be low in those situations and significant oxygen inputs can lead to excessive growths of low DO filamentous bacteria.

In spite of the effects of oxygen entry into anoxic and anaerobic zones, both are often uncovered and simply rely on low surface turbulence to minimize oxygen transfer rates. As a consequence, care must be exercised in the selection and placement of mixers to minimize surface turbulence. In situations where interfacial oxygen transfer must be kept as low as possible, covers of various construction can be used. Since they need not exclude all air, but merely need to reduce surface transfer, opportunities exist for the design engineer to be innovative in solving the problem.

The mixing energy provided to anoxic and anaerobic zones must be sufficient to keep the MLSS in suspension, but not so great as to cause significant surface turbulence, as discussed above. This means that the mixers in such zones should be selected with care. It is more difficult to generalize about the volumetric power input required to keep solids in suspension in anoxic and anaerobic zones than in aerobic systems because it depends very heavily on the type and placement of the mixer used. Consequently, mixer selection should be done in close cooperation with a qualified vendor. Nevertheless, as a rough approximation, vertical mixers often require around 12–16 kW/1000 m^3 whereas horizontal mixers require about 8 kW/1000 m^3.

Bioreactor and aerator configurations also influence the environment produced in a BNR system. Both reaction kinetics and organism selection are improved by staging the reactor zones. This can

be done by building separate tanks, by using curtain walls in a single tank, or by a long narrow aerobic zone such as typically used with CAS. Aerator configuration will influence localized DO concentrations, as discussed previously, which can influence organism selection and process reaction rates in the aerobic zone.

12.3 PROCESS DESIGN

This section describes procedures and presents illustrative examples for the design of BNR systems. The design of these systems builds upon and utilizes all of the procedures and principles presented in Chapters 10 and 11. The major difference is that more interrelated processes are at work in BNR systems, making it impossible to derive accurate analytical expressions for them. As a consequence, the only way to accurately predict the performance of any proposed BNR system design is through simulation with a model like IWA ASM No. 1 or No. 2d. Nevertheless, it is very useful to have approximate procedures with which to estimate the required sizes of the various zones to use as starting points for simulations or to use with experience to initiate a design. The procedures and equations presented herein are provided for that purpose.

The design process for BNR systems is organized in exactly the same way as activated sludge design, with all of the same constraints and limitations, such as on mixing energy input, oxygen transfer rates, MLSS concentrations, and so on. Therefore, to save space and to allow the reader to focus on the unique aspects of a BNR process design, that material will not be repeated here. This makes it imperative for the reader to be thoroughly familiar with the material in Chapters 10 and 11 before proceeding. To provide continuity, the examples in this chapter are based on the same wastewater characteristics and parameter values used for the examples in Chapter 11. The wastewater characteristics are presented in Table E9.4, the kinetic and stoichiometric parameters are presented in Table E11.2, and the temperature correction factors are presented in Table E11.1. In addition to the parameter values given in Table E11.2, for the design of denitrification facilities, information is needed about the values of the anoxic growth and hydrolysis coefficients, η_g and η_h, which were introduced in Section 6.1.2 and Table 6.3. Those values can be readily obtained during treatability studies by using the procedures presented in Section 9.5.3. For the examples in this chapter, η_g is assumed to have a value of 0.8 and η_h is assumed to have a value of 0.4, the default values in IWA ASM No. 1.

The section is organized according to the objectives of the BNR process. First we consider processes for the removal of nitrogen, then those for the removal of phosphorus, and finally those that must remove both nitrogen and phosphorus. Table 12.5 summarizes the steps used to design BNR processes for any of these objectives.

12.3.1 Biological Nitrogen Removal Processes

Biological nitrogen removal processes require the oxidation of ammonia-N to nitrate-N through nitrification and the reduction of nitrate-N to nitrogen gas (N_2) through denitrification, thereby removing the nitrogen from the wastewater and transferring it to the atmosphere in an innocuous form. Because of the critical role that nitrification plays in nitrogen removal processes, a key decision during design is the selection of an aerobic SRT that will ensure reliable nitrification year-round. If carbon oxidation and nitrification are to occur in an ordinary activated sludge system, followed by separate stage denitrification, then the activated sludge system SRT will be equal to the aerobic SRT and design of that system proceeds exactly as outlined in Chapter 11. On the other hand, if carbon oxidation, nitrification, and denitrification are to be accomplished in a single-sludge system, then the system SRT will be larger than the aerobic SRT, with the difference being determined by the number of anoxic zones needed and the degree of nitrogen removal required. This increase in system SRT will decrease the excess solids production rate. In addition, the presence of denitrification in the system will decrease the oxygen requirement and the net destruction of

TABLE 12.5
Summary of BNR Process Design Procedure

1. Summarize process design and loading conditions including maximum, minimum, and average sustained temperature; maximum, minimum, average, and design process influent flows and pollutant loadings; and desired effluent quality.
2. Where necessary, convert process loadings, expressed in conventional parameters (such as BOD_5) into the units used in the process design (such as biodegradable COD).
3. Select the process configuration. This includes determination of the number and type (anaerobic, anoxic, aerobic) of zones required to provide the necessary process functions as defined in Table 12.1.
4. Select the design SRTs for the individual process zones. The total SRT is the sum of the SRTs of the individual zones.
5. If nitrification will occur, calculate the nitrate produced.
6. If denitrification is to be provided, calculate the nitrate removed in each of the anoxic zones. Size necessary recirculation streams.
7. Calculate the net alkalinity consumption considering both nitrification and denitrification. Compare to available alkalinity, recognizing the need for an effluent of at least 50 mg/L as $CaCO_3$.
8. Calculate the steady-state oxygen requirement for maximum, minimum, and average sustained temperature conditions based on the CMAS configuration.
9. Calculate the diurnal maximum and minimum oxygen requirements for the conditions above.
10. Determine the range of allowable bioreactor volumes based on the CMAS model. The minimum volume can be limited either by the maximum achievable volumetric oxygen transfer rate or by floc shear. The maximum bioreactor volume will be limited by mixing. In some cases it may be necessary to compromise as not every condition can be accommodated.
11. Using the maximum and minimum volumes determined in step 7, calculate the maximum and minimum MLSS concentrations using the CMAS model.
12. Considering cost trade-offs between the bioreactor and the final clarifier (or membrane system), choose the MLSS concentration within the allowable range and calculate the associated bioreactor volume.
13. Calculate the waste sludge production rate using the CMAS model.
14. For processes with MLSS concentrations that vary through the bioreactor, such as UCT, VIP, and JHB, calculate the MLSS distribution.
15. For processes with variations in oxygen requirements, calculate the distribution of oxygen requirements through the bioreactor.
16. If biological phosphorus removal is to be provided, estimate effluent total phosphorus concentration by estimating the effluent soluble phosphorus, the phosphorus content of the mixed liquor suspended solids, and the effluent particulate phosphorus.
17. Based on the above, make any necessary adjustments in the process design and summarize the results in tabular form.

alkalinity. In this section we will first review the requirements for nitrification, and then we will examine the additional requirements for design of an anoxic selector, a MLE system, a four-stage Bardenpho system, and a separate-stage denitrification system.

12.3.1.1 Nitrification

The factors that must be considered in the selection of the aerobic SRT required to achieve stable nitrification are covered in Sections 11.2.2 and 11.3.2. For a fully aerobic system to be followed by separate-stage denitrification, the process design proceeds from there exactly as outlined in Section 11.3. In this section, we consider the design of a fully aerobic system.

Because nitrification, which destroys alkalinity, is required in all nitrogen removal systems and the activity of nitrifying bacteria is sensitive to pH, consideration must be given to the need for

chemical addition for pH control. In general, a residual alkalinity of 1 mM (50 mg/L as $CaCO_3$) will result in an adequate pH value. Thus, the need for chemical addition will depend on the influent alkalinity and the amount of ammonia-N oxidized. As shown in Equation 3.30, 6.71 g HCO_3^- are removed for each g of NH_4^+ removed. This corresponds to 7.07 g of alkalinity (expressed as $CaCO_3$) per g of ammonia-N removed by nitrifying bacteria. Because some of the ammonia-N removed is incorporated into the nitrifying biomass (Equation 3.30), only 0.98 g nitrate-N is formed for each g of ammonia-N removed. Consequently, the amount of alkalinity destroyed corresponds to 7.23 g of alkalinity (as $CaCO_3$) per g nitrate-N formed. This latter value is sometimes useful when determining the alkalinity destruction in a system performing both carbon oxidation and nitrification because ammonia-N is also removed for incorporation into heterotrophic biomass, thus making it simpler to determine the alkalinity requirement from the amount of nitrate-N formed. The concentration of nitrate-N formed, S_{NO}, is determined by the difference between the concentration of nitrogen available to the nitrifiers, $S_{N,a}$, as given by Equation 11.17, and the effluent soluble nitrogen concentration:

$$S_{NO} = 0.98(S_{N,a} - S_{NH} - S_{NS}), \tag{12.5}$$

where the factor 0.98 is the mass of nitrate-N formed per mass of ammonia-N oxidized, as discussed above. The effluent ammonia-N concentration, S_{NH}, can be calculated with Equation 11.19 for a completely mixed activated sludge system, but will generally be lower for other systems, as discussed in Section 11.3.2 and illustrated in Figure 11.8. If the bioreactor has plug-flow characteristics, S_{NH} will often be negligible. The effluent soluble biodegradable organic nitrogen concentration, S_{NS}, must either be measured directly or estimated by assuming that the ratio of soluble biodegradable organic nitrogen to soluble biodegradable organic matter in the effluent is the same as the ratio in the influent. For domestic wastewater, the value of S_{NS} is often assumed to be negligible. It should be noted that the concentration of nitrogen available to the nitrifiers, $S_{N,a}$, will change with temperature because the heterotrophic nitrogen requirement, as given by Equation 5.46, depends on the observed yield, which is a function of temperature. The value of $S_{N,a}$ will be slightly larger in summer, making the effluent nitrate-N concentration, the amount of alkalinity destroyed, and the amount of oxygen required slightly higher then.

If a nitrifying system is to be configured as a CAS system, in which the oxygen requirement must be distributed along the bioreactor, special consideration must be given to the distribution of the autotrophic oxygen requirement. During the winter, when temperatures are low, the maximum nitrification rate will be low and nitrification will occur throughout a greater proportion of the bioreactor than in the summer, when the maximum nitrification rate is higher. The required distribution under each condition can be determined by applying the procedures of Section 11.3.4 as illustrated in Example 11.3.4.1. Because two different oxygen transfer systems would be needed to meet both oxygen distributions, the potential benefits of two systems must be carefully weighed against the additional expense incurred. During the summer, if the oxygen supply is not limiting, nitrification will generally be complete within the first half (or less) of the system. However, if oxygen were supplied according to the winter distribution (but in total amount sufficient to meet the higher summer requirement), the result would be a decreased DO concentration in the first part of the bioreactor. Because of the sensitivity of nitrification to the DO concentration, this would decrease the nitrification rate in the first part of the bioreactor, increasing the ammonia loading on the latter part, and forcing more of the oxygen requirement to it, where oxygen is available. As long as the DO concentration in the first part of the system is maintained sufficiently high to avoid problems with low DO filaments, such a procedure is entirely acceptable. Consequently, many designers specify only a single oxygen transfer system.

The examples presented in Chapter 11 have considered all of the aspects of the design of a system for combined carbon oxidation and nitrification, except for the determination of the amount of

nitrate-N produced and the need for pH control. The following example illustrates those procedures. In addition, it presents the determination of the required SRT and results from determination of the steady-state oxygen requirement, solids wastage rate, and required bioreactor volume because they will be used as baselines against which to make comparisons for the remainder of the examples in this section. As in the examples in Chapter 11, all organic substrate concentrations are given as biodegradable COD and all suspended matter concentrations are as TSS, unless noted otherwise.

Example 12.3.1.1

Consider the wastewater used for all examples in Chapter 11, with characteristics given in Table E9.4, kinetic and stoichiometric parameters in Table E11.2, and temperature correction factors in Table E11.1. As in Chapter 11, the average wastewater flow rate is 40,000 m^3/day and the diurnal peak pollutant loadings are 2.5 times the average. The characterization of the wastewater is considered to be sufficient to allow the safety factor for uncertainty, ς_U, to be set to 1.0. Determine the major design characteristics of an activated sludge system to perform both carbon oxidation and nitrification year round, while maintaining an MLSS concentration of 3000 mg/L and a DO concentration of 2.0 mg/L. A conventional bioreactor with hydraulic characteristics equivalent to four tanks in series is to be used.

a. What is the design SRT?
The design SRT should be chosen for winter operating conditions. The approach used is exactly the same as in Example 11.3.2.1, except that the peak load safety factor, ς_{PL}, is 2.5. Since a CAS system is to be used, Equation 11.11 is the appropriate expression with which to calculate the required SRT. The value of the minimum SRT can be obtained from Figure 10.4, which gives a value of 3.5 days at a temperature of 15°C. The safety factor for uncertainty, ς_U, has a value of 1.0. Because the half-saturation coefficient for DO for nitrifiers is 0.75 mg/L (Table E11.2), the DO safety factor can be calculated with Equation 11.9, giving a value of 1.375. Thus,

$$\Theta_{c,r} = (3.5)(1.0)(2.5)(1.375) = 12.0 \text{ days.}$$

b. What concentration of nitrate-N will be formed?
This will be calculated for summer conditions since it will be slightly larger then because more nitrogen will be available to the nitrifiers. The concentration of nitrate-N formed can be calculated with Equation 12.5. Since the bioreactor has hydraulic characteristics equivalent to four tanks in series, the concentrations of ammonia-N and soluble organic-N in the effluent can be considered to be negligibly small, even in winter. Thus, the concentration of nitrate-N formed will just be 98% of the concentration of nitrogen available to the nitrifiers, as calculated with Equation 11.17. Use of Equation 11.17 requires knowledge of the nitrogen requirement of the heterotrophs, which can be calculated with Equation 5.46. This calculation was illustrated in part c of Example 11.3.3.2 and will not be repeated here. However, it should be noted that the heterotrophic nitrogen requirement for the SRT of 12 days is 0.022 mg N used/mg COD removed at 25°C, which is lower than the value in Example 11.3.3.2 because the longer SRT results in less net biomass synthesis. Therefore, the available nitrogen concentration is

$$S_{N,a} = 25 + 6.5 + 8.5 - 0.022(115 + 150)$$
$$= 34.2 \text{ mg/L as N.}$$

Thus, the amount of nitrate-N formed in summer will be (0.98)(34.2) = 33.5 mg/L. Slightly less will be formed in winter because the heterotrophic nitrogen requirement will be slightly higher because of the decreased decay coefficient.

c. Will it be necessary to provide alkalinity to control pH in the process?
The wastewater contains 200 mg/L as $CaCO_3$ of alkalinity (Table E9.4) and the residual alkalinity must be 50 mg/L as $CaCO_3$ to provide a stable pH. Thus, the alkalinity available to neutralize the hydrogen ions formed during nitrification is 150 mg/L as $CaCO_3$. As noted above, nitrification destroys 7.23 g of alkalinity (as $CaCO_3$) per g nitrate-N formed. Since 33.5 mg/L of nitrate-N are formed:

$$\text{Alkalinity destroyed} = (7.23)(33.5) = 242 \text{ mg/L as } CaCO_3.$$

This exceeds the alkalinity available by 92 mg/L. Thus, 92 mg/L of alkalinity as $CaCO_3$ must be provided to the process through chemical addition in the summer. The amount required in winter will be slightly less because slightly less nitrate-N will be formed.

d. What is the required bioreactor size?
As in other cases, the bioreactor size must be calculated for winter conditions because that is when solids production is highest. The first task is to calculate the mass of MLSS in the system at 15°C with Equation 10.8, as illustrated in part b of Example 11.3.3.3. The mass of autotrophic nitrifiers can be neglected because it is small, as seen in Chapter 11. The only difference in the calculation is the use of an SRT of 12 days. The resulting mass of MLSS is 51,800,000 g.

The MLSS concentration for this design is 3000 mg/L, which is a higher value than was possible in the Examples in Chapter 11. This is permissible here because once nitrification is fully established, the mass of MLSS in the system increases faster than the oxygen requirement as the SRT is increased. Thus, the constraints due to oxygen transfer rates are encountered at higher MLSS concentrations. For an MLSS concentration of 3000 mg/L (g/m^3), the bioreactor volume will be

$$V = \frac{51{,}800{,}000}{3000} = 17{,}300 \text{ m}^3.$$

e. What is the average process oxygen requirement (RO) for design of the oxygen transfer system?
This is determined at the highest sustained wastewater temperature of 25°C because that is when the oxygen requirement is maximum. This requires calculation of the heterotrophic oxygen requirement with Equation 10.10 and the autotrophic oxygen requirement with Equation 11.16. The procedure was illustrated in Example 11.3.3.2, so it will not be repeated here. However, application of the procedure results in a heterotrophic oxygen requirement (RO_H) of 7930 kg O_2/day and an autotrophic oxygen requirement (RO_A) of 6080 kg O_2/day, for a total requirement (RO) of 14,010 kg O_2/day.

h. What is the excess solids production rate, $W_{M,T}$?
Because the maximum solids production occurs in the winter, this is calculated at 15°C using Equation 10.9, as illustrated in Example 11.3.3.4. Application of the equation for the conditions of this design gives a value for $W_{M,T}$ of 4310 kg TSS/day.

12.3.1.2 Design of an Anoxic Selector

An anoxic selector is an initial anoxic zone similar to that used in the MLE (Figure 1.13) and four-stage Bardenpho processes (Figure 1.14). As in those processes, it receives the influent wastewater and the recirculated flow from a downstream aerobic zone where nitrification occurs. The influent wastewater contains biodegradable organic matter, which serves as the electron donor for the growth of heterotrophic denitrifying bacteria, whereas the recirculated flow from the aerobic zone provides nitrate-N, which serves as the electron acceptor for those bacteria. Unlike the initial anoxic zone in the MLE and four-stage Bardenpho processes, however, the primary purpose of an anoxic selector is the control of solids settleability. As discussed in Section 11.2.1, many filamentous bacteria use readily biodegradable organic matter very efficiently, but most are not able to use nitrate-N

as an electron acceptor. Consequently, if the readily biodegradable organic matter in a wastewater is removed in an anoxic zone, the growth of many types of filamentous bacteria is prevented. Moreover, since readily biodegradable organic matter can be removed easily, a relatively small anoxic zone can be used. Initial anoxic zones designed to control the growth of filamentous bacteria are called anoxic selectors because their primary purpose is to select for floc forming bacteria and to selectively prevent the growth of filamentous bacteria.

Just as with aerobic selectors, the purpose of an anoxic selector is to remove the bulk of the readily biodegradable substrate so that little passes through to the aerobic zone where filamentous bacteria have a competitive advantage because of the long SRT and associated low substrate concentration. Few guidelines are available concerning the readily biodegradable substrate concentration that should leave the selector, so it is difficult to specify the specific growth rate that should be maintained there. Rather, most design experience has resulted from the application of a particular process loading factor, as has been done with aerobic selectors. Values that have been used successfully are 0.8 to 1.2 kg BOD_5/(kg MLSS·day) for operation at temperatures above about 20°C and 0.7 to 1.0 kg BOD_5/(kg MLSS·day) for operation at temperatures below about 17°C.[44] These correspond to biodegradable COD loadings of approximately 1.4 to 2.1 kg COD/(kg MLSS·day) and 1.2 to 1.7 kg COD/(kg MLSS·day), respectively. Thus, design of an anoxic selector could be approached in the same way as design of an aerobic selector, as outlined in Section 11.3.4, except that a single mixed basin is often used. However, since the total SRT is not known at the outset, an iterative procedure is required. This can be avoided if the anoxic SRT can be specified as the design criterion. Reexamination of the information leading to the process loading factors given above reveals that they correspond to anoxic SRTs of about 1.0 day at temperatures over about 20°C and of about 1.5 days at temperatures less than about 17°C. Thus, in the absence of site-specific data, these SRTs can be used to perform a preliminary process design. If additional selective pressure is desired, then the anoxic selector can be staged by configuring it as multiple cells in series as is done for aerobic selectors as described in Section 11.3.4. A one-quarter, one-quarter, one-half configuration is often used.[35,59]

The determination of the bioreactor size proceeds in the same manner as for activated sludge systems as discussed in Chapter 11. The anoxic SRT is added to the aerobic SRT, giving the total SRT to be used for the design. The mass of MLSS in the system is then calculated using the total SRT in Equation 10.8, and it is apportioned to the anoxic and aerobic zones in proportion to the anoxic and aerobic zone SRTs. Since the MLSS concentration is uniform in this system, the total system volume can be calculated after deciding upon an MLSS concentration that will give a reasonable settler size, as discussed in Chapters 10 and 11. Finally, that volume can be allocated to the two zones in proportion to their respective SRTs, as suggested by Equations 12.1 and 12.2 for systems with uniform MLSS concentrations.

Once the anoxic SRT has been selected, the next task is to determine the amount of nitrate-N that must be recirculated to the anoxic zone by the solids recycle and MLR flows. The required quantity must be equal to the amount of electron acceptor that is needed for biomass synthesis and decay. Because this will be greatest at a higher temperature, the computation should be made for summer conditions. As discussed in Section 6.4.2, the amount of electron donor required to reduce a given amount of nitrate-N to nitrogen gas ($\Delta S/\Delta N$) is a function of the SRT in the anoxic zone and it can be calculated using Equation 6.4 for situations in which no oxygen enters the system. Since oxygen is used preferentially over nitrate-N as an electron acceptor, the oxygen equivalents of nitrate-N required will be decreased in direct proportion to the amount of oxygen entering the bioreactor. Even though all of the design equations assume no oxygen leakage, this effect should be kept in mind. During selection of the RAS and MLR flows we are interested in the amount of nitrate-N required, so the equation should be inverted. Making this modification and using only the anoxic SRT in the equation gives the expression for the amount of nitrate-N (ΔN) required to remove a given amount of COD (ΔS):

$$\frac{\Delta N}{\Delta S} = \frac{1 + (b_H \cdot \Theta_{c,ANX}) - (Y_{H,T})(i_{O/XB,T})(1 + f_D \cdot b_H \cdot \Theta_{c,ANX})}{2.86[1 + (b_H \cdot \Theta_{c,ANX})]}. \quad (12.6)$$

All other symbols have been defined previously. The larger the amount of COD removed in the anoxic selector, the larger the amount of nitrate-N that must be provided, and the larger the MLR flow rate must be. Since the anoxic selector has been sized to remove the bulk of the readily biodegradable organic matter, the least acceptable MLR flow rate will result when ΔS is taken to be the mass flow rate of readily biodegradable substrate into the system, $F \cdot S_{SO}$. However, it is likely that some of the slowly biodegradable substrate will also be oxidized, with the amount depending on the SRT of the system and the nitrate-N concentration, as indicated in process 7 of Table 6.1. Hydrolysis of slowly biodegradable organic matter under anoxic conditions is thought to be slower than aerobic hydrolysis, as reflected by the fact that a typical value of η_h, the anoxic hydrolysis factor, is 0.4, as given in Table 6.3. Because of this and because the kinetics of hydrolysis are still not well defined, it is difficult to predict with certainty the degree of hydrolysis that will occur in an anoxic selector. Furthermore, as discussed previously, it is difficult to know exactly how much oxygen will enter the anoxic zone. Since hydrolysis of slowly biodegradable substrate increases the nitrate-N requirement over the amount calculated with Equation 12.6 and oxygen entry decreases it and since both are hard to predict, the recommended procedure is to limit ΔS in Equation 12.6 to readily biodegradable substrate and to consider the resulting MLR flow rate to be the minimum allowable value. A value in slight excess of that can be adopted for the design, depending on the experiences and preferences of the design engineer. Care should be exercised in the selection of the MLR flow rate, however, because incomplete denitrification will occur whenever the electron accepting capacity of the recirculated nitrate exceeds the amount of electron donor available. Such a situation leads to the release of nitrous oxide, an important greenhouse gas, and should be avoided.[33,34,38,54]

Because the concentration of nitrate-N in the MLR stream depends on several factors, it is easiest to determine the MLR flow rate by considering the fraction of nitrate-N formed that must be denitrified. In the absence of denitrification, the concentration of nitrate-N in the process effluent would be given by Equation 12.5. Multiplication of that concentration by the flow rate through the system, F, gives the mass formation rate of nitrate-N in the system. Therefore, since ΔN represents the mass of nitrate-N that is denitrified, the fraction of the nitrate-N denitrified in the initial anoxic zone, $f_{NO,D}$, is

$$f_{NO,D} = \frac{\Delta N}{0.98F(S_{N,a} - S_{NH} - S_{NS})}. \quad (12.7)$$

Furthermore, for the flow diagram depicted in Figure 1.13, the fraction of the nitrate-N produced in the aerobic zone that is recirculated to the anoxic zone ($f_{NO,R}$) can be calculated by simple mass balance. The result is

$$f_{NO,R} = \frac{\alpha + \beta}{1 + \alpha + \beta}, \quad (12.8)$$

where β is the mixed liquor recirculation ratio and α is the solids recycle ratio. The desired MLR rate results when the fraction of nitrate-N recirculated is equal to the fraction denitrified. Higher recirculation rates will result in higher nitrate-N concentrations in the anoxic zone and the potential for nitric and nitrous oxide production.[33,34,38,54] Lower recirculation rates will allow readily biodegradable substrate to pass through into the aerobic zone, where filamentous bacteria can compete effectively for it. Substitution of $f_{NO,D}$ for $f_{NO,R}$ in Equation 12.8 and rearrangement gives an

expression that can be used to calculate the MLR and RAS flow rate ratios to achieve denitrification of a specified fraction of the nitrate-N formed:

$$\alpha + \beta = \frac{f_{NO,D}}{1 - f_{NO,D}}. \tag{12.9}$$

While solids recycle from the final clarifier is done by pumping through a pipe, this may not be necessary for mixed liquor recirculation, depending on the system configuration. If the exit from the aerobic zone is located some distance from the anoxic tank, then MLR is usually pumped through a pipe. On the other hand, if the anoxic and aerobic zones share a common wall, then it may be possible to recirculate mixed liquor by using a propeller pump in an opening in the common wall. Such pumps are similar to the horizontal mixer shown in Figure 12.2b. They are low head devices and work with lower power requirements than centrifugal pumps.

As discussed in Section 6.4.1 and shown in Table 6.1, 1/14 moles of alkalinity are produced for each gram of nitrate-N reduced to N_2. This amounts to 3.5 g of alkalinity (as $CaCO_3$) per g of nitrate-N reduced. The production of this alkalinity in the anoxic zone helps to offset somewhat the destruction of alkalinity due to nitrification in the aerobic zone. This can be another important benefit from using an anoxic selector. There will be a slight difference in the alkalinity recovery under summer and winter conditions, just as there will be for alkalinity destruction. Generally, however, the estimation is not precise enough to warrant recalculating it for both conditions. Either can be used to give a general idea.

The oxidation of biodegradable organic matter in the anoxic zone decreases the amount of substrate entering the aerobic zone, causing a corresponding decrease in the oxygen requirement. The decrease in the aerobic zone oxygen requirement can be calculated using one of two general procedures. In one, the oxygen requirement associated with the synthesis of biomass in the aerobic zone is reduced by the amount of substrate that is oxidized in the anoxic zone. However, it should be noted that the oxygen requirement associated with decay of the biomass resulting from that substrate is not reduced. In the other, the nitrate-N reduced in the anoxic zone is converted to its oxygen equivalent (using the stoichiometric factor –2.86 g COD/g NO_3-N as shown in Table 3.1) and subtracted from the aerobic zone oxygen requirement calculated assuming that the anoxic zone is not present. In making this computation, the heterotrophic oxygen requirement should be calculated as if the total SRT were aerobic to account for the added decay, but the autotrophic requirement should be calculated only for the aerobic SRT. The two procedures will arrive at the same answer.

The procedure outlined above for the design of an anoxic selector for the removal of readily biodegradable substrate is approximate because of uncertainties about the amount of slowly biodegradable substrate used in the anoxic tank and the amount of oxygen leakage into the anoxic zone. In addition, it contains several approximations. Nevertheless, it should give sufficiently accurate estimates to allow a pilot study to be designed or to provide starting conditions for a series of simulations to refine the design. Its use is illustrated below.

Example 12.3.1.2

An anoxic selector is to be added to the nitrifying activated sludge system designed in Example 12.3.1.1. The purpose of this selector is to remove the readily biodegradable organic matter so that solids with good settling properties will be produced. Determine the approximate requirements for the system.

a. What size should the anoxic selector be?
   As discussed above, experience suggests that the readily biodegradable organic matter can be removed from a typical domestic wastewater under cold weather conditions using an

SRT for the anoxic zone of 1.5 days. Thus, since the aerobic SRT is 12.0 days, the total SRT will be 13.5 days. Using Equation 10.8 to recalculate the mass of MLSS in the system at 15°C for this slightly larger SRT gives a value of 56,700,000 g. At the design MLSS concentration of 3000 mg/L (g/m^3), the total bioreactor volume will be

$$V = \frac{56{,}700{,}000}{3000} = 18{,}900 \text{ m}^3.$$

Because the MLSS concentration is uniform, the volume of the anoxic zone is equal to the total bioreactor volume times the fraction of the total SRT associated with the anoxic zone, as suggested by Equations 12.1 and 12.2. Thus:

$$V_{ANX} = 18{,}900\left(\frac{1.5}{13.5}\right) = 2100 \text{ m}^3.$$

By difference:

$$V_{AER} = 18{,}900 - 2{,}100 = 16{,}800 \text{ m}^3.$$

In comparison to Example 12.3.1.1, the aerobic zone is slightly smaller due to the slightly larger total SRT, which results in additional decay.

b. What should the MLR flow rate be?
First, the mass of nitrate-N required to accept the electrons from use of the readily biodegradable substrate at the anoxic SRT is calculated using Equation 12.6. This calculation is performed at 25°C as it represents the greatest value and will dictate the MLR pump sizing. Using the values for the various parameters from Table E11.2 and an anoxic SRT of 1.5 days gives:

$$\frac{\Delta N}{\Delta S} = \frac{1+(0.22)(1.5)-(0.5)(1.2)[1+(0.2)(0.22)(1.5)]}{(2.86)[1+(0.22)(1.5)]}$$

$$= 0.182 \text{ g nitrate-N/g COD.}$$

Since the concentration of readily biodegradable organic matter in the influent wastewater is 115 mg COD/L (g/m^3) and the influent flow is 40,000 m^3/day, the mass of nitrate-N required is

$$\Delta N = (0.182)(115)(40{,}000) = 837{,}000 \text{ g nitrate-N/day.}$$

The fraction of nitrate-N denitrified in the anoxic selector can be calculated with Equation 12.7. The value of available nitrogen, $S_{N,a}$, should be recalculated for 25°C and an SRT of 13.5 days because the longer SRT will reduce the nitrogen requirement for the heterotrophs. Following the procedure in Examples 11.3.3.2 and 12.3.1.1 gives a value of 0.021 mg N used/mg COD removed for the heterotrophic nitrogen requirement, which gives a value for $S_{N,a}$ of 34.4 mg/L as N. Substituting this into Equation 12.7 and assuming that the effluent concentrations of ammonia-N and soluble organic-N are negligible, gives:

$$f_{NO,D} = \frac{837{,}000}{(0.98)(40{,}000)(34.4)} = 0.62.$$

The sum of the recycle and recirculation ratios can now be calculated with Equation 12.9:

$$\alpha + \beta = \frac{0.62}{1-0.62} = 1.63.$$

Because some slowly biodegradable organic matter may be hydrolyzed and oxidized in the anoxic zone, the sum of the selected ratios should exceed 1.63, with higher values being used for systems that exclude more oxygen. Assuming an anoxic zone with little oxygen entry, use a value of 1.5 for β, the MLR ratio, which, when combined with typical values for α of 0.5 to 1.0, will provide a conservatively high recirculation of nitrate-N to the anoxic zone. The MLR flow rate will thus be (1.5) (40,000 m^3/day) or 60,000 m^3/day. The capability should be provided for adjusting the MLR flow during operation to provide the correct amount of nitrate to the anoxic selector. Excessive amounts will lead to the formation of nitrous oxides, as discussed above, whereas insufficient amounts will allow readily biodegradable substrate to pass through the selector to the aerobic bioreactor.

c. What will the effluent nitrate-N concentration be at 25°C?
Since the available nitrogen concentration is 34.4 mg/L as N, the concentration of nitrate-N formed is (0.98)(34.4), or 33.7 mg/L as N. At least 62% of this will be destroyed by denitrification, leaving an effluent concentration of less than 12.8 mg/L of nitrate-N. The exact value will depend on the amount of slowly biodegradable substrate used in the anoxic selector and the amount of oxygen entering it.

d. What is the effect of denitrification on the alkalinity requirement at 25°C?
Using the logic of part c of Example 12.3.1.1, production of 33.7 mg/L of nitrate-N will destroy 244 mg/L as $CaCO_3$ of alkalinity. Since 150 mg/L of alkalinity are available in the wastewater, it would be necessary to add 94 mg/L in the absence of denitrification. However, destruction of 20.9 mg/L of nitrate-N by denitrification produces (3.5)(20.9) = 73 mg/L as $CaCO_3$ of alkalinity, thereby reducing the amount of alkalinity that must be provided by chemical addition to 21 mg/L.

e. By what percentage is the average oxygen requirement in the aerobic zone decreased by the presence of the anoxic selector at 25°C?
The reduction in the oxygen requirement in the aerobic zone can be approximated by subtracting the oxygen equivalents of the amount of nitrate reduced from the oxygen requirement for a totally aerobic system, as calculated with an SRT of 13.5 days for the heterotrophs and 12.0 days for the autotrophs. The autotrophic oxygen requirement, RO_A, will be slightly larger than calculated in part e of Example 12.3.1.1 because the available nitrogen concentration is slightly higher due to the reduced nitrogen requirement for the heterotrophs associated with the additional SRT. Using a value of 34.4 mg/L for $S_{N,a}$ and an SRT of 12 days in Equation 11.16 gives an autotrophic requirement of 6120 kg O_2/day. Likewise, using an SRT of 13.5 days in Equation 10.10 gives a value of 8050 kg O_2/day for the heterotrophic requirement. Thus, the total requirement would be 14,170 kg O_2/day, from which would be subtracted the mass removal of nitrate-N, expressed as oxygen equivalents. From part b above, the mass of nitrate-N reduced is 837 kg/day. This is equivalent to (2.86)(837) = 2390 kg O_2/day. Thus, the total oxygen requirement is 11,780 kg O_2/day, which is 84% of the oxygen requirement in the totally aerobic system of Example 12.3.1.1.

f. What is the WAS production rate?
This is calculated at 15°C as illustrated in part h of Example 12.3.1.1 using Equation 10.9 and an SRT of 13.5 days. The result is 4200 kg TSS/day, which is 97% of the quantity to be disposed of from the totally aerobic system of Example 12.3.1.1.

12.3.1.3 Design of an MLE System to Achieve a Desired Effluent Nitrate-N Concentration

The design of an MLE system to achieve a particular effluent nitrate-N concentration usually involves degradation of some of the slowly biodegradable substrate in the anoxic zone. Because of the interactions between the reaction rates and various concentrations in the system, the only way to accurately predict the performance of such a system is by simulation with a model like IWA ASM No. 1. Various anoxic SRTs and MLR rates can be tried in a systematic manner until a suitable design is developed. In the absence of complete kinetic data, or when a preliminary design is needed as input to a simulation program, an alternative approach is required, just as it was for an anoxic selector. Although the preceding discussion deals with the design of anoxic selectors for the removal

of readily biodegradable substrate, the same general procedure can be used when a particular effluent nitrate-N concentration must be achieved. The primary difference is that the initial anoxic zone must be sized to remove a specified mass of nitrate-N rather than a specific mass of biodegradable organic matter.

The estimation of the organic substrate utilization and associated nitrate-N reduction for an MLE system requires two steps, one for readily biodegradable substrate, and one for slowly biodegradable substrate. First, because the anoxic SRT usually exceeds 1.5 days, we can assume that all readily biodegradable substrate will be used, even in the winter. Thus, we can calculate the nitrate-N reduction associated with that activity by using a $\Delta N/\Delta S$ value. However, because the approach we will use to estimate the nitrate-N reduction associated with slowly biodegradable substrate utilization includes the effects of biomass decay, the reduction of nitrate-N associated with the removal of readily biodegradable substrate should consider only the electron acceptor requirement for biomass synthesis. Using Equation 11.24 as a guide, the nitrate-N utilization for this case, ΔN_{SS}, is

$$\Delta N_{SS} = \Delta S\left[\frac{1-(Y_{H,T})(i_{O/XB,T})}{2.86}\right], \tag{12.10}$$

in which ΔS is the mass rate of readily biodegradable substrate entry to the system. In other words, Equation 12.10 came from eliminating all of the decay-associated terms in Equation 12.6. Next, an empirical expression that has been widely used to size initial anoxic zones[55,61] can be used to estimate the nitrate-N utilization associated with biodegradation of slowly biodegradable substrate and biomass decay. This expression was developed with domestic wastewaters that contained high proportions of slowly biodegradable organic matter,[12] and thus is appropriate for estimating the nitrate-N utilization associated with hydrolysis and oxidation of that type of substrate in this context. It allows direct calculation of the specific nitrate-N utilization rate, $q_{NO/XS}$, as a function of the process loading factor for the anoxic zone, U_{ANX}, when that loading factor is expressed in terms of slowly biodegradable substrate alone. In other words, since the effects of readily biodegradable substrate utilization are calculated with Equation 12.10, readily biodegradable substrate should not be included in U_{ANX}. The expression is

$$q_{NO/XS} = 0.018U_{ANX} + 0.029. \tag{12.11}$$

The units of the specific nitrate-N utilization rate are mg NO_3-N/(mg MLSS·day). The units of the process loading factor in Equation 12.11 are mg COD/(mg MLSS·day). In the original equation given by Burdick et al.,[12] the slope was 0.030 and the units of the process loading factor were mg BOD_5/(mg MLSS·day). For consistency within this text, Equation 12.11 is expressed in biodegradable COD units by using Equation 9.31 to convert BOD_5 to equivalent biodegradable COD. The relationship of Burdick et al.[12] was developed at 20°C and $q_{NO/XS}$ can be adjusted to other temperatures by using Equation 3.99 with an appropriate value of the temperature coefficient, θ, in the range of 1.04 to 1.08. It was developed using domestic wastewater and should not be used for a wastewater containing a high percentage of industrial wastes. In order to use Equation 12.11, the mass of MLSS in the anoxic zone must be known. This requires the use of an iterative approach to size the anoxic zone for an MLE system.

The approach to preparing a preliminary design of an MLE system is as follows. The selection of the anoxic SRT must be done for winter temperatures because that is when denitrification rates are slowest. First, an initial value for the anoxic SRT must be selected and added to the aerobic SRT to give the total. The total SRT is then used to calculate the nitrogen available to the nitrifiers using Equation 11.17, which in turn is used to calculate the mass rate of nitrate-N formation by multiplying Equation 12.5 by the influent flow rate, F. The mass rate of nitrate-N reduction by biomass synthesis

on readily biodegradable substrate, ΔN_{SS}, can be estimated with Equation 12.10 from the known mass input rate of readily biodegradable substrate for ΔS. Using the total SRT, the mass of MLSS in the system $(X_{M,T} \cdot V)_{System}$ is calculated with Equation 10.8. The mass of MLSS in the anoxic zone can then be calculated with Equation 12.2. Once it is known, the process loading factor on the anoxic zone due to slowly biodegradable substrate can be calculated from its definition, Equation 5.48, in which slowly biodegradable substrate is the only substrate considered. This allows estimation of the specific denitrification rate in the anoxic zone with Equation 12.11. Multiplication of it by the mass of MLSS in the anoxic zone gives the mass rate of nitrate-N reduction by slowly biodegradable substrate and decay, ΔN_{XS}. Summing ΔN_{SS} and ΔN_{XS} gives the total mass rate of nitrate-N reduction. Subtraction of this value from the mass production rate of nitrate-N gives the mass rate at which nitrate-N is leaving the process, from which the effluent nitrate-N concentration can be calculated when dividing by the flow rate. Repetition of these calculations for a number of anoxic SRTs provides information on the effluent nitrate-N concentration as a function of anoxic SRT, from which the anoxic SRT can be selected. Once that is fixed, the system volume is determined by considering the mass of MLSS in the system under winter temperature conditions and the design MLSS concentration, which is uniform throughout. The design MLR rate can be calculated with Equation 12.9 using the sum of ΔN_{SS} and ΔN_{XS} for ΔN. This should be done for the summer temperature, since it will give the largest value. However, as mentioned previously, the need for chemical addition to control pH can be calculated for either winter or summer temperatures, because it is not precise enough to distinguish between the needs under the two conditions. The rest of the design can be accomplished exactly as in previous cases, with summer conditions determining oxygen requirements and winter conditions excess solids production. This procedure is illustrated in the following example.

Example 12.3.1.3

The process design developed in Example 12.3.1.2 is being revised to produce an effluent with a nitrate-N concentration of 6 mg/L or less. Determine the approximate requirements for the system. Assume that the temperature coefficient for the specific denitrification rate is 1.05.

a. What anoxic zone SRT is required for this application?
The determination of the anoxic SRT must be made at 15°C since the effluent goals must be met under all conditions and low temperatures lead to the lowest rates of denitrification. Since the effluent nitrate-N concentration in Example 12.3.1.2 was 12.8 mg/L as N when the anoxic SRT was 1.5 days, and since anoxic hydrolysis of slowly biodegradable substrate is slow, it is likely to require a considerably longer anoxic SRT to achieve an effluent nitrate-N concentration of 6.0 mg/L as N or less. Thus, select an anoxic SRT of four days as an initial guess. This will give a total system SRT of 16 days since the aerobic SRT is 12 days.

The first task is to calculate the concentration of nitrogen available to the nitrifiers when the SRT is 16 days. This is done exactly as in part b of Example 12.3.1.1, except that it must be done for a temperature of 15°C, giving a value of 34.0 mg/L.

The second task is to estimate the mass rate of nitrate-N production. This is done by multiplying Equation 12.5 by the influent flow rate. Because of the configuration of the aerobic zone, we can consider the effluent ammonia-N and organic-N concentrations to be negligible. Thus, in the absence of denitrification, the effluent nitrate-N concentration would be 33.3 mg/L as N and the mass rate of nitrate-N production would be 1,332,000 g/day.

The mass rate of nitrate-N reduction by heterotrophic biomass synthesis on readily biodegradable substrate, ΔN_{SS}, can be calculated with Equation 12.10 by making use of the fact that ΔS is the mass input rate of readily biodegradable substrate, or (115 g COD/m^3) (40,000 m^3/day):

$$\Delta N_{SS} = (115)(40{,}000)\left[\frac{1-(0.50)(1.20)}{2.86}\right] = 643{,}000 \text{ g/day}.$$

Next, the mass of MLSS in the anoxic zone must be estimated to allow the specific denitrification rate in the zone to be approximated. The total mass of MLSS in the system at 15°C can be estimated with Equation 10.8 for an SRT of 16 days, giving a value of 64,750,000 g. The mass of MLSS in the anoxic zone is in proportion to the fraction of the SRT in the anoxic zone, as given by Equation 12.2:

$$(X_{M,T} \cdot V)_{ANX} = 64{,}750{,}000\left(\frac{4.0}{16.0}\right) = 16{,}190{,}000 \text{ g}.$$

This can be used to calculate the process loading factor for addition of slowly biodegradable substrate to the anoxic zone. Since the concentration of slowly biodegradable substrate in the influent is 150 mg COD/L (g/m^3) and the flow rate is 40,000 m^3/day, the process loading factor is

$$U_{ANX} = \frac{(150)(40{,}000)}{16{,}190{,}000} = 0.37 \text{ g COD/(g MLSS} \cdot \text{day)}.$$

Substitution of the anoxic process loading factor into Equation 12.11 allows calculation of the specific denitrification rate in the anoxic zone for 20°C:

$$q_{NO/XS,20} = (0.018)(0.37) + 0.029 = 0.036 \text{ g } NO_3 - N/(\text{g MLSS} \cdot \text{day}).$$

Since the design is being performed for 15°C, the specific denitrification rate must be corrected to that temperature using Equation 3.99:

$$q_{NO/XS,15} = (0.036)(1.05)^{15-20} = 0.028 \text{ g } NO_3\text{-N/(g MLSS·day)}.$$

The mass rate of denitrification associated with the utilization of slowly biodegradable substrate and decay, ΔN_{XS}, is obtained by multiplying $q_{NO/XS}$ by the mass of MLSS in the anoxic zone:

$$\Delta N_{XS} = (0.028)(16{,}190{,}000) = 453{,}000 \text{ g } NO_3\text{-N/day}.$$

The total mass rate of denitrification is the sum of ΔN_{SS} and ΔN_{XS}, or

$$\Delta N = 643{,}000 + 453{,}000 = 1{,}096{,}000 \text{ g } NO_3\text{-N/day}.$$

Since the mass rate of nitrate-N production is 1,332,000 g NO_3-N/day, the mass rate of nitrate release from the system is 1,322,000 − 1,096,000 = 226,000 g NO_3-N/day. Therefore, the effluent nitrate-N concentration is

$$S_{NO} = \frac{226{,}000}{40{,}000} = 5.7 \text{ mg N/L}.$$

This value is acceptable and can be used for preliminary design. If it had not been adequate, it would have been necessary to adopt a new anoxic SRT and repeat the process until an acceptable value was found.

b. What are the sizes of the anoxic and aerobic zones?
The total mass of MLSS in the system at 15°C is 64,750,000 g. Since the design MLSS concentration is 3000 mg/L (g/m^3) throughout the system, the total system volume is

$$V = \frac{64{,}750{,}000}{3000} = 21{,}600 \text{ m}^3.$$

The mass of MLSS in the anoxic zone is 16,190,000 g, making its volume:

$$V_{ANX} = \frac{16{,}190{,}000}{3000} = 5400 \text{ m}^3.$$

Therefore, the volume of the aerobic zone is

$$V_{AER} = 21{,}600 - 5400 = 16{,}200 \text{ m}^3.$$

c. What MLR flow is needed at 15°C?
The specific denitrification rate is lowest in the winter, and thus the MLR flow for winter operation will be lower than that required for summer operation, which governs design of the mixed liquor recirculation system. Nevertheless, because the degree of denitrification has only been calculated for the winter temperature, we will use it to demonstrate the procedure. The sum of the recycle and recirculation ratios can be calculated with Equation 12.9. Since 1,322,000 g NO_3-N is produced per day and 1,096,000 g NO_3-N is denitrified per day, the fraction of the nitrate-N denitrified is 0.83. Therefore:

$$\alpha + \beta = \frac{0.83}{1 - 0.83} = 4.88.$$

This is a relatively high value, but necessary to meet the process goals. For economic final settler performance, the solids recycle ratio, α, typically lies between 0.5 and 1.0. This means that the MLR ratio, β, should be approximately 4.38. If a value of 4.5 is chosen, the MLR flow rate must be (4.5) (40,000 m^3/day) or 180,000 m^3/day. A final decision can be made only after the requirement for summer operation is determined. This will require recalculation of the mass of MLSS in the system in the summer, thereby giving the MLSS concentration associated with the system volume. This fixes the process loading factor for the anoxic zone, which determines the specific denitrification rate. The nitrogen available to the nitrifiers must also be recalculated for 25°C, allowing computation of the mass production rate of nitrate-N. Finally, the information can be combined to determine the fraction of the nitrate-N that is denitrified, allowing $\alpha + \beta$ to be calculated. As with the anoxic selector design, care must be taken not to return excessive amounts of nitrate to the anoxic zone because of the potential for nitrous oxide production when the electron acceptor is present in excess. Because of the difficulty of predicting whether nitrous oxide production will occur, flexibility should be designed into the MLR system to allow an adjustment of MLR flow rates during operation to achieve the desired effluent goals without release of greenhouse gases. Determination of the MLR rate needed for summer operation is left as an exercise for the reader.

d. What is the net alkalinity consumption at 15°C?
As calculated in part a, the concentration of nitrate-N leaving the system in the absence of nitrification would be 33.3 mg/L as N. Its formation would lead to the destruction of (7.23) (33.3) = 241 mg/L as $CaCO_3$ of alkalinity. Since the effluent nitrate-N concentration is 5.7 mg/L as N, the amount of nitrate-N denitrified is equivalent to 27.6 mg/L as N. This would return (3.5)(27.6) = 96.6 mg/L as $CaCO_3$ of alkalinity to the system. Thus, the net alkalinity destruction will be 241 − 96.6 = 144.4 mg/L as $CaCO_3$. Since 150 mg/L of the alkalinity in the influent could be destroyed while maintaining the needed residual of 50 mg/L, it would not be necessary to add chemicals in the winter. This is in contrast to the system with the anoxic selector, for which a chemical addition was still required, although in a lesser amount than the system without denitrification.

The remainder of the process design would proceed in exactly the same way as for the other systems. The oxygen requirement would have to be calculated for summer conditions and the savings due to denitrification determined in the same way as in Example 12.3.1.2. The heterotrophic oxygen requirement in the aerobic zone will be relatively low because the majority of the degradable

organic matter will have been removed in the anoxic zone. This fact, coupled with the fact that the tank is large to accommodate the biomass needed to achieve a long SRT, means that the power requirements are associated primarily with solids suspension rather than with oxygen transfer. Investigation of these issues will be left as an exercise for the reader.

Examples 12.3.1.2 and 12.3.1.3 illustrate that initial anoxic zones of two quite different sizes may be used in systems of the same basic configuration, depending on process objectives. If an anoxic selector is used to remove readily biodegradable organic matter, then a relatively small anoxic zone results due to the rapid utilization of that substrate. Meaningful removal of nitrate-N will occur as a result of the presence of an anoxic selector, but the effluent nitrate-N concentration may still be significant. Relatively modest MLR rates may be adequate for these applications because the fraction of the nitrate-N that must be recirculated to the selector is fairly modest. Furthermore, a noticeable reduction in the oxygen requirement in the aerobic zone and in the net alkalinity consumption will occur. The reduction in power requirements associated with the decrease in the oxygen requirement will partially or completely offset the power required to mix the selector and to recirculate mixed liquor. A larger initial anoxic zone may be required if the objective is to meet a specified effluent nitrate-N concentration through use of the MLE process. This occurs because both readily and slowly biodegradable organic matter may be required as electron donors to remove the larger mass of nitrate-N, and utilization of slowly biodegradable substrate results in slower denitrification rates. The greater degree of nitrate-N utilization in such applications results in increased MLR flow rates since a higher fraction of the nitrate-N formed must be recirculated to the initial anoxic zone. Greater reductions in the oxygen requirement and the associated power requirement in the aerobic zone occur, and net alkalinity consumption is reduced more. However, the further reduction in the power requirement for oxygen transfer may be small in comparison to the increase in power required to mix the larger anoxic zone and to pump the increased MLR flow.

12.3.1.4 Four-Stage Bardenpho Process—Addition of Second Anoxic and Aerobic Zones

A second anoxic zone, such as in the four-stage Bardenpho process (Figure 1.14), may be used to further reduce the nitrate-N concentration below that which can be economically achieved using an MLE system. However, because essentially all of the readily and slowly biodegradable organic matter will have been removed in the initial anoxic and aerobic zones, the primary source of electrons for the additional denitrification is biomass decay. Since the decay of heterotrophs is a relatively slow process, particularly under anoxic conditions, and since not all of the heterotrophic bacteria are capable of denitrification, the resulting specific rate of denitrification will be low. This means that the size of a second anoxic zone can be significant. In addition, a small aerobic zone is required to prepare the MLSS for settling. Typically, a hydraulic residence time (HRT) of 30 minutes is used, which is sufficient to strip entrained gases from the denitrified mixed liquor exiting the second anoxic zone and to add dissolved oxygen to it before it enters the clarifier. The MLSS concentration is uniform throughout the four-stage Bardenpho system, making the SRT in each zone directly proportional to its volume.

As with the design of a MLE system, the best way to estimate the impact of adding second anoxic and aerobic zones is through the use of simulation with a model like the IWA ASM No. 1. Care should be exercised in the selection of the anoxic hydrolysis factor, η_h, because it will have a large impact on the rate of nitrate-N utilization associated with decay in the anoxic zones.

In the absence of a kinetic characterization sufficient to allow simulation, or as a prelude to simulation, it is possible to roughly estimate the nitrate-N utilization in the second anoxic zone by partitioning the electron acceptor requirement due to decay in much the same way that the oxygen requirement due to decay was partitioned to the various reactors in a CAS system, as illustrated in Section 11.3.4. The major uncertainty associated with this is the effect that the anoxic conditions have on the rate of decay.

An alternative to estimating the denitrification rate due to decay is to take an empirical approach. As they did for the specific denitrification rate in the first anoxic zone, Burdick et al.[12] have reported an empirical relationship for the specific rate of denitrification due to decay in a second anoxic zone, $q_{NO/XB}$. In this case, since no substrate enters the second anoxic zone, they were able to correlate the specific denitrification rate with the system SRT:

$$q_{NO/XB} = 0.12\Theta_c^{-0.706}. \tag{12.12}$$

This relationship indicates that the specific denitrification rate will decrease as the SRT is increased, as expected. It was developed at a temperature of 20°C, but it can be corrected to other temperatures using Equation 3.99 with an appropriate θ value; a value around 1.02 appears reasonable. This relationship has been widely reported and used,[55,61] and it is appropriate when other data are not available.

As with Equation 12.11, use of Equation 12.12 requires an iterative procedure. As with all previous nitrogen removal systems, selection of the system size should be done for winter operating conditions. After the first anoxic and aerobic zones have been sized using the procedures for the MLE system, an SRT is assumed for the second anoxic zone, thereby increasing the system SRT. Since the MLSS concentration is uniform throughout the system, its volume is calculated in proportion to the volume of the first anoxic and aerobic zones as suggested by Equations 12.1 and 12.2. Typically, the second aerobic zone is sized to give an HRT of around 30 minutes, as mentioned previously, thereby increasing the system volume. The SRT of this zone is also calculated by proportion, giving the new system SRT. The new system SRT allows computation of $q_{NO/XB}$ with Equation 12.12 and the mass of MLSS in the system with Equation 10.8. The mass of MLSS in the second anoxic zone can then be calculated by proportioning with respect to its SRT as a fraction of the total SRT. Multiplication of that mass by $q_{NO/XB}$ allows calculation of the mass denitrification rate due to biomass decay in the second anoxic zone, ΔN_{XB}. Division of that value by the flow rate gives the additional amount by which the effluent nitrate-N concentration is reduced. Subtraction of that value from the MLE effluent nitrate-N concentration will determine whether the required effluent nitrate-N concentration is attained. If it is not, then another SRT must be assumed for the second anoxic zone and the process repeated. Generally, the volume of the second aerobic zone is held constant at a value giving an HRT of about 30 minutes. The system size determined by this procedure will be approximate because each time the system SRT is increased the amount of nitrogen available to the nitrifiers and the rate of denitrification in the first anoxic zone will change. However, recalculation around the entire system is not justified because of the approximate nature of the empirical relationships used for denitrification in both anoxic zones. As stated earlier, the most accurate method of arriving at a final design is through simulation, provided the needed kinetic information is available.

The presence of a second anoxic zone will increase the amount of alkalinity recovery in the system, which will increase the alkalinity of the final effluent. However, the alkalinity produced in the second anoxic zone will have little impact on the alkalinity in the main aerobic zone because little of it will be recirculated through the system. Thus, it will not change the amount of chemical required to achieve stable nitrification from that required by an MLE system.

The inclusion of a second anoxic zone in a nitrogen removal process will increase the mixed liquor recirculation requirements, which should be calculated for summer conditions. This occurs because the second anoxic zone will reduce the nitrate-N concentration entering the secondary clarifier, which reduces the nitrate-N concentration in the RAS flow to the initial anoxic zone, thereby decreasing the mass rate of nitrate-N return. If it is assumed that the nitrate-N concentration in the RAS is zero, Equation 12.8 can be modified to show the effect of the RAS and MLR ratios on the fraction of nitrate-N formed in the aerobic zone that is recirculated to the initial anoxic zone:

$$f_{NO,R} = \frac{\beta}{1+\alpha+\beta}. \tag{12.13}$$

Likewise, Equation 12.9 can be modified to allow calculation of the MLR ratio required to allow denitrification of a specified fraction of the nitrate-N in the initial anoxic zone:

$$\beta = \frac{f_{NO,D}(1+\alpha)}{1-f_{NO,D}}. \tag{12.14}$$

It is important to note that the value of $f_{NO,D}$ used in Equation 12.14 is exactly the same as the value used in Equation 12.9 because the fractional nitrate removal in the first anoxic zone is the same. As noted previously, the MLR flow rate should be calculated for summer conditions because it will be largest then.

The presence of the second anoxic zone will have little effect on the oxygen requirement in the first aerobic zone because the additional decay resulting from the increase in system SRT provides the electrons for denitrification in the second anoxic zone. The oxygen requirement in the second aerobic zone will be low, and thus the aeration rate in it will be governed primarily by the need to keep solids in suspension. The impact of the additional SRT on the solids wastage rate can be calculated for winter conditions using the same procedures as used in all other designs. Little is known about the potential production of nitric and nitrous oxide when denitrification is occurring via biomass decay.

The design of a second anoxic zone is illustrated in the following example.

Example 12.3.1.4

A second anoxic zone is to be added to the MLE process sized in Example 12.3.1.3. It is to reduce the nitrate-N concentration from 6 mg/L as N to 2 mg/L as N at 15°C. The second aerobic zone will have an HRT of 30 minutes. Assume that the temperature coefficient for the specific denitrification rate in the second anoxic zone is 1.02.

a. What is the required size of the second anoxic zone if the sizes and SRTs of the first anoxic and aerobic zones remain the same as in the MLE system?
As a first guess, assume an SRT of four days for the second anoxic zone. From Example 12.3.1.3, the SRT of the MLE system is 16 days and its volume is 21,600 m^3. Thus, the volume of the second anoxic zone, $V_{ANX,2}$, is given by proportion:

$$V_{ANX,2} = \left(\frac{4}{16}\right) 21{,}600 = 5400 \text{ m}^3.$$

The second aerobic zone has an HRT of 30 minutes. Since the influent flow rate is 40,000 m^3/day, its volume is 833 m^3. The total system volume is

$$V = 21{,}600 + 5400 + 833 = 27{,}833 \text{ m}^3.$$

By proportion, the total system SRT is

$$\Theta_c = \left(\frac{27{,}833}{21{,}600}\right)16 = 20.6 \text{ days}.$$

Therefore, the SRT of the second aerobic zone is 0.6 day.

Substitution of the system SRT into Equation 12.12 allows calculation of the specific denitrification rate in the second anoxic zone at 20°C:

$$q_{NO/XB,20} = (0.12)(20.6)^{-0.706} = 0.0142 \text{ g } NO_3\text{-N/(g MLSS·day)}.$$

Since the design is being performed for 15°C, the specific denitrification rate must be corrected to that temperature using Equation 3.99:

$$q_{NO/XB,15} = (0.0142)(1.02)^{15-20} = 0.013 \text{ g } NO_3\text{-N/(g MLSS·day)}.$$

This must now be multiplied by the mass of MLSS in the second anoxic zone to determine the mass removal rate of nitrate-N in the zone. The mass of MLSS in the system can be estimated with Equation 10.8, as has been done several times, giving a value of 79,000,000 g. Since the SRT in the second anoxic zone is 4 days and the system SRT is 20.6 days, the mass of MLSS in the second anoxic zone is calculated by proportion to be 15,300,000 g. Therefore, the mass rate of denitrification in the second anoxic zone is

$$\Delta N_{XB} = (0.013)(15{,}300{,}000) = 199{,}000 \text{ g } NO_3\text{-N/day}.$$

The mass rate of NO_3-N release from the MLE system was calculated in part a of Example 12.3.1.3 to be 226,000 g NO_3-N/day. Since the second anoxic zone can remove 199,000 g NO_3-N/day, the discharge into the second aerobic zone is 27,000 g NO_3-N/day. This corresponds to a concentration of 0.67 mg/L as N. Because the concentration is below the target value of 2.0 mg/L as N, it is possible that a smaller second anoxic zone could be used. However, several uncertainties exist in the design that suggest that it would be prudent to maintain an SRT of four days in the second anoxic zone. First, some additional nitrification will occur in the second aerobic zone on the ammonia-N released by the decay reactions in the second anoxic zone. Second, because of the longer SRT, the mass of MLSS in the first anoxic zone will be somewhat smaller than the amount calculated in Example 12.3.1.3. This means that a little less denitrification will occur there, increasing the mass flow rate of nitrate-N into the second anoxic zone. Consequently, it would be prudent to retain an SRT of four days in the second anoxic zone. Of course, if possible, simulations with a model like IWA ASM No. 1 should be done to refine the design.

b. What MLR flow rate to the first anoxic zone is required for this application at 15°C?
Equation 12.14 is used to perform this calculation. From part c of Example 12.3.1.3, 83% of the nitrate-N produced in the first aerobic zone must be recirculated to the first anoxic zone for denitrification. Typical solids recycle ratios for applications such as these range from 0.5 to 1.0. Calculate β for these two values of α. For α = 0.5:

$$\beta = \frac{0.83(1+0.5)}{1-0.83} = 7.32,$$

and for α = 1.0:

$$\beta = \frac{0.83(1+1)}{1-0.83} = 9.76.$$

These are high values, which may not be practical. The alternative would be to reduce the size of the initial anoxic zone to reduce the fraction of the nitrate-N that would be reduced in that zone. This will necessitate an increase in the size of the second anoxic zone since more nitrate-N must be removed there. The entire computational procedure of this and the preceding example would have to be repeated to arrive at an estimate of the performance of an alternative system. Doing this several times would provide the information required to choose the optimal system design. Consideration of the effort involved in doing this demonstrates clearly the benefits associated with being able to use simulation to investigate alternative designs.

12.3.1.5 Simultaneous Nitrification and Denitrification

As discussed in Section 12.1.2, simultaneous nitrification and denitrification can be a significant nitrogen removal mechanism in a nitrifying activated sludge system oxygenated with a point source aerator or in a system with a uniformly low DO concentration, even though the bioreactor does not

have a distinct and separate anoxic zone. Denitrification of as much as 50% of the nitrate-N produced has been reported in some applications.[51,62] The occurrence of simultaneous nitrification and denitrification requires three factors: (1) an oxygen transfer system that allows the development of zones of high and low DO concentration (either on a macroscopic or microscopic scale), (2) control of the oxygen input rate to the process, and (3) a sufficiently long SRT to allow full nitrification to occur even though parts of the bioreactor contain very low DO concentrations.

The impact of an oxygen transfer system that develops zones of high and low DO concentration was discussed in detail in Section 12.1.2. However, even if such a system is being used, regions of low DO concentration will not develop if the potential oxygen transfer rate to the system greatly exceeds the oxygen requirement. Rather, they will only develop when the potential oxygen transfer rate is less than the oxygen requirement. Furthermore, the mass rate of denitrification will be determined by the difference between the oxygen requirement for a totally aerobic system and the mass rate at which oxygen is actually being transferred to the liquid as an electron acceptor, TO:

$$\Delta N = \frac{RO_H + RO_A - TO}{2.86}. \tag{12.15}$$

When denitrification is occurring, the effluent nitrate-N concentration is determined by the difference between the mass rate of nitrate-N formation and its utilization. The mass formation rate is just the flow rate times the concentration of nitrate-N that would be in the effluent from a totally aerobic system as given by Equation 12.5. Therefore, the effluent concentration from a system experiencing simultaneous nitrification and denitrification is

$$S_{NO} = \frac{0.98F(S_{N,a} - S_{NH} - S_{NS}) - \Delta N}{F}. \tag{12.16}$$

Since S_{NO} cannot be negative, the ΔN from Equation 12.15 cannot exceed the mass formation rate.

In order to have simultaneous nitrification and denitrification, the SRT must be sufficiently long for nitrification to occur even though the average DO concentration is low. Because the bioreactor contains regions (either microscopic or macroscopic) of high and low DO concentration and these regions are dynamic, it is difficult to estimate exactly the degree of nitrification. Thus, an approximation must be used. First, it is assumed that nitrification occurs at a rate consistent with the average DO concentration in the bioreactor. This allows computation of a DO safety factor for the system (Equation 11.9) that can be used in Equation 11.10 or 11.11 (depending on the bioreactor configuration) to estimate the system SRT required for stable nitrification. As long as that SRT is less than the equivalent aerobic SRT, $\Theta_{c,AER,eq}$, of the system, stable nitrification will occur, thereby allowing simultaneous nitrification and denitrification. The term equivalent aerobic SRT is used to signify that distinct aerobic and anoxic zones do not exist, preventing a purely aerobic SRT from being calculated as it was for the other nitrogen removal systems. Thus, some other approach must be used to approximate the equivalent aerobic SRT.

Estimation of the equivalent aerobic SRT requires the assumption that anoxic and aerobic regions (either microscopic or macroscopic) are distributed equally throughout the entire bioreactor. Under that condition, if nitrate-N were used as effectively as oxygen as an electron acceptor, the fraction of the system that was anoxic would be equivalent to the fraction of the heterotrophic oxygen requirement that was being met by nitrate-N. However, only a fraction of the heterotrophs are able to denitrify. If we take that fraction as being represented by the anoxic growth factor, η_g, then the fraction of the system volume that is anoxic, $f_{V,ANX}$, would be given by

$$f_{V,ANX} = \frac{2.86 \cdot \Delta N}{\eta_g \cdot RO_H}. \tag{12.17}$$

Therefore, the equivalent aerobic SRT is

$$\Theta_{c,AER,eq} = (1 - f_{V,ANX})\Theta_c. \tag{12.18}$$

As stated above, as long as the required SRT as given by Equation 11.10 or 11.11 (depending on the bioreactor configuration) is less than the equivalent aerobic SRT of the system, as given by Equation 12.18, then stable nitrification will occur and the process will work. However, it should be recognized that these computations are approximate and that it is difficult to predict with certainty the degree of nitrification that will occur.

Savings in the amount of oxygen that must be supplied and in the amount of alkalinity that will be destroyed can be calculated using the same procedures as those used for the other denitrification systems.

One other important point about simultaneous nitrification and denitrification in a single vessel should be emphasized. That is the potential for filamentous bulking from growth of Group IV bacteria (Table 11.2).[30,35] Currently, techniques are not available for preventing such growth without chemical control as discussed in Section 11.4.3. Thus, the potential impacts of poor settleability on overall process performance should be considered before implementing this strategy for nitrogen removal.

Example 12.3.1.5

Consider the nitrifying activated sludge system designed in Example 12.3.1.1. The design SRT, which was chosen for winter operation, was 12 days and it is maintained year round. This SRT provides a substantial safety factor during warm temperature operation. Consequently, the feasibility of operating the oxygen transfer system to achieve 50% denitrification during warm weather conditions is to be assessed. In making this assessment, assume that the reduction in oxygen input will result in an average DO concentration of 0.5 mg/L, that the value of the anoxic growth factor, η_g, is 0.8, and that the safety factor for uncertainty is 1.0.

a. What SRT will be required to achieve stable nitrification in summer?
   The approach is the same as in part a of Example 12.3.1.1. Since a CAS system is being used, Equation 11.11 is the appropriate expression with which to calculate the required SRT. From Figure 10.4, the minimum aerobic SRT for the growth of nitrifiers is 1.3 days. The peak load safety factor, ζ_{PL}, is 2.5. The safety factor for uncertainty, ζ_U, has a value of 1.0. Because the half-saturation coefficient for DO for nitrifiers is 0.75 mg/L (Table E11.2), and the average DO concentration is 0.5 mg/L, the DO safety factor (Equation 11.9) has a value of 2.5. Thus:

$$\Theta_{c,r} = (1.3)(1.0)(2.5)(2.5) = 8.12 \text{ days}.$$

   Thus an SRT of at least 8.12 days is needed to ensure that stable nitrification will occur even during peak loads. The absolute minimum SRT for nitrification, however, is obtained by multiplying the DO safety factor by the minimum SRT, giving a value of 3.25 days. At that SRT, nitrification will be less stable during transients, with significant breakthrough of ammonia-N occurring. It will become more stable, however, as the SRT is increased toward 8.12 days.

b. What is the equivalent aerobic SRT at 25°C?
   From Example 12.3.1.1, the effluent nitrate-N concentration in the summer is 33.5 mg/L as N. Since the flow rate is 40,000 m^3/day, the nitrate production rate is 1,340,000 g NO_3-N/day. For 50% denitrification, ΔN would be half of that or 670 kg NO_3-N/day. The heterotrophic oxygen requirement at 25°C was calculated in part e of Example 12.3.1.1, where

it was found to be 7930 kg O_2/day. Substituting these into Equation 12.17 gives the fraction of the system volume that is anoxic:

$$f_{V,ANX} = \frac{(2.86)(670)}{(0.8)(7{,}930)} = 0.30.$$

The equivalent aerobic SRT can then be calculated with Equation 12.18:

$$\Theta_{c,AER,eq} = (1 - 0.30)12.0 = 8.4 \text{ days.}$$

The equivalent aerobic SRT of 8.4 days exceeds the required SRT of 8.12 days. Thus, stable nitrification will occur at the reduced DO concentration, allowing significant denitrification to occur. Thus, the proposed operating mode is feasible.

c. What oxygen transfer rate, TO, is required to achieve 50% denitrification at 25°C?
The summer oxygen requirement for a completely aerobic system was calculated in part e of Example 12.3.1.1. The heterotrophic requirement, RO_H, was found to be 7930 kg O_2/day, as noted above. The autotrophic requirement, RO_A, was found to be 6080 kg O_2/day. The denitrification rate, ΔN, was found to be 670 kg NO_3-N/day in part b above. The required oxygen transfer rate can be calculated with a rearranged form of Equation 12.15:

$$TO = 7930 + 6080 - (2.86)(670) = 12{,}090 \text{ kg } O_2\text{/day.}$$

This rate is 86.3% of the rate in the totally aerobic system.

12.3.1.6 Separate Stage Denitrification

Separate stage denitrification systems, such as illustrated in Figure 1.12, can be designed by the same procedures as those presented in Section 11.3 for the design of aerobic processes. In this case, however, the terminal electron acceptor (nitrate-N) is present in the influent wastewater and it is the electron donor that must be added. Since the required mass rate of nitrate-N removal, ΔN, is known, the mass rate at which the electron donor must be supplied, ΔS, can be calculated with Equation 6.4 by setting the SRT equal to the anoxic SRT. Division of this rate by the influent flow rate gives the minimum electron donor concentration that, expressed as COD, must be provided. In practice, a slightly higher quantity should be supplied to ensure that electron donor is not limiting in the anoxic zone, thereby minimizing the potential for the formation of nitric and nitrous oxide. As discussed in Section 12.1.3, methanol is often used as the electron donor. An appropriate value for the anoxic SRT is selected, generally on the order of three to five days. The mass of MLSS in the system is then calculated for the selected anoxic SRT using Equation 10.8 and the MLSS concentration can be chosen to give an economical settler, thereby fixing the volume of the anoxic zone. Since oxygen transfer is not required, the final settler size is the only factor that needs to be considered in choosing the MLSS concentration, unless the influent nitrate-N concentration is so high that the evolution of N_2 gas would cause floc shear. For domestic wastewater treatment, mixing energy with a submerged mixer will generally be required to keep the MLSS in suspension. An aerobic zone with an HRT of around 45 to 60 minutes is added to the process to remove the residual electron donor, strip the nitrogen gas, and oxygenate the mixed liquor prior to entering the clarifier. The total SRT for the system can be estimated from the total system volume by proportioning, as illustrated in the examples of this chapter. The mass of MLSS in the system can then be recalculated and the MLSS concentration reconsidered if warranted. The quantity of waste solids is calculated as illustrated in Section 11.3.3. As indicated in Section 12.1.3, separate stage suspended growth denitrification systems are not used widely in practice. The interested reader is referred to the U. S. EPA *Process Design Manual for Nitrogen Control*[61] for a detailed description of the design of such systems.

12.3.2 Biological Phosphorus Removal Processes

This section deals with those BNR processes that are designed to remove only organic matter and phosphorus. No nitrogen removal occurs. In fact, nitrification is undesirable because of the negative impact of nitrate-N recycle on the processes occurring in the anaerobic zone, as discussed in Sections 7.7.2 and 7.7.3. Section 12.3.3 will consider those BNR systems in which both nitrogen and phosphorus must be removed, or in which nitrification must occur in addition to phosphorus removal.

The mechanisms of biological phosphorus removal are discussed in Sections 2.4.6 and 3.7, while simulations showing the interactions between PAOs and the other bacteria commonly found in activated sludge systems are presented in Sections 7.7.2 and 7.7.3. It is apparent from all of this information that BPR is the most complicated suspended growth biochemical operation in use today, with many complex interactions possible between the different populations in the microbial community. Consideration of those interactions makes it clear that simulation is the best way to evaluate potential designs, short of actually testing them in pilot-scale facilities. While procedures for the design of BPR processes are not as well developed as those for biological nitrogen removal systems and are more empirically based, the fundamental design approaches for activated sludge systems presented in Section 11.3 can be coupled with appropriate values of the SRT to establish the preliminary size and configuration of a BPR process. That approach will be presented here.

Processes such as the A/O process are sized using the principles presented in Section 11.3. The SRTs in the anaerobic and aerobic zones are selected as described in Section 12.2.1 and added to give the total SRT. The mass of MLSS, the solids wastage rate, and the oxygen requirement are calculated for the selected total SRT and the bioreactor is sized using the procedures illustrated in Section 11.3.3.

Selection of the anaerobic SRT is influenced strongly by the amount and nature of the readily biodegradable substrate in the wastewater. Phosphate accumulating organisms sequester VFAs in the anaerobic zone, thereby providing themselves with the energy source that they will use to grow and store phosphorus in the aerobic zone. Approximately 7 to 10 mg of acetic acid are required to remove 1.0 mg of phosphorus.[71,74,75] This corresponds to 7.5 to 10.7 mg of COD due to VFAs for each mg of phosphorus that must be removed. Storage of VFAs as PHAs in the anaerobic zone is a very rapid process.[73] Consequently, if the needed amount of VFAs is present in the wastewater, the anaerobic SRT can be as short as 0.5 day at 20°C.[40] If no VFAs are present in the influent, but the amount of readily biodegradable substrate is sufficient to produce the needed quantity by fermentation, then the anaerobic SRT needs to be around 1.5 days at 20°C.[40] For situations in which some VFAs are present, but partial fermentation is still required, SRTs between 0.5 and 1.5 days should be sufficient, depending on the quantity of VFAs available. On the other hand, should the amount of readily biodegradable substrate be insufficient, hydrolysis of slowly biodegradable substrate must occur before fermentation can form the VFAs and the anaerobic SRT must be much longer, on the order of 2.5 to 3 days. In that case, it may be more economic to carry out fermentation reactions off-stream, thereby allowing a VFA-rich stream to be added to the influent.

If nitrification is occurring in the aerobic zone, nitrate-N will be returned to the anaerobic zone through the RAS flow (making it anoxic) and the denitrifying OHOs will be able to outcompete the PAOs for VFAs. (While denitrifying PAOs can compete with the denitrifying OHOs for VFAs, they will metabolize those VFAs, rather than storing them as PHAs. Thus, in effect, they act like denitrifying OHOs. Consequently, their role is not considered explicitly here, giving a conservative design.) The return of nitrate-N to the anaerobic zone will also interfere with fermentation since the OHOs will simply use as much readily biodegradable substrate as is possible for the amount of nitrate-N returned, rather than fermenting it to VFAs. Equation 6.4 allows calculation of the amount of readily biodegradable COD (or VFAs) removed per unit of nitrate-N returned to the anaerobic zone. Using typical parameter values, it suggests that about 6 mg of COD will be lost per mg of nitrate-N recycled to the anaerobic zone. Consequently, if nitrification is occurring in the aerobic

zone, this additional substrate requirement must be considered when making a judgment about the size of the anaerobic SRT. To do this, the fraction of nitrate-N formed that is returned to the anaerobic zone via the RAS flow must be known. It can be calculated with Equation 12.8 in which β has been set equal to zero since no mixed liquor recirculation is used.

Because of these negative impacts of nitrification, processes like the A/O process should be designed with short aerobic SRTs to minimize the likelihood of nitrification. This practice will also maximize the excess biomass production, thereby maximizing phosphorus removal. Selection of the aerobic SRT is complicated by the fact that the minimum SRTs required for growth of PAOs and nitrifiers are very similar at high temperatures, as illustrated in Figure 10.4. Therefore, to avoid nitrification at high temperature while providing sufficient aerobic SRT for growth of PAOs at low temperature, it may be necessary to use different aerobic SRTs in winter and summer. Generally, a safety factor of 1.5 to 2.0 is applied to the minimum SRT for PAOs to arrive at an aerobic SRT that will allow PAOs to grow while minimizing the chances for stable nitrification.

Since no substrate oxidation occurs in the anaerobic zone (VFAs are simply stored as PHAs), the oxygen requirement can be calculated with Equation 10.10 using the total SRT. However, some evidence suggests that the oxygen requirement in a BPR process may be less than that calculated value.[3,64] The mechanism for such a reduction is not known, but the observed magnitude of the effect appears to be on the order of 10 to 30%. Until this issue is resolved, it is prudent to design the oxygen transfer system based on the results of Equation 10.10, but to incorporate additional turndown capacity into the design to allow plant operators to save energy if actual oxygen requirements are less than calculated.

Since the MLSS concentration is uniform in this type of BPR process, the sizes of the aerobic and anaerobic zones will be proportional to their SRTs, as indicated by Equations 12.1 and 12.3. Likewise, the mass of MLSS in each zone will be proportional to its SRT. Consequently, once the mass of MLSS in the system has been calculated by using the total SRT in Equation 10.8, the mass in the aerobic zone can be calculated by proportion. It can then be used in the selection of the aerobic zone volume, and the associated MLSS concentration, by considering floc shear, mixing, and oxygen transfer rates, just as was done in Section 11.3.3. The size of the anaerobic zone is then determined by proportionality. Benefits accrue from staging the anaerobic zone, with two tanks in series being a recommended configuration.[23,55] The mixing requirements in the anaerobic zone are similar to those in the anoxic zone of a nitrogen removal system, which are discussed in Section 12.3.1.

Because of the complexity of the events occurring in biological phosphorus removal, the only way to predict the effluent soluble phosphorus concentration with any certainty is to fully characterize the wastewater and use the resulting parameters in simulations. However, when that cannot be done, the organic matter to phosphorus removal ratio for the particular type of process (Table 12.4) can be used to approximate the value, or alternatively, the stoichiometry based on VFAs and readily biodegradable substrate reported earlier can be used. As long as more than enough organic matter is present in the wastewater to remove all of the phosphorus (i.e., the wastewater is phosphorus limited), then an effluent low in soluble phosphorus will be produced. On the other hand, if the process is carbon limited, the amount of phosphorus that can be removed will be determined primarily by the available organic matter, with the effluent soluble phosphorus concentration being determined by the difference between the amount entering the system and the amount removed. The total phosphorus concentration in the effluent will be the soluble phosphorus plus the phosphorus in the effluent suspended solids, which will depend on the effluent suspended solids concentration, as illustrated in Figure 12.8. The phosphorus content of the suspended solids can be determined on the basis of the phosphorus removed and the excess solids production. It is equal to the difference between the mass flow rates of phosphorus in and out of the process divided by the solids wastage rate.

The following example illustrates the design of a biological phosphorus removal process.

Example 12.3.2.1

Consider the design of a BPR process like the A/O process to remove phosphorus from the wastewater that was the subject of all of the examples throughout Sections 11.3 and 12.3.1. The phosphorus content of the wastewater is 7.5 mg/L as P but the VFA concentration is negligible. A safety factor of 1.5 is considered to be adequate for the selection of the aerobic SRT to ensure stable PAO growth.

a. What SRTs are appropriate for cold and warm temperature operating conditions?
   First consider cold temperature operation (15°C). Because the wastewater contains no VFAs, the anaerobic SRT will need to be at least 1.5 days to allow fermentation of the readily biodegradable organic matter. From Figure 10.4, the minimum aerobic SRT is 2 days at 15°C. Using a factor of safety of 1.5, the design aerobic SRT would be 3 days. Also from Figure 10.4, the minimum aerobic SRT for nitrification 3.5 days. Thus, for winter operation, an SRT of 3 days would allow the system to operate without nitrification. The total SRT would be 4.5 days.

   Next, consider warm temperature operation (25°C). At 25°C it should be possible to reduce the anaerobic SRT to 1 day while still performing the necessary fermentation. From Figure 10.4, the minimum aerobic SRT is 1.3 days. Using a safety factor of 1.5, this gives a design aerobic SRT of 2 days. Unfortunately, it can be seen from Figure 10.4 that the minimum aerobic SRT for nitrifiers at 25°C is also around 1.3 days, so some nitrification will occur during the warmest operating condition. The extent of nitrification will depend on a number of factors, including the bioreactor configuration, and can only be accurately predicted from pilot studies. However, complete nitrification would definitely occur if the SRT were left at the winter value. Thus, in summer, the total SRT should be 3 days.

   We now need to check to determine whether these two operating conditions are compatible with each other. Since the fraction of the SRT in each zone is equal to the fraction of the system volume in each zone, the fractions must be the same in summer and winter. Calculation of the anaerobic SRT as a fraction of total SRT for both conditions reveals that it is one-third both times. Thus, the choices are acceptable.

b. Selection of bioreactor volume and MLSS concentrations.
   The mass of MLSS in the system should be calculated for winter conditions using Equation 10.8 because winter conditions require the longer SRT and have the lower decay coefficient, both of which maximize the mass of MLSS. Once that is done, the bioreactor volume and the MLSS concentration must be selected in exactly the same way as in Example 11.3.3.2. Because this procedure has been demonstrated several times, it will not be repeated here. However, it should be recognized that the MLSS concentration will be lower in summer because of the shorter SRT and higher decay coefficient and this should be considered in the choices.

c. What are the solids wastage rate and the oxygen requirement for heterotrophs?
   Because of the use of different SRTs in summer and winter, we cannot assume that winter conditions will give the greater solids wastage rate, $W_{M,T}$, and summer the higher heterotrophic oxygen requirement, RO_H. Consequently, they must be calculated for each operating condition. As has been done several times, the solids wastage rate is calculated using Equation 10.9, while the heterotrophic oxygen requirement is calculated using Equation 10.10. The results are provided in Table E12.1. The differences are insignificant. It should be noted that such small differences will not always be the case.

d. What effluent soluble phosphorus concentration would be expected under winter operating conditions?
   Nitrification will not occur in the winter, so no nitrate-N will be recycled to the anaerobic zone in the RAS. We can expect to use 7.5 to 10.7 mg of VFA COD for each mg of phosphorus that must be removed. Little COD is lost during fermentation of readily biodegradable organic matter to VFAs, so we would expect to get a mg of VFA COD for each mg of readily biodegradable COD fermented. To be conservative, assume that 10.7 mg of readily biodegradable COD will be needed to remove a mg of phosphorus after fermentation. Since the wastewater contains 115 mg/L of readily biodegradable COD, the maximum

phosphorus removal capability is 115 ÷ 10.7 = 10.7 mg/L. Furthermore, since the wastewater contains 7.5 mg/L of phosphorus, the system would be expected to remove essentially all of it in winter. Thus, we would expect the effluent soluble phosphorus concentration to be low.

Another way of addressing this question is with the $BOD_5/\Delta P$ ratios given in Table 12.4. Since the A/O process is considered to be a high efficiency process when nitrification is not occurring, the $BOD_5/\Delta P$ ratio will range from 15 to 20 mg BOD_5/mg P. Again, to be conservative, assume that the worst case applies (i.e., a ratio of 20). Since the wastewater contains 155 mg/L of BOD_5 (Table E9.4), the maximum phosphorus removal capability is 155 ÷ 20 = 7.8 mg/L. This, too, suggests that excellent phosphorus removal should occur, giving a very low effluent soluble phosphate concentration.

e. What effluent soluble phosphorus concentration would be expected under summer operating conditions?

During the warmest part of the summer, it is likely that nitrification will occur in spite of the reduction in the SRT to three days. In Example 11.3.3.2, the concentration of nitrogen available to the nitrifiers at an SRT of three days was calculated to be 30.5 mg/L as N. Thus, we can expect the maximum amount of nitrate-N formed to be 29.9 mg/L as N (Equation 12.5). The amount of nitrate-N recycled to the anaerobic zone depends on the RAS flow rate. Typically, the RAS flow rate is one-half the influent flow rate, or α is 0.5. Using that in Equation 12.8 with $\beta = 0$ tells us that one-third of the nitrate-N formed will be returned, or an effective concentration of 10 mg/L as N. Since about 6 mg of COD will be required to denitrify each mg of nitrate-N, the recycle of the nitrate-N will reduce the readily biodegradable COD by 60 mg/L, to a value of 55 mg/L. If only 7.5 mg of readily biodegradable COD were required to remove a mg of phosphorus, the process would be capable of removing 55 ÷ 7.5 = 7.3 mg/L of P, which would be sufficient to remove almost all of the phosphorus. However, if 10.7 mg of COD were required, the system could only remove 55 ÷ 10.7 = 5.1 mg/L of P. Although a small amount of phosphorus would be used in biomass synthesis, this suggests that a significant concentration of residual phosphorus would be left in the summer. The exact amount could only be determined with pilot studies. This analysis did not consider the slowly biodegradable substrate because relatively little of it will be available at the short SRTs involved. Again, only pilot studies or simulations will reveal whether any slowly biodegradable substrate would become available.

As in part d, another way of addressing this question is with the $BOD_5/\Delta P$ ratio. From Table 12.4 we see that the ratio is 20 to 25 mg BOD_5/mg P for an A/O process in which nitrification is occurring. A ratio of 20 would lead us to believe that almost all of the phosphorus could be removed, as we saw in part d. If the ratio were 25, we would expect the removal to be 155 ÷ 25 = 6.2 mg/L as P, which would leave a significant residual. Thus this analysis, like the previous, gives mixed results, with one assumption suggesting full phosphorus removal while the other suggests incomplete removal. Therefore, pilot studies should be performed to more accurately define the capability.

f. What is the phosphorus content of the MLSS in the winter?

Since all of the phosphorus will be removed via the waste solids and the solids are homogeneous throughout the system, the phosphorus content of the MLSS can be determined by calculating the phosphorus content of the waste solids. The mass flow rate of phosphorus in the influent to the process is (7.5 g/m^3 as P) (40,000 m^3/day) ÷ 1000 g/kg = 300 kg/day of P. As discussed in part d above, this phosphorus will be almost completely removed. Therefore, to be conservative assume that the effluent soluble PO_4-P concentration is negligible. Therefore, the mass of phosphorus removed per day is 300 kg. From Table E12.1, the solids wastage rate in winter is 5330 kg/day. Therefore, the phosphorus content of the mixed liquor is 300 kg P/day ÷ 5330 kg TSS/day = 0.056 mg P/mg TSS.

g. What will be the concentration of particulate phosphorus in the effluent if the effluent suspended solids concentration is 10 mg/L as TSS?

Since the phosphorus content of the solids is 0.056 mg P/mg TSS and the effluent suspended solids concentration is 10 mg/L as TSS, the effluent particulate phosphorus concentration is 0.56 mg/L as P. The total phosphorus concentration will be the sum of this value plus any soluble phosphorus that is present.

TABLE E12.1
Effect of Temperature on the Excess Solids Production Rate and the Oxygen Requirement in Example 12.3.2.1

	Warm Weather	Cold Weather
Temperature (°C)	25	15
SRT (days)	3.0	4.5
$W_{M,T}$ (kg TSS/day)	5350	5330
RO_H (kg O_2/day)	6260	6290

12.3.3 Processes That Remove Both Nitrogen and Phosphorus

As discussed in Section 12.1.3, several processes are available that remove both nitrogen and phosphorus. This would be of significant concern if unique design procedures were required for each. Fortunately this is not the case, and all processes that remove both nitrogen and phosphorus can be designed using the procedures already presented in Sections 12.3.1 and 12.3.2. Because those procedures are approximate, so is the procedure for designing systems to remove both nitrogen and phosphorus. The degree of interaction among the different components of the microbial community in a BNR system makes it impossible to develop exact analytical procedures. However, just as for systems that remove nitrogen or phosphorus, they are sufficiently exact to provide designs that can be verified through pilot studies and simulation. Just a few additional factors must be considered when designing a process that removes both nutrients.

Because complete nitrification must occur in the course of removing nitrogen, one consideration of particular importance is the minimization of nitrate-N recycle to the anaerobic zone. As discussed in Section 7.7, nitrate-N recycle adversely impacts phosphorus removal by allowing increased growth of OHOs and by reducing fermentation in the anaerobic zone. Elimination of nitrate-N recycle is primarily a process selection issue, and several processes have been developed that minimize or eliminate it. The issues involved in process selection are discussed in Section 12.1.2 and the benefits and drawbacks of the alternative processes are presented in Table 12.2.

Another consideration in adapting the approaches of Sections 12.3.1 and 12.3.2 to the design of a process to remove both nutrients is the impact of the upstream anaerobic zone on denitrification in the downstream anoxic zone. It might be thought that the removal of readily biodegradable substrate in the anaerobic zone would result in reduced rates of denitrification in the anoxic zone because of the need to use slowly biodegradable substrate. Experience indicates that this is not the case.[15] It is hypothesized that fermentation of slowly biodegradable organic matter in the anaerobic zone results in the formation of readily biodegradable substrate, which produces a rapid rate of denitrification by OHOs in the anoxic zone. The rapid denitrification rate could also be due to the presence of denitrifying PAOs, which would store VFAs in the anaerobic zone and then metabolize them for growth and phosphate storage in the anoxic zone. While additional research is needed to elucidate the mechanism, it is clear that similar anoxic zone sizes can be used in processes that remove nitrogen alone and in processes that remove both nitrogen and phosphorus.

A final consideration is that some systems have different MLSS concentrations in the anaerobic zone than in the anoxic and aerobic zones. Examples are the UCT and VIP processes, shown in Figures 1.16 and 12.9, respectively, which add the RAS flow to the anoxic zone and provide biomass to the anaerobic zone by recirculating denitrified anoxic mixed liquor (AR) from the anoxic zone to the anaerobic zone. Although this is done to minimize nitrate-N recirculation to the anaerobic zone, it will result in lower MLSS concentrations in the anaerobic zone. The MLSS concentration in the anaerobic zone, $X_{M,T,ANA}$, can be estimated from the MLSS concentration in the anoxic zone,

$X_{M,T,ANX}$, by performing a mass balance on the anaerobic zone. Neglecting any change in the MLSS concentration that occurs in the zone, it can be shown that:

$$X_{M,T,ANA} = X_{M,T,ANX}\left(\frac{\delta}{1+\delta}\right), \tag{12.19}$$

where δ is the AR rate expressed as a fraction of the influent flow rate. The JHB process (Figure 12.4) also has a nonuniform MLSS concentration, but in this case only the first anoxic zone is different, having approximately the same concentration as in the RAS.

As with the other BNR processes, design begins with a selection of the SRTs for the three environments. Anaerobic SRTs can be reduced in processes that remove both nitrogen and phosphorus in comparison to the anaerobic SRT needed in a process that removes phosphorus alone. This is because readily biodegradable substrate is removed in both the anaerobic and anoxic zones of a process that removes both nutrients. Consequently, the anaerobic zone need not be relied upon to remove all of this material. However, reduction of the anaerobic SRT will result in reduced fermentation and reduced phosphorus removal capability for the process. It is prudent to provide the flexibility to adjust the relative sizes of the anaerobic and anoxic zones, thereby adjusting their SRTs. Because it is the input of nitrate-N that distinguishes an anoxic zone from an anaerobic zone, one way to do this is to construct those zones as several completely mixed tanks in series, as in the VIP process (Figure 12.9), and to provide several possible discharge points for the RAS and nitrified mixed liquor recirculation, NR. In this way, a tank can be changed from anaerobic to anoxic, and vice versa, simply by moving the RAS and NR input point. In addition, both mixing and oxygen transfer equipment can be installed in the initial sections of the aerobic zone to provide the flexibility to extend the anoxic SRT should the need exist. Staging of the bioreactor provides several advantages including improved reaction rates and the selection of microorganisms with higher maximum specific growth rates. Thus, incorporation of staging not only provides operational flexibility, but also enhances overall process performance.

Just as with biological nitrogen removal processes, systems that remove both nitrogen and phosphorus can be designed for a widely varying degree of nitrogen removal. The principles presented in Section 12.3.1 can be directly applied here, depending on effluent total nitrogen objectives. The minimum degree of nitrogen removal required is that which will eliminate nitrate-N recycle to the anaerobic zone.

The design proceeds in exactly the same way that the other designs have proceeded. First, the SRTs of the three zones must be selected using the criteria presented above and in the preceding two sections, as well as in Section 12.2.1. This may require iteration, as discussed in Section 12.3.1. Using the total SRT, the mass of MLSS in the system is estimated with Equation 10.8, giving the mass of MLSS in each zone by application of Equations 12.1–12.3. The RAS and NR flow rates are then selected to give the desired degree of nitrogen removal through the anoxic zone, using the principles articulated in Section 12.3.1 for an initial anoxic zone. The

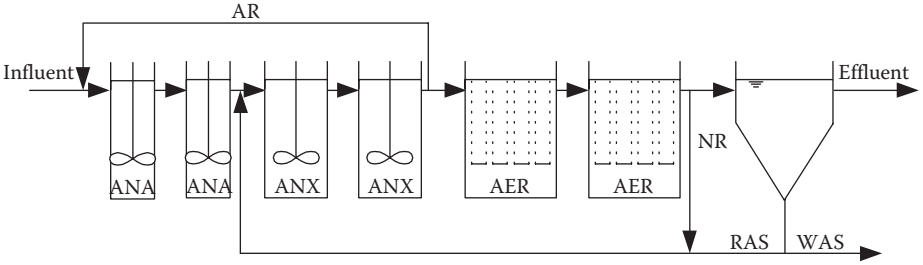

FIGURE 12.9 Virginia Initiative Plant (VIP) process.

oxygen requirement in the absence of denitrification can be calculated with Equations 10.10 and 11.16, and corrected for denitrification to determine the net oxygen requirement as illustrated in Example 12.3.1.2. An MLSS concentration is then selected for the aerobic zone following consideration of the requirements for sedimentation, mixing, and oxygen transfer. This allows the total volume of the aerobic zone to be calculated. Since the MLSS concentration is the same in the anoxic zone as in the aerobic zone, the chosen MLSS concentration can also be used to calculate the volume of the anoxic zone from the mass of MLSS present. A denitrified anoxic mixed liquor recirculation (AR) rate can then be selected, thereby setting the anaerobic MLSS concentration by Equation 12.19. That, in turn, is used to calculate the volume of the anaerobic zone from the known mass of MLSS present. Higher AR rates will result in higher anaerobic MLSS concentrations, thereby reducing the volume of the anaerobic zone. Thus, the opportunity exists for selecting a least-cost combination. Each zone can be subdivided into stages, if desired. If an effluent is desired with a lower nitrate-N concentration than can be accomplished with only an initial anoxic zone, then a second anoxic zone and a second aerobic zone can be added using the approach from Example 12.3.1.4.

Example 12.3.3.1

A facility to remove both nitrogen and phosphorus is to be designed to treat the wastewater considered throughout Sections 11.3 and 12.3, which contains 7.5 mg/L of phosphorus. Process objectives include oxidizing ammonia-N, maximizing phosphorus removal, and obtaining a moderate degree of nitrogen removal to reduce the net oxygen requirement and alkalinity consumption. The VIP process was selected based on these objectives.

a. What SRTs are appropriate for this application?
   As discussed in part a of Example 12.3.1.1, the required aerobic SRT is 12.0 days. Although an anaerobic SRT as low as 0.5 days could be used, a value of 1.0 is selected to allow some fermentation to occur. An anoxic SRT of 1.5 days is selected to ensure complete removal of readily biodegradable organic matter, which will result in good sludge settling characteristics. Thus, the total SRT will be 1.0 + 1.5 + 12.0 or 14.5 days.
b. If the design MLSS concentration in the anoxic and aerobic zones is 3000 mg/L as TSS and the AR rate is equal to the influent flow rate, what are the sizes of the bioreactor and its individual components, as well as the HRTs?
   First, calculate the mass of MLSS in the system at 15°C using Equation 10.8 and an SRT of 14.5 days. Use of the procedure demonstrated in Example 11.3.3.3 gives a value of 60,000,000 g. The application of Equations 12.1–12.3 gives the mass in each zone:

$$\Sigma(X_{M,T} \cdot V)_{AER} = 49{,}700{,}000 \text{ g},$$

$$\Sigma(X_{M,T} \cdot V)_{ANX} = 6{,}200{,}000 \text{ g},$$

   and

$$\Sigma(X_{M,T} \cdot V)_{ANA} = 4{,}100{,}000 \text{ g}.$$

   Since the MLSS concentration in the aerobic and anoxic zones is 3000 mg/L, the volumes of those zones are 16,600 and 2100 m^3, respectively.

   The MLSS concentration in the anaerobic zone is obtained by the application of Equation 12.19. Since the AR flow rate is equal to the influent flow rate, δ has a value of 1.0. Therefore:

$$X_{M,T,ANA} = 3000\left(\frac{1.0}{1+1.0}\right) = 1500 \text{ mg/L}.$$

TABLE E12.2
Volumes and Hydraulic Residence Times for the Zones in the VIP Process of Example 12.3.3.1

Zone	Volume m^3	HRT Day
Anaerobic	2700	0.067
Anoxic	2100	0.053
Aerobic	16,600	0.415
Total	21,400	0.535

Using this value gives a volume for the anaerobic basin of 2700 m^3.

The HRTs of each zone can be calculated using the design flow of 40,000 m^3/day. The volumes and resulting HRTs are summarized in Table E12.2. Note that the size and HRT of the anaerobic zone are larger than those of the anoxic zone, even though its SRT is smaller. This is a result of the lower MLSS concentration in the anaerobic zone.

Appropriate values for the anaerobic and anoxic zone SRTs are only approximately known, and thus staging of these zones is desirable, as discussed above. Consequently, use a total volume of 4800 m^3 for the anaerobic and anoxic zones and configure the volume as five stages, each with a volume of 960 m^3. Provide flexibility in the NR and RAS discharge locations so that either two or three of the cells can function as the anaerobic zone, with the remainder functioning as the anoxic zone. Furthermore, because of the benefits of staging the aerobic zone with respect to nitrification, divide it into four cells, each with a volume of 4150 m^3.

c. What total nitrogen removal can be expected, and what NR flow rate is appropriate for this application?
As shown in part b of Example 12.3.1.2, an anoxic selector with an SRT of 1.5 days is expected to remove 837 kg NO_3-N/day based on the availability of readily biodegradable substrate. As discussed above, experience indicates that we should get about the same amount of denitrification in this application in spite of the presence of the anaerobic zone preceding the anoxic zone. It was also shown in part b of Example 12.3.2.2 that the sum of the RAS and MLR rates should be 163% of the influent flow rate to provide sufficient nitrate-N to remove all of the readily biodegradable substrate. Thus, it is appropriate to provide an equivalent amount of RAS and NR flow in this case. However, because the goal is to fully denitrify in the anoxic zone to prevent recirculation of nitrate-N to the anaerobic zone, considerable flexibility should be provided in the possible pumping rates to ensure that this is accomplished.

d. How much phosphorus should this process be able to remove?
Because the PAOs are protected from nitrate by the process configuration, the system should be capable of removing equal quantities of phosphorus in winter and summer. Thus, the amount of phosphorus removal should be similar to that estimated for the A/O process in the winter in part d of Example 12.3.2.1. Thus, essentially all of the influent phosphorus should be removed.

12.3.4 Process Optimization by Dynamic Simulation

As discussed in Sections 10.4.3 and 11.3.7, process optimization using dynamic simulation is an extremely valuable tool for refining the design and operation of biological processes. This is particularly true for BNR systems because of the complex interactions among components of the system. The most widely used models for BNR processes are IWA ASMs No. 1 and 2d, as described in Section 6.1. Activated Sludge Model No. 1 was used in Chapter 6 to illustrate the

impacts of nitrification and denitrification on the operational and performance characteristics of suspended growth biochemical operations, and both models were used in Chapter 7 to illustrate the performance of several BNR systems. Both ASM No. 1 and 2d have been widely used to model and optimize suspended growth processes for nitrogen, phosphorus, and combined nitrogen and phosphorus removal.[2,19,31,32,51,55,61] Experience with their use suggests that they can be used successfully if limited wastewater characterization data are available.[2,31,32] Of particular importance are the fraction of the biodegradable organic matter in the influent wastewater that is readily biodegradable and, when biological phosphorus removal is to be provided, the concentration of VFAs. Appropriate kinetic parameters for autotrophic biomass are also important as they have been observed to vary from one application to the next. If site-specific values of the kinetic parameters are not available for the nitrifying bacteria, standard values can be used with caution to provide approximate estimates of their performance in BNR processes. Due to the numerous interactions, the use of models such as ASM No. 1 and 2d is encouraged to optimize the design of nutrient removal systems.

12.4 PROCESS OPERATION

The operation of BNR systems builds on the operating principles for the activated sludge process presented in Section 11.4.[67] Long-term control of the process is achieved by control of the SRT; the techniques described in Section 11.4 for accomplishing this are equally applicable to BNR systems. Furthermore, the visual observations discussed in that chapter are also equally applicable to BNR systems.[65]

Complete nitrification is necessary if efficient nitrogen removal is to be achieved. Nitrification is controlled by adjustment of the overall SRT to attain the aerobic SRT necessary to maintain the nitrifying bacteria in the system at the given temperature. Adequate DO concentrations must be maintained in the aerobic zone. However, as discussed in Sections 12.1.2 and 12.3.1, if the SRT is long enough, the system can be operated at lower DO concentrations to encourage denitrification in the aerobic zone. The key is the maintenance of a sufficient nitrification safety factor to retain the nitrifiers in the process and to ensure an adequate level of activity.

Efficient denitrification is achieved by excluding DO from the anoxic zones and recirculating sufficient quantities of nitrate-N to initial anoxic zones. Because the MLR flow comes from the aerobic zone, avoiding high DO concentrations in that zone helps to minimize the recirculation of DO with the nitrate-N. Entry of this alternative electron acceptor into the anoxic zone will allow some of the biodegradable organic matter to be consumed aerobically, resulting in less being available for denitrification. Moreover, since the DO concentration in an initial anoxic zone will be low, the recycle of significant quantities of DO to that zone may result in the growth of Group I filamentous bacteria (see Table 11.2), creating a poorly settling sludge. Mixed liquor recirculation provides the operator with considerable control over the mass of nitrate-N that can be denitrified in a BNR system, with increased mass flow rates resulting from increased MLR rates. Increased recirculation rates will increase the power requirements for recirculation pumping, but this must be balanced against the nitrogen removal requirements and the fact that increased denitrification will reduce both the oxygen and alkalinity requirements in the aerobic basin. Mixed liquor recirculation rates that provide nitrate-N in excess of that required can also lead to emissions of nitrous oxide, which should be avoided.

The provision of anaerobic and anoxic zones provides metabolic selection, which effectively controls the growth of Group I filamentous bacteria. However, other filaments, especially the Group III and IV filaments (see Table 11.2), can grow in biological nitrogen and phosphorus removal systems. Experience demonstrates that the adverse impacts of Group III and IV filaments can be controlled by staging the various zones of the bioreactor, especially the aerobic zone.[35,59] Adequate DO must also be provided, especially at the inlet end of the aerobic zone, where the process oxygen demand will be the greatest.

Biological phosphorus removal is optimized by excluding DO and nitrate-N from the anaerobic zone. Preventing the addition of these electron acceptors allows the maximum possible mass of organic matter to be stored by PAOs, thereby providing them with the maximum possible competitive advantage. Aeration of the wastewater upstream of the bioreactor should be avoided, both because it can add DO to the anaerobic zone, and because it can lead to stripping or aerobic biodegradation of VFAs. If fermentation facilities are available, their operation can significantly improve the performance of a BPR system.

In some instances, specialized operational techniques can be applied to BNR systems. Measurement of nutrient profiles through a staged bioreactor system can provide insight into the reactions occurring there. For example, phosphorus release in the anaerobic zone and phosphorus uptake in the aerobic zone can be monitored. These profiles can be used as an early indicator of process upsets, and they can be used to adjust process operating parameters. Nitrate-N profiles through the anoxic zone can be used to indicate incomplete denitrification that can result in nitrate-N recycle to the anaerobic zone, while nitrate-N and ammonia-N profiles through the aerobic zone will indicate the extent of nitrification. Furthermore, the ammonia-N profile can be used to adjust the aerobic SRT to obtain the necessary degree of nitrification. In some systems, alkalinity consumption and/or effluent turbidity are monitored and used to control aeration input to minimum values, allowing optimization of denitrification in the aerobic zone.

As discussed above, nutrients contained in recycle streams from solids handling systems can adversely affect the performance of BNR processes. Consequently, the entire wastewater treatment plant must be viewed as an integrated system. The plant operator needs to monitor and understand the interaction between the liquid and solids treatment trains to optimize overall process performance and cost.

12.5 KEY POINTS

1. Segregation of the bioreactor into anaerobic, anoxic, and aerobic zones distinguishes biological nutrient removal (BNR) systems from other activated sludge systems. Anaerobic zones allow for selection of phosphorus accumulating organisms (PAOs), thereby increasing the phosphorus content of the mixed liquor suspended solids (MLSS) and allowing phosphorus removal in the waste solids. Anoxic zones allow for denitrification, thereby removing nitrogen as N_2. Aerobic zones are necessary for the growth of nitrifying bacteria and PAOs, although denitrifying PAOs can grow when cycled between anaerobic and anoxic zones.
2. Several processes are emerging to treat high-nitrogen solids processing recycle streams. These can be grouped into four categories: (1) select addition of the recycle stream to the mainstream process, (2) sidestream systems that seed the mainstream process with nitrifying bacteria, (3) processes that reduce oxygen requirements by 25% and carbon requirements by 40% by converting ammonia only to nitrite, and (4) autotrophic nitrogen removal processes such as anammox.
3. Phosphorus removal is adversely impacted by the recycle of nitrate-N to the anaerobic zone of a biological phosphorus removal (BPR) process. Nitrate-N allows consumption of volatile fatty acids (VFAs) by heterotrophic denitrifying bacteria, thereby reducing the competitive advantage of the PAOs. It also decreases the production of VFAs by fermentation of readily biodegradable substrate by ordinary heterotrophic organisms (OHOs), thereby lowering the amount of VFAs available.
4. Incorporation of initial anoxic zones into nitrifying activated sludge systems provides several benefits: reduced alkalinity consumption, decreased oxygen requirement, diminished denitrification in the clarifier, and selection of floc forming bacteria at the expense of filaments. The incorporation of anaerobic zones into nonnitrifying systems also offers the benefit of selection of floc forming bacteria.

5. Anoxic zones can be incorporated into nitrogen removal systems both upstream and downstream of the main aerobic zone. The rate of denitrification in an upstream anoxic zone is rapid due to the utilization of readily biodegradable substrate, while it is slow in a downstream anoxic zone because it must depend on decay reactions and utilization of slowly biodegradable substrate.
6. Biological phosphorus removal is maximized by operation at short solids retention times (SRTs). This occurs because: nitrification, and the resulting potential for recycle of nitrate-N to the anaerobic zone, is minimized; sludge production and the resulting wasting of phosphorus from the system is maximized; and uptake of phosphorus by the biomass is maximized.
7. The SRT of a BNR system impacts organism selection. The aerobic SRT must be long enough to allow PAOs to grow if phosphorus removal is desired, while it must be increased further to allow nitrifying bacteria to grow if nitrification is desired. Anaerobic and anoxic SRTs affect phosphorus and nitrogen removal, as well as control of filamentous bacteria.
8. Organic matter to nutrient ratios provide an indication of the suitability of wastewater for biological nutrient removal. The organic matter to nitrogen ratio indicates whether sufficient organic matter is available to denitrify the nitrate-N generated. The organic matter to phosphorus ratio indicates whether sufficient organic matter is available to grow an adequate quantity of PAOs to remove the influent phosphorus. If both nitrogen and phosphorus are to be removed in one system, the organic matter requirement for each should be considered and the larger amount determined. It will generally control overall performance.
9. The composition of the organic matter in a wastewater affects the performance of BNR systems. Volatile fatty acids are utilized by PAOs, and highly efficient phosphorus removal can be achieved with a small anaerobic zone when a high proportion of the influent organic matter is VFAs. If insufficient VFAs are present, readily biodegradable organic matter can be fermented to generate VFAs, but the reaction is slow, and thus the anaerobic zone must be larger. Readily biodegradable organic matter also results in a high denitrification rate in initial anoxic zones.
10. Wastewaters can be fermented to increase their VFA content and improve the performance of BNR systems. Fermentation can occur in the wastewater collection systems and in specially designed facilities at the wastewater treatment plant.
11. Due to the high phosphorus content of the MLSS in a BPR system, the effluent total suspended solids (TSS) can contribute a significant quantity of phosphorus to the process effluent. They also contribute some organic-N.
12. Biological nutrient removal processes are designed using the same procedures as for other activated sludge systems. Solids retention times must be chosen for each of the zones (aerobic, anoxic, and anaerobic) and summed to obtain the system SRT. Using the system SRT, the mass of MLSS, the solids wastage rate, and the total electron acceptor requirement are calculated using the equations from Chapter 10. The bioreactor is sized based on the mass of MLSS and the design MLSS concentration, and divided into the separate zones in proportion to their SRTs. Aerobic zone volumetric power inputs must be checked relative to the requirements for floc shear and mixing, and mixing energy inputs must be determined for anaerobic and anoxic zones. Mixed liquor recirculation (MLR) rates must also be calculated.
13. A key requirement for a biological nitrogen removal system is stable nitrification. Thus, selection of the aerobic SRT for such a system is governed by the maintenance of a stable population of nitrifiers.
14. Anoxic selectors are sized to remove readily biodegradable substrate, which typically requires an anoxic SRT of 1 day in summer and 1.5 days in winter for domestic wastewater.

The MLR rate is set to return enough nitrate-N to the selector to accept the electrons removed during growth of heterotrophs on the readily biodegradable substrate. Greater MLR rates create the potential for the formation of the greenhouse gas nitrous oxide in the anoxic selector.

15. The initial anoxic zone in the modified Ludzak-Ettinger (MLE) process is designed to remove sufficient nitrate-N to achieve a desired effluent concentration, which typically requires utilization of some slowly biodegradable substrate in addition to the readily biodegradable substrate. Because of uncertainties about the utilization rate of slowly biodegradable substrate, an empirical expression is used to select the anoxic SRT. The MLR rate is selected to return sufficient nitrate-N to the anoxic zone to meet the desired effluent concentration, provided sufficient electron donor is available.
16. The second anoxic zone in a four-stage Bardenpho system is sized to remove additional nitrate-N by anoxic biomass decay. Selection of its SRT requires the use of an empirical relationship. The introduction of the second anoxic zone essentially eliminates the recirculation of nitrate-N to the initial anoxic zone via the return activated sludge (RAS) flow, thereby requiring a larger MLR rate to achieve the same degree on nitrate-N removal in the initial anoxic zone.
17. Careful control of the oxygen transfer rate to a nitrifying activated sludge system provides the opportunity for simultaneous nitrification and denitrification in a single vessel, provided that the SRT is sufficiently long for stable nitrification under reduced dissolved oxygen (DO) conditions. However, it is possible that such an activity will lead to filamentous bulking by Group IV bacteria (Table 11.2).
18. Selection of the aerobic SRT for a process to remove phosphorus but not nitrogen requires careful consideration of the minimum SRT required for both PAOs and nitrifiers. Under winter operating conditions it is usually possible to select an aerobic SRT that will allow growth of the PAOs while excluding the nitrifiers, but this will be difficult to do in summer.
19. Selection of the anaerobic SRT for a BPR process is influenced strongly by the nature of the readily biodegradable substrate and the recycle of nitrate-N to the anaerobic zone. If the readily biodegradable substrate is high in VFAs, then short anaerobic SRTs can be used. Fermentation of readily biodegradable substrate to form VFAs requires longer anaerobic SRTs, however.
20. The amount of phosphorus removed by a BPR process is determined by the amounts of readily biodegradable substrate and phosphorus in the influent, and the recycle of nitrate-N to the anaerobic zone by the RAS flow. Approximately 7.5 to 10.7 mg of VFA chemical oxygen demand (COD) are needed to remove 1.0 mg of phosphorus. In addition, around 6 mg of readily biodegradable COD are needed to remove 1.0 mg of recycled nitrate-N.
21. The design of BNR processes for the removal of both nitrogen and phosphorus is accomplished by combining the techniques used to design processes that remove either nitrogen or phosphorus. Because nitrification is necessary in systems removing both nutrients, a key task during the design is to minimize the entry of nitrate-N into the anaerobic zone.
22. Because of the complex interactions among the members of the microbial community in them, dynamic simulation is a valuable tool for optimizing the design of BNR systems. International Water Association (IWA) Activated Sludge Models (ASMs) No. 1 and 2d have been used for this purpose and have proven useful in a number of full-scale applications.
23. The operation of BNR systems is similar to that of other activated sludge processes, and the operational techniques applied to the activated sludge process are equally applicable. The SRT is controlled to provide long-term regulation of the process using the same

approaches used with activated sludge systems. Visual observations of the bioreactor, the clarifier, the settleometer, and under the microscope can also be used to optimize the process. Specialized operational techniques used with BNR processes include measurement of nutrient profiles, alkalinity consumption, and effluent turbidity.

24. The operators of wastewater treatment plants containing a BNR process must recognize the interactions between the liquid and solids processing trains.

12.6 STUDY QUESTIONS

1. Prepare a table comparing the bioreactor configuration, SRT, HRT, recycle ratio, MLR ratio, and design approach for the following: MLE, four-stage Bardenpho, five-stage Bardenpho, A/O, A^2/O, VIP, and UCT processes.
2. Describe the biochemical transformations occurring in a BNR process that achieves both nitrogen and phosphorus removal. Where do these transformations occur, and how do they interact to allow nutrient removal to occur?
3. Describe the benefits that biological nutrient removal can provide to the operation of an activated sludge system and list the circumstances under which nutrient removal capabilities would be incorporated into an activated sludge system design even though the effluent quality criteria did not require it.
4. Describe the mechanisms that allow denitrification to occur in a supposedly aerobic bioreactor. How can denitrification be increased? What are the impacts of this denitrification on the operational and performance characteristics of the system?
5. Make a list of all the process design and operational factors that maximize phosphorus removal in a BNR system. Make a similar list of factors that maximize nitrogen removal. Compare the two lists and identify those factors that are similar for the two systems and those that are different. Discuss the impacts of these similarities and differences on the design of a system that removes both nitrogen and phosphorus.
6. Make a list of factors that can affect the characteristics of a wastewater as it enters the first bioreactor in a BNR system and the resulting impact of those characteristics on the performance of a BNR system. Identify which factors are under the control of the process designer and which are under the control of the process operator. What does this analysis indicate about the design and operation of a BNR system?
7. Describe the concept of the organic matter to phosphorus removal ratio. How can this ratio be used as a screening tool to identify the BPR processes to be considered for a particular application?
8. List the factors that affect the size of the initial anoxic zone required for a particular nitrogen removal application. Under what circumstances would a relatively small zone be used? When would a relatively large zone be used?
9. List the factors that affect the size of the anaerobic zone required for a particular BPR application. Under what circumstances would a relatively small zone be used? When would a relatively large zone be used?
10. Derive the equation used to calculate the fraction of nitrate-N recirculated from the aerobic zone of a biological nitrogen removal system to an upstream anoxic zone. Do this for both the MLE and the four-stage Bardenpho systems and explain why the expressions are different.
11. Using the wastewater characteristics in Table E9.4, the stoichiometric and kinetic parameters in Table E11.2, and the temperature correction factors in Table E11.1, design a CAS system with an MLSS concentration of 2750 mg/L as TSS to produce a fully nitrified effluent year-round while treating an average wastewater flow rate of 30,000 m^3/day. The lowest sustained winter temperature is 13°C and the highest sustained summer temperature is 24°C. Use a

diffused air oxygen transfer system and assume that the in-process oxygen transfer efficiency is 12%. Also assume that the hydraulic characteristics of the CAS bioreactor are equivalent to three tanks in series and that the safety factor for uncertainty is 1.0. The diurnal peak loading on the system is twice the average loading. Use this information in selecting the aerobic SRT for the system, but for simplicity, base all decisions about tank volumes and oxygen transfer rates on average loading conditions. Justify all assumptions and decisions.

12. Add an anoxic selector capable of removing all readily biodegradable substrate to the CAS system designed in Study Question 11. Determine its volume and the MLR rate. Justify all assumptions and decisions. Compare the following for the systems with and without the selector: total system volume, aerobic bioreactor volume, oxygen transfer rate and air flow rate to the aerobic zone, and alkalinity that must be added to maintain a residual concentration of 50 mg/L as $CaCO_3$.
13. Use a computer code implementing ASM No. 1 or a similar model to evaluate the effluent nitrate-N concentration and the oxygen requirement in the system designed in Study Question 12. Comment on any differences between the results from Study Question 12 and the simulation results, suggesting possible reasons for differences.
14. For the situation described in Study Question 11, design an MLE system to produce an effluent containing no more than 10 mg/L as N of nitrate-N, determining the volume of the anoxic zone and the MLR rate. Justify all assumptions and decisions. Compare the following to the system designed in Study Question 12: total system volume, aerobic bioreactor volume, oxygen transfer rate and air flow rate to the aerobic zone, and alkalinity that must be added to maintain a residual concentration of 50 mg/L as $CaCO_3$.
15. Use a computer code implementing ASM No. 1 or a similar model to evaluate the effluent nitrate-N concentration and the oxygen requirement in the system designed in Study Question 14. Comment on any differences between the results from Study Question 14 and the simulation results, suggesting possible reasons for differences.
16. Add a second anoxic zone to the process considered in Study Question 14, to lower the effluent nitrate-N concentration to 3 mg/l as N. Determine the size of the second anoxic and aerobic zones and calculate the MLR required. Justify all assumptions and decisions. Compare the following to the system designed in Study Question 14: total system volume, aerobic bioreactor volume, oxygen transfer rate and air flow rate to the aerobic zone, alkalinity that must be added to maintain a residual concentration of 50 mg/L as $CaCO_3$.
17. Use a computer code implementing ASM No. 1 or a similar model to evaluate the effluent nitrate-N concentration and the oxygen requirement in the system designed in Study Question 16. Comment on any differences between the results from Study Question 16 and the simulation results, suggesting possible reasons for differences.
18. Redo Study Question 16, but in this case, size the first anoxic zone to remove the amount of nitrate-N that can be returned with an MLR rate of four times the influent flow rate. Compare this system to the one designed in Study Question 16 and comment on the differences.
19. Evaluate the possibility of operating the oxygen transfer system in the CAS system designed in Study Question 11 in such a way that 45% of the nitrate-N formed during summer operation is denitrified. At what oxygen transfer rate would the system have to be operated to achieve the desired degree of denitrification? In making your assessment, assume that the reduction in oxygen input will result in an average DO concentration of 0.5 mg/L and that the anoxic growth factor, η_g, is 0.75.
20. For the situation described in Study Question 11, prepare the design of an A/O process operating at a temperature of 20°C at steady state. Assume that the MLSS concentration is 2500 mg/L as TSS. If the total phosphorus concentration of the influent wastewater is 8 mg/L as P and the effluent suspended solids concentration is 15 mg/L as TSS, what is the estimated effluent total phosphorus concentration for this process? What is the estimated effluent nitrate-N concentration? State and justify all assumptions.

21. For the situation described in Study Question 11, prepare the design of a VIP process to produce an effluent with minimal ammonia-N and phosphorus concentrations while operating at steady state at 20°C. Assume that the MLSS concentration is 3500 mg/L as TSS. If the total phosphorus concentration of the influent wastewater is 12 mg/L as P and the effluent suspended solids concentration is 10 mg/L as TSS, what is the estimated effluent total phosphorus concentration for this process? What is the estimated effluent nitrate-N concentration? State and justify all assumptions.
22. Using a computer code implementing ASM No. 1 or a similar model, systematically investigate the effects of the anoxic SRT and MLR rate on the performance of the MLE process developed in Study Question 14. Also determine the impact of staging the bioreactors on overall process performance. Discuss the implications of your findings to an optimal system design.
23. Using a computer code implementing ASM No. 1 or a similar model, systematically investigate the effects of the distribution of the anoxic SRT between the first and second anoxic zones on the performance of the four-stage Bardenpho process developed in Study Question 16. Adjust the MLR rate to match the nitrate-N need in the first anoxic zone for each size investigated. Also determine the impact of staging the bioreactors on overall process performance. Discuss the implications of your findings to an optimal system design.
24. Using a computer code implementing ASM No. 2d or a similar model, systematically investigate the effects of the anaerobic and aerobic SRTs on the performance of the A/O process developed in Study Question 20. Also determine the impact of staging the bioreactors on overall process performance. Discuss the implications of your findings to an optimal system design.
25. Using a computer code implementing ASM No. 2d or a similar model, systematically investigate the effects of the anaerobic and anoxic SRTs on the performance of the VIP process developed in Study Question 21. Adjust the MLR rate to match the nitrate-N need in the anoxic zone for each size investigated. Also determine the impact of staging the bioreactors on overall process performance. Discuss the implications of your findings to an optimal system design.
26. Discuss the impact of solids handling recycles on the performance of a BNR system. What steps can be taken to mitigate these impacts?

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13 Aerobic Digestion

The term aerobic digestion refers to the use of aerobic bioreactors to stabilize particulate organic matter arising from primary clarification (predominantly biodegradable organic matter) and biological treatment (predominantly biomass) of wastewaters. The solids are oxidized using either dissolved oxygen or nitrate-N as the terminal electron acceptor. The resulting residue consists primarily of a relatively inert, humus-like material that degrades quite slowly (months to years) in both aerobic and anaerobic environments. The destruction of pathogens is also an objective in municipal wastewater treatment facilities.

13.1 PROCESS DESCRIPTION

Aerobic digestion has been used for several decades to stabilize the waste solids produced at municipal and industrial wastewater treatment plants.[41] Initially its popularity increased but this trend was halted in the mid-1970s as rapidly escalating energy costs adversely impacted its overall cost-effectiveness relative to other solids stabilization options. Then, in 1979, federal regulations governing the management of solids from municipal wastewater treatment plants were issued that set new requirements controlling pathogens when solids are to be reused.[24,40] This further decreased the attractiveness of aerobic digestion for municipal wastewater treatment plants since its rates of pathogen inactivation are generally lower than anaerobic digestion. Nevertheless, aerobic digestion remained a popular option for small plants treating industrial wastewaters because of its simplicity. In addition, thermophilic aerobic digestion processes, which have higher solids stabilization and pathogen inactivation rates, were developed allowing aerobic digestion to remain a viable option for the stabilization of waste solids.[10,42,43]

13.1.1 General Description

Figure 13.1 summarizes the biochemical transformations occurring in an aerobic digester. Biodegradable particulate organic matter is hydrolyzed and converted into biodegradable soluble organic matter, releasing nutrients such as ammonia-N and phosphate. The biodegradable soluble organic matter is then converted into carbon dioxide, water, and active biomass through the action of heterotrophic bacteria. The active biomass, in turn, undergoes decay, resulting in the generation of additional carbon dioxide and water along with inactive biomass (i.e., debris). Nonbiodegradable particulate organic matter in the influent is not affected by the digestion process and becomes a portion of the digested solids. Figure 13.1 is based on the traditional decay model for biomass destruction, as discussed in Section 3.3.1. The lysis:regrowth model, described in Section 3.3.2, is equally applicable and, in fact, International Water Association (IWA) Activated Sludge Model (ASM) No. 1, with its explicit treatment of hydrolysis, nitrification, and denitrification, provides a more accurate description for some aerobic digestion process options. Nevertheless, the simplified models often used to design aerobic digesters are directly related to the traditional decay model and thus it is emphasized herein.

Observations of aerobic digestion processes provide the following conceptual framework upon which design models are based:

- The suspended solids in the influent stream can be segregated into biodegradable and nonbiodegradable components.[14,36] The biodegradable components include particulate organic matter, X_S, and active biomass, both heterotrophic and autotrophic $X_{B,H}$ and $X_{B,A}$. The

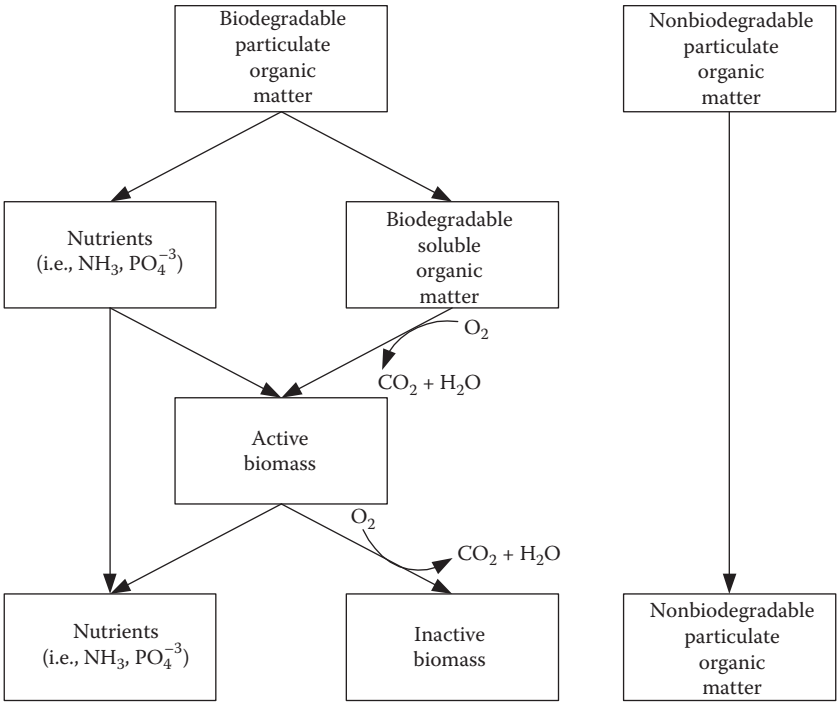

FIGURE 13.1 Schematic diagram of the events occurring during aerobic digestion.

nonbiodegradable component consists of particulate inert organic matter, X_I, and biomass debris, X_D.

- A nonbiodegradable residue will result from aerobic digestion, even if no nonbiodegradable particulate matter is present in the influent solids stream because biomass debris results from the decay of active biomass.[36,45]
- Aerobic digestion results in the destruction of both volatile suspended solids (VSS) and fixed suspended solids (FSS).[7,36] This occurs because both the organic and inorganic materials in the biodegradable suspended solids are solubilized and/or oxidized as the solids are digested. However, the volatile and fixed components of the biodegradable and nonbiodegradable suspended solids are not equal. Consequently, VSS and FSS will not generally be destroyed in the same proportion. However, in spite of the loss of fixed solids during aerobic digestion, most designers focus on loss of VSS.
- The biodegradable fraction of solids is a function of their source.[25,36,45] This is clearly illustrated by the models discussed throughout previous chapters. For example, both primary solids and waste activated sludge from a system with a short solids retention time (SRT) will contain relatively high fractions of biodegradable material, whereas waste activated sludge from a system with a long SRT will contain a low fraction of biodegradable material and a high fraction of biomass debris.
- The destruction of biodegradable suspended solids can be characterized as a first-order reaction.[14,22,36] This occurs because the decay of active biomass is a first-order reaction. Biodegradable particulate organic matter is rapidly converted to active biomass. Then that biomass, as well as any active biomass present in the influent, decays in a first-order manner, resulting in an overall first-order reaction for loss of biodegradable suspended solids. As a result of this relationship, the destruction of biodegradable suspended solids is often referred to as decay, and the first-order reaction rate coefficient is called a decay coefficient.

- For solids containing a relatively high proportion of active biomass, the value of the decay coefficient for biodegradable suspended solids is relatively independent of the SRT at which the waste solids were produced.[25,36,45] This is because the decay coefficient for the biodegradable suspended solids will be influenced strongly by the decay coefficient for heterotrophic bacteria, which is relatively constant.

Mathematically, these relationships can be summarized as follows. Because design of aerobic digesters is typically concerned with VSS destruction, the concentrations will be expressed as VSS, although it should be recognized that they could also be expressed as total suspended solids (TSS) or chemical oxygen demand (COD). The VSS undergoing aerobic digestion, $X_{M,V}$, can be subdivided into biodegradable and nonbiodegradable components, $X_{M,V,b}$ and $X_{M,V,n}$, respectively:

$$X_{M,V} = X_{M,V,b} + X_{M,V,n}. \tag{13.1}$$

Furthermore, loss of biodegradable VSS occurs in a first-order manner:

$$r_{XMVb} = -b_{MV} \cdot X_{M,V,b}, \tag{13.2}$$

where b_{MV} is the first-order decay coefficient based on the loss of VSS. Decay coefficients can also be determined on the basis of COD or TSS, depending upon how the solids concentration is expressed. Because the COD/VSS ratio can be considered to be constant, the decay coefficients based on VSS and COD loss have the same numerical values. However, because VSS and FSS are not generally destroyed in proportion to each other, the decay coefficient on a TSS basis will have a different numerical value. The nonbiodegradable solids are considered to be totally inert so that nothing happens to them during aerobic digestion.

Sometimes it is desirable to consider the solids in terms of the constituents used in the simplified model of Chapter 5 and in ASM No. 1. In those terms, the VSS consists of

$$X_{M,V} = X_{S,V} + X_{B,H,V} + X_{B,A,V} + X_{D,V} + X_{I,V}. \tag{13.3}$$

All symbols have been defined previously; the subscript V simply indicates that they are expressed on a VSS basis. The particulate substrate and the biomass are considered to be biodegradable, although decay of biomass leads to debris. The debris and the inert material are, of course, nonbiodegradable. The fate of these components can be modeled in ASM No. 1, just as the various activated sludge systems are modeled in Chapter 7. This has been done by Marais and coworkers[25,45] using the model upon which much of ASM No. 1 was based. The simplified model of Chapter 5 does not contain a term for particulate substrate and thus it is not as accurate as ASM No. 1. Nevertheless, it has been used to model aerobic digestion by lumping the particulate substrate with the biomass and considering the system to be a bioreactor receiving only biomass and nonbiodegradable solids. In that situation, the autotrophic biomass is generally neglected as being insignificant. This is done in Section 5.2.3 for TSS units, with Equations 5.76 and 5.77 expressing the heterotrophic biomass and debris concentrations, respectively, in a single CSTR. Equation 5.80 gives the oxygen requirement for such a bioreactor when nitrification is not considered. Figure 5.10 shows the effects of SRT on the theoretical performance of a single CSTR receiving only active heterotrophic biomass and debris. There it can be seen that a point of diminishing returns is reached at which further increases in the SRT have little effect. When that point is reached, most of the suspended solids will be nonbiodegradable, with only a small fraction of active biomass.

Since the purpose of aerobic digestion is to stabilize the biodegradable organic matter in an influent waste solids stream, criteria must be available for quantifying the degree of stabilization. Although several could be proposed, two frequently used ones are the VSS destruction efficiency

(expressed as the percentage of VSS reduction) and the specific oxygen uptake rate (SOUR) of the digested solids (typically expressed as mg O_2/(g VSS·hr)).[12] A VSS reduction of 38% and an SOUR of 1.0 to 1.5 mg O_2/(g VSS·hr) are values typically used to represent stabilized solids.[24,40]

The simple first-order model presented above can be used to estimate the effect of the size of a completely mixed aerobic digester on the degree of solids stabilization achieved. The VSS destruction efficiency, E_{XMV}, in a CSTR can be calculated as a function of its SRT, Θ_c:

$$E_{XMV} = 100\left[\frac{X_{M,VO} - X_{M,V,nO}}{X_{M,VO}}\right]\left[\frac{b_{MV}\cdot\Theta_c}{1+b_{MV}\cdot\Theta_c}\right], \tag{13.4}$$

where the subscript O represents the influent concentrations. From Equation 13.1:

$$X_{M,V,bO} = X_{M,VO} - X_{M,V,nO}. \tag{13.5}$$

Thus, it can be seen that the term $X_{M,VO} - X_{M,V,nO}$ is just the concentration of biodegradable VSS entering the digester. Consideration of Equation 13.4 in the limit as the SRT becomes very large reveals that the first bracketed term on the right side represents the highest possible VSS destruction efficiency since it is just the fraction of the influent VSS that can be degraded biologically. It is an important determinant of whether a target destruction efficiency of 38% can be economically achieved.

Calculation of the actual VSS destruction efficiency in an operating digester should be based on the mass flow rates of VSS entering and leaving it. While some sources[42] use the percentage VSS content of the feed and effluent solids to make this calculation, such a procedure can give an inaccurate measure because of changes in the FSS as discussed above. Consequently, it should not be used.

The SOUR for a completely mixed aerobic digester can also be calculated as a function of the SRT:

$$SOUR = 1000\left[\frac{b_{MV}\cdot i_{O/XM,V}\left(X_{M,VO} - X_{M,V,nO}\right)}{\left(X_{M,VO} - X_{M,V,nO}\right) + X_{V,M,nO}\left(1+b_{MV}\cdot\Theta_c\right)}\right], \tag{13.6}$$

where $i_{O/XM,V}$ is the conversion factor from VSS units to oxygen units. The numerical value for $i_{O/XM,V}$ depends on whether the ammonia-N released through digestion of biodegradable VSS is nitrified in the aerobic digester. If the biodegradable VSS can be assumed to have the same elemental composition as biomass (i.e., $C_5H_7O_2N$), then the values of $i_{O/XM,V}$ can be determined theoretically. If the released ammonia-N is not oxidized to nitrate-N by autotrophic biomass, the value will be that given in Table 3.1 or 1.42 g O_2/g VSS. If, on the other hand, the released ammonia-N is nitrified, the value of $i_{O/XM,V}$ will be increased by 40%, to 1.98 g O_2/g VSS. Table 13.1 summarizes these values as well as the values the conversion factors would have if the solids concentrations were expressed in COD or TSS units. If the biodegradable solids cannot be assumed to have the same elemental composition as biomass, the values of the conversion factor must be determined experimentally.

TABLE 13.1
Oxygen Mass Equivalents for Biomass

Condition	$i_{O/XM}$ g O_2/g COD	$i_{O/XM,V}$ g O_2/g VSS	$i_{O/XM,T}$ g O_2/g TSS
With nitrification	1.40	1.98	1.68
Without nitrification	1.00	1.42	1.20

Although Equations 13.4 and 13.6 are for a CSTR, many aerobic digester studies are done in batch reactors. The VSS concentration in a batch reactor at any time t can be estimated by writing the mass balance equation on biodegradable VSS with Equation 13.2 as the loss term, integrating it, and adding the initial nonbiodegradable VSS concentration to give the total VSS concentration:

$$X_{M,V} = X_{M,V,nO} + (X_{M,VO} - X_{M,V,nO})\exp(-b_{MV} \cdot t). \tag{13.7}$$

This equation indicates that the VSS concentration will decline exponentially over time, but will approach a residual concentration equal to the nonbiodegradable solids.

The effects described in this section have been observed experimentally, as illustrated by the data of Reece et al.[36] who performed aerobic digestion studies in batch reactors using solids produced in an activated sludge system. Since the wastewater was totally biodegradable, the waste solids consisted primarily of active biomass and biomass debris. The SRT of the activated sludge system in which the waste solids were produced was varied to produce waste solids with varying proportions of active and inactive biomass. As indicated in Figure 13.2, both VSS and FSS (referred to in the figure as nonvolatile suspended solids or NVSS) were destroyed during aerobic digestion. In fact, the proportions of VSS and FSS destroyed were similar. However, Figure 13.3 shows that the decay coefficient was relatively constant and independent of the SRT of the activated sludge system

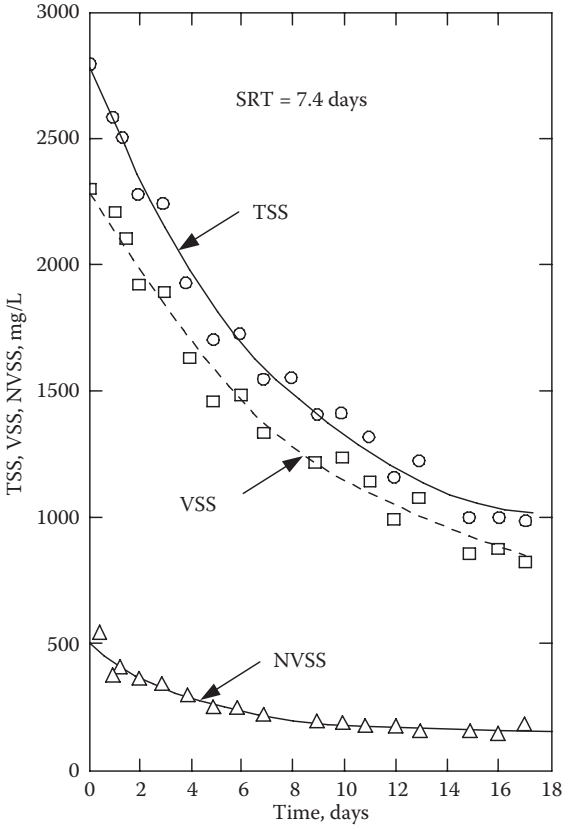

FIGURE 13.2 Destruction of TSS, VSS, and FSS (called NVSS) during batch aerobic digestion of waste activated sludge. (From Reece, C. S., Roper, R. E., and Grady, C. P. L., Jr., Aerobic digestion of waste activated sludge. *Journal of the Environmental Engineering Division, ASCE,* 105:261–72, 1979. Copyright © American Society of Civil Engineers. Reprinted with permission.)

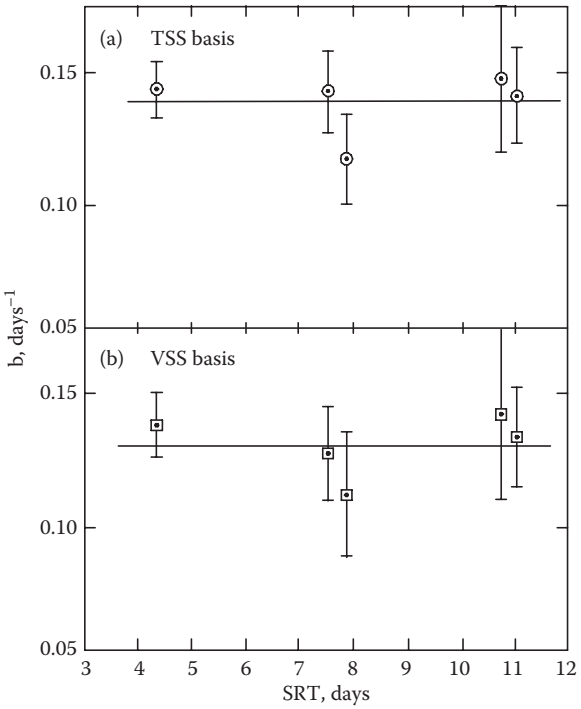

FIGURE 13.3 Effect of the SRT of an activated sludge system on the decay coefficient describing aerobic digestion of the resulting waste solids. (a) TSS basis and (b) VSS basis. (From Reece, C. S., Roper, R. E., and Grady, C. P. L., Jr., Aerobic digestion of waste activated sludge. *Journal of the Environmental Engineering Division, ASCE,* 105:261–72, 1979. Copyright © American Society of Civil Engineers. Reprinted with permission.)

in which the waste solids were generated, although its numerical value depended on whether it was quantifying the loss of TSS or VSS. In contrast, the nonbiodegradable suspended solids content of the waste solids increased as the SRT of the activated sludge system was increased, as illustrated in Figure 13.4. Similar effects have been observed by others.[33,45]

If waste solids from municipal wastewater treatment plants are to be reused, either the sludge stabilization process or the quality of the biosolids produced must meet certain requirements. Although requirements vary from one country to another, those implemented in the United States are typical.[24,40,42,46,47] Referred to as the "503 Regulations" because of the section in the Federal Register in which they were published, they establish requirements for the reduction of biodegradable organic matter (also referred to as the stability of the solids and called the "vector reduction" requirements in the regulations) and pathogens. Aerobic digestion processes can be used to produce "Class A" or "Class B" biosolids, which can be reused in various ways. Both can be applied in agriculture as a fertilizer and soil conditioner, but public access and agricultural practices are further restricted for Class B biosolids (for example, public access must be restricted for 12 months following biosolids application). Table 13.2 summarizes the 503 Regulations relevant to aerobic digestion.

13.1.2 Process Options and Comparison

Three basic aerobic digestion process options exist. They are conventional aerobic digestion (CAD), anoxic/aerobic digestion (A/AD), and autothermal thermophilic aerobic digestion (ATAD) as discussed in Section 1.3.1. Hybrid options incorporating some of the features of these basic options are also being developed.

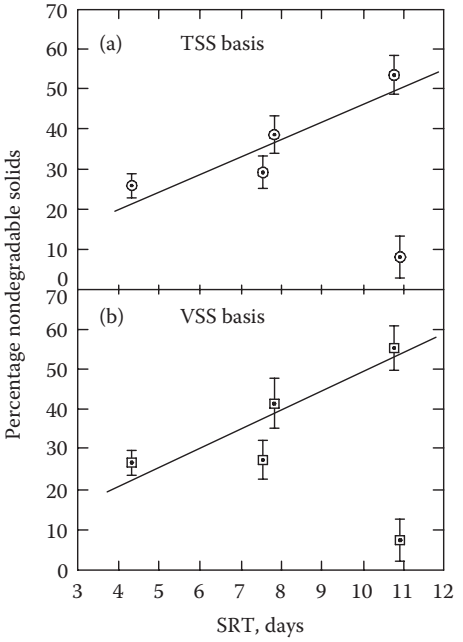

FIGURE 13.4 Effect of the SRT of an activated sludge system on the nonbiodegradable suspended solids content of the resulting waste solids. (a) TSS basis and (b) VSS basis. (From Reece, C. S., Roper, R. E., and Grady, C. P. L., Jr., Aerobic digestion of waste activated sludge. *Journal of the Environmental Engineering Division, ASCE,* 105:261–72, 1979. Copyright © American Society of Civil Engineers. Reprinted with permission.)

TABLE 13.2
503 Regulations as Applied to Aerobic Digestion[a]

Vector reduction

- 38% volatile solids (VS) reduction, or
- SOUR less than 1.5 mg O_2/[g total solids (TS)·hr] at 20°C, or
- Less than 15% VS reduction following 30 days of additional digestion at 20°C

Class B pathogen reduction

- An SRT of at least 40 days at 20°C and 60 days at 15°C, or
- Fecal coliform less than 2×10^6 MPN/g TS (MPN = Most Probable Number)

Class A pathogen reduction

- An SRT of at least 10 days at 55°C–60°C, or
- Fecal coliform less than 1000 MPN/g TS, or
- *Salmonella* less than 3 MPN/4 g TS

[a] From U.S. Environmental Protection Agency, *Control of Pathogens in Municipal Wastewater Sludge,* EPA/625/10-89/006, U.S. Environmental Protection Agency, Cincinnati, OH, 1989.

13.1.2.1 Conventional Aerobic Digestion

Conventional aerobic digestion is quite simple. It consists of the addition of solids to an aerated vessel and their retention there for a period of time equal to the SRT, as illustrated in Figure 1.17 where two CAD bioreactors, one with intermittent and one with continuous addition of feed solids, are shown.

In the intermittent process (Figure 1.17a) solids are added and removed from the digester periodically. This process is often used in conjunction with biological wastewater treatment systems in which solids are wasted periodically over the day, often over a relatively short time period. Digested solids are removed from the digester as necessary, depending on the downstream solids handling system. In some moderate size wastewater treatment facilities, solids are thickened prior to addition to the digester. In those cases, the SRT of the digester will be equal to its hydraulic residence time (HRT). When thickening is not used, the aerobic digester is operated like a sequencing batch reactor (SBR) to provide both solids thickening and digestion. Consequently, the SRT is greater than the HRT. The typical steps for operation of an SBR are shown in Figure 1.4. In this application digested solids are withdrawn for further processing at the end of the supernatant draw period. A suspended solids concentration in the range of 17,500 to 22,500 mg/L can typically be achieved if waste activated sludge is being digested. Somewhat higher concentrations can be achieved if a mixture of primary solids and waste activated sludge (20,000 to 30,000 mg/L) or primary solids alone (30,000 to 40,000 mg/L) is being digested. Although feeding is intermittent, feed is added many times during one SRT, making the bioreactor perform like a continuous stirred tank reactor (CSTR) rather than like an SBR.

Solids may also be wasted from a biological wastewater treatment system on a more continuous basis, a practice often used in larger plants. Figure 1.17b illustrates an aerobic digestion system that receives feed on a continuous basis. It looks like an activated sludge system, with feed solids displacing digesting solids to a gravity thickener. Supernatant overflows the thickener, while thickened solids are withdrawn from its bottom and returned to the digester. Thickened solids are also periodically directed to solids handling, with the rate of thickened solids removal being adjusted to maintain the desired SRT. Suspended solids concentrations in the thickened solids are similar to those achieved with the intermittent feed process. Consequently, suspended solids concentrations within the continuous feed digester are lower.

Other operating modes are possible for the intermittent and the continuous feed systems. For example, because VSS destruction is first-order, arrangement of aerobic digesters in series can increase the efficiency of the process, as discussed in Section 13.2.5. Series operation is being used increasingly because of the increased efficiency of both VSS and pathogen destruction.

Conventional aerobic digesters are constructed using facilities and equipment similar to those used for activated sludge systems.[41,46,47] In fact, the aerobic digester may simply be one or more of the aeration basins provided in an activated sludge system. The bioreactors can be concrete or steel or they can be lined earthen basins. Submerged aeration systems, such as diffused air, may be more desirable in cold climates than mechanical surface aerators because they minimize heat losses, which can be quite significant because of the relatively long HRTs. However, surface aeration systems can be used, particularly in warm climates. Decant devices are required for intermittently fed digesters and several approaches have been successfully utilized. Solids retention times on the order of 20 days will usually produce a significant VSS destruction efficiency, although longer SRTs may be required to achieve an SOUR less than 1 mg O_2/(g VSS·hr). Relatively long SRTs may also be required to meet pathogen destruction requirements, depending on the operating temperature, as illustrated in Table 13.2.

13.1.2.2 Anoxic/Aerobic Digestion

One difficulty often experienced with CAD is destruction of alkalinity by nitrification, as discussed in Section 6.3.3. The destruction of organic matter, particularly active biomass, results in liberation of organic nitrogen as ammonia-N. Because of the long SRTs required to accomplish solids stabilization, nitrifying bacteria typically grow in the digester even if they are not present in the feed solids. Furthermore, because of the relatively high feed solids concentrations typically used, the ammonia-N concentrations that develop are high, causing nitrification to deplete the alkalinity in the system, dropping the pH. Figure 13.5 illustrates typical pH profiles during digestion of a waste activated sludge.[30] Three operational modes and three temperatures were considered in this study.

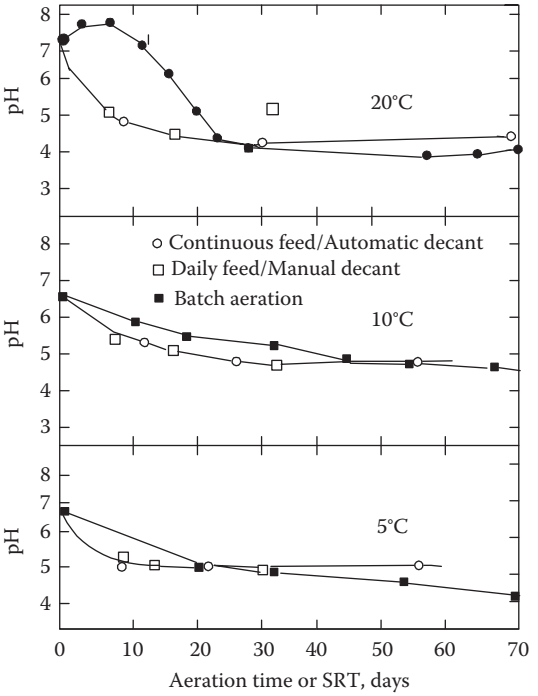

FIGURE 13.5 Depression of the pH during aerobic digestion. (From Mavinic, D. S. and Koers, D. A., Fate of nitrogen in aerobic sludge digestion. *Journal, Water Pollution Control Federation,* 54:352–60, 1982. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

Two modes involved feeding of the digester, either on a continuous or a daily basis. The third was batch operation with only one feed addition. In all cases the pH dropped to relatively low values (4.5 to 5.5), which are typical of those that can be experienced in aerobic digesters without pH control. Although digestion will continue at these lower pH values, the rate will be reduced. The pH of the digester can be adjusted through the addition of bases such as lime, but this is an additional operating cost.

Just as is done with anoxic selectors (Section 12.3.1), anoxic/aerobic digestion incorporates an anoxic cycle to allow alkalinity production by denitrification to partially offset that consumed through nitrification.[45] An additional benefit is a small reduction in process energy requirements since some of the organic matter is oxidized by using nitrate-N generated through nitrification, rather than oxygen, as the electron acceptor.

As discussed above, digestion of cellular material using dissolved oxygen as the terminal electron acceptor, with concurrent oxidation of the ammonia-N to nitrate-N, leads to the net destruction of alkalinity. When the production of autotrophic biomass is neglected (since it is small), the molar stoichiometry is

$$C_5H_7O_2N + 7\ O_2 + HCO_3^- \rightarrow 6\ CO_2 + \ NO_3^- + 4\ H_2O. \tag{13.8}$$

Consequently, one mole of bicarbonate alkalinity is destroyed for each mole of biomass destroyed. Furthermore, seven moles of oxygen are used for each mole of biomass destroyed, which is equivalent to 1.98 g O_2/g VSS destroyed as shown in Table 13.1. However, if a situation could be devised so that all of the nitrogen released was nitrified and then converted to nitrogen gas through denitrification with biomass as the electron donor, the overall molar stoichiometry would be

$$4\ C_5H_7O_2N + 23\ O_2 \rightarrow 20\ CO_2 + 2\ N_2 + 14\ H_2O. \tag{13.9}$$

In that case there would be no alkalinity destruction and the oxygen requirement would be reduced to 5.75 moles per mole of biomass destroyed or 1.63 g O_2/g VSS destroyed, which represents an 18% savings compared to the fully aerobic process. Nitrogen removal is also provided, which can be important if the digested sludge is subsequently dewatered and the separated clear water returned to the main liquid stream of a biological nutrient removal (BNR) plant. Considerations such as these led to the concept of the A/AD process. Evidence indicates that pH values near neutrality can be maintained with this process if a significant degree of denitrification is obtained.

Figure 1.18 illustrates the A/AD process options. The option in Figure 1.18a is a modification of intermittent CAD in which the oxygen transfer system is cycled on and off to create aerobic and anoxic periods during the digester operational cycle. Mixers may be provided to maintain solids in suspension during the anoxic periods, but if a sufficiently high suspended solids concentration is maintained in the bioreactor only limited settling will occur, making mixing unnecessary. Other options involve the use of separate anoxic and aerobic zones with recirculation of mixed liquor from the aerobic to the anoxic zone, just as in the modified Ludzak-Ettinger process, as shown in Figures 1.18b and 1.18c. Solids retention times and suspended solids concentrations in A/AD systems are similar to those in CAD systems. Furthermore, the facilities required for A/AD systems are similar to those used with CAD systems, with the exception of the addition of mixing equipment where necessary for anoxic operation.

Because of the need to recirculate nitrified mixed liquor for denitrification, as shown in Figure 1.18b and c, complete denitrification will not generally be achieved in continuous feed A/AD processes. Consequently, the reductions in alkalinity destruction and oxygen utilization will not be as great as suggested by Equation 13.9. This means that chemical addition for pH control may still be required for locations with very low alkalinity water. However, nearly complete denitrification can be achieved with intermittent A/AD (Figure 1.18a). In either case the implementation of A/AD is still worthwhile because of the savings involved.

Research has demonstrated the practicality of the A/AD process. It has also indicated that solids destruction rates under anoxic conditions are less than the rates under aerobic conditions. This phenomenon can be explained adequately using ASM No. 1 with its reduced rate of hydrolysis of particulate substrate under anoxic conditions. However, application of this model also indicates that the overall stabilization of organic matter in the A/AD process will be the same as achieved in a CAD system of equal SRT. Although a buildup of particulate substrate occurs in the anoxic zone of an A/AD process, it is rapidly oxidized in the subsequent aerobic zone.[17,20] The rate of degradation of active biomass to particulate substrate, which is rate limiting, is the same in the anoxic and aerobic zones.

13.1.2.3 Autothermal Thermophilic Aerobic Digestion

As discussed in Section 11.2.7, heat is released when organic matter is oxidized, and Table 13.3 summarizes the amount associated with biomass destruction.[43] Autothermal thermophilic aerobic digestion (ATAD) takes advantage of that heat to elevate the temperature of an aerobic digester. Many, but not all, CAD systems do not experience a significant temperature increase because of the large mass of water flowing through them, the conductive heat loss through the walls of the bioreactor, and the evaporative heat loss associated with operation of the oxygen transfer system.

TABLE 13.3
Energy Released by Oxidation of Biomass

Units	COD Basis	VSS Basis
kJ/kg	13,200	18,800
kcal/kg	3500	5000

In ATAD elevated temperatures are achieved by thickening the feed solids to a concentration of 40,000–60,000 mg/L to minimize the mass of water that must be heated, covering and insulating the bioreactor to minimize conductive heat losses, and using a high efficiency oxygen transfer device to minimize evaporative heat losses. As a result, it is practical to achieve bioreactor temperatures in the thermophilic range (45 to 65°C) without external heating. Smaller bioreactors may be used for ATAD because of the smaller feed volumes resulting from the thickening of the feed solids and the higher digestion rates associated with the elevated temperature. The elevated temperatures also accelerate the inactivation rate of pathogens,[8,39,43] as suggested by Table 13.2.

Figure 1.19 presents a schematic of ATAD. Two tanks in series are often used; significant digestion and heating occur in the first tank, with further digestion and heating to a temperature of about 55°C in the second tank. Feeding is often intermittent, with removal of digested solids from the second tank, transfer of digesting solids from the first to the second tank, and addition of feed solids to the first tank. This promotes temperature elevation and minimizes short-circuiting of feed solids to the digested solids, thereby giving better pathogen inactivation.

The bioreactors may be constructed of steel or concrete, but covers and insulation are generally required to control conductive heat losses. Proprietary oxygen transfer devices achieve the necessary oxygen transfer while minimizing associated heat losses. Bioreactor geometry is typically constrained by the requirements of these devices. High purity oxygen may also be used for oxygen transfer, resulting in even less heat loss.[27,39] Foaming can be severe because of the high feed solids concentrations and high temperatures. Consequently, specialized foam control devices are provided with many designs. Nitrification of the released ammonia-N typically does not occur because the elevated temperature minimizes the growth of nitrifying bacteria, which are mesophilic. Thus, oxygen requirements are reduced compared to CAD and pH depression is generally not a problem. In fact, the destruction of biomass in the absence of nitrification produces alkalinity, which results in pH values in the 7.5 to 8.0 range. This increased pH, coupled with elevated ammonia-N concentrations, can result in increased inactivation of viruses.[38] Because of the accelerated destruction rates of VSS and pathogens, SRTs on the order of five to six days are often used.

Autoheating can also occur in CAD systems if the feed solids concentration exceeds 20,000 mg/L, the feed solids are sufficiently biodegradable, and a high efficiency oxygen transfer system is used.[19,26] In such cases the temperature in the digester will depend on the heat loss characteristics of the bioreactor and the oxygen transfer device, along with waste activated sludge and ambient temperatures. Significant variations in digester operating temperature will occur in locations with significant seasonal ambient temperature variations. Periodic heating beyond the mesophilic temperature range (40°C) can result in excessive foaming, similar to the ATAD process, which can disrupt the process. Volatile suspended solids destruction is also adversely affected because mesophilic organisms are inactivated within the 40 to 45°C range, as discussed in Section 11.2.7. Procedures for computing the heat balance necessary to estimate temperatures in aerobic digesters have been presented elsewhere.[31,44]

Another variation is the dual digestion process,[5,27,43] which uses a single-stage, high-rate ATAD system to heat solids for subsequent feeding to an anaerobic digester, as shown in Figure 13.6. Pure oxygen is typically used to provide oxygen to the ATAD unit, thereby minimizing heat losses from it. As a consequence, an HRT of about one day can be used to heat the solids to a temperature of 55 to 65°C. The short HRT minimizes the mass of biodegradable organic matter oxidized aerobically, thereby maximizing the mass fed to the downstream anaerobic digester. Oxygen feed is regulated to achieve the desired ATAD reactor temperature.[31] Significant solubilization of VSS occurs because of the high ATAD reactor loadings, and COD reduction, rather than VSS destruction, correlates best with heat generation. Excellent pathogen destruction occurs because of the elevated temperatures developed in the ATAD reactor.

A further variation is the staged aerobic digestion process with prethickening of feed solids for temperature control, as illustrated in Figure 13.7.[10] Feed can be continuous or intermittent. The upstream cells can be operated at constant volume, with the volume of the final cell varied

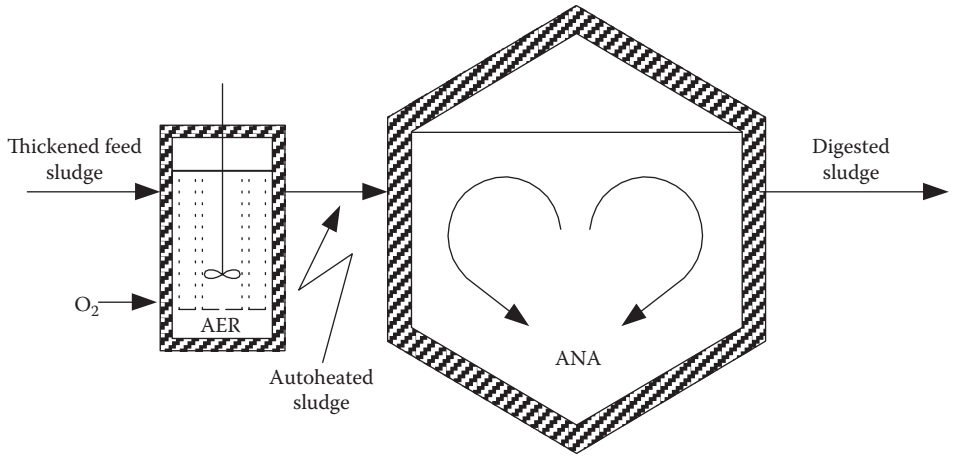

FIGURE 13.6 Dual digestion process.

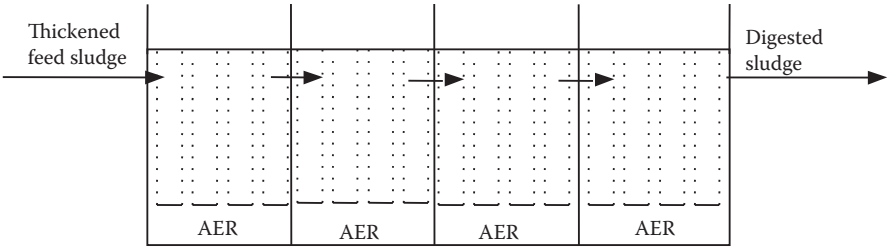

FIGURE 13.7 Staged aerobic digestion process.

to accommodate downstream processes (such as solids removal and dewatering). The sludge is prethickened to control temperature through autoheating but in the mesophilic range, generally between 20 and 35°C to accelerate VSS and pathogen destruction while avoiding excessive foaming associated with heating beyond 35°C. Staged operation allows increased VSS reduction and pathogen destruction. Aeration can also be cycled to achieve A/AD to control pH depression and reduce power requirements. Mixing may not be needed during anoxic operation as minimal settling occurs due to the high suspended solids concentrations maintained as a result of prethickening.

Table 13.4 summarizes the benefits and drawbacks of the various aerobic digestion process options. Conventional aerobic digestion is a demonstrated, proven process. It is mechanically simple and it is simple to operate. It is possible to incorporate both solids thickening and digestion into a single vessel, and the supernatant is of reasonably good quality. In contrast, the power costs for CAD are relatively high, and the rates of pathogen destruction are low. Relatively long SRTs are required, resulting in relatively large tank volumes. The pH will drop because of nitrification, and the solids that are produced will generally not dewater readily by mechanical means. The benefits and drawbacks of A/AD are similar to those of CAD, except that it provides pH control and affords a moderate reduction in power requirements. Staged A/AD with prethickening of feed solids for temperature control represents a further improvement. It allows a lower SRT because of the elevated temperatures that can be maintained: staging results in greater pathogen destruction and increased volatile solids reduction.

Autothermal aerobic digestion offers significantly reduced bioreactor volumes because of the smaller feed flow rates associated with the thickened solids that are used and the much lower SRTs that result from the higher operating temperature. Nitrification does not occur, so the pH is generally above neutral and, while the process oxygen requirement is reduced, the power requirement can be equal or greater due to the difficulty of transferring oxygen to the thickened solids. The

TABLE 13.4
Aerobic Digestion Process Comparison

Process	Benefits	Drawbacks
Conventional aerobic digestion (CAD)	• Demonstrated process, proven • Mechanically simple • Simple to operate • Both thickening and stabilization can be provided in one vessel • Supernatant is of reasonable quality	• High power cost • Low rates of pathogen inactivation • Long SRTs required • Relatively large reactor volumes • pH drops due to nitrification • Digested solids dewater poorly
Anoxic/aerobic digestion (A/AD)	• pH control provided • Mechanically simple • Simple to operate • Power requirements less than CAD • Both thickening and stabilization can be provided in one vessel • Supernatant is of reasonable quality	• Power costs still relatively high • Low rates of pathogen inactivation • Long SRTs required • Relatively large reactor volumes • Digested solids dewater poorly
Autothermal thermophilic aerobic digestion (ATAD)	• Low SRT • Small reactor volume • No pH drop • Excellent pathogen inactivation • Lower oxygen requirement than CAD or A/AD	• Mechanically more complex • Foaming • Newer process, less experience than with CAD • Power costs still relatively high • Requires separate thickening • Digested solids dewater quite poorly • Requires adequate biodegradable solids in feed
Staged anoxic/aerobic digestion (staged A/AD)	• Lower SRT than CAD or A/AD due to elevated temperature • Small reactor volume due to prethickening and lower SRT • pH control • Excellent pathogen inactivation • Lower power requirement than CAD • No supernatant (prethickening)	• Power costs still relatively high • Reactor volume larger than ATAD • Newer process, less experience than other aerobic digestion options • Requires separate thickening • Digested solids dewater poorly

higher temperature, elevated pH, and increased ammonia-N concentration result in greater rates of pathogen inactivation. In exchange for these benefits, ATAD is mechanically more complex and subject to severe foaming. The dewatering characteristics of the solids are poor due to the formation of colloidal solids that interfere with sludge conditioning.[32] It is a newer process, and engineers have significantly less experience with it than with CAD. Consequently, its performance and operational characteristics are less predictable. Separate thickening of the feed solids is required to achieve a sufficiently high concentration to achieve autothermal conditions. The successful operation also requires a sufficient fraction of biodegradable solids in the feed.

13.1.3 Typical Applications

Conventional aerobic digestion is widely used to stabilize the solids at small to medium sized wastewater treatment plants (less than 20,000 to 40,000 m^3/day). Many hundreds of examples exist. Conventional aerobic digestion is utilized in such circumstances because of its mechanical and process simplicity, and because solids thickening and stabilization can be incorporated into a single vessel. Aerobically digested solids dewater quite poorly by mechanical means, but waste solids from small wastewater treatment plants are often handled in a liquid, not dewatered, form, thereby eliminating the problem. Even though power costs are relatively high, capital costs are relatively low,

resulting in generally favorable economics. Pathogen destruction requirements may dictate the use of longer SRTs, thereby increasing capital and operating costs and reducing somewhat the use of CAD. Anaerobic digestion often offers cost advantages in larger wastewater treatment plants. Even though its capital costs are higher, power costs are lower, resulting in more favorable overall economics. Consequently, anaerobic digestion is frequently used in larger wastewater treatment plants.

Operational difficulties exist with CAD, particularly in colder climates where VSS reduction efficiencies are often low and freezing of equipment—and even of a portion of the bioreactor contents—can occur. Design approaches are available to address these difficulties, such as covering and heat tracing of equipment, and thickening of solids prior to digestion to allow some autoheating. Other operational difficulties exist because of poor understanding of the process and the factors that affect its performance. In spite of these difficulties, successful operation and acceptable performance can be obtained through the application of proper design and operational principles, and many successful case histories exist (e.g., see Pizarro[35]).

Anoxic/aerobic digestion is a relatively new option but it is being increasingly applied because it generally offers the benefits of CAD, along with minimal pH depression and reduced energy requirements. Existing CAD processes can easily be modified to operate in the A/AD manner by changing the operation of the oxygen transfer system to an intermittent mode. This can result in energy cost savings, along with improved control over bioreactor pH.

Increased concern over pathogen destruction and the desire for reduced bioreactor volumes is fueling interest in advanced digestion processes such as ATAD and staged A/AD with prethickening for temperature control. Autothermal thermophilic aerobic digestion has been used widely in Europe, as has the dual digestion process. However, interest in the ATAD process, either separately or as a component of the dual digestion process, is waning due to its complexity, odors caused by the difficulty of achieving sufficient oxygen transfer into the thick sludge necessary to achieve autothermal conditions, and poor sludge dewatering characteristics. In contrast, staged A/AD with prethickening for temperature control is receiving increased interest because it retains many of the desirable features of CAD and eliminates many of its drawbacks.

The benefits, drawbacks, and costs of aerobic digestion must be compared to other solids stabilization options and the optimal one selected for a particular application. It appears that the use of aerobic digestion will continue, particularly at the small to medium sized wastewater treatment plants where it has traditionally been used. Design of staged CAD and A/AD processes, with prethickening to achieve autoheating, is also likely, thereby providing the benefits of plug flow and operation at elevated temperature (perhaps still in the mesophilic range), while retaining the benefits of CAD and A/AD.[15]

13.2 FACTORS AFFECTING PERFORMANCE

13.2.1 Solids Retention Time and Temperature

Figure 5.10 illustrates the effect of SRT on the performance of CAD of the type shown in Figure 1.17b, as simulated with the simple model of Chapter 5. There it can be seen that the total solids concentration decreases as the SRT of the digester is increased, which is to be expected. Perhaps more importantly, however, the figure demonstrates that the buildup of biomass debris in the system limits the percentage of solids destruction that can be achieved. In this case about 53% solids destruction is achieved at an SRT of 1000 hrs or about 42 days, as shown in Figure 13.8. At that SRT, debris accounts for about 83% of the remaining solids, suggesting that little additional solids destruction will occur, even if the SRT is extended greatly. This can also be seen from the slope of the VSS destruction curve in Figure 13.8. This characteristic should be kept in mind when selecting the SRT for an aerobic digester. A point may quickly be reached at which further increases in SRT will have minimal effect, making expenditures for additional tank volume questionable. That point will depend on the nature of the influent solids, their biodegradability, and the temperature of the system.

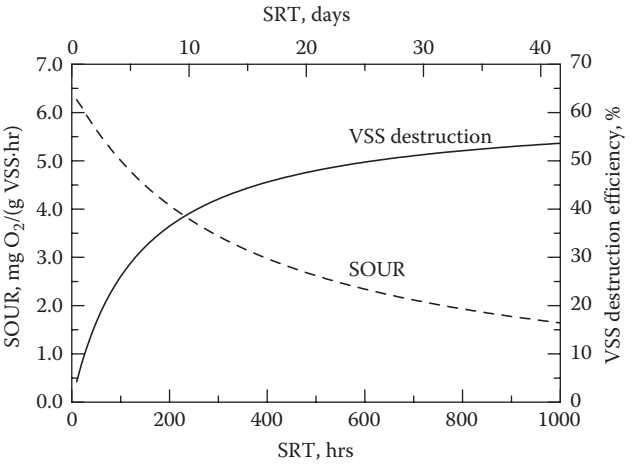

FIGURE 13.8 Effect of SRT on the percent VSS destruction and SOUR in a completely mixed aerobic digester. The curves are theoretical and were generated from the information in Figure 5.10.

Figure 5.10 also shows the effect of SRT on the oxygen requirement of the digester and that curve has been converted into the SOUR, as shown in Figure 13.8. Because the oxygen requirement curve in Figure 5.10 does not consider nitrification of the released ammonia-N, neither does the SOUR curve in Figure 13.8. Should the pH and other factors allow nitrifiers to grow, the SOUR would be 40% greater than shown. Like the VSS destruction curve, the SOUR curve quickly reaches a point where further increases in SRT have little effect. Just how rapidly that occurs also depends on the nature of the influent solids, their biodegradability, and the temperature of the system. This, too, must be considered when an SRT is being chosen. These effects of aerobic digester SRT on the VSS destruction and the SOUR of the solids have been confirmed by laboratory results.[33]

Figure 13.8 illustrates a very important point: sometimes it is possible to meet one criterion of stabilized solids without meeting the other. In this particular case 38% VSS destruction is achieved at a very reasonable SRT, but an SOUR of 1.0 mg O_2/(g VSS·hr) is not. This follows directly from the nature of the solids being digested, which were waste activated sludge from a completely mixed activated sludge system treating totally soluble substrate. As a consequence, it contained little nonbiodegradable VSS, which allowed a high degree of VSS destruction, but also made the SOUR high. Had the solids been high in nonbiodegradable VSS, just the opposite effect would have been seen; that is, it would have been possible to reach an SOUR of less than 1.0 mg O_2/(g VSS·hr) at a relatively short SRT, but it would have been difficult to achieve 38% destruction of the VSS. Consequently, it can be seen that what constitutes stable solids depends to a large degree on their source and characteristics. This point is discussed further in Section 13.2.4 because it cannot be overemphasized.

Temperature has an important impact on the destruction of VSS during aerobic digestion because of its effect on the rate coefficient, b_{MV}. That effect is typically expressed with Equation 3.99, using a θ value of 1.029 for mesophilic digestion. Less certainty is associated with the temperature coefficient for thermophilic conditions. Nevertheless, no matter what the temperature range, lower temperatures mean that longer SRTs are required, just as with activated sludge systems.

Simulations such as those presented above can be used for design when sufficient information is available. Otherwise, performance correlations from the literature must be used. Experience indicates that temperature exerts such an important effect that both temperature and SRT must be considered together in evaluating the performance of an aerobic digester.[15,18] Consequently, Koers and Mavinic[21,29] suggested that the VSS destruction efficiency be plotted as a function of the bioreactor temperature times the SRT. Figure 13.9 provides an example of such a plot that is widely used in aerobic digester design.[29,41] It suggests that the operation of a mesophilic digester at

a temperature-SRT product of 400 to 500°C-days will result in substantially complete VSS destruction. Further increases in the temperature-SRT product may result in additional VSS destruction, but the increase will be relatively small. Of course, as discussed above, the actual VSS destruction efficiency obtained will depend on the biodegradability of the solids being digested.

The effect of SRT and temperature on the SOUR is illustrated in Figure 13.10.[2] An SRT in excess of 60 days may be sufficient at temperatures above 10°C to reduce the SOUR below 1 mg O_2/(g VSS·hr), whereas at temperatures below 10°C, SRTs in excess of 100 days may be required.

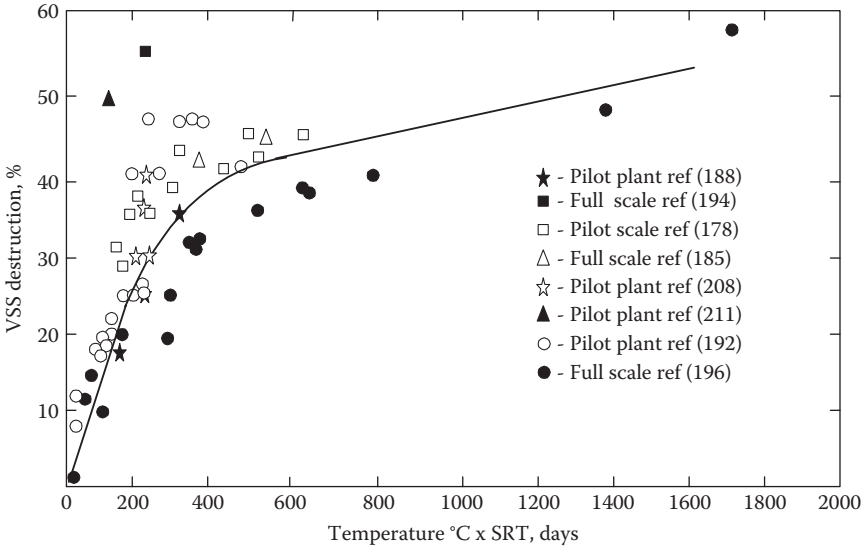

FIGURE 13.9 Effect of the temperature-SRT product on the VSS destruction efficiency during aerobic digestion. Reference numbers refer to original source. (From U.S. Environmental Protection Agency, *Process Design Manual for Sludge Treatment and Disposal*, EPA/625/1-79/011, U.S. Environmental Protection Agency, Cincinnati, OH, 1979.)

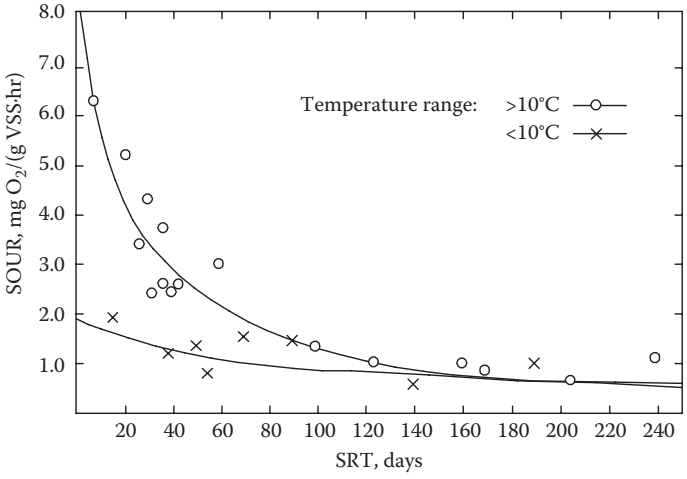

FIGURE 13.10 Effect of temperature and aerobic digester SRT on the SOUR of digested solids. (From Ahlberg, N. R., and Boyko, B. I., Evaluation and design of aerobic digesters. *Journal, Water Pollution Control Federation,* 44:634–43, 1972. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

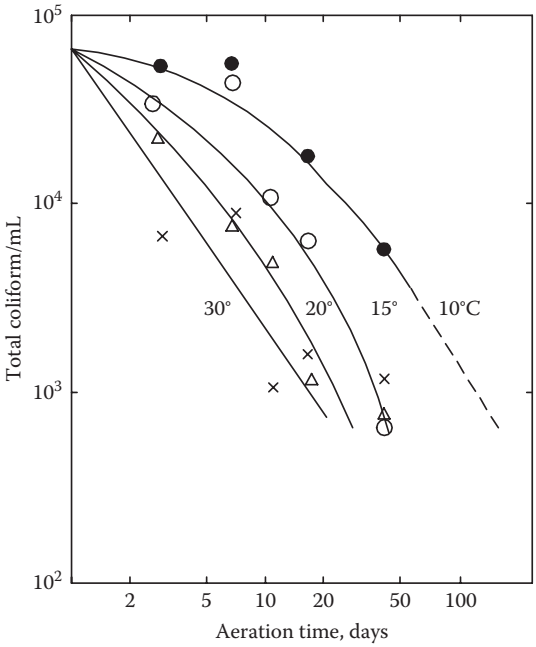

FIGURE 13.11 Effect of temperature on the inactivation of total coliforms during aerobic digestion in batch reactors. (From Novak, J. T., Eichelberger, M. P., Banerji, S. K., and Yaun, J., Stabilization of sludge from an oxidation ditch. *Journal, Water Pollution Control Federation,* 56:950–54, 1984. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

These operating conditions correspond to temperature-SRT products ranging from 600 to well over 1000°C-days. This again illustrates that it sometimes may be difficult to meet both criteria of stable solids. Similar results have been observed by others.[21,29,33]

Pathogen destruction occurs in aerobic digesters as a result of their natural die-off.[6,23,26,33] Consequently, digestion per se does not affect pathogen destruction; rather, it is the environment created in the aerobic digester and the retention of the pathogens in that environment that results in their destruction. Aeration is necessary primarily to maintain aerobic conditions and avoid nuisance conditions. Figure 13.11[33] illustrates that rates of pathogen die-off are significantly affected by digester temperature. Relatively long aerobic digester SRTs are required for operation at lower temperatures (10°C), while much shorter SRTs are required at higher temperatures, particularly when they are in the thermophilic range. This observation provided one justification for the development of ATAD.

Since the inactivation of pathogens during aerobic digestion is a function of both temperature and SRT, regulatory approaches have specified the required digester SRT to achieve compliance with pathogen control requirements as a function of digester temperature, as illustrated in Table 13.2.[24,40,42] The available data indicate that operating temperature has a stronger effect on pathogen destruction than the digester SRT. Consequently, the requirements for pathogen destruction cannot be expressed as a simple temperature-SRT product. The requirements listed in typical regulatory requirements are for completely mixed digesters. Operation of digesters in a batch or tanks-in-series mode will significantly improve pathogen destruction, allowing a significantly reduced SRT to be used at any particular temperature.[10,15]

13.2.2 pH

As discussed in Section 12.3.1, a residual alkalinity of around 50 mg/L as $CaCO_3$ is required to maintain a stable pH near neutrality. However, alkalinity is destroyed if the ammonia-N released

during aerobic digestion is nitrified. As shown in Equation 13.8, the amount is one mole of HCO_3^- per mole of biomass destroyed. This corresponds to 0.44 g of alkalinity (expressed as $CaCO_3$) per g VSS destroyed. Consequently, as illustrated in Figure 13.5, unless pH control is practiced, the pH will decrease during CAD if the released ammonia-N is nitrified. Without pH control, pH values in operating aerobic digesters routinely drop to 5 or below. Although the destruction of biodegradable organic matter proceeds at these low pH values, the rate is reduced compared to the rate at pH 7.[22] More rapid digestion has been demonstrated through the addition of lime and other pH control chemicals or application of A/AD to recover alkalinity to maintain the pH near neutrality.[3,4,10]

13.2.3 Mixing

Adequate mixing energy must be provided in aerobic digesters to maintain solids in suspension. Solids settlement will reduce the effective volume of the bioreactor and result in anaerobic conditions in the settled solids. The provision of adequate mixing can be a challenge because of the high suspended solids concentrations commonly maintained. Design and operational manuals typically recommend an air input rate to aerobic digesters of 20 to 40 $m^3/(min \cdot 1000\ m^3)$ when diffused air systems are used.[46,47] Reynolds[37] has provided the following equation to calculate the volumetric power input (Π, expressed in units of kW/1000 m^3 of bioreactor volume) required to maintain solids in suspension during aerobic digestion:

$$\Pi = 0.935\ \mu_{wc}^{0.3} X_{M,T}^{0.298}, \tag{13.10}$$

where μ_{wc} is the viscosity of water in centipoise and $X_{M,T}$ is the MLSS concentration in the digester, expressed as mg/L as TSS. At 25°C and an MLSS concentration of 10,000 mg/L, Equation 13.10 gives a volumetric power input of 13.8 kW/1000 m^3, which is near the minimum recommended for mechanical aeration devices in activated sludge systems, as discussed in Section 11.2.5. While Equation 13.10 can be used to estimate required mixing energy inputs for CAD applications, care should be exercised when applying it to the high suspended solids concentrations used in ATAD. For example, mixing energy inputs in the range of 85 to 105 kW/1000 m^3 are typically used with ATAD,[43] whereas Equation 13.10 would give lower values.

Evidence suggests the difficulty in maintaining aerobic conditions throughout an aerobic digester. For example, several researchers have indicated that the specific decay rate of a variety of waste solids declined as the suspended solids concentration was increased.[11,16,22,37] Since there is no biological basis for such an observation, the most likely explanation is the increasing difficulty of transferring oxygen and maintaining aerobic conditions throughout the digesting solids particles. Furthermore, a loss of nitrogen that can be attributed to denitrification has been observed in aerobic digesters, even though measurable dissolved oxygen concentrations were maintained on a continuous basis.[28,30] These observations suggest that it may not be possible to maintain fully aerobic conditions in many full-scale aerobic digesters. In fact, experience suggests that the oxygen transfer capacity of many devices declines noticeably when the solids concentration exceeds 20,000 to 25,000 mg/L as TSS. Thus, the TSS concentration in the digester should generally be maintained below this value to avoid inadequate oxygen transfer and DO concentrations.

13.2.4 Solids Type

Waste solids being sent to aerobic digestion vary in the proportions of their biodegradable and nonbiodegradable components. For example, if waste activated sludge is being digested, one factor that can influence this is the SRT of the activated sludge system from which it came. This can be seen in Figure 5.7, where the active fraction, which is numerically equivalent to the biodegradable fraction for the situation simulated, decreases as the activated sludge SRT is increased. It can also be seen in Figure 13.4, where experimental determinations of the nonbiodegradable fraction are shown as

a function of the SRT of the activated sludge system from which the solids came. The implication of this is that it will be more difficult to achieve a given percentage of VSS destruction in solids from an activated sludge system with a long SRT, simply because a smaller fraction of the solids is biodegradable. The nature of the influent to an activated sludge system will also influence the biodegradability of the solids wasted from it. This can also be seen in Figure 5.7, where two cases are compared, one with and one without inert (i.e., nonbiodegradable) suspended solids in the influent. Clearly, the mixed liquor suspended solids (MLSS) in the system receiving inert solids has a lower active (i.e., biodegradable) fraction, which will make it more difficult to achieve a given percentage of VSS destruction. For other solids, the biodegradable content will depend on their source. On the order of 60 to 80% of the VSS in domestic primary solids will be biodegradable,[22] but no such generalization can be made for solids produced during the treatment of industrial wastewaters. Their biodegradable content must be measured.

Waste solids also differ with respect to the nature of their biodegradable component, depending on their source. This is important because the nature of that biodegradable component determines the rate at which it is stabilized. Theoretically the rate of degradation of waste biomass will be nearly independent of the solids source.[25,33,45] The data presented in Figure 13.9 suggest that this may be true, since they came from several sources. Likewise, the data in Figure 13.3 show that the rate coefficient for destruction of waste activated sludge is independent of the SRT of the system from which it came.[36] In general, the rate of destruction of domestic primary solids will be lower than the rate of destruction of waste biomass.[22] This occurs because of the need to first convert the particulate organic matter contained in the primary solids to active biomass that is subsequently oxidized.

13.2.5 Bioreactor Configuration

Because the destruction of biodegradable organic matter can be characterized as a first-order reaction, the efficiency of an aerobic digester can be improved by configuring it as a series of CSTRs, provided that there is no solids recycle around the reactor chain. This provides the basis for the staged aerobic digestion process, as discussed above.[10] The impacts of bioreactor staging on VSS destruction efficiency can be estimated with

$$E_{XMV} = 100\left[\frac{X_{M,VO} - X_{M,V,nO}}{X_{M,V,O}}\right]\left[1 - \frac{1}{\left(1 + \frac{b_{MV} \cdot \Theta_c}{N}\right)^N}\right]. \tag{13.11}$$

The term Θ_c is the total SRT for the bioreactor system and N is the equivalent number of equal sized tanks in series. It is analogous to Equation 13.4, which applies only to a single CSTR. If the tanks are not equal in size, the performance of the bioreactor system can be determined by sequential application of Equation 13.4. In a similar fashion, an expression equivalent to Equation 13.6 can be developed to predict the SOUR of the treated solids from a tanks-in-series system:

$$SOUR = 1000\left[\frac{b_{MV} \cdot i_{O/XM,V}\left(X_{M,VO} - X_{M,V,nO}\right)}{\left(X_{M,VO} - X_{M,V,nO}\right) + X_{M,V,nO}\left(1 + \frac{b_{MV} \cdot \Theta_c}{N}\right)^N}\right]. \tag{13.12}$$

The impact of the number of CSTRs in series on the performance of CAD is illustrated in Figure 13.12. This figure was developed for the situation depicted in Figures 5.10 and 13.8, except

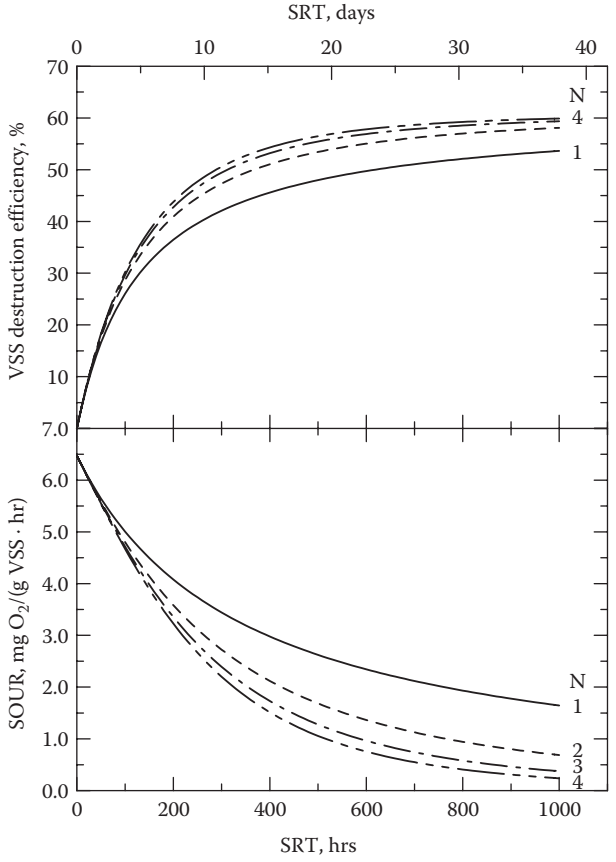

FIGURE 13.12 Effects of SRT and number of equivalent CSTRs in series on the percentage VSS reduction and SOUR in a conventional aerobic digester. The curves are theoretical and were generated from Equations 13.11 and 13.12 using the same kinetics and solids characteristics used to generate Figures 5.10 and 13.8.

that the bioreactor was divided into one to four equal sized compartments. Performance is significantly improved by configuring the bioreactor as two CSTRs in series rather than as a single CSTR, allowing a given degree of stabilization to be achieved at a lower SRT. However, compared to the two CSTR systems, less improvement is obtained by going to three and four CSTR systems. Similarly, the pathogen destruction efficiency of an aerobic digester is improved by configuring it as a series of CSTRs. This has been clearly demonstrated for the ATAD process, where two CSTRs in series are typically used, and for the staged aerobic digestion process, with significant improvements in bioreactor performance.

A CSTRs-in-series configuration can be obtained in several ways. Consider, for example, the intermittent feed CAD system illustrated in Figure 1.17a. If more than one bioreactor is available, they can be operated in an alternating fashion in which one is fed and decanted for a period of time while another is off-line to allow digestion and pathogen inactivation to proceed. After digested solids are removed from the second bioreactor, feed is then directed to it while the first is taken off-line for further reaction. The advantages of CSTRs in series can be achieved with the continuous feed process by splitting the bioreactor into two compartments and directing the underflow from the settler to the second one. This approach is necessary because, as discussed in Section 7.2.2, the recycle of solids around the entire system would make it completely mixed with respect to biomass, which would make it behave like a single CSTR. Consequently, care must be taken to ensure that plug-flow type conditions are truly achieved with regard to the flow of solids through the bioreactor. Because

the solids concentration in the first compartment will be less than that in the second, the volume of the first compartment should be larger to fully gain the benefits of the tanks-in-series configuration. Finally, a staged system without solids recycle, such as illustrated in Figure 13.7, can be used.

13.3 PROCESS DESIGN

13.3.1 Overview

The design of an aerobic digester can be accomplished by an application of the principles presented in this and previous chapters, as enumerated in Table 13.5. Several procedures can be used to perform the necessary process calculations. Among them are those based on empirical correlations, those using batch data in simple models, and those using the simplified model of Chapter 5 or IWA ASM No. 1. The basic approaches are similar to those used to design activated sludge and BNR systems. Consequently, all of the procedures presented in those chapters will not be repeated here. Rather, just the unique points will be emphasized.

13.3.2 Design from Empirical Correlations

Empirical correlations such as Figure 13.9 can be used to select a design temperature-SRT product. Then an estimation of the digester operating temperature allows direct calculation of the required

TABLE 13.5
Summary of Aerobic Digestion Process Design Procedure

1. Select the process option; that is, whether the system will be CAD, A/AD, or ATAD. In addition, a decision must be made as to whether operation will be intermittent or continuous.
2. Selection of the bioreactor configuration; that is, whether it will be a single CSTR or a series of CSTRs.
3. Select the bioreactor feed solids concentration and physical reactor configuration. Both of these factors affect the heat balance for the bioreactor and determine whether significant autoheating will occur. If significant autoheating is expected, a heat balance should be performed to estimate the bioreactor operating temperature. Procedures for performing a heat balance are available elsewhere.[a] If ATAD is to be designed, ensure that the temperature will be elevated into the thermophilic range. In all cases, the temperature should be either in the mesophilic or thermophilic range—operation in the intermediate temperature zone between mesophilic and thermophilic conditions must be avoided due to reduced sludge stabilization efficiency and excessive foaming.
4. Select the bioreactor SRT. This is done on the basis of the desired percentage of VSS destruction or SOUR to be achieved, using Equations like 13.4, 13.6, 13.11, or 13.12. Alternatively, graphical information like that in Figure 13.9 can be used.
5. Determine the required bioreactor volume. This determination is based on the selected SRT, influent flow rate, and desired bioreactor solids concentration.
6. Calculate the oxygen requirement. The power input required to meet the oxygen requirement can then be calculated using the procedures presented in Section 11.2.5.
7. Determine the power input required to achieve adequate mixing and to maintain solids in suspension. This can be done with Equation 13.10 or with the procedures presented in Section 11.2.5. Just as in activated sludge design, the power required for mixing must be compared to the power required for oxygen transfer and the larger of the two provided.
8. Evaluate the need for supplemental alkalinity for pH control. As discussed in Section 13.1.2, nitrification of released ammonia-N will result in destruction of alkalinity if CAD is used. If the amount of alkalinity available is insufficient, the pH in the digester will drop, reducing the rate at which digestion occurs. Consequently, the amount of alkalinity available should be compared to the amount of alkalinity likely to be destroyed to determine the need for supplementation.

[a] Messenger, J. R., de Villiers, H. A., and Ekama, G. A., Oxygen utilization rate as a control parameter for the aerobic stage in dual digestion. *Water Science and Technology*, 22(12):217–27, 1990; Vismara, R., A model for autothermic aerobic digestion—Effects of scale depending on aeration efficiency and sludge concentration. *Water Research*, 19:441–47, 1985.

SRT. If the solids are to be thickened prior to digestion, the SRT is equal to the HRT, allowing the bioreactor volume to be calculated from the influent solids flow rate using the definition of HRT, as given by Equation 4.15. If solids are to be thickened during digestion, in either the intermittent or the continuous process, then the VSS concentration in the digester is given by

$$X_{M,V} = \left(\frac{\Theta_c}{\tau}\right)\left[X_{M,VO}\left(\frac{100 - E_{XMV}}{100}\right)\right]. \tag{13.13}$$

By substituting Equation 4.15 for the HRT (τ), the required digester volume may be calculated as

$$V = \left(\frac{F \cdot \Theta_c}{X_{M,V}}\right)\left[X_{M,VO}\left(\frac{100 - E_{XMV}}{100}\right)\right]. \tag{13.14}$$

In determining the degree of thickening that can be accomplished during digestion it is often desirable to know the suspended solids concentration on a TSS basis. The TSS is just the sum of the VSS and the FSS, $X_{M,F}$:

$$X_{M,T} = X_{M,V} + X_{M,F}. \tag{13.15}$$

Although some FSS are solubilized during digestion, as discussed in Section 13.1.1, many designers assume that they remain unchanged, thereby giving a conservative estimate of the TSS concentration. In that case, the FSS act like inert suspended solids, so that:

$$X_{M,F} = \left(\frac{\Theta_c}{\tau}\right)X_{M,FO}, \tag{13.16}$$

where $X_{M,FO}$ is the FSS concentration in the digester feed. Thus, when solids are expressed on a TSS basis, the reactor volume is given by

$$V = \left(\frac{F \cdot \Theta_c}{X_{M,T}}\right)\left[X_{M,FO} + X_{M,VO}\left(\frac{100 - E_{XMV}}{100}\right)\right]. \tag{13.17}$$

The oxygen requirement, RO, can be calculated on the basis of the estimated VSS destruction efficiency:

$$RO = F \cdot i_{O/XM,V} \cdot X_{M,VO}\left(\frac{E_{XMV}}{100}\right). \tag{13.18}$$

The value of $i_{O,XM,V}$ depends on whether nitrification is occurring, as expressed in Table 13.1. Once RO is known, the air flow rate or power requirement for oxygen transfer can be estimated as was done in Section 11.2.5. That can then be compared to the power required for mixing as given by Equation 13.10 or other appropriate information.

Alkalinity destruction can be estimated by recognizing that approximately 0.44 g of alkalinity (expressed as $CaCO_3$) will be used for each g VSS destroyed if the released ammonia-N is nitrified. This can be compared with the available alkalinity to determine whether pH control will be needed. As discussed previously, a residual alkalinity of around 50 mg/L as $CaCO_3$ is required to maintain a stable pH near neutrality. Aeration can be cycled as one method to recover alkalinity and reduce net alkalinity consumption.

The following example illustrates the use of empirical correlations to accomplish a preliminary design.

Example 13.3.2.1

A preliminary design is needed for a conventional aerobic digester using intermittent feeding with in-basin thickening for stabilization of waste biomass from an aerobic biological wastewater treatment system. At least 40% VSS destruction is desired year-round. The waste biomass concentration is 8000 mg/L as TSS and the flow rate is 500 m^3/day. The solids are 75% volatile, but little else is known about their characteristics. The alkalinity of the carriage water is 150 mg/L as $CaCO_3$. The lowest operating temperature of the digester is expected to be 10°C based on similar installations in the region. Since those installations can achieve a TSS concentration of 20,000 in the digester, that will be used as a target in the design. Assume that an oxygen transfer device with a transfer efficiency of 1.2 kg O_2/(kW·hr) will be used.

a. What SRT might be appropriate for this application?
   Figure 13.9 indicates that 40% stabilization can be achieved with a temperature-SRT product of 600°C-days. Since the lowest expected temperature is 10°C, the SRT should be 60 days.
b. What volume should the digester have?
   The digester volume can be determined with Equation 13.17. To use that equation the influent VSS and FSS concentrations must be known. Since the waste solids are 75% volatile:

$$X_{M,VO} = (0.75)(8000) = 6000 \text{ mg/L}.$$

   Therefore,

$$X_{M,FO} = 8000 - 6000 = 2000 \text{ mg/L}.$$

   Recognition of the fact that 1.0 mg/L = 1.0 g/m^3 and substitution of these values into Equation 13.17 gives:

$$V = \left(\frac{500 \cdot 60}{20{,}000}\right)\left[2000 + 6000\left(\frac{100-40}{100}\right)\right] = 8400 \text{ m}^3.$$

c. What is the oxygen requirement?
   The oxygen requirement can be estimated with Equation 13.18 after an appropriate value has been chosen for $i_{O/XM,V}$. Provided that pH control is practiced, nitrification will occur even in the winter because of the long SRT. Consequently, from Table 13.1, the appropriate value for $i_{O/XM,V}$ is 1.98 mg O_2/mg VSS destroyed. Consequently, the oxygen requirement is

$$RO = (500)(1.98)(6000)\left(\frac{40}{100}\right) = 2{,}376{,}000 \text{ g/day} = 99 \text{ kg/hr}.$$

d. What is the power requirement for oxygen transfer?
   Assuming the use of an oxygen transfer device with a transfer efficiency of 1.2 kg O_2/(kW·hr), the required power input is

$$P = \frac{99}{1.2} = 82.5 \text{ kW}.$$

e. What is the power requirement for mixing?
   The volumetric power input for mixing can be calculated with Equation 13.10. At 10°C, $\mu_{wc} = 1.310$ cp. Consequently:

$$\Pi = (0.935)(1.31^{0.3})(20{,}000^{0.298}) = 19.4 \text{ kW}/1000 \text{ m}^3.$$

Since the volume is 8400 m^3, the power requirement for mixing is (19.4)(8.4) = 163 kW. Since this exceeds the power required for oxygen transfer, the larger power for mixing must be provided.

f. Will pH control be required to maintain a neutral pH?
The destruction rate of VSS in the digester is (500)(6000)(0.40) = 1,200,000 g/day. Since 0.44 g of alkalinity is destroyed for each gram of VSS destroyed when nitrification occurs, the mass rate of alkalinity destruction will be 528,000 g/day. The waste solids contain 150 mg/L of alkalinity, but 50 mg/L must be retained as a residual. Therefore the mass of alkalinity available per day is (500)(100) = 50,000 g/day. This is inadequate, so pH control must be used, either through the addition of supplemental alkalinity or through cyclic aeration.

13.3.3 Design from Batch Data

An alternative design approach involves the use of a batch reactor to characterize the solids to be digested, and it is commonly used (e.g., see Reece et al.[36] and Reynolds[37]). A sample of raw solids is placed in a well-mixed vessel, aerobic conditions are maintained, and pH is controlled at seven. Data are then collected and used in Equation 13.7 to determine the decay coefficient b_{MV} and the nonbiodegradable VSS concentration, $X_{M,V,nO}$. To obtain the most accurate assessment of b_{MV}, data should be collected on the oxygen uptake rate (OUR) and the TSS and VSS concentrations over time. Care should be exercised to add distilled water to replace any evaporation losses and to scrape any solids that accumulate on the inside walls of the reactor back into the liquid to avoid changes in suspended solids concentrations not attributable to biological reaction. The batch reactor should be operated long enough to ensure that the majority of the biodegradable organic matter is destroyed, thereby allowing an accurate estimate of the nonbiodegradable VSS concentration to be made. This requires that the batch digestion time be greater than five times the reciprocal of the decay coefficient, b_{MV}. To determine b_{MV} the OUR data are analyzed in the same manner as described in Section 9.3.2 for the determination of the decay coefficient, b_H. Once b_{MV} is known, it can be substituted into Equation 13.7 and the VSS data can be analyzed according to that equation for estimation of the nonbiodegradable VSS concentration. If OUR data cannot be collected, then b_{MV} and $X_{M,V,nO}$ can be determined simultaneously by fitting Equation 13.7 to data on the VSS concentration over time. However, it should be recognized that an inaccurate estimate of the nonbiodegradable VSS concentration will result in an inaccurate estimate of b_{MV}.

While measured parameter values are frequently used for sizing aerobic digesters, care must be exercised for the following reasons:

- Significant variation in the measured decay coefficient and the nonbiodegradable fraction of the waste solids may occur with time. Consequently, it is recommended that several batch digestion tests be conducted over time. Then a statistical approach can be used to select the design values for the decay coefficient and the nonbiodegradable proportion of the waste solids.[1]
- Conditions in the batch tests can differ significantly from those anticipated for the full-scale digester. Factors that may differ include pH, temperature, and suspended solids concentrations that can lead to oxygen transfer limitations, as discussed previously. While the data from lab-scale batch reactors can be used successfully to predict the performance of full-scale continuous flow bioreactors, the batch results may significantly underestimate full-scale performance.[13] This can be due to acclimation and/or to the maintenance of more favorable conditions in the full-scale bioreactor.
- If the waste solids to be studied come from a nonnitrifying activated sludge system, they may give erroneous results when the OUR technique is used to determine the decay coefficient. When the waste solids come from a fully nitrifying activated sludge system, ammonia-N will be nitrified as it is released during the batch digestion test.[45] Consequently, the oxygen demand per unit of VSS destroyed will remain reasonably constant during the

test and the decrease in OUR will be proportional to the destruction of biodegradable organic matter as assumed. On the other hand, if the waste solids are from a nonnitrifying system, the released ammonia-N will not be nitrified until a sufficient population of nitrifiers has developed.[28,30] In this situation, the oxygen demand per unit of VSS destroyed will not be constant and the change in OUR will not be proportional to the destruction of biodegradable organic matter. Ammonia-N concentrations should be monitored during the batch digestion test to detect whether consistent nitrification is occurring. If it is not, then either nitrification should be inhibited during the OUR measurements or nitrifiers should be added so that complete nitrification of ammonia-N occurs as it is released.

After the values of b_{MV} and $X_{M,V,nO}$ have been determined, the SRT required to achieve either a desired percentage of VSS destruction or a desired SOUR can be calculated. The SRT required for a given percentage of VSS destruction in a single-stage digester can be obtained with a rearranged form of Equation 13.4:

$$\Theta_c = \frac{(E_{XMV}/100)X_{M,VO}}{b_{MV}[X_{M,VO} - X_{M,V,nO} - (E_{XMV}/100)X_{M,VO}]}. \tag{13.19}$$

Similarly, the SRT required to achieve a given SOUR in a single-stage digester can be obtained with a rearranged form of Equation 13.6:

$$\Theta_c = \frac{1000 \cdot i_{O/XM,V}(X_{M,VO} - X_{M,V,nO})}{SOUR \cdot X_{M,V,nO}} - \frac{X_{M,VO}}{b_{MV} \cdot X_{M,V,nO}}. \tag{13.20}$$

Consideration must be given to the SRT required to meet each criterion when deciding on the design SRT. If both are reasonable, then the larger of the two should be used. On the other hand, if one is inordinately high, then the SRT should be selected from consideration of both criteria.

Once the required SRT has been determined, the remainder of the design proceeds in exactly the same manner as described in Section 13.3.2.

Example 13.3.3.1

Batch tests with a waste activated sludge have revealed that the decay coefficient for its aerobic decomposition has a value of 0.216 day^{-1} at 20°C. The solids to be digested are wasted from the bottom of the final settler at a concentration of 12,000 mg/L as VSS. Solids at that concentration were used to run the batch tests, revealing that the nonbiodegradable VSS concentration was 5400 mg/L. If the lowest temperature expected in the digester is 12°C, what SRT would be required to achieve at least 38% VSS destruction and reduce the SOUR to 1.0 mg O_2/(g VSS·hr) or less? Is it realistic to meet both criteria? Assume that the temperature coefficient for b_{MV} has a value of 1.029.

a. What is the value of the decay coefficient at 12°C?
The decay coefficient can be corrected for temperature with Equation 3.99:

$$b_{MV,12} = 0.216\ (1.029)^{12-20} = 0.172\ day^{-1}.$$

b. What SRT is required to meet the percentage of VSS destruction criterion at 12°C?
This may be determined with Equation 13.19:

$$\Theta_c = \frac{(38/100)(12{,}000)}{0.172[12{,}000 - 5400 - (38/100)(12{,}000)]} = 13.0 \text{ days}.$$

c. What SRT is required to meet the SOUR criterion at 12°C?
This may be determined with Equation 13.20. Note that SOUR and b_{MV} must have consistent time units. To be conservative, assume that nitrification occurs, making the value of $i_{O/XM,V}$ equal to 1.98.

$$\Theta_c = \frac{(1000)(1.98)(12{,}000 - 5400)}{(1.0)(24)(5400)} - \frac{12{,}000}{(0.172)(5400)} = 87.9 \text{ days.}$$

d. Is it realistic to meet both criteria?
An SRT of 88 days is very long and may not be realistic unless the waste activated sludge flow is very small. The reason that it is easy to meet the VSS destruction criterion but not the SOUR criterion is that the waste activated sludge has a fairly high percentage of biodegradable solids. The shorter of the two SRTs can be selected as either of these criteria indicates a stabilized sludge.

13.3.4 Design by Simulation

If an activated sludge system is being designed with the simple model of Chapter 5 or with IWA ASM No.1, the output may be used directly in the design of an aerobic digester by simulation. In either case, the characteristics of the waste solids are first determined from the simulations used in the activated sludge design. For the simple model of Chapter 5 the components will include active biomass, biomass debris, and inert organic matter. Only the active biomass will be degraded in the digester, leading to additional debris. The concentrations of active biomass and debris in a completely mixed digester can be calculated with Equations 5.76 and 5.77, respectively, while the MLSS concentration can be calculated with Equation 5.78. The values of the influent biomass ($X_{B,H,TO}$) and debris ($X_{D,TO}$) concentrations in those equations are the concentrations in the waste activated sludge entering the digester. If the waste solids also contain inert organic matter that originated in the influent to the activated sludge system, its concentration should be added within the bracket of Equation 5.77 to reflect its presence in the digester.

When using ASM No. 1, the output from the activated sludge simulation will provide the concentrations of active biomass, biomass debris, slowly biodegradable substrate, and inert organic matter in the waste activated sludge. These components can be used directly as inputs into a model for the aerobic digester. This is particularly useful when A/AD is being considered since ASM No. 1 can handle both nitrification and denitrification. Simulations conducted with different bioreactor configurations and different recirculation ratios will allow the designer to select a system capable of optimal performance.

Even if the activated sludge system has been designed by simulation, the aerobic digester can be designed with the simple first-order model presented earlier in this chapter. In that case the total VSS concentration entering the digester would be calculated with Equation 13.3, whereas the influent nonbiodegradable VSS concentration would be calculated as

$$X_{M,V,nO} = f_D \cdot X_{B,H,VO} + X_{D,VO} + X_{I,VO}, \tag{13.21}$$

where the subscript O refers to the influent. The biodegradable solids can then be calculated with Equation 13.5. Once those terms are known, everything can proceed exactly as presented in Sections 13.3.1–13.3.3.

13.4 PROCESS OPERATION

Historically, aerobic digestion has been regarded as a simple process with very modest operational requirements. In fact, aerobic digesters have frequently been used merely as solids holding and

thickening tanks that are aerated to avoid nuisance conditions. In such situations, digestion occurs simply because of the aerobic conditions maintained. As a consequence, troubleshooting guides for aerobic digestion typically emphasize activities to keep equipment in working order and to avoid nuisance conditions, rather than activities aimed at process control.[46] This situation will change as more sophisticated performance requirements are imposed, including the need to achieve specified pathogen inactivation standards or adequate solids stabilization, as evidenced by either a stipulated VSS destruction efficiency or a specified SOUR. More stringent requirements will cause increased emphasis on the maintenance of desired values of the SRT, pH, and temperature. Many of these objectives can be accomplished at existing aerobic digestion facilities. For example, the operation of CAD systems, particularly intermittently fed ones, as A/AD systems, can avoid the precipitous drop in pH often associated with aerobic digestion. In addition, waste solids can be thickened prior to their addition to the digester so that autoheating will occur, resulting in improved digestion performance and pathogen inactivation. Minor physical modifications to reduce heat loss during the winter can also result in elevated digester temperatures. In either case, care should be exercised to ensure that the digester operates either in the mesophilic range (15 to 40°C) or in the thermophilic range (45 to 65°C), but not between the two ranges, where operation will be erratic. Such simple changes can significantly improve process performance. Furthermore, the use of oxidation-reduction potential as a technique for real-time control of the A/AD process should facilitate its operation and optimization and encourage greater full-scale use.[34]

The supernatant from aerobic digestion is often of such poor quality that it cannot be discharged directly to the environment, requiring it to be recycled to the head of the liquid treatment train. This is due in part to the fact that the destruction of biomass results in the release of soluble cellular constituents including nitrogen, phosphorus, micronutrients, and nonbiodegradable organic matter.[9,30,41] In addition, the settleability of aerobically digested solids can be poor, making the suspended solids concentration in the supernatant high. Care must be exercised to minimize the suspended solids content of the supernatant and to control the timing of the return of supernatant to the liquid treatment train so as to minimize any adverse impacts on its performance. If the liquid treatment process is highly loaded, it may be desirable to return supernatant during low nighttime loading periods. In contrast, for nutrient removal systems it may be desirable to return the supernatant during the high loading period when an increased mass of organic matter is available to remove the recycled nutrients. Analysis of the entire treatment system will allow the operator to select the optimal approach.

13.5 KEY POINTS

1. Aerobic digestion has two primary objectives: the destruction of biodegradable particulate organic matter and the inactivation of pathogens present in waste solids.
2. Aerobic digestion is most applicable to the stabilization of waste biological solids, such as those generated by activated sludge and trickling filter facilities. It can also be used to stabilize primary solids, but aerobic digestion of such solids is often less economic than anaerobic digestion.
3. The influent to an aerobic digester contains both biodegradable and nonbiodegradable particulate organic matter. The relative proportions of each depend on the loading and operating characteristics of the process producing the solids.
4. The destruction of biodegradable particulate organic matter can be characterized as a first-order reaction.
5. Both volatile and fixed suspended solids are destroyed during aerobic digestion, although the relative proportions destroyed may not be the same. Fixed suspended solids are lost as they are solubilized and released from the biodegradable particulate organic matter destroyed.
6. Solids stabilization is typically quantified as either the percentage of volatile suspended solids (VSS) destruction achieved during digestion or the specific oxygen uptake rate (SOUR) of the digested solids.

7. In conventional aerobic digestion (CAD) the solids are maintained under aerobic conditions at the ambient temperature for a period of time adequate to achieve the desired degree of solids stabilization and pathogen inactivation. Both intermittent and continuous feed options are available. Nitrification of released ammonia-N typically occurs, resulting in the destruction of alkalinity and depression of the pH.
8. Anoxic/aerobic digestion (A/AD) includes an anoxic and aerobic sequence in the digestion process. Alkalinity produced through denitrification can offset that consumed in the nitrification of the ammonia-N released. Oxygen requirements are also reduced in comparison to CAD.
9. Autoheating of the digester can be achieved if the solids are thickened prior to digestion and the vessel is designed to minimize heat loss. In autothermal thermophilic aerobic digestion (ATAD) such approaches are used to achieve bioreactor temperatures in the 45 to 65°C range. This results in increased rates of solids stabilization and pathogen inactivation. Because nitrification does not occur under thermophilic conditions, pH depression is avoided and oxygen requirements are reduced.
10. The destruction of biodegradable organic matter in an aerobic digester can be characterized using a variety of approaches. Mathematical approaches include a first-order decay model, the simplified model presented in Chapter 5, and International Water Association Activated Sludge Model No. 1. Another approach uses empirical correlations, such as those that relate percentage of VSS destruction to the operating temperature-solids retention time product.
11. Aerobic digestion is most efficient at neutral pH. Maintenance of a neutral pH can be accomplished by the use of A/AD to denitrify any nitrate-N generated, ATAD to eliminate nitrification, or chemical pH control in CAD.
12. The mixing energy required to maintain solids in suspension increases as the suspended solids concentration in the aerobic digester is increased.
13. The performance of an aerobic digester can be improved by designing and operating it as a series of continuous stirred tank reactors (CSTRs) rather than as a single CSTR.
14. Data collected using batch tests can provide the basis for the design of aerobic digesters. Because of the variability associated with the solids' characteristics, a series of tests should be run and a statistical approach used as the basis for the design. The batch tests must be conducted under conditions reflective of those anticipated in the full-scale system.
15. The physical design of an aerobic digester can significantly influence its operation and performance. Heat loss can be a particularly significant problem in colder climates.

13.6 STUDY QUESTIONS

1. Prepare a table that summarizes the typical SRT values, bioreactor suspended solids concentrations, operating temperatures, and pH values for the three types of aerobic digestion. Describe how the differences in operating conditions result in differences in performance.
2. List the benefits and drawbacks of aerobic digestion and describe where it is typically applied.
3. Discuss how ASM No. 1 can be used to evaluate A/AD. Describe the steps required to apply the model, including procedures to calibrate it.
4. What factors determine the biodegradable fraction of waste solids leaving an activated sludge system? Express the results in terms of the simplified model of Chapter 5.
5. Describe the batch technique used to determine the decay coefficient and nonbiodegradable fraction of waste solids, including the data analysis. Why must constant temperature and pH values be maintained during the test?

TABLE SQ13.1
Batch Aerobic Digestion Data

Time, Days	TSS, mg/L	VSS, %
0.00	3080	87.0
0.34	3000	86.6
0.84	2890	85.2
1.9	2630	85.7
2.9	2320	85.9
3.9	2140	87.3
4.9	1926	87.5
5.9	1710	88.0
6.9	1690	87.9
7.9	1590	88.5
8.9	1520	87.7
9.9	1490	88.2
10.9	1320	88.0
11.9	1260	87.3
12.9	1280	87.8
13.9	1140	88.6
14.9	980	86.6
15.9	1060	87.7
16.9	1030	87.3

6. A batch aerobic digestion test was performed on waste activated sludge from a pilot plant treating a soluble wastewater. The temperature was 20°C and the pH was maintained at a value above 6.5. The results are presented in Table SQ13.1. (a) Determine the concentrations of the volatile and fixed nonbiodegradable suspended solids. (b) Determine the decay coefficients for both volatile and fixed suspended solids.
7. A completely mixed aerobic digester is to be designed to treat the waste solids characterized in Study Question 6. The solids will be thickened to 10,000 mg/L prior to digestion but no additional thickening will be practiced in the digester. The winter operating temperature will be 10°C, and the temperature correction factor, θ, is 1.04. The flow rate of the thickened solids will be 500 m^3/day. (a) What SRT is required to provide 38% VSS destruction during winter operating conditions? (b) The temperature of the bioreactor can be elevated to 25°C by insulation and selection of the oxygen transfer device. What SRT is required to achieve a 38% VSS destruction at this temperature? (c) What is the SOUR of the digested solids at 25°C? (d) What is the oxygen requirement for the aerobic digester? (e) What volume must the digester have at 25°C? (f) Assuming that FSS are lost during digestion in a first-order manner, what is the percentage of reduction in TSS at 25°C? (g) How much power is required to mix the bioreactor at 25°C?
8. Use the simple model of Chapter 5 to define the characteristics of the waste solids produced by an activated sludge system operating at 20°C and an SRT of five days. For these calculations use the wastewater characteristics presented in Table E9.4 and the stoichiometric and kinetic parameters in Table E11.2. Assume that all slowly biodegradable substrate is solubilized and converted to readily biodegradable substrate, as was done in using the simple model in Chapter 11. Clearly state all other assumptions. The wastewater flow rate to the activated sludge system is 12,000 m^3/day. How many kg/day of waste solids are produced?

Express your answer both as VSS and TSS. What fraction of the VSS is biodegradable? What is the SOUR of the waste solids? Does the estimated SOUR indicate that the solids are stabilized?

9. Using the information developed in Study Question 8, size an aerobic digester to reduce the SOUR to 1.0 mg O_2/(g VSS·hr). Assume that b_{MV} is numerically equal to b_H. Also assume that FSS are conserved. Compare the SRTs required for configurations consisting of one and two tanks in series. Size the bioreactors, calculate the oxygen requirements, and compare the power required for oxygen transfer to that required for mixing for each configuration. Assume that the digesters will be operated without solids recycle and that feed solids are thickened to 15,000 mg/L as TSS. Also assume that the efficiency of the oxygen transfer device is 1.2 kg O_2/(kW·hr). What percentage of VSS destruction will each digester achieve?
10. Use ASM No. 1 to define the characteristics of the waste solids produced by an activated sludge system operating at 20°C and an SRT of five days. For these calculations use the wastewater characteristics presented in Table E9.4 and the stoichiometric and kinetic parameters in Table 6.3. Clearly state all assumptions. The wastewater flow rate to the activated sludge system is 12,000 m^3/day. How many kg VSS/day of waste solids are produced? What fraction of the waste solids is biodegradable? What is the SOUR of the waste solids? Does the estimated SOUR indicate that the solids are stabilized?
11. Use ASM No. 1 to evaluate the effect of SRT on the performance of a single CSTR CAD system receiving the waste solids characterized in Study Question 10. Model it as a continuous feed process with solids recycle. After preparing a graph of percentage solids destruction versus SRT, choose an SRT to give 38% solids destruction and size the bioreactor to maintain a solids concentration of 15,000 mg/L on a VSS basis. How much alkalinity would have to be supplied to maintain a residual alkalinity of 50 mg/L as $CaCO_3$?
12. Reconsider the CAD system sized in Study Question 11. Maintaining the same SRT and total bioreactor volume, reconfigure the system as an A/AD system like that shown in Figure 1.18c. Then use ASM No. 1 to investigate the effects of the recirculation flow rate and the relative sizes of the anoxic and aerobic zones on the performance of the system. Specifically, investigate the effects of those variables on the percentage of solids destruction, the effluent nitrate-N concentration, the oxygen requirement, and the alkalinity required to maintain a residual alkalinity of 50 mg/L as $CaCO_3$.

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14 Anaerobic Processes

The term anaerobic process refers to a diverse array of biological wastewater treatment systems from which dissolved oxygen and nitrate-N are excluded. In most instances they are operated to convert biodegradable organic matter, both soluble and particulate, to methane and carbon dioxide. Since methane is a sparingly soluble gas, most is evolved and recovered, thereby removing organic matter from the liquid phase and stabilizing any solids present in the influent or produced in the process. Anaerobic digestion of municipal wastewater solids also results in inactivation of pathogens, a step that is usually required prior to ultimate solids disposal. In some cases, anaerobic processes are operated to convert biodegradable particulate organic matter into volatile fatty acids (VFAs), which are subsequently separated from the particulate matter and fed to biological nutrient removal (BNR) systems to enhance their performance. Anaerobic digestion, high-rate suspended growth anaerobic processes, and fermentation processes are addressed in this chapter. Low-rate anaerobic processes such as anaerobic lagoons are addressed in Chapter 15 and attached growth anaerobic processes are addressed in Chapter 21.

14.1 PROCESS DESCRIPTION

14.1.1 General Description

A general description of the microbiology and biochemistry of anaerobic processes is presented in Chapters 2 and 8, while the kinetics of the transformations are summarized in Section 10.3.2. Although the chemistry, biochemistry, and microbiology of anaerobic decomposition are quite complex, it can be conceptualized as comprising three steps, as summarized in Figure 2.4: (1) hydrolysis of particulate organic matter to soluble substrates; (2) fermentation of those soluble substrates to produce acetic acid, carbon dioxide, and H_2; and (3) conversion of the acetic acid, the H_2, and a portion of the carbon dioxide to methane.[51,63,66] Methane is a sparingly soluble gas, which is evolved from solution and collected for subsequent use. The evolution of methane decreases the chemical oxygen demand (COD) of the waste stream and provides the mechanism for stabilization of the biodegradable organic matter contained in it. Only minimal COD reduction occurs without methane production and it is associated with the formation and evolution of H_2. As discussed in Sections 2.3.3 and 10.3.2, the H_2-oxidizing methanogens are fast growing organisms and are present in most anaerobic treatment systems, resulting in conversion of most of the H_2 produced to methane.[63,66,70] However, since the greatest proportion of the methane produced comes from acetic acid, growth of aceticlastic methanogens is required to achieve significant waste stabilization.

Since COD stabilization in anaerobic processes is directly related to methane evolution, methane production can be calculated from the COD removed in the process, just as the oxygen requirement in an aerobic system can be calculated from a COD balance. As discussed in Section 2.3.3, two moles of oxygen are required to oxidize one mole of methane to carbon dioxide and water. Thus, the COD equivalent of methane is 4 kg COD/kg methane. At standard temperature and pressure (0°C and one atmosphere) this corresponds to 0.35 m^3 of methane produced per kg of COD converted to methane.[51,63] For municipal primary solids, the methane equivalent is 0.7 m^3 of methane produced per kg of volatile solids (VS) destroyed.[63] The carbon dioxide content of the gas produced in anaerobic processes ranges between about 30 and 50% and varies depending on the nature of the substrate. For example, the carbon dioxide content is higher when carbohydrates are being treated than when proteins are treated.[63]

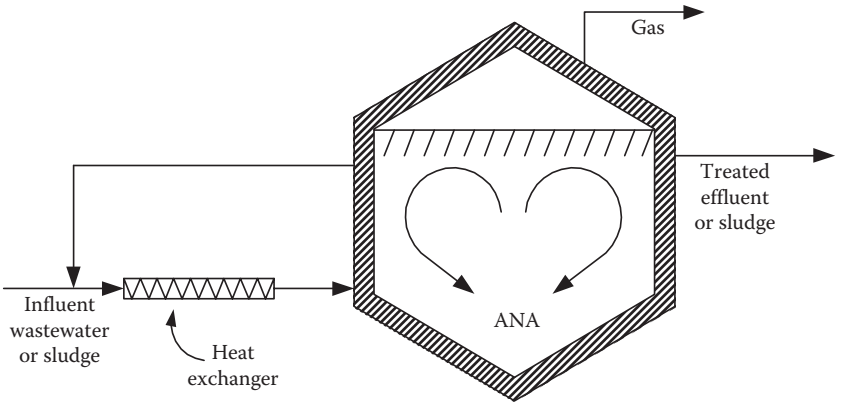

FIGURE 14.1 Anaerobic bioreactor.

Figure 14.1 provides a schematic of an anaerobic bioreactor that illustrates its four major components: (1) a closed vessel, (2) a mixing system, (3) a heating system, and (4) a gas-liquid-solids separation system. Anaerobic bioreactors are typically constructed of either concrete or steel, although earthen basins are used for some low-rate processes as described in Chapter 15. An enclosed vessel is used to exclude dissolved oxygen and ensure the development of anaerobic conditions. The bioreactor is often insulated to minimize heat loss. Mixing is provided to increase the homogeneity of the reaction environment and to reduce the resistance to mass transfer. Uniform bioreactor conditions minimize the impacts of the inhibitory materials produced as metabolic intermediates, keep bioreactor physicochemical parameters within limited ranges, and minimize the impacts of influent flow and composition fluctuations. Due to the high affinity of the reactions for their substrates, performance is not severely impacted by the uniform bioreactor environment. Several methods are used to mix the bioreactor, including devices such as gas recirculation or mechanical mixers, recirculation of bioreactor effluent to the influent, or bioreactor configurations that use the influent and recirculation flows to mix the contents. Gas evolution during treatment results in a degree of mixing that can be significant in certain bioreactor configurations. The configuration of the feed distribution system can also encourage mixing. Heating is provided in many, but not all, cases to maintain temperatures that are constant and near the optimum values for the biomass. Methane gas produced by the system is generally used to fire boilers that provide the necessary heat.

Relatively long solids retention times (SRTs) are required in anaerobic processes because of the low maximum specific growth rates of methanogens. Long SRTs also minimize the buildup of inhibitory reaction intermediates and allow the process to respond better to fluctuations in wastewater flow and composition. In some instances the necessary SRT is achieved by providing a sufficiently long hydraulic residence time (HRT).[49,63] In other cases, the necessary SRT is provided by separating solids from the treated effluent and retaining them in the bioreactor, thereby achieving an SRT that is significantly longer than the HRT.[15,29,73] The gas-liquid-solids separation device is critical to the performance of such systems because the efficiency of liquid-solids separation determines the extent to which active biomass can be accumulated. Gas separation from the solids is necessary to facilitate liquid-solids separation. Several approaches are used to retain active biomass in anaerobic treatment systems, they are described in Section 14.1.3.

14.1.2 Anaerobic Digestion

Anaerobic digestion (AD) is used for the stabilization of particulate organic matter. As illustrated in Figure 1.24, an anaerobic digester is well mixed with no liquid-solids separation.[49,77] Consequently, the bioreactor can be treated as a continuous stirred tank reactor (CSTR) in which the HRT and SRT are identical. An SRT of 15 to 20 days is typically used, although SRTs as low as 10 days have been

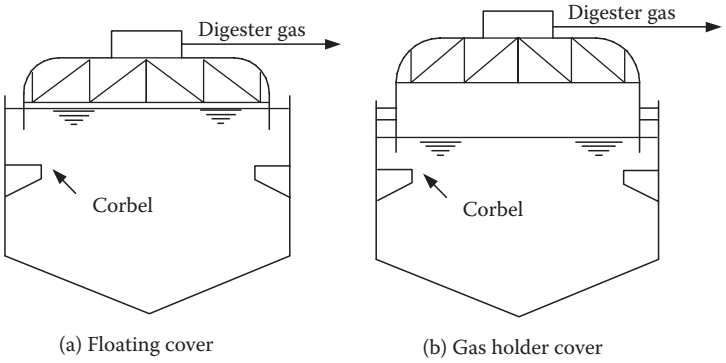

FIGURE 14.2 Gas storage covers for anaerobic digesters.

used successfully and longer SRTs are employed when greater waste stabilization is required.[49,77,85] Many anaerobic digesters are cylindrical concrete tanks with cone-shaped bottoms and steel or concrete covers, although other materials and configurations can be used. Diameters range from 10 to 40 m and sidewall depths from 5 to 10 m. Mixing is required and is provided by internal mechanical mixers, external mechanical mixers that recirculate the tank contents, gas recirculation systems of various types, or pumped recirculation of the tank contents. Historically, relatively low volumetric power inputs have been used to mix anaerobic digesters. However, more recent experience suggests that such practices may result in a significant portion of the bioreactor volume being inactive, as well as in significant short-circuiting of feed to the effluent.[57] In contrast, tracer testing has demonstrated that newer approaches can produce essentially completely mixed conditions, thereby minimizing inactive volume and short-circuiting.[14,57,90]

Methane produced by the process is combusted and used to heat the feed stream and digester contents. Bioreactor temperatures in the mesophilic range (~35°C) are typically maintained,[49,63,77,85] although numerous investigations of the use of thermophilic operating temperatures (~55°C) have been conducted.[7,63] Gas storage is typically provided to accommodate variations in gas production rates, thereby facilitating the operation of boilers and other equipment using the gas as a fuel source. External pressurized storage is sometimes used, but more frequently gas is stored in the digester under a cover that floats on the digester contents, as illustrated in Figure 14.2.[49,54,77,85]

Historically, anaerobic digesters treating municipal wastewater solids have experienced operating problems associated with the accumulation of grit in the bottom and floating scum on the surface.[49,77,85] Consequently, bioreactor configurations have been developed that reduce these problems. One is the egg-shaped digester, illustrated in Figure 14.3.[49,85] The large height-to-diameter ratio and the steeply sloped lower and upper sections of the vessel result in improved mixing, reduced grit and scum accumulation, and easier removal of any that does accumulate. The waffle bottom digester is another configuration that facilitates grit and heavy solids removal.[77]

Many reference works and textbooks discuss two-stage anaerobic digestion, in which two digesters are operated in series.[49,54,77,85] Heating and mixing are provided in the first stage, where active digestion occurs, while quiescent conditions are provided in the second stage for liquid-solids separation. Supernatant from the second stage is recycled to the liquid process train while thickened, settled solids are directed to further processing or ultimate disposal. Although appropriate when primary sludge and/or attached growth biomass is digested, use of the two-stage process is not appropriate when suspended growth biomass is digested. This is because suspended growth biomass does not settle very well, leading to poor quality supernatant and excessive recycle of digested solids back to the liquid stream. Consequently, current practice is to thicken the feed solids prior to single-stage, high-rate anaerobic digestion, which is the process illustrated in Figure 1.24.

One purpose of anaerobic digestion is the stabilization of biodegradable particulate organic matter. Consequently, its performance can be quantified by the percentage of VS destruction. At an

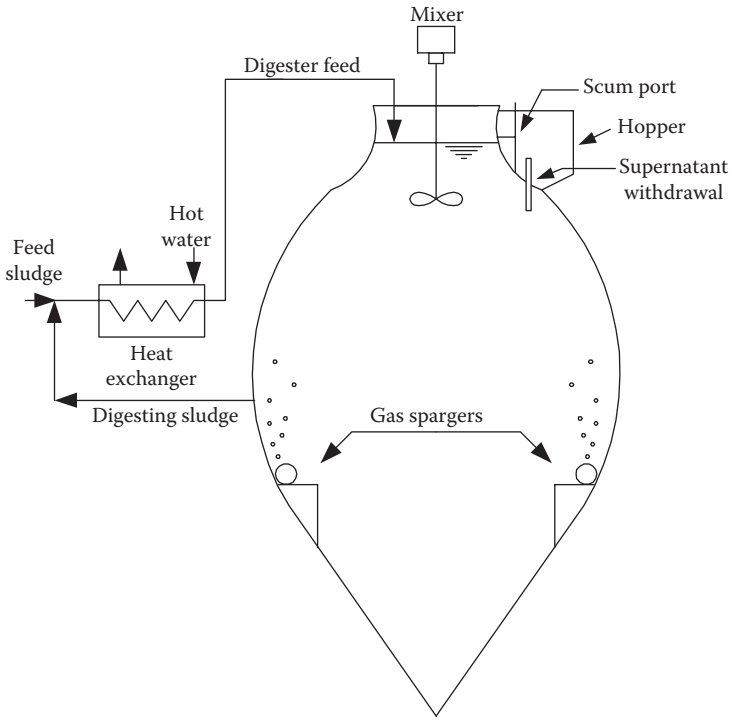

FIGURE 14.3 Egg shaped anaerobic digester.

SRT of 15 to 20 days, 80 to 90% of the influent biodegradable particulate organic matter will be converted to methane gas.[63] This corresponds to destruction of about 60% of the VS contained in primary solids and 30 to 50% of the VS contained in waste activated sludge, as described further in Section 14.2.9.[27,49,63,77] In general, the more stable solids are; that is, the greater the reduction in VS content, the less attractive they will be to disease vectors such as insects and rodents. Consequently, if anaerobically digested biosolids are to be used in various ways, such as a fertilizer in agriculture, they must meet criteria indicating that sufficient stabilization has occurred. Just as with aerobic digestion, discussed in Chapter 13, these practices are governed by regulations. In the United States the 503 regulations apply.[78] The 503 regulation requirements for anaerobic digestion are summarized in Table 14.1 and for vector reduction include at least 38% destruction of VS or no more than 17% VS reduction following 40 days of additional batch digestion at mesophilic temperature.

A second purpose of anaerobic digestion, which is particularly important when the biosolids are to be used, is the destruction of pathogens. Consequently, the degree of pathogen destruction required in the United States is also specified by the 503 regulations.[78] Just as with aerobic digestion, two levels of pathogen control are defined: Class A and Class B. As indicated in Table 14.1, Class B biosolids must contain fecal coliform levels less than 2×10^6 most probable number (MPN)/g total solids (TS). Alternatively, the 503 regulations specify that, if an SRT of 15 days is maintained in a conventional mesophilic (35°C) anaerobic digester (60 days at 20°C) it is presumed that a Class B product is being produced with regard to pathogen control. As indicated in Table 14.1, the regulations for Class A biosolids are more stringent, requiring that the fecal coliform level be less than 1000 MPN/g TS or the Salmonella level less than 3 MPN/4 g TS. Alternatively, it is presumed that a Class A product is being produced with regard to pathogen control if one of a series of time and temperature requirements is met as specified in Table 14.1.

Increased concern about pathogens in biosolids has resulted in significant efforts to develop anaerobic digestion processes that can meet Class A pathogen standards.[67,86] These studies have also demonstrated that increased VS destruction can be obtained, at least in some cases, thereby reducing

TABLE 14.1
503 Regulations as Applied to Anaerobic Digestion

Vector reduction

- 38% volatile solids reduction, or
- Less than 17% volatile solids reduction following 40 days of additional anaerobic digestion at 30°C–37°C

Class B pathogen reduction

- An SRT of at least 15 days at 35°C–55°C and 60 days at 20°C, or
- Fecal coliform less than 2×10^6 MPN/g TS

Class A pathogen reduction

- Sludge at less than 7% solids is held in a batch mode at any one of the combination of times and temperatures defined by the following equation: $D = 50{,}700{,}000/10^{0.1400T}$ where T is the temperature in °C, specified to be at least 50°C and D is the batch holding time in days, specified to be greater than 0.0208 days (30 min). In any case, a minimum of 30 min must be provided. Application of this equation, and applying the 30 min minimum requirement, results in the following combinations:

Temperature (°C)	Time
70	30 min
65	57 min
60	4.8 hr
55	1 day
50	5 day

- Fecal coliform less than 1000 MPN/g TS, or
- *Salmonella* less than 3 MPN/4 g TS

Note: From U.S. Environmental Protection Agency, Control of Pathogens and Vector Attraction in Sewage Sludge, EPA/625/R-92/013, U.S. Environmental Protection Agency, Washington, DC, 2003.

the amount of residual biosolids that must be managed and increasing biogas production. Generally referred to as advanced anaerobic digestion (AAD) processes, development and testing is continuing and is likely to result in more widely accepted process options. These systems generally incorporate tanks-in-series configurations that are quite different from the outdated two-stage system discussed above. In one variation, the first stage is operated at an SRT that allows fermentation but prevents methane formation from the volatile fatty acids (VFAs) formed. The second stage is operated at a sufficiently long SRT for methanogenesis. This option is referred to as phase separation because the two "phases" of anaerobic digestion occur in separate bioreactors.[25,26] Multiple tanks-in-series can also be provided in either the acidogenic or methanogenic stages.[67,86] In other options, methanogenesis occurs in each stage, with some operated in the thermophilic temperature range and others in the mesophilic temperature range. Pathogen destruction is enhanced in these processes through elevated temperature (if thermophilic stages are provided), nonionized VFAs in acid phase stages (due to both the elevated VFA concentration and the depressed pH), and increased plug flow through the process. Some stages can also be operated in a batch mode to eliminate short-circuiting through an individual stage.

14.1.3 High-Rate Anaerobic Processes

High-rate anaerobic processes utilize bioreactor configurations that provide significant retention of active biomass, resulting in large differences between the SRT and the HRT.[15,29,73] Three mechanisms

TABLE 14.2
Typical High Rate Anaerobic Process Performance

Parameter	Value
BOD_5 removal, percent	80%–90%
COD removal, mass	$1.5 \times BOD_5$ removed
Biogas production	0.5 m^3/kg COD removed
Methane production	0.35 m^3/kg COD removed
Biomass production	0.05–0.10 g VSS/g COD removed

Note: Adapted from Hall, E. R., Anaerobic treatment of wastewaters in suspended growth and fixed film processes. Design of Anaerobic Processes for the Treatment of Industrial and Municipal Wastes, eds. J. F. Malina Jr. and F. G. Pohland, 41–118, Technomics Publishing Co., Inc., Lancaster, PA, 1992.

are used to retain biomass: (1) the formation of settleable particles that are retained by sedimentation, (2) the use of reactor configurations that retain suspended solids, and (3) the growth of biofilms on surfaces within the bioreactor. In many instances, more than one mechanism is operating in a bioreactor. Consequently, high-rate anaerobic processes represent a spectrum of bioreactor types ranging from suspended growth to attached growth, with hybrid bioreactors, which contain significant quantities of both suspended and attached biomass, in between. One of the earliest high-rate anaerobic processes was anaerobic contact, which was essentially configured like the activated sludge process but with an anaerobic bioreactor and sludge de-gassing to allow the anaerobic biomass to settle in the downstream solids separator. This process is seldom used today and has been superseded by more economical process options. Three bioreactor types that span the range of options in practical application today are described in this section: (1) upflow anaerobic sludge blanket (UASB), (2) anaerobic filters (AF), and (3) hybrid UASB and anaerobic filters (UASB/AF). The more recently developed expanded granular sludge blanket (EGSB) process is also described. Fluidized bed/expanded bed (FB/EB) processes are also used on occasion and are addressed in Chapters 18 and 21.

Hall[29] has summarized the typical performance of high-rate anaerobic processes, as presented in Table 14.2. A relatively high level of biodegradable organic matter removal can be achieved, as indicated by typical five-day biochemical oxygen demand (BOD_5) removal efficiencies of 80 to 90%. Biogas production is about 0.5 m^3/kg COD removed, corresponding to a methane production of 0.35 m^3/kg COD removed. Solids production is low, typically ranging from 0.05 to 0.10 kg VSS/kg COD removed. These performance levels can be achieved by all of the processes discussed in this section if appropriate organic loading rates are used.

14.1.3.1 Upflow Anaerobic Sludge Blanket

The UASB process uses suspended growth biomass, but the gas-liquid-solids separation system is integral with the bioreactor. More importantly, the environmental conditions created in the bioreactor can result in the development of large, dense, readily settleable particles called granules, which allow very high concentrations of suspended solids, on the order of 20 to 30 g/L as VSS, to be accumulated.[29,46,47] These high suspended solids concentrations allow significant separation between the SRT and HRT, and operation at relatively short HRTs, often on the order of two days or less, even when the SRT is long. No single theory has been accepted for explaining the formation of granules in the UASB process.[36] However, there is consensus that the initial stages of adhesion are similar to the initial stages of biofilm formation, discussed in Section 16.1, and most theories confirm that the aceticlastic methanogen *Methanosaeta* plays a key role. Once granulation begins, the hydrodynamic conditions in the bioreactor play an important role by washing out dispersed

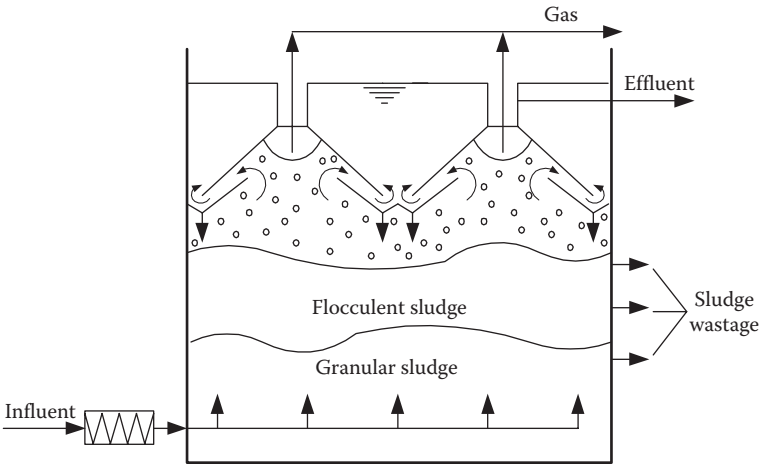

FIGURE 14.4 Upflow anaerobic sludge blanket (UASB) bioreactor.

biomass, thereby favoring the growth of the bacteria in the granules. The complex ecology of methanogenic cultures is reflected in the structure of the granules, with producers and consumers of H_2 and acetic acid growing in close proximity to one another, thereby facilitating their rapid utilization.

Figure 14.4 provides a schematic of the process. Influent wastewater enters the bottom of the bioreactor through a distribution system that is designed to provide relatively uniform flow across its cross section. A dense slurry of granules forms in the lower portion of the bioreactor, and the combined effects of the influent wastewater distribution and gas production result in mixing of the influent wastewater with the granules. Treatment occurs within the dense blanket of granules. For some wastewaters, a much less dense flocculent sludge also develops, and this accumulates on top of the blanket of granules provided the upflow velocity is insufficient to carry it away. Other wastewaters contain suspended solids that are not trapped in the granular sludge, and these solids can also accumulate as a flocculent sludge blanket overlying the granules. Treated effluent exits the granular and flocculent sludge zones and flows upward into the gas-liquid-solids separator. A variety of configurations can be used for this device, and the one illustrated in Figure 14.4 is only meant to represent the basic concepts used by several manufacturers. The device often consists of a gas collection hood with a settler section above it. Gas bubbles and the upward flowing liquid cause some granular and flocculent solids (particularly small granules) to rise through the bioreactor and enter the gas-liquid-solids separator. Gas separation occurs in the hood area, thereby allowing some of this suspended material to return directly to the solids blanket. Gas collects in the upper inverted V section of the hood and is removed from the bioreactor. Liquid with some entrained solids flows out of the hood into the settler section where liquid-solids separation occurs. Clarified effluent overflows the weirs and is discharged while separated solids settle back into the reaction zone. Design of the gas-liquid-solids separation device requires insight into the physical processes occurring there and experience with specific devices in a variety of applications.

Bioreactor dimensions are affected by process loadings, constraints on maximum upflow velocities, wastewater type, and the settling characteristics of the solids that develop in the process.[29,46,47] The solids inventory increases as treatment occurs and new biomass is grown. Consequently, provisions must be made for solids wastage, as illustrated in Figure 14.4. The relative proportions of flocculent and granular sludge can be controlled by the wasting locations used. Bioreactor HRTs in the 0.2 to 2 day range are typical, along with volumetric organic loading (VOL) rates of 2 to 25 kg COD/(m^3·day), depending on wastewater characteristics and whether granular or flocculent solids develop.

14.1.3.2 Anaerobic Filter

Anaerobic filter systems use upflow bioreactors that are filled with media. The packing is the same as that used with aerobic plastic media trickling filters, discussed in Chapter 19. Example media are illustrated in Figure 1.21; the specific surface area is typically 100 m^2/m^3 with a void volume of 90 to 95%. The presence of packing allows for the growth of some attached biomass, but the primary role of the media is to retain suspended growth.[29,72,88] The media may be thought of as performing like a set of tube settlers, which provide enhanced liquid-solids separation and retention of suspended biomass within the bioreactor. Gas-solids separation is also facilitated within the packed section. Although several types have been used successfully in AF systems, direct comparisons indicate advantages for the cross-flow modular media because of its superior gas-liquid-solids separation capabilities.[72,88]

Figure 1.21 provides a schematic of the overall AF process. Influent wastewater and recirculated effluent are distributed across the bioreactor cross section and flow upward through the media. Treatment occurs as a result of the suspended and fixed biomass retained by the media. Effluent exits the top of the media section and is collected for discharge. Gas is collected under the bioreactor cover and is conveyed to subsequent use. Effluent is typically recirculated to maintain a reasonably uniform hydraulic loading on the bioreactor in spite of varying influent flow rates, thereby maintaining uniform bioreactor hydrodynamic conditions. Although performance is determined by the SRT maintained, accurate assessment of the bioreactor suspended solids inventory is not generally possible. Consequently, bioreactor designs are based on the HRTs and VOLs used successfully in other applications. Hydraulic residence times between 0.5 and 4 days are typical, along with VOLs in the 5 to 15 kg COD/(m^3·day) range. The biomass inventory is typically controlled by the hydrodynamic conditions that develop in the media as a result of the influent (wastewater plus recirculation) flow applied. Excess biomass is washed out of the system as it develops and becomes a part of the effluent. In some instances the capability to remove settled solids from the bioreactor bottom may be provided since heavy solids and precipitates can accumulate there. Solids removal from this location does not constitute an SRT control mechanism, however, because most of the active biomass is retained within the media section.

14.1.3.3 Hybrid Upflow Anaerobic Sludge Blanket/Anaerobic Filter

Hybrid UASB/AF systems combine aspects of the UASB process with aspects of the AF process.[29] Influent wastewater and recirculated effluent are distributed across the bioreactor cross section and flow upward through granular and flocculent sludge blankets where anaerobic treatment occurs. The effluent from the sludge blanket zone enters a section of media identical to that used in AF systems where gas-liquid-solids separation occurs. Treated effluent then exits the media section and is collected for discharge from the bioreactor. Gas collects under the bioreactor cover and is transported to storage and/or use. The hybrid UASB/AF process primarily uses suspended biomass, and process loadings are similar to those used with the UASB process. The solids removal system is similar to that used with the UASB process. This process is illustrated in Figure 1.23.

14.1.3.4 Expanded Granular Sludge Bed

Expanded granular sludge bed (EGSB) systems represent an extension of the UASB process wherein the granular sludge bed is expanded by the use of a deeper and narrower bioreactor to increase the upflow velocity. The narrower bioreactor also increases the upflow impact of biogas production because the production rate per unit of bioreactor cross-sectional area is increased.[91] The net result is increased mass transfer, thereby allowing increased VOLs. In general, the VOL for an EGSB system is about double that for a comparable UASB.[24] Specialized solids and gas separators are used with these systems to accommodate the higher hydraulic and organic loading rates. Internal recirculation of process flow is also applied in some instances to increase mass transfer.[24,81]

The basic process options described in this section can be combined in a variety of ways to produce a wide range of additional anaerobic treatment systems. For example, interest currently exists in the use of membranes as a means of further separating the SRT and the HRT, thereby producing an even more compact anaerobic process.[19,31] Staged systems are also being evaluated for specific applications, such as when separation of fermentation and methanogenesis or separation of the biodegradation of readily and slowly biodegradable organics is desired.

14.1.4 Solids Fermentation Processes

Solids fermentation processes are used to solubilize particulate organic matter in primary solids and ferment the soluble products to VFAs, particularly acetic and propionic acid, for use in BNR processes.[71,83] The objectives of solids fermentation processes are different from those of the anaerobic stabilization processes discussed previously. They are to maximize the production of VFAs and recover them in a stream that can be delivered to a BNR system. The first objective is achieved by controlling the SRT to a value that allows the growth of hydrolytic and fermentative bacteria but prevents the growth of aceticlastic methanogens, which would consume the VFAs.[17,18,71] As indicated in Figure 10.5, at 35°C this requires an SRT in the 2 to 3 day range. In general, the feed solids and bioreactor contents are not heated, so the SRT must be increased to compensate for the lower temperature. Some methane will be produced as a result of the growth of H_2-utilizing methanogens, but the amount will be small. The second objective is achieved when the VFAs are separated from the residual primary solids by passing the bioreactor effluent through a liquid-solids separation step.

Figure 14.5 illustrates schematically the concepts of fermentation systems. Feed solids are fed to a mechanically mixed bioreactor where fermentation occurs. The SRT is controlled by adding dilution water in sufficient quantities so that the HRT, which equals the SRT, is maintained at the desired value. The use of gravity sedimentation to achieve liquid-solids separation is illustrated in Figure 14.5. The option of adding elutriation flow to the bioreactor effluent is provided to ensure sufficient supernatant to effectively recover the produced VFAs. Typically, the settled solids are removed from the settler and taken to further processing. However, the capability to recycle a portion of those solids to the bioreactor may be provided to increase its SRT above its HRT.

Figure 14.6 illustrates how the concepts in Figure 14.5 have been implemented at several full-scale wastewater treatment plants. In an activated primary clarifier (Figure 14.6a), primary solids are accumulated in a sludge blanket where fermentation occurs. The settled solids are then recycled to an upstream mixing/elutriation tank where the soluble VFAs are washed from the fermented primary solids and into the primary clarifier effluent. The SRT is controlled by wasting settled solids from the process. In the completely mixed fermenter (Figure 14.6b), solids are removed from the

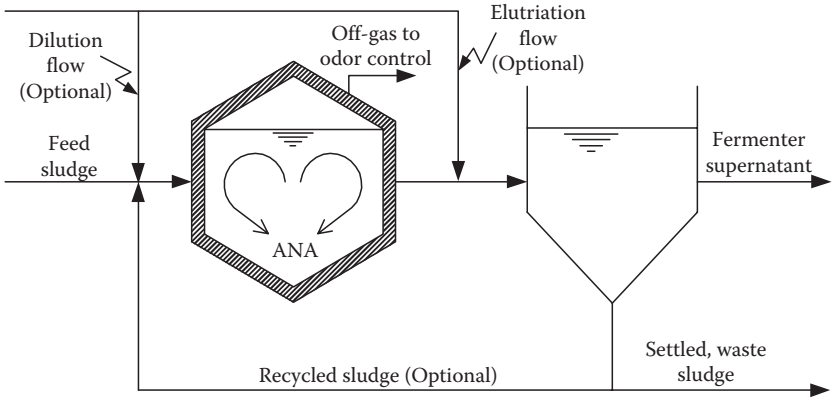

FIGURE 14.5 Solids fermentation process.

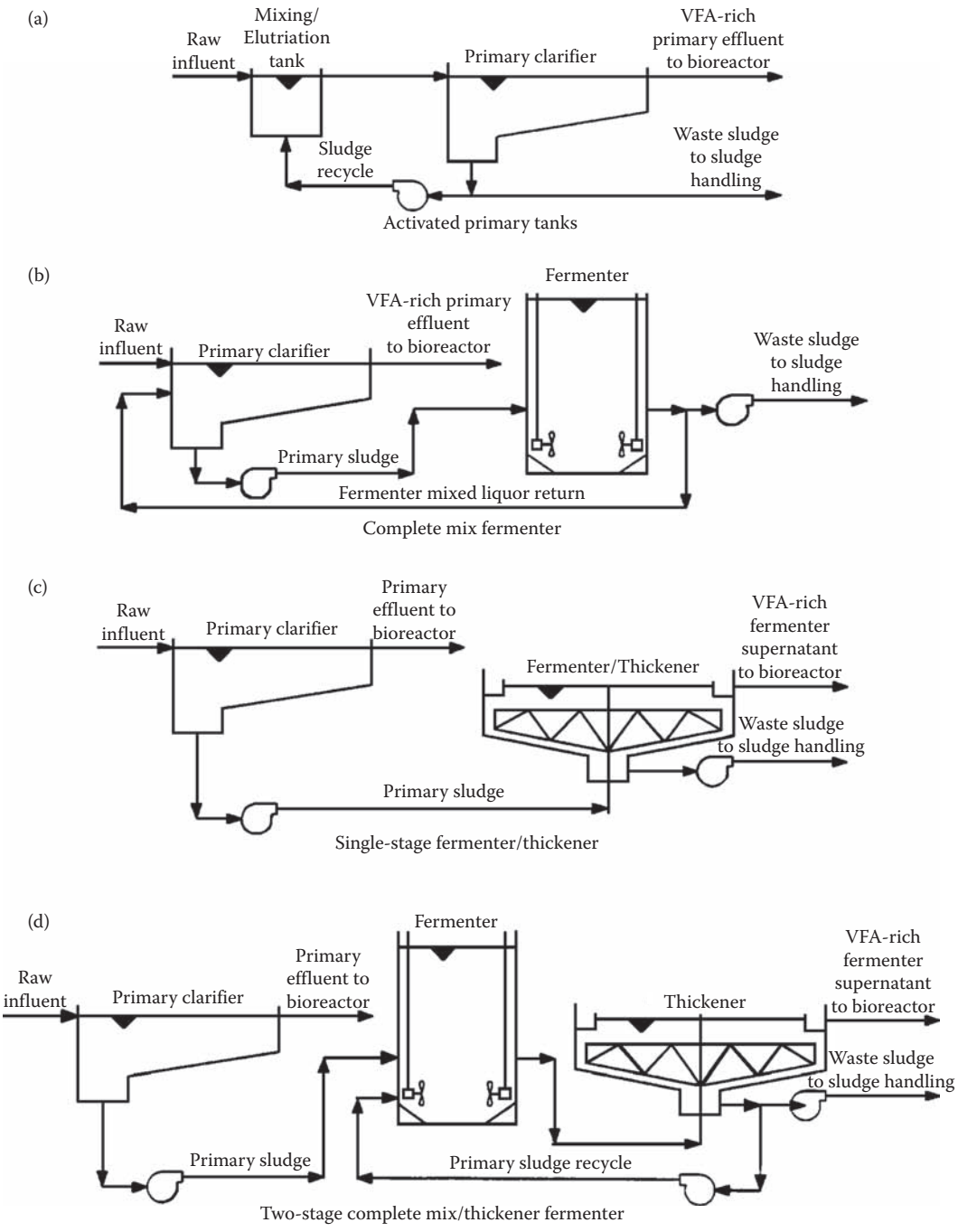

FIGURE 14.6 Alternative configurations for the solids fermentation process: (a) activated primary clarifier, (b) completely mixed fermenter, (c) single-stage fermenter/thickener, and (d) two-stage completely mixed fermenter/thickener. (From Water Environment Federation, Use of fermentation to enhance biological nutrient removal, *Proceedings of the Conference Seminar, 67th Annual Water Environment Federation Conference and Exposition*, Water Environment Federation, Alexandria, Virginia, 1994, Copyright © Water Environment Federation. Reprinted with permission.)

primary clarifier and fermented in a separate bioreactor. However, the fermented solids are then recycled to the primary clarifier where VFAs are removed by the wastewater flow and fermented solids are wasted to control the SRT. In the single-stage fermenter/thickener (Figure 14.6c), primary solids are added to an oversized gravity thickener where both solids fermentation and liquid-solids separation occur. Solids wasting is controlled to achieve the desired SRT, and VFAs are removed in the overflow. Finally, in the two-stage completely mixed fermenter/thickener system (Figure 14.6d), the fermentation and solids thickening steps are separated into two unit operations. Optional primary effluent addition points provide operational flexibility to control the SRT and VFA elutriation. This option is functionally identical to the prototype solids fermentation process illustrated in Figure 14.5.

All the options presented in Figure 14.6 use gravity liquid-solids separation. Relatively poor solids thickening has been experienced in some instances, probably as a result of the gases produced. Such operating problems have been controlled in some cases by the operation at a reduced fermentation reactor solids concentration and/or by dilution of the bioreactor effluent prior to gravity separation. Alternatively, centrifuges have been used.[8,39] Other options include static fermenters or fermentation basins upstream of biological reactors. Details of the process configurations developed to date are presented elsewhere.[83]

14.1.5 Comparison of Process Options

Table 14.3 summarizes the primary benefits and drawbacks of the anaerobic treatment systems used to stabilize organic matter. Conventional mesophilic anaerobic digestion is suitable for a wide range of wastewaters, particularly those with high concentrations of suspended solids. Well-mixed conditions are provided within the bioreactor, resulting in a uniform environment that produces predictable and stable performance. Process performance is not dependent on solids settleability since anaerobic digesters use completely mixed bioreactors with no biomass recycle. The long HRTs required to achieve adequate SRTs result in large bioreactor volumes that can effectively dilute toxic materials. However, they can cause high capital costs. Effluent quality can be poor if the influent contains high concentrations of nonbiodegradable organic matter. Process stability will be good if a sufficient SRT is provided, but it will be poor at shorter SRTs. The process requires a separate mixing system. Pathogen destruction can be limited for the conventional mesophilic process. Advanced anaerobic digestion processes are being developed to improve pathogen destruction to meet Class A biosolids standards and also to improve VS destruction. These systems are more complex than conventional mesophilic anaerobic digestion processes and their performance characteristics and design criteria are not yet well developed.

All high-rate anaerobic processes share certain characteristics. High biomass concentrations are maintained, thereby allowing long SRTs to be achieved while keeping the HRTs short. The high biomass concentrations allow high VOLs to be applied, resulting in relatively small bioreactors, and the long SRTs provide good process stability. Although the systems are compact and require relatively small land areas, a high quality effluent can generally be achieved. Well-mixed conditions are typically produced in the bioreactor, resulting in a uniform reaction environment. High-rate processes are best suited for the treatment of wastewaters containing soluble organic matter, and are adversely impacted by the presence of high concentrations of influent suspended solids. The relatively short HRTs possible with these systems mean that less equalization is provided and less dilution is available for toxic inputs. Important differences also exist among the various high-rate processes.

Upflow anaerobic sludge blanket systems are mechanically simple and easy to operate, but their performance is dependent on the formation of dense, settleable biomass granules. Furthermore, the allowable organic loading rate is adversely impacted by the presence of suspended solids in the influent wastewater, special bioreactor configurations are required, and little process control is possible.

TABLE 14.3
Anaerobic Treatment Process Comparison—Organic Stabilization

Process	Benefits	Drawbacks
Conventional mesophilic anaerobic digestion (AD)	• Suitable for a wide range of wastewaters • Efficiently handles high suspended solids wastewaters • Easy to mix, thereby creating uniform reaction environment • Large bioreactor volume to dilute inhibitors • Performance not dependent on sludge settleability • Capable of accepting waste aerobic biomass	• Large bioreactor volumes required • Effluent quality can be poor if nondegradable organic matter is present or if a large concentration of anaerobic organisms is generated • Process stability and performance poor at short SRTs • Requires separate mechanical mixing
Advanced anaerobic digestion (AAD)	• Suitable for a wide range of wastewaters • Efficiently handles high suspended solids wastewaters • Easy to mix, thereby creating uniform reaction environment • Large bioreactor volume to dilute inhibitors • Performance not dependent on sludge settleability • Capable of accepting waste aerobic biomass • Increased control of pathogens compared to AD • Increased organic matter stabilization compared to AD	• Large bioreactor volumes required • More complex than AD • Process performance not well characterized • Effluent quality can be poor if nondegradable organic matter is present or if a large concentration of anaerobic organisms is generated • Process stability and performance poor at short SRTs • Requires separate mechanical mixing
Upflow anaerobic sludge blanket (UASB)	• High biomass concentrations and long SRTs achievable • Small bioreactor volumes due to high volumetric organic loading rates • High quality effluent achievable • Mechanically simple • Compact system, relatively small land area • Well-mixed conditions produced	• Performance dependent on development of dense, settleable solids • Much lower process loading required if wastewater contains suspended solids • Special bioreactor configuration required that is based on experience • Little process control possible • Shorter bioreactor HRTs mean less equalization and dilution of inhibitors
Anaerobic filter (AF)	• High biomass concentrations and long SRTs achievable • Small bioreactor volumes due to high volumetric organic loading rates • High quality effluent achievable • Mechanically simple • Compact system, relatively small land area • Performance not dependent on development of dense, settleable solids • Well-mixed conditions produced in bioreactor	• Suspended solids accumulation may negatively impact performance • Not suitable for high suspended solids wastewaters • Little process control possible • High cost for media and support • Shorter bioreactor HRTs mean less equalization and dilution of inhibitors

TABLE 14.3 (CONTINUED)
Anaerobic Treatment Process Comparison—Organic Stabilization

Process	Benefits	Drawbacks
Hybrid UASB/AF	• High biomass concentrations and long SRTs achievable • Small bioreactor volumes due to high volumetric organic loading rates • High quality effluent achievable • Mechanically simple • Compact system, relatively small land area • Performance partially dependent on development of dense, settleable solids • Well-mixed conditions generally produced in bioreactor • Reduced media cost	• Lower process loadings required if wastewater contains suspended solids • Little process control possible • Shorter bioreactor HRTs mean less equalization and dilution of inhibitors
Expanded granular sludge bed (EGSB)	• High biomass concentrations and long SRTs achievable • Smallest bioreactor volumes due to highest volumetric organic loading rates • High quality effluent achievable • Mechanically simple • Compact system, relatively small land area • Well-mixed conditions produced	• Performance most dependent on development of dense, settleable solids • Process will not function if granular sludge does not develop. Thus, application depends on specific wastewater characteristics • Special bioreactor configuration required that is based on experience • Little process control possible • Shorter bioreactor HRTs mean less equalization and dilution of inhibitors

The design of AF systems is quite straightforward. Furthermore, they are not dependent on the development of a dense, settleable biomass because the media provides the primary mechanism for biomass retention. However, excessive accumulations of suspended solids can lead to plugging, which negatively impacts process performance; therefore, they are not suitable for wastewaters containing high concentrations of suspended solids. Little process control is possible, and the cost of the media and its associated supports can be relatively high.

Hybrid UASB/AF systems combine the advantages of their parent systems. However, they are still adversely impacted by the presence of suspended solids in the influent wastewater and little process control is possible.

Expanded granular sludge bed systems provide the principal advantages of UASB systems but allow even greater VOLs, thereby resulting in reduced reactor volumes. However, performance is even more dependent on the development of a readily settleable granular sludge. Consequently, these systems are applicable to a smaller number of wastewaters.

Although high organic removal rates can be achieved with all of the high-rate anaerobic processes, differences exist for soluble materials. The highest rates of soluble substrate removal are generally achieved in EGSB systems because of their high biomass concentrations and excellent mass transfer characteristics. High soluble substrate removal rates can also be achieved in UASB and hybrid UASB/AF systems, particularly when a dense, readily settleable, granular sludge develops. This is because of the high biomass concentrations in the granular sludge bed and the mixing caused by the introduction of influent wastewater and the evolution of gas. Soluble substrate removal rates are lower in AF systems because of their lower biomass concentrations and poorer mixing conditions.

14.1.6 Typical Applications

As discussed in Chapter 10, anaerobic processes are typically used to stabilize the organic matter present in wastewaters with biodegradable COD concentrations greater than about 1000 mg/L. Compared to aerobic systems, the advantages of anaerobic processes include less solids production, lower nutrient requirements, lower energy requirements, and the production of a potentially useful product, methane. On the other hand, the effluent quality from anaerobic processes is generally not as good as from aerobic processes, and aerobic polishing may be required to achieve effluent quality goals. Anaerobic processes can be more sensitive than aerobic processes to shock loads and toxic materials, although the anaerobic process technology developed in the past 10 years has demonstrated significant resistance to them. Finally, anaerobic processes are capable of metabolizing some organic compounds not readily biodegraded in aerobic systems. Examples include chlorinated organics, which can be dechlorinated in anaerobic treatment systems even though they are not readily biodegraded in aerobic systems.[73,79] These advantages have led to the application of UASB technology to the treatment of domestic wastewater in tropical locations where wastewater temperatures are generally 20°C or greater. Today several hundred such systems are in use in developing countries and are achieving BOD_5 and COD removal efficiencies in the range of 60 to 80% at VOLs of 3 kg COD/(m^3·day) and HRTs of 6 to 10 hours.[22,23,80] This interesting development is one that should be watched for potentially broader application.

Several factors affect the choice between anaerobic and aerobic treatment systems for wastewaters with biodegradable COD concentrations in the 1000 to 4000 mg/L range. One is wastewater temperature. Anaerobic processes are often operated at temperatures near the optimum for either mesophilic (30 to 40°C) or thermophilic (50 to 60°C) microorganisms because deviations from these ranges can result in significant reductions in microbial activity, which must be compensated for by increases in the design SRT. In general, the impact of temperature on the design SRT is greater for anaerobic than for aerobic processes. This is offset somewhat because the methane produced in the anaerobic process can be used to heat the influent wastewater. Because the quantity of methane produced is a function of the concentration of biodegradable organic matter in the influent wastewater, the potential heat rise depends on the wastewater strength, as illustrated in Figure 14.7. Two cases are considered. One incorporates recovery of the heat in the bioreactor effluent for heating the influent wastewater, while the other does not. Sufficient energy is available to achieve a significant temperature increase only for wastewaters with biodegradable COD concentrations greater than about 2000 mg/L if heat recovery is practiced and around 7000 mg/L if heat recovery is not practiced. The use of anaerobic treatment should not be ruled out for wastewaters of lower strength, however, because anaerobic systems can achieve reliable low temperature operation by increasing the SRT.[59,81] Wastewater flow rate also affects the choice between aerobic and anaerobic systems for wastewaters containing 1,000 to 4,000 mg/L of biodegradable COD. The simplicity of aerobic systems generally favors their use for smaller wastewater flows, while the significant energy and solids production savings available favors the use of anaerobic systems for larger wastewater flows.[29]

Wastewater composition also affects the choice between anaerobic and aerobic systems. High-rate anaerobic treatment technology was developed for the treatment of soluble organic matter, and the presence of significant quantities of suspended solids adversely impacts its efficiency. As indicated in Figure 10.5, either acidogenesis or methanogenesis can be the rate limiting step in the anaerobic stabilization of soluble organic matter, with the nature of the organic matter determining which is slower. If the organic matter is predominantly simple carbohydrates and proteins, methanogenesis will be slower, but can still be accomplished at short SRTs. In fact, some of the high-rate anaerobic systems were developed for food processing wastes containing such constituents. On the other hand, wastes high in lipids require much longer SRTs for acidogenesis, which can increase the SRT required in an anaerobic system. Hydrolysis and fermentation are generally the rate limiting steps in the anaerobic stabilization of particulate organic matter and longer SRTs are required for them as well. Furthermore, some of the suspended solids are likely to be nonbiodegradable and

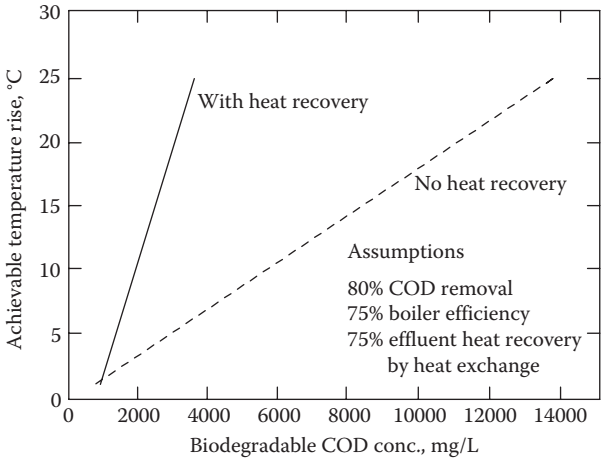

FIGURE 14.7 Relationship between wastewater strength and achievable temperature rise for anaerobic processes. (From Hall, E. R., Anaerobic treatment of wastewaters in suspended growth and fixed film processes. *Design of Anaerobic Processes for the Treatment of Industrial and Municipal Wastes*, eds. J. F. Malina Jr. and F. G. Pohland, Technomics Publishing Co., Inc., Lancaster, Pennsylvania, pp. 41–118, 1992. Copyright © Taylor and Francis Group, LLC, a division of Informa plc. Reprinted with permission.)

will accumulate in the bioreactor, thereby reducing the specific activity of the anaerobic biomass. Both of these factors significantly affect the VOLs that can be applied to an anaerobic bioreactor and negatively impact its economics. The presence of suspended solids can also interfere with the development of granular sludge, thereby adversely impacting the VOL that can be achieved in processes that are dependent on sludge granulation. The presence of inhibitory or toxic chemicals also results in significant increases in the required SRT, which negatively impacts the economics of anaerobic processes. Finally, as illustrated in Figure 10.2, differences also exist among anaerobic processes with respect to the waste strengths for which they are suited.

Anaerobic digestion is generally applied to the treatment of high strength wastewaters, particularly those with high suspended solids concentrations. In fact, historically it has been one of the most widely used processes for stabilizing organic solids produced in wastewater treatment plants. As a consequence, several thousand operating facilities exist around the world. The uniform reaction conditions and long SRTs used provide the conditions necessary for hydrolysis and stabilization of these materials. Anaerobic digesters are capital intensive, but with low operating costs. Consequently, they are generally found in larger wastewater treatment plants where the savings in operating cost more than offset the increased capital investment. For example, at current unit costs it is often found that anaerobic digestion is not cost-effective in wastewater treatment plants with capacities less than 40,000 to 100,000 m^3/day. Nevertheless, because cost relationships and available options were different in the past, they are often found in many older wastewater treatment plants with lower capacities. The increased availability of alternative approaches to solids management and stabilization has reduced the use of anaerobic digestion. However, it remains a viable solids stabilization technology, and continues to be widely used.

The conventional, mesophilic anaerobic digestion process (single-stage, operated at 35°C) has traditionally been the "workhorse" for stabilizing organic solids, but increased concern about pathogens in digested biosolids, coupled with the desire for increased levels of stabilization, is leading to changes in digestion practice. One way to produce Class A biosolids from conventional mesophilic anaerobic digestion is by pasteurization of the sludge prior to its addition to the digester. Typically referred to as "prepasteurization," this is accomplished by heating the sludge to 70°C and holding it in a batch mode for 30 minutes, thereby meeting the time and temperature requirements for Class A biosolids listed in Table 14.1. No significant digestion occurs during pasteurization

because of the elevated temperature (outside the physiological range) and the limited holding time. Other approaches involve the development and application of advanced digestion processes, as discussed in Section 14.1.2. The temperature phased anaerobic digestion (TPAD) process is typical. It contains a thermophilic first stage, generally operated at a temperature of 55°C and an SRT of 5 to 10 days, followed by a mesophilic second stage operated at 35°C with an SRT of 10 days or more.[30,86] The purpose of the first stage is pathogen destruction and partial solids stabilization, whereas the purpose of the second stage is further solids stabilization. Mesophilic conditions are used in the second stage because it has generally been observed that the product of mesophilic digestion is less odorous than the product from thermophilic digestion. The second stage is typically operated as a CSTR, but the first stage is operated as a modified sequencing batch reactor (SBR) to achieve the time and temperature relationship required to meet the Class A pathogen reduction requirements of the 503 regulations. As seen in Table 14.1, at a temperature of 55°C, all sludge solids must be held for at least one day to achieve the required pathogen destruction. To accomplish this, the first stage is operated as an SBR with three periods: fill, hold, and discharge. Because all solids added must be held in the reactor one day, the cycle time must be three days, with fill occurring on the first day, hold on the second day, and discharge to the second stage on the third day. Because fill only occurs on one day of the cycle, to process sludge on a continuous basis the first stage must have three reactors in parallel, with one receiving sludge, one holding, and one discharging on any given day. The volume of the reactors in the first stage is determined by the desired SRT for the stage and is determined using the principles in Section 7.9.2. Other process options are also available and in various levels of development, as discussed previously.

High-rate anaerobic processes are applied most often for the treatment of moderate to high strength wastewaters (those with biodegradable COD concentrations up to about 20,000 mg/L) containing mostly soluble organic matter. It is estimated that about 2000 applications existed worldwide in 65 countries by the year 2001, with significant numbers of applications existing in Europe, North and South America, and Asia.[24] The principal industries served include breweries and beverages, distilleries and fermentation, chemical, pulp and paper, food, and landfill leachate. Nearly 80% of these facilities are UASB type processes, including hybrid and EGSB systems. While the majority of existing facilities use UASB processes, EGSB systems are increasing in popularity and represent the majority of the facilities installed in recent years. This is because of the increased VOL possible with EGSB processes, averaging 20 kg COD/(m^3·day) in comparison to 10 kg COD/(m^3·day) for conventional UASB processes.

If a wastewater contains solids that require stabilization, two options exist for handling them if a high-rate anaerobic process is to be used to treat the wastewater; they can be removed either before or after treatment. An advantage of the latter option is that solids produced during anaerobic treatment will also be removed, but some systems are not amenable to that alternative. For example, prior removal is generally required for AF and EGSB systems, in which case the removed solids can be processed in a separate anaerobic digester designed specifically for that purpose. Sedimentation is also often used prior to UASB and hybrid UASB/AF systems, but a larger bioreactor could be provided to allow operation at a lower organic loading rate to encourage hydrolysis and stabilization of the solids that accumulate. Stabilization of the accumulated solids can be further encouraged by periodically diverting wastewater away from the bioreactor to allow them to digest and/or by periodically increasing the bioreactor temperature to increase the rates of hydrolysis and acidogenesis. This may be particularly attractive if wastewater is not continuously applied to the treatment system.

Fermentation of organic solids to produce VFAs is well developed for BNR systems receiving wastewaters with insufficient VFAs. The proceedings of a workshop sponsored by the Water Environment Federation provides a summary of current experience.[83]

14.2 FACTORS AFFECTING PERFORMANCE

Many factors affect the performance of anaerobic treatment systems. They range from process loading factors such as the SRT, VOL, and hydraulic loading rate; to environmental factors such

as temperature, pH, nutrient supply, and the presence of toxics; to operational factors such as mixing and the characteristics of the waste being treated. Historically, the stability and performance of anaerobic treatment systems have been considered to be poor in comparison to aerobic systems. However, with improved understanding of the factors that affect their performance, it has been possible to obtain stable and reliable performance. Consequently, a thorough understanding of these factors is critical to successful design and operation.

14.2.1 Solids Retention Time

The role of the SRT in controlling the performance of anaerobic processes was discussed briefly in Chapter 10 and has been referred to in the previous sections of this chapter. Solids retention time controls the types of microorganisms that can grow in the process and the extent to which various reactions will occur. While SRT is the fundamental control parameter, it is difficult to routinely determine it in some anaerobic processes. Determination of the SRT is straightforward in flow-through systems such as anaerobic digesters, where it simply equals the HRT. In some instances solids are separated and recycled back to the digester to increase the SRT relative to the HRT. A mechanical separation system such as a centrifuge or sludge flotation is generally used due to the nature of the solids in anaerobic digesters. Often referred to as "recuperative thickening," this approach is not generally used for routine operation. Rather, recuperative thickening is provided for unusual operating conditions, such as unusually high process loadings or the removal of one of several parallel units from service for maintenance. In the latter case, recuperative thickening allows the remaining units to maintain sufficient capacity to treat the influent solids. Standby thickening equipment can often be used for this purpose. The SRT can also be measured and controlled in UASB and hybrid UASB/AF systems, but more often solids are simply wasted to maintain a set level for the granular and flocculent sludge layers. While it is possible to determine the biomass concentration in pilot-scale AF systems by removing sections of the media, this is not a practical approach for routine operation of full-scale systems. Thus, process control for some systems is achieved by controlling the VOL, as discussed in the next section.

When the SRTs in pilot-scale anaerobic treatment systems are calculated, it is not unusual to find values of 30 to 40 days, with some systems ranging up to over 100 days.[29,38,73] Such values are significantly higher than required for wastewater treatment and represent the accumulation of excess biomass. Experience indicates that very stable performance can be obtained from some anaerobic treatment systems, particularly if long SRTs are used. It also indicates that anaerobic systems can be shut down for extended periods of time (up to several months) and that good performance can be restored shortly after they are restarted.[29,38,73] In spite of these desirable features, it is possible that these long SRTs represent underloaded systems that could have been constructed more economically using shorter SRTs, while still achieving acceptable performance.

One benefit of increased SRTs is increased hydrolysis and stabilization of particulate organic matter. This can be particularly important for the stabilization of certain types of wastewater solids. More information will be provided on this topic in Section 14.2.9.

14.2.2 Volumetric Organic Loading Rate

Even though the VOL is not a fundamental parameter determining the performance of anaerobic treatment systems, it is related to the SRT through the active biomass concentration in the bioreactor. It is also a relatively easy parameter to calculate, and it has been used historically to characterize the loading on anaerobic treatment systems. Knowledge of the VOLs that can typically be achieved for a particular process quantifies how effectively the bioreactor volume is being utilized. Used in this fashion, the VOL provides useful information for the design and

operation of anaerobic processes. The volumetric organic loading rate, $\Gamma_{V,S}$, can be calculated in units of kg COD/(m^3·day) as

$$\Gamma_{V,S} = \frac{F(S_{SO} + X_{SO})}{V}, \tag{14.1}$$

where $(S_{SO} + X_{SO})$ is the influent wastewater strength in g COD/L (kg COD/m^3), F is the influent wastewater flow rate in m^3/day, and V is the bioreactor volume in m^3. Substitution of Equation 4.15 into Equation 14.1 relates the VOL to the HRT, τ:

$$\Gamma_{V,S} = \frac{S_{SO} + X_{SO}}{\tau}. \tag{14.2}$$

This shows that the VOL is inversely proportional to the HRT, as illustrated in Figure 14.8. As discussed in Section 14.1, VOLs typically range from 2 to 40 kg COD/(m^3·day) for high-rate processes.

The SRT (Θ_c) is defined by Equation 5.1, just as for all other biochemical operations. However, we saw in Section 10.4.1 that it is often convenient to use the net process yield to relate the biomass inventory to the mass input rate of substrate and the SRT. Rearranging that equation and expressing the biomass concentration (X_M) and net yield (Y_n) on a VSS basis gives:

$$\Theta_c = \frac{X_{M,V} \cdot V}{Y_{n,V} \cdot F(S_{SO} + X_{SO})}. \tag{14.3}$$

Combining Equations 14.2 and 14.3 gives:

$$\Theta_c = \frac{X_{M,V}}{Y_{n,V} \cdot \Gamma_{V,S}}. \tag{14.4}$$

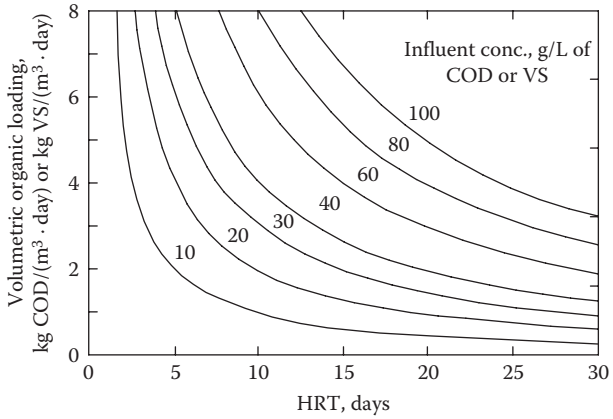

FIGURE 14.8 Effects of HRT and influent wastewater concentration on the volumetric organic loading of an anaerobic process.

Thus, it can be seen that the SRT and the VOL are inversely proportional to each other. Equation 14.4 also shows that for a fixed SRT, the VOL is increased as the biomass concentration is made larger, thereby allowing the bioreactor to be made smaller.

A similar approach is used for the solids stabilization systems, such as anaerobic digesters, except that the VOL is expressed in terms of the mass of volatile solids applied, rather than COD, typically having units of kg VS/(m^3·day). Figure 14.8 also presents the relationship between the anaerobic process HRT, the influent volatile solids concentration, and the resulting volatile solids VOL. For single stage, anaerobic digestion processes (i.e., systems operated without solids recycle), the HRT and the SRT are identical. In these instances, the volatile solids VOL simply indicates how effectively the digester volume is being utilized. Volatile solids VOLs typically range from 2 to 6 kg VS/(m^3·day).[49,63,77,85]

Interestingly, experience indicates that a maximum COD stabilization activity of 1 kg COD/(kg VSS·day) is achieved in a wide variety of anaerobic treatment processes.[29] Although higher values have been reported, especially in conjunction with the treatment of wastewaters rich in acetate, this value can be used to develop an initial estimate of the capability of a particular anaerobic process to stabilize organic matter.

14.2.3 Total Hydraulic Loading

In contrast to the suspended growth systems considered in Part II and the rest of Part III, some of the high-rate anaerobic processes are influenced by the total hydraulic loading (THL) applied to them. This is characteristic of the attached growth processes considered in Parts IV and V, and a detailed discussion of the effects of THL on them is presented in Chapters 17, 18, 19, and 21. This section presents the most important impacts of THL on UASB, AF, and hybrid UASB/AF processes.

The THL (Λ_H) is simply the total flow applied to the bioreactor (including recirculation) divided by the bioreactor cross-sectional area perpendicular to the flow. It is calculated as

$$\Lambda_H = \frac{F + F_R}{A_c}, \tag{14.5}$$

where F_R is the recirculation flow rate and A_c is the cross-sectional area. The THL is a superficial velocity (i.e., a theoretical velocity based on the empty bed cross-sectional area).

The THL affects process performance in several ways. For upflow processes with sludge blankets, such as UASB and hybrid UASB/AF systems, maximum allowable values of the THL correspond to the settling velocity of the particles to be retained in the bioreactor. If the THL exceeds these values, the particles will be washed out of the bioreactor. As a result, the desired biomass inventory and associated SRT cannot be maintained and the process will fail.

For UASB and hybrid UASB/AF processes, the maximum allowable THL depends on the nature of the solids developing in the bioreactor.[46,47] For granular solids, the daily average THL should not exceed 72 m/day when treating fully soluble wastewater, and 24 to 30 m/day when treating partially soluble wastewater. The THL can be temporarily increased to 144 m/day for fully soluble wastewaters and 48 m/day for partially soluble wastewaters. For flocculent solids, the daily average THL should not exceed 12 m/day and the maximum THL should not exceed 48 m/day. The factors that led to development of granular versus flocculent solids are discussed in Section 14.2.9. Knowledge concerning appropriate THL values for specific wastewaters continues to evolve.

For AF processes a minimum THL is needed to achieve uniform distribution of flow across the bioreactor cross section to minimize short-circuiting. Values in the range of 10 to 20 m/day appear to be appropriate.[87] As with the UASB and hybrid UASB/AF processes, THL criteria for specific wastewaters continue to evolve.

Total hydraulic loading constraints can affect the configuration of the anaerobic bioreactor. The bioreactor cross-sectional area must be adjusted to produce THL values within the necessary range. In some instances, recirculation must be initiated to maintain the minimum required THL. The impacts of the THL constraints on the design of anaerobic processes are discussed briefly in Section 14.3.2.

14.2.4 Temperature

As with all biological processes, the performance of anaerobic processes is significantly affected by operating temperature. Best performance is typically obtained by operation in the optimal region of one of the two temperature ranges (i.e., 30 to 40°C for mesophilic or 50 to 60°C for thermophilic), and most anaerobic processes are designed to do so. These two regions generally represent the optima for growth of the methanogens. Nevertheless, it is possible to grow methanogens at lower temperatures, provided that longer SRTs are used to compensate for the lower maximum specific growth rates. Although anaerobic activity can be sustained at temperatures approaching 10°C, operating temperatures in the 20 to 25°C range currently appear to be the lower limit from a practical perspective,[42,49,52,63,66,73] although recent developments may extend the practical range.[59,81]

Although the preceding paragraph focused on methanogens, operating temperature affects hydrolytic and acidogenic reactions as well. For wastewaters consisting largely of simple, readily biodegradable organic matter, the effect of temperature on methanogenesis is the primary concern. However, for wastewaters consisting largely of complex organic compounds or particulate materials, the effects of temperature on hydrolysis and acidogenesis will be the primary concern. Table 14.4 presents $\hat{\mu}$ and K_S values for biodegradation of VFAs at temperatures of 25, 30, and 35°C. These data may be used to characterize the impact of temperature on the anaerobic biodegradation of simple organic compounds.

Figure 14.9 shows the combined effects of SRT and temperature on the anaerobic digestion of municipal primary solids. Essentially complete stabilization of biodegradable volatile solids is achieved at an SRT of 10 days when operating at a temperature of 35°C. A moderate increase in SRT to about 15 days is required when operating at a temperature of 25°C, but the stabilization is not complete, as indicated by a residual VS concentration at SRT values as long as 60 days. The required SRT increases to about 25 days when operating at a temperature of 20°C, and a higher residual VS concentration is observed. At 15°C an SRT of about 30 days is required to obtain stable operation, and only about one-half of the biodegradable volatile solids are destroyed at SRTs as long as 60 days. The curves showing the correspondence between VS destruction and methane

TABLE 14.4
Average Values of Kinetic Parameters for Anaerobic Enrichment Cultures Grown on Various Volatile Fatty Acids

	35°C		30°C		25°C	
Volatile Fatty Acid	$\hat{\mu}$ day^{-1}	K_S mg/L as COD	$\hat{\mu}$ day^{-1}	K_S mg/L as COD	$\hat{\mu}$ day^{-1}	K_S mg/L as COD
Acetic	0.36	165	0.26	356	0.24	930
Propionic	0.31	60	—	—	0.38	1145
Butyric	0.38	13	—	—	—	—

Note: Adapted from Lawrence, A. W., Application of process kinetics to design of anaerobic processes. Anaerobic Biological Treatment Processes, American Chemical Society Advances in Chemistry Series, 105:163–89, 1971.

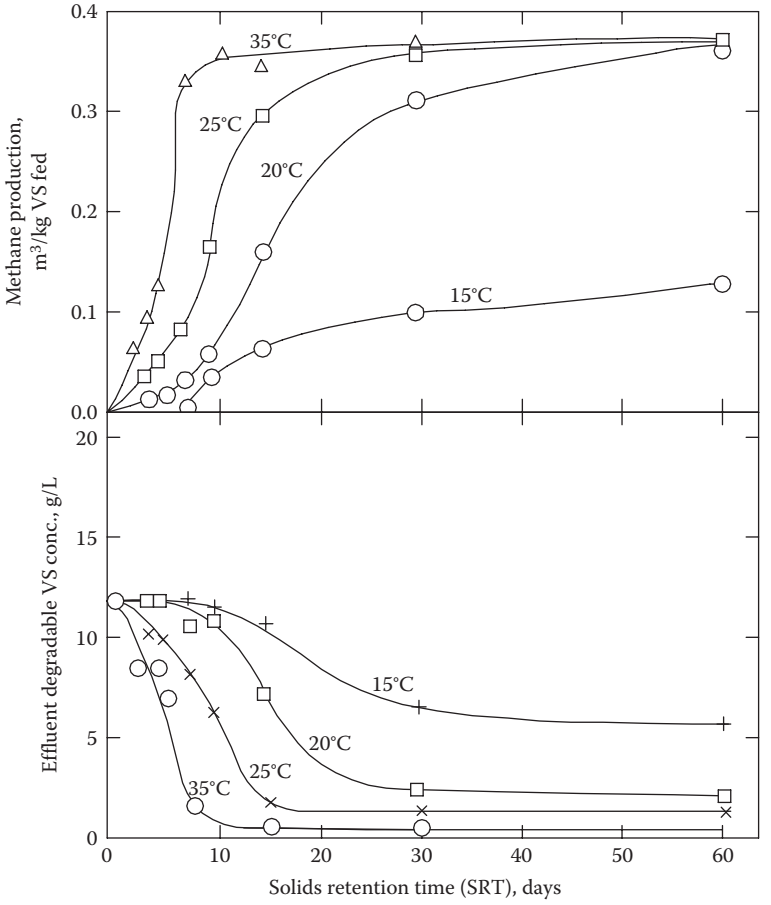

FIGURE 14.9 Effects of SRT and temperature on the anaerobic digestion of municipal primary solids. (From Lawrence, A. W., Application of process kinetics to design of anaerobic processes. *Anaerobic Biological Treatment Processes*, American Chemical Society Advances in Chemistry Series, 105:163–89, 1971. Copyright © American Chemical Society. Reprinted with permission.)

production suggest that hydrolysis of the solids is generally the rate limiting step at these temperatures. Taken together, these data suggest that a temperature of about 25°C is the practical minimum for the anaerobic stabilization of municipal primary solids.

Temperature variations are also of concern, and it is typically recommended that systems be designed and operated to achieve variations of less than ±1°C each day.[13,76,77,84] Some research indicates that anaerobic processes are capable of reacting successfully to temperature variations; although reaction rates decrease when the temperature is reduced, activity is restored quickly when the temperature returns to the optimum value. In contrast, experience with full-scale systems indicates that performance is adversely impacted by rapid temperature variations of as little as 2 to 3°C. This may be because of factors such as mixing and stratification within the bioreactor. Regardless of the mechanism, it appears prudent to adhere to recommended practice and to design and operate anaerobic processes to minimize short-term temperature variations.

Although opinions have historically varied concerning the benefits of operation under thermophilic conditions,[7,63,77,85] it is clear today that the principal benefit is increased inactivation of pathogenic organisms, which increases the options for disposing of treated solids. Potential drawbacks include the increased energy required to achieve thermophilic operating temperatures, increased odors, and decreased process stability. Decreased process stability because of increased VFA concentrations,

increased sensitivity to temperature variations, increased ammonia toxicity, increased foaming, and increased odor potential are all areas of concern. Because of its uncertain benefits and numerous drawbacks, designs based on thermophilic operation should be approached with caution unless site-specific pilot test results and/or full-scale experience are available. As discussed above, advanced anaerobic digestion systems are being intensively investigated and the evolving knowledge about them can be used for their implementation.

14.2.5 pH

Like all biochemical operations, pH has a significant impact on the performance of anaerobic processes, with activity decreasing as the pH deviates from an optimum value. This effect is particularly significant for anaerobic processes because the methanogens are affected to a greater extent than the other microorganisms in the microbial community.[29,32,49,53,63,65,66] As a consequence, there is a greater decrease in methanogenic activity as the pH deviates from their optimum value. A pH range of 6.8 to 7.4 generally provides optimum conditions for the methanogens, whereas a pH between 6.4 and 7.8 is considered necessary to maintain adequate activity. The pH will also affect the activity of the acidogenic bacteria; however, the effect is less significant and primarily influences the nature of their products. In a single-stage system, a decrease in pH increases the production of higher molecular weight VFAs, particularly propionic and butyric acid, at the expense of acetic acid. As discussed in Section 2.3.3, one mechanism causing this is the buildup of H_2 in the system. As its utilization by methanogens is slowed, it begins to accumulate, which then slows down the production of acetic acid by the acidogens and shifts their metabolism toward other VFAs. In the acidogenic reactor of a two-phase system, however, lower pH values do not favor the production of propionic or butyric acid over acetic acid. Rather pH values down to 5.5 favor acetic acid formation.[5] The activity of the hydrolytic microorganisms is affected the least by pH deviations from neutrality.

The pH sensitivity of the methanogens, coupled with the fact that VFAs are intermediates in the stabilization of organic matter, can result in an unstable response by single-stage anaerobic systems to a decrease in pH.[29,63,76,77] The unstable response may be triggered by a high VOL that results in an increase in the production of VFAs by the acidogenic bacteria. If the increased VFA production rate exceeds the maximum capacity of the methanogens to use acetic acid and H_2, excess VFAs will begin to accumulate, decreasing the pH. The decreased pH will reduce the activity of the methanogens, thereby decreasing their use of acetic acid and H_2, causing a further accumulation of VFAs and a further decrease in the pH. If this situation is left uncorrected, the result is a precipitous decrease in the pH, the accumulation of higher molecular weight VFAs, and a near cessation of methanogenic activity. This condition is known as a "sour" or "stuck" anaerobic process. It can be corrected in its early stages by resolving the environmental factors causing the imbalance between the acidogenic bacteria and the methanogens. In the case considered above, this could be accomplished by reducing the VOL to the point where the VFA production rate is less than their maximum consumption rate. This will allow consumption of the excess VFAs in the system, thereby causing the pH to return to neutrality and the activity of the methanogens to increase. The VOL can then be increased as the process recovers until the full loading capability is utilized. In extreme cases, decreases in loading must be coupled with the addition of chemicals for pH adjustment, as discussed below.

For an anaerobic process functioning within the acceptable pH range, the pH is controlled primarily by the bicarbonate buffering system. Bicarbonate alkalinity is produced by the destruction of nitrogen-containing organic matter and the reaction of the released ammonia-N with the carbon dioxide produced in the reaction. This is illustrated by Equation 14.6 for the conversion of primary solids (represented as $C_{10}H_{19}O_3N$) to methane, carbon dioxide, biomass, and ammonium bicarbonate:[63]

$$\begin{aligned} C_{10}H_{19}O_3N + 4.69\ H_2O \rightarrow 5.74\ CH_4 + 2.45\ CO_2 \\ + 0.2\ C_5H_7O_2N + 0.8\ NH_4HCO_3. \end{aligned} \tag{14.6}$$

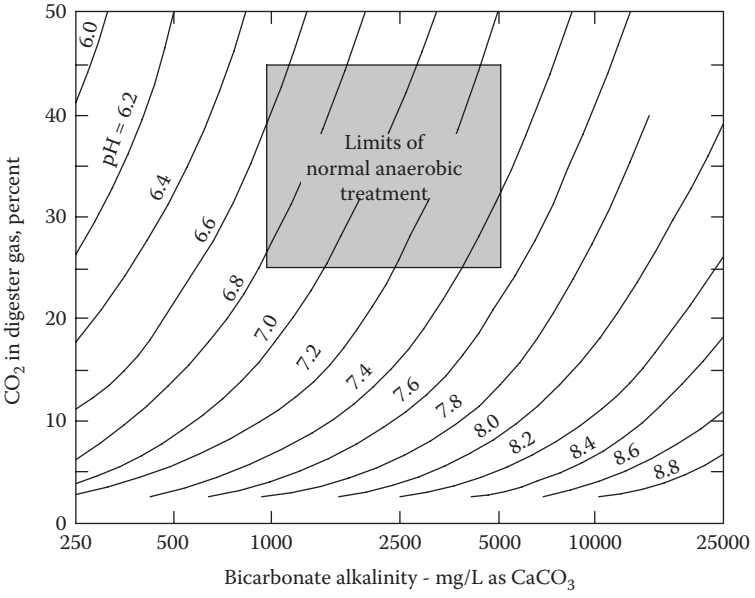

FIGURE 14.10 Effect of pH on the relationship between the bicarbonate alkalinity of the liquid phase and the carbon dioxide content of the gas phase in an anaerobic process. (From Parkin, G. F. and Owen, W. F., Fundamentals of anaerobic digestion of wastewater sludges. *Journal of the Environmental Engineering Division, ASCE,* 112:867–920, 1986. Copyright © American Society of Civil Engineers. Reprinted with permission.)

As illustrated, bicarbonate alkalinity is produced in direct relation to the ammonia-N released. A strong base is needed to react with the carbon dioxide produced in the system to form the bicarbonate. In most instances ammonia is the strong base, although the cations associated with soaps or the salts of organic acids can also serve to maintain electroneutrality in the reaction with carbon dioxide.

The concentration of bicarbonate alkalinity in solution is related to the carbon dioxide content of the gas space in the bioreactor and the bioreactor pH:

$$S_{BAlk} = 6.3 \times 10^{-4} \left(\frac{\overline{p}_{CO_2}}{10^{-pH}} \right), \tag{14.7}$$

where S_{BAlk} is the bicarbonate alkalinity expressed as mg/L as $CaCO_3$ and $\overline{p}_{CO_2}$ is the partial pressure of carbon dioxide in the gas space expressed in atmospheres.[9,51,63] This relationship is presented in Figure 14.10 and illustrates that typical anaerobic processes operate with bicarbonate alkalinities in the range of 1000 to 5000 mg/L as $CaCO_3$ and carbon dioxide partial pressures of 25 to 45%.

When VFAs begin to accumulate in an anaerobic process, they are neutralized by the bicarbonate alkalinity present. For example, consider acetic acid. Acetic acid is released by the acidogenic bacteria in nonionized form, but exists as acetate ion at neutral pH. The reaction of acetic acid with bicarbonate alkalinity to convert it to acetate is

$$HCO_3^- + HAc \rightleftarrows H_2O + CO_2 + Ac^-, \tag{14.8}$$

where HAc represents nonionized acetic acid and Ac^- represents acetate ion. When a pH end point of 4.0 is used in the alkalinity analysis, acetate will be partially converted to acetic acid and will, therefore, register as alkalinity. Thus, if VFAs are present, the total alkalinity will represent the

concentration of both bicarbonate ion and VFAs. If the concentration of VFAs is known and is expressed as acetic acid, the bicarbonate alkalinity can be calculated from the total alkalinity as

$$S_{BAlk} = S_{TAlk} - 0.71(S_{VFA}), \tag{14.9}$$

where S_{TAlk} is the total alkalinity expressed as $CaCO_3$ and S_{VFA} is the concentration of VFAs expressed as acetic acid. The factor 0.71 converts the VFA concentration expressed as acetic acid to $CaCO_3$ and corrects for the fact that approximately 85% of the VFA anions are titrated to the acid form at a pH of 4.0.[51,63] Other organic and inorganic bases, such as sulfides, can also be titrated to their acid form and, consequently, measured as alkalinity. The concentrations of these anions are typically small relative to the bicarbonate concentration, but the potential for such interferences with bicarbonate alkalinity measurement should be recognized.

As discussed above, under stable operating conditions, bicarbonate is the primary form of alkalinity in anaerobic processes. However, under unstable operating conditions VFAs will react with bicarbonate alkalinity, both reducing its concentration and producing carbon dioxide (see Equation 14.8), which increases the carbon dioxide content of the gas space. Reference to Figure 14.10 illustrates that both of these changes act to decrease the pH in the bioreactor. Stable operation of anaerobic processes is generally achieved by the maintenance of a relatively high concentration of bicarbonate alkalinity so that increased VFA production can be tolerated with a minimal decrease in bioreactor pH.

Adverse pH conditions can be corrected by the addition of appropriate chemicals, but care must be exercised in their selection because of the complex interactions that can occur and the potential for adding toxicants. Commonly used chemicals include sodium bicarbonate, sodium carbonate, lime, sodium or potassium hydroxide, and ammonia.

Sodium bicarbonate is preferred for pH adjustment because its impact is longer lasting and its toxicity potential is low. It adjusts the pH by the direct addition of bicarbonate ions that, as illustrated in Figure 14.10, will result in a direct increase in the pH without affecting the carbon dioxide content of the gas space.

The addition of hydroxide ions by adding lime, sodium hydroxide, or potassium hydroxide, adjusts the pH because the hydroxide ion reacts with carbon dioxide to form bicarbonate alkalinity. Using lime as the example pH adjustment chemical gives:

$$Ca(OH)_2 + 2CO_2 \rightleftarrows Ca^{+2} + 2HCO_3^-. \tag{14.10}$$

This reaction is accompanied by a decrease in the carbon dioxide content of the gas space, which further contributes to the rise in the bioreactor pH. Unfortunately, further production of carbon dioxide by the microorganisms in the process will restore the original gas space carbon dioxide content and reduce the pH.

The use of carbonate based chemicals reduces the magnitude of the pH variation, as follows:

$$Na_2CO_3 + H_2O + CO_2 \rightleftarrows 2Na^+ + 2HCO_3^-. \tag{14.11}$$

Comparison to Equation 14.10 illustrates that only one mole of carbon dioxide is required to produce two moles of bicarbonate from carbonate while two moles of carbon dioxide are required to produce two moles of bicarbonate from hydroxide. Thus, when carbonate-based chemicals are used for pH adjustment, the immediate consumption of carbon dioxide from the gas space is one-half of that when hydroxide-based chemicals are used.

These changes in pH are illustrated in Figure 14.11. Consider an initial condition represented by a gas phase carbon dioxide content of 40% and a bicarbonate alkalinity of 500 mg/L as $CaCO_3$ (point 1), which corresponds to a pH of about 6.3. The addition of sufficient sodium bicarbonate to elevate the bioreactor bicarbonate alkalinity to 2100 mg/L as $CaCO_3$ would directly increase the

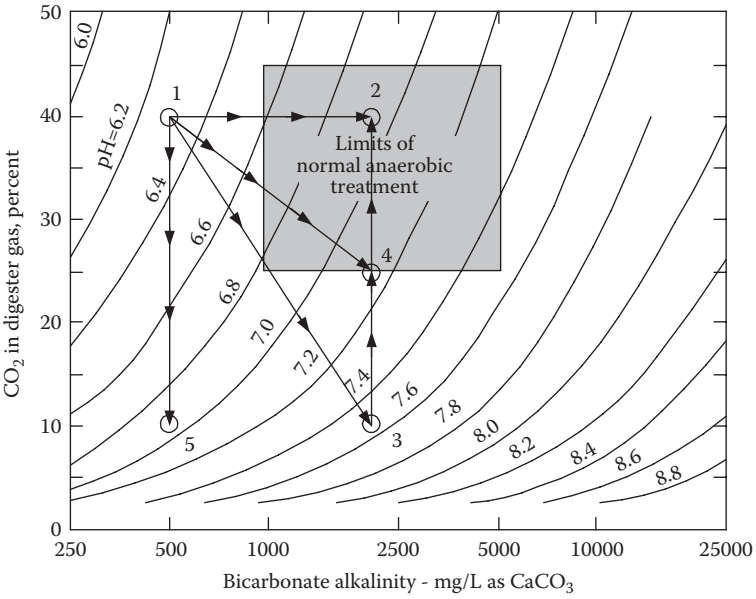

FIGURE 14.11 Illustration of the effects of changes in the bicarbonate alkalinity of the liquid phase and/or the carbon dioxide content of the gas phase on the pH in an anaerobic process.

bioreactor pH to about 6.9 (point 2). The addition of an equivalent amount of the hydroxide-based chemical will result in not only an increase in the bicarbonate alkalinity to 2100 mg/L as $CaCO_3$, but also in an immediate decrease in the carbon dioxide content of the gas space as it is removed to produce the bicarbonate (Equation 14.10). The actual decrease in the carbon dioxide content of the gas will depend on the relative gas and liquid volumes in the bioreactor. If the requirement for carbon dioxide is large relative to the amount available, a negative pressure can be created, causing air to be drawn into the gas space, creating an explosive mixture of methane and oxygen. Furthermore, under extreme conditions, removal of carbon dioxide can cause a sufficiently strong negative pressure to collapse the structure. However, for the purposes of this example a decrease to 10% is assumed, and no other adverse consequences are experienced. This results in a pH of approximately 7.5 immediately after addition of the chemical (point 3). However, as additional carbon dioxide is produced by the biomass, the carbon dioxide content of the gas space will increase to its equilibrium value of 40% and the pH will decrease to 6.9, as illustrated by line 3-2 in Figure 14.11. This dramatic variation in pH, from 6.3 (point 1) to 7.5 (point 3) to 6.8 (point 2), can be detrimental to the process. Moreover, it makes pH control difficult from an operational perspective because the relationship between chemical addition and the resulting pH is not straightforward. The effect of adding a carbonate chemical, such as sodium carbonate, is illustrated by point 4 in Figure 14.11, where it is observed that the pH immediately after its addition will be about 7.1. The pH will decrease to 6.9 as carbon dioxide is produced and the carbon dioxide content of the gas is increased back to 40% (point 2). Thus, it can be seen that the addition of carbonate chemicals causes less drastic swings in pH than the addition of hydroxide chemicals.

The cations associated with pH adjustment chemicals can also impact an anaerobic process. Use of lime increases the calcium concentration and if it becomes too high, calcium carbonate can precipitate and reduce the effective volume of the bioreactor. As discussed in Section 14.2.6, sodium, potassium, and calcium are toxic if concentrations become high enough. Ammonia can also be used to adjust the pH because it reacts with carbon dioxide to form ammonium bicarbonate:

$$NH_3 + H_2O + CO_2 \rightleftarrows NH_4HCO_3. \tag{14.12}$$

However, this results in the same variations in pH produced by the addition of hydroxide chemicals. Moreover, as discussed in Section 14.2.6, ammonia is also toxic at high concentrations. Thus, it can be seen that no ideal pH adjustment chemical exists and that some degree of care is required when any of them are used.

The control of pH by the removal of carbon dioxide from the gas space has been suggested.[28] As illustrated by point 5 in Figure 14.11, this would require reduction of the carbon dioxide content of the gas to about 10% on a sustained basis to achieve the same pH as achieved by increasing the bicarbonate alkalinity to 2100 mg/L as $CaCO_3$. This would require continuous removal of carbon dioxide from the gas phase.

14.2.6 Inhibitory and Toxic Materials

As discussed in Section 14.1.1, one characteristic of anaerobic processes is their sensitivity to inhibition by chemicals present in the wastewater or produced as process intermediates. As discussed in Section 3.2.7, inhibition causes a reduction in the maximum specific growth rate of microorganisms, thereby requiring an increase in the SRT of a biochemical operation to produce the same effluent that would be produced in the absence of the inhibitor. However, if the inhibitor concentration is increased sufficiently, a toxic response is exhibited and the microorganisms are killed, thus causing total process failure. Unfortunately, the literature has not always made a clear distinction between inhibition and toxicity. Consequently, in the information that follows the two terms should not be interpreted strictly. However, it should be recognized that, in general, inhibition precedes toxicity as the concentration of a compound is increased. Several inorganic materials can cause an inhibitory response; the materials of greatest concern are light metal cations, ammonia, sulfide, and heavy metals. In addition, sulfate interferes with methane production by providing an alternate electron acceptor. Not only is the sulfide produced an offensive and dangerous gas, but soluble sulfide exerts an oxygen demand that reduces the amount of COD stabilized. Finally, many organic compounds are also inhibitory, particularly to methanogens.

14.2.6.1 Light Metal Cations

The light metal cations include sodium, potassium, calcium, and magnesium. They may be present in the influent, released by the breakdown of organic matter (such as biomass), or added as pH adjustment chemicals. They are required for microbial growth and, consequently, affect specific growth rate like any other nutrient. Consequently, they must be available if anaerobic treatment is to occur. Nevertheless, their inhibitory nature has been known for over four decades.[51,73] While moderate concentrations stimulate microbial growth, excessive amounts slow it, and even higher concentrations can cause severe inhibition or toxicity. Table 14.5 indicates the concentration ranges over which these various responses occur.

The light metal cations exhibit complex interactions in their effects on microbial growth.[51,63,77] For example, inhibition can be increased when two light metal cations are present at their moderately inhibitory concentrations. This is known as a synergistic response because the combined effects of the two light metal cations exceeds that of either individually. Secondly, the inhibition caused by one light metal cation can be increased if the other light metal cations are present at concentrations below their stimulatory concentrations. Finally, the presence of one light metal cation at its stimulatory concentration can reduce the inhibition of another. This phenomenon is known as antagonism, since the effect is reduced. Table 14.6 summarizes antagonistic responses for the light metal cations and ammonia.

14.2.6.2 Ammonia

Ammonia-N is a required nutrient and stimulates bacterial growth at low concentrations. For anaerobic processes, ammonia concentrations between 50 and 200 mg/L as N are generally within the stimulatory range.[51,63] However, ammonia is inhibitory at higher concentrations and

TABLE 14.5
Stimulatory and Inhibitory Concentrations of Light Metal Cations

	Concentration, mg/L		
Cation	Stimulatory	Moderately Inhibitory	Strongly Inhibitory
Sodium	100–200	3500–5500	8000
Potassium	200–400	2500–4500	12,000
Calcium	100–200	2500–4500	8000
Magnesium	75–150	1000–1500	3000

Note: Adapted from McCarty, P. L., Anaerobic waste treatment fundamentals. Public Works, 95 (9): 107–12; (10): 123–26; (11): 91–94; (12): 95–99, 1964.

TABLE 14.6
Antagonistic Responses for Light Metal Cations and Ammonia

Inhibitor	Antagonist
Na^+	K^+
K^+	Na^+, Ca^{+2}, Mg^{+2}, NH_4^+
Ca^{+2}	Na^+, K^+
Mg^{+2}	Na^+, K^+

Note: Adapted from Kugelman, I. J. and Chin, K. K., Toxicity, synergism, and antagonism in anaerobic waste treatment processes. Anaerobic Biological Treatment Processes, American Chemical Society Advances in Chemistry Series, 105:55–90, 1971.

toxic if the concentration is high enough. Ammonia may be present in the influent wastewater, or it may be formed as a result of the breakdown of organic materials that contain nitrogen, such as proteins. The production of ammonia by the breakdown of primary solids is illustrated in Equation 14.6.

Ammonia is a weak base and dissociates in water:

$$NH_3 + H_2O \rightleftarrows NH_4^+ + OH^-. \tag{14.13}$$

Free ammonia (NH_3) is the principal inhibitory species and can cause a toxic response at concentrations of about 100 mg/L as N.[63] Ammonium ion (NH_4^+) concentrations as high as 7000 to 9000 mg/L as N have been successfully treated without a toxic response with an acclimated culture,[63] although concentrations as low as 1500 mg/L as N have been reported to be toxic.[51] This type of response has been observed for many materials that can ionize and the nonionized species is often the more inhibitory of the two. As noted in Table 14.6, ammonium ion is also an antagonist for inhibition by potassium.

The pK_a for the dissociation of ammonia is approximately 9.3, so ammonia is present primarily as the ionized species at the pH values typically occurring in anaerobic processes. However, if the total ammonia ($NH_3 + NH_4^+$) concentration is high enough, a sufficient concentration of free ammonia

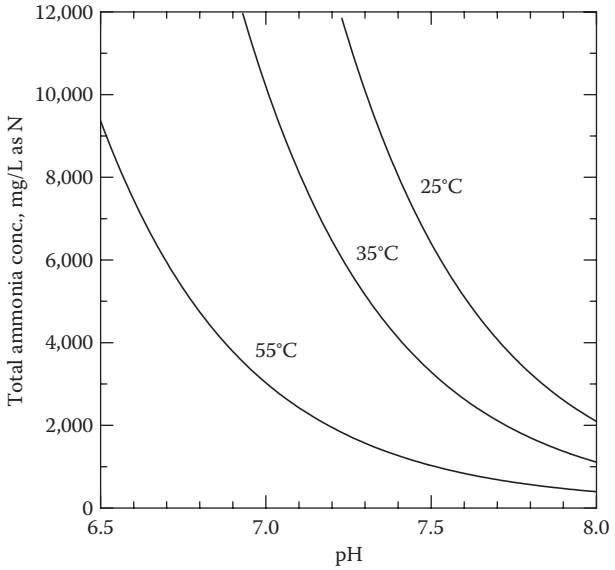

FIGURE 14.12 Effects of pH and temperature on the total ammonia-N concentration necessary to give a free ammonia concentration of 100 mg/L as N. The curves were generated from equilibrium and thermodynamic constants given in *Water Chemistry.* (Snoeyink, V. L. and Jenkins, D., John Wiley & Sons, Inc., New York, 1980).

can be present to cause an inhibitory or toxic response. The proportion of total ammonia that is present as free ammonia increases with both pH and temperature. As illustrated in Figure 14.12, vastly different total ammonia concentrations can result in a toxic free ammonia concentration (100 mg/L as N), depending on the pH and temperature. For mesophilic conditions (25 and 35°C), the total ammonia concentration can exceed 10,000 mg/L as N and the free ammonia concentration will still be below 100 mg/L as N at pH values of about seven. However, for thermophilic conditions (55°C), the total ammonia concentration must be maintained below 2000 mg/L as N to keep free ammonia concentrations below toxic levels. Even for mesophilic operating conditions, total ammonia concentrations of about 2000 mg/L can result in toxic free ammonia concentrations (100 mg/L as N) as the pH approaches 7.5 to 8.0. In some instances, systems can acclimate to the higher total and nonionized ammonia concentrations.

Three strategies are available for reducing ammonia inhibition in anaerobic processes: reduce the temperature, reduce the pH, or reduce the total ammonia concentration. As discussed in the preceding paragraph, Figure 14.12 illustrates the impact of mesophilic versus thermophilic operating temperatures on allowable total ammonia concentrations. This difference should be carefully considered when selecting the temperature range. Moreover, a reduction in operating temperature within a temperature range must also be carefully evaluated. For example, reducing the temperature from 35 to 25°C causes a noticeable increase in the allowable total ammonia concentration. However, a temperature decrease of this magnitude would also result in a significant reduction in the maximum specific growth rate of the anaerobic biomass, which may be more detrimental to the process than the reduced inhibition associated with the reduction in free ammonia concentration. As further indicated in Figure 14.12, if pH values are relatively high, significant reductions in ammonia toxicity can result from their decrease. High pH values can occur when high strength wastewaters or solids are treated because of the high concentrations of bicarbonate alkalinity that result. If such wastewaters also contain high concentrations of ammonia or organic nitrogen, then the concentration of total ammonia will be elevated as well. The pH in such a bioreactor can be decreased by the addition of an acid. Hydrochloric acid is the ideal chemical for this purpose because chloride ion

has little or no impact on anaerobic biomass. In addition, the total ammonia concentration can be reduced by dilution of the wastewater or solids with clean water. Care must be exercised if this is done because the larger flow rate may compromise the SRT of the system. However, if an adequate SRT can be maintained, this approach can be used quite successfully.

14.2.6.3 Sulfide

Sulfide is produced in an anaerobic process through the reduction of sulfate present in the influent and by the degradation of sulfur-containing organic matter (e.g., proteins). Only soluble sulfides are inhibitory and concentrations greater than 200 mg/L cause strong inhibition, while concentrations up to 100 mg/L can be tolerated with little or no acclimation. Concentrations between 100 and 200 mg/L may be tolerated after acclimation.[43,44,63] Sulfide reacts with heavy metal cations, including iron, forming highly insoluble precipitates. In fact, iron sulfide gives anaerobic processes their characteristic black color. Consequently, the concentration of soluble sulfide can be reduced by the addition of iron to the bioreactor, thereby reducing sulfide inhibition.

Hydrogen sulfide is a weak acid and, consequently, at neutral pH is present in equilibrium with the sulfide anion. Hydrogen sulfide is sparingly soluble in water, so it will partition between the liquid and gas phases. Hydrogen sulfide increases the corrosivity of anaerobic process gas and results in the formation of sulfur oxides when the gas is burned. Consequently, control of the hydrogen sulfide content of the product gas is desirable. This too can be done by adding iron to the bioreactor to precipitate the sulfide anion as iron sulfide.

Sulfate itself is not inhibitory to anaerobic bacteria, but it impacts anaerobic processes by providing an electron acceptor that can be used by sulfate reducing bacteria, allowing them to compete with methanogens for the electrons available in the organic matter. This has several effects. First, it produces sulfide, which is inhibitory, as discussed above. Second, it reduces the amount of methane produced because the electrons used to reduce the sulfate are not available for the reduction of carbon dioxide to methane. Third, it reduces the value of the product gas, as discussed above. Fourth, it decreases the removal of COD from the wastewater being treated. Although the organic matter is still oxidized to carbon dioxide, much of the sulfide produced remains in the process stream, where it represents an oxygen demand. Approximately two mg of carbonaceous COD are consumed by sulfate reducing bacteria for each mg of sulfate-S reduced to sulfide-S,[29,55] but any of that sulfide that is still present in the liquid phase exerts a COD. This can have a major impact when anaerobic processes are used to treat relatively dilute wastewaters.

The competition between methanogens and sulfate reducing bacteria is very complex and is influenced by many factors, including the nature of the wastewater and the type of bioreactor being used.[10,12,37,55,64,73,74] Although it would be desirable to design an anaerobic reactor receiving a wastewater containing sulfate in such a way that the methanogens outcompete the sulfate reducing bacteria, thereby excluding the latter from the system, methods for doing this are currently not available.[37] In fact, given the relative growth kinetics of the various bacteria, it is probably not possible.[74] This means that when wastewaters containing elevated sulfate levels are to be treated anaerobically it is generally prudent to assume that all of the influent sulfate will be reduced to sulfide, provided that the influent COD/SO_4^- ratio is sufficiently high, and to make provisions to deal with the resultant sulfide levels. Some of the produced sulfide will be precipitated by heavy metals and some will partition into the gas phase. Both of these mechanisms will reduce the soluble sulfide concentration and, consequently, the potential for the development of an inhibitory sulfide concentration. Nevertheless, when the influent COD is low relative to the influent sulfate concentration, insufficient methane gas may be produced to strip the sulfide produced from the liquid phase, resulting in soluble sulfide concentrations that are inhibitory or toxic. Experience suggests that inhibitory soluble sulfide concentrations may develop when treating wastewaters with a COD/SO_4^- ratio less than about 7.5 g/g.[29] Furthermore, Hulshoff Pol et al.[37] have stated that as long as the COD/SO_4^- ratio is above 10 g/g, anaerobic treatment always proceeds successfully because the dissolved sulfide levels will be maintained below inhibitory levels. Both Hall[29] and Hulshoff Pol et al.[37]

TABLE 14.7
Soluble Heavy Metal Concentrations Exhibiting 50% Inhibition of Anaerobic Digesters

Cation	Concentration mg/L
Fe^{+2}	1–10
Zn^{+2}	10^{-4}
Cd^{+2}	10^{-7}
Cu^{+}	10^{-12}
Cu^{+2}	10^{-16}

Note: Adapted from Mosey, F. E. and Hughes, D. A., The toxicity of heavy metal ions to anaerobic digestion. Water Pollution Control, 74:18–39, 1975.

list sulfide control strategies that can be applied when the COD/SO_4^- ratio is insufficient, including elevating the bioreactor pH and temperature, adding iron salts to precipitate sulfide from solution, purging hydrogen sulfide from the bioreactor liquid, scrubbing hydrogen sulfide from the biogas and recirculating it to the bioreactor to remove sulfide, and using biological sulfide oxidation and sulfur recovery. A review of sulfur treatment technology has recently been provided.[45]

14.2.6.4 Heavy Metals

As with other biochemical operations, heavy metals have strong effects on anaerobic processes, as indicated in Table 14.7 by the low concentrations causing 50% inhibition. Fortunately, only the soluble metal ions are inhibitory and the metal sulfides are extremely insoluble, giving residual heavy metal concentrations much less than the concentrations in Table 14.7. Consequently, heavy metal inhibition to anaerobic processes is often prevented by the sulfide produced in the process. In situations where inadequate sulfide is produced, sulfur can be added. Approximately 0.5 mg of sulfide is needed to precipitate one mg of heavy metal.[49,63] Ferrous sulfide is an ideal chemical to provide supplemental sulfide. Table 14.7 shows that ferrous iron is much less inhibitory than other heavy metals. In addition, the sulfide precipitates of the more inhibitory heavy metals are more insoluble than ferrous sulfide, and consequently the added sulfide will maintain the concentration of those heavy metals at low concentrations. Furthermore, the presence of residual iron will maintain soluble sulfide concentrations at low values. Finally, as long as the pH is 6.4 or above, any excess iron will precipitate as iron carbonate, thereby preventing any inhibition caused by soluble iron.

14.2.6.5 Volatile Acids

Early evidence suggested that VFA concentrations above 2000 mg/L were inhibitory to methanogens, but when the pH was held near neutral, neither acetic nor butyric acid inhibited methane formation at concentrations up to 10,000 mg/L.[35] Propionic acid was inhibitory at a concentration of 6000 mg/L at neutral pH.[41] A number of studies have now found that very high concentrations of acetic and butyric acid can be tolerated at neutral pH, but that much lower concentrations of propionic acid are tolerated.[16] Andrews and coworkers suggested that it is the nonionized form of the VFAs that is actually inhibitory, with concentrations on the order of 30 to 60 mg/L having an effect.[2–4] Volatile fatty acids are weak acids that are largely dissociated at neutral pH. For example, a total acetic acid concentration of approximately 5500 mg/L is required to produce a nonionized acetic acid concentration of 30 mg/L at pH 7. On the other hand, at pH 6.5 a total acetic acid concentration of only 1800 mg/L produces the same nonionized acetic acid concentration,[63] showing that pH and VFA concentration are interrelated in their effects. Because of the high concentrations of VFAs tolerated, inhibition

TABLE 14.8
Concentrations of Organic Compounds Reported to be Inhibitory to Anaerobic Processes

Compound	Inhibitory Concentration mg/L
Formaldehyde	50–200
Chloroform	0.5
Ethyl benzene	200–1000
Ethylene dibromide	5
Kerosene	500
Linear ABS (detergent)	1% of dry solids

Note: Adapted from Parkin, G. F. and Owen, W. F., Fundamentals of anaerobic digestion of wastewater sludges. *Journal of Environmental Engineering*, 112:867–920, 1986.

caused by VFAs will be of little concern as long as the pH remains within the normal range for the growth of methanogens (6.8 to 7.4). For pH values below this range, pH impacts themselves will be significant and will be compounded by any inhibition caused by nonionized VFAs.

14.2.6.6 Other Organic Compounds

As with aerobic processes, a wide range of organic compounds can inhibit anaerobic processes. Also like aerobic processes, significant biodegradation of these chemicals can occur with sufficient acclimation.[6,20,21,28,63,68,73,79] Table 14.8 summarizes inhibitory concentrations of some typical organic compounds, while Table 14.9 compares the relative effects of several organic compounds on anaerobic processes. The concentration ranges presented in these tables represent the response of anaerobic cultures upon initial exposure to the compounds. However, it has been found that, with acclimation, anaerobic cultures can tolerate concentrations of 20 to 50 times those values while successfully metabolizing the compounds.[63] Table 14.10 demonstrates the biodegradative capability of anaerobic systems by summarizing petrochemical wastewater components that were inhibitory initially, but biodegradable following acclimation. During acclimation, the activity of a methanogenic community may nearly cease. However, even after long periods of inactivity (50 days or more), a community capable of degrading the target compound can develop. This suggests that some organisms survived and served as seed for the development of a healthy community capable of degrading the target compound. Procedures have been developed to assess the effects of compounds on anaerobic cultures, and they may be used to determine the concentration range over which an inhibitory response may be observed.[40,73,89] They may also be used to develop cultures capable of biodegrading a target compound.

The response of both aerobic and anaerobic processes to inhibitory organic chemicals is an area of continued research, and the reader is urged to consult the literature for ongoing developments. This topic is discussed further in Chapter 22.

14.2.7 Nutrients

Like all other biochemical operations, nutrients are required by anaerobic processes because they are essential components of the biomass produced. However, biomass yields are much lower in anaerobic processes than in aerobic ones, and this results in reduced nutrient requirements.[29,66] While the nutrient requirements in Table 10.3 are appropriate for anaerobic processes, only about 4 to 10% of the COD removed is converted into biomass, and thus the nutrient quantities required will be

TABLE 14.9
Relative Inhibition of Selected Organic Compounds to Anaerobic Processes

Compound	Concentration Causing 50% Inhibition, mM
1-Chloropropene	0.1
Nitrobenzene	0.1
Acrolein	0.2
1-Chloropropane	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-Chloropropionic acid	8
Vinyl acetate	8
Acetaldehyde	10
Ethyl acetate	11
Acrylic acid	12
Catechol	24
Phenol	26
Aniline	26
Resorcinol	29
Propanal	90

Note: Adapted from Parkin, G. F. and Owen, W. F., Fundamentals of anaerobic digestion of wastewater sludges. *Journal of Environmental Engineering*, 112:867–920, 1986.

much lower. Consequently, adequate nutrients will generally be available when complex wastes are being treated. However, nutrient addition may be required when carbon rich industrial wastes are being treated. Such wastewaters may be deficient in the macronutrients nitrogen and phosphorus. The concentrations of micronutrients such as iron, nickel, cobalt, sulfur, and calcium may also be limiting.[63,66,73] Nickel and cobalt are particularly important for growth of methanogens.

14.2.8 Mixing

An effective mixing system is critical to the successful operation of an anaerobic process. It provides intimate contact between the microorganisms and their substrates, reduces resistance to mass transfer, minimizes the buildup of inhibitory reaction intermediates, and stabilizes environmental conditions. Mixing is an integral part of the design of many high-rate systems. For example, introduction of the influent wastewater directly into the sludge bed in a UASB bioreactor promotes intimate contact between the wastewater and the granules. Mixing is less efficient in other high-rate anaerobic processes, such as AF, and this is one of the factors restricting their loading.

Mechanical or gas mixing is an integral component of some anaerobic processes, such as anaerobic digestion. Several systems have been developed to mix these processes, and the reader is referred to design references for a detailed discussion.[49,77,85] The contents of such processes are viscous, thixotropic slurries, and mixing criteria applied to other processes are not generally applicable. The solids and wastewaters treated may contain rags and hair, which can wrap around and damage mixing equipment, and inorganic solids such as grit, which can accumulate and reduce

TABLE 14.10
Petrochemicals Metabolized by Enriched Methanogenic Cultures

Petrochemical		
Acetaldehyde	Formaldehyde	Phthalic acid
Acetone	Formic acid	Propanal
Adipic acid	Fumaric acid	Propanol
1-Amino-2-propanol	Glutaric acid	2-Propanol
4-Aminobutyric acid	Glycerol	Propionic acid
Benzoic acid	Hexanoic acid	Propylene glycol
Butanol	Hydroquinone	Resorcinol
Butyraldehyde	Isobutyric acid	Sec-butanol
Butyric acid	Maleic acid	Sec-butylamine
Catechol	Methanol	Sorbic acid
Crotonaldehyde	Methyl acetate	Succinic acid
Crotonic acid	Methyl ethyl ketone	Tert-butanol
Ethyl acetate	Nitrobenzene	Valeric acid
Ethyl acrylate	Phenol	Vinyl acetate

Note: Adapted from Parkin, G. F. and Owen, W. F., Fundamentals of anaerobic digestion of wastewater sludges. *Journal of Environmental Engineering*, 112:867–920, 1986.

the effective volume of the bioreactor if mixing is inadequate. Floating material can accumulate in a scum layer, which also reduces effective volume. Given these challenges, it is interesting that anaerobic digester volumetric power inputs are often lower than those used in aerobic suspended growth processes, such as activated sludge and aerated lagoons. Volumetric power inputs in anaerobic digesters are often in the range of 5 to 8 kW/1000 m^3, but successful performance has been obtained at inputs as low as 1 kW/1000 m^3. The importance of the configuration and efficiency of the mixing system is illustrated by the fact that power densities as high as 20 kW/1000 m^3 have been ineffective in some instances. Egg shaped digesters, shown in Figure 14.3, have superior mixing characteristics and can be properly mixed using lower than normal volumetric power inputs.

Specialized techniques are used to determine the mixing pattern within full-scale anaerobic digesters.[14,57,90] The most frequently used technique involves the pulse addition of lithium into the digester and the monitoring of its concentration in the effluent for at least three SRTs. The results are then analyzed as discussed in Section 4.3.2 to determine the residence time distribution, from which the effective volume of the bioreactor and the proportion of the feed that short-circuits can be estimated. Application of this technique allows the effectiveness of various mixing systems to be determined and compared. It has revealed that significant differences exist in the effectiveness of such systems.

14.2.9 Waste Type

The nature of the wastewater being treated significantly affects its performance in an anaerobic process. One consideration is the relative amounts of soluble and particulate organic matter. Some anaerobic processes are better suited to treat wastewaters containing primarily particulate matter, while others are ideally suited to remove soluble substrates. For example, anaerobic digesters and solids fermentation systems were developed specifically to handle particulate organic matter. They effectively retain particulate material and allow slow hydrolysis reactions to proceed. On the other hand, UASB and hybrid UASB/AF systems do not retain particulate organic matter as effectively,

allowing it to pass through the bioreactor with little hydrolysis and stabilization. They are better for soluble wastes.

Soluble organic matter can be further subdivided into readily and slowly biodegradable components. Slowly biodegradable soluble substrate consists of high molecular weight and/or recalcitrant materials requiring significant metabolism to convert them to the simple monomers that are the substrates of the acidogenic bacteria. Examples include polymers such as carbohydrates and proteins, as well as the complex organic compounds found in many industrial wastewaters. Long SRTs may be required to metabolize these materials.[6,47,75] One characteristic of high-rate processes is their ability to accumulate high concentrations of biomass, which allows maintenance of long SRTs even though their HRTs are short. Thus, effective metabolism of slowly biodegradable soluble organic matter can be achieved in them. Long HRTs are required to degrade such substrates in anaerobic process that are unable to achieve such an effective separation of SRT and HRT. In contrast, a wide range of bioreactor types and process loadings can be used to treat wastewaters containing primarily soluble, readily biodegradable substrates.

The nature of the wastewater has a strong impact on the performance of UASB systems because it affects granule development.[46,47,73,79] Research is still under way to characterize all of the factors that affect the development of granules and there is still no consensus about the determining mechanism triggering granulation.[36] While further research will undoubtedly define the conditions that facilitate granule formation, experience indicates that it is encouraged during the treatment of wastewaters consisting primarily of carbohydrates and retarded during the treatment of wastewaters consisting primarily of VFAs or proteins.[46,47] Granule formation may also be impeded when the wastewater contains a large proportion of particulate or slowly biodegradable organic matter.

The extent and rate of biodegradation of organic solids varies and this can affect the performance of anaerobic digesters. Approximately 70% of the organic matter in municipal primary solids, measured as either COD or VS, is biodegradable in an anaerobic environment.[42,63,77] In contrast, the biodegradability of waste solids from aerobic biochemical operations depends on how much stabilization they have undergone in the operations from which they came. For example, Gossett and Belser[27] found that the biodegradable fraction of waste activated sludge under anaerobic conditions is equal to the active fraction, as defined in Section 5.1.5. Furthermore, the active biomass degraded under anaerobic conditions in a first-order manner, with a rate coefficient of 0.22 day^{-1} at 35°C. In short, the anaerobic stabilization of waste activated sludge is qualitatively and quantitatively similar to its aerobic stabilization, as described in Chapter 13. However, the rate coefficient for waste activated sludge is lower than the rate coefficient for primary solids. Moreover, because the active fraction of waste activated sludge is often on the order of 50%, and only about 80% of the active mass will be stabilized, only a small fraction of the total organic matter in waste activated sludge will be stabilized during anaerobic digestion. These effects are illustrated in Figure 14.13 where the COD reduction efficiencies of municipal primary solids, waste activated sludge, and a mixture of primary solids and waste activated sludge are plotted as a function of anaerobic digester SRT. More recently, Novak and colleagues have demonstrated that the nature of the organic matter affects its degradability under anaerobic conditions.[62]

14.3 PROCESS DESIGN

As we saw in Section 14.1, a wide range of anaerobic process options exists. Although the various processes operate according to a unified set of principles, they differ in many ways. In some, such as anaerobic digestion, the bioreactor functions as a CSTR without biomass recycle so that the SRT is equal to the HRT. In others, significant quantities of biomass are accumulated, allowing long SRTs to be maintained at relatively short HRTs. However, because of the mechanisms used to accumulate biomass in some anaerobic processes, it is impossible to predict or calculate the resulting biomass concentration or SRT. In these instances, empirical correlations between the VOL and the

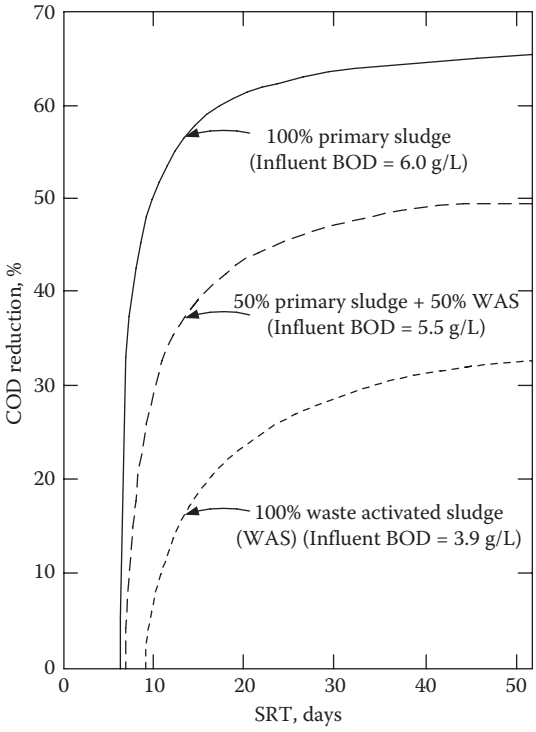

FIGURE 14.13 Effect of SRT on the stabilization of primary solids and waste activated sludge (WAS). (From Parkin, G. F. and Owen, W. F., Fundamentals of anaerobic digestion of wastewater sludges. *Journal of the Environmental Engineering Division, ASCE,* 112:867–920, 1986. Copyright © American Society of Civil Engineers. Reprinted with permission.)

performance must be used for design purposes. In short, a wide range of design procedures must be used to accommodate the wide range of anaerobic processes.

14.3.1 Anaerobic Digestion

The design of an anaerobic digester to stabilize solids is quite straightforward. Since anaerobic digesters are simple CSTRs, the SRT is equal to the HRT. Consequently, the process design consists simply of selecting an appropriate SRT and calculating the bioreactor volume directly from the solids flow rate and the definition of HRT, as given by Equation 4.15. Principal concerns in choosing the SRT include the degree of stabilization and pathogen inactivation required, digester mixing efficiency, requirements for equipment and digester redundancy, and variations in solids flow rates. Table 14.11 summarizes the general procedure.

Several factors must be considered when selecting the minimum acceptable SRT, including washout of methanogens, hydrolysis of particulate organic matter, and pathogen inactivation. As indicated in Figure 10.5, growth of aceticlastic methanogens can be maintained at SRTs as low as five days at 35°C, which is the most common digester operating temperature. While full-scale digesters have been successfully operated at SRTs this low,[85] it really is a lower limit and operation at such an SRT places the digester at risk for rapid washout of methanogens and process failure. Furthermore, the hydrolysis of particulate organic matter and its conversion to acetic acid will generally be the rate limiting steps when treating complex organic material. Consequently, longer SRTs are usually used.

A distinction must be made between the design of anaerobic digesters for treatment of primary solids and waste activated sludge, as discussed in Section 14.2.9. Figure 14.9 demonstrates that for

TABLE 14.11
Summary of Anaerobic Digestion Process Design Procedure

1. Characterize feed sludge characteristics, including type, total solids, volatile solids, flow rate, and expected degradability.
2. Determine treatment objective, including degree of waste stabilization and pathogen control.
3. Select process configuration and SRT.
4. Tabulate sludge flow rates for various operating conditions.
5. Using sludge flow rates and established SRT values, calculate required active digestion volumes.
6. Consider maintenance needs and determine options for number and sizes of digesters. Perform economic analyses, as necessary, to select between the available options.
7. Calculate biogas production based on expected VS mass reduction and 0.7 m^3/kg VS destroyed.
8. Perform a heat balance and determine the need to insulate the digesters, along with supplemental heat to maintain the design temperature (mesophilic or thermophilic conditions) under the range of feed flow rates and sludge and ambient temperatures.
9. Summarize the results of the process design in a table of process loadings and required facilities.

municipal primary solids, an SRT of 8 to 10 days is needed at 35°C to ensure reasonably complete stabilization. Figure 14.14 presents information about the degradation of the various components of those solids and shows that the overall performance is limited by the degradation of lipids. This is consistent with Figure 10.5, which shows that anaerobic oxidation of long and short chain VFAs requires an SRT of about 10 days. The data presented in Figures 14.9 and 14.14 are also consistent with Figure 14.13, where it is observed that an SRT of 10 days results in reasonably complete stabilization of primary solids. Thus, an SRT of at least 10 days is needed to stabilize primary solids at 35°C. However, the hydrolysis of the biomass in waste activated sludge occurs at a slower rate than the hydrolysis of primary solids. As a consequence, a longer SRT is required if waste activated sludge is to be stabilized. Using Figure 14.13 as a guide, an SRT on the order of 15 to 20 days is required to achieve substantial stabilization of waste activated sludge. These conclusions are consistent with observations at full-scale plants.[77,85]

Pathogen control is a relatively new requirement for anaerobic digesters. It has been known for some time that digestion reduces the concentration of indicator organisms. In fact, that is one reason anaerobic digestion has been used. However, the purposeful design of digesters to achieve a specific degree of pathogen control is new. As described in Section 14.1.2 and presented in Table 14.1, the U.S. 503 regulations specify a minimum SRT of 15 days for anaerobic digesters operating at 35°C to ensure pathogen reduction in municipal wastewater solids to meet Class B standards. The design to Class A standards is more involved because the criteria are more stringent. The operation of anaerobic digesters in series also increases pathogen destruction, just as it does in aerobic digesters. As further described in Sections 14.1.2 and 14.1.6, advanced digestion processes are being developed that can accomplish destruction of pathogens to meet the Class A requirements of the 503 regulations, as listed in Table 14.1.

Once the SRT (i.e., HRT) has been selected, the effective volume of the digester is calculated by multiplying the design solids flow rate by the SRT. The design solids flow rate should be for the month or week in which the highest volume of solids is produced to allow the digester to function properly under all reasonable operating conditions. Variations in both the mass of solids produced and the performance of upstream solids thickening devices should be considered in choosing that flow rate. The total volume is then calculated considering the relationship between the effective and total volumes. The effective volume is less than the total volume because of ineffective mixing, leading to the accumulation of grit in the bottom and scum at the top of the digester. In digesters with older style mixing systems, the effective volume can be less than 50% of the total volume, but in digesters with modern mixing systems the effective volume is generally at least 90% of the total volume. Volume should also be allocated to grit and scum accumulations. Typical designs allocate

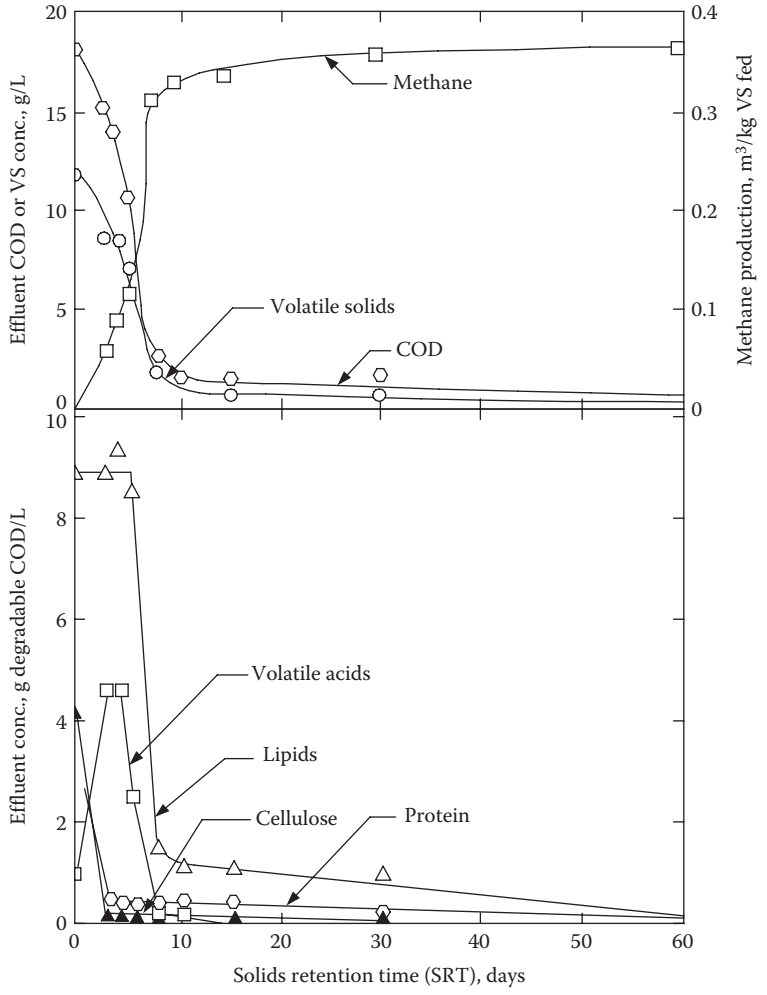

FIGURE 14.14 Fate of various components of municipal primary solids during anaerobic digestion at 35°C. (From Lawrence, A. W., Application of process kinetics to design of anaerobic processes. *Anaerobic Biological Treatment Processes*, American Chemical Society Advances in Chemistry Series, 105:163–89, 1971. Copyright © American Chemical Society. Reprinted with permission.)

the volume of the floor cone (see Figures 1.24 and 14.2) to grit accumulation and the top 0.6 m of the digester to scum accumulation.[54,77,85] Minimal grit and scum accumulations occur in egg shaped digesters (Figure 14.3), so the total and effective volumes can generally be assumed to be the same.

Once the total bioreactor volume has been determined, the number of individual units and their dimensions must be selected. Provisions must be made for units to be removed from service for maintenance, so a minimum of two units should be provided. The impacts on performance of having a unit out of service must also be considered, and this may dictate the number of units provided and/or the total volume. In doing so, it may be assumed that the unit will be removed from service during average, rather than peak, solids production. Considerable latitude exists in the selection of digester sizes, with individual digesters with volumes approaching 20,000 m^3 having been constructed. The gas production rate is estimated based on the mass of volatile solids stabilized and the conversion factor of 0.7 standard m^3 of methane produced/kg of VS destroyed, as presented in Section 14.1.1. The process design for an anaerobic digester to stabilize the waste solids produced at a municipal wastewater treatment plant is illustrated in the following example.

Example 14.3.1.1

An anaerobic digestion system is to be designed to stabilize the solids produced by a municipal wastewater treatment plant. It must be capable of destroying pathogens to Class B standards, implying that the SRT must be at least 15 days at 35°C. The estimated masses of primary solids and waste activated sludge to be produced daily under various conditions are given in Table E14.1. After blending and thickening, the solids concentration entering the digestion system is expected to average 60 g/L (kg/m^3) and to range from 50 to 70 g/L. The volatile solids concentration is 75% of the total solids concentration. Design the system with multiple digesters, but assume that one will be taken out of service for cleaning only under average loading conditions.

a. What solids flow rates must be processed by the system?
   The mass flow rates of dry solids under various conditions are given in Table E14.1. These may be converted to volumetric flow rates by assuming solids concentrations. It is likely that the thickener can maintain the average solids concentration under average and maximum month conditions, but that performance will deteriorate during the maximum solids production week. Consequently, the average solids concentration is used to calculate the average and maximum month volumetric flow rates but the minimum solids concentration is used to calculate the maximum week volumetric flow rate. The results are summarized in Table E14.2.
b. What SRT should be used in the design?
   Because an excellent degree of solids stabilization is desired under average loading conditions, an SRT of 20 days is appropriate, based on Figure 14.13. This value should be attained even during the maximum solids production month, but it is unrealistic to maintain it during the maximum week. However, to ensure pathogen destruction to Class B standards under all conditions, an SRT of at least 15 days must be maintained even during the maximum week.
c. What effective total digester volume must be provided?

TABLE E14.1
Solids Production Rates for Design of the Anaerobic Digester in Example 14.3.1.1

	Mass of Dry Solids, kg/Day		
Type of Solids	Average	Maximum Month	Maximum Week
Primary	18,000	22,500	27,000
Waste activated sludge	16,000	20,000	24,000
Total	34,000	42,500	51,000

TABLE E14.2
Anticipated Volumetric Solids Flow Rates under Various Conditions for Example 14.3.1.1

Condition	Solids Mass kg/Day	Concentration kg/m^3	Flow Rate m^3/Day
Average	34,000	60	567
Maximum month	42,500	60	708
Maximum week	51,000	50	1020

TABLE E14.3
Required Total Digester Volumes under Various Conditions with All Units in Service for Example 14.3.1.1

Condition	Flow Rate m^3/Day	SRT Days	Effective Volume m^3
Average	567	20	11,340
Maximum month	708	20	14,160
Maximum week	1020	15	15,300

The required effective digester volume must be calculated in two steps. First the volume required for each flow rate must be calculated based on the assumption that all units are in service. Since the SRT is the same as the HRT, this is done by multiplying the volumetric flow rate by the SRT. The results are given in Table E14.3. If all units could be kept in service all of the time, then the maximum week would control the design and a total volume of 15,300 m^3 would be required. However, it must be possible to take a unit out of service for maintenance during average conditions, so this must also be considered. If two units were used, then one would have to have a volume of 11,340 m^3 under average conditions to maintain the 20-day SRT, making the total volume 22,680 m^3. This is larger than the volume required during the maximum month or maximum week since both units would be in service then, and would control. Similarly, if three units were used, two would have to have a total volume of 11,340 m^3 under average conditions, making the system volume 17,100 m^3. This, too, is larger than the volume required during the maximum month or maximum week and would control. In this case, using three units reduces the total volume by 25%.

Some savings in digester volume could be achieved by allowing the SRT to decrease to 15 days during the period when one unit was out of service for maintenance. This would have only a minimal impact on performance, as seen by Figure 14.13, and would still ensure pathogen destruction to Class B standards. If this were done, the total effective volume for a two unit system under average conditions would be 17,010 m^3. This is larger than the volumes required for the maximum month and maximum week, which would remain unchanged from the values in Table E14.3 and thus would control. However, for a three unit system, the total effective volume under average conditions would be 12,760 m^3. This is smaller than the volume required for the maximum month and maximum week, so the maximum week would control. Thus, a three unit system would have to have a total volume of 15,300 m^3. Consequently, in this case, using three units only reduces the total volume by 10%.

The choice between these possible designs would have to be made on the basis of economics. However, given the small sacrifice in performance associated with short-term operation at a 15 day SRT, a reasonable decision would be to allow the SRT to drop to 15 days when one unit is out of service for maintenance and to use two units, with a total volume of 17,010 m^3.

d. What volatile solids destruction efficiency and methane production rate would be achieved under the three loading conditions with all units in service?
The volatile solids destruction efficiencies for the primary solids and waste activated sludge must be estimated separately and then combined to obtain the overall digester performance. From Figure 14.13, the COD destruction efficiency for primary solids, which equals the volatile solids destruction efficiency, will be about 60% for SRTs of 15 to 20 days. The effect of SRT over that range is so small that it need not be considered. Although the volatile solids (or COD) destruction efficiency of waste activated sludge depends on its biodegradable fraction, which depends in turn on the operating conditions of the activated sludge system, we will assume that the curve in Figure 14.13 is applicable. It suggests that

TABLE E14.4
Estimation of the Volatile Solids Destruction Efficiency and Methane Production Rate for the Anaerobic Digester in Example 14.3.1.1

	Average	Maximum Month	Maximum Week
Primary solids			
Total solids, kg/day	18,000	22,500	27,000
Volatile solids,[a] kg/day	13,500	16,875	20,250
Volatile solids destroyed,[b] kg/day	8100	10,125	12,150
Methane,[c] m^3/day	5670	7090	8505
Waste activated sludge			
Total solids, kg/day	16,000	20,000	24,000
Volatile solids,[a] kg/day	12,000	15,000	18,000
Volatile solids destroyed,[d] kg/day	2400	3000	3600
Methane,[c] m^3/day	1680	2100	2520
Total solids			
Total solids, kg/day	34,000	42,500	51,000
Volatile solids, kg/day	25,500	31,875	38,250
Volatile solids destroyed			
kg/day	10,500	13,125	15,750
percent	41	41	41
Methane, m^3/day	7350	9190	11,025

[a] 0.75 × total solids.
[b] 0.6 × volatile solids.
[c] 0.7 m^3/kg VS destroyed.
[d] 0.2 × volatile solids.

the volatile solids destruction efficiency of the waste activated sludge would be 20 to 25% at an SRT of 15 to 20 days. Consequently, to be conservative, we will use 20%. Using this information, the overall volatile solids destruction efficiency can be calculated as shown in Table E14.4. It would be around 41% for all three situations and would represent stable solids, as defined in Section 13.1.1. Furthermore, by assuming that destruction of 1 kg of VS results in the formation of 0.7 m^3 of methane under standard conditions, the methane production rate can also be estimated, as shown in the table. The average production rate would be 7350 m^3/day. This information can be used to plan for use of the methane.

The procedures described above and illustrated in Example 14.3.1.1 have historically been used to design anaerobic digestion processes. As discussed in Chapter 8, the International Water Association (IWA) anaerobic digestion model (ADM) has become available and applied to a variety of anaerobic processes. It has been successfully applied to characterize the performance of full-scale anaerobic digestion processes,[69] suggesting that, in the future, it may be used as a design tool.

Another important design consideration for anaerobic digesters is the feeding frequency. Wastewater sludges are thixotropic, and they contain grit, rags, and other debris that can clog piping if adequate velocities are not maintained. Furthermore, the presence of debris requires the use of minimum pipe diameters of 10 cm. The need to maintain minimum velocities in sludge piping often precludes continuous feeding of digesters. Fortunately, because of the relatively long SRTs and HRTs used, periodic feeding will not adversely impact digester performance if the time between feedings is sufficiently short.[76,77,84,85] A feeding frequency of several times per day spaced relatively uniformly will provide acceptable performance. The feed sludges to multiple digesters should be well blended to ensure that

TABLE 14.12
Summary of High-Rate Process Design Procedure

1. Characterize the wastewater to be treated. Characterization involves determination of conventional wastewater parameters such as total and soluble BOD_5 and COD, total and volatile solids, total and volatile suspended solids, pH, alkalinity, temperature, and nutrient concentrations. It also involves assessment of the nature of the organic matter present (i.e., readily or slowly biodegradable; carbohydrate, protein, or synthetic organic compounds of a particular type), and the potential presence of inhibitory materials.
2. Summarize experience with treatment of the particular type of wastewater in anaerobic processes.
3. Compare the subject wastewater with the characteristics of the high-rate anaerobic processes. Some preliminary cost analyses may be conducted to help discriminate among the various process options. Based on the available information, the option or options most appropriate for treating the subject wastewater can then be selected.
4. Determine the need for bench or pilot-scale studies of treatment of the subject wastewater in the selected process(es). This will depend on the available knowledge base.
5. Conduct bench or pilot-scale studies, as appropriate.
6. Develop a correlation between process performance (generally COD removal efficiency) and the VOL or some other measure of loading. Also characterize other pertinent design parameters such as HRT, THL, and bioreactor geometry.
7. Use the process performance relationships developed above to size and configure the bioreactor.
8. Calculate the methane production rate using the stoichiometric relationship 0.35 standard m^3 of methane per kg of COD stabilized.
9. Perform a heat balance and determine the need to insulate the bioreactor and/or to provide supplemental heating. If supplemental heating is needed, size the system.
10. Determine the need for any ancillary facilities such as nutrient addition, pH adjustment or alkalinity addition, iron addition to control sulfide, or sulfur addition to control heavy metal toxicity.
11. Summarize the results of the process design in a table of process loadings and required facilities.

a uniform mixture is fed to parallel units. This can be accomplished by providing a blend tank for feed sludges or by separately distributing the various sludges in equal proportions to each unit.

14.3.2 High Rate Anaerobic Processes

High-rate anaerobic processes are used primarily to treat industrial wastewaters and occasionally to treat municipal wastewaters. Because of the mechanisms that some use to accumulate active biomass, it is often impossible to precisely determine the degree of biomass accumulation and the resulting SRT. Consequently, process performance is commonly correlated with the volumetric organic loading rate. In many instances, this is done by operating a pilot plant for a particular bioreactor type treating a specific wastewater. In other instances, the correlation can be based on experience. Further information on the use of these procedures and detailed examples are presented in Chapter 21.

The general procedure used to design a high-rate anaerobic process is summarized in Table 14.12. Many anaerobic processes are directly analogous to other suspended and attached growth systems discussed elsewhere in this book, and thus the procedures used to size and configure them are similar.

Upflow anaerobic sludge blanket and hybrid UASB/AF processes are designed using the procedures for submerged attached growth bioreactors described in Chapter 21. These processes behave essentially as upflow packed bed bioreactors, and can be sized and configured using the procedures described in Section 21.3.2. These procedures involve selection of appropriate VOLs and THLs and determination of the necessary bioreactor volume, cross-sectional area, and recirculation flow rates (when necessary) using these criteria. The waste solids mass flow rate is calculated based on the biodegradable COD stabilized, along with any nonbiodegradable solids present in the influent wastewater. The methane production rate is calculated using Equation 10.6.

Anaerobic filter processes are similar in configuration to trickling filters, and are designed using procedures like those described in Chapter 19. Volumetric organic loading and THL criteria are used most often, so the approach presented in Section 19.3.2 is appropriate. The waste solids mass flow rate and methane production rate can be calculated using the procedures described above for the UASB and hybrid UASB/AF processes.

Rather than repeat information presented in greater detail elsewhere, the reader is referred to the appropriate sections of this book for detailed descriptions of the process design procedures. Further information on the design of anaerobic processes is provided in books devoted entirely to anaerobic systems.[50,73]

14.3.3 Fermentation Systems

The primary objective of most anaerobic processes is stabilization of biodegradable organic matter through its conversion to methane. In contrast, the objective of fermentation processes is conversion of biodegradable organic matter to VFAs and the harvesting of those VFAs for addition to BNR systems.

Several differences exist between fermentation systems and other anaerobic processes. The first is the SRT. Conversion of biodegradable organic matter to VFAs is accomplished by operation at SRTs that allow the growth of hydrolytic and acidogenic bacteria but preclude the growth of aceticlastic methanogens. The latter is necessary because acetic acid is the most desirable VFA for nutrient removal systems, and thus we do not want it to be converted to methane. Analysis of Figure 10.5 suggests that growth of H_2-utilizing methanogens is likely at the SRTs used, so that some methane will be produced. This results in the loss of some COD in the form of methane, but this loss is beneficial because the consumption of H_2 minimizes its partial pressure in the bioreactor and allows the fermentation reactions to proceed with acetic acid as the main product. Suppression of the growth of sulfate reducing bacteria is also desirable since they will consume acetic acid. Experience with full-scale fermentation systems indicates that their growth can also be controlled by appropriate selection of the SRT.

A second difference is that fermentation bioreactors are generally operated at ambient temperature without heating. Since only limited quantities of methane are produced, heating generally requires an external energy source. Because fermentative bacteria can grow at significantly lower SRTs than aceticlastic methanogens, the economic benefit of reduced bioreactor volume as a result of elevated temperature is much less for fermentation systems. Operation at reduced temperature also makes it easier to limit the growth of aceticlastic methanogens and maximize the production of VFAs.

Another difference is the operating pH. Since the primary objective of a fermenter is production of VFAs, the bioreactor pH will be significantly less than the pH in methanogenic anaerobic processes. The pH values are typically less than six and may be less than five. Reduced pH values also aid in controlling the growth of aceticlastic methanogens.

Although the production of methane is limited, other gases are produced. Significant quantities of carbon dioxide are generated in the hydrolytic and acidogenic reactions, along with some H_2, which will be converted to methane by H_2-utilizing methanogens. Limited quantities of hydrogen sulfide will be produced if sulfate reducing bacteria are able to grow, and nitrogen may also be present because of the entrance of air into the bioreactor or the reduction of any nitrate-N present. Thus, the gas produced will consist primarily of carbon dioxide with small quantities of methane and trace quantities of nitrogen, H_2, and hydrogen sulfide. Consequently, the gas will not generally be combustible.

The typical feed to a fermentation process is primary solids collected from the influent wastewater. At the SRTs and temperatures used, only a portion of the biodegradable organic matter in those solids is converted to VFAs, with yields on the order of 0.05 to 0.3 g VFA produced/g VS fed to the fermenter.[17,18,71,83] Consequently, the solids still contain significant quantities of biodegradable organic matter that must be stabilized prior to final disposal. Primary solids also contain inert

TABLE 14.13
Summary of Fermentation Process Design Procedure

1. Characterize feed sludge characteristics, including type, total solids, volatile solids, flow rate, and expected VFA production.
2. Determine treatment objectives, including VFA production and recovery.
3. Select process configuration, SRT, and elutriation flow requirements.
4. Tabulate sludge flow rates for various operating conditions.
5. Using sludge flow rates, SRT values, and elutriation requirements, size process units.
6. Consider maintenance needs and determine options for number and sizes of required process units. Perform economic analyses, as necessary, to select between the available options.
7. Summarize the results of the process design in a table of process loadings and required facilities.

suspended solids (both volatile and fixed). Thus, the VFAs are normally separated from the solids stream for addition to the BNR process, while the remaining solids are sent for further treatment in the solids processing train.

As illustrated in Figure 14.6, gravity settling is often used for liquid-solids separation in fermentation systems, with the overflow carrying the VFAs for use in the downstream BNR system. Since the concentration of VFAs, which are soluble, is the same in the overflow and underflow from the settler, the recovery of VFAs will be equal to the fraction of flow leaving the settler via the overflow. However, because the solids in the bioreactor are highly concentrated, little additional concentration can occur in the settler, which means that, without dilution, the overflow rate will be a small fraction of the inflow rate, thereby limiting VFA recovery. This problem is overcome by adding an elutriation stream to increase the total flow and make the overflow a larger fraction of the total.

The steps in the design of a solids fermentation process are summarized in Table 14.13. The design of a two-stage, completely mixed thickener/fermenter, shown schematically in Figures 14.5 and 14.6, illustrates the approach. It is presented in the following example.

Example 14.3.3.1

A primary solids flow of 385 m^3/day at a solids concentration of 25 g/L (75% volatile) is to be fermented to produce VFAs to add to a BNR process. A two-stage, completely mixed thickener fermenter is to be used, as illustrated in Figure 14.6d. The process will be sized based on the following assumptions: primary solids are added directly to the fermenter, no overflow from the primary clarifier is added to the fermenter, no thickened solids are recycled from the thickener to the fermenter, and overflow from the primary clarifier is added to the fermenter effluent to dilute it prior to the thickener. Experience with fermentation of these solids indicates that a conversion efficiency of 0.12 g VFA/g VS fed can be achieved at an SRT of five days. In addition, the fermented solids will thicken to 40 g/L. Design the system for 80% recovery of the VFAs produced.

a. What is the volume of the fermenter?
   At the design condition the completely mixed fermenter operates as a CSTR. For a CSTR the volume is just the flow rate times the SRT. Therefore:

$$V = (385)(5) = 1925 \text{ m}^3.$$

b. How many kg/day of VFAs will be produced?
   The mass of volatile solids fed to the process is

$$(25)(385)(0.75) = 7220 \text{ kg VS/day}$$

$$\text{VFA production} = (0.12)(7220) = 866 \text{ kg VFA/day}.$$

c. What volume of overflow from the primary clarifier must be added to the fermenter effluent if 80% of the VFAs are to be recovered in the gravity thickener overflow?
The thickened solids concentration will be 40 g/L. Minimal destruction of solids will occur in the fermenter. Therefore, the mass of solids in the thickener underflow will be approximately equal to the solids fed to the process. From a mass balance, the thickened solids flow rate will be

$$\text{thickened solids flow} = \frac{(385)(25)}{40} = 241\ \text{m}^3/\text{day}.$$

To achieve 80% VFA recovery, the thickener overflow must be 80% of the total flow leaving the thickener and the thickened solids flow must be the other 20%. Since the total flow out must equal the flow in, the flow to the thickener must be

$$\text{thickener influent flow} = \frac{241}{0.2} = 1205\ \text{m}^3/\text{day}.$$

The thickener influent flow consists of primary solids plus overflow from the primary clarifier. Thus, the primary overflow required is

$$\text{primary overflow} = 1205 - 385 = 820\ \text{m}^3/\text{day}.$$

14.3.4 Other Design Considerations

Once the process design is completed, the design of the other components of the system can commence. Mixing and recirculation systems must be selected and sized, with the type depending on the particular anaerobic process being designed. A heat balance must also be done, as discussed previously. If that balance shows that more methane will be produced than is needed to heat the process, then plans can be made for the use of the gas. One potential use is in an engine-driven electric generator, with waste heat from the engine being used to heat the anaerobic process. Alternatively, excess gas can be used directly for various heating purposes at the facility, or it can be processed and sold as a fuel. Insulation of the bioreactor will reduce its heat requirements, and the cost of the insulation can be compared to the value of the extra gas made available to determine whether insulation is justified. Heat transfer can be a daunting task, especially for thermophilic processes where significant amounts of heat must be transferred. Some advanced anaerobic digestion processes use a thermophilic stage followed by a mesophilic stage. In such cases the sludge must be cooled as it is transferred from the thermophilic to the mesophilic stage. Sludge-to-sludge heat exchangers can be used for this purpose.

Materials selection is of particular concern in the design of anaerobic processes. Corrosion of process components is minimal as long as the environment in which they are housed remains completely anaerobic. For example, concrete inside of anaerobic digesters that have been in service for several decades is generally in excellent condition because the environment has consistently been anaerobic. However, corrosion can be excessive at interfaces between anaerobic and aerobic environments because reduced anaerobic reaction products can be oxidized to acidic products as they come in contact with oxygen. One example is hydrogen sulfide, which can be oxidized to sulfuric acid that will attack metal and concrete bioreactor components, causing rapid deterioration. Care must also be exercised in the handling of anaerobic process gas because a combustible mixture can result if it mixes with air. Standard safety equipment is available to prevent atmospheric air from entering anaerobic bioreactors and to suppress an explosion if one begins. However, these devices are not foolproof, and care must be exercised in anaerobic bioreactor design and operation. Further details on the physical design of anaerobic processes are available elsewhere.[54,77,85]

14.4 PROCESS OPERATION

The great variety of anaerobic processes results in a corresponding multiplicity of process monitoring and control techniques, as well as numerous operating problems. Nevertheless, because all of the processes employ similar microbial communities, a number of similarities exist between the monitoring and control techniques used and the operating problems encountered. These similarities are discussed below.

14.4.1 Process Monitoring and Control

Control of anaerobic processes is accomplished primarily by maintaining appropriate loadings and operating conditions. Loadings are controlled by controlling the rate at which biodegradable organic matter is added to the process. Operating conditions of particular concern include temperature and pH. As discussed in Section 14.2.4, temperature must be maintained in an optimum range, but, more importantly, changes in temperature must be held to less than 1°C per day. Optimum performance is generally obtained at pH values between about 6.8 and 7.4. Lower pH values lead to inhibition of methanogens, while higher pH values can lead to ammonia toxicity because of increased free ammonia concentrations. Fortunately, most anaerobic processes operate naturally within this pH range as a result of the carbonate/bicarbonate buffering system. The pH may deviate from this desirable range during process upsets, and pH adjustment chemicals must be added as discussed in Section 14.2.5. Process upsets can result from temporarily high loadings, deviations in the environment provided, or the presence of toxic or inhibitory materials in the bioreactor influent. While pH adjustment is necessary to prevent process failure, the root cause of the upset should be identified and corrected to ensure long-term process stability.

Several parameters can be monitored to assess anaerobic process performance. As indicated by the previous discussion, deviations in bioreactor pH are associated with process upsets. However, both experience and theoretical analysis indicate that pH is not a good indicator of process upsets.[76,84] Because of the buffering capacity inherent in the system, by the time that a noticeable decline in bioreactor pH occurs, the upset may be well under way. Other indicators, such as the relative proportions of VFAs and alkalinity, or the methane production rate, are better indicators of impending failure.

The ratio of VFAs to alkalinity indicates the relative proportion of compounds acting to lower the pH and of buffering capacity acting to maintain it. Any change that suggests an increase in acids or a decrease in buffering capacity indicates an imbalance between the acid forming and consuming microbial populations and an impending upset. Alkalinity concentrations are generally measured by titration, whereas VFA concentrations can be measured directly by gas chromatography or by titration.[1] The VFA to alkalinity ratio and the bioreactor pH should be plotted chronologically to allow detection of trends indicative of an impending upset and to assist in the identification of potential causes.

Methane production is another good indicator of process performance because it is proportional to the mass of biodegradable organic matter stabilized.[51,63,76,84] It is also a direct indicator of the activity of the methanogens. Various specific indicators have been used to quantify process performance. One is the methane content of the gas produced. A decrease in this parameter suggests a decrease in the activity of the methanogens. Another parameter is the volume of methane produced per unit of COD or VS fed to the bioreactor. As long as the composition of the bioreactor feed remains constant, then a fixed proportion of it should be converted into methane if operating conditions remain constant. Therefore, deviations in this ratio suggest deviations in operating conditions. Methane production will change more quickly than bioreactor effluent COD or VS concentrations, and consequently, the ratio of methane produced to COD or VS fed provides an early indication of decreased bioreactor performance. This ratio can also be plotted chronologically and the trends used to identify the onset of upsets.

For many anaerobic processes, the collection of operating data and calculation of process performance indicators on a daily basis is adequate. For these, the HRT is on the order of several days, causing process changes to occur over the course of days. For the high-rate processes, however, HRTs are just a few days and significant changes can occur from one day to the next. These processes benefit from the increased data collection and analysis frequency provided by online analysis. Research is ongoing to evaluate the use of various online control strategies, such as those based on gas flow rate and composition and bicarbonate alkalinity measurements.[11,33,56]

14.4.2 Common Operating Problems

Two of the most common operating problems in anaerobic processes are foaming and the formation of precipitates. Two types of foaming can occur in anaerobic processes. One can occur in all of them. It is associated with incomplete metabolism of the influent organic matter, which leads to the production of intermediates with surface-active properties. Because of the surfactants, the gas produced by the process forms bubbles, to which particulate matter attaches, forming foam. The foam can plug gas piping, interfering with proper operation of the bioreactor, and can escape from the bioreactor, resulting in unsightly and unsafe operating conditions. Moreover, the removal of active biomass by the foam reduces the SRT, thereby decreasing the treatment capacity. A downward spiral of performance can result as the decreased treatment capacity causes more surfactant production, which produces more foaming, thereby removing more biomass, causing more surfactant production, and so on. The primary corrective measure in such instances is to reduce the organic loading to a value that allows complete treatment to occur so that intermediates with surfactant properties are no longer produced.

The second type of foaming occurs in anaerobic digesters that are stabilizing solids from activated sludge systems containing significant quantities of the nuisance foam-causing group of microorganisms known as nocardioforms (see Section 11.4.2). Nocardioforms are a group of branched, filamentous microorganisms that cause foaming in activated sludge systems. When the waste solids from such a system are added to an anaerobic digester, the nocardioforms retain their physical integrity and characteristics, thereby causing foaming in the digester.[34,82] The effects are similar to those described in the preceding paragraph. In severe cases, significant disruption of the digestion process can occur. Correction of anaerobic digester foaming caused by the presence of nocardioforms requires their elimination from the feed solids. This requires correction of the conditions causing the nocardioforms to grow in the upstream activated sludge system.

The other major operational problem in anaerobic processes is the formation of precipitates resulting from the liberation of inorganic constituents during stabilization of complex organic matter like wastewater solids. This follows from the high feed concentrations to many anaerobic processes, which result in high concentrations of inorganic constituents. Precipitation of metal sulfides and their impact on heavy metal solubility were discussed in Section 14.2.6. Other precipitates can form in large quantities, forming scale on surfaces and clogging pipes. Two precipitates of particular concern are struvite ($MgNH_4PO_4$) and calcium carbonate ($CaCO_3$).

Struvite precipitation occurs most frequently when waste biomass and vegetable matter are digested because their stabilization releases inorganic cell constituents, including magnesium, ammonia, and phosphate.[76,84] Its formation is exacerbated when waste activated sludge from a biological phosphorus removal process is digested because such sludges contain increased concentrations of phosphorus and magnesium, which are easily released under anaerobic conditions. Struvite is moderately soluble (pK_{sp} of 12.6 at 25°C), but the high biomass concentrations in the feed to many anaerobic digesters result in a supersaturated solution with respect to struvite precipitation.[48] The kinetics of struvite precipitation are slow and precipitation does not occur immediately or uniformly. Instead, it begins at some location within the digester system, and further precipitation occurs rapidly at that location, resulting in the formation of a scale. Struvite precipitates occur frequently at points of turbulence, such as in overflow structures, in piping immediately adjacent to pumps, and in heat exchangers or heat exchanger piping. These points

of turbulence strip carbon dioxide, resulting in a localized increase in pH. Struvite solubility decreases with increasing pH, thereby increasing the degree of supersaturation and the propensity for a precipitate to form.

Work is ongoing to characterize the precise conditions under which struvite precipitation occurs.[48,60,61] Design features to minimize the impacts of struvite precipitation include the use of plastic or glass lined pipes to minimize the adherence of precipitates, the design of piping systems with long radius elbows and other features to minimize turbulence, and the incorporation of features to allow easy cleaning of piping.[84] Iron addition to precipitate phosphorus as iron phosphate rather than as struvite is a standard practice that has proven to be highly effective in controlling struvite scaling in a wide variety of anaerobic processes. A one-to-one molar dose of iron to the phosphorus released (5 mg $FeCl_3$/mg P) provides a reasonable initial estimate of the iron dose needed. Iron addition also reduces the sulfide content of biogas because iron precipitates dissolved sulfide, thereby reducing the amount that will be present in the produced gas.

Calcium carbonate precipitates can form when treating high strength wastewaters that also contain high concentrations of calcium (for example, dairy wastes).[47] Carbonate formed as a result of stabilization of the organic matter reacts with the calcium to form the precipitate. The precipitate may form on surfaces, but it may also form within biomass flocs or biofilms. For example, calcium carbonate precipitates have been observed in the granules in UASB systems. These precipitates may or may not cause operating problems. However, they introduce a nonreactive solid phase that reduces treatment capacity by reducing the unit activity of the biomass in the process. These effects must be accounted for in the process design.

14.5 KEY POINTS

1. Anaerobic processes stabilize biodegradable organic matter by converting it to methane gas. At standard temperature and pressure, 0.35 m^3 of methane are produced per kg of chemical oxygen demand (COD) removed. For primary solids this is equivalent to 0.7 m^3 of methane per kg of volatile solids (VS) removed.
2. Anaerobic processes consist of four major components: (1) a closed bioreactor, (2) a mixing system, (3) a heating system, and (4) a gas-liquid-solids separation system.
3. Anaerobic processes can be grouped into three principal types: (1) anaerobic digesters, (2) high-rate anaerobic processes, and (3) solids fermentation processes.
4. Anaerobic digesters are used to stabilize biodegradable organic matter and inactivate pathogens in slurries that contain high concentrations of particulate matter. They are completely mixed bioreactors with no cell recycle and can be characterized as single continuous stirred tank reactors (CSTRs).
5. Advanced anaerobic digestion processes are being developed that provide for increased destruction of biodegradable organic matter and pathogens. Such processes involve tanks-in-series configurations and can involve separation of the acetogenic and methanogenic phases, operation at thermophilic or staged thermophilic and mesophilic temperatures, and batch operation to meet specified time and temperature relationships.
6. High-rate anaerobic processes incorporate a variety of biomass retention mechanisms, including the formation of readily settleable particles that are retained by sedimentation and the use of bioreactor configurations that retain suspended solids. High-rate anaerobic processes are generally able to remove 80 to 90% of the five-day biochemical oxygen demand (BOD_5) applied (COD removed is approximately 1.5 times the mass of BOD_5 removed), to produce 0.35 m^3 of methane per kg COD removed, and to produce 0.05 to 0.10 kg of biomass (as volatile suspended solids [VSS]) per kg COD removed.
7. The various high-rate anaerobic processes differ in their ability to successfully treat wastewaters with high concentrations of particulate matter and in the volumetric organic loadings (VOLs) applied. Anaerobic filter (AF) systems can treat wastewaters with moderate

levels of suspended solids. The upflow anaerobic sludge blanket (UASB) and hybrid UASB/AF bioreactors do not generally respond well to wastewaters containing high concentrations of particulate matter. High VOLs can be applied to UASB and hybrid UASB/AF bioreactors if the wastewater contains primarily soluble organic matter.

8. Solids fermentation systems differ from other anaerobic processes since their objective is the production of volatile fatty acids (VFAs) and their separation from the waste stream for feeding to biological nutrient removal processes.
9. Anaerobic processes are generally competitive with aerobic processes for the treatment of wastewaters with biodegradable COD concentrations greater than 1000 mg/L, and are usually the process of choice when the biodegradable COD concentration exceeds 4000 mg/L. Factors affecting the choice include waste strength, flow rate, and temperature. Combustion of the methane produced and recovery of heat from the effluent can be used to achieve the temperatures required for effective anaerobic treatment.
10. The solids retention time (SRT) is the primary factor determining the performance of anaerobic processes. An SRT of 15 to 20 days is generally required to achieve stable, reliable performance at 35°C. Many high-rate processes have SRTs in excess of 30 to 50 days and sometimes even 100 days. These very high SRTs may partially account for their stability.
11. It is not possible to precisely determine the SRT for some bioreactors. In such cases, the VOL is used to characterize process performance. The VOL is related to the SRT through the process yield and the biomass concentration.
12. The performance of some high-rate anaerobic processes, such as AF, UASB, and hybrid UASB/AF, is affected by the total hydraulic loading (THL). The THL is the total bioreactor influent flow rate, including recirculation, divided by the cross-sectional area perpendicular to the flow. Total hydraulic loading criteria include maximum values to prevent biomass washout and minimum criteria to ensure good flow distribution.
13. Optimum performance of anaerobic processes is generally achieved by operation at a temperature near the optimum for mesophilic (30 to 40°C) or thermophilic (50 to 60°C) microorganisms. Acceptable performance can be achieved at temperatures below these values if an increased SRT is provided and if sufficient time is allowed for acclimation. However, operation between the mesophilic and thermophilic ranges should be avoided. In addition, short-term temperature fluctuations must be avoided, with a typical goal of no more than 1°C/day.
14. The optimum pH range for growth of aceticlastic methanogens is 6.8 to 7.4, with their activity decreasing significantly at pH values below this range. In contrast, the growth of acidogenic bacteria is much less sensitive to low pH. This difference in pH sensitivity can result in a downward spiral in process performance in which retardation of the aceticlastic methanogens causes reduced consumption of VFAs relative to their formation, which results in further reductions in pH, and so on. This condition can be resolved by pH adjustment and reduction of the loading.
15. The pH in anaerobic processes is determined by the bicarbonate buffering system. Excess carbon dioxide is generally produced, and the quantity of bicarbonate is determined by the concentration of a strong base available to react with it. Ammonia-N present in the wastewater or produced through biodegradation of nitrogen-containing organic matter is the base carbon dioxide reacts with most often.
16. Sodium bicarbonate is the most desirable chemical for pH adjustment. Other options result in pH variations as the added chemical reacts with carbon dioxide, removing it from the gas space. When the carbon dioxide balance is restored through continued metabolic activity, a second pH shift occurs. Addition of calcium-based chemicals can also cause the precipitation of calcium carbonate.
17. The light metal cations sodium, potassium, calcium, and magnesium are required nutrients in anaerobic processes and their presence at low concentrations causes a stimulatory effect

on microbial growth. However, elevated concentrations can cause moderate inhibition, and even higher concentrations can cause severe inhibition or toxicity. Interactions between the light metal cations can cause either increased or decreased inhibition.

18. Total ammonia (the sum of the free plus ionized ammonia species) concentrations of 50 to 200 mg/L as N stimulate microbial growth in anaerobic processes. However, free ammonia (NH_3) can be inhibitory if it reaches concentrations of about 100 mg/L as N. The fraction of total ammonia present as free ammonia increases with increasing temperature and pH.
19. Three strategies are available for reducing ammonia toxicity, including: (1) reducing the temperature, (2) reducing the pH, or (3) reducing the total ammonia concentration. The pH can be reduced by the addition of hydrochloric acid.
20. Dissolved sulfide is toxic to anaerobic processes at a concentration of about 100 mg/L (200 mg/L with acclimation). Sulfide is formed by the destruction of sulfur-containing organic matter and by the reduction of sulfate. The possibility of sulfide inhibition must be considered for wastewaters with COD/SO_4^- ratios less than about 7.5 but is generally not a problem when the ratio exceeds 10. Sulfide reacts with heavy metals, forming insoluble precipitates that are not inhibitory. The reduction of sulfate requires electrons from biodegradable organic matter, thereby decreasing the number available for methane production. Sulfide production also decreases the degree of waste stabilization because soluble sulfide exerts an oxygen demand.
21. Dissolved heavy metals can be quite toxic to anaerobic processes. However, the presence of dissolved sulfides minimizes their effect since the sulfide precipitates of heavy metals are quite insoluble.
22. The nonionized forms of the VFAs are inhibitory, with concentrations on the order of 30 to 60 mg/L having an effect. At neutral pH relatively high total VFA concentrations are required to cause nonionized concentrations in that range. Propionic acid is much more inhibitory than either acetic or butyric acid.
23. A wide variety of organic compounds can inhibit anaerobic processes. However, biomass can become acclimated to many of these compounds and cultures can acquire the ability to biodegrade many of them.
24. Because the net process yield is low in anaerobic systems, nutrient limitations are seldom encountered when treating complex wastewaters, but they may occur when treating certain high strength industrial wastewaters. Nutrients of concern include the macronutrients nitrogen and phosphorus and the micronutrients iron, nickel, cobalt, sulfur, and calcium.
25. Several approaches are used to mix anaerobic processes, including effluent recirculation, gas recirculation, and mechanical mixing. Mixing in anaerobic digesters is particularly challenging because of the thixotropic nature of the solids processed.
26. The nature of the organic matter fed to an anaerobic process can dramatically affect its performance. For example, the biodegradable portion of the particulate organic matter in primary solids is typically about 70%, while the biodegradable portion of the particulate organic matter in waste activated sludge typically ranges from 30 to 50%, depending on the SRT of the activated sludge system from which it came.
27. Anaerobic digesters are typically designed with SRTs on the order of 15 to 20 days at 35°C to achieve good stabilization of biodegradable organic matter and to meet Class B pathogen control requirements. Design procedures must also consider variations in influent flow rates and requirements to periodically remove units from service for maintenance.
28. High-rate anaerobic processes are often designed using VOL and other parameters based on pilot-scale and full-scale experience. Pilot tests may be needed to design a specific installation, or previous experience with the subject wastewater in anaerobic processes may provide sufficient information for design. The procedures used to design many anaerobic process options are analogous to those used to design other processes considered in this book.

29. Several solids fermentation process configurations are available. In general, fermentation processes are operated at SRTs that are short enough to preclude the growth of aceticlastic methanogens.
30. Control procedures for anaerobic processes generally require monitoring the bioreactor pH, the volatile acids to alkalinity ratio, and the methane production rate.
31. Foaming removes active biomass from the liquid phase in the bioreactor, thereby interfering with anaerobic treatment. It can be caused by incomplete metabolism of influent organic matter or by the presence of nocardioforms. Foaming caused by incomplete metabolism can be reduced by reducing the process loading. Foaming caused by nocardioforms requires their elimination from the feed by appropriate control of the activated sludge system producing them.
32. Precipitates can form in anaerobic processes and cause scaling of surfaces and plugging of piping. One frequently encountered precipitate is struvite ($MgNH_4PO_4$). It is encountered when complex wastes are degraded, resulting in releases of high concentrations of magnesium, ammonia, and phosphate. Struvite formation is often addressed by iron addition to precipitate phosphorus. Calcium carbonate can form when wastes that are high in calcium (such as dairy wastes) are treated.

14.6 STUDY QUESTIONS

1. Prepare a table summarizing the advantages and disadvantages of anaerobic digestion compared to aerobic digestion for the stabilization of waste solids.
2. Prepare a table summarizing the advantages and disadvantages of high-rate anaerobic wastewater treatment processes relative to aerobic processes. When is the anaerobic process typically used?
3. Discuss the roles of H_2 utilizing and aceticlastic methanogens in anaerobic processes.
4. Prepare a table summarizing the typical design criteria for the various anaerobic processes. Contrast those processes in terms of the fate of soluble and particulate organic matter within them.
5. List the principal biomass retention mechanisms used in each of the anaerobic processes.
6. Prepare a table summarizing the advantages and disadvantages of mesophilic versus thermophilic anaerobic processes.
7. Discuss the impact of SRT on the reactions occurring in anaerobic processes. What SRT should be selected for various applications and why?
8. A wastewater with a biodegradable COD concentration of 20 g/L is being treated in an anaerobic process operated at an HRT of five days. What is the VOL? If the bioreactor biomass concentration is 10 g VSS/L, what is the SRT? What net yield value did you use in the calculation of the SRT and why?
9. An anaerobic digester is treating waste solids with a volatile solids concentration of 40 g/L. If the VOL is 3 kg VS/(m^3·day), what is the HRT?
10. Municipal primary solids with characteristics similar to those used to develop Figure 14.9 are to be treated in an anaerobic digester. A minimum of 70% of the biodegradable organic matter is to be converted to methane. What SRT is required to achieve this objective at temperatures of 35, 25, and 20°C? Should somewhat larger SRTs be used in some cases to increase process stability? If so, when and why?
11. An anaerobic digester treating municipal primary solids with a volatile solids concentration of 60 g/L has an HRT of 25 days. If it is operating at 35°C, what volatile solids destruction efficiency would be expected? Explain how you arrived at your answer.
12. Prepare a diagram demonstrating the downward spiral that occurs as a "stuck" or "sour" anaerobic process develops.

13. An anaerobic process is operating with a bicarbonate alkalinity concentration of 750 mg/L as $CaCO_3$ and a gas carbon dioxide content of 40%. What is the bioreactor pH?
14. For the anaerobic process described in Study Question 13, how much sodium bicarbonate, lime, sodium carbonate, or ammonia must be added to the bioreactor to adjust the pH to 7.0?
15. For the anaerobic process described in Study Question 13, to what value must the gas carbon dioxide content be adjusted to produce a bioreactor pH of 7.0?
16. Describe what is meant by "stimulatory" and "inhibitory" concentrations of a chemical. Describe what is meant by "synergistic" and "antagonistic" interactions of inhibitors.
17. Discuss the relationship between temperature and pH as it affects ammonia toxicity in anaerobic processes.
18. A wastewater with a flow rate of 1000 m^3/day and a biodegradable COD concentration of 25 g/L is to be treated in an anaerobic process. Assuming typical performance for a high-rate anaerobic process, what will the methane production rate be?
19. A second waste stream with a flow rate of 300 m^3/day, a sulfate concentration of 5 g SO_4^-/L, and a biodegradable COD concentration of 1000 mg/L is to be added to the anaerobic process described in Study Question 18. If all of the sulfate is reduced to sulfide and the sulfide is precipitated with iron, how will the addition of this waste stream affect the methane production rate from the anaerobic process?
20. What can be done to reduce the dissolved sulfide concentration in an anaerobic process?
21. Heavy metal toxicity is occurring in an anaerobic process. What chemicals can be added to eliminate this toxicity? What are the relative advantages and disadvantages of these chemicals?
22. A wastewater with a biodegradable COD concentration of 20 g/L is to be treated in an anaerobic process. What concentrations of ammonia-N and phosphorus are required to achieve efficient treatment?
23. The primary solids and waste activated sludge (WAS) from a wastewater treatment plant are to be stabilized by anaerobic digestion. For the primary solids, 75% of the total solids are volatile and 70% of the volatile solids are biodegradable. For the WAS, 80% of the total solids are volatile and 40% of the volatile solids are biodegradable. The solids masses and thickened solids concentrations are given in Table SQ14.1. For this plant, do the following:
    a. Select an appropriate SRT to stabilize the biodegradable organic matter and inactivate pathogens, and calculate the total digester effective volume required under average, maximum month, and maximum week conditions. Assume that the operating temperature is 35°C.
    b. Evaluate options that provide two, three, or four digesters and determine the option that requires the minimum total bioreactor volume. Assume that digester cleaning occurs only under average loading conditions.
    c. Calculate the methane production rate under all loading conditions.

TABLE SQ14.1
Data for Study Question 23

	Average		Maximum Month		Maximum Week	
Type of Solids	Mass kg/Day	Conc. g/L	Mass kg/Day	Conc. g/L	Mass kg/Day	Conc. g/L
Primary	25,000	50	27,500	50	30,000	40
WAS	20,000	45	22,500	40	25,000	35

24. Reconsider Study Question 23. The SRT in the activated sludge system is to be reduced, resulting in a 25% increase in the mass of biodegradable volatile solids in the WAS stream. How much more methane will be produced when the solids are anaerobically digested?
25. Discuss when treatability tests and a pilot study should be conducted prior to the design of an anaerobic process to treat an industrial waste.
26. Consider a wastewater with a flow of 125,000 m^3/day and a TSS concentration of 200 mg/L (75% volatile). Primary treatment of this wastewater results in removal of 60% of the TSS. Solids are removed from the primary clarifier at a concentration of 10 g/L. A completely mixed/thickener fermenter is to be designed to produce VFAs to add to a BNR system. The fermented primary solids can be gravity thickened to 25 g/L. You wish to recover 85% of the VFAs produced. Do the following:
    a. Size the completely mixed fermenter.
    b. Determine the flow rate of any elutriation streams required.
    c. Determine the mass of VFAs formed in the process and the mass elutriated for addition to the BNR system.

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15 Lagoons

The term lagoon refers to a diverse array of suspended growth biochemical operations with the common characteristic that they do not include downstream clarifiers and associated settled solids recycle. Their name comes from the technique historically used to construct them, as inground earthen basins that resemble shallow ponds. Lagoons are typically used to stabilize biodegradable organic matter, although nitrogen removal (by nitrification/denitrification and ammonia stripping) and phosphorus removal (by chemical precipitation) are observed in some instances. Several process options are available, depending on the type of metabolism occurring and the mechanism used to provide the terminal electron acceptor. This chapter provides an overview of lagoon options, with particular focus on aerobic lagoons.

15.1 PROCESS DESCRIPTION

Lagoons represent one of the oldest forms of biological wastewater treatment, having been used in some form for more than 3000 years.[43] They have been used as the only means of treatment prior to discharge to surface waters and for pretreatment and/or storage prior to treatment in a conventional system or a wetland. A wide range of industrial and municipal wastewaters has been treated in lagoon systems.

Lagoons are mechanically simple, which often translates into low capital and operating costs. However, this mechanical simplicity masks a degree of physical, chemical, and biological complexity unparalleled by other biochemical operations, resulting in a poor understanding of the factors that affect process performance. As a consequence, the effluent quality from lagoons has often been relatively poor, relegating them to uses where high quality effluent is not necessary. Algal growth is a particular problem. Algae generally settle slowly and, consequently, pass into the effluent where they increase the concentration of suspended solids and biodegradable organic matter. Recently, the rational application of fundamental principles, including understanding of the factors influencing algal growth, has led to the development of aerated lagoon systems with significantly improved effluent quality.[28–32,35,37,38,40,47] Background will be provided in this chapter on the variety of lagoons used in practice, but emphasis will be placed on design of the newer generation of aerobic lagoons.

15.1.1 General Description

Figure 15.1 presents a schematic diagram of a lagoon. The structure is typically an earthen basin constructed with sloping sidewalls. To minimize construction costs, the lagoon is often configured so that the soil needed to construct the sidewalls is excavated from the interior (i.e., cut and fill are balanced). Natural sealing will occur to some extent as wastewater solids enter the pores of the soil and reduce the seepage rate. However, it is now common practice to provide a liner for positive seepage control. Materials used include: natural clays (such as bentonite), asphalt, synthetic membranes, and concrete. Regardless of the liner material used, a concrete apron is often provided at the water line to simplify maintenance. The remainder of the sidewalls above the water level is often covered by grass. Details on lagoon construction are provided elsewhere.[21,40,43] Influent and effluent structures complete the lagoon. Influent enters at one end, and the treated wastewater is collected in an effluent structure, generally located at the opposite end. No formal mechanisms are provided to retain biomass within the lagoon. Consequently, the solids retention time (SRT) approaches the hydraulic residence time (HRT), and therefore HRTs on the order of several days are typically used.

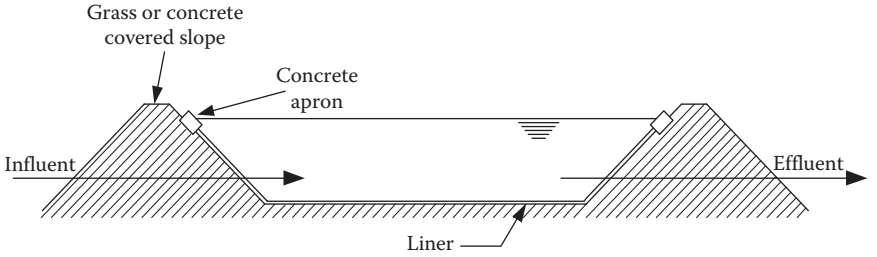


FIGURE 15.1 Schematic diagram of a lagoon (vertical dimension exaggerated).

The term lagoon refers to a configuration, not to a set of controlled environmental conditions within a bioreactor. In fact, the environmental conditions depend on the process loading and operating conditions, causing significant variations in the physical, chemical, and biochemical conversions occurring, and in the treatment efficiency obtained. In short, lagoons can be used in a variety of ways and a wide range of lagoon options exists.

Many approaches can be used to characterize lagoons.[6,21,22,40,43,45] We use a simplified method based on the type of metabolism occurring and the mechanism by which the terminal electron acceptor is supplied. In anaerobic lagoons, biodegradable organic matter is stabilized by its conversion to methane and carbon dioxide. In other lagoons, oxygen is provided as an electron acceptor, and biodegradable organic matter is stabilized by its conversion to carbon dioxide and water. Two principal mechanisms are used to provide oxygen: (1) by the growth of algae, which produce oxygen through photosynthesis; and (2) by mechanical means, such as with surface aerators, diffused aeration by blowers, and unique systems specifically designed for lagoons. These differences provide the basis for the three types of lagoons discussed in this chapter: (1) anaerobic, (2) facultative and facultative/aerated, and (3) aerobic.

15.1.2 Process Options and Comparison

15.1.2.1 Anaerobic Lagoon

An anaerobic lagoon (ANL) is a low rate anaerobic process in which biodegradable organic matter is stabilized through its conversion to carbon dioxide and methane. They often use earthen basins, although rectangular concrete vessels have also been used. Compared to other lagoons, ANLs are constructed as relatively deep structures, typically ranging from 2 to 6 m deep. This minimizes the lagoon surface area for a given volume, thereby minimizing oxygen transfer, odor release, and heat loss from the surface, which is important because ANLs are commonly not covered.[21,40,43,45] Historically, materials in the wastewater were allowed to float to the surface and form a scum mat that provided some insulation and odor control, although gas would pass through it and escape to the atmosphere. However, they can be covered to collect the methane gas produced and to eliminate odor release, as illustrated in Figure 1.28. Another alternative is to provide oxygen to the top layer of the lagoon so that the odoriferous compounds are oxidized to less odoriferous materials before they leave the lagoon. This aerated "cap" can be provided by directly aerating the top layer of the lagoon or by recirculating oxygenated process flow to it. Although some oxygen transfer can occur across the surface of uncovered ANLs, anaerobic conditions develop because the addition rate of biodegradable organic matter greatly exceeds the oxygen transfer rate. The loading rates also preclude the growth of algae, which would produce oxygen if they were present. Influent digestion chambers are sometimes incorporated into ANLs, particularly those treating more dilute wastewaters such as municipal wastewater.[21] Mechanical mixing can be provided, but is not in many systems. In all cases, gas evolution from the digesting organic matter provides some mixing. Some systems have incorporated settled solids recycle from a downstream settling zone to an upstream reaction zone,

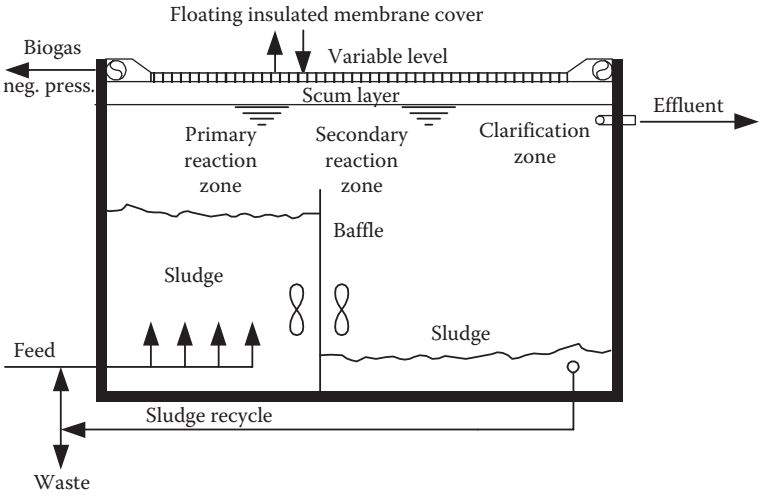


FIGURE 15.2 Anaerobic lagoon using a rectangular concrete structure with solids recycle to upstream reaction zone.

as illustrated in Figure 15.2. Heating is typically not provided. Consequently, either the wastewater must be sufficiently warm or an adequately long SRT must be maintained to allow treatment to occur at ambient temperature.

Environmental conditions within ANLs are not well regulated and, even though active biomass accumulates, accurate control of the SRT is generally not possible. Solids and floating materials accumulate in the lagoon, thereby retaining biomass and providing surfaces for microbial growth. The resulting retention of biomass makes the SRT greater than the HRT, but the difference is difficult to quantify. Hydraulic residence times can be as high as 20–50 days, but many ANLs are designed and operated with HRTs less than 10 days, with values in the 5–10 day range often quite appropriate.[21,40,43,45] As discussed in Chapter 14, an SRT in excess of 20 days (more likely approaching 40 or 50 days) is required to produce a stable, effective anaerobic treatment process, even when bioreactor temperature and other environmental conditions are carefully controlled. No such controls are provided in an anaerobic lagoon. Consequently, the fact that ANLs work suggests that significant retention of biomass occurs to achieve the required SRT. Organic loading rates of 1–2 kg chemical oxygen demand (COD)/($m^3 \cdot$ day) are often found to be appropriate. Process performance varies, but approaches that are achieved by high rate anaerobic processes are described in Chapter 14.

15.1.2.2 Facultative and Facultative/Aerated Lagoon

As their name suggests, facultative and facultative/aerated lagoons (F/ALs) are systems in which biodegradable organic matter is stabilized by both anaerobic and aerobic processes. As illustrated in Figure 1.27, the lower portion of a facultative lagoon is anaerobic and biodegradable organic matter is stabilized there by anaerobic processes that convert it into methane and carbon dioxide. The upper portion of the lagoon is aerobic because oxygen is provided by algal growth (major contribution) and by surface reaeration across the air-liquid interface (minor contribution). Because of the aerobic environment, biodegradable organic matter is stabilized by aerobic metabolism in the upper zone. The aerobic zone also provides a "cap" that oxidizes reduced compounds produced in the underlying anaerobic zone, minimizing odor release and oxidizing soluble oxygen-demanding compounds. Effluent is withdrawn from the aerobic zone.

A synergistic relationship exists between the bacteria and the algae in a facultative lagoon. Bacteria stabilize organic matter via anaerobic and aerobic metabolism, resulting in new biomass.

During daylight hours, algae produce oxygen in the upper portion of the lagoon and the bacteria there use it as their electron acceptor. Carbon dioxide produced by the bacteria serves as the carbon source for algal growth, while sunlight provides the necessary energy. However, when light is not available, algae use molecular oxygen to oxidize biodegradable organic matter and obtain energy by heterotrophic metabolism. Although the presence of algae produces most of the oxygen needed by the bacteria for aerobic metabolism, it also results in a decrease in waste stabilization because a portion of the carbon dioxide produced by the bacteria is converted back into particulate organic matter in the form of algal cells. Experience indicates that many algae do not settle well and pass into the lagoon effluent where they contribute biodegradable organic matter and total suspended solids.[28–30,32,35,40,47]

Algal growth is promoted by constructing facultative lagoons as shallow basins, generally 1–2 m deep, thereby allowing maximum exposure of the lagoon contents to sunlight; minimizing mixing, so that light can penetrate the upper layers of the lagoon; and balancing the organic loading with the production of oxygen by the algae. Because oxygen production is generally limited by the light available to the algae, and the light available is determined by the lagoon surface area, the organic loading is generally expressed as the mass of biodegradable organic matter applied per day per unit of surface area. The allowable organic loading rates generally result in HRTs of 25 days or more.

Diurnal variations in incident light cause significant changes in the environmental conditions within facultative lagoons.[21,40,43,45] During the day, when light is available and algae produce oxygen, the size of the aerobic zone is significant. During the night, however, when light is not available, the size of the aerobic layer is reduced, perhaps to zero. In addition, the diurnal variation in algal activity causes the carbon dioxide concentration to vary, which produces pH variations. During the day, the pH in the aerobic zone can reach values as high as 10 as carbon dioxide is consumed by the algae. During the night, on the other hand, the pH decreases to seven or below as carbon dioxide is produced by both bacterial and algal respiration. The long HRTs in facultative lagoons, coupled with the high pH values, result in excellent pathogen destruction. In fact, in some instances, facultative lagoons have been used for the disinfection of municipal wastewater.[20,25,40] Sedimentation of nematode eggs is another important pathogen removal mechanism.[3,40]

Significant variations in facultative lagoon performance occur because of ambient conditions, which vary on both a seasonal and a geographical basis. For example, ambient temperatures vary and this affects the temperature in the lagoon. The availability of sunlight also varies seasonally and geographically. Thus, wide ranges in environmental conditions can exist within a lagoon, resulting in a wide range in allowable loadings.[21,40] Consequently, care must be used in extrapolating allowable loadings from one location to another. In some areas, lagoons freeze during the winter, which disrupts performance. This problem can be overcome by making the lagoon large enough to accumulate the wastewater during the portion of the year when performance is unsatisfactory, allowing its discharge only when the effluent quality is acceptable and the receiving water has sufficient assimilative capacity. Lagoons can also be designed to prevent surface water discharges; the water either seeps into the groundwater or evaporates.

Facultative lagoons can remove nitrogen and phosphorus from wastewaters. Nitrogen is removed by two mechanisms: nitrification and denitrification, and ammonia stripping. Because zones of both high and low oxygen concentration exist within a facultative lagoon, the environments required for both nitrification and denitrification are present. Ammonia stripping occurs because of the high pH in the aerobic zone of the lagoon. Elevated pH results in conversion of ammonium ion to free ammonia, as illustrated in Equation 14.13. Free ammonia can be easily volatilized to the atmosphere. Although both mechanisms operate, their relative importance is not known.[14,24,40] Elevated pH can also result in the precipitation of phosphorus, thereby removing it from the liquid phase. Although these conversions occur in lagoons, they may not occur consistently. Consequently, effluent nutrient concentrations may fluctuate.

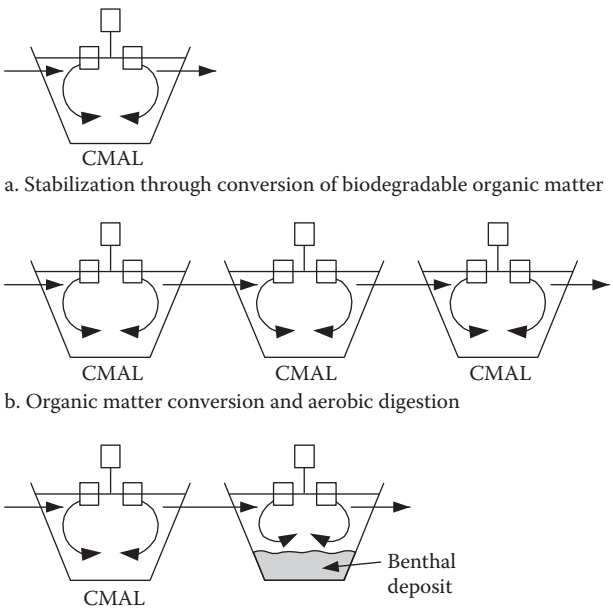
The organic loading on a facultative lagoon can be increased by providing additional oxygen by mechanical means. If only a low level of mixing energy is introduced by the oxygen transfer device,

insufficient to completely mix the lagoon, the two zones will be maintained and light penetration will be sufficient for the algae to grow in the same fashion as in facultative lagoons. This provides the basis for facultative/aerated lagoons, which have operation and performance characteristics similar to facultative lagoons but with somewhat higher allowable loadings.

15.1.2.3 Aerobic Lagoon

Aerobic lagoons (AELs) are designed and operated to exclude algae. This is accomplished by two means. First, sufficient mixing is used to keep all biomass in suspension, thereby providing turbidity that restricts penetration of light into the water column. The mixing also has the effect of making the SRT equal to the HRT. Second, the HRT is controlled to values less than the minimum SRT for algal growth (about two days).[28–30,32,35,37,47] Because algae are excluded, oxygen must be delivered by mechanical means.

Aerobic lagoon systems can be designed to meet a variety of objectives, including the removal of biodegradable organic matter through its conversion to biomass, the stabilization of organic matter (including synthesized biomass) by aerobic digestion, and the removal of synthesized biomass by gravity settling.[28–30,35,37,38,47] Figure 15.3 illustrates these process options. Regardless of the objective, the first step in an AEL system is a completely mixed aerated lagoon (CMAL) where sufficient mixing is provided to keep all biological solids in suspension. Just as in activated sludge systems, aerobic bacteria oxidize a portion of the biodegradable organic material into carbon dioxide and water, and convert a portion into new biomass. Consequently, the overall waste stabilization accomplished is the difference in the oxygen demands of the original wastewater and the synthesized biomass. As discussed in Chapter 5, this is equal to $1 - (Y_{Hobs,T}\, i_{O/XB,T})$, which typically represents stabilization of about 40%. As discussed in Section 10.3.2, nearly complete conversion of biodegradable organic matter into biomass can be accomplished aerobically at SRTs on the order of two to three days. Experience confirms this for a CMAL, where the HRT equals the SRT.[28,29,31,35,37]



a. Stabilization through conversion of biodegradable organic matter

b. Organic matter conversion and aerobic digestion

c. Organic matter conversion plus benthal stabilization

FIGURE 15.3 Types of aerobic lagoons (vertical dimension exaggerated): (a) stabilization through conversion of biodegradable organic matter, (b) organic matter conversion and aerobic digestion, and (c) organic matter conversion plus benthal stabilization.

Further removal and stabilization of biodegradable organic matter can be accomplished in a variety of ways. One approach, illustrated in Figure 15.3b, is simply to provide a larger HRT to allow aerobic digestion of the synthesized biomass and any organic solids that entered via the influent. This can be accomplished by constructing a larger CMAL or by constructing several CMALs in series. Lagoons in series provide a slight benefit in terms of overall stabilization, as discussed in Section 13.2.5. Another approach, illustrated in Figure 15.3c, is to provide a lagoon with lower mixing energy in which the biosolids leaving the initial CMAL are removed by gravity sedimentation, stabilized by benthal processes, and stored for later disposal.[37,38] Benthal stabilization involves anaerobic digestion and the end products are methane, carbon dioxide, organic acids, and nutrients such as ammonia-N. If an oxygen concentration of at least 2 mg/L is maintained in the clear water zone overlying the benthal layer, the reduced products (such as organic acids) will be oxidized as they pass through the upper portion of the settled solids.[37,38] Steps must also be taken to minimize the growth of algae in the settling lagoon. This is generally accomplished by providing a minimal level of mechanical aeration and by limiting the HRT in the overlying clear water zone to a value less than the minimum SRT for algal growth.[30,37,42]

15.1.2.4 Comparison of Lagoon Systems

Table 15.1 summarizes the benefits and drawbacks of the various lagoon systems. Anaerobic lagoons are inexpensive, easy to operate, and effective. They can provide significant wastewater flow and load equalization for downstream treatment processes since their HRT is relatively large. They can also be used to degrade waste biomass from downstream aerobic processes. Solids production is low and methane is produced, which can be collected and used as an energy source. Effective destruction of pathogens is also obtained because of the relatively long SRTs, a benefit with many wastewater types. On the other hand, process control is poor because of the lack of mixing and biomass retention systems. Many ANL systems are not covered, except for scum and debris that accumulates at the lagoon surface, and thus they can be significant odor sources. Lagoons can be covered using membranes and other devices, but odor release can still occur from the reactor inlets and outlets. Furthermore, odors will be released when the cover is removed for periodic lagoon cleaning. Although significant removal of biodegradable organic matter can be accomplished with an anaerobic lagoon, effluent quality may be relatively poor, requiring further treatment. Significant land area is also required due to the relatively low allowable process loadings.

Facultative and facultative/aerated lagoons are inexpensive to construct and operate, and are also simple to operate. Solids production is low because some of the biodegradable organic matter is stabilized anaerobically, and residual solids are stored in the lagoon. The large solids storage capacity allows solids to be managed on a periodic basis, which simplifies operations. Flow can also be equalized. Significant pathogen destruction can occur as a result of natural die-off and because the wastewater is exposed to sunlight. On the other hand, large land areas are required and process control is relatively poor. Furthermore, process performance is unreliable; while good quality effluent can be produced, these periods are interspersed between periods of excessive algal discharge with resulting poor effluent quality. Nutrient removal can be achieved in F/AL systems, but performance is not predictable and effluent quality is inconsistent. Odor potential is also significant because of the large land area and the diurnal variations in oxygen concentration caused by the diurnal availability of sunlight.

Aerobic lagoons are simple and inexpensive to construct, and are simple to operate. Operating costs are moderate, but higher than for anaerobic or facultative lagoons because of the aeration equipment. However, unlike the other lagoon processes, a high quality effluent can be reliably produced. Periodic solids management is possible in systems that incorporate benthal stabilization basins. Solids production from such systems is also low because of anaerobic stabilization of the removed suspended solids. No solids management is required for systems that maintain solids in suspension and discharge them with the effluent, but effluent quality is much poorer. Odor production

TABLE 15.1
Lagoon Process Comparison

Process	Benefits	Drawbacks
Anaerobic lagoon (ANL)	• Simple construction • Low capital and operation and maintenance costs • Simple operation • Low solids production • Effective and efficient • Energy (methane) recovery possible • Flow and load equalization • Large bioreactor volume to dilute inhibitors • Pathogen destruction • Suitable for a wide range of wastewaters • Efficiently handles high suspended solids wastewaters • Capable of accepting waste aerobic biomass	• Relatively large bioreactor volumes required • Large land area required • Poor process control • Significant odor potential • Effluent may require further treatment • Significant land area required
Facultative and facultative/aerated lagoon (F/AL)	• Simple construction • Low capital and operation and maintenance costs • Simple operation • Flow equalization • Low solids production • Periodic solids management • Pathogen destruction	• Large land areas required • Poor process control • Unreliable process performance • Odor potential
Aerobic lagoon (AEL)	• Simple construction • Low capital costs • High effluent quality • Simple operation • Periodic solids management • Low sludge production possible • Little odor production	• Significant land area required • Moderate operation and maintenance costs

is minimal because aerobic conditions are maintained on a consistent basis. Land requirements are also significant, although much less than for a facultative lagoon.

15.1.3 Typical Applications

Lagoons are widely used for the treatment of both municipal and industrial wastewater and their use is expected to continue. Table 15.2 summarizes typical lagoon applications.

Because of their low cost and effectiveness, ANLs are often used to pretreat industrial and municipal wastewaters. Their ability to accept wastewaters with high concentrations of suspended solids represents an advantage relative to the anaerobic processes described in Chapter 14. The significant removal of biodegradable organic matter that occurs in an ANL results in a significant reduction in the size of a downstream aerobic biological treatment system, with associated reductions in energy requirements and solids production. Industrial wastewaters containing high concentrations

TABLE 15.2
Typical Lagoon Applications

Anaerobic Lagoons	Facultative and Facultative/Aerated Lagoons	Aerobic Lagoons
Pretreatment of industrial wastewater prior to: downstream mechanical treatment system downstream natural treatment system discharge to municipal wastewater treatment system	Pretreatment of municipal and industrial wastewater prior to: downstream mechanical treatment system downstream natural treatment system	Pretreatment of industrial wastewater prior to: discharge to municipal wastewater treatment system
Pretreatment of municipal wastewater prior to: downstream F/AL system downstream natural treatment system	Treatment prior to surface water discharge (existing systems only)	Pretreatment of municipal and industrial wastewater prior to: downstream natural treatment system Treatment prior to surface water discharge

of readily biodegradable organic matter, on the order of 20,000–30,000 mg/L as COD, are frequently pretreated in ANLs. The downstream treatment system can be located at the industry or it may be a municipal system that the industry discharges to. Anaerobic lagoons are used less frequently to pretreat municipal wastewaters because of their lower strength, but such applications are known.[21,43,45] Anaerobic lagoons are also used frequently to pretreat and store animal wastes, such as from feedlots, prior to land application.

Facultative and facultative/aerated lagoons have been used historically to treat municipal and industrial wastewater.[40,43] Many municipalities served by lagoon systems are small (less than 7500 m^3/day), although some relatively large systems exist (in excess of 75,000 m^3/day). A similar wide range in sizes exists for F/ALs serving industries. The storage capacity of F/ALs is particularly attractive for land application systems where significant wastewater storage may be required to accommodate cropping cycles and weather patterns. The treatment provided in the F/AL minimizes nuisances associated with wastewater storage, and the aerobic "cap" on the lagoon minimizes the discharge of odoriferous materials from the underlying anaerobic zone. Facultative and facultative/aerated lagoons also provide a measure of pathogen control for municipal wastewaters, a benefit for these applications. Facultative and facultative/aerated lagoons, sometimes coupled with ANLs, are frequently used for municipal and industrial wastewater treatment in developing countries because of their low cost and good pathogen control.

Given today's stringent surface water discharge standards, F/AL effluents must generally be further treated before they can be discharged to lakes and streams. U.S. water pollution control regulations recognize the generally poor effluent quality produced by F/AL systems and establish specific discharge limits for this technology.[43] However, these revised limits apply only to cases where the receiving stream is not water quality limited. Mechanical treatment systems such as dissolved air flotation and filters or microscreens have been applied to F/AL effluents, but with generally poor results. Intermittent sand filters and rock filters have been used successfully, along with wetland treatment systems.[21,36,43]

Aerobic lagoons were developed initially to treat wastewater from the pulp and paper industry, where they were referred to as aerated stabilization basins. Since then they have been used to treat a wide variety of industrial[1,8–10,16–19,22] and municipal wastewaters.[4,7,15,21,22,37,43] Many of these systems are actually F/ALs because the aerator power is inadequate to fully suspend biosolids, resulting in significant algal growth and poor effluent quality. However, improved understanding of the factors that control algal growth and affect effluent quality has made it possible to design AEL systems that produce an effluent acceptable for discharge to surface waters.[36,37,40] As a consequence, AEL

systems are increasingly being used for complete treatment of municipal wastewater prior to surface water discharge, particularly those that couple a CMAL with a downstream benthal stabilization basin. They can also be used as pretreatment systems for industrial or municipal wastewaters.

15.2 FACTORS AFFECTING PERFORMANCE

The factors that affect the performance of lagoons are similar to those that affect the performance of other biochemical operations.

15.2.1 Solids Retention Time/Hydraulic Residence Time

As with other biochemical operations, the primary factor determining the performance of a lagoon system is the SRT. Because a CMAL behaves like a continuous stirred tank reactor (CSTR) without biomass recycle, its SRT is equal to its HRT; that is, the reactor volume divided by the flow rate, as shown in Equation 4.15. Consequently, CMAL performance relationships are often expressed as functions of the HRT. Experience confirms that the performance of a CMAL is as predicted by the simple model presented in Chapter 5. Typical performance is illustrated in Figure 15.4, which shows that the removal of soluble five day biochemical oxygen demand (BOD_5) is essentially complete for the lowest HRT tested of five days.[4] Total BOD_5, which includes both residual biodegradable organic matter and synthesized biomass, is removed to a much lower extent, although its removal increases with increasing HRT because of increased biomass decay and destruction of biodegradable particulate organic matter. The BOD_5 of the settled effluent, which includes residual biodegradable soluble organic matter plus nonsettleable biomass and nonsettleable residual biodegradable particulate organic matter, is also reduced to a significant extent. Its removal increases only slightly with increasing HRT. Although a settling basin or pond would be required to produce the settled effluent, a significant improvement in effluent quality results from its addition.

Rich and coworkers[29–32,37,38,47] have demonstrated that the concentration of biodegradable soluble organic matter can be accurately predicted using the simplified model presented in Chapter 5. This is demonstrated in Figure 15.5, which compares model predictions to full-scale CMAL performance data.[33] The comparison suggests that the model is conservative (i.e., the actual concentrations are less than or equal to predicted concentrations), thereby showing its utility in practice. Consistent with our understanding of other aerobic suspended growth systems, such comparisons also show

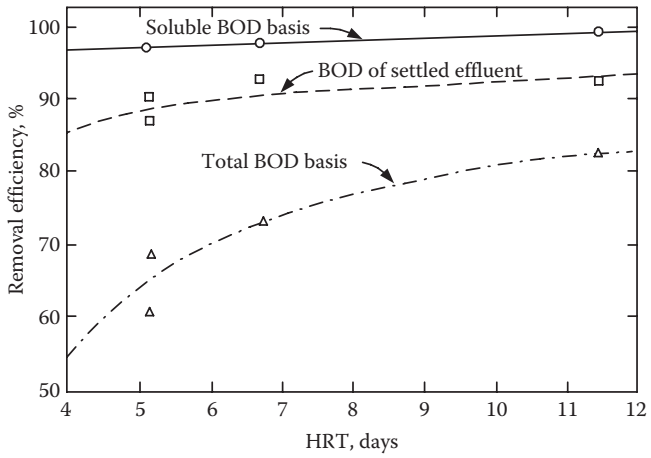


FIGURE 15.4 Performance of a CMAL treating municipal wastewater. (Adapted from Balasha, E. and Sperber, H., Treatment of domestic wastes in an aerated lagoon and polishing pond. *Water Research*, 9:43–49, 1975.)

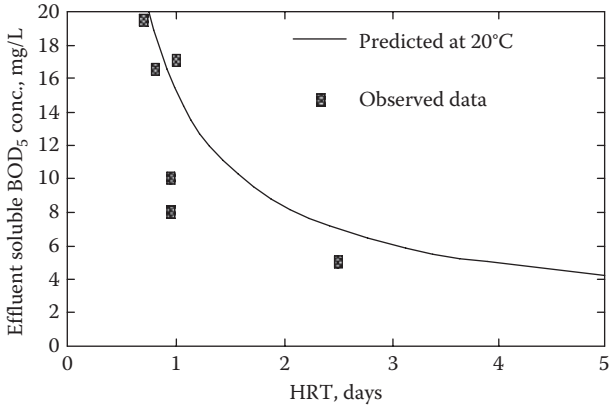


FIGURE 15.5 Comparison of simplified model of Chapter 5 to observed CMAL performance. The data shown were taken by L. G. Rich from Fleckseder, H. R., and Malina, J. F., *Performance of the Aerated Lagoon Process*, Technical Report CRWR-71, Center for Research in Water Resources, University of Texas, Austin, TX, 1970. (From Rich, L. G., *Technical Note No. 1, Aerated Lagoons,* Office of Continuing Engineering Education, Clemson University, Clemson, SC, 1993. Reprinted with permission.)

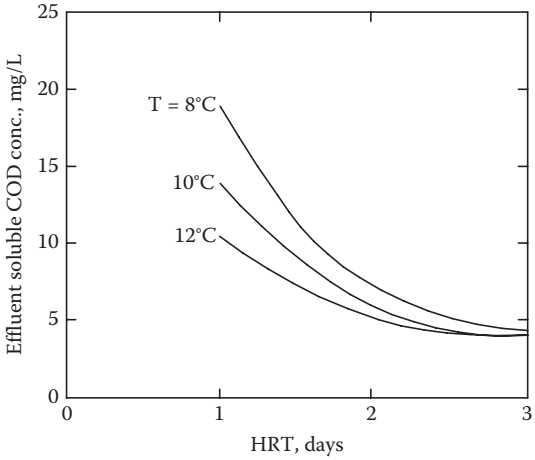


FIGURE 15.6 Effluent soluble COD from a CMAL as predicted by IWA ASM No. 1. (From Rich, L. G., *Technical Note No. 3, Aerated Lagoons,* Office of Continuing Engineering Education, Clemson University, Clemson, SC, 1994. Reprinted with permission.)

that substantial removal of soluble organic matter can generally be obtained at SRTs on the order of two to three days.

Rich[35,37] further demonstrated that International Water Association (IWA) activated sludge model (ASM) No. 1 can be used to predict the fate of readily and slowly biodegradable organic matter in CMAL systems. Predictions of the fate of readily (i.e., soluble) biodegradable COD are illustrated in Figure 15.6.[34] They demonstrate that its removal will be nearly complete at HRTs as short as two to three days, even at relatively low temperatures. Similar HRTs also result in essentially complete degradation of slowly biodegradable organic matter.[35,37]

A further consideration for an AEL process in which settleable solids are to be removed in a settling basin is the production of a settleable biomass. Experience with activated sludge systems, as discussed in Chapters 10 and 11, suggests that this can occur at SRTs as low as about one day,

and that effective flocculation will be accomplished at SRTs on the order of two to three days. Experience with full-scale AEL systems confirms these expectations.

The kinetics of algal growth in lagoon systems are not as well characterized as the kinetics of bacterial growth. However, data from Toms et al.[42] suggests an effective, average algal maximum specific growth rate of 0.48 day^{-1}, which corresponds to a minimum SRT of about two days. This is consistent with the observation that algal growth can typically be minimized in a single lagoon by maintaining an HRT of less than two days. Thus, an SRT (HRT) of two days is a reasonable "benchmark" of whether algae can grow.

Although the SRT is equal to the HRT for CMALs, the bioreactor configuration is not as well characterized or controlled for other types of lagoons, making calculation of the SRT less straightforward.[21,40,43] Anaerobic lagoons could be sized by assuming complete mixing with no biomass recycle or retention, but this would give a very conservative design because experience indicates that significant biomass retention occurs due to the settling of particulate matter, which provides surfaces for growth of anaerobic bacteria. This makes the SRT greater than the HRT. The mixing pattern in facultative and facultative/aerated lagoons is quite complex and changes on a diurnal, daily, and seasonal basis. Consequently, it is not possible to calculate the SRT for such systems. However, since long HRTs are used, the SRTs are also quite long.

15.2.2 Volumetric Organic Loading Rate

Just as for anaerobic processes, the volumetric organic loading (VOL) is the mass of biodegradable organic matter applied to the lagoon system per unit time per unit volume. It is denoted symbolically by $\Gamma_{V,S}$ and is expressed in units of kg COD or $BOD_5/(m^3 \cdot day)$. It is calculated with Equation 14.1 in terms of the flow rate and reactor volume or with Equation 14.2 in terms of the HRT. As illustrated by Equation 14.4, the VOL for a suspended growth process is related to the SRT by the concentration of active biomass in the reactor and the biomass net process yield. For processes such as anaerobic and facultative lagoons, it is not possible to determine the active biomass concentration. However, the VOL can be used to characterize process performance just as it was used for the various anaerobic processes described in Chapter 14 for which biomass concentrations could not be determined. Thus, it is useful as a design parameter for these more poorly defined systems. Volumetric organic loadings for anaerobic lagoons generally range from 0.2–1 kg $BOD_5/(m^3 \cdot day)$,[21,40,43,45] although somewhat higher loadings can be used in some instances. For facultative and facultative/aerated lagoons they generally range from 0.1–0.3 kg $BOD_5/(m^3 \cdot day)$.[21,40,43,45] Since the SRT is well defined for CMAL processes, VOL need not be used to design them. However, the VOL is related to the volumetric oxygen requirement, and can be used to estimate whether adequate oxygen transfer can be achieved.

Lagoon performance will deteriorate slightly as the VOL is increased within the applicable loading range, but will deteriorate rapidly as the VOL is increased beyond the applicable range. This is consistent with the typical relationship observed between process performance and SRT. Process performance is relatively independent of the SRT for values significantly in excess of the minimum SRT for the process, but becomes quite sensitive to SRT as it approaches the minimum value. Limiting VOLs corresponds to ensuring adequate SRTs.

15.2.3 Areal Organic Loading Rate

The areal organic loading (AOL) is the mass of biodegradable organic matter applied per unit time per unit lagoon surface area, A_L. It is denoted symbolically as $\Gamma_{A,S}$, and is typically expressed in units of kg COD or $BOD_5/(ha \cdot day)$. It is calculated as

$$\Gamma_{A,S} = \frac{F(S_{SO} + X_{SO})}{A_L}. \tag{15.1}$$

The logic behind this loading parameter is made obvious when it is recognized that it is used for lagoons in which the growth of algae is necessary. It represents the balance between the loading of biodegradable organic matter and the production of oxygen by algal growth, which is dependent on the penetration of light across the lagoon surface area. Consequently, it is typically used to characterize the performance of facultative and facultative/aerated lagoons.[21,40,43,45]

As with the other performance parameters, facultative lagoon performance is relatively insensitive to AOL over a wide range of loadings, but deteriorates rapidly as the normal range is exceeded. Acceptable values of AOL vary widely from one geographic location to another. For example, values on the order of 50–70 kg BOD_5/(ha · day) can be used quite successfully in the southeastern portion of the United States, while loadings of no more than 20–40 kg BOD_5/(ha · day) must be used in the more northern regions. This range reflects variations in ambient temperature and solar radiation over the annual cycle. Local experience and practice should be consulted to select the appropriate AOL for a particular application.

The AOL can also be used to characterize loadings when particulate biodegradable organic matter is being stabilized anaerobically in benthal deposits. Again, the process oxygen demands are being balanced with the supply of oxygen to the process. Typically, the AOL should not exceed 80 g biodegradable volatile solids (VS)/[m^2 · day; about 115 g biodegradable COD/(m^2 · day)] for AEL processes with benthal stabilization.[13,37,38] Mechanical aeration is used to maintain an oxygen concentration in the overlying liquid of 2 mg/L or more. At this loading and residual oxygen concentration, the reduced products (such as organic acids, hydrogen sulfide, etc.) diffusing from the anaerobic benthal deposits are oxidized by a thin aerobic layer at the top of the deposits. Higher loadings or lower oxygen concentrations result in anaerobic conditions throughout the benthal deposit and the release of odoriferous compounds to the overlying water column and, subsequently, to the atmosphere. Likewise, facultative solids storage lagoons, which provide long-term storage of anaerobically digested solids and maintain an aerobic "cap" by algal growth in the overlying clear fluid, are typically sized based on an AOL of 22.5 kg VS/(ha · year).[43]

15.2.4 Mixing

The degree of mixing provided in lagoons differs dramatically from type to type. These differences, coupled with variations in lagoon configuration, result in differences in flow patterns.[21,23,37]

Mechanical mixing is generally not provided in anaerobic lagoons, although some mixing is contributed by gas evolution from the digesting material. The absence of intentional mixing results in settling of solids within the lagoon and the retention of biomass. However, the absence of controlled mixing also causes complex flow patterns and the potential for short-circuiting. Inlet and outlet locations and configurations can be selected to minimize short-circuiting.[21,23]

Mixing is provided in facultative lagoons by several mechanisms, including wind action, gas evolution, and thermal gradients caused by diurnal heating and cooling of the lagoon surface. As with anaerobic lagoons, the uncontrolled nature of the mixing results in poorly defined flow patterns and the potential for short-circuiting. Facultative lagoons tend to be well mixed, and must be configured as tanks in series to achieve some degree of plug flow. As with anaerobic lagoons, short-circuiting is controlled by proper selection of the inlet and outlet location and configuration. The reader is referred elsewhere for a more complete discussion of facultative lagoon configuration.[21,28,45]

Mixing is provided in F/ALs by an oxygen transfer device. The power input to that device generally determines the degree of mixing, but different volumetric power inputs are required to achieve complete mixing of soluble versus particulate constituents. A volumetric power input of about 1 kW/1000 m^3 is adequate to achieve a uniform concentration of dissolved species such as oxygen within the zone of influence of the aerator,[5,9] although this power input may not produce a residence time distribution reflective of complete mixing. In contrast, a volumetric power input on the order of 6–10 kW/1000 m^3 may be required to achieve uniform concentrations of settleable solids, with the input depending on the concentration.[11] Another consideration is the volumetric power input

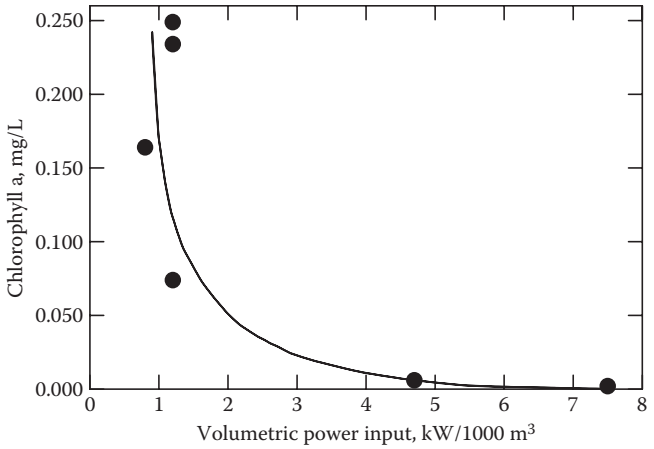


FIGURE 15.7 Relationship between aerator power density and chlorophyll a concentration for aerated lagoons. (From Rich, L. G., *Solids Control in Effluents from Aerated Lagoon Systems,* Report No. 73, Water Resources Research Institute, Clemson University, Clemson, SC, 1978. Reprinted with permission.)

required to suspend settleable solids. At power densities below 2 kW/1000 m^3 all settleable solids will be removed from suspension, leaving only nonsettleable solids.[28,29,31,32,37] The impacts of this are illustrated in Figure 15.7.[27] There it can be seen that algal concentrations in aerated lagoons (as indicated by the chlorophyll a concentration) decrease significantly at power densities above 2 kW/1000 m^3. This is because light penetration is reduced by the solids in suspension. A volumetric power input of at least 6 kW/1000 m^3 is required to completely suspend settleable solids when domestic wastewater is being treated.[5,9]

A more precise estimate of the volumetric power input required to suspend settleable solids is obtained from

$$\Pi = 0.004X_{M,T} + 5, \tag{15.2}$$

where Π is the volumetric power input in kW/1000 m^3 and $X_{M,T}$ is the suspended solids concentration in mg/L as total suspended solids (TSS).[29,31,37] This equation is valid for suspended solids concentrations less than about 2000 mg/L. At a suspended solids concentration of 250 mg/L, Equation 15.2 stipulates that a volumetric power input of 6 kW/1000 m^3 is needed.

Volumetric power inputs to facultative/aerated lagoons are generally below the level required to maintain solids in suspension. As a consequence, sunlight can penetrate the water column and algae will grow. In an aerobic lagoon, on the other hand, the volumetric power input will be sufficiently high to maintain solids in suspension, light penetration will be prevented, and algae will not grow. Thus, the volumetric power input dramatically affects the environmental conditions in the lagoon, and the resulting treatment efficiency. As discussed in Section 15.1, algae do not generally settle in lagoons and pass into the effluent, deteriorating its quality.

The presence of mechanical mixing in aerobic lagoons results in more controlled mixing conditions, and these bioreactors can be characterized as completely mixed. Aerobic lagoons are generally constructed as rectangular structures with sloping sidewalls. Length-to-width ratios between 1:1 and 3:1 are typical. The reader is referred to Murphy and Wilson[23] for further discussion of the factors affecting mixing patterns in aerobic lagoons.

The number and placement of aerators also affects the performance of facultative/aerated and aerobic lagoons. The turnover time for an aerator, which is an indication of its mixing ability, is a function of its size. The lower the turnover time, the more closely the lagoon approaches complete mixing. The turnover time of the aerator is determined, in turn, by its pumping capacity. The

pumping capacity for a surface aerator varies inversely with its size and approaches a maximum value of about 4.2×10^{-3} m^3/(sec · kW).[22] Thus, it is normally better to use several small units, rather than one large unit, as this will result in a higher overall pumping rate and a smaller turnover time. Surface aerators also have an effective volume and distance, which are the volume and horizontal distance they can mix effectively. The effective distance also varies with the aerator size[12] and has a reported maximum value of 0.8 m/kW. Regardless of the aerator size, the distance between adjacent aerators should not exceed about 75 m. Minimum spacings should also be maintained to avoid interactions between adjacent aerators.[26,48] Surface aerators also have effective depths. Consequently, as the depth increases, a draft tube may be required to allow the aerator to mix the entire lagoon depth. Conversely, in a shallow lagoon a concrete pad may be needed beneath the aerator to prevent erosion of soil or destruction of the lagoon liner. Surface aerator manufacturers can provide further information on the specific characteristics of their aerators.

15.2.5 Temperature

Temperature affects the biological activity in a lagoon in the same fashion as in other biochemical operations. A primary difference is that the large lagoon surface area results in higher heat losses and, typically, in lower temperatures. This is exacerbated in facultative/aerated and aerobic lagoons by the action of the oxygen transfer device if a mechanical surface aerator is used. Freezing will typically occur in cold climates, resulting in a significant decline in treatment efficiency. Significant cooling will occur even in warm climates, resulting in reduced reaction rates that must be compensated for in the process design.

A heat balance for a lagoon requires consideration of a number of components. Potential heat inputs include the influent wastewater, solar radiation, the aeration device (if one is provided), and biological reactions. Potential heat losses include the effluent wastewater, radiation to the atmosphere, evaporation, surface convection, convection/conduction through the wall of the lagoon, and the aeration device. Nevertheless it is possible to obtain a preliminary estimate of lagoon temperature by considering just the lagoon surface area and the appropriate temperatures:[11]

$$\frac{T_L}{T_i} = \frac{62.4(F/A_L) + H'(T_A - 2)/T_i}{62.4(F/A_L) + H'}. \tag{15.3}$$

This is a dimensional equation in which T_L is the weekly average lagoon temperature in °F, T_i is the weekly average influent wastewater temperature in °F, T_A is the weekly average air temperature in °F, F is the influent wastewater flow rate in ft^3/day, A_L is the surface area of the lagoon in ft^2, and H′ is the average heat transfer coefficient, which is generally taken to be 145 BTU/(ft^2 · day · °F). Experience indicates that predictions using this model are relatively insensitive to the specific value of the heat transfer coefficient.

Other workers have developed more sophisticated lagoon temperature models that require more input data; examples include the models of Talati and Stenstrom[41] and Argaman and Adams.[2] Table 15.3 summarizes the required inputs for the model of Talati and Strenstrom.[41] These models can provide quite precise estimates of bioreactor temperature when adequate information is available.[39]

15.2.6 Other Factors

Like other biochemical operations, the performance of lagoons is affected by other environmental conditions such as pH and nutrients. Lagoons are often used to treat industrial wastewaters, which can be nutrient deficient and nutrients may have to be added. One advantage of many lagoon processes is that the produced solids are accumulated and anaerobically digested, which will allow the removed nutrients to be recycled. However, the rate of digestion may vary seasonally, with

TABLE 15.3
Inputs to the Heat Loss Model of Talati and Stenstrom

Site-Specific Data	Process Data	Physical Properties
• Latitude of plant site, degrees	• Tank dimensions (L × W × H), m	• Air density, kg/m^3
• Ambient air temperature, °C	• Wastewater flow rate, m^3/day	• Water density, kg/m^3
• Wind speed, m/s	• Influent temperature, °C	• Specific heat of air, cal/(kg·°C)
• Relative humidity, percent	• Airflow rate (for diffused air), m^3/s	• Specific heat of water, cal/(kg·°C)
• Cloud cover, 10ths	• Number of aerators	• Emissivity of water
• Atmospheric radiation factor	• Aerator spray area, m^2	• Reflectivity of water
	• Power input to aerator/compressor, hp	
	• Efficiency of air compressor (for diffused aeration), percent	
	• Substrate removal rate, kg COD removed/day	
	• Overall heat transfer coefficient for tank walls, cal/(m^2·day)	
	• Humidity factor for exit air	

Source: Talati, S. N., and Stenstrom, M. K., Aeration-basin heat loss. *Journal of Environmental Engineering,* 116:70–86, 1990.

increased rates during warmer temperatures and reduced rates during colder temperatures so seasonal variations in nutrient recycle may occur. This can result in seasonal variations in net nutrient requirements. Significant pH variations can be observed on a diurnal basis in lagoons where algae are active. In such systems, pH values can easily range from 7 to 10 on a diurnal basis. Such pH variations do not appear to inhibit the biological activity within a lagoon, although operation with sustained pH values greater than eight or less than six should be avoided. The presence of toxic materials will adversely impact the performance of a lagoon. The large volume and HRT of a typical lagoon provides significant equalization, which dampens adverse short-term variations in the influent wastewater characteristics.

15.3 PROCESS DESIGN

The design of anaerobic, facultative, and facultative/aerated lagoons is based on experience and various rules of thumb and will not be discussed here. Such a design approach must be used because of the poorly defined bioreactor flow patterns and the complexity of the biochemical processes occurring. The design of anaerobic lagoons is similar to the design of high-rate anaerobic processes, as discussed in Chapter 14, and the design of facultative and facultative/aerated lagoons is discussed in various design manuals.[21,43,45]

Aerobic lagoons use well-mixed bioreactors, and the biochemical reactions occurring within them are primarily bacterial. As a consequence, the principles presented previously in this book can be used for their design. This is the focus of this section and the steps involved are listed in Table 15.4. The examples will consider a wastewater with the characteristics listed in Table E15.1. Section 9.6 discusses how to translate between the various ways of expressing wastewater characteristics.

15.3.1 Completely Mixed Aerated Lagoons

A completely mixed aerated lagoon (illustrated in Figure 15.3a) can be characterized as a CSTR with no biomass recycle. It is normally sized to produce an effluent with a specified concentration of soluble organic substrate. The design is typically based on the simplified model presented in

TABLE 15.4
Summary of Completely Mixed Aerated Lagoon (CMAL) Process Design Procedure

1. Summarize process design and loading conditions including maximum, minimum, and average sustained temperature; maximum, minimum, average, and design process influent flows and pollutant loadings; and desired effluent quality.
2. Where necessary, convert process loadings, expressed in conventional parameters (such as BOD_5) into the units used in the process design (such as biodegradable COD).
3. Define the process objectives and select the process configuration (CMAL, CMAL with sludge stabilization, or CMAL with benthal stabilization).
4. Select or calculate the design HRT for the CMAL.
5. Calculate the CMAL temperature for minimum, average, and maximum temperature conditions. If necessary, adjust CMAL size to ensure that treatment objectives are met over the full range of operating conditions.
6. Calculate the steady-state oxygen requirement for maximum, minimum, and average sustained temperature conditions. Consider whether nitrification will occur under these various conditions when completing these calculations.
7. Calculate the diurnal maximum and minimum oxygen requirements for the conditions above.
8. Calculate the power required to transfer the oxygen needed as determined in step 7.
9. Calculate the TSS concentration and, based on this, the volumetric power input required to maintain solids in suspension. Provide the greater of this power or the power required to transfer necessary oxygen as computed in step 8.
10. For benthal stabilization basins, calculate the area required based on the biodegradable suspended solids loading.
11. Calculate the depth for storage of inert organic suspended solids and for wintertime storage of biodegradable suspended solids, based on the desired storage period.
12. Calculate the process oxygen requirement caused by the stabilizing suspended solids.
13. Calculate the volume necessary to transfer the process oxygen requirement from step 12 while simultaneously providing adequate mixing and allowing suspended solids to settle.
14. Determine whether the clear water zone needs to be staged to minimize algae growth.
15. Based on the above, make any necessary adjustments in the process design and summarize the results in tabular form.

TABLE E15.1
Wastewater Strength and Treatment Characteristics for Design Examples

Wastewater Strength		Kinetic and Stoichiometric Coefficients	
X_{SO}	84 mg/L as COD	$\hat{\mu}_{H,20}$	10 day^{-1}, $\theta = 1.08$
S_{SO}	440 mg/L as COD	K_S	120 mg/L as COD
$X_{IO,T}$	45 mg/L as TSS	$Y_{H,T}$	0.52 mg biomass TSS/mg biodegradable COD
S_{NHO}	35 mg/L as N	$b_{H,20}$	0.15 day^{-1}, $\theta = 1.04$
X_{NSO}	7 mg/L as N	f_D	0.20 mg debris TSS/mg biomass TSS
Temperature	25°C	$i_{O/XB,T}$	1.2 mg COD/mg TSS
Flow	Equalized to 2400 m^3/day	$\hat{\mu}_{A,20}$	0.8 day^{-1}, $\theta = 1.11$
Air temperature	−5°C minimum weekly average	$K_{NH,20}$	1.0 mg/L as N, $\theta = 1.14$
		$Y_{A,T}$	0.20 mg biomass TSS/mg N
		$b_{A,20}$	0.05 day^{-1}, $\theta = 1.04$

Chapter 5, although IWA ASM No. 1 can also be used.[35] The process design of a CMAL requires three decisions: the bioreactor volume required, the amount of oxygen that must be supplied, and the minimum power input required to achieve adequate mixing and oxygen transfer. The basis for these decisions was discussed in Chapter 5 and Section 11.3, and it will be reviewed in this section.

For a CMAL, the SRT is equal to the HRT. Equation 5.22 provides the effluent soluble substrate concentration from a CSTR as a function of the SRT. Setting the SRT equal to the HRT, inserting Equation 4.15 for the HRT, and rearranging Equation 5.22 gives:

$$V = \frac{F(K_S + S_S)}{S_S(\hat{\mu}_H - b_H) - K_S b_H}. \tag{15.4}$$

The design volume is generally determined for wintertime operating temperatures because biological activity will be the lowest then and the largest volume will be required. Because the temperature in the lagoon depends on its size and the kinetics of biomass growth depends on the temperature, an iterative procedure is required using Equation 15.4 to calculate the bioreactor volume with temperature corrected coefficients and Equation 15.3 to calculate the bioreactor temperature. Example 15.3.1.1 illustrates the use of the simplified model to determine the required bioreactor volume.

Example 15.3.1.1

Consider a wastewater with the characteristics given in Table E15.1. A CMAL is to be designed to reduce the soluble organic substrate concentration to 10 mg/L as COD under all conditions. The lagoon is to have an average depth of 3.0 m. Determine the lagoon volume needed to meet the effluent requirement year round.

a. What bioreactor volume would be required at 20°C?
   Since the kinetic parameters are given for 20°C, it is convenient to start the process by determining the volume required at that temperature. Using Equation 15.4 and the values from Table E15.1:

$$V = \frac{(2400)(120+10)}{10(10-0.15)-(120)(0.15)} = 3880 \text{ m}^3.$$

   Since the flow rate is 2400 m^3/day, this provides an HRT of 1.62 days. In addition, the surface area would be 1293 m^2.

b. What would the lagoon temperature be when the air temperature is at its minimum weekly average value of –5°C?
   A rough estimate of the lagoon temperature can be obtained with Equation 15.3, which is dimensional using English units. Therefore, the inputs must be converted using the factors in Appendix C:

$$F = 2400 \text{ m}^3/\text{day} = 84{,}755 \text{ ft}^3/\text{day},$$

$$A_L = 1293 \text{ m}^2 = 13{,}917 \text{ ft}^2,$$

$$T_i = 25°C = 77°F,$$

   and

$$T_A = -5°C = 23°F.$$

   Taking the heat transfer coefficient H′ to be 145 BTU/(ft^2·day·°F), Equation 15.3 gives:

$$\frac{T_L}{77} = \frac{62.4(84{,}755/13{,}917)+145(23-2)/77}{62.4(84{,}755/13{,}917)+145} = 0.799.$$

   Therefore, the minimum weekly average lagoon temperature will be

$$T_L = 61.5°F = 16.4°C.$$

This is less than 20°C, the value that was used to calculate the bioreactor volume. Consequently, an iterative procedure must be used until the assumed and calculated temperatures are equal.

c. What bioreactor volume is required for the minimum weekly average temperature?
The last iteration, with an assumed temperature of 12°C (54°F), is summarized below. The kinetic parameter values are corrected for temperature using the temperature coefficients in Table E15.1 and Equation 3.99:

$$\hat{\mu}_{H,12} = (10)(1.08)^{12-20} = 5.40 \text{ day}^{-1},$$

$$b_{H,12} = (0.15)(1.04)^{12-20} = 0.11 \text{ day}^{-1},$$

and

$$V = \frac{(2400)(120+10)}{[10(5.40-0.11)-(120)(0.11)]} = 7860 \text{ m}^3.$$

This gives an HRT of 3.28 days.

$$A_L = \frac{7860}{3} = 2560 \text{ m}^2 = 27{,}556 \text{ ft}^2,$$

$$\frac{T_L}{77} = \frac{62.4(84{,}755/27{,}556)+145(23-2)/77}{62.4(84{,}755/27{,}556)+145} = 0.687,$$

$$T_L = 53°\text{F} \ (12°\text{C}).$$

Given the approximate nature of Equation 15.3, this is close enough to the assumed value. Thus, the calculated lagoon volume can be used. For design purposes use a volume of 8000 m^3, which corresponds to an HRT of 3.3 days. The difference between this volume and the one calculated in (a) demonstrates that temperature has an important impact on the design.

The oxygen requirement for heterotrophic activity must include both the average (steady state) and transient state requirements, just as required for activated sludge design, as discussed in Sections 11.3.2 and 11.3.3. The average oxygen requirement for the heterotrophs, RO_H, can be calculated using Equation 10.10 with the HRT substituted for the SRT. The resulting relationship is

$$RO_H = F(S_{SO} + X_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \tau)Y_{H,T}i_{O/XB,T}}{1 + b_H \cdot \tau}\right]. \tag{15.5}$$

The transient state oxygen requirement for the heterotrophs, $RO_{H,TS}$, is calculated using Equation 11.12.

If the HRT is sufficiently long for the operating temperature, nitrifying bacteria will be able to grow and oxygen must be provided to meet their metabolic requirements. If this is not done, adequate oxygen concentrations will not be maintained, resulting in impaired performance and potential odor problems. Just as for activated sludge design, the minimum SRT required for nitrification can be estimated from Figure 10.4 or calculated with Equation 5.25a, which appears in Section 11.3.2. Recognizing that for a CMAL the minimum SRT is equal to the minimum HRT (τ_{min}) allows Equation 5.25a to be modified to

$$\tau_{min} = \frac{K_{NH} + S_{NHO}}{S_{NHO}(\hat{\mu}_A - b_A) - K_{NH} \cdot b_A}, \tag{15.6}$$

where S_{NHO} is the influent ammonia-N concentration and the parameters are all for autotrophic biomass. If the HRT exceeds τ_{min}, then the effluent ammonia-N concentration can be calculated using Equation 11.19, with the HRT substituted for the SRT. The effects of dissolved oxygen (DO) concentration on the effluent ammonia-N concentration, S_{NH}, are considered in Equation 11.19 through inclusion of the DO safety factor, ς_{DO}. Unlike activated sludge systems, however, engineering control is not often exerted over the oxygen concentration in a CMAL. Consequently, the DO safety factor is normally set equal to 1.0 when calculating average effluent ammonia-N concentrations from CMALs. Incorporating these changes into Equation 11.19 gives:

$$S_{NH} = \frac{K_{NH}(1/\tau + b_A)}{\hat{\mu}_A - (1/\tau + b_A)}. \tag{15.7}$$

The average oxygen requirement for nitrification, RO_A, can be calculated using Equation 11.16 with the HRT inserted in place of the SRT. The result is

$$RO_A = F(S_{N,a} - S_{NH})\left[4.57 - \frac{(1 + f_D \cdot b_A \cdot \tau)Y_{A,T}i_{O/XB,T}}{1 + b_A \cdot \tau}\right]. \tag{15.8}$$

The nitrogen available to the nitrifiers, $S_{N,a}$, is calculated as discussed in Section 11.3.3 and given by Equation 11.17. The transient state oxygen requirement for the nitrifiers, $RO_{A,TS}$, is calculated using Equation 11.13 or 11.15, as appropriate. The total oxygen requirement for nitrification is the sum of the average and transient state oxygen requirements.

As with other aerobic processes, the oxygen requirements are calculated for the high summertime temperature, because that is when biological activity is greatest, resulting in the highest requirements. Calculation of the oxygen requirement is illustrated in the following example.

Example 15.3.1.2

The oxygen requirement for the CMAL considered in Example 15.3.1.1 is to be calculated. The average influent concentrations are given in Table E15.1. Even though the flow is equalized, transient loadings result in elevated pollutant concentrations in the CMAL influent. The peak influent concentrations are

$$X_{SO} = 84 \text{ mg/L as COD},$$

$$S_{SO} = 677 \text{ mg/L as COD},$$

$$S_{NHO} = 54 \text{ mg/L as N},$$

$$X_{NSO} = 7 \text{ mg/L as N}.$$

The oxygen requirement must be calculated at the high summertime temperature. To be conservative, assume that no cooling of the influent wastewater occurs, resulting in a lagoon temperature of 25°C.

a. What is the average oxygen requirement for the heterotrophs?
   Equation 15.5 is used. The influent slowly, X_{SO}, and readily biodegradable, S_{SO}, organic matter concentrations are 84 and 440 mg/L as COD, respectively, for a total of 524 mg/L (g/m^3). During the wintertime, the effluent soluble substrate concentration, S_S, is 10 mg/L as COD, and it will be even smaller at the higher summertime temperatures. Thus, S_S can be neglected relative to the influent substrate concentration when calculating the oxygen

requirement. The value of the heterotrophic decay coefficient, b_H, must be adjusted to 25°C by using the temperature coefficient given in Table E15.1 and Equation 3.99:

$$b_{H,25} = (0.15)(1.04)^{25-20} = 0.18 \text{ day}^{-1}.$$

The steady-state heterotrophic oxygen requirement is then calculated with Equation 15.5 using an HRT of 3.3 days:

$$RO_H = (2400)(524)\left\{1 - \frac{[1+(0.20)(0.18)(3.3)](0.521)(1.20)}{1+(0.18)(3.3)}\right\}$$

$$= 706{,}000 \text{ g } O_2/\text{day} = 706 \text{ kg } O_2/\text{day}.$$

b. What is the transient oxygen requirement for the heterotrophs?
The transient state oxygen requirement for the heterotrophs is calculated using Equation 11.12. This requires knowledge of the transient state increases in the mass flow rates of readily and slowly biodegradable substrate, $\Delta(F \cdot S_{SO})$ and $\Delta(F \cdot X_{SO})$, respectively:

$$\Delta(F \cdot S_{SO}) = (2400)(677) - (2400)(440) = 569{,}000 \text{ g COD/day} = 569 \text{ kg COD/day},$$

and

$$\Delta(F \cdot X_{SO}) = (2400)(84) - (2400)(84) = 0.0 \text{ kg COD/day}.$$

Substitution into Equation 11.12 gives:

$$RO_{H,TS} = [1 - (0.521)(1.2)](569) = 213 \text{ kg } O_2/\text{day}.$$

c. What is the total heterotrophic oxygen requirement?
The total heterotrophic oxygen requirement is $RO_H + RO_{H,TS}$, or

$$706 + 213 = 919 \text{ kg } O_2/\text{day}.$$

d. Will the process nitrify in the summer?
Determine whether autotrophs will grow in the summer using Equation 15.6 and the average influent ammonia-N concentration. Since the summer lagoon temperature is 25°C, the kinetic parameters for the autotrophs must be corrected to that value with the temperature coefficients in Table E15.1 and Equation 3.99:

$$\hat{\mu}_{A,25} = (0.8)(1.11)^{25-20} = 1.35 \text{ day}^{-1},$$

$$K_{NH,25} = (1.0)(1.14)^{25-20} = 1.92 \text{ mg/L as N},$$

$$b_{A,25} = (0.05)(1.04)^{25-20} = 0.06 \text{ day}^{-1}.$$

Substitution of these values into Equation 15.6 gives:

$$\tau_{min} = \frac{1.92+35}{35(1.35-0.06)-(1.92)(0.06)} = 0.82 \text{ day}.$$

Since this is well below the HRT of 3.3 days, nitrification will occur.

e. What will the effluent ammonia-N concentration be?
Calculate the effluent ammonia-N concentration with Equation 15.7:

$$S_{NH} = \frac{1.92\left(1/3.3 + 0.06\right)}{1.35 - \left(1/3.3 + 0.06\right)} = 0.7 \text{ mg/L as N.}$$

f. What is the average oxygen requirement for nitrification?
The average oxygen requirement for nitrification can be calculated with Equation 15.8. This requires knowledge of the concentration of nitrogen available to the nitrifiers, which can be calculated with Equation 11.17. The use of Equation 11.17 requires knowledge of the nitrogen requirement for the heterotrophs, NR, which can be calculated with Equation 11.18:

$$NR = 0.087\left\{\frac{\left[1+(0.20)(0.18)(3.3)\right](0.521)(1.20)}{1+(0.18)(3.3)}\right\}$$

$$= 0.038 \text{ mg N/mg COD removed.}$$

This can now be used in Equation 11.17. Recognizing that the effluent soluble organic substrate concentration is negligible compared to the influent gives:

$$S_{N,a} = 35 + 7 - (0.038)(440 + 84) = 22 \text{ mg/L as N.}$$

This, then, can be substituted into Equation 15.8 to calculate the steady-state autotrophic oxygen requirement, RO_A:

$$RO_A = (2400)(22 - 0.7)\left\{4.57 - \frac{\left[1+(0.20)(0.06)(3.3)(0.24)\right]}{1+(0.06)(3.3)}\right\}$$

$$= 223{,}000 \text{ g } O_2\text{/day} = 223 \text{ kg } O_2\text{/day.}$$

g. What is the transient oxygen requirement for nitrification?
The transient oxygen requirement for nitrification is calculated using Equation 11.13, which requires knowledge of the change in the nitrogen available to the nitrifiers, $\Delta(F \cdot S_N)_{a,TS}$. It must be calculated with Equation 11.14, which requires that several changes in mass flow rates be known:

$$\Delta(F \cdot S_{NHO}) = (2400)(54) - (2400)(35) = 45{,}600 \text{ g N/day} = 45.6 \text{ kg N/day,}$$

$$\Delta(F \cdot S_{NSO}) = (2400)(0) - (2400)(0) = 0.0 \text{ kg N/day,}$$

$$\Delta(F \cdot X_{NSO}) = (2400)(7) - (2400)(7) = 0.0 \text{ kg N/day,}$$

$$\Delta(F \cdot S_{SO}) = (2400)(677) - (2400)(440) = 569{,}000 \text{ g COD/day} = 569 \text{ kg COD/day,}$$

and

$$\Delta(F \cdot X_{SO}) = (2400)(84) - (2400)(84) = 0.0 \text{ kg COD/day.}$$

Substitution of these values into Equation 11.14 gives:

$$\Delta(F \cdot S_N)_{a,TS} = 45.6 - (0.087)(0.521)(1.20)(569) = 14.7 \text{ kg N/day.}$$

Substitution of this value into Equation 11.13 gives the transient state oxygen requirement associated with nitrification:

$$RO_{A,TS} = [(4.57 - (0.20)(1.20)](14.7) = 63.4 \text{ kg } O_2/\text{day}.$$

h. What is the total oxygen requirement for nitrification?
The total oxygen requirement for nitrification is $RO_A + RO_{A,TS}$, or

$$223 + 63.4 = 286.4 \text{ kg } O_2/\text{day}.$$

i. What is the total oxygen requirement for the lagoon?
The total oxygen requirement is the sum of the heterotrophic and autotrophic requirements, or

$$919 + 286 = 1205 \text{ kg } O_2/\text{day}.$$

After the lagoon volume and oxygen requirement are known, the power input required to achieve complete mixing and suspension of settleable solids is then calculated and compared to the power input required to transfer the needed oxygen. The required volumetric power input for mixing is calculated with Equation 15.2 and multiplied by the lagoon volume to obtain the total power input for mixing. Equation 15.2 requires knowledge of the bioreactor suspended solids concentration. That concentration can be calculated using Equation 10.8 after simplification by recognizing that, for a CMAL, the SRT and HRT are identical. The modified equation is

$$X_{M,T} = X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \tau) Y_{H,T} (S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \tau}. \tag{15.9}$$

Suspended solids concentrations should be calculated for the lowest sustained bioreactor temperature since that will result in the highest value because of the low biomass decay. If sufficient information is not available to use Equation 15.9, a volumetric power input of 6 KW/1000 m^3 can be used as an approximation.

The power input for oxygen transfer can be calculated based on the efficiency of the oxygen transfer device used. High speed surface aerators are often used in CMALs, and these devices have in-process efficiencies on the order of 1.0–1.1 kg O_2/(kW · hr). Values for other oxygen transfer devices are available from the manufacturers of the equipment and in various manuals of practice.[44,46] These calculations are illustrated in the following example.

Example 15.3.1.3

Determine the power that must be supplied to the CMAL considered in Examples 15.3.1.1 and 15.3.1.2. This must be done by calculating the power required to mix the lagoon and keep the solids in suspension and the power required to deliver the needed amount of oxygen, and then choosing the larger of the two.

a. What is the TSS concentration?
The TSS concentration is calculated using Equation 15.9 for the average process loading condition and the low weekly CMAL temperature of 12°C. We saw in Example 15.3.1.1 that b_H has a value of 0.11 day^{-1} at that temperature. Using the HRT of 3.3 days, Equation 15.9 gives:

$$X_{M,T} = 45 + \frac{[1 + (0.20)(0.11)(3.3)](0.521)(440 + 84 - 10)}{1 + (0.11)(3.3)}$$

$$= 256 \text{ mg/L as TSS}.$$

b. What is the power input required for mixing and biomass suspension?
The volumetric power input for mixing and biomass suspension can be estimated with Equation 15.2 using the solids concentration of 256 mg/L:

$$\Pi = (0.004)(256) + 5 = 6 \text{ kW/1000 m}^3.$$

At a bioreactor volume of 8000 m^3, the total power required for mixing is

$$(8000)(6) = 48 \text{ kW}.$$

c. What power input is required for oxygen transfer?
The peak oxygen requirement, as calculated in Example 15.3.1.2, is 1205 kg O_2/day, or 50.2 kg O_2/hr. Assuming the use of an aerator with an efficiency of 1.0 kg O_2/(kW · hr), the power required for oxygen transfer is 50.2/1.0 = 50.2 kW. Consequently, the power required for oxygen transfer will control. When that amount of power is supplied, the lagoon will be adequately mixed. The selection of specific aerators and the determination of their locations in the actual lagoon geometry may result in the need for some adjustment to the design and power input.

15.3.2 Completely Mixed Aerated Lagoon with Aerobic Solids Stabilization

Conversion of organic matter into biomass results in stabilization of a portion of the oxygen demand contained in the wastewater. This occurs because a fraction of the organic matter, roughly $[1 - (Y_{H,T}\ i_{O/XB,T})]$, must be oxidized to provide energy for biomass synthesis. Since the numerical value of $Y_{H,T}\ i_{O/XB,T}$ is on the order of 0.6 mg biomass COD synthesized/mg COD removed, about 40% of the biodegradable organic matter in the influent wastewater will be stabilized by a CMAL sized for conversion of organic matter to biomass. As demonstrated above, organic matter can be converted to biomass at relatively short HRTs in CMALs. Additional stabilization of organic matter can be obtained by increasing the HRT (SRT) to allow the synthesized biomass to decay.

Aerobic stabilization is accomplished by oxidation of biodegradable organic matter, whether in the form of the original substrate or in the form of biomass derived from that substrate. Thus, the efficiency of stabilization in a lagoon can be calculated as the steady-state oxygen requirement associated with the oxidation of organic matter divided by the loading of biodegradable organic matter to the lagoon. The fraction of the influent biodegradable organic substrate aerobically stabilized, Ξ, can be calculated as follows:

$$\Xi = \frac{RO_H}{F(S_{SO} + X_{SO})}. \tag{15.10}$$

The bioreactor volume required to achieve a specified degree of stabilization can be calculated with the following equation, which follows from substituting Equation 15.5 into Equation 15.10 and assuming that the residual soluble substrate concentration is negligible:

$$V = \frac{F[(Y_{H,T} \cdot i_{O/XB,T}) - (1 - \Xi)]}{b_H[1 - \Xi - (f_D \cdot Y_{H,T} \cdot i_{O/XB,T})]}. \tag{15.11}$$

Again, the bioreactor volume is calculated at the lowest sustained operating temperature since reactions rates are slowest then. Bioreactor volumes can escalate significantly if a high degree of stabilization is desired under all weather conditions because larger lagoons lead to more cooling, which decreases reaction rates. Sizing a CMAL for solids stabilization is illustrated in the following example.

Example 15.3.2.1

Reconsider the wastewater used in the previous examples and size a CMAL to stabilize 55% of the biodegradable organic matter contained in the wastewater.

a. What degree of stabilization is achieved in the summer in the lagoon designed in Examples 15.3.1.1–15.3.1.3, and what does it suggest about the operating temperature in this situation?
The degree of stabilization can be estimated with Equation 15.10 by using the RO_H value of 706,000 g O_2/day calculated in Example 15.3.1.2:

$$\Xi = \frac{706{,}000}{(2400)(440+84)} = 0.56.$$

Thus, 56% stabilization is achieved under summer operating conditions. Since the new design must achieve almost this degree of stabilization in the winter, a larger lagoon will be required, which will make the winter operating temperature lower. Since the winter temperature was 12°C in the examples of Section 15.3.1, assume a winter temperature of 5°C for this situation. This assumption will have to be checked.

b. What size CMAL will be required to achieve 55% stabilization in winter?
The decay coefficient is the only kinetic parameter appearing in Equation 15.11, so only its value must be adjusted to the assumed temperature of 5°C:

$$b_{H,5} = (0.15)(1.04)^{(5-20)} = 0.08 \text{ day}^{-1}.$$

That coefficient is then used in Equation 15.11 to calculate the required volume.

$$V = \frac{(2400)[(0.521)(1.20)-(1-0.55)]}{(0.08)[1-0.55-(0.2)(0.521)(1.20)]} = 16{,}150 \text{ m}^3.$$

This is approximately twice as large as the CMAL required to give an effluent soluble substrate concentration of 10 mg/L, as determined in Example 15.3.1.1. The larger volume gives an HRT of 6.7 days.

Before the calculated volume can be considered to be acceptable, the assumed temperature must be verified through use of Equation 15.3, just as was done in Example 15.3.1.1. Such an exercise predicts that the lagoon temperature will be 6°C in the winter. This is sufficiently close to the assumed temperature to be acceptable. As with Example 15.3.1.1, if the calculated temperature had not been consistent with the assumed temperature an iterative procedure would have been required to calculate the required volume and resulting temperature.

Oxygen requirements and mixing requirements are calculated in the same manner as in Section 15.3.1. Oxygen requirements are calculated for summertime operating conditions, and the long HRTs required to achieve solids stabilization generally mean that nitrification will occur. The increased volume required for solids stabilization also means that bioreactor power inputs are more likely to be controlled by mixing and settleable solids suspension than by oxygen transfer.

Completely mixed aerated lagoon processes for solids stabilization can be configured as either a single CSTR or as CSTRs in series, as illustrated in Figure 15.3b. Significant improvements in performance result by using two lagoons in series rather than a single lagoon, but additional improvements are modest as the number of lagoons is increased further, just as was the case for aerobic digesters, as illustrated in Figure 13.12. The treatability of the wastewater will generally decrease as it moves from lagoon to lagoon because the more easily biodegradable organic matter is removed in the upstream lagoons. Consequently, the potential improvement in performance may be overestimated by equations that assume constant kinetics throughout a staged system,[22] such as those used

to generate Figure 13.12. Thus, a maximum of three lagoons in series is generally used. Caution should be exercised to avoid an initial stage that is so small that the oxygen requirement controls the power input, rather than the mixing requirement. It is not unusual to use unequal size lagoons in series, with the initial CMAL being larger so that power requirements for mixing and oxygen transfer are balanced. The initial CMAL may be 50–60% of the total lagoon volume, with the subsequent two lagoons of equal size.

15.3.3 Completely Mixed Aerated Lagoon with Benthal Stabilization and Storage

A CMAL with benthal stabilization, illustrated in Figure 15.3c, represents another approach to incorporating solids stabilization into aerobic lagoons. It offers the further benefit of providing a clarified effluent since settleable solids are removed in the benthal stabilization basin. The CMAL (i.e., the first basin) is designed as described in Section 15.3.1, while the benthal stabilization basin is designed for solids sedimentation, stabilization, and storage.

As illustrated in Figure 15.3c, settleable solids accumulate in the lower portion of the benthal stabilization basin where they are anaerobically digested. Aeration and mixing are provided to the upper portion to supply oxygen to serve as the electron acceptor for bacteria degrading the organic matter released from the anaerobically digesting solids. Aeration of the overlying water column also allows aerobic biodegradation of residual organic matter that escapes the upstream CMAL. Mixing promotes fluid movement, which minimizes stagnant regions where algae can grow. However, mixing levels must be kept low enough so that settleable solids do not remain in suspension. Furthermore, the HRT in the clear water zone above the benthal layer must be sufficiently short to minimize the growth of algae. From considering the above, it can be seen that requirements in the design of a benthal stabilization basin include: (1) providing sufficient volumetric power inputs to mix the upper portion of the basin and to meet oxygen requirements, while still allowing the sedimentation of settleable solids; (2) supplying a sufficiently large clear water zone to allow sedimentation of suspended solids, while also making it small enough to avoid algal growth; and (3) furnishing a sufficient solids digestion and storage zone for stabilization of the particulate organic matter applied. Experience indicates that these multiple, conflicting objectives can generally be achieved.

Although sedimentation of settleable solids is an important function of the clear water zone, it does not generally control the sizing of the basin since the basin's HRT and depth result in overflow rates that are at least an order of magnitude lower than those required for good solids sedimentation. Thus, other factors control its size.

Consider first the solids stabilization and storage requirements. The input of suspended solids into the benthal layer can be thought to consist of two components, one biodegradable and the other nonbiodegradable. The surface area of the benthal stabilization basin is determined by the input of biodegradable solids, while the depth of the benthal zone is determined primarily by the input and storage of nonbiodegradable solids.

First consider the surface area. If the biodegradable solids loading is sufficiently small, then those solids will not accumulate during the warm months when biodegradation is proceeding at its maximal rate. Rather, they will be anaerobically degraded into carbon dioxide and methane in the benthal zone. However, during cold winter months, biodegradable solids may accumulate because the lagoon temperature may be too cold for effective anaerobic decomposition. As a consequence, areal solids loadings are generally made small enough to allow time for biodegradable solids deposited during the winter to decompose during the subsequent summer. The steady-state areal loading rate for biodegradable suspended solids, $\Gamma_{A,XB}$, depends on the amount of biomass synthesized in the preceding CMAL. Consequently:

$$\Gamma_{A,XB} = \frac{F\left[\dfrac{(S_{SO} + X_{SO} - S_{S1})Y_{H,T} \cdot i_{O/XB,T}}{1 + b_H \cdot \tau_1}\right]}{A_L}. \tag{15.12}$$

The units on $\Gamma_{A,XB}$ are g biodegradable COD/($m^2 \cdot$ day). The bracketed term in the numerator of Equation 15.12 is the active heterotrophic biomass concentration (as COD) in the CMAL preceding the benthal stabilization basin, where τ_1 is the HRT of that CMAL. Settleable solids are generally deposited on both the bottom and the sloping sides of the benthal stabilization basin. Consequently, the entire water surface of that basin, A_L, is used in the calculation of the biodegradable areal solids loading. The required surface area, A_L, is calculated by adopting a reasonable value for the areal loading and using a rearranged form of Equation 15.12. A conservative value of about 115 g biodegradable COD/(m^2·day) [80 g biodegradable VSS/($m^2 \cdot$ day)] has been recommended by Rich[28,29,31,32,37,38] based on pilot tests and experience with full-scale systems.

Inert solids entering the benthal stabilization basin will not be biodegraded and will accumulate. The depth of the benthal zone depends on their rate of accumulation, the length of time over which they are allowed to accumulate, their concentration in the benthal layer, and the surface area of the basin. Inert solids include those present in the influent wastewater as well as those produced by decay of biomass produced in the upstream CMAL. The steady-state input (accumulation) rate for inert organic solids as TSS, $RX_{I,T}$, can be calculated as

$$RX_{I,T} = F\left[X_{IO,T} + \frac{f_D \cdot b_H \cdot \tau_1 \cdot Y_{H,T}(S_{SO} + X_{SO} - S_{SI})}{1 + b_H \cdot \tau_1}\right], \tag{15.13}$$

in which the right term within the bracket represents biomass debris arising from biomass grown in the preceding CMAL. Multiplication of this accumulation rate by the time period over which accumulation will occur gives the mass of inert organic solids to be stored. That time period depends on the means by which the solids are disposed of, as well as on the cost of land for the CMAL and benthal stabilization basin. However, three years is a common value. Space must also be provided for accumulation of the biodegradable suspended solids during the winter months when colder temperatures reduce anaerobic activity. Typical practice is to remove solids from a benthal stabilization basin in the fall before biodegradable solids start to accumulate because that provides the minimum quantity of unstabilized solids. Consequently, only enough storage volume is required for six months accumulation of biodegradable solids. The accumulation rate of biodegradable solids during the winter in COD units is just $\Gamma_{A,XB} \cdot A_L$ from Equation 15.12. It can be converted to TSS units by division by the conversion factor $i_{O/XB,T}$. After the mass of accumulated solids is known, the volume required to house them can be calculated by dividing by their concentration in the benthal layer. Concentrations in the 30–50 g/L (kg/m^3) range are commonly obtained. After the required volume is known, the depth can be calculated by dividing the volume by the basin area, A_L, calculated with Equation 15.12. Finally, some additional depth must be provided to allow space for the digesting biodegradable suspended solids. The following example illustrates the sizing of a benthal stabilization basin.

Example 15.3.3.1

Consider the CMAL from Example 15.3.1.1, which has an HRT of 3.3 days. A benthal solids stabilization basin is to be provided downstream to remove and stabilize settleable suspended solids. Determine the required surface area for the basin as well as the depth of the benthal zone if the basin is to be cleaned every three years in the early fall. Base the design on an areal loading rate of 115 g biodegradable COD/($m^2 \cdot$ day), a benthal solids concentration of 40 g/L as TSS of organic solids, and a temperature of 20°C.

a. What is the surface area of the benthal stabilization basin?
   The basin surface area can be calculated with Equation 15.12 by using a value of 115 g biodegradable COD/($m^2 \cdot$ day) for $\Gamma_{A,XB}$. The value of S_S may be neglected.

$$A_L = \frac{(2400)\left[\frac{(440+84)(0.521)(1.20)}{1+(0.15)(3.3)}\right]}{115} = 4570 \text{ m}^2.$$

b. What mass of inert organic suspended solids must be stored if the basin is cleaned every three years?
The mass of inert organic solids to be stored is the steady-state input rate times the storage period. The steady-state input rate of inert organic solids can be calculated with Equation 15.13:

$$RX_{I,T} = (2400)\left[45 + \frac{(0.2)(0.15)(3.3)(0.521)(440+84)}{1+(0.15)(3.3)}\right]$$
$$= 151{,}000 \text{ g/day} = 151 \text{ kg/day}.$$

Since the lagoon will be cleaned every three years, the total mass accumulated is

$$(151)(3)(365) = 165{,}000 \text{ kg of inert organic solids as TSS.}$$

c. What mass of biodegradable solids must be provided for?
Space must be provided for six months accumulation of biodegradable organic solids. Because this space must be available during the winter and the basin will be cleaned in the fall, not all of the inert solids will have accumulated when this space is required. However, by adding the mass of biodegradable solids to the total mass of inert solids calculated above, additional space will be provided to accommodate digesting solids.

The accumulation rate of biodegradable solids in COD units is just $\Gamma_{A,XB} \cdot A_L$. Dividing this by the conversion factor between COD and TSS units, $i_{O/XB,T}$, gives the accumulation rate in TSS units. Using this concept, the accumulation rate of biodegradable solids during the winter is

$$\frac{(115)(4570)}{1.20} = 438{,}000 \text{ g/day} = 438 \text{ kg/day as TSS.}$$

Since these solids will accumulate for up to six months, the total mass accumulated will be

$$(438)(0.5)(365) = 80{,}000 \text{ kg of biodegradable solids as TSS.}$$

d. What will the volume of the solids storage zone be?
Solids in the digesting benthal solids blanket will thicken to approximately 40 g/L (kg/m^3). Thus, the maximum solids volume will be approximately:

$$\frac{165{,}000 + 80{,}000}{40} = 6100 \text{ m}^3.$$

e. What will the depth of the solids storage zone be?
Since the basin surface area is 4570 m^2 the solids will accumulate to an average depth of

$$\frac{6100}{4570} = 1.3 \text{ m.}$$

f. What mass of solids must be disposed of?
The mass of solids removed at the end of three years is three years accumulation of inert solids. Because the solids will be removed from the basin at the end of the summer, all of the biodegradable solids will have been digested. The mass of inert solids accumulated over three years was calculated in part b. Thus, 165,000 kg as TSS of inert organic solids must be disposed of.

The clear water zone above the benthal solids is sized next. Several factors determine the volume of this zone. One is the trade-off between oxygen transfer and mixing. Sufficient oxygen must be transferred to meet peak demands exerted by the feedback of reduced compounds from the digesting benthal solids layer. Mixing levels must also be sufficiently low so that settleable solids do not remain in suspension.

Peak oxygen demands will generally be experienced in the spring and the fall of the year as the water temperature reaches about 16°C. Experience indicates a peak oxygen requirement during these periods of about 80 g $O_2/(m^2 \cdot day)$, again based on the lagoon surface area.[28,29,31,32,37,38] The power input required to meet this demand will depend on the efficiency of the aerator used. Just as with a CMAL, the mechanical surface aerators typically have an oxygen transfer efficiency on the order of 1.0–1.1 kg $O_2/(kW \cdot hr)$.

Once the power input for oxygen transfer is known, the volume of the clear water zone is chosen to keep the volumetric power input between 1.0 and 2.0 kW/1000 m^3. As discussed previously, at volumetric power inputs below 1.0 kW/1000 m^3 dissolved oxygen and other soluble constituents will not be uniformly distributed, while at volumetric power inputs above 2.0 kW/1000 m^3 settleable solids will remain in suspension.

The sizing of the clear water zone based on these considerations is illustrated in the following example.

Example 15.3.3.2

Size the clear water zone for the benthal stabilization basin considered in Example 15.3.3.1. Assume that the peak oxygen demand is 80 g $O_2/(m^2 \cdot day)$ and that the oxygen transfer device has a transfer efficiency of 1.0 kg $O_2/(kW \cdot hr)$. Maintain the volumetric power input between 1.0 and 2.0 kW/1000 m^3 during all operating conditions.

a. How much power is needed for oxygen transfer?
The lagoon surface area is 4570 m^2. At a peak oxygen demand of 80 g $O_2/(m^2 \cdot day)$, the peak oxygen requirement is

$$(4570)(80) = 365{,}600 \text{ g } O_2/\text{day} = 15.2 \text{ kg } O_2/\text{hr}.$$

Assuming the use of high speed surface aerators with an efficiency of 1.0 kg $O_2/(kW \cdot hr)$, the power required is 15.2 kW.

b. What is the volume of the clear water zone?
The minimum clear water zone volume will result when the volumetric power input to the clear water zone is maximized. At a maximum volumetric power input of 2 kW/1000 m^3, the volume of the clear water zone is $15.2 \div 2 = 7.6$, which has units of 1000 m^3. Thus, the volume is 7600 m^3.

c. What is the volumetric power input immediately after cleaning before solids have accumulated?
Before solids have accumulated, the entire basin acts like a clear water zone. The total basin volume is the volume of the clear water zone calculated in part b (7600 m^3) plus the maximum volume of the benthal solids zone (6100 m^3) for a total of 13,700 m^3. Thus, the initial volumetric power input will be $15.2 \div 13.7 \times 10^3 = 1.1$ kW/1000 m^3. This is sufficiently high to uniformly disperse oxygen throughout the basin.

A final consideration is the HRT in the clear water zone. An excessive HRT can lead to the growth of algae, which will pass into the effluent and deteriorate effluent quality. As discussed in Section 15.2.1, an HRT of two days or greater will generally result in excessive algal growth. In cases where the HRT exceeds two days, a series of completely mixed cells can be used to prevent the HRT in any one cell from exceeding two days. This works well from a practical perspective since multiple benthal stabilization basins are typically provided so that one can be taken out of service to allow the removal of stabilized solids while the remaining cells continue to remove solids coming from the upstream CMAL. Based on the work of Toms et al.,[42] Rich[30,35,37] developed an equation for the growth of algae in equal-sized completely mixed cells in series:

$$\frac{X_{PN}}{X_{PO}} = \frac{1}{\left[1 - \mu_P \frac{\tau}{N}\right]^N}, \tag{15.14}$$

where X_{PN} and X_{PO} are the concentrations of photosynthetic microorganisms (algae) in the effluent from the last completely mixed cell and the influent to the first completely mixed cell, respectively, N is the number of equal-sized completely mixed cells, μ_P is the specific growth rate of the photosynthetic microorganisms (taken to be 0.48 day^{-1}), and τ is the total HRT of the clear water zones of the series of cells. This equation is applicable for values of individual cell HRTs (τ/N) less than $1/\mu_P$ or 2.08 days. For values of individual cell HRTs greater than this, algal growth will not be controlled and X_{PN}/X_{PO} cannot be predicted by this equation. Based on experience with CMAL systems followed by benthal stabilization basins, Rich[35] recommended that X_{PN}/X_{PO} ratios be kept less than 25. Equation 15.14 is plotted in Figure 15.8 where it illustrates that the total HRT in the clear water zone can be increased significantly when multiple cells in series are used while still minimizing the growth of algae. The effect is not linear, since an increase from one to two cells produces a much larger increase in allowable total basin clear water zone HRT than does an increase from two

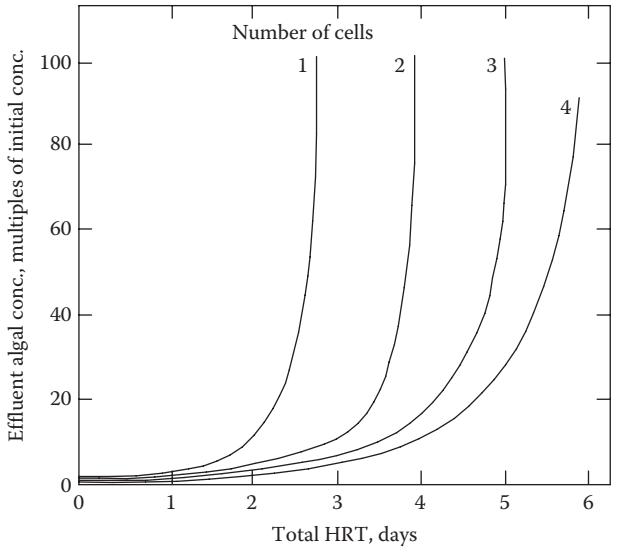


FIGURE 15.8 Effects of total clear water zone HRT and the number of equal size completely mixed cells in series on algal growth in lagoons. Theoretical curves from Equation 15.14. (From Rich, L. G., Modification of design approach to aerated lagoons. *Journal of Environmental Engineering*, 122:149–53, 1996. Copyright © American Society of Civil Engineers. Reprinted with permission.)

to three cells, and so on. The following example illustrates the use of this equation to configure a benthal stabilization basin.

Example 15.3.3.3

Determine what configuration should be used for the benthal stabilization basin sized in Examples 15.3.3.1 and 15.3.3.2 to minimize algal growth.

a. What are the HRTs in the basin and what do they suggest about the required configuration of the system?
   Since the flow to be treated is 2400 m^3/day, the HRT in the clear water zone is 7600 ÷ 2400 = 3.1 days and the HRT based on the total basin volume is 13,700 ÷ 2400 = 5.7 days. Since both of these exceed a value of two days, excessive algal growth is likely if the basin is configured as a single cell.
b. What is the minimum number of cells in series required to control algal growth?
   We saw in Section 15.2.1 that the minimum SRT for algal growth was around two days. This means that the maximum SRT per cell is around two days. Consequently, considering only the clear water zone, the stabilization basin would have to be divided into at least two equal size cells. Immediately after cleaning, light could penetrate throughout the basin, so the entire volume would be available for algal growth. Since the total HRT of the basin is 5.7 days, it would have to be divided into at least three equal sized cells to keep the HRT in any given cell below two days.
c. Based on Equation 15.14, how many equal sized cells in series should be used to prevent excessive algae growth?
   An HRT of two days in each cell does not guarantee that the effluent algal concentration will be acceptable. Equation 15.14 must be used to evaluate the effect of the number of cells on algal growth potential for the full range of operating conditions. Consequently, use it to calculate the X_{PN}/X_{PO} ratio based on the clear water zone and total basin volume for three, four, five, and six cells in series. For example, using the clear water zone volume that gives a total HRT of 3.1 days and three cells in series:

$$\frac{X_{PN}}{X_{PO}} = \frac{1}{\left[1 - 0.48\left(\frac{3.1}{3}\right)\right]^3} = 7,$$

   which is acceptable. The values of X_{PN}/X_{PO} are calculated for HRTs of 3.1 and 5.7 days and summarized in Table E15.2. The results shown there indicate that effective control of algae (X_{PN}/X_{PO} less than 25) can be achieved using three cells if the clear water zone HRT is restricted to 3.1 days. Algae control, on the other hand, is quite difficult if the full basin volume is used (i.e., such as immediately after cleaning). Consequently, use three cells along with the capability to adjust the lagoon depth to control the clear water zone HRT. The combination of these features should provide ample flexibility to control algal growth.

TABLE E15.2
Evaluation of the Potential for Algal Growth in the Benthal Stabilization Basin Considered in Example 15.3.3.3

	X_{PN}/X_{PO}			
	N			
Total HRT, Days	3	4	5	6
3.1	7	6	6	6
5.7	2680	117	58	42

This example illustrates the need for flexibility in the design of a benthal stabilization basin. It may be desirable to reduce the basin depth (and consequently the volume) immediately after solids have been removed from the basin so that the clear water zone HRT, and the corresponding potential for the growth of algae, is reduced. The depth would be increased later as solids accumulated to maintain the necessary clear water zone.

While degradation of most of the readily biodegradable organic matter occurs in the CMAL, further biodegradation will occur in the clear water zone of the benthal stabilization basin. This can be significant when a high quality effluent is desired. Although the bulk of the active heterotrophs are removed from the clear water zone by sedimentation, a residual concentration of about 4–6.5 mg/L as TSS remains. With the long HRT in the clear water zone and particularly when the benthal stabilization basin consists of multiple cells in series, effective removal of readily biodegradable substrate can be obtained. Because of the already low concentration of biodegradable organic matter, the first-order approximation of the Monod equation (Equation 3.38) can be used and applied to the clear water zone of the multicell benthal stabilization basin for prediction of the soluble substrate concentration in the N^{th} cell, S_{SN}. The relationship is

$$S_{SN} = \frac{S_{SCMAL}}{\left[1 + \left(\frac{\hat{\mu}_H}{Y_{H,T} \cdot K_S}\right) X_{B,H} \cdot \frac{\tau}{N}\right]^N}, \tag{15.15}$$

where S_{SCMAL} is the concentration of readily biodegradable organic matter in the effluent from the upstream CMAL, τ is the total HRT of the clear water zone of the multicell benthal stabilization basin, and N is the number of cells in the basin. Again, $X_{B,H}$ will be about 4 to 6.5 mg/L as TSS.

15.4 PROCESS OPERATION

Lagoon operation is quite straightforward. It generally involves placing the required number of units in service to achieve the desired VOL or AOL and performing necessary maintenance so that facilities are available for service.

One of the principal difficulties experienced with lagoon processes other than CMALs is odors. The potential for odor production is an inherent part of anaerobic and facultative lagoons because the anaerobic metabolism occurring in them forms odoriferous organic and inorganic end products. Emission of these products can be controlled by minimizing transport to a receptor. For some wastewaters, scum and other materials accumulate on the surface of the lagoon, sealing it, and restricting odor emissions. In other cases an overlying layer of relatively clean, oxygenated water is provided by encouraging the growth of algae in the overlying clear water (facultative lagoons), by providing devices that transfer oxygen into the overlying clear water (benthal stabilization basins), or by recirculating oxygenated water from a downstream aerobic unit. This latter approach is typically used when an anaerobic lagoon is followed by a facultative lagoon. Covers can also be added to anaerobic lagoons to contain odoriferous materials and collect the methane gas.

Organic overloads can cause odor problems when oxygen demands exceed oxygen supplies in any lagoon. Such upsets require that additional electron acceptor be added to match the increased organic loading. Effluent recirculation and the addition of mechanical aeration are two options. Other options include addition of hydrogen peroxide, which degrades to water and oxygen or nitrate-N, which serves directly as an alternative electron acceptor. The addition of nitrate-based fertilizers offers an easily implemented and effective option. The expense of these latter approaches relative to mechanical aeration generally relegates them to temporary use.

Solids accumulations occur in most lagoons and they must be removed when they begin to interfere with performance. One method of removing solids involves taking the lagoon out of service, removing the relatively clean water, and allowing the remaining solids to dry. Although a dry product

is produced for ultimate disposal, this method will keep the lagoon out of service for several weeks, which may adversely impact overall wastewater treatment. Alternatively, dredging can be used to remove accumulated solids while the lagoon remains in service. Odor emissions are generally a problem during lagoon cleaning since the solids are anaerobic and are exposed to the atmosphere. Solids accumulations can also occur in CMALs if the volumetric power inputs are inadequate or if the location and arrangement of aerators are inappropriate.[1,16,18]

The control of mixing energy levels and HRTs to control the growth of algae is particularly critical in aerobic lagoons. Sufficient mixing must be provided in CMALs to maintain settleable solids in suspension and block light penetration to minimize algal growth. In basins where suspended solids are allowed to settle and light can penetrate the water column, algal growth is controlled by maintaining the HRT of each cell in the basin sufficiently low so that algae are washed out of the system. Techniques available to control the HRT include adjusting the number of cells in service and controlling the basin depth. If the guidelines presented in Section 15.3.3 are followed, algal growth can be controlled and the associated adverse impacts on effluent quality can be avoided.

Icing of aerators can be a particular problem in cold climates.[41] Mechanical surface aerators spray bioreactor contents into the air and mist can condense and freeze on the aerator floats. Accumulations of ice can result in unbalanced loads on the floats, which cause them to tip over and sink. Heat tracing can be added to the aerator to prevent such accumulations.

Further information on the operation of lagoons is provided in various standard references.[21,37,40,43,45]

15.5 KEY POINTS

1. The term lagoon refers to a diverse array of suspended growth biochemical operations with no downstream clarifier. They are typically constructed as inground earthen basins, resulting in their name.
2. Lagoons are mechanically simple, but this masks a degree of physical, chemical, and biological complexity that exceeds that of other biochemical operations. They stabilize biodegradable organic matter by aerobic and anaerobic processes; remove nitrogen by nitrification, denitrification, and stripping; and remove phosphorus by precipitation.
3. In anaerobic lagoons (ANLs) organic matter is stabilized by conversion to carbon dioxide and methane. Mixing is provided by gas evolution and results in settling and retention of suspended solids. Consequently, the solids retention time (SRT) is longer than the hydraulic residence time (HRT).
4. In facultative and facultative/aerated lagoons organic matter is stabilized by both aerobic and anaerobic metabolism. Oxygen is provided by algae and, in some instances, also by mechanical means. Anaerobic metabolism occurs in the lower portion of the lagoon, and odoriferous compounds formed there are oxidized in the aerobic upper portion.
5. In aerobic lagoons organic matter is stabilized using oxygen as the electron acceptor. The growth of algae is limited by maintaining solids in suspension to limit light penetration and by maintaining the HRT in the clear water zone below the minimum value required for algal growth (generally about two days).
6. Lagoons are widely used to pretreat municipal and industrial wastewaters prior to treatment in a downstream treatment system.
7. In a completely mixed aerated lagoon (CMAL) the SRT and HRT are equal. Its performance can be characterized by the relationships presented in Chapter 5 or by the International Water Association (IWA) activated sludge model (ASM) No 1.
8. The volumetric organic loading (VOL) rate is used to characterize the performance of processes such as ANLs where mixing conditions allow the settlement and accumulation of biomass.

9. The areal organic loading (AOL) rate is used to characterize the performance of lagoons in which algae provide the oxygen. It is also used to characterize benthal stabilization processes where odiferous products released to the overlying clear water zone must be oxidized.
10. The volumetric power input provided by mechanical aeration systems determines the degree of mixing and the suspension of solids in an aerobic lagoon. A value of 1 kW/1000 m^3 provides uniform dispersion of dissolved species such as oxygen. A value less than 2 kW/1000 m^3 allows settleable suspended solids to settle, while 6–10 kW/1000 m^3 is generally needed to maintain all suspended solids in suspension.
11. The large surface area and long HRT generally used with lagoons result in significant heat loss during cold weather. Reduced temperatures cause lower biological activity and decreased treatment efficiency.
12. Three decisions must be made when a CMAL is designed: (1) the bioreactor volume, (2) the amount of oxygen to be supplied, and (3) the minimum power input required to achieve adequate mixing.
13. The effluent from a CMAL can be treated in a benthal stabilization basin where settleable solids are removed by sedimentation and stabilized by anaerobic processes.
14. Proper operation of a lagoon is achieved by maintaining appropriate process loadings. Temporary overloads can be mitigated by the addition of alternative electron acceptors such as hydrogen peroxide and nitrate-N. Solids are disposed of on a periodic basis, but accumulated solids must be removed when they begin to interfere with lagoon performance. Icing of aerators can be a particular problem in cold climates.

15.6 STUDY QUESTIONS

1. Prepare a table summarizing the benefits and drawbacks of lagoon processes. List the characteristics of the various lagoon processes and discuss where they are typically used.
2. Summarize the design approaches and types of design criteria used to size anaerobic lagoons, facultative and facultative/aerated lagoons, and aerobic lagoons. List typical values of the design criteria for each process.
3. Prepare a figure illustrating the impact of volumetric power input on mixing in an aerated lagoon. Distinguish between various mixing regimes and discuss the potential for algal growth in each of these regimes. What other approaches are available to control algal growth?
4. Using the wastewater characteristics listed in Table E9.4, the stoichiometric and kinetic parameters in Table E11.2, and the temperature correction factors in Table E11.1, size a CMAL to treat an average flow rate of 10,000 m^3/day to reduce the concentration of readily biodegradable substrate to 10 mg/L as COD. The range of weekly average air temperatures is 0–35°C, with an average of 20°C. The diurnal flow and pollutant concentration variations are as illustrated in Figure 6.2. Determine the required volume of the CMAL.
5. Determine the average and diurnal peak oxygen requirements and required power input for aeration for the CMAL sized in Study Question 4. Assume an oxygen transfer efficiency for the oxygen transfer device of 1.1 kg O_2/(kW·hr). Also determine the power required to adequately mix the CMAL. Which power input controls the design?
6. Use a simulation package implementing IWA ASM No. 1 to evaluate the design developed in Study Questions 4 and 5.
7. The CMAL designed in Study Questions 4 and 5 must be upgraded by adding another CMAL in series with the existing unit to achieve 50% stabilization of biodegradable organic matter at the minimum weekly average temperature. Determine the size of the CMAL that must be added. What will the efficiency of stabilization of biodegradable organic matter be under average temperature conditions; under maximum weekly temperature conditions?

8. Determine the oxygen requirement and required aeration power input for the new CMAL considered in Study Question 7. Assume an oxygen transfer efficiency for the oxygen transfer device of 1.1 kg O_2/(kW·hr). Also determine the power required to adequately mix the CMAL. Which power input controls the design?
9. Use a simulation package implementing IWA ASM No. 1 to evaluate the design developed in Study Questions 7 and 8.
10. The CMAL designed in Study Questions 4 and 5 is to be upgraded by the addition of a benthal stabilization basin. Size and configure the basin and determine the power input required to meet oxygen and mixing requirements.
11. Develop a figure that demonstrates the factors affecting the heat balance and resulting temperature of a lagoon. The work of Argaman and Adams[2] and of Stenstrom and coworkers[39,41] may be helpful. Discuss the relative importance of the various components of the heat balance.
12. Prepare a table comparing the benefits and drawbacks of the three lagoon processes considered in this chapter relative to their ability to meet effluent quality requirements for discharge to surface waters.

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Part IV

Theory: Modeling of Ideal Attached Growth Reactors

All of the biochemical operations in the preceding chapters have been suspended growth systems in which the biomass is suspended uniformly throughout the liquid phase. One assumption made in their modeling is that they behave as homogeneous systems; that is, that all microorganisms experience the dissolved constituents at the concentration of the liquid phase surrounding them. Even though the biomass exists in floc particles in most of those operations, no consideration was given to changes in concentration within the floc. This approach was acceptable because the effects of concentration gradients within floc particles are accounted for through quantification of the half-saturation coefficient, K_S, provided that the physical characteristics of the biofloc used in kinetic testing mimic closely the characteristics in the system being modeled. The modeling of attached growth systems is more complicated because this simplifying assumption cannot be made. The biomass in these systems grows as a biofilm attached to a solid support (usually impermeable) with the result that substrate and other nutrients can only get to the bacteria within the biofilm by mass transport mechanisms. Consequently, biofilms must be treated as heterogeneous systems in which the combined effects of reaction and transport are explicitly accounted for. In Chapter 16 we explore two approaches for doing this, the effectiveness factor and pseudoanalytical approaches. In Chapter 17 we use the pseudoanalytical approach to investigate the performance of packed towers, which behave like plug-flow systems, and rotating disc reactors, which behave like biofilms in continuous stirred tank reactors. Finally, in Chapter 18, we examine the theoretical performance of fluidized bed biological reactors, in which the biomass grows on particles suspended in the liquid phase, requiring us to also consider the hydrodynamics of particle fluidization.

16 Biofilm Modeling

The biochemical operations we have considered so far have all employed suspended growth cultures of microorganisms. However, as discussed in Section 1.2.3, attached growth bioreactors have been used extensively in environmental engineering practice. As the name implies, microorganisms in such bioreactors grow attached to a solid surface rather than being freely suspended in the wastewater undergoing treatment. That surface may be fixed in space with the wastewater flowing over it in thin sheets, such as in a packed tower, it may rotate about an axis, thereby moving through the fluid in the bioreactor, as in a rotating disc reactor, or it may be in the form of small particles that are held in suspension by the upward flow of water, as in a fluidized bed biological reactor. In all cases, however, the key distinguishing characteristic is that the microorganisms live in a biofilm attached to a surface. This means that the electron donor, the electron acceptor, and all other nutrients must be transported to the microorganisms within the biofilm by diffusional and other mass transport processes. It is the necessity to consider the combined effects of mass transport and reaction that makes the modeling of biofilm systems different from and more complicated than the modeling of suspended growth systems. In this chapter we will briefly examine the structure of biofilms and then review the ways in which the combined effects of transport and reaction are considered during modeling.

16.1 NATURE OF BIOFILMS

The existence of biofilms can be traced back 3.25 billion years, indicating that microbial cells have the intrinsic ability to "self-assemble" into highly integrated and structured communities.[22,39] Biofilms are very complex, both physically and microbiologically. In fact, they are so complex that it is impossible to fully explore all aspects of them in the space available here. Therefore, those interested in a more detailed explanation of their properties should consult the reviews by Costerton and colleagues.[10,22,39]

The basic conceptualization of a biofilm system is shown in Figure 16.1.[8] The biofilm grows attached to a solid support, which is usually impermeable, although it need not be. In this book, only impermeable supports will be considered. The solid support may be natural material, such as rock in old trickling filters, or it may be synthetic, such as the plastic packing in modern ones. Furthermore, it may range in configuration from corrugated sheets, such as in packed towers, to small particles, such as in fluidized beds. In general, the biofilm can be divided into two zones, the base film and the surface film. Both contain an assemblage of microorganisms and other particulate material bound together by a matrix of extracellular polymeric substances (EPS; see Section 2.4.3). Those polymers, which are excreted by the microorganisms, are thought to be the same as the polymers involved in bioflocculation (see Section 2.3.1). The base film consists of a structured accumulation, with well-defined boundaries. Transport in the base film has historically been viewed as being by molecular processes (diffusion), although, as we will see later that view is changing. The surface film provides a transition between the base film and the bulk liquid, and transport within it is dominated by advection. The relative thicknesses of the base and surface films depend largely on the hydrodynamic characteristics of the system, but also on the nature of the microorganisms in the biofilm. Consequently, one biofilm may have almost no surface film whereas another may be entirely surface film. There is normally relative motion between the biofilm and the bulk liquid, with the one moving depending upon the configuration of the attached growth process. For example, in packed towers the bulk fluid moves down over the biofilm in a thin sheet, whereas in a rotating disc reactor the biofilm support moves through the bulk liquid.

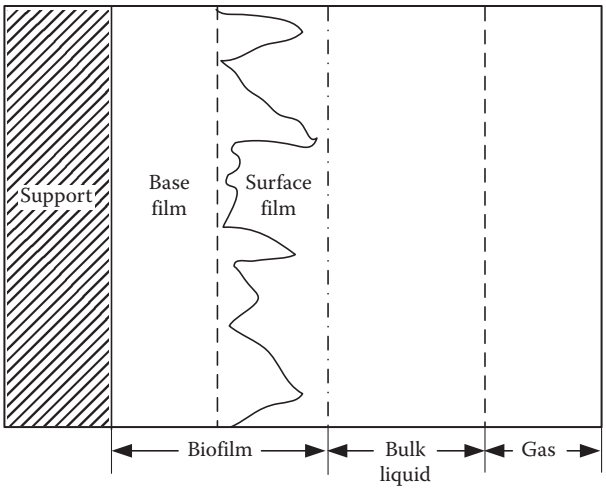

FIGURE 16.1 Conceptualization of a biofilm system. The base film and the surface film constitute the biofilm. (After Characklis, W. G. and Marshall, K. C., Biofilms: A basis for an interdisciplinary approach. *Biofilms*, eds. W. G. Characklis and K. C. Marshall, 3–15, Wiley, New York, 1990.)

In either case, however, mass transfer from the bulk fluid to the biofilm depends on the hydrodynamic regime. Finally, some biofilm systems contain a gas phase that provides oxygen or serves as a sink for gaseous products.

Simpler mathematical models of biofilm systems, such as the ones used later in this chapter, consider the surface film to be negligible, and thus consider only the base film. Furthermore, unless they are specifically trying to model a variety of events such as carbon oxidation, nitrification, and denitrification, they usually reflect a single species biofilm. Figure 16.2[49] shows such a biofilm. The bacterial cells can be seen to be suspended within a polymeric matrix, much the way fruit is held in a jello salad. From such pictures the concept developed that the transport of substrates, nutrients, electron acceptor, and so on to and from the bacteria within the biofilm is by molecular diffusion

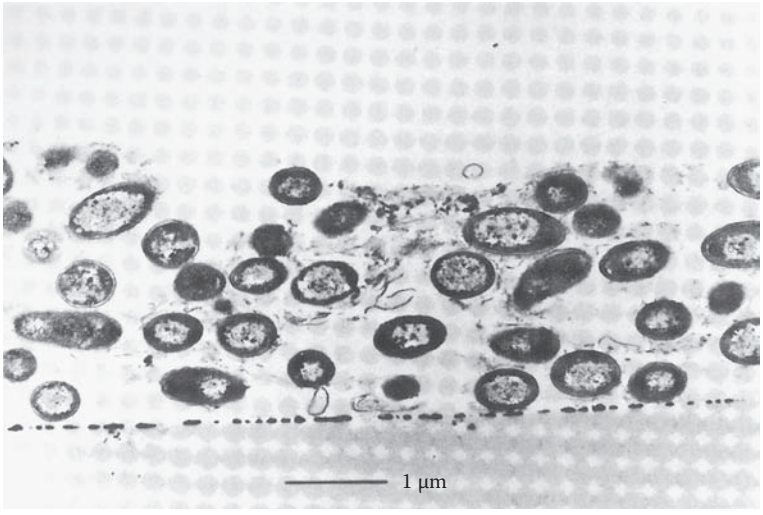

FIGURE 16.2 A transmission electron micrograph of a *Pseudomonas aeruginosa* biofilm consisting almost entirely of a base film. (From Wilderer, P. A. and Characklis, W. G., Structure and function of biofilms. *Structure and Function of Biofilms*, eds. W. G. Characklis and P. A. Wilderer, 5–17, Wiley, New York, 1989. Reprinted with permission from John Wiley & Sons, Inc.)

alone.[21] Transport between the bulk fluid and the biofilm, on the other hand, is dominated by advection and turbulent diffusion.[21] These concepts dominate all mathematical models today.

Due to the development of new tools for the study of biofilms, a different picture of the internal structure of the base film is emerging.[10,11,13] Figure 16.3a is an artist's conceptualization of the architecture of a biofilm based on the observations of several researchers.[10] This type of architecture has been confirmed experimentally, as shown in Figure 16.3b. Mathematical models now are able to recreate ab initio the complex architecture of biofilms, as illustrated in Figure 16.3c. The similarity of the three panels in Figure 16.3 is striking, indicating that the main factors responsible for defining biofilm structure are now reasonably well understood.

Biofilms are nonuniform structures consisting of discrete cell clusters attached to each other and to the solid support with EPS.[13,38] The spaces between clusters form vertical and horizontal voids, with the vertical voids acting as pores and the horizontal voids acting as channels. As a result, biomass distribution within a biofilm is not uniform,[10,26] nor are physical factors such as porosity and density.[52] In addition to the hydrodynamic conditions surrounding and within the biofilm, biological factors play important roles in defining the biofilm's overall architecture. Cells embedded within biofilms are phenotypically different from their planktonic (dispersed in solution) counterparts, as seen through different patterns of gene expression.[39] The cells in a biofilm do not behave independently of each other; rather, they communicate through the production of molecules that signal their presence, a process known as quorum sensing. Quorum sensing is thought to be involved in regulating EPS production,[39] which in turn determines biofilm structure. The cell clusters are microbial aggregates cemented with EPS, whereas the voids are open structures relatively free of it. The significance of the voids is that liquid can flow through them.[11] This has a profound effect on mass transfer in the biofilm because it suggests that it can occur by both diffusion and advection, with

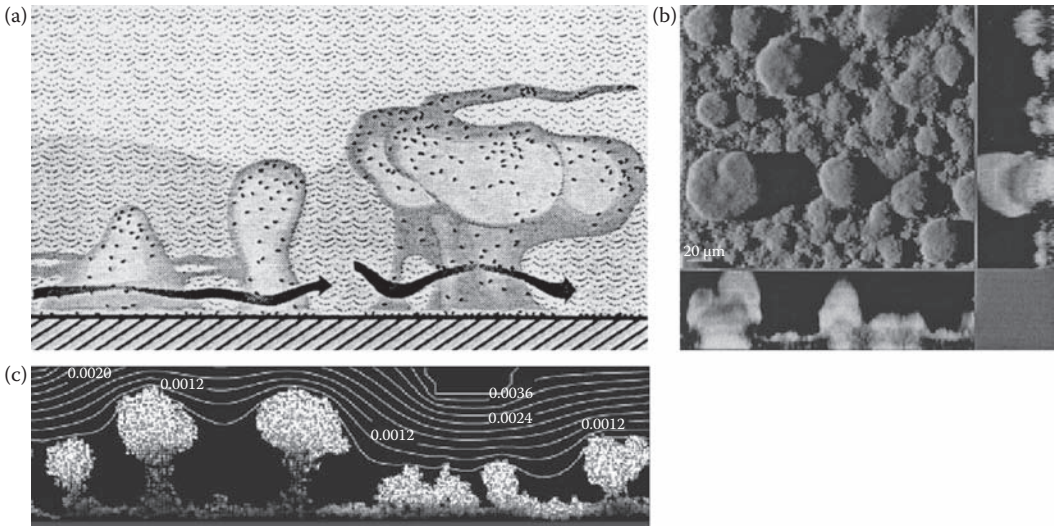

FIGURE 16.3 Architecture of a biofilm: (a) Artist's conceptualization of the architecture of a biofilm. (From Costerton, J. W., Lewandowski, Z., Caldwell, D. E., Korber, D. R., and Lappin-Scott, H. M., Microbial biofilms. *Annual Review of Microbiology*, 49:711–45, 1995. Copyright © Annual Reviews, Inc. Reprinted with permission.) (b) Confocal laser scanning micrograph of a *Pseudomonas aeruginosa* biofilm. (From Klausen, M., Heydorn, A., Ragas, P., Lambertsen, L., Aaes-Jorgensen, A., Molin, S., and Tolker-Nielsen, T., Biofilm formation by *Pseudomonas aeruginosa* wild type, flagella and type IV pili mutants. *Molecular Microbiology*, 48:1511–24, 2003. Copyright © Blackwell Publishing Ltd. Reprinted with permission.) (c) Two-dimensional biofilm architecture generated using a mathematical model describing the growth of a heterotrophic organism that generates extracellular polymeric substances. (From de B. Xavier, J., Picioreanu, C., and van Loosdrecht, M. C. M., A general description of detachment for multidimensional modelling of biofilms. *Biotechnology and Bioengineering*, 91:651–69, 2005. Copyright © Wiley Interscience. Reprinted with permission.)

diffusion dominating in the cell clusters. However, because advection brings materials to the clusters, diffusion can occur from almost any direction into a cluster, rather than just from the liquid-biofilm interface. In addition, it also appears that the cell clusters have small conduits through them, adding another level of complexity to the biofilm.[51] Finally, many factors, such as the texture of the substratum, the nature of the flow past the biofilm, and the geometry of the bioreactor, influence the heterogeneity of the biofilm that develops.[18] These observations suggest that the commonly accepted use of a single transport parameter, such as an effective diffusion coefficient, for describing the transport of substrate, electron acceptor, and so on within a biofilm is inadequate.[12,13] In fact, various researchers have shown that effective diffusion coefficients vary with biofilm depth,[37,52,55] which is consistent with changes in the structure of the biofilm. Nevertheless, because several mathematical models for biofilm reactors assume that transport within a biofilm is by diffusion alone with a constant diffusion coefficient, then that is the approach we will take herein. However, the reader should be aware of the limitations of such an approach.

The conceptual models presented above are for a simple heterotrophic biofilm in which the bacteria are using a single electron donor with a single electron acceptor. However, just as heterotrophic and autotrophic bacteria can grow together in suspended growth bioreactors, they can also grow together in attached growth reactors. In this instance they have different electron donors (organic matter and ammonia-N), but compete for the same electron acceptor (oxygen). They also must compete for space in the biofilm. The assumed spatial arrangements of the competing species within the biofilm can take several forms in mathematical models.[21] However, the most realistic approach assumes that all types of bacteria are available for growth at any point within a biofilm, but that their ultimate distribution is determined by their competition for shared nutrients and space,[21,31,44,45] which is consistent with observation.[54] Although the mathematical models for this competition were developed before the advent of the conceptual model in the preceding paragraph, we will consider them because of the importance of the interactions between heterotrophs and autotrophs in attached growth reactors.

The importance of competition for space in determining the ultimate distribution of competing species within a biofilm can be visualized by considering the traditional conceptualization of a base biofilm. Consider first a single species biofilm. Because substrate can only move into the biofilm by diffusion, a substrate concentration gradient will exist through the biofilm as illustrated by Figure 16.4. This means that bacteria near the liquid-biofilm interface are growing faster than those in the interior. However, as bacteria in the interior grow, they occupy more space, pushing those that are closer to the liquid-biofilm interface further away from the solid support. In addition, all of the bacteria are subject to decay, regardless of their position in the biofilm, resulting in the

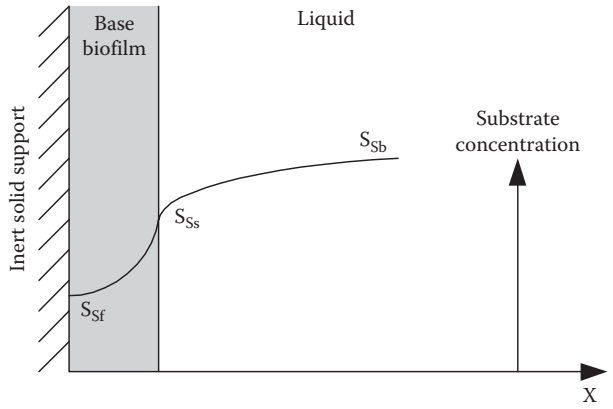

FIGURE 16.4 Traditional conceptualization of a base biofilm showing a typical concentration profile for a single limiting nutrient.

accumulation of biomass debris. The net effect of both processes is to cause a migration of particles from the interior of the film to the exterior where surface shear forces remove them, allowing a biofilm of constant thickness to develop. Even for a single species biofilm, however, the distribution of active organisms will not be the same throughout the depth of the biofilm.[52] Rather, active biomass will predominate in the outer regions of the film and biomass debris in the inner regions, as shown by the simulation results in Figure 16.5.[44]

If we have two species that do not compete for any nutrient, but only for space, their ultimate distribution will depend upon their relative specific growth rates at any point within the biofilm. Consider two species, A and B, growing on different substrates, but sharing oxygen as the electron acceptor. Oxygen is assumed to be present in excess, so as not to limit either species. Species A has a higher maximum specific growth rate coefficient on its substrate than species B does on its substrate. Species A will dominate the outer regions and species B will dominate the inner regions, as shown by the simulation results in Figure 16.6.[31] Species B is confined to the inner regions because there the substrate concentration for species A will have been diminished sufficiently to allow species B to grow as fast as, or faster than, species A. When the two species compete for a resource, such as oxygen, the distribution of organisms can become even more complex, depending upon the relative values of the half-saturation coefficient (K_S) for the shared resource, as well as the growth kinetics of each species on its individual substrate.

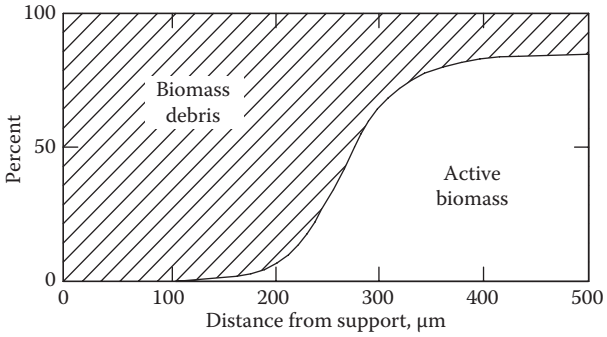

FIGURE 16.5 Simulation results showing the relative distribution of active biomass and biomass debris in a single species biofilm. (After Wanner, O., and Gujer, W., Competition in biofilms. *Water Science and Technology*, 17 (2/3): 27–44, 1984.)

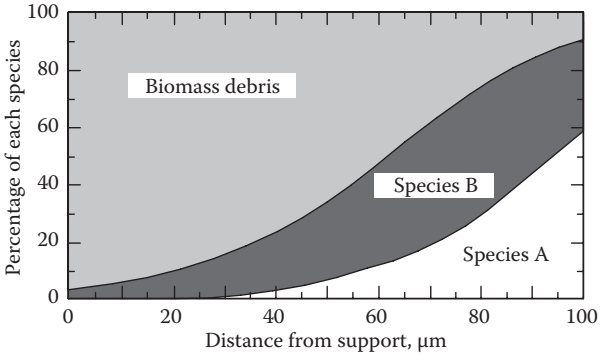

FIGURE 16.6 Simulation results showing the relative distributions of rapidly (Species A) and slowly (Species B) growing bacteria in a dual species biofilm when their only shared resource is space. (After Rittmann, B. E., and Manem, J. A., Development and experimental evaluation of a steady-state, multispecies biofilm model. *Biotechnology and Bioengineering*, 39:914–22, 1992.)

16.2 EFFECTS OF TRANSPORT LIMITATIONS

All current mathematical models for biofilms assume that the electron donor, the electron acceptor, and all nutrients are transported to the biomass within the biofilm by diffusional processes alone. In addition, consideration must also be given to the transport of those constituents from the bulk fluid to the biofilm. In this section we will examine these processes and the techniques used to model them. This examination will be limited to transport of a single electron donor (i.e., the substrate) to one type of biomass. It will be assumed that the electron acceptor and all other nutrients are provided in sufficiently high concentration in the bulk liquid so as not to be limiting within the biofilm.

16.2.1 Mass Transfer to and within a Biofilm

Consider a flat plate covered with a base biofilm. If this plate is placed into a substrate solution, the concentration of the substrate at the surface of the biofilm will be less than the concentration in the bulk of the fluid because of the substrate consumption by the microorganisms within the biofilm. Furthermore, because of that consumption, the substrate concentration will continue to drop with depth in the biofilm. In order for the consumption to continue, substrate must be transported from the bulk fluid to the liquid-biofilm interface by molecular and turbulent diffusion. It must also be transported within the biofilm. As discussed above, although both diffusion and advection are involved in internal transport, the phenomenon is modeled as if it were due to diffusion alone. Nevertheless, the net effect of these events is to cause a substrate concentration profile that looks something like the one in Figure 16.4. In this situation, the observed substrate consumption rate depends on the rate of mass transport external to and within the biofilm as well as on the true, intrinsic substrate consumption rate of the biomass (i.e., the true reaction rate without any mass transfer limitations). Consequently, if one were to observe the substrate consumption rate of a biofilm as a function of the substrate concentration in the bulk liquid, it would differ from the intrinsic relationship between substrate consumption rate and substrate concentration that could be measured when the microorganisms were dispersed throughout the liquid phase (thereby eliminating mass transfer effects). Thus, the effects of mass transfer obscure the true reaction rate relationship in a biofilm and any attempt to model the situation without incorporating the effects of mass transfer would be futile.

External mass transfer is typically modeled by idealizing the substrate concentration profile in the bulk liquid as shown in Figure 16.7. The variation in substrate concentration is restricted to a hypothetical stagnant liquid film of thickness L_w through which substrate must be transported to

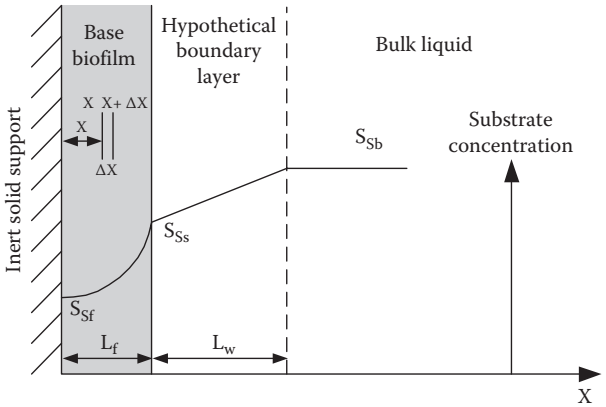

FIGURE 16.7 Traditional conceptualization of a base biofilm showing an idealized concentration profile for a single limiting nutrient.

reach the biofilm. As a consequence, the substrate concentration throughout the remaining fluid (i.e., the bulk liquid phase) is constant. All resistance to mass transfer from the bulk fluid to the biofilm is assumed to occur in the stagnant liquid film.

Two approaches are commonly used to model external mass transfer. One assumes that transport across the liquid layer is by molecular diffusion, with diffusivity D_w. In that case, the thickness L_w is defined as the equivalent depth of liquid through which the actual mass transfer can be described by molecular diffusion alone. Consequently, the flux, J_S, or mass of substrate transported per unit area per unit time, is given by

$$J_S = \frac{D_w}{L_w}(S_{Sb} - S_{Ss}), \tag{16.1}$$

where S_{Sb} is the substrate concentration in the bulk liquid and S_{Ss} is its concentration at the biofilm surface, as illustrated in Figure 16.7. Because the diffusivity is an intrinsic characteristic of the material being transported (the fluid is assumed to be water), L_w becomes the parameter that must be evaluated before Equation 16.1 can be used to depict the rate of transport of the substrate to the biofilm. Its value must be deduced from Equation 16.1 using measured fluxes coupled with known diffusivities and concentration gradients. The second approach employs a liquid phase mass transfer coefficient, k_L, which incorporates all of the effects of diffusive and convective mass transfer into one parameter. In that approach

$$J_S = k_L (S_{Sb} - S_{Ss}). \tag{16.2}$$

The value of k_L must also be deduced from measured fluxes and concentration gradients. It is apparent from comparison of Equations 16.1 and 16.2 that

$$k_L = \frac{D_w}{L_w}. \tag{16.3}$$

Thus, measured values of k_L may be used to estimate L_w and vice versa. The value of k_L (and L_w) will depend on the properties of the fluid (such as its viscosity, μ_w, and its density, ρ_w), the diffusivity of the substrate in the fluid, and the nature of the turbulence, which can be represented in part by the bulk fluid velocity past the biofilm, v. Figure 16.8[53] illustrates how that velocity influences the gradient in the bulk fluid. Here, the material being transported is oxygen, which is being used by the biofilm as it consumes the substrate. The numbered arrows in the figure show how the fluid velocity affects the thickness of the actual boundary layer, and thus illustrates the impact on L_w or k_L. Many relationships are available for relating k_L to the system characteristics. They are usually defined in terms of the Reynolds number ($v\rho_w d/\mu_w$) and the Schmidt number ($\mu_w/\rho_w D_w$) and are discussed in texts covering mass transfer (for example, see Weber and DiGiano[48]), as well as elsewhere.[9] Examination of common relationships reveals that many predict that k_L will increase with the square root of the fluid velocity. However, because flow situations in attached growth reactors are complex, it is usually necessary to determine experimentally how k_L depends on fluid velocity, or some factor affecting it, like mixing intensity in an agitated vessel or speed of rotation of a disc in a quiescent fluid.

Mass transfer within the biofilm is normally characterized by Fick's first law, which for free diffusion in an aqueous solution is

$$J_S = D_w \frac{dS_S}{dx}, \tag{16.4}$$

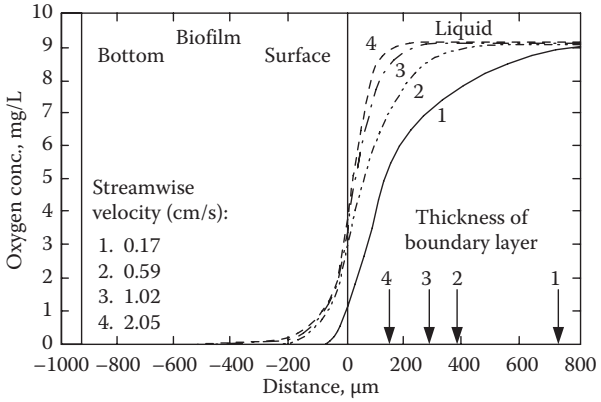

FIGURE 16.8 Effect of the fluid velocity past a biofilm on the thickness of the boundary layer for external mass transfer. (From Zhang, T. C. and Bishop, P. L., Experimental determination of the dissolved oxygen boundary layer and mass transfer resistance near the fluid-biofilm interface. *Water Science and Technology*, 30 (11): 47–58, 1994. Copyright © IWA Publishing. Reprinted with permission.)

in which D_w is a diffusivity and dS_S/dx is a concentration gradient. However, it is obvious from the previous discussion that a biofilm is more complex than the situation of free diffusion for which Fick's first law was developed. Therefore, the approach generally taken by modelers is to retain Fick's first law as the governing equation, but to replace the diffusivity with an effective diffusivity, D_e:

$$J_S = D_e \frac{dS_S}{dx}. \tag{16.5}$$

The effective diffusivity is usually smaller than the free diffusivity due to the presence of the EPS surrounding the cells in the biofilm. However, some researchers have measured effective diffusivities that are greater than the corresponding free diffusivities,[25] which are consistent with the presence of advection within the biofilm as discussed in Section 16.1. Thus, while Equation 16.5 continues to be used to describe transport within the biofilm, D_e should be thought of as being due to more than diffusion alone.

Having determined how transport to and within the biofilm can be modeled, the next step is to combine transport with the reactions occurring within the biofilm to establish the relationship between the bulk substrate concentration and the rate of substrate removal by the biofilm. The resulting relationship can then be combined with the appropriate process model to simulate the performance of an attached growth bioreactor. Three techniques are in common use for combining the biofilm model with a process model: the direct technique, the effectiveness factor technique, and the pseudoanalytical technique.[19] In the direct technique, the differential equations describing reaction within the biofilm are combined directly with the differential equations describing the bioreactor, giving a set of differential equations that must be solved by numerical methods. This technique is commonly employed for modeling systems involving multiple species carrying out multiple reactions, such as carbon oxidation, nitrification, and denitrification. The effectiveness factor technique pretends that the reaction rate at any point in a bioreactor can be defined by the intrinsic reaction rate expressed in terms of the bulk substrate concentration, multiplied by a factor (the effectiveness factor) that corrects for the effects of mass transport. Relationships between the effectiveness factor and the system characteristics are then coupled with the differential equations of the process model to allow simulation of bioreactor performance. The pseudoanalytical technique is similar to the effectiveness factor technique in concept, in that it develops a relationship between reaction rate and bulk substrate concentration that can then be used in the bioreactor model. In this case, however, the

differential equations representing transport and reaction within the biofilm are solved numerically and the output is used to develop simplified general relationships for transport and reaction that can be solved analytically, thereby allowing them to be coupled with the process equations. In the following two sections we will investigate the effectiveness factor and pseudoanalytical techniques.

16.2.2 Modeling Transport and Reaction: Effectiveness Factor Approach[20]

16.2.2.1 Effectiveness Factor

The basic concept in the modeling of biofilms is that the flux of substrate to and through the liquid-biofilm interface must equal the overall utilization rate per unit of biofilm planar area. Because the local substrate utilization rate depends on the substrate concentration at that location, it is clear from Figure 16.7 that the utilization rates at various points in the biofilm will be different. The overall utilization rate by the biofilm must consider this by integrating the reaction rate over the biofilm depth. Because of this averaging and because of the requirement for substrate transport from the bulk fluid to the biofilm surface, any observed relationship between the overall substrate removal rate and the bulk substrate concentration will be different from the intrinsic reaction rate expression for substrate removal. It is often convenient, however, to express the substrate removal rate as a function of the bulk substrate concentration, S_{Sb}, using a correction factor that takes into account the effects of transport. The correction factor is called the effectiveness factor[1] and is given the symbol η_e. If the Monod equation (Equation 3.36) expresses the intrinsic relationship between the specific substrate removal rate, q_H, and the substrate concentration, this can be expressed as

$$J_S = \eta_e \cdot X_{B,Hf} \cdot L_f \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}} \right), \tag{16.6}$$

in which $X_{B,Hf}$ is the mass of biomass per unit volume of biofilm and L_f is the biofilm thickness. Note that the substrate concentration is expressed in terms of the bulk substrate concentration, S_{Sb}. The units for $X_{B,Hf}$ and $\hat{q}_H$ must be consistent. Here, both are shown for biomass expressed in chemical oxygen demand (COD) units, and this convention will be followed throughout this chapter. This equation, as well as those that follow, could also be written with biomass expressed in total suspended solids (TSS) units, provided that appropriate conversion factors are used in COD mass balance equations. To use Equation 16.6 in mass balance equations for various types of biofilm reactors, information must be available about the effectiveness factor, η_e.

The nature of η_e can be determined by writing the mass balance equation for substrate for a differential element within the biofilm (Figure 16.7) and solving it to obtain the actual flux of substrate into the biofilm. The actual flux can then be used with Equation 16.6 to deduce the effects of the system kinetic and transport parameters on η_e. If the transport parameters include transport both to and within the biofilm, the effectiveness factor is called an overall effectiveness factor, denoted as η_{eO}. Assume that the biofilm in Figure 16.7 has reached a steady state in which it has a constant thickness L_f, a constant biomass concentration $X_{B,Hf}$, and uses substrate at a constant rate when exposed to a bulk substrate concentration S_{Sb}. A mass balance on substrate around a differential element in the biofilm yields

$$-D_e A_s \left. \frac{dS_S}{dx} \right|_x + D_e A_s \left. \frac{dS_S}{dx} \right|_{x+\Delta x} - X_{B,Hf} \cdot A_s \cdot \Delta x \left(\frac{\hat{q}_H \cdot S_S}{K_S + S_S} \right) = 0, \tag{16.7}$$

where A_s is the planar surface area normal to the direction of diffusion and x is the distance into the biofilm from the inert solid support. If D_e is constant, dividing both sides by A_s and Δx, and taking the limit as Δx approaches zero yields

$$D_e \frac{d^2S_S}{dx^2} - X_{B,Hf}\left(\frac{\hat{q}_H \cdot S_S}{K_S + S_S}\right) = 0, \tag{16.8}$$

which must be solved with two boundary conditions, one at the biofilm-support interface (x = 0) and the other at the liquid-biofilm interface ($x = L_f$). At the biofilm-support interface there is no transfer of substrate because the solid support is inert and impermeable. Thus, the appropriate boundary condition is

$$\frac{dS_S}{dx} = 0 \text{ at } x = 0. \tag{16.9}$$

As mentioned previously, having a permeable support is possible, but that case is not considered herein. However, it is apparent that a different boundary condition would be required. The boundary condition at the liquid-biofilm interface is more complicated. It is written by recognizing that the substrate flux at that interface must equal the substrate flux through the stagnant liquid layer. Consequently, the appropriate boundary condition is

$$J_S = D_e \frac{dS_S}{dx} = k_L\left(S_{Sb} - S_{Ss}\right) \text{ at } x = L_f. \tag{16.10}$$

The development of an equation for the overall effectiveness factor, η_{eO}, requires the solution of Equation 16.8 with Equations 16.9 and 16.10 as boundary conditions, giving substrate flux, J_S, and hence the substrate removal rate per unit of biofilm planar area, as a function of the bulk substrate concentration, S_{Sb}. The resulting relationship can then be used in Equation 16.6 to obtain η_{eO}.

In order to develop a generalized relationship between the overall effectiveness factor and the physical and biochemical characteristics of a biofilm system, Fink et al.[16] solved Equation 16.8 with its associated boundary conditions. They did this by a transformation of the two-point boundary value problem into an initial value problem. In doing so they used the following dimensionless quantities:

$$Bi = \frac{k_L \cdot L_f}{D_e}, \tag{16.11}$$

$$\phi = \left(\frac{X_{B,Hf} \cdot \hat{q}_H \cdot L_f^2}{K_S \cdot D_e}\right)^{0.5}, \tag{16.12}$$

$$\kappa = \frac{S_{Sb}}{K_S}, \tag{16.13}$$

and

$$\phi_f = \phi\left(\frac{1}{1+\kappa}\right)^{0.5}. \tag{16.14}$$

Bi is a Sherwood number, called the Biot number.[6] Recalling that a diffusivity divided by a length is equivalent to a mass transfer coefficient, it can be seen that the term D_e/L_f represents an internal mass transfer coefficient. Thus, the Biot number is the ratio of the external mass transfer rate to the

internal mass transfer rate. This means that when Bi is large, the external mass transfer coefficient is large relative to the internal coefficient so that all resistance to mass transfer can be considered to reside within the biofilm. In other words, the external resistance to mass transfer is negligible. This situation can arise when the flow rate past the biofilm is high. Conversely, when Bi is small, all of the resistance to mass transfer can be considered to be external to the biofilm. This situation can arise when the biofilm is very thin.

The parameter ϕ is a Thiele modulus. The physical significance of the Thiele modulus may be seen by squaring it, multiplying both the numerator and the denominator by $(S_{Sb} \cdot A_s)$, and rearranging:

$$\phi^2 = \left(\frac{X_{B,Hf} \cdot \hat{q}_H \cdot L_f^2}{K_S \cdot D_e} \right) \left(\frac{S_{Sb} \cdot A_s}{S_{Sb} \cdot A_s} \right) = \frac{(A_s \cdot L_f \cdot X_{B,Hf})(\hat{q}_H / K_S) S_{Sb}}{A_s (D_e / L_f) S_{Sb}}. \tag{16.15}$$

The term $(\hat{q}_H / K_S) S_{Sb}$ in the numerator is the first-order approximation of the Monod equation (Equation 3.38) for the specific substrate removal rate. For first-order kinetics, the maximum possible removal rate will occur when the substrate concentration surrounding the bacteria is the bulk substrate concentration. Consequently, the numerator represents the maximum possible first-order reaction rate. Likewise, the maximum possible diffusion rate will occur when the gradient is maximized, so that the denominator represents a maximum diffusion rate within the biofilm. Therefore, the Thiele modulus is the ratio of the maximum first-order reaction rate to the maximum diffusion rate. A large value of the Thiele modulus represents a situation in which the reaction rate is large relative to the diffusion rate. Such a situation is said to be diffusion limited. Conversely, a small value of ϕ represents a situation in which the diffusion rate is larger than the reaction rate. Such a situation is said to be reaction limited.

The parameter ϕ_f is a modified Thiele modulus. The purpose of the parameter κ is to take into consideration the deviation of the Monod equation from first-order kinetics, which was the basis for the Thiele modulus. It will be recalled from Section 3.2.7 that when the substrate concentration is small relative to the half-saturation coefficient, the Monod equation simplifies to an expression that is first-order with respect to substrate concentration, which is consistent with the basis of the Thiele modulus. Thus, when κ is small, the substrate removal rate behaves in a first-order manner so that ϕ_f equals ϕ, and the Thiele modulus adequately describes the relative importance of reaction versus diffusion. On the other hand, when κ is large, the substrate concentration is large relative to the half-saturation coefficient and the Monod equation does not behave in a first-order manner. In that situation, the deviation from first-order kinetics is large and ϕ_f is smaller than ϕ.

The results of Fink et al.[16] giving the overall effectiveness factor as a function of these dimensionless groups are shown in Figure 16.9. These values of η_{eO} may be used to calculate the overall flux of substrate into a biofilm of thickness L_f containing microorganisms at concentration $X_{B,Hf}$ under conditions where both internal and external mass transfer resistances exist. Thus:

$$J_S = \eta_{eO} \cdot X_{B,Hf} \cdot L_f \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}} \right). \tag{16.16}$$

The two curves for Bi = ∞ represent the case when the rate of external mass transfer is much higher than the rate of internal mass transfer, whereas the two curves for Bi = 0.01 represent the case where internal mass transfer is much more rapid than external mass transfer, due to a large external mass transfer resistance. Comparison of two groups of curves with different Bi values but the same ϕ_f value demonstrates that the existence of external mass transfer resistance has a strong effect on the overall effectiveness factor. For example, when $\phi_f = 1.0$, a 10-fold decrease in Bi (from 1.0 to 0.1) results in almost a 10-fold reduction in the overall effectiveness factor. Moreover, a comparison of curves with the same Biot number but different values of the Thiele modulus shows the relative importance of

reaction versus diffusion. When Bi = 0.01, the external mass transfer coefficient is much smaller than the internal mass transfer coefficient. Consequently, external mass transfer controls and the effectiveness factor is influenced little by the relative importance of reaction versus diffusion. Thus, the value of the Thiele modulus, ϕ, has little effect. Under these circumstances, the effectiveness factor is dominated by external mass transfer resistance and is often called an external effectiveness factor and given the symbol η_{eE}. On the other hand, when Bi = ∞, there is no external resistance to mass transfer and thus the relative importance of reaction versus diffusion has a strong impact on η_{eO}, as evidenced by the strong impact of the Thiele modulus. Under circumstances when Bi = ∞, the effectiveness factor is often called an internal effectiveness factor and given the symbol η_{eI}.

Although graphical representations like Figure 16.9 are convenient for some applications, for most occasions being able to determine the effectiveness factor analytically would be better. Consequently, it is common for investigators to develop functional relationships for the limited range of conditions they are interested in and this has been done for external, internal, and overall effectiveness factors. For example, for the case in which external mass transfer resistance is negligible (i.e., Bi is very large), Atkinson and Davies[2] developed both complex and simplified functional relationships for the internal effectiveness factor that agree quite well with the numerical results. In the interest of brevity, their equations will not be presented here. Rather, the reader is referred to other sources.[1,2]

16.2.2.2 Application of Effectiveness Factor

Equation 16.16 can be used with Figure 16.9 to determine the performance of a bioreactor containing a biofilm of known depth and biomass density. To illustrate how this is done, we will consider a continuous stirred tank reactor (CSTR) containing a biofilm.

Assume that steady-state conditions prevail over a reasonable time in a CSTR containing a solid surface covered by a biofilm of thickness L_f containing biomass at concentration $X_{B,Hf}$. To maintain a constant biofilm thickness, the cells generated by substrate consumption must be detached from the surface, dispersed throughout the liquid phase, and washed out in the bioreactor effluent. Because cells are in the bulk of the liquid as well as the biofilm, they are consuming substrate from both locations. Thus, the steady-state mass balance equation for substrate is

$$F \cdot S_{SO} - F \cdot S_{Sb} - J_S \cdot A_s - q_H \cdot X_{B,Hb} \cdot V = 0, \tag{16.17}$$

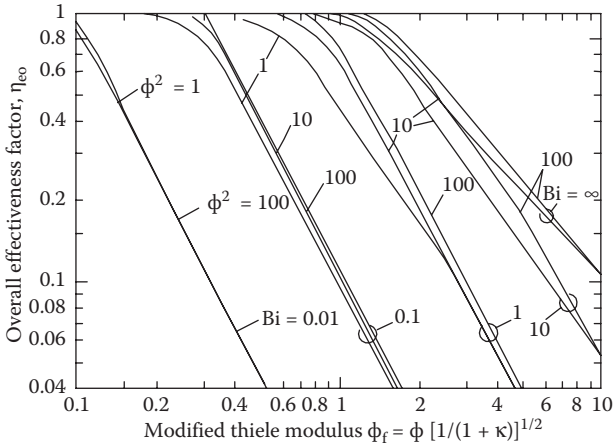

FIGURE 16.9 Overall effectiveness factor for Monod kinetics within a flat biofilm with external mass transfer resistance. (Adapted from Fink, D. J., Na, T.-Y., and Schultz, J. S., Effectiveness factor calculations for immobilized enzyme catalysts. *Biotechnology and Bioengineering*, 15:879–88, 1973.)

where J_S is the substrate consumption rate per unit surface area of biofilm, which is equivalent to the substrate flux, A_s is the biofilm surface area in the reactor, $q_H \cdot X_{B,Hb}$ is the substrate consumption rate per unit volume of reactor by dispersed bacteria, V is the bioreactor volume, F is the flow rate of influent and effluent, S_{SO} is the influent substrate concentration, and S_{Sb} is the effluent or bulk liquid substrate concentration. Substitution of Equation 16.16 for J_S and the Monod equation (Equation 3.36) for q_H yields

$$\frac{1}{\tau}(S_{SO} - S_{Sb}) - \eta_{eO} \cdot X_{B,Hf} \cdot L_f \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}} \right) \frac{A_s}{V} - X_{B,Hb} \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}} \right) = 0, \tag{16.18}$$

where τ is the hydraulic residence time (HRT) and $X_{B,Hb}$ is given by

$$X_{B,Hb} = Y_{Hobs}(S_{SO} - S_{Sb}). \tag{16.19}$$

Equation 16.19 assumes that the influent contains no biomass. It is an approximation of the biomass concentration in a CSTR without biomass recycle in which Y_{Hobs} is an observed yield that accounts for decay in both the biofilm and the dispersed bacteria. The rationale for its use is that biomass can only arise from the utilization of substrate and that the biofilm is at steady state. Therefore, Equation 16.18 may be rewritten as

$$\frac{1}{\tau} = \eta_{eO} \cdot X_{B,Hf} \cdot L_f \left(\frac{A_s}{V} \right) \left[\frac{\hat{q}_H \cdot S_{Sb}}{(K_S + S_{Sb})(S_{SO} - S_{Sb})} \right] + \frac{Y_{Hobs} \cdot \hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}}. \tag{16.20}$$

The solution of Equation 16.20 to determine the value of S_{Sb} associated with a given HRT requires an iterative approach since the value of η_{eO} depends on S_{Sb}. Thus, a value must be assumed for S_{Sb} and the value of η_{eO} determined from Figure 16.9 or an associated approximate equation. Equation 16.20 can then be solved for S_{Sb} and the solution compared with the assumed value. The procedure is repeated until the assumed and calculated values of S_{Sb} agree. An iterative solution is not required when the bulk substrate concentration is fixed at a desired value and the bioreactor HRT or biofilm surface area per unit volume (A_s/V) required to achieve that value is being calculated. Under that circumstance, the effectiveness factor may be determined directly from Figure 16.9 for use in Equation 16.20.

To show the effect of external mass transfer resistance on the performance of a CSTR containing a biofilm, Figure 16.10 was prepared by using Equation 16.20 with Figure 16.9 and the parameter values in Table 16.1.[20] As shown in Table 16.1, the value of D_e used to generate Figure 16.10 was extremely large ($\approx D_w \times 10^4$) to remove all internal mass transfer resistance. Three curves are presented, one with a very large k_L value to represent the absence of external mass transfer resistance, and two with k_L values that might be encountered in practice. Each of these curves represents effects that might be caused by changes in the velocity of the fluid past the biofilm. An examination of Figure 16.10 shows that the effect of a decrease in k_L is to reduce the activity of the microbial film, thereby making the effluent substrate concentration greater than it would be in a bioreactor with less mass transfer resistance.

Figure 16.11 was prepared to show the effects of internal mass transfer resistance, again by using Equation 16.20 with Figure 16.9 and the parameter values in Table 16.1. In this case, however, the value of k_L was made very large to remove all external mass transfer resistance. As in Figure 16.10, three curves are presented. One curve has a very large D_e value to represent the absence of internal mass transfer resistance. It is essentially the same as the curve with a k_L value of 20,000 cm/hr in Figure 16.10 and can be used for comparing the relative effects of the two types of resistance. In other words, that curve represents a situation in which the overall effectiveness factor is 1.0 for all

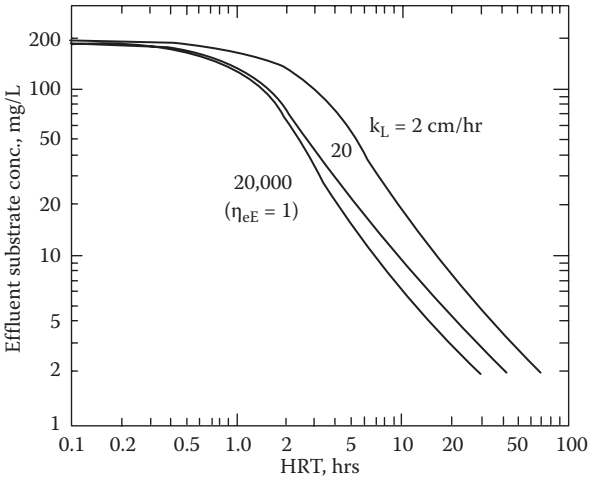

FIGURE 16.10 Effects of external mass transfer on the removal of soluble substrate by a CSTR containing a biofilm. (From Grady, C. P. L., Jr. and Lim, H. C., *Biological Wastewater Treatment: Theory and Applications*, Marcel Dekker, Inc., New York, 1980.)

TABLE 16.1
Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used to Generate Figures 16.10 and 16.11

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD·hr)	0.44
K_S	mg/L as COD	30
Y_{Hobs}	mg biomass COD/mg substrate COD	0.50
k_L	cm/hr	as noted in Figure 16.10 20,000 in Figure 16.11
D_e	cm^2/hr	2484 in Figure 16.10 as noted in Figure 16.11
D_w	cm^2/hr	0.2484
$X_{B,Hf}$	mg biomass COD/cm^3	32
L_f	cm	0.05
V	cm^3	1000
A_s	cm^2	100
S_{SO}	mg/L as COD	200

Source: Grady, C. P. L., Jr. and Lim, H. C., *Biological Wastewater Treatment: Theory and Applications*, Marcel Dekker, Inc., New York, 1980.

conditions. An examination of Figure 16.11 shows that the general effects of internal mass transfer resistance are similar to those of external mass transfer resistance; that is, a reduction in the amount of substrate that can be removed by the biofilm. One difference that should be noted, however, is that whereas the external mass transfer resistance is subject to change by engineering factors such as the velocity of flow past the biofilm, the internal mass transfer resistance is not. Instead, it depends on the physical and chemical properties of the wastewater and the microorganisms in the system.

Many factors can influence the values of the mass transfer coefficients in biofilm systems. Unfortunately, space does not permit a discussion of them here. Nevertheless, it is apparent from the above that an accurate estimation of the coefficients is a requirement for proper application of

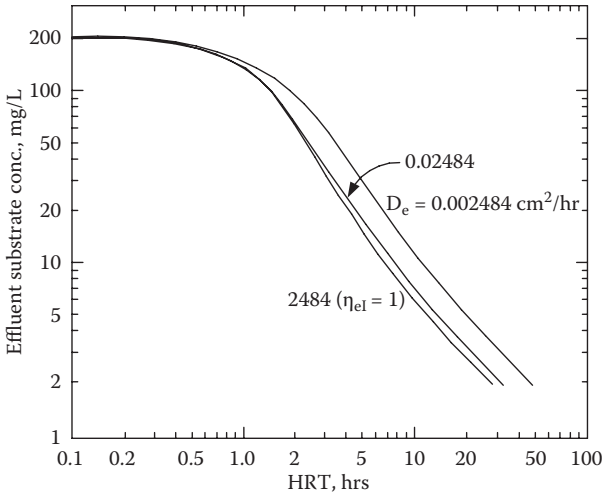

FIGURE 16.11 Effects of internal mass transfer on the removal of soluble substrate by a CSTR containing a biofilm. (From Grady, C. P. L. Jr. and Lim, H. C., *Biological Wastewater Treatment: Theory and Applications*, Marcel Dekker, Inc., New York, 1980.)

mathematical models. Consequently, readers should consult the work of others for more information on this important topic.[9,21,37]

Equation 16.16 can also be used in the mass balance on substrate in a plug-flow bioreactor. The mass balance must be written around an infinitesimal section and the limit taken to get the differential equation describing the change of substrate concentration along the bioreactor length. In this case, because the substrate concentration varies along the bioreactor length, so will the overall effectiveness factor. Thus, a solution of the equation requires a functional relationship for η_{eO} that can be used with numerical methods to solve the problem. Applications of this method are presented elsewhere[20] and the reader is referred there for more information. However, it is important to note that the change in the effectiveness factor can be appreciable from one end of a plug-flow bioreactor to the other. Thus, assuming a constant effectiveness factor throughout the bioreactor is inappropriate.

Many advantages are associated with the use of effectiveness factors for modeling attached growth systems and they have found reasonably wide use, particularly to model fluidized bed systems.[19] Unfortunately, the effectiveness factor approach requires knowledge of the steady-state biofilm thickness. Thus, the biofilm and bioreactor models must be coupled with some means of obtaining the steady-state biofilm thickness. In addition, the effectiveness factor approach gets much more complex when one needs to consider dual nutrient limitation and competition for space by multiple bacterial types. These situations are more easily handled by an alternative approach. We will examine one, the pseudoanalytical approach.

16.2.3 Modeling Transport and Reaction: Pseudoanalytical Approach

16.2.3.1 Pseudoanalytical Approach

The pseudoanalytical approach uses simple algebraic expressions for the flux of substrate into a biofilm. Those expressions are based on an analysis of the results from the numerical solution of the differential equations describing transport and reaction in a biofilm. The availability of simple algebraic equations eliminates the need to repetitiously solve numerically a set of nonlinear differential equations while modeling the performance of a biofilm reactor. Pseudoanalytical solutions have been developed by several authors,[32,34,36,40] but the approach of Sáez and Rittmann[34,36] is particularly accurate.

A key characteristic of the pseudoanalytical approach is that it allows calculation of the bulk substrate concentration and the biofilm thickness for a steady-state biofilm. A steady-state biofilm is one in which the gains in biofilm mass due to biomass growth are just balanced by the losses in biofilm mass due to the combined effects of microbial decay within the biofilm and detachment by shear at the liquid-biofilm interface.[32] Decay is treated in the traditional manner as presented in Chapter 5. Because of the balance between growth and loss, the biofilm attains a uniform thickness, L_f. For a purely heterotrophic biofilm, that thickness is given by[32]

$$L_f = \frac{J_S Y_H}{(b_H + b_D) X_{B,Hf}}, \tag{16.21}$$

where Y_H is the true growth yield for heterotrophs (COD units), b_H is the traditional decay coefficient for heterotrophs, b_D is the loss coefficient due to detachment caused by surface shear, and the other terms have their usual meaning. The value of the detachment coefficient varies with the shear stress on the biofilm, which depends on the hydrodynamic regime surrounding the biofilm.[30,42]

An important characteristic of a steady-state biofilm is the existence of a minimum bulk substrate concentration below which it cannot be maintained.[32] If the bulk substrate concentration is below that value, growth cannot occur rapidly enough to replace the losses to decay and detachment and the biofilm will decrease in thickness until it ceases to exist. The minimum bulk substrate concentration, S_{Sbmin}, is given by[32]

$$S_{Sbmin} = \frac{K_S (b_H + b_D)}{Y_H \cdot \hat{q}_H - (b_H + b_D)}. \tag{16.22}$$

An examination of Equation 16.22 reveals that it is analogous to Equation 5.23, the minimum attainable substrate concentration in a CSTR. This is because both represent the substrate concentration required to drive the growth reactions at a rate that will just balance loss by decay (and detachment, in the case of the biofilm). Since S_{Sbmin} is determined solely by parameters that depend on the biomass and substrate ($\hat{q}_H$, K_S, Y_H, and b_H) and the fluid regime (b_D), it takes on special significance as a parameter in the pseudoanalytical approach.

The equations upon which the pseudoanalytical solution of steady-state biofilm kinetics is based differ somewhat from those used to develop the effectiveness factor approach and are the result of the necessity to compute the biofilm thickness, L_f. They are[34]

$$D_e \frac{d^2 S_S}{dx^2} - X_{B,Hf}\left(\frac{\hat{q}_H \cdot S_S}{K_S + S_S}\right) = 0, \tag{16.23}$$

$$\frac{dS_S}{dx} = 0 \text{ at } x = 0, \tag{16.24}$$

$$S_S = S_{Ss} \text{ at } x = L_f, \tag{16.25}$$

$$\frac{dL_f}{dt} = \int_0^{L_f} \left[\frac{Y_H \cdot \hat{q}_H \cdot S_S}{K_S + S_S} - (b_H + b_D)\right] dx, \tag{16.26}$$

$$J_S = D_e \left.\frac{dS_S}{dx}\right|_{x=L_f}, \tag{16.27}$$

and

$$S_{Sb} = S_{Ss} + \frac{J_S}{k_L}, \tag{16.28}$$

where t is time. All other symbols are as defined previously. Equation 16.28 is just a rearranged form of Equation 16.2, the flux across the hypothetical boundary layer. After the introduction of several dimensionless variables, Sáez and Rittmann[34,36] solved Equations 16.23 through 16.28 using the numerical method of orthogonal collocation.[17] This was done for 500 initial conditions, covering the entire region of feasible solutions.[36] The output was then used to develop the pseudoanalytical solution.

The pseudoanalytical solution is based on the flux into a deep biofilm, which is defined as one in which the substrate concentration at the biofilm-support interface is zero.[23] The reason for using a deep biofilm as the reference case is because the dimensionless flux into a deep biofilm, $J^*_{S,deep}$ can be calculated analytically with[23]

$$J^*_{S,deep} = \left\{2\left[S^*_{Ss} - \ln\left(1 + S^*_{Ss}\right)\right]\right\}^{0.5}, \tag{16.29}$$

where S^*_{Ss} is the dimensionless substrate concentration at the liquid-biofilm interface:

$$S^*_{Ss} = \frac{S_{Ss}}{K_S}. \tag{16.30}$$

Thus, for any value of S_{Ss}, $J^*_{S,deep}$ can be calculated. Once that has been done, the dimensionless flux into an actual biofilm with that value of S_{Ss}, J^*_S, can be computed as some function of $J^*_{S,deep}$:

$$J^*_S = \xi J^*_{S,deep}, \tag{16.31}$$

provided an expression is available for ξ. The dimensionless flux into the actual biofilm, J^*_S, is defined as

$$J^*_S = \frac{J_S}{\left(K_S \cdot \hat{q}_H \cdot X_{B,Hf} \cdot D_e\right)^{0.5}}. \tag{16.32}$$

Consequently, once the dimensionless flux has been determined from Equation 16.31, the actual flux, J_S, associated with the liquid-biofilm interface substrate concentration, S_{Ss}, can be calculated with Equation 16.32 since all of the parameters in the relationship are known.

In developing the pseudoanalytical approach, Sáez and Rittmann[34] defined a new dimensionless group, S^*_{Sbmin}, the dimensionless minimum bulk substrate concentration. Its value is given by

$$S^*_{Sbmin} = \frac{S_{Sbmin}}{K_S} = \frac{b_H + b_D}{Y_H \cdot \hat{q}_H - \left(b_H + b_D\right)}. \tag{16.33}$$

The S^*_{Sbmin} is important because of its physical significance.[23] Recognition of the fact that $Y_H \cdot \hat{q}_H$ is equal to $\hat{\mu}_H$ (Equation 3.44), suggests that the value of S^*_{Sbmin} is an indication of the relative importance of biomass loss (by decay and detachment) and biomass growth. A large value of S^*_{Sbmin} (>1) implies that the maximum specific growth rate is not much larger than the specific loss rate of biomass from the biofilm, suggesting that the biofilm may be difficult to maintain. A small value (<1),

on the other hand, suggests a potentially high net growth rate relative to losses, thereby making the biofilm easy to maintain. It is important to recognize that the value of b_D, the specific detachment coefficient, depends on the flow velocity past the face of the biofilm and, thus, is under engineering control. Consequently, the term S^*_{Sbmin} represents both biological and physical factors. Because the significance of this term to the pseudoanalytical approach is similar to the significance of the dimensionless groups in the effectiveness factor approach, we think that it should be a named dimensionless group. Therefore, we propose and use the name Rittmann number, with the symbol Ri:

$$\mathrm{Ri} = \frac{b_H + b_D}{Y_H \cdot \hat{q}_H - (b_H + b_D)}. \tag{16.34}$$

Thus, the Rittmann number is the ratio of the specific loss rate of biomass from a biofilm to the net potential growth rate. As stated above, a large value means that a biofilm will be difficult to maintain, whereas a small value means that it will be easy to maintain.

Examination of the results from the 500 conditions studied by Sáez and Rittmann,[36] revealed that ξ could be adequately represented by[36]

$$\xi = \tanh\left[\alpha'\left(\frac{S^*_{Ss}}{\mathrm{Ri}} - 1\right)^{\beta'}\right], \tag{16.35}$$

where

$$\alpha' = 1.5557 - 0.4117 \tanh(\log_{10}\mathrm{Ri}) \tag{16.36}$$

and

$$\beta' = 0.5035 - 0.0257 \tanh(\log_{10}\mathrm{Ri}). \tag{16.37}$$

An examination of Equations 16.35 through 16.37 shows that they all depend upon the Rittmann number, showing why it is an important parameter in the pseudoanalytical solution. Sáez and Rittmann[36] examined the accuracy of the pseudoanalytical solution technique by using it to compute the flux into the biofilm for each of the 500 initial conditions and comparing the values with those obtained with the full numerical solution for each of the same conditions. The error depended somewhat on the value of the Rittmann number, but gave a standard error on the order of 2%, with 2.6% being the greatest observed. Thus, the pseudoanalytical solution is quite accurate.

To summarize, the calculation of the substrate flux associated with a given substrate concentration at the liquid-biofilm interface proceeds in the following manner. First, that concentration is put into dimensionless form with Equation 16.30, allowing calculation of the dimensionless flux into a deep biofilm, $J^*_{S,deep}$, with Equation 16.29. Then the Rittmann number is calculated with Equation 16.34, allowing the parameter ξ to be determined with Equations 16.35–16.37. Once ξ is known, the dimensionless flux into the biofilm can be calculated with Equation 16.31, allowing the actual flux to be determined with Equation 16.32.

16.2.3.2 Application of Pseudoanalytical Approach

The pseudoanalytical approach allows direct calculation of the flux into a steady-state biofilm associated with a given liquid-biofilm interface substrate concentration. However, what we really want to know is the flux associated with a given bulk substrate concentration, since it is the concentration that can be measured. The pseudoanalytical approach can be used to calculate it, as well as the

biofilm's thickness, by combining Equation 16.28 (in dimensionless form) with Equations 16.29, 16.31, and 16.35, giving

$$S_{Sb}^* = S_{Ss}^* + \frac{\tanh\left[\alpha'\left(\frac{S_{Ss}^*}{Ri} - 1\right)^{\beta'}\right]\{2[S_{Ss}^* - \ln(1 + S_{Ss}^*)]\}^{0.5}}{k_L^*}, \tag{16.38}$$

where the Rittmann number, Ri, the dimensionless bulk substrate concentration, S_{Sb}^*, and the dimensionless external mass transfer coefficient, k_L^*, are calculated with Equations 16.34, 16.39, and 16.40, respectively,

$$S_{Sb}^* = \frac{S_{Sb}}{K_S} \tag{16.39}$$

and

$$k_L^* = k_L\left(\frac{K_S}{\hat{q}_H \cdot X_{B,Hf} \cdot D_e}\right)^{0.5}. \tag{16.40}$$

After calculating α' and β' with Equations 16.36 and 16.37, respectively, Equation 16.38 can be solved iteratively for S_{Ss}^* using Newton's root-finding technique, which converges rapidly.[34] Sáez and Rittmann[34] recommend using $Ri + 10^{-6}$ as the initial guess for S_{Ss}^*. Once S_{Ss}^* is known, S_{Ss} can be calculated from its definition (Equation 16.30), allowing the flux into the biofilm, J_S, to be calculated from Equation 16.2. Finally, the biofilm thickness, L_f, can be calculated with Equation 16.21.

Example 16.2.3.1 (Adapted from Sáez and Rittmann[34])

A steady-state biofilm, described by the kinetic and stoichiometric coefficients given in Table E16.1, exists in an environment where the bulk substrate concentration is 0.5 mg/L. Determine the substrate flux into the biofilm (i.e., the substrate utilization rate per unit area of biofilm) and the biofilm thickness.

TABLE E16.1
Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used in Example 16.2.3.1

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD·hr)	0.2667
K_S	mg/L as COD	10
Y_H	mg biomass COD/mg substrate COD	0.625
b_H	hr^{-1}	0.0025
$X_{B,Hf}$	mg biomass COD/cm^3	50
k_L	cm/hr	3.333
D_e	cm^2/hr	0.02667
b_D	hr^{-1}	0.0017

Note: Data from Sáez, P. B. and Rittmann, B. E., Improved pseudo-analytical solution for steady-state biofilm kinetics. *Biotechnology and Bioengineering*, 32:379–85, 1988.

Because k_L, D_e, and $X_{B,Hf}$ all use cm in their units, S_{Sb} and K_S should be expressed as mg/cm³ for consistency. Thus, S_{Sb} has a value of 0.0005 mg/cm³ and K_S has a value of 0.01 mg/cm³. The values of S^*_{Sb}, k^*_L, and Ri are calculated with Equations 16.39, 16.40, and 16.34, respectively, yielding

$$S^*_{Sb} = \frac{0.0005}{0.01} = 0.05,$$

$$k^*_L = (3.33)\left[\frac{0.01}{(0.2667)(50)(0.0266)}\right]^{0.5} = 0.559,$$

and

$$Ri = \frac{0.0025 + 0.0017}{(0.625)(0.2667) - (0.0025 + 0.0017)} = 0.025641.$$

Equations 16.36 and 16.37 are used to calculate the parameters α' and β':

$$\alpha' = 1.5557 - 0.4117 \tanh(\log_{10} 0.025641) = 1.9346$$

and

$$\beta' = 0.5035 - 0.0257 \tanh(\log_{10} 0.025641) = 0.5272.$$

Equation 16.38 is then solved iteratively, giving a value of S^*_{Ss} of 0.027577. The dimensionless substrate concentration at the liquid-biofilm interface may be transformed into the physical domain by using the definition of the dimensionless variable, Equation 16.30:

$$S_{Ss} = (0.01)(0.027577) = 0.00027577 \text{ mg COD/cm}^3 = 0.2758 \text{ mg COD/L}.$$

The flux of substrate into the biofilm can then be calculated with Equation 16.2:

$$J_S = 3.333(0.0005 - 0.0002758) = 0.000747 \text{ mg COD/}(\text{cm}^2 \cdot \text{hr}).$$

Finally, the steady-state biofilm thickness can be calculated with Equation 16.21:

$$L_f = \frac{(0.000747)(0.625)}{(0.0025 + 0.0017)(50)} = 0.00222 \text{ cm} = 22.2 \text{ μm}.$$

Once the flux of substrate into a biofilm is known, it is a simple matter to calculate the biofilm area required to achieve the desired bulk substrate concentration in a CSTR, provided that all substrate removal is due to the biofilm alone. In other words, the contribution of suspended biomass to substrate removal is assumed to be negligible, which is a reasonable assumption given the HRTs normally associated with attached growth reactors. Under that condition, a steady-state mass balance on substrate gives:

$$A_s = \frac{F(S_{SO} - S_{Sb})}{J_S}. \tag{16.41}$$

Biofilm media is generally characterized by its specific surface area; that is, the surface area per unit of bioreactor volume, a_s:

$$a_s = \frac{A_s}{V}, \tag{16.42}$$

where V is the volume of the bioreactor actually containing the media. Thus, once the total required surface area is known, calculating the bioreactor volume required for a given media is easy.

Example 16.2.3.2

A synthetic wastewater with a biodegradable COD of 10 mg/L (0.010 mg/cm^3) is flowing at a rate of 1.0 L/hr (1000 cm^3/hr) into a CSTR containing a biofilm media with a specific surface area of 90 m^2/m^3 (0.90 cm^2/cm^3). The wastewater, the bioreactor, and the associated biofilm have the characteristics listed in Table E16.1. What total surface area of biofilm would be required to reduce the biodegradable COD to 0.5 mg/L (0.0005 mg/cm^3)? What bioreactor volume is required to house the media?

Because the bioreactor is a CSTR, the bulk substrate concentration is 0.5 mg COD/L throughout, and thus the substrate flux into the biofilm is the same as in Example 16.2.3.1. As a result, the total biofilm area can be calculated with Equation 16.41:

$$A_s = \frac{(1000)(0.010 - 0.0005)}{0.000747} = 12{,}700 \text{ cm}^2.$$

The bioreactor volume can then be calculated from a rearranged form of Equation 16.42:

$$V = \frac{12{,}700}{0.90} = 14{,}100 \text{ cm}^3 = 14.1 \text{ L}.$$

A simple CSTR without biomass recycle or biofilm would require a volume of 184 L to achieve the same effluent substrate concentration. Thus, the benefit of the biofilm is apparent.

As with other systems we have studied, biofilm systems also benefit from being housed in a plug-flow or tanks-in-series configuration. The pseudoanalytical approach can also be used to determine the performance of such systems by considering the plug-flow systems to be a series of completely mixed compartments.[23] Although the number of compartments required to adequately simulate a plug-flow system depends on its hydraulic characteristics, considering the bioreactor to contain six is usually adequate.[23] Regardless of the situation considered, a number of compartments in series should be assumed, as well as the volume of each compartment. Starting with the last compartment and the desired effluent substrate concentration, it is possible to calculate the flux into the biofilm using the procedure in Example 16.2.3.1. Once that flux is known, it can be used to determine the influent substrate concentration that could be treated with a given media by using the procedure in Example 16.2.3.2. Since the influent to the last compartment is the effluent from the next-to-last compartment, the procedure can be repeated to determine the concentration of organic matter in the influent to the next-to-last compartment, and so on. The procedure is repeated until the influent to the first compartment is calculated. If it is equal to the known influent concentration, the system is the correct size. If it is larger than the known concentration, the system is larger than required and the calculations should be repeated with a smaller volume for each compartment. (Alternatively, the

excess volume may be acceptable as a factor of safety.) If the calculated influent substrate concentration is less than the known value, the system is too small and the calculations must be repeated with larger compartment volumes.

While the procedure in the preceding paragraph is straightforward, it is tedious because of the need to solve Equation 16.38 repeatedly. Consequently, Heath et al.[23,24] have proposed an even simpler method based on normalized loading curves.

16.2.3.3 Normalized Loading Curves

A graphical representation forms the basis for the normalized loading curve approach to biofilm reactor analysis and design. Heath et al.[23,24] used the pseudoanalytical approach to solve for the substrate flux associated with various bulk substrate concentrations. This was done for many conditions and the results were presented in generalized form by normalizing the bulk substrate concentration relative to S_{Sbmin} and the flux relative to a reference flux, J_{SR}. The curves are called loading curves because the flux to a biofilm is approximately equal to the rate of substrate input per unit of biofilm (i.e., the loading). The reference flux is the minimum flux just required to maintain a steady-state biofilm that is deep. For computational purposes, it is defined as the flux resulting when $\xi = 0.99$ in Equation 16.31.[23] The reference flux depends on the value of the Rittmann number and can be presented in generalized form by a plot of J^*_{SR}/Ri versus Ri, where J_{SR} is made dimensionless in the same way as J_S, as indicated in Equation 16.32.[7] Figure 16.12 shows such a plot.[36] It can be entered with a known value of the Rittmann number, giving the corresponding value of J^*_{SR}, which can be put back into the physical domain (i.e., J_{SR}) by using Equation 16.32.

Normalized loading curves were plotted for fixed values of Ri and k^*_L.[24] The Rittmann number is used as a parameter because of its fundamental importance, as discussed previously. The dimensionless external mass transfer coefficient, k^*_L, was chosen as the second parameter because it represents the importance of external mass transfer resistance in the performance of the biofilm.

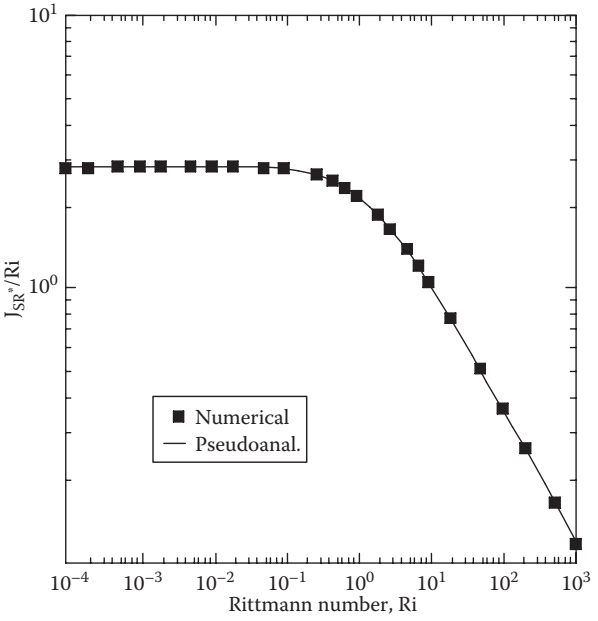

FIGURE 16.12 Curve for the determination of the dimensionless reference flux, J^*_{SR}/Ri, from the Rittmann number. (From Sáez, P. B. and Rittmann, B. E., Accurate pseudoanalytical solution for steady-state biofilms. *Biotechnology and Bioengineering*, 39:790–93, 1992. Copyright © John Wiley & Sons. Reprinted with permission.)

As can be seen in Equation 16.40, if k_L^* is large (>10), external mass transfer resistance is of little importance relative to reaction and internal mass transfer. Conversely, when k_L^* is small (<1) external mass transport is likely to play an important role in biofilm performance. Figures 16.13 through 16.17 present normalized loading curves for Ri values of 0.01, 0.1, 1, 10, and 100, respectively.[24] It should be noted that S_{Sb}/S_{Sbmin} is equivalent to S_{Sb}^*/Ri and J_S/J_{SR} is equivalent to J_S^*/J_{SR}^*.

The use of the normalized loading plots is very straightforward for a biofilm in a completely mixed bioreactor or bioreactor compartment. For a given situation, the value of Ri is calculated with Equation 16.34 and Figure 16.12 is used to determine the dimensionless reference flux, J_{SR}^*. The family of normalized loading curves corresponding most closely to the value of Ri is then used, with

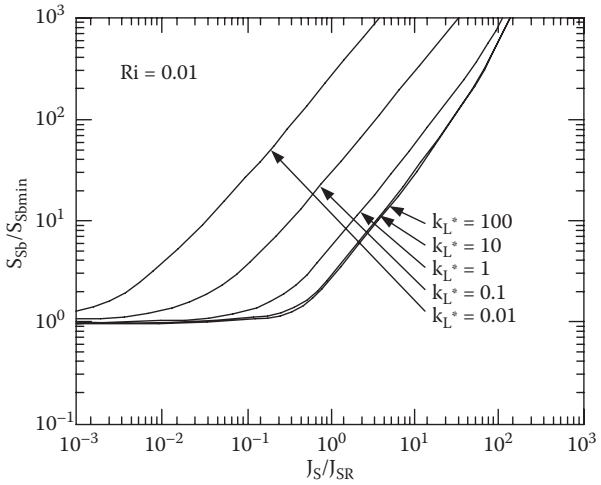

FIGURE 16.13 Normalized loading curves for Ri = 0.01. (From Heath, M. S., Wirtel, Rittmann, B. E., and Noguera, D. R., Closure to discussion of "Simplified design of biofilm processes using normalized loading curves." *Research Journal, Water Pollution Control Federation*, 63:91–92, 1991. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

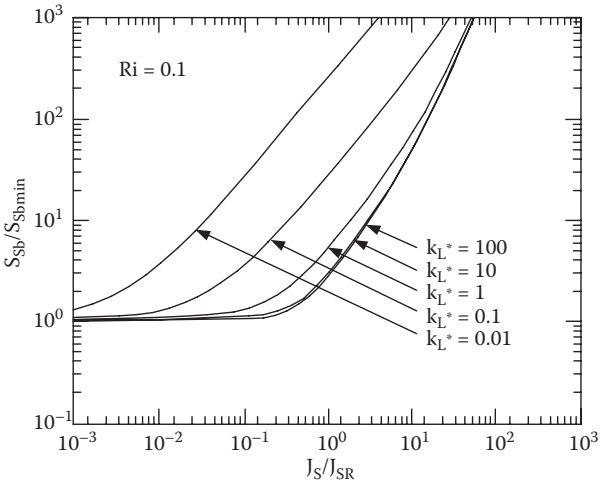

FIGURE 16.14 Normalized loading curves for Ri = 0.1. (From Heath, M. S., Wirtel, S. A., Rittmann, B. E., and Noguera, D. R., Closure to discussion of "Simplified design of biofilm processes using normalized loading curves." *Research Journal, Water Pollution Control Federation*, 63:91–92, 1991. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

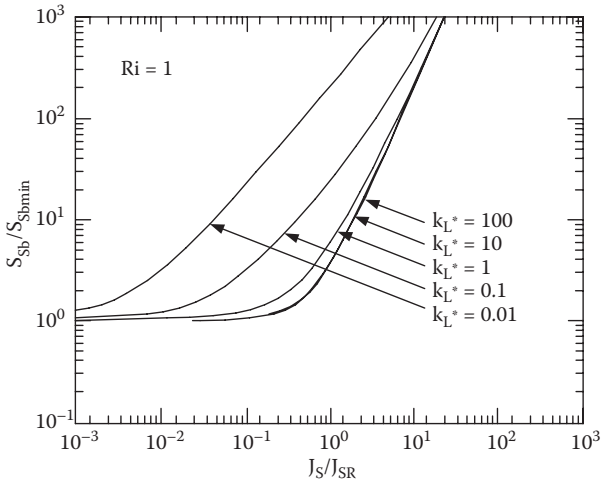

FIGURE 16.15 Normalized loading curves for Ri = 1. (From Heath, M. S., Wirtel, S. A., Rittmann, B. E., and Noguera, D. R., Closure to discussion of "Simplified design of biofilm processes using normalized loading curves." *Research Journal, Water Pollution Control Federation*, 63:91–92, 1991. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

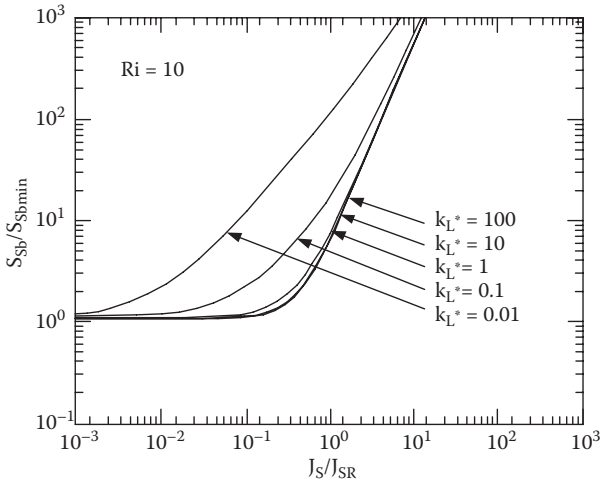

FIGURE 16.16 Normalized loading curves for Ri = 10. (From Heath, M. S., Wirtel, S. A., Rittmann, B. E., and Noguera, D. R., Closure to discussion of "Simplified design of biofilm processes using normalized loading curves." *Research Journal, Water Pollution Control Federation*, 63:91–92, 1991. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

the particular curve depending on the value of k_L^*. If no curve corresponds exactly to the Ri and k_L^* values for the system, then interpolation can be used.[23] Using the desired bulk substrate concentration, S_{Sb}, the value of S_{Sb}/S_{Sbmin} can be calculated and used to determine J_S/J_{SR} (or J_S^*/J_{SR}^*) from the appropriate curve. Since the value of J_{SR}^* is known, the value of J_S is fixed. Equation 16.41 can then be used to calculate the required biofilm area, A_s, and Equation 16.42 to calculate the associated bioreactor volume. This gives a direct solution for a single CSTR, or allows the iterative procedure discussed previously to be used for a plug-flow system approximated as compartments in series.

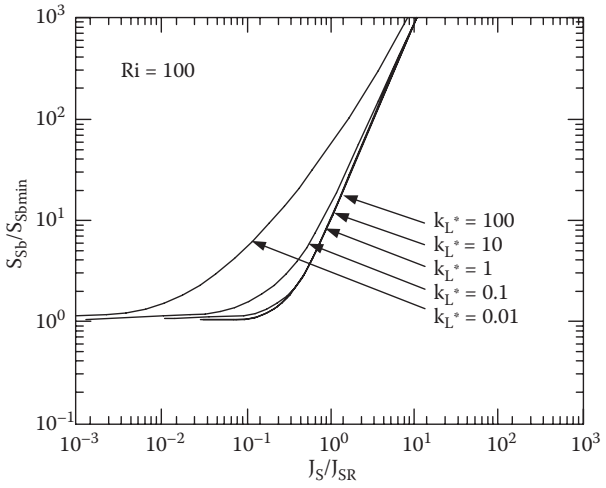

FIGURE 16.17 Normalized loading curves for Ri = 100. (From Heath, M. S., Wirtel, S. A., Rittmann, B. E., and Noguera, D. R., Closure to discussion of "Simplified design of biofilm processes using normalized loading curves." *Research Journal, Water Pollution Control Federation*, 63:91–92, 1991. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

Example 16.2.3.3

A synthetic wastewater with a biodegradable COD of 10 mg/L (0.010 mg/cm^3) is flowing at a rate of 1.0 L/hr (1000 cm^3/hr) into a CSTR containing a biofilm media with a specific surface area of 90 m^2/m^3 (0.90 cm^2/cm^3). The wastewater, the bioreactor, and the associated biofilm have the characteristics listed in Table E16.2. What total surface area of biofilm would be required to reduce the biodegradable COD to 1.5 mg/L (0.0015 mg/cm^3)? What bioreactor volume is required to house the media?

The first task is to determine the dimensionless reference flux, J_{SR}^* from Figure 16.12. This requires a calculation of Ri with Equation 16.34:

$$\text{Ri} = \frac{0.0075 + 0.0076}{(0.625)(0.2667) - (0.0075 + 0.0076)} = 0.10.$$

Entering Figure 16.12 with Ri = 0.10 gives a value of J_{SR}^*/Ri of 2.8, thereby fixing J_{SR}^* at 0.28.

The next task is to determine J_S from the appropriate normalized loading curve. This requires calculation of S_{Sb}/S_{Sbmin} and k_L^*. The S_{Sb} is given as 1.5 mg COD/L. The S_{Sbmin} can be calculated with Equation 16.22 or by multiplying Ri by K_S (see Equations 16.33 and 16.34), which is 10 mg COD/L giving $S_{Sbmin} = 1.0$ mg COD/L. Thus, S_{Sb}/S_{Sbmin} is 1.5. The value of k_L^* is calculated with Equation 16.40:

$$k_L^* = 6.00\left[\frac{0.01}{(10.2667)(50)(0.02667)}\right]^{0.5} = 1.0.$$

Thus, the appropriate curve to use to obtain J_S is the curve for $k_L^* = 1.0$ in Figure 16.14. Entering that curve with $S_{Sb}/S_{Sbmin} = 1.5$ gives J_S/J_{SR} (or J_S^*/J_{SR}^*) = 0.2. Since J_{SR}^* is 0.28, J_S^* is 0.056. The value of J_S can then be determined from Equation 16.32:

TABLE E16.2
Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used in Example 16.2.3.3

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD·hr)	0.2667
K_S	mg/L as COD	10
Y_H	mg biomass COD/mg substrate COD	0.625
b_H	hr^{-1}	0.0075
$X_{B,Hf}$	mg biomass COD/cm^3	50
k_L	cm/hr	6.00
D_e	cm^2/hr	0.02667
b_D	hr^{-1}	0.0076

$$J_S = 0.056\left[(0.010)(0.2667)(50)(0.02667)\right]^{0.5} = 0.00334 \text{ mg COD}/(\text{cm}^2 \cdot \text{hr}).$$

Once the flux is known, the required biofilm surface area can be calculated with Equation 16.41:

$$A_s = \frac{(1000)(0.010 - 0.0015)}{0.00334} = 2545 \text{ cm}^2.$$

The bioreactor volume can then be calculated from a rearranged form of Equation 16.42:

$$V = \frac{2545}{0.90} = 2840 \text{ cm}^3 = 2.83 \text{ L}.$$

A simple CSTR without biomass recycle or biofilm would require a volume of 70 L to achieve the same effluent substrate concentration. As in Example 16.2.3.2, the benefit of the biofilm is apparent.

16.2.3.4 Parameter Estimation

Before the pseudoanalytical approach can be used, values must be available for the parameters in the model. Since the Monod kinetic parameters represent the kinetics of the biomass in the absence of mass transfer limitations, theoretically, one could use kinetic parameters determined with suspended growth cultures. However, the growth of biomass in a biofilm alters its physiological state.[10] Consequently, the parameters should be determined for biomass growing as a biofilm. Two techniques are available, one based on steady-state experiments[55] and the other on transient-state experiments.[33] Space does not allow us to discuss those procedures, but the reader should be aware of their existence, should the need to measure biofilm kinetics arise.

16.2.4 Modeling Transport and Reaction: Limiting-Case Solutions

Although the pseudoanalytical approach to modeling transport and reaction has greatly simplified the computation of the flux into a steady-state biofilm, it still does not lead to closed form solutions to bioreactor mass balances. Many circumstances exist, however, in which direct calculation of the bulk substrate concentration in a biofilm reactor would be advantageous. Consequently, several

investigators have proposed simplifying assumptions for limiting cases that allow closed analytical solutions. There are four of them for steady-state biofilms.[35]

16.2.4.1 Deep Biofilm

As we saw in Section 16.2.3, it is possible to solve directly for the flux into a deep biofilm, which is one in which the substrate concentration at the solid-biofilm interface is zero. Under that condition, the dimensionless flux is given by Equation 16.29.[34,41] Its use results in a different flux for each liquid-biofilm interface substrate concentration.

16.2.4.2 Fully Penetrated Biofilm

A fully penetrated biofilm is one in which the change in the substrate concentration with depth within the biofilm is negligible. In other words, the entire biofilm contains substrate at almost the same concentration as the liquid-biofilm interface. By assuming that the substrate concentration at the biofilm-support interface, S_{Sf}, is equal to the substrate concentration at the liquid-biofilm interface, S_{Ss}, an analytical solution can be obtained for the dimensionless flux, $J^*_{S,fp}$:[41]

$$J^*_{S,fp} = \left(\frac{S^*_{Ss}}{1 + S^*_{Ss}}\right) L^*_f, \tag{16.43}$$

where L^*_f, the dimensionless biofilm thickness, is given by

$$L^*_f = L_f \left(\frac{\hat{q}_H \cdot X_{B,Hf}}{D_e \cdot K_S}\right)^{0.5}. \tag{16.44}$$

16.2.4.3 First-Order Biofilm

When the substrate concentration at all points within the biofilm is much less than the half-saturation coefficient, the Monod equation can be approximated as first-order with respect to substrate concentration (Equation 3.38). This allows the differential equation describing reaction within a steady-state biofilm to be rewritten as

$$D_e \frac{d^2 S_S}{dx^2} - \frac{X_{B,Hf} \cdot \hat{q}_H \cdot S_S}{K_S} = 0. \tag{16.45}$$

For the boundary conditions that

$$\frac{dS_S}{dx} = 0 \text{ at } x = 0 \tag{16.46}$$

and

$$S_S = S_{Ss} \text{ at } x = L_f, \tag{16.47}$$

this equation can be solved analytically, giving the following when placed into dimensionless form:[34]

$$J^*_{S,first} = S^*_{Ss} \tanh\left[\left(\frac{1 + Ri}{Ri}\right) J^*_{S,first}\right]. \tag{16.48}$$

16.2.4.4 Zero-Order Biofilm

When the substrate concentration at all points within the biofilm is much greater than the half-saturation coefficient, the Monod equation can be approximated as zero-order with respect to substrate concentration (Equation 3.37). This allows the differential equation describing reaction within a steady-state biofilm to be rewritten as

$$D_e \frac{d^2S_S}{dx^2} - X_{B,Hf} \cdot \hat{q}_H = 0. \tag{16.49}$$

Solution of this equation with the boundary conditions expressed by Equations 16.46 and 16.47 gives the following when placed into dimensionless form:[40]

$$J^*_{S,zero} = L^*_f \tag{16.50}$$

16.2.4.5 Other Cases

It is theoretically possible for first-order or zero-order biofilms to be either deep or fully penetrated. In such cases, both requirements must be satisfied. A biofilm that is neither first-order nor zero-order has been called a Monod biofilm.[35] A biofilm that is neither deep nor fully penetrated has been called a shallow biofilm.[35]

16.2.4.6 Error Analysis

The limiting-case solutions are fine provided the simplifying assumptions are appropriate for the conditions encountered. If they are not, their application can lead to gross errors.[35,41] Thus, one must be sure a limiting-case solution is appropriate before using it. To help in that assessment, Sáez and Rittmann[35] conducted an error analysis of the limiting-case solutions and prepared Figure 16.18 showing conditions under which a limiting-case solution of the steady-state biofilm model differs from the full pseudoanalytical solution by less than 1.0%. Several important points arise from the figure. First, fully penetrated biofilms are difficult to attain, occurring only when the Rittmann number is greater than 10; that is, only when the relative growth potential is low. This suggests that the fully penetrated case is of limited utility. First-order biofilms occur when Ri is small, but Monod biofilms cover a broad range of Ri values. In fact, the Monod-deep zone could be expanded into the first-order-deep zone, since first-order kinetics is just a limiting case of Monod kinetics and the deep-biofilm equation is straightforward. Perhaps the most important point, however, is that the full pseudoanalytical approach must be used over a broad range of conditions (i.e., the Monod-shallow zone), which covers much of the expected range of Ri values.[35]

In summary, while several limiting-case solutions are available in the literature, most are applicable only under very restricted conditions. The exception to this is the case of the deep biofilm. Thus, Equation 16.29 represents the most useful limiting-case solution. Nevertheless, for the practical range of Ri values, many problems will require use of the full pseudoanalytical solution.

16.3 EFFECTS OF MULTIPLE LIMITING NUTRIENTS

All of the models presented in Section 16.2 consider only a single limiting nutrient, the electron donor. However, in aerobic systems it is quite likely that the concentration of electron acceptor (i.e., oxygen) will drop to sufficiently low values within the biofilm for the microorganisms to become limited by both the electron acceptor and the electron donor. This follows from the fact that for most biofilm systems, the maximum concentration of oxygen in the bulk liquid phase is limited to the saturation concentration associated with air at atmospheric pressure, which is roughly 8 to 10 mg/L.

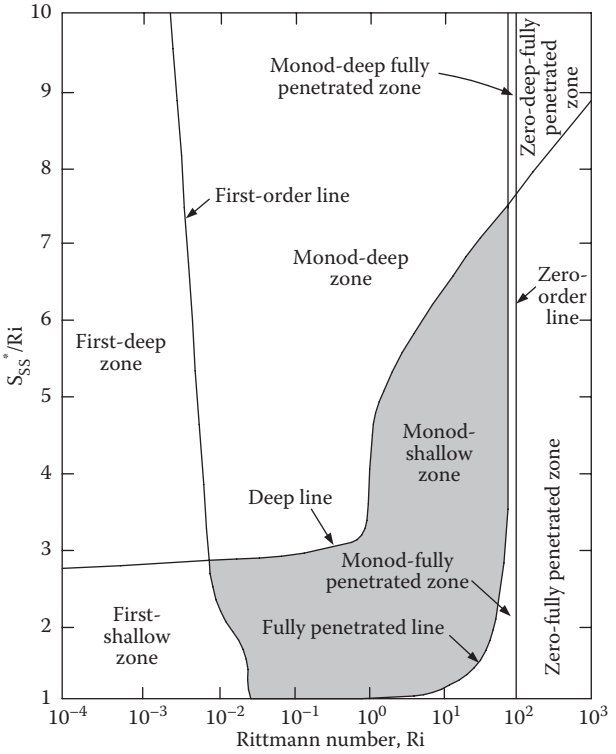

FIGURE 16.18 Conditions under which a limiting case solution to the steady-state biofilm model differs from the full pseudoanalytical solution by less than 1.0%. Each unshaded area represents a limiting case solution. The shaded area, labeled Monod shallow zone, indicates the conditions under which none of the limiting case solutions are accurate. (From Sáez, P. B. and Rittmann, B. E., Error analysis of limiting-case solutions to the steady-state-biofilm model. *Water Research*, 24:1181–85, 1990. Copyright © Elsevier Science Ltd. Reprinted with permission.)

In addition, in some attached growth reactors, the concentration of oxygen in the bulk liquid is likely to be much lower than saturation.

The importance of the dissolved oxygen (DO) concentration to the behavior of a biofilm is shown clearly in Figure 16.19.[54] The biofilm depicted was grown with ammonia-N as the sole electron donor, so nitrifying bacteria were the predominant types present. We saw earlier that they have a relatively high half-saturation coefficient for oxygen and thus their activity is influenced strongly by the DO concentration. After growth, the biofilm was placed in a test chamber and the profiles of ammonia-N, nitrate-N, and DO concentrations were measured with microelectrodes for different DO concentrations in the bulk liquid phase.[54] The ammonia-N profile is shown in Figure 16.19a and the DO profile is shown in Figure 16.19b. Examination of the figure reveals that when the DO concentration in the bulk liquid was high (15 mg/L), ammonia-N was exhausted before the oxygen. However, as the concentration of DO in the bulk liquid was decreased, oxygen became exhausted at shallower depths in the biofilm, thereby limiting the amount of ammonia-N converted to nitrate-N. This suggests that for accurate depiction of substrate removal in a biofilm, consideration must be given to the concentration of the electron acceptor as well as the electron donor.

Modeling of dual nutrient limitation requires the use of an interactive double Monod kinetic expression like Equation 3.46 in the rate equations for both the electron donor and the electron acceptor, with the two rates being linked stoichiometrically by Equation 3.34. More importantly, it requires transport of both constituents to be considered. This means that another mass balance equation like Equation 16.8, with its associated boundary conditions, must be written for the electron

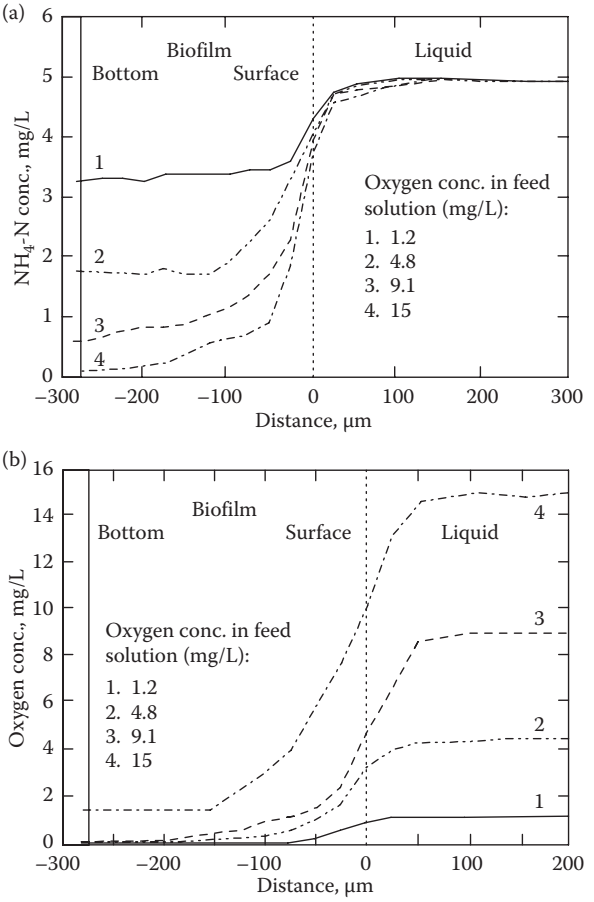

FIGURE 16.19 Concentration profiles for: (a) ammonia-N and (b) dissolved oxygen in nitrifying biofilms subjected to different concentrations of dissolved oxygen in the bulk liquid phase. (Adapted from Zhang, T. C., Fu, Y. C., and Bishop, P. L., Competition in biofilms. *Water Science and Technology*, 29 (10/11): 263–70, 1994. Copyright © IWA Publishing. Reprinted with permission.)

acceptor and solved along with the mass balance equation on substrate. The boundary conditions would be analogous to Equations 16.9 and 16.10, but would be written in terms of the electron acceptor. Furthermore, the diffusivity and mass transfer coefficient for that equation and its boundary conditions must be for the electron acceptor. Because of the increased number of parameters, this set of equations does not lend itself to the effectiveness factor or the pseudoanalytical approach. Rather, the full equations must be solved numerically for each situation.

Because modeling of systems under dual nutrient limitation is complex, having a simple way to decide when a biofilm can be considered to be limited by either the electron donor or the electron acceptor alone would be convenient because then the techniques of Section 16.2 could be applied. Unfortunately, this question has only been addressed with the noninteractive dual substrate model (Equation 3.47), which is not an accurate representation of the effects of two limiting nutrients.[3] Furthermore, the question is sufficiently complex so as to make the answer nonintuitive. Thus, at this time it is impossible to establish clear guidelines as to when a biofilm can be considered limited by only the electron donor or the electron acceptor, without performing numerical simulations with more complex models.[47] However, it can be stated that when the dimensionless concentration of one is in stoichiometric excess relative to the dimensionless concentration of the other in the bulk

liquid, then the one present in the least amount can be considered to be a single limiting constituent, allowing the approaches of Section 16.2 to be used. This suggests that only in dilute substrate solutions saturated with DO can the electron donor be considered to be the rate limiting substance in aerobic environments.

One important use of biofilm reactors is the treatment of contaminated groundwater and the removal of trace organic compounds from drinking water. Because both of those situations often involve dilute substrate solutions, a single substrate biofilm model may be accurate enough for them. Furthermore, because anoxic and anaerobic systems often have high electron acceptor concentrations, it is likely that higher electron donor concentrations can be considered rate limiting in them, thereby extending the applicability of single substrate biofilm models. Nevertheless, because of the benefits associated with the use of the effectiveness factor and pseudoanalytical techniques, it is apparent that the question of when dual substrate models must be applied needs more systematic study, thereby helping establish simple rules for determining whether growth is limited by the electron donor or by the electron acceptor.

16.4 MULTISPECIES BIOFILMS

All of the models we have considered so far in this chapter have considered only a single type of bacteria growing on a single electron donor. Furthermore, they have considered only one type of electron acceptor. However, as we saw when we studied suspended growth systems, if both organic carbon and ammonia-N are available, then both heterotrophs and autotrophs will grow if the environmental conditions are appropriate. In addition, if an electron donor is present but environmental conditions change from aerobic to anoxic, then facultative heterotrophs can change the nature of their electron acceptor. We have already seen that the oxygen concentration decreases with depth in a biofilm and can approach zero. Thus, if nitrate is present, the potential for denitrification exists in the interior of a biofilm. Furthermore, when biofilms are exposed to liquids containing high ammonia concentrations, the oxygen concentration can go to zero in the biofilm, making it possible for ammonia oxidation to occur with nitrite as the terminal electron acceptor by the obligate anaerobic anammox bacteria (Section 2.3.3). In other words, just as the potential existed in suspended growth systems for many of the events discussed in Chapter 2, so too does the potential exist in attached growth systems. Consequently, in order for a model to have general utility, it should consider all of these possibilities.

The modeling of multiple events within a biofilm is a good deal more complex than their modeling in suspended growth systems. There are many reasons for this. First is the necessity to consider both transport and reaction simultaneously. We have already seen how that is handled for a single electron donor and electron acceptor provided in the bulk fluid. Extension of those concepts to multiple donors and acceptors in the bulk fluid is not complicated; it just increases the number of differential equations that must be solved. However, it must be recognized that when an electron acceptor such as nitrite-N or nitrate-N is generated within the biofilm, transport can occur in either or both directions from the point of generation, depending on the concentration gradient established. This, too, must be considered in the equations, complicating them somewhat. Another event that complicates the modeling of multispecies biofilms is the competition between the various types of bacteria for the electron acceptor. Autotrophs require molecular oxygen as their electron acceptor and heterotrophs will use it in preference to nitrate-N and nitrite-N when it is present. However, we saw earlier that heterotrophs have a lower half-saturation coefficient for oxygen than autotrophs do. This means that heterotrophs can lower the oxygen concentration within the biofilm to the point that autotrophs cannot grow. This puts the autotrophs at a disadvantage and limits the region in which they can grow. Perhaps the most important complication arises, however, from the competition for space within the biofilm. In a suspended growth system the biomass is distributed uniformly and is lost from the system in proportion to its concentration. In other words, all of the biomass has the

same residence time. This is not true in an attached growth system. Rather, biomass grows outward from a solid support and is removed by detachment at the liquid-biofilm interface. This displacement toward the interface must be considered in any multispecies model. In addition, as we saw in Figure 16.6, the distribution of biomass is not uniform throughout the biofilm. This means that different types of bacteria have different residence times in the biofilm, and the model must be structured to consider this as well.

As one might deduce from the description above, the modeling of multispecies biofilms is among the most complicated activities in the modeling of biochemical operations for wastewater treatment. Consequently, space does not permit us to explore the subject to the same depth that we explored the single species, single substrate models. Rather, we will consider only a few key concepts. Those wishing to develop a greater understanding of the subject should read Chapter 11 in *Biofilms*[21] and any of the papers describing modeling efforts.[14,28,31,43–47] Those wishing to use a computer code implementing a general-purpose multispecies biofilm model should acquire a copy of AQUASIM.[29] In Section 16.5, more complex models will also be discussed.

To understand any of the multispecies biofilm models in the literature, it is important to understand why one type of bacteria can be displaced by another within a biofilm. That is, one must understand competition for space. Consider a control volume within a biofilm as shown in Figure 16.7. When net growth is positive in the control volume, the biomass increases. If the biomass density is constant, biomass must cross the control volume's boundary, giving a biomass flux.[21,44,45] The mass balance on the control volume must incorporate the flux of biomass into it from deeper within the biofilm and the flux of biomass out toward the liquid-biofilm interface. The flux of biomass into the control volume is proportional to the integrated net growth of biomass (i.e., growth minus decay) deeper in the biofilm.[31] Biomass debris will also be generated within the biofilm due to decay (Equation 3.57) and it will occupy space, also contributing to the flux. This flux causes the biofilm to increase in depth over time until the flux of biomass due to net growth is just balanced by the loss per unit area due to surface detachment. One effect of this flux, which starts at the biofilm support interface and increases with distance from that interface, is a net migration of particles from the interior of the biofilm to the liquid-biofilm interface. If a particular species of bacteria cannot grow rapidly enough in the biofilm to equal this displacement, it will eventually be lost. Thus, the question of coexistence is one of whether the net growth rate of one species is great enough to allow it to compete with other species for space.

Rittmann and Manem[31] have developed a model for a steady-state biofilm containing i bacterial species that are competing only for space (i.e., they each have their unique electron donor, S_{Si}, and the concentration of the electron acceptor is not rate limiting). It is an extension of the steady-state biofilm model presented in Section 16.2.3 and requires combining a new mass balance on biomass with transport and reaction equations for each of i substrates. In a multispecies biofilm, the density of any one species (including biomass debris, X_{Df}) is a fraction f_{XBi} of the total biomass density, X_{Bf}:

$$X_{B,if} = f_{XBi} X_{Bf}, \tag{16.51}$$

in which $X_{B,if}$ is the density of species i at a point in the biofilm and X_{Bf} is assumed constant. (For this development, X_{Df} will be considered to be one value of $X_{B,if}$) The sum of the densities must equal the total density, thus,

$$\sum_{i=1}^{n} f_{XBi} = \sum_{i=1}^{n} \left(\frac{X_{B,if}}{X_{Bf}} \right) = 1, \tag{16.52}$$

in which n is the total number of biomass types. A steady-state mass balance on species i in a fixed control volume of biofilm leads to:[31]

$$\frac{df_{XBi}}{dx}\sum_{i=1}^{n}\int_0^x(\mu_i - b_i f_D)f_{XBi}dx = (\mu_i - b_i f_D)f_{XBi} - f_{XBi}\sum_{i=1}^{n}(\mu_i - b_i f_D)f_{XBi}, \tag{16.53}$$

in which the values of f_{XBi} and μ_i not inside the integral are for position x only, and f_D is the fraction of active biomass contributing to biomass debris, as defined in Equation 3.53. The rate coefficient b_i is the traditional decay coefficient and μ_i is the specific growth rate for the particular type of biomass represented by i. For example, if species 1 is a heterotroph, μ_1 will be a μ_H and if species 2 is an autotroph, μ_2 will be a μ_A. The boundary condition for Equation 16.53 is that there is no flux of any species into the attachment surface:

$$J_X = 0 \text{ at } x = 0. \tag{16.54}$$

For a steady-state biofilm, the growth of all species due to substrate utilization is just equal to the loss by detachment and decay. For multispecies, this can be written as

$$\sum_{i=1}^{n}(J_{Si}\cdot Y_i) = X_{Bf}\left[b_D\cdot L_f + \sum_{i=1}^{n}\int_0^{L_f}(f_{XBi}f_D b_i)dx\right]. \tag{16.55}$$

This equation can be used in place of Equation 16.26 when writing a model for multispecies biofilm growth and substrate utilization. It is used in combination with equations like Equation 16.23 with associated boundary conditions for each of the i substrates.

Rittmann and Manem[31] then used the concept of S_{Sbmin} to decide whether coexistence of two species is possible. First, S_{Sbmin} should be calculated for each species using Equation 16.22. The species whose bulk substrate concentration is closest to its S_{Sbmin} value is the one whose ability to exist in the biofilm is questionable. For this development, that species will be defined as species 2. Two conditions must be met by species 2 for it to coexist with species 1 in the biofilm. First, it must have a net positive specific growth rate somewhere in the biofilm. As a consequence:

$$S_{S2} > \frac{K_{S2}\cdot b_2}{\hat{q}_2\cdot Y_2 - b_2}, \tag{16.56}$$

at some position in the biofilm. Second, it must have a sufficiently fast specific growth rate to allow it to compete with species 1 for space. Because S_{Sb1}/S_{Sbmin1} is greater than S_{Sb2}/S_{Sbmin2}, the most favorable location for species 2 to compete with species 1 is near the attachment surface, where S_{S1} is the lowest. Because $J_X = 0$ at the attachment surface (Equation 16.54), Equation 16.53 gives for species 2:

$$f_{XB2}\left[(\mu_2 - b_2) - \sum_{i=1}^{n} f_{XBi}(\mu_i - b_i f_D)\right] = 0 \text{ at } x = 0. \tag{16.57}$$

Because f_{XB2} must be greater than zero for species 2 to be present, the terms inside the brackets in Equation 16.57 must equal zero. Consequently,

$$\mu_{2as} - b_2 = b_C, \tag{16.58}$$

in which

$$b_C = f_{XB1as}(\mu_{1as} - b_1 f_D) + f_{XB2as}(\mu_{2as} - b_2 f_D), \tag{16.59}$$

and the subscript "as" refers to the location at the attachment surface. The term b_C is called the competition coefficient. At the limit of coexistence, f_{XB2as} approaches zero, in which case b_C becomes:

$$b_C = f_{XB1as}(\mu_{1as} - f_D b_1). \tag{16.60}$$

The minimum bulk liquid concentration of substrate 2 that will just allow coexistence of the two species, $S_{Sbmin2C}$, can be obtained by noting that it must provide a specific growth rate for species 2 at the attachment surface that will satisfy Equation 16.58. Consequently:

$$S_{Sbmin2C} = \frac{K_{S2}(b_2 + b_C)}{\hat{q}_2 Y_2 - (b_2 + b_C)}, \tag{16.61}$$

where b_C is defined by Equation 16.60. In other words, one can use a single substrate, single species model to estimate μ_{1as}, allowing b_C to be quantified. This will allow quantification of $S_{Sbmin2C}$. If S_{Sb2} is greater than that value, then the two species will be able to coexist in the biofilm.

An examination of Equation 16.61 shows that it is similar to Equation 16.22 for a single species biofilm, except that b_C replaces b_D. This follows from the fact that at the limit of coexistence, species 2 is protected from detachment by species 1, but must grow rapidly enough to compete with it for space. For the special case in which the biofilm is deep for species 1 (i.e., $S_{S1} = 0$ at $x > 0$), f_{XB1as} approaches zero, which makes b_C approach zero.[31] In this case, $S_{Sbmin2C}$ is

$$S_{Sbmin2C} = \frac{K_{S2} \cdot b_2}{\hat{q}_2 Y_2 - b_2}, \tag{16.62}$$

which is the same as for a single-species biofilm undergoing no detachment. When the biofilm is fully penetrated with respect to substrate 1, f_{XB1as} approaches 1, b_C approaches b_D, and $S_{Sbmin2C}$ is given by the normal equation for S_{Sbmin}, Equation 16.22.[31] For all cases between a fully penetrated and a deep biofilm, $S_{Sbmin2C}$ will gradually decrease from the value given by Equation 16.22 to the value given by Equation 16.62. Thus, having a deeper biofilm with respect to species 1 protects species 2 from detachment and lowers its effective S_{Sbmin} value.

The above analysis suggests that the more slowly growing species can be protected from loss provided that b_C is less than b_D. The caveat in this analysis is that it assumes that the two species do not compete for the electron acceptor at limiting concentrations. If that occurs, the specific growth rates of the two species within the biofilm will also be influenced by the electron acceptor concentration profile and their relative half-saturation coefficients for it. This greatly complicates the analysis and makes it difficult to come up with a single criterion as was done above. However, Wanner and Gujer[44] have investigated the question of competition between heterotrophs and autotrophs competing for DO being supplied from the bulk liquid and have examined the conditions for coexistence. The interactive double Monod equation, Equation 3.46, was used to express the effects of the electron donor and the electron acceptor on the specific growth rate of each type of organism. Nitrification was assumed to occur in one step as was done in International Water Association (IWA) Activated Sludge Model No. 1 (ASM; see Chapter 6). Analysis of the question with the kinetic and stoichiometric coefficients in Table 16.2 gave the results in Figure 16.20. Any combination of bulk liquid biodegradable COD and ammonia-N concentrations that lays in zone H will result in a fully heterotrophic biofilm and no nitrification will occur. This suggests that nitrification can only occur in a biofilm process after the bulk of the organic matter has been removed. Any combination that lies in zone A will result in a fully autotrophic biofilm. This suggests that fully autotrophic biofilms will be rare. Finally, combinations of bulk liquid COD and ammonia-N concentrations that lie in zone AH will result in a two-species biofilm in which both nitrification and carbon oxidation occur. However, the larger the bulk liquid COD concentration associated with

TABLE 16.2
Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used to Generate Figure 16.20

Symbol	Units	Value
$\hat{\mu}_H$	hr^{-1}	0.20
K_S	mg/L as COD	5.0
$K_{O,H}$	mg/L as O_2	0.10
Y_H	mg biomass COD/mg substrate COD	0.40
b_H	hr^{-1}	0.0083
$\hat{\mu}_A$	hr^{-1}	0.040
K_{NH}	mg/L as N	1.0
$K_{O,A}$	mg/L as O_2	0.10
Y_A	mg biomass COD/mg N	0.22
b_A	hr^{-1}	0.0021
D_{eS}	cm^2/hr	0.035
D_{eN}	cm^2/hr	0.062
D_{eO}	cm^2/hr	0.073

Note: Data from Wanner, O. and Gujer, W., Competition in biofilms. *Water Science and Technology*, 17 (2/3): 27–44, 1984.

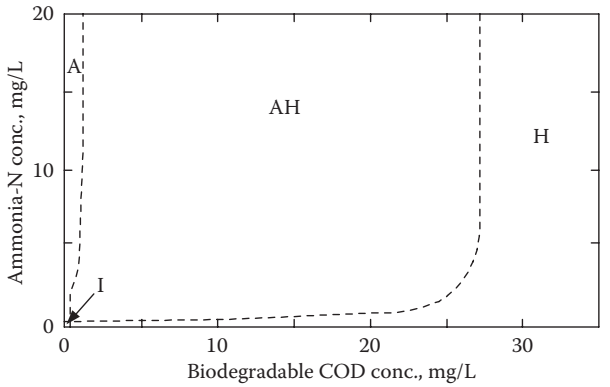

FIGURE 16.20 Effects of bulk fluid ammonia-N and biodegradable COD concentrations on the coexistence of autotrophic and heterotrophic biomass in a biofilm. The kinetic and stoichiometric coefficients used are given in Table 16.2. (After Wanner, O. and Gujer, W., Competition in biofilms. *Water Science and Technology*, 17 (2/3): 27–44, 1984.)

a given ammonia-N concentration, the greater the contributions of the heterotrophs to the biofilm. These effects of competition for space and oxygen have a strong effect on the design of biofilm processes for carbon oxidation and nitrification as we will see in Chapters 19 through 21.

16.5 MULTIDIMENSIONAL MATHEMATICAL MODELS OF BIOFILMS

In the previous sections, relatively simple approaches for modeling biofilms were presented. The discussion was limited to one-dimensional (1-D) systems and most considered the presence of one

electron donor, one nonlimiting electron acceptor, and one organism. The main motivation for doing this was to introduce the reader to the concept of combining mass transfer and reaction (growth) to describe biofilm systems. Biofilms found in natural and engineered environments are significantly more complex in structure and composition than the simple approach, consisting of several organisms coexisting within an EPS matrix with multiple electron donors and acceptors present. In addition, the biofilms are surrounded by fluids that generate a variety of hydrodynamic conditions. Mathematical models for biofilm formation have become more sophisticated in an effort to capture these features and are now able to generate two- and three-dimensional (2-D and 3-D) biofilm structures.[14,28,43] The success of the new models can be seen in the similarity between their results (Figure 16.3c) and the architecture of real biofilms (Figure 16.3b). The multidimensional models are highly comprehensive and account for convection within the biofilm, diffusion of substrates and nutrients, substrate utilization, and biomass generation. They also incorporate the distribution of newly formed biomass into the biofilm and the detachment of biomass from the surface.

In response to the very rapid development of biofilm modeling, IWA appointed a task group on the subject. As stated in the task group report, "a model should be as simple as possible, and only as complex as needed."[47] Guided by this principle the two main objectives of the group were to compare the various levels of modeling complexity against a set of benchmark problems and to provide guidance on how to select the appropriate model for a given set of objectives to be achieved. The models considered by the task group were

1. Analytical models (1-D, single species, and single substrate) and pseudoanalytical models for a single species, as well as for multiple species.
2. Multi-species 1-D numerically solved models (e.g., see references 45 and 46), which provide concentration gradients for all species in the model in the direction perpendicular to the biofilm.
3. Multispecies 2-D and 3-D numerically solved models (e.g., see references 4, 14, 28, 43, and 50) in which the architecture of the biofilm is simulated in detail and external mass transfer limitations are predicted. In addition, advective and diffusive mass transfer mechanisms are considered.

For a detailed description of the results obtained by the IWA task group, the reader should consult their report.[47]

The IWA task group found that simple 1-D models, such as the ones presented in Sections 16.2.2 and 16.2.3, are capable of describing basic substrate removal rates as well as the more complex models. If one needs to describe the behavior of a system at the macroscale, then simple models can (and should) be used. More complex multidimensional models are needed in studying processes at the microscale, such as describing the complex architecture of a biofilm. Multidimensional models incorporating multiple species are becoming powerful tools in the study of microbial ecology. Two examples from the literature are highlighted.

As discussed in Section 2.3.3, H_2 partial pressure is a critical parameter for stable operation of anaerobic systems. If it is higher than 10^{-4} atmospheres, the reactions leading from long chain fatty acids, volatile acids, amino acids, and carbohydrates to acetic acid and H_2 become thermodynamically unfeasible. Batsone et al.[4] used fluorescence in situ hybridization (FISH) with 16S rRNA-directed oligonucleotide probes to determine the spatial distribution of H_2-producing and H_2-utilizing organisms in granules from upflow anaerobic sludge blanket reactors and observed a close syntrophic association between them. Simulations using a 3-D model demonstrated the same type of colocation between H_2-producing and H_2-consuming organisms.[5]

In Section 2.3.3 we also introduced the anammox bacteria, which are obligate anaerobes able to oxidize ammonia using nitrite as the electron acceptor, with formation of nitrogen gas and nitrate. Egli et al.[15] used FISH to characterize the biofilm of a rotating biological contactor (RBC) treating leachate from a hazardous waste landfill containing a high ammonia concentration

(500 mg N/L) and a low organic carbon concentration. They selected this RBC for study because a nitrogen balance across it revealed an apparent loss of 70% of the nitrogen from the liquid phase, something that would not be anticipated from autotrophic nitrification and heterotrophic denitrification based on the leachate characteristics. The biofilm from the RBC was probed for the existence of ammonia oxidizing bacteria (AOB), nitrite oxidizing bacteria (NOB), and anammox bacteria. The results revealed that 20–30% of the biomass was anammox bacteria and 20–30% was AOB, whereas NOB accounted for less than 5% of the biomass. In terms of the spatial distribution, AOB and NOB were mainly at the surface of the biofilm and the anammox bacteria were in the bottom layers. This distribution can be explained by oxygen being the limiting nutrient in this system.

In the top sections of the biofilm, AOB oxidize ammonia to nitrite, resulting in oxygen consumption. Because of the high ammonia concentration, the AOB will likely be growing at their maximum specific growth rate and oxygen will rapidly be depleted. Under these conditions NOB will be displaced from the biofilm because they cannot grow as rapidly as the AOB. Consequently, nitrite will accumulate and diffuse into the biofilm below the aerobic layer. Furthermore, because oxygen is the limiting nutrient, not all ammonia will be oxidized to nitrite, causing it to also diffuse into the oxygen depleted areas of the biofilm. Under these conditions, anammox bacteria will be able to grow because both their electron donor and their electron acceptor are present, resulting in the formation of nitrate and nitrogen gas (the reason for the apparent loss of nitrogen from the liquid phase). Picioreanu et al.[27] developed a 2-D model capable of simulating a biofilm containing AOB, NOB, and anammox bacteria. Two sets of simulations were performed. The bulk ammonia-N concentration was set to 30 mg-N/L in both, but the oxygen concentration was set to either 2 or 10 mg/L. The results of these simulations are very consistent with the observations of Egli et al.[15] A biofilm composed primarily of AOB and anammox bacteria was obtained for the case in which the oxygen concentration in the bulk fluid was low.

These examples illustrate that models are now able to capture the ecological relationships observed in real biological systems. As a result, they will undoubtedly have an important role to play in the field of microbial ecology.

16.6 KEY POINTS

1. Biofilms are nonuniform structures consisting of discrete cell clusters attached to each other and to a solid support with extracellular polymeric material. The spaces between clusters form vertical and horizontal voids, with the vertical voids acting as pores and the horizontal voids acting as channels.
2. In an attached growth bioreactor, the substrate concentration at which the reaction takes place is less than the substrate concentration in the bulk liquid phase because of the resistances encountered in the transport of substrate from the bulk liquid into the biofilm. Therefore, it is necessary to combine physical mass transport with microbial reactions when modeling attached growth reactors.
3. Mass transfer from the bulk liquid phase to the liquid-biofilm interface is considered to occur across a hypothetical stagnant liquid film adjacent to the interface. The rate of mass transfer across that film is proportional to the drop in substrate concentration across it and the proportionality constant, k_L (T^{-1}), is called the mass transfer coefficient.
4. The transport of substrate within a biofilm may be described by Fick's first law with the free diffusion coefficient, D_w (L^2T^{-1}), replaced by an effective diffusion coefficient, D_e.
5. The term flux refers the mass flow rate of a substance per unit area. The flux of substrate across the stagnant liquid layer to the liquid-biofilm interface must equal the flux of substrate into the biofilm by diffusion, which must equal the rate of substrate consumption per unit of biofilm planar surface area. Thus, the terms flux and substrate utilization rate are used interchangeably for biofilms.

6. The effectiveness factor, η_e, is defined as the ratio of the actual, observed substrate removal rate per unit of biofilm planar surface area to the theoretical rate that would occur in the absence of mass transfer resistance. Thus, the effectiveness factor is a correction factor, which, when multiplied by the homogeneous reaction rate without mass transfer resistance, gives the actual rate in the presence of mass transfer resistance.
7. The overall effectiveness factor, η_{eO}, accounts for the presence of both external and internal mass transfer resistance. Its values must be determined numerically and are commonly presented graphically in terms of a Sherwood number called the Biot number, Bi, the Thiele modulus, ϕ, and a modified Thiele modulus, ϕ_f. They are functions of the bulk liquid substrate concentration. In general, the overall effectiveness factor decreases rapidly with increases in ϕ_f, increases slightly with increases in ϕ, and increases rapidly with increases in Bi.
8. The pseudoanalytical approach to modeling transport and reaction utilizes simple algebraic expressions for the flux of substrate into a biofilm. The simple algebraic expressions are based on an analysis of the results from the numerical solution of the differential equations describing transport and reaction in a biofilm. The availability of simple algebraic equations eliminates the need to repetitiously solve numerically a set of nonlinear differential equations while modeling the performance of a biofilm reactor.
9. A steady-state biofilm is one in which the gains in biofilm mass due to biomass growth are just balanced by the losses in biofilm mass due to the combined effects of microbial decay within the biofilm and detachment by shear at the liquid-biofilm interface. As a consequence, a minimum bulk liquid substrate concentration, S_{Sbmin}, is required to drive the growth reactions to balance the losses and concentrations below that value cannot support a steady-state biofilm.
10. The dimensionless minimum bulk liquid substrate concentration, S^*_{Sbmin}, is determined solely by parameters that depend on the biomass, substrate, and fluid regime. It represents the ratio of the biomass specific loss rate to the biomass maximum net specific growth rate, giving it special significance for steady-state biofilms. Consequently, it has been named the Rittmann number and given the symbol Ri.
11. Implementation of the pseudoanalytical approach for the analysis and design of biofilm reactors is made easier by the use of normalized loading curves, which present a normalized flux as a function of the normalized minimum bulk substrate concentration. Normalization of the plots makes them applicable for a broad range of parameter values.
12. Closed form solutions allowing direct computation of the substrate flux into a biofilm are possible for several limiting cases: a deep biofilm, a fully penetrated biofilm, a first-order biofilm, and a zero-order biofilm. Error analysis has shown that all but the deep biofilm case are applicable only under very restricted conditions.
13. Multispecies biofilm modeling requires that all electron donors and electron acceptors for both heterotrophs and autotrophs be considered, just as they were in the International Water Association activated sludge models. Furthermore, appropriate transport and reaction equations must be incorporated for each component. The complexity involved prevents the effectiveness factor and pseudoanalytical approaches from being used. Rather, a set of differential equations must be solved simultaneously.
14. Because there is a net flux of particulate material from the biofilm support toward the liquid-biofilm interface, more rapidly growing bacteria can displace slowly growing bacteria from a biofilm. However, because bacteria in the interior of the biofilm are protected from surface shear, rapidly growing bacteria can protect slowly growing bacteria, allowing them to survive in a biofilm under conditions where they would not be able to form a biofilm alone.

15. Multispecies modeling results suggest that because of competition for space and dissolved oxygen, nitrifying bacteria cannot survive in biofilms when the bulk organic substrate concentration exceeds approximately 27 mg/L as chemical oxygen demand. However, nitrifiers and heterotrophs can coexist in biofilms for concentrations below that value.
16. Complex multidimensional models describing the detailed structure of biofilms allow capturing complex interactions among different organisms within biofilms and they are powerful tools in the field of microbial ecology.

16.7 STUDY QUESTIONS

1. Describe the structure of a biofilm as it is now conceived and contrast that conceptualization with the conceptualization used in the simpler mathematical models of biofilms.
2. Define the mass transfer coefficient, k_L, and relate it to the thickness of the hypothetical stagnant liquid layer that is assumed to lie between the bulk liquid and the liquid-biofilm interface.
3. Describe the transport of substrate within a biofilm, including in your description an explanation of the effective diffusivity.
4. Draw a sketch depicting the substrate concentration profile within a biofilm. Then tell what the slope of the profile must be at the liquid-biofilm interface as well as at the biofilm support interface (assuming an impermeable support) and explain why.
5. What is the role of an effectiveness factor?
6. Name and define the three parameters that characterize the overall effectiveness factor for Monod kinetics in a flat biofilm. Describe briefly the effects of these parameters on the value of the overall effectiveness factor.
7. A CSTR with a volume of 1.0 L receives a sterile feed with a substrate concentration of 200 mg/L as COD at a flow rate of 1.0 L/hr. The CSTR contains a biofilm with the kinetic and stoichiometric parameters listed in Table SQ16.1. The system variables are also listed there. Prepare a graph showing the surface area of biofilm required to reduce the effluent substrate concentration to values between 1 and 100 mg/L. Compute at least four data points and then explain why the curve has the shape that it does. Use the effectiveness factor approach.
8. Describe a steady-state biofilm and state the conditions required to maintain it.
9. Explain why S_{Sbmin} has special significance as a parameter in the pseudoanalytical approach to solving the equations quantifying substrate flux into a steady-state biofilm. Then explain the significance of the Rittmann number, telling what values greater than and less than 1.0 suggest.

TABLE SQ16.1
Kinetic Parameters, Stoichiometric Coefficients, and System Variables for Study Question 7

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD·hr)	0.33
K_S	mg/L as COD	20
Y_{Hobs}	mg biomass COD/mg substrate COD	0.60
$X_{B,Hf}$	mg biomass COD/cm^3	40
k_L	cm/hr	5.0
D_e	cm^2/hr	0.025
L_f	cm	0.020

TABLE SQ16.2
Kinetic Parameters, Stoichiometric Coefficients, and System Variables for Study Questions 10 and 11

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD·hr)	0.75
K_S	mg/L as COD	15.0
Y_H	mg biomass COD/mg substrate COD	0.54
b_H	hr^{-1}	0.004
$X_{B,Hf}$	mg biomass COD/cm^3	45
k_L	cm/hr	5.0
D_e	cm^2/hr	0.025
b_D	hr^{-1}	0.003

10. A steady-state biofilm, described by the kinetic and stoichiometric coefficients given in Table SQ16.2, exists in an environment where the bulk substrate concentration is maintained at a constant value. Determine how the bulk substrate concentration affects the substrate flux into the biofilm and the biofilm thickness. Calculate at least four points over the range of bulk substrate concentrations from 0.1 to 1.0 mg/L.
11. A synthetic wastewater with a biodegradable COD of 20 mg/L (0.020 mg/cm^3) is flowing at a rate of 1.0 L/hr (1000 cm^3/hr) into a single CSTR containing a biofilm media with a specific surface area of 150 m^2/m^3 (1.50 cm^2/cm^3). The wastewater, the bioreactor, and the associated biofilm have the characteristics listed in Table SQ16.2. What total surface area of biofilm would be required to reduce the biodegradable COD to 0.5 mg/L (0.0005 mg/cm^3)? What bioreactor volume is required to house the media? Determine your answer by the pseudoanalytical approach.
12. Repeat Study Question 11 using the normalized loading curves.
13. Consider the situation presented in Examples 16.2.3.1 and 16.2.3.2. If a synthetic wastewater with the same kinetic characteristics was introduced at a rate of 1.0 L/hr into a chain of three CSTRs containing the same media, what influent COD concentration could be treated to a final effluent concentration (i.e., from bioreactor number three) of 0.5 mg/L of COD if each bioreactor had a volume of 1.0 L?
14. Describe the situations under which each of the limiting case solutions of the biofilm flux equation (deep, fully penetrated, first-order, and zero-order biofilms) are applicable.
15. Describe the situations under which it is acceptable to assume that a biofilm is behaving as if it were limited by only the electron donor or the electron acceptor.
16. Explain why a fast growing species of bacteria is able to prevent a slow growing species from growing in a biofilm even when they are not competing for either the electron donor or the electron acceptor.
17. What is the competition coefficient and what is its significance to the determination of whether two species can coexist in a biofilm when they are competing only for space and not for the electron donor or acceptor?
18. Describe the conditions under which autotrophs and heterotrophs can coexist in a biofilm in which they are competing for dissolved oxygen and explain why they are able to do so.

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17 Biofilm Reactors

As seen in Chapter 1, there are several types of attached growth bioreactors. In this chapter we will focus on two bioreactor configurations: packed towers (referred to as trickling filters in practice) and rotating disc reactors (referred to as rotating biological contactors (RBCs) in practice). The performance of these systems is affected by growth kinetics, as well as by the transport of the electron donor(s) and electron acceptor(s) across the biofilm. To better understand the behavior of these two popular attached growth systems, simple mathematical models describing their behavior are presented and used to demonstrate their general performance. In addition, other models available in the literature are discussed.

Throughout this chapter, all substrate and biomass concentrations will be expressed in chemical oxygen demand (COD) units. Conversion to other units can be accomplished with the techniques discussed previously.

17.1 PACKED TOWERS

17.1.1 Description and Simplifying Assumptions for Model Development

Among attached growth bioreactors, the most widely used is the packed tower, which contains microorganisms growing on an immobile support over which wastewater flows in thin sheets. Although packed towers are called trickling filters in practice, we will refer to them as packed towers herein to focus on the type of reactor employed. Most recently installed packed towers use plastic media as the immobile support. Two types are in current use, random packing that is typically in the form of cylinders approximately 5 cm in diameter and 5 cm long, and bundle media that consists of sheets formed into self-supporting modules with vertical surfaces. Clarified wastewater is distributed uniformly over the top of the media by the distribution system, which may be either rotary or fixed. Fixed nozzles may discharge either continuously or intermittently, in which case they have hydraulic characteristics similar to rotary distributors. After passing over the media in thin sheets, the treated wastewater is collected in the underdrain system, which is open to the atmosphere to allow free movement of air through the tower. From there, it flows to a gravity settler for removal of biomass.

When wastewater containing organic matter, ammonia-N, or other electron donors flows through a packed tower, microorganisms consume the substrates and grow attached to the media as a biofilm. The flow of water over the biofilm imparts a shear force to it, keeping its thickness relatively constant. The suspended biomass is then removed by gravity settling before discharge of the treated effluent. In some cases, effluent from the tower is recirculated to the top to allow control of the flow rate through the tower and to influence the concentration of substrate. The recirculated flow is generally clarified effluent, but not always.

Although a packed tower is mechanically simpler than most suspended growth systems, it is more difficult to model for a number of reasons. First, both the electron donor and the electron acceptor must be transported into the biofilm for reaction, as discussed in Chapter 16. This means that the models for substrate removal are much more complicated. Second, although flow is generally assumed to be in thin sheets in most models, in reality, the flow patterns are quite complex over all of the media types. Third, various types of bacteria must compete for nutrients and space in the biofilm, as discussed in Section 16.4, rather than being homogeneously distributed as in a suspended growth culture. Fourth, the biofilm will not be evenly distributed over all of the media;

rather, the distribution will depend on the flow patterns of the fluid and the concentration of the substrate in it.

As a consequence of the various factors in the preceding paragraph, no consensus models yet exist for biofilm processes that are comparable to the International Water Association (IWA) activated sludge models. Models exist that account for transport and reaction in the manner discussed in Chapter 16, but they are generally limited to a single substrate and consider fluid flow to be in thin sheets with no intermixing. At the other extreme, models exist that seek to account for the complexity in fluid flow through packed towers, but they use limiting case solutions for transport and reactions, thereby limiting their generality, as discussed in Section 16.2.4. As a consequence, it is currently impossible to use mechanistic modeling to investigate the theoretical performance of packed towers with the same degree of confidence that modeling was used in Part II to investigate suspended growth systems. Nevertheless, much can be learned from models, as long as one views their results qualitatively rather than quantitatively. Consequently, we will use such a model herein to learn how certain factors influence packed tower performance and then will consider other important aspects not included in the model.

By making certain simplifying assumptions, both the effectiveness factor and the pseudoanalytical approach can be used to investigate the theoretical performance of a packed tower. Grady and Lim[7] used the effectiveness factor approach, but the pseudoanalytical approach will be used herein. Before developing the model, some assumptions must be made about the packed tower and its operation. The first assumption is that the electron acceptor is present in excess so that the electron donor acts as a single limiting nutrient. The second is that only one type of microorganism is present; heterotrophs if the electron donor is organic matter and autotrophs if it is ammonia-N. The results will be presented in terms of heterotrophs removing soluble chemical oxygen demand (COD), but they will be qualitatively similar to those for autotrophs oxidizing ammonia-N in the absence of heterotrophs. The third is that steady-state conditions prevail for the biofilm, which means that the rate at which biomass is formed within the biofilm is equal to the rate at which it is lost from the biofilm by the combined effects of decay and detachment. The fourth assumption is that the density of the biofilm is not dependent on the ratio of active biomass to biomass debris. The fifth is that the detached biomass is carried along with the liquid flow and contributes to substrate removal. The sixth is that the surface area of biofilm is equal to the surface area of the media present in the tower, which means that there are no patches of bare media. The seventh is that there is no limit on the maximum thickness of biofilm within the tower; in other words, there is not a prespecified limit on the total mass of biomass in the packed tower. The eighth assumption is that only clarified effluent is recirculated around the tower and that the clarifier is perfect so that the recirculation contains no biomass. The ninth is that the flow pattern in the packed tower can be described by considering any number (from 1 to N) of continuous stirred tank reactors (CSTRs) in series. All of these assumptions have important implications to the output from the modeling effort and prevent the model from being used as a quantitative design tool. Nevertheless, the output is sufficiently qualitatively accurate to allow the major characteristics of packed towers to be observed and understood.

17.1.2 Model Development

The schematic diagram of a packed tower is shown in Figure 17.1. A wastewater at flow rate F containing a soluble biodegradable substrate with a concentration S_{SO} flows to the tower where it is intermixed with recirculated effluent from the clarifier containing residual substrate at concentration S_{Se}, thereby diluting the influent concentration to S_{Sa}, which is the applied concentration. The value of S_{Sa} may be calculated from a mass balance on the mixing point:

$$S_{Sa} = \frac{S_{SO} + \alpha S_{Se}}{1 + \alpha}, \tag{17.1}$$

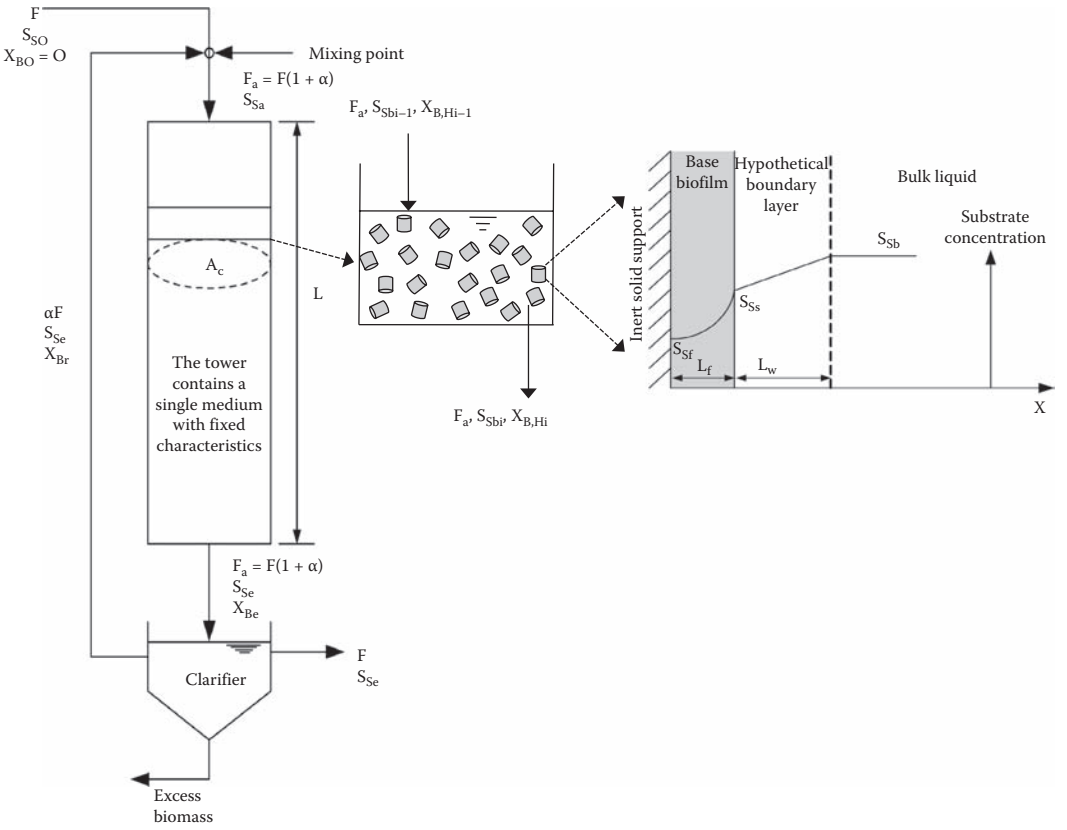

FIGURE 17.1 Schematic diagram of a packed tower with recirculation of clarified effluent. The tower was simulated as 25 CSTRs in series, with biofilm growing attached to the media in each CSTR. A liquid boundary layer exists between the bulk liquid and the surface of the biofilm.

where α is the fraction of the influent flow that is recirculated. The total flow rate applied to the tower, F_a, is $F(1 + \alpha)$ and it is applied uniformly and continuously across the entire cross-sectional area, A_c. The cross-sectional area for fluid flow is $f_{Ac}A_c$, where f_{Ac} is the fraction of the tower cross-sectional area occupied by the liquid film. The total volume of liquid in the packed tower is V_L, which is equal to $f_{Ac}A_c$ multiplied by the height of the tower, L. The influent wastewater is considered to be devoid of biomass, as is the recirculated flow. Thus, no biomass enters the tower. However, because the biofilm in the tower is considered to be at steady state, biomass is continuously removed from its surface and carried down through the tower to the clarifier. This means that the concentration of biomass in suspension increases as the fluid moves through the tower, giving a concentration X_{Be} in the flow entering the clarifier. The suspended biomass will contribute to substrate removal, but the attached biofilm will be responsible for most of it. The surface area of the media is A_s, which is assumed to be equal to the total surface area of biofilm in the tower, and the total media volume is V_M. The biofilm thickness is L_f.

The packed tower was modeled as a series of CSTRs and one is shown in Figure 17.1. There are two reasons for taking this approach. The first concerns the flow regime. Grady and Lim[7] assumed that fluid flow through a packed tower behaved in a perfect plug-flow manner, with no vertical intermixing. We now know that intermixing occurs via the flow of droplets from protrusions on the biofilm;[15] thus, the assumption of perfect plug flow is not correct. As discussed in Section 4.4.1, modeling the tower as a series of CSTRs allows a broad range of flow regimes to be described, ranging from a single CSTR to a plug-flow reactor (PFR). The second reason for modeling the system

as a series of CSTRs is to simplify the numerical solution technique. The method chosen is that of Heath et al.,[14] who combined a model for a simple CSTR with the pseudoanalytical approach for modeling reaction in a biofilm.

Model development is started by performing a steady-state material balance on substrate for the i^{th} CSTR in the series of N CSTRs used to mimic the packed tower:

$$F_a \cdot S_{Sbi-1} - F_a \cdot S_{Sbi} - \frac{\hat{\mu}_H}{Y_H}\left(\frac{S_{Sbi}}{K_S + S_{Sbi}}\right) X_{B,Hi} \cdot V_{Li} - J_{Si} \cdot A_{si} = 0, \tag{17.2}$$

where S_{Sb} is the bulk liquid phase substrate concentration and the subscripts i and i – 1 represent the concentrations in the i^{th} CSTR and the feed to it, respectively. The symbol V_{Li} is the liquid volume, J_{Si} is the substrate flux into the biofilm, A_{si} is the planar surface area of biofilm, and $X_{B,Hi}$ is the biomass concentration, all in the i^{th} CSTR. Equation 17.2 states that the total rate of substrate removal is due to two processes: substrate utilization for the growth of suspended biomass and growth of the biomass in the biofilm. The substrate flux into the biofilm, J_S, is defined as the mass of substrate entering the biofilm, per unit time, per unit area of biofilm. As discussed in Chapter 16, the substrate flux is related to the bulk substrate concentration. For a steady-state biofilm, the pseudoanalytical approach (Section 16.2.3) can be used to find the flux associated with any given bulk substrate concentration. The top of the tower is represented by the first reactor in the series of CSTRs, where $S_{Sbi-1} = S_{Sa}$. At the bottom of the tower, the last reactor in the series generates an effluent $S_{SbN} = S_{Se}$.

The next step in the development of the model is to perform a steady-state material balance on biomass suspended in the liquid phase of the i^{th} CSTR:

$$\begin{aligned} F_a \cdot X_{B,Hi-1} - F_a \cdot X_{B,Hi} + \hat{\mu}_H\left(\frac{S_{Sbi}}{K_S + S_{Sbi}}\right) X_{B,Hi} \cdot V_{Li} \\ - b_H \cdot X_{B,Hi} \cdot V_{Li} + b_D \cdot X_{B,Hf} \cdot A_{si} \cdot L_{fi} = 0. \end{aligned} \tag{17.3}$$

This equation states that the mass of suspended biomass in the i^{th} CSTR increases due to biomass entering the reactor with the feed from the previous CSTR (i – 1), biomass growth due to substrate utilization by suspended cells, and biomass detachment from the biofilm (b_D). Suspended biomass is lost from the reactor through the effluent and through biomass decay (b_H). At steady state, the rate of biomass accumulation in the reactor is equal to zero. For the top of the tower (i = 1), $X_{B,Hi-1} = X_{B,HO} = 0$ and for the bottom of the tower (i = N), $X_{B,HN} = X_{B,He}$. In Chapter 5, the material balance on biomass across a single CSTR containing only suspended biomass led to a key relationship between the solids retention time (SRT) and the specific growth rate of the organisms growing in the reactor (Equation 5.21). Equation 17.3 is substantially more complex and does not directly lead to such a relationship, because biomass can originate from two different processes: biofilm detachment and suspended growth. Moreover, Equation 17.3 shows that washout of the suspended biomass will not occur because each CSTR after the first is constantly seeded by biomass detaching from the biofilm. An additional variable present in Equation 17.3 is the biofilm thickness (L_f). This variable can be expressed as a function of the substrate flux into the biofilm, J_S, by performing a material balance on the biomass in the biofilm in the i^{th} CSTR:

$$J_{Si} \cdot Y_H \cdot A_{si} - (b_D + b_H) X_{B,Hfi} \cdot A_{si} \cdot L_{fi} = 0. \tag{17.4}$$

For a steady-state biofilm, the rate at which biomass is formed in the biofilm (through substrate consumption) is equal to the rate at which biomass is lost from the biofilm through decay and biomass

detachment. The term $X_{B,Hfi}$ is the biofilm density in the i[th] CSTR. At steady state, the biofilm thickness in the i[th] CSTR can be calculated as

$$L_{fi} = \frac{J_{Si} Y_H}{(b_H + b_D) X_{B,Hfi}} \tag{16.21a}$$

(The addition of "a" to the equation numbers from Chapter 16 indicates that they have been modified by the addition of subscript "i.") Combination of Equations 17.2, 17.3, and 16.21a results in

$$S_{Sbi} - S_{Sbi-1} + \tau_i \left\{ \frac{\hat{\mu}_H}{Y_H} \cdot \frac{S_{Sbi}}{(K_S + S_{Sbi})} \frac{\left[X_{B,Hi-1} + \frac{J_{Si} Y_H b_D \tau_i}{(b_D + b_H)} \left(\frac{A_{si}}{V_{Li}} \right) \right]}{\left[1 + b_H \tau_i - \hat{\mu}_H \left(\frac{S_{Sbi}}{K_S + S_{Sbi}} \right) \tau_i \right]} + J_{Si} \left(\frac{A_{si}}{V_{Li}} \right) \right\} = 0, \tag{17.5}$$

where τ_i is the hydraulic residence time of the i[th] CSTR based on the total flow through it, F_a.

Inspection of Equation 17.5 reveals that there are two variables that remain to be defined: the bulk substrate concentration in the liquid phase, S_{Sb}, and the substrate flux into the biofilm, J_S. These can be calculated by combining Equation 17.5 with the pseudoanalytical approach described in Section 16.2.3. To use this approach, we start by calculating the following dimensionless variables:

$$Ri = \frac{b_H + b_D}{Y_H \cdot \hat{q}_H - (b_H + b_D)}, \tag{16.34}$$

$$\alpha' = 1.5557 - 0.4117 \tanh(\log_{10} Ri), \tag{16.36}$$

$$\beta' = 0.5035 - 0.0257 \tanh(\log_{10} Ri), \tag{16.37}$$

and

$$k_L^* = k_L \left(\frac{K_s}{\hat{q}_H \cdot X_{B,Hf} \cdot D_e} \right)^{0.5}, \tag{16.40}$$

where Ri is the Rittmann number, α' and β' are empirical parameters, k_L is the external mass transfer coefficient, and D_e is the effective diffusivity in the biofilm. Assuming an initial value for the bulk substrate concentration in the i[th] CSTR, S_{Sbi}, the assumed dimensionless bulk substrate concentration in that CSTR, S_{Sbi}^*, can be calculated:

$$S_{Sbi}^* = \frac{S_{Sbi}}{K_S}. \tag{16.39a}$$

Using S_{Sbi}^*, Equation 16.38a can be solved implicitly to determine the dimensionless substrate concentration at the surface of the biofilm, S_{Ssi}^*:

$$S_{Sbi}^* = S_{Ssi}^* + \frac{\tanh\left[\alpha' \left(\frac{S_{Ssi}^*}{Ri} - 1 \right)^{\beta'} \right] \{2[S_{Ssi}^* - \ln(1 + S_{Ssi}^*)]\}^{0.5}}{k_L^*}. \tag{16.38a}$$

The dimensionless variable S_{Ssi}^* can then be rescaled with Equation 16.30a to determine the substrate concentration at the biofilm surface, S_{Ssi}:

$$S_{Ssi} = K_S \cdot S_{Ssi}^*. \tag{16.30a}$$

Having found an S_{Ssi} value that is consistent with the assumed S_{Sbi} value, we can now calculate the substrate flux into the biofilm in the i^{th} CSTR using:

$$J_{Si} = k_L (S_{Sbi} - S_{Ssi}). \tag{16.2a}$$

Using this value of J_{Si}, Equation 17.5 can then be solved to find the value of S_{Sbi} that satisfies the equation. If the value obtained is different from the initially assumed value for S_{Sbi}, then a new value of S_{Sbi} needs to be assumed and the procedure repeated until convergence is attained. For a single CSTR, a spreadsheet can be used to obtain the value for S_{Sb}. For simulation of a series of CSTRs, a more robust approach is required.

For use herein, a program was developed in MATLAB® using a procedure equivalent to the one described above. For cases in which no flow was recirculated around the tower ($\alpha = 0$), the computations began with the first CSTR using an applied substrate concentration of S_{SO}, the concentration in the influent flow. After the bulk substrate concentration in the first CSTR was obtained, it was used as the influent substrate concentration to the second CSTR, and so forth down the chain to the last CSTR. The bulk substrate concentration in it was the effluent substrate concentration, S_{Se}. For cases in which flow was recirculated around the tower (i.e., $\alpha \neq 0$), it was necessary to iterate around the entire tower because the applied substrate concentration depended upon the effluent substrate concentration, as indicated by Equation 17.1. The procedure was started by assuming an initial estimate of S_{Se} and iterating until the calculated values of S_{Se} from two consecutive iterations were the same, indicating that convergence was achieved.

17.1.3 Dependence of Substrate Flux on Bulk Substrate Concentration

A key concept in understanding the performance of any bioreactor is how the substrate removal rate is influenced by the bioreactor's bulk substrate concentration. Because the substrate removal rate is stoichiometrically linked to the growth rate of the microorganisms by the yield coefficient (Equation 3.34), the two measures of biomass activity are influenced by the substrate concentration surrounding the biomass in qualitatively similar ways. We have used Monod kinetics (Equation 3.36) to represent the relationship between biomass activity and substrate concentration for suspended growth cultures (Parts II and III). Furthermore, in suspended growth cultures we have assumed that all biomass is in contact with the bulk substrate concentration, so that the Monod equation adequately describes the relationship between biomass activity and the bulk substrate concentration. We will also use the Monod equation to represent the relationship between the activity of biomass and the substrate concentration surrounding them for attached growth bioreactors, but with an additional level of complexity because in attached growth bioreactors the substrate concentration surrounding the biomass is not the bulk substrate concentration. Rather, it is less than the bulk substrate concentration because of the need to transport substrate up to and through the biofilm, as discussed in Section 16.2.1. Thus, the overall rate of substrate removal is determined not only by the kinetics of microbial growth but by mass transfer effects as well. We will briefly examine how this impacts the apparent relationship between bulk substrate concentration and activity.

The overall rate of substrate removal is best described by the substrate flux, which is the final result of the combination of microbial growth and mass transfer phenomena. Figure 17.2 shows how the substrate flux into a biofilm depends on the bulk substrate concentration. It was generated using the kinetic and stoichiometric parameter values in Table 17.1 by assuming a bulk substrate concentration, S_{Sb}, solving Equation 16.38 for the corresponding substrate concentration at the biofilm surface, S_{Ss}, and using the values of S_{Sb} and S_{Ss} to calculate the substrate flux with Equation 16.2. The first reaction upon seeing Figure 17.2 is to compare the curve to a Monod curve with a very large half-saturation coefficient, K_S, thereby letting the large K_S value account for the mass transport limitations. This approach is frequently found in the literature, but it is misleading. To see how mass transport and reaction kinetics affect Figure 17.2, we will make use of the fact that Monod

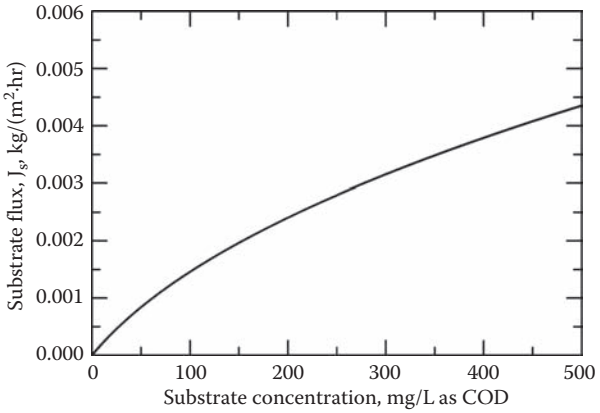

FIGURE 17.2 Effect of the bulk substrate concentration on the substrate flux for a steady-state biofilm. The values of the kinetic parameters and stoichiometric coefficients are given in Table 17.1.

TABLE 17.1
Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used to Generate Figures 17.2–17.12

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD·hr)	0.27
$\hat{\mu}_H$	hr^{-1}	0.135
K_S	mg/L as COD	10
Y_H	mg biomass COD/mg substrate COD	0.50
b_D	hr^{-1}	0.0017
b_H	hr^{-1}	0.0025
$X_{B,Hf}$	mg biomass COD/cm^3	40
D_e	m^2/hr	2.667×10^{-6}
k_L	m/hr	0.033
A_c	m^2	50
f_{AC}	—	0.4
L	m	5
V_L	m^3	100
A_s	m^2	15,000
V_M	m^3	100
F	m^3/hr	100
α	—	0
S_{SO}	mg/L as COD	100
$X_{B,HO}$	mg/L as COD	0

kinetics represents a dependence of the rate on the substrate concentration that transitions from first-order, with rate coefficient $\hat{q}_H/Ks$, to zero-order, with coefficient $\hat{q}_H$, as the substrate concentration increases. Analytical solutions for first-order and zero-order kinetics are available and can be used to see the influence of reaction kinetics on Figure 17.2.

For first-order kinetics, the analytical solution for the substrate flux into a deep biofilm (a biofilm in which the substrate concentration becomes zero before the surface upon which the biofilm is growing is reached) is[12]

$$J_{first,deep} = \left(\frac{D_e \cdot \hat{q}_H \cdot X_{Hf}}{K_S} \right)^{0.5} (S_{Ss}). \tag{17.6}$$

In the presence of external mass transfer resistance (Equation 16.2), the flux becomes:

$$J_{first,deep} = \frac{D_e \cdot S_{Sb}}{\frac{D_e}{k_L} + \left(\frac{D_e \cdot K_S}{\hat{q}_H \cdot X_{Hf}} \right)^{0.5}}. \tag{17.7}$$

The most important point to take from Equation 17.7 is that for first-order kinetics the flux depends in a first-order manner on the bulk substrate concentration, which is expected since all three process acting on the substrate (mass transfer to the biofilm, diffusion through the biofilm, and substrate consumption within the biofilm) are first-order processes. Thus, at low bulk substrate concentrations, where Monod kinetics behaves in a first-order manner, the overall rate combining transport and reaction also behaves in a first-order manner. This is illustrated in Figure 17.3, which shows that for low bulk substrate concentrations (inset), the substrate flux for first-order kinetics is equal to the flux for Monod kinetics. In other words, the first-order line is tangent to the curve for Monod kinetics. Figure 17.3 clearly shows that as the bulk substrate concentration increases, Monod kinetics deviates from zero-order kinetics and, hence, Equation 17.7 is no longer applicable.

At substrate concentrations much larger than K_S, Monod kinetics behaves in a zero-order manner. Thus, we will also consider this case. The analytical solution for the substrate flux into a deep biofilm with zero-order substrate removal kinetics is[12]

$$J_{zero,deep} = \left(2D_e \cdot \hat{q}_H \cdot X_{Hf} \cdot S_{Ss}\right)^{0.5}. \tag{17.8}$$

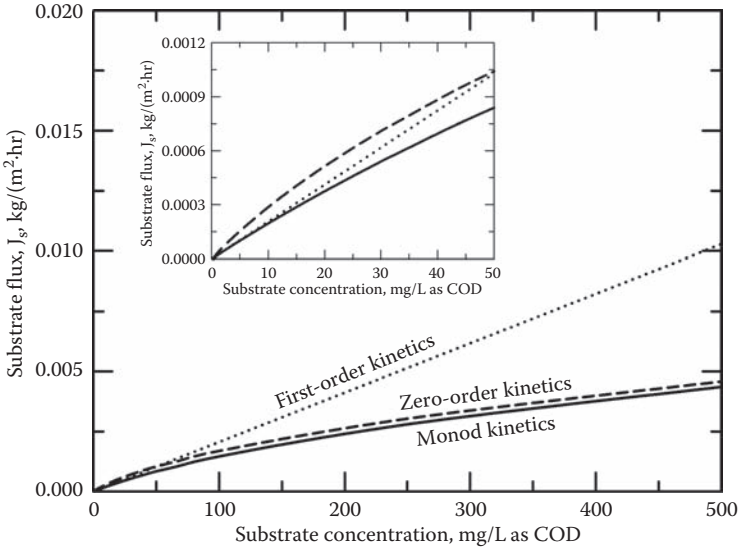

FIGURE 17.3 Comparison of the dependence of substrate flux on the bulk substrate concentration for zero-order, first-order, and Monod kinetics for substrate removal. The inset shows the lower bulk substrate concentration range in detail. The first-order case was obtained with Equation 17.7 and the zero-order case with Equation 17.9. The Monod curve is from Figure 17.2. The values of the kinetic parameters and stoichiometric coefficients are given in Table 17.1.

In the presence of external mass transfer resistance (Equation 16.2), the flux becomes:

$$J_{zero,deep} = \frac{(2D_e \cdot \hat{q}_H \cdot X_{Hf})^{0.5}}{2k_L}\left[(2D_e \cdot \hat{q}_H \cdot X_{Hf} + 4k_L^2 \cdot S_{Sb})^{0.5} - (2D_e \cdot \hat{q}_H \cdot X_{Hf})^{0.5}\right]. \quad (17.9)$$

The most important point to take from Equation 17.9 is that for zero-order kinetics, the substrate flux into the biofilm follows a half-order dependence on the bulk substrate concentration. The reason for this is that even though substrate consumption in the biofilm is a zero-order process, mass transport within the biofilm by diffusion is a first-order process. When the bulk substrate concentration approaches zero, the substrate flux also approaches zero because the concentration gradient driving diffusion is very low. As the bulk substrate concentration is increased, the ability of a given thickness of biofilm to remove substrate remains constant (i.e., zero-order with respect to substrate concentration). This means that as the substrate concentration entering that slice of biofilm increases, the concentration leaving it also increases. In other words, the concentration gradient driving transport to and through the slice increases more slowly than the substrate concentration entering the slice increases. The net effect is that while the substrate flux into the biofilm increases with increased bulk substrate concentration, the effect is nonlinear, and is, in fact, half-order. This is also illustrated in Figure 17.3.

Figure 17.3 shows that the substrate flux for Monod kinetics results from a combination of the impacts of first-order and zero-order substrate removal kinetics. The substrate flux for Monod kinetics is always lower than the fluxes associated with either first-order or zero-order kinetics. First-order kinetics is applicable only for low bulk substrate concentrations (inset). For the parameter values assumed (Table 17.1), the Monod curve parallels the zero-order curve over most of the substrate concentration range. This suggests that for this particular case, substrate removal occurs mostly following zero-order kinetics.

To gain a better sense of how mass transfer processes affect the apparent kinetics of a biofilm system, let's reconsider Figure 17.2, which represents a steady-state biofilm following Monod kinetics. Because the biofilm resulting in Figure 17.2 is a steady-state biofilm, the biofilm thickness associated with any substrate flux (and its associated bulk substrate concentration) can be calculated with Equation 16.21. To see the impact of mass transfer on the performance of the biofilm, we can calculate the flux that would result if the entire biofilm thickness was exposed to substrate at the bulk substrate concentration. The substrate flux into such a hypothetical biofilm is

$$J_{Monod,intrinsic} = \hat{q}_H \frac{S_{Sb}}{K_S + S_{Sb}} X_{Hf} \cdot L_f. \quad (17.10)$$

The results of this exercise are plotted in Figure 17.4, along with the curve from Figure 17.2. (Note that Figure 17.4 uses a log scale on the ordinate, which makes the shapes of the curves appear different from Figure 17.2). The most striking observation from Figure 17.4 is that the flux associated with the hypothetical intrinsic biofilm is around 30 times larger than the flux in the mass transfer limited biofilm. To evaluate the contribution of external mass transfer resistance to this difference, the curve from Figure 17.2 was recalculated using a very high value for the external mass transfer coefficient, k_L, and the resulting curve is also shown in Figure 17.4. A comparison of the curves shows that while the effects associated with external mass transfer are significant, they cannot account for the large difference between the intrinsic curve and the mass transfer limited curve. This suggests that internal mass transfer resistance is responsible. The requirement for substrate to be transported by diffusion within the biofilm sets up a concentration gradient within the biofilm. For Monod kinetics, as the substrate concentration decreases within the biofilm, so

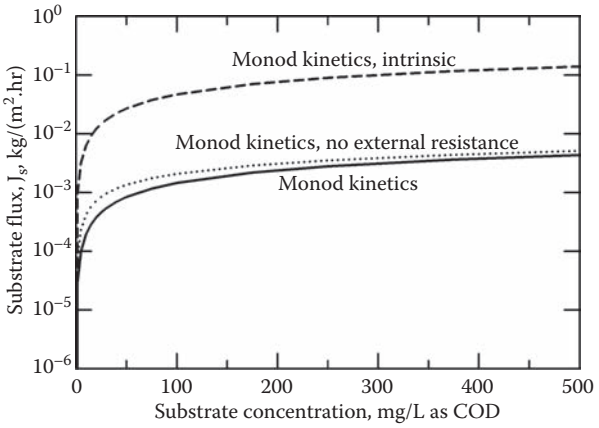

FIGURE 17.4 The effect of mass transfer on the substrate flux to a biofilm. The intrinsic curve was calculated with Equation 17.10. The values of the kinetic parameters and stoichiometric coefficients are given in Table 17.1.

will the substrate removal rate. Thus, most of the substrate removal within the biofilm occurs at substrate concentrations less than the bulk substrate concentration. Furthermore, in steady-state biofilms, which tend to be deep, only part of the biomass is exposed to substrate and is actively removing it. The remainder of the biomass is buried within the biofilm where the substrate concentration is equal to zero, and thus it is contributing nothing to the actual flux, even though it contributed to the intrinsic curve calculated with Equation 17.10. This leads to the question of how much of the difference between the intrinsic and the mass transfer limited curves can be attributed to this "inactive" biomass. Unfortunately, for Monod kinetics, there is no analytical solution that allows calculation of the fraction of the biomass that is actively involved in substrate removal in the biofilm. However, we saw in Figure 17.3 that for higher substrate concentrations, the plot for Monod kinetics paralleled the plot for zero-order kinetics, and for zero-order kinetics, such an expression is available:[12]

$$L_{f,zero,active} = \left[\frac{2D_e}{\hat{q}_H \cdot X_{Hf}} \left(S_{Sb} - \frac{J_{zero,deep}}{k_L} \right) \right]^{0.5}. \tag{17.11}$$

It can be used to approximate the active film thickness at higher substrate concentrations where Monod kinetics behaves in a zero-order manner. Application of Equation 17.11 reveals that for the parameter values used here, only 3% of the biofilm is active in that range. Thus, we can see that the major reason for the difference between the mass transfer limited Monod curve and the intrinsic Monod curve in Figure 17.4 is that only a fraction of the biofilm is active due to the need to transport substrate within it by diffusion. This means that maintaining a thick biofilm has no advantage; hence, many biofilm reactors are operated in such a way as to keep the biofilm thin. Inspection of Equation 17.11 tells us that the depth of active biofilm increases as the external mass transfer coefficient, k_L, is increased. This shows that the hydrodynamic regime imposed on the biofilm will influence its active thickness.

The dependence shown in Figure 17.2 holds true for all of the bioreactor simulations presented in this chapter for packed towers, since it is defined by the kinetic and stoichiometric coefficients associated with the growth of the microorganisms. The model used herein does not consider the dependence of the values of the external mass transfer coefficient (k_L) and the detachment coefficient (b_D) on the hydraulics of the system (this will be discussed in Section 17.1.6).

17.1.4 Performance of a Packed Tower without Flow Recirculation ($\alpha = 0$)

Simulations of tower performance were obtained using the parameter values in Table 17.1. These values were selected to be consistent with typical design parameters, such as the hydraulic and organic loading applied to the packed tower, as discussed in Chapter 19. The packed tower was simulated as a series of 25 CSTRs with a total volume, V_L, and total media surface area for biofilm growth, A_s, as indicated in Table 17.1.

17.1.4.1 Performance as a Function of Tower Depth

Typical profiles of substrate concentration, biofilm thickness, and suspended biomass concentration as a function of tower depth are shown in Figure 17.5 for a tower without recirculation. As indicated in Table 17.1, the total tower depth is 5 m.

The substrate concentration decreases in a nearly exponential manner with depth. At the top of the tower, the substrate concentration is the highest, resulting in the highest substrate flux into the biofilm, as indicated by the greater slope of the substrate concentration curve. This dependence of flux on concentration is consistent with Figure 17.2. As the wastewater moves down the tower, the substrate concentration decreases, resulting in lower substrate fluxes into the biofilm. There are two reasons for this. First, the concentration gradients driving substrate transport across the stagnant boundary layer and diffusion throughout the biofilm are lower. Second, lower substrate concentrations result in lower specific growth rates for the microorganisms, hence lower substrate utilization rates. The observed exponential decrease in substrate concentration with depth is a general characteristic of packed towers.

As seen in Figure 17.5, the biofilm thickness decreases as a function of the depth in the tower. Furthermore, the biofilm is deep ($\xi > 0.99$) throughout the entire tower. This means that the substrate concentration within the biofilm goes to zero before the solid support is reached. The main implication of having a deep biofilm is that the substrate flux is independent of the biofilm thickness. The maintenance of a deep biofilm—even though its thickness decreases—illustrates an important point. Whether a biofilm is deep is not determined by its actual physical thickness. Rather, it is determined by the penetration of substrate into it. Even the thinner biofilm at the bottom of a tower can be deep when the substrate concentration is low, causing its exhaustion within the biofilm. The profiles depicted in Figure 17.5 are unique to the parameter and operational values used in the simulations, as given in Table 17.1. Not all towers can be assumed to have deep biofilms. However, the decrease in biofilm thickness with depth is a general occurrence because it follows from the balance between growth and loss associated with a steady-state biofilm. The rate of biomass growth decreases with depth in the tower due to the lower substrate concentration, which causes a lower substrate flux. However, the rate at which biomass is lost from the biofilm is constant over the entire tower depth because detachment is governed by the hydrodynamic conditions. The combined effect of these two processes results in the observed decrease of biofilm thickness with depth.

The suspended biomass concentration increases as the flow progresses through the tower for two reasons: biomass detachment from the biofilm and growth of suspended biomass through substrate utilization in the liquid phase. The contribution of suspended growth to the total substrate removal in a packed tower is low, as shown in Figure 17.5b. Suspended growth only becomes important when very small effluent substrate concentrations must be achieved.

17.1.4.2 Effect of Biofilm Surface Area on Tower Performance

A key parameter in the design of biofilm reactors is the total media surface area available for growth of biomass. This is because substrate removal is directly linked to the substrate flux, which is the rate at which substrate is transported per unit area. The effect of media surface area on the effluent substrate concentration from a packed tower of fixed depth is shown in Figure 17.6, which was generated with the information in Table 17.1. When the media surface area is equal to zero, the effluent substrate concentration is equal to the influent substrate concentration because the biomass

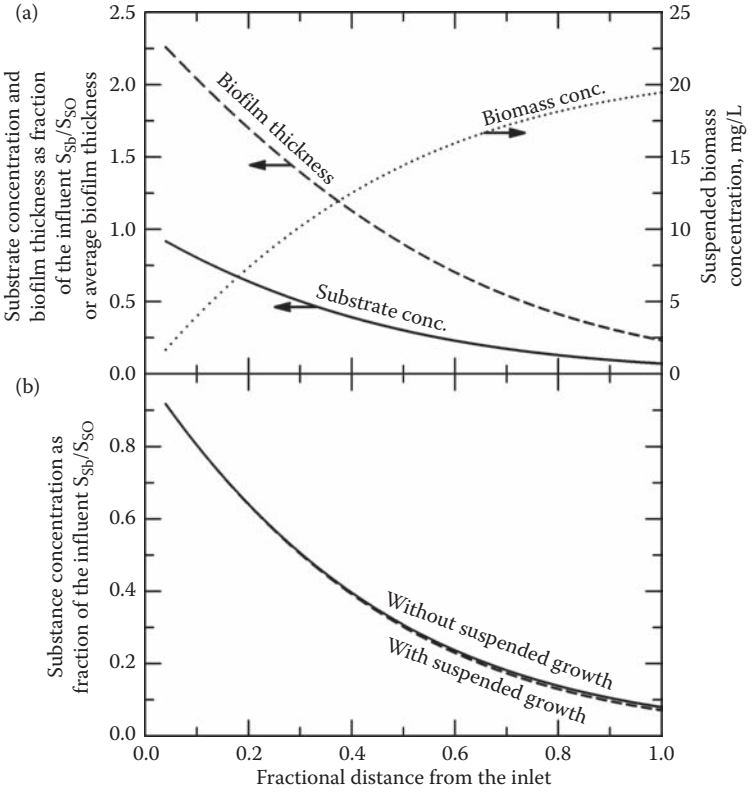

FIGURE 17.5 Panel a: effect of reactor depth on the substrate concentration, biofilm thickness, and suspended biomass concentration in a packed tower without recirculation. Panel b: the solid curve represents the values obtained by setting the maximum specific growth rate of the suspended biomass to zero, thereby eliminating their activity. The dashed curve is the same as the substrate concentration curve in Panel a. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.1.

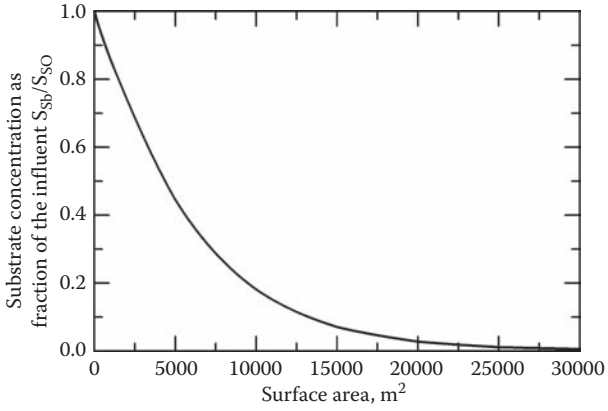

FIGURE 17.6 The effect of the media surface area available for biofilm growth on the effluent substrate concentration from a packed tower. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.1, except for the media surface area, A_s, which was varied as shown in the figure.

has nowhere to grow. As the surface area is increased, more microbial growth can occur, allowing a greater mass of biomass to be present in the reactor. The greater mass of biomass results in a lower specific substrate removal rate, which is equivalent to a lower substrate flux into the biofilm. Lower substrate fluxes are required to obtain lower substrate concentrations in attached growth systems, as can be seen from an examination of Figure 17.2.

There is a minimum bulk substrate concentration that can be achieved in an attached growth bioreactor, and it is S_{Sbmin}:

$$S_{Sbmin} = \frac{K_S(b_{Hf} + b_D)}{Y_H \cdot \hat{q}_H - (b_{Hf} + b_D)}. \tag{16.22}$$

As can be seen in Equation 16.22, S_{Sbmin} is independent of the media surface area for growth. It only depends on the kinetic parameters for microbial growth and the rates at which biomass is lost from the biofilm due to detachment (b_D) and decay (b_H). Once S_{Sbmin} has been reached, providing additional media has no impact because the bulk substrate concentration is too low to allow biomass to grow on the media fast enough to replace that lost by detachment and decay. Thus, even though the curve in Figure 17.6 appears to approach zero as the media surface area is made very large, in reality it approaches S_{Sbmin}, which for the parameters in Table 17.1 is 0.325 mg COD/L. Because the influent substrate concentration is 100 mg COD/L, the lower limit in Figure 17.6 is 0.00325, a value too small to be visible on the scale used.

17.1.4.3 Effect of Influent Substrate Concentration on Tower Performance

For suspended growth in a CSTR at a fixed SRT, the steady-state effluent substrate concentration is independent of the influent substrate concentration because the specific growth rate is fixed. An increase in influent substrate concentration is simply offset by a proportional increase in biomass present. This is not the case for an attached growth bioreactor. Rather, for a fixed media surface area, the area of biomass that can be present is fixed. Consequently, if the influent substrate concentration is increased, the only way for more substrate to be removed is for the flux to increase. An increase in flux, in turn, requires an increase in the bulk substrate concentration. Thus, we would expect the effluent substrate concentration to increase as the influent substrate concentration is increased and this is exactly what happens, as illustrated in Figure 17.7.

Figure 17.7 presents the results of simulations for a packed tower with the characteristics in Table 17.1. Focusing first on the highest influent substrate concentration (1500 mg COD/L), we

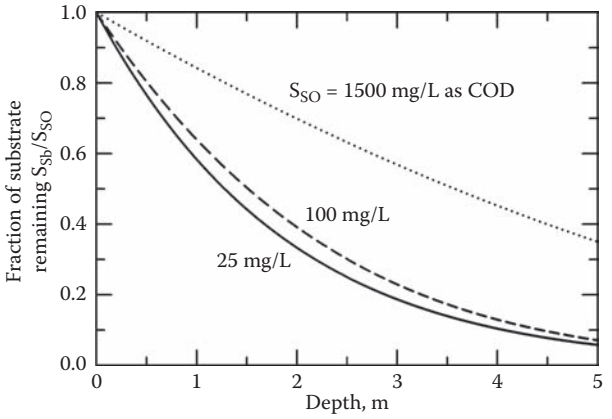

FIGURE 17.7 Substrate removal profiles through a packed tower without recirculation for three different influent substrate concentrations. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.1 unless otherwise specified.

see that the substrate concentration decreases almost linearly with depth. As seen in Figure 17.2, when the bulk substrate concentration is high, the substrate flux does not change much as the bulk substrate concentration is changed. The main reason for this is that at high substrate concentrations Monod kinetics behaves as a zero-order process, making the flux of apparent half-order, as discussed in Section 17.1.3. For influents with lower substrate concentration, the behavior is quite different. The concentration profiles look very similar when the influent concentration is either 25 or 100 mg COD/L. In both cases the substrate concentration drops in an exponential manner as a function of the tower depth, following an essentially first-order behavior. In fact, the tower achieves almost the same percentage of removal for the two influent substrate concentrations, which means that the one with the higher concentration will have a proportionally higher effluent substrate concentration. As seen in Figure 17.3 (inset), at lower substrate concentrations the flux is almost a linear function of the bulk substrate concentration. This follows from the fact that Monod kinetics approaches first-order behavior at low substrate concentrations. Since mass transfer is also a first-order process, the overall performance of the tower is first-order, as discussed in Section 17.1.3.

The effects of influent substrate concentration can be seen more directly in Figure 17.8, where the fractional substrate removal and the mass rate of substrate removal are shown as functions of the influent substrate concentration. At very low influent substrate concentrations the fractional substrate removal is essentially constant. This is the range over which the tower behaves in a first-order manner. For higher influent concentrations, however, the fractional substrate removal decreases as the influent substrate concentration is increased. This means that the tower behaves in a less than first-order manner. As the mass rate of substrate addition increases, higher substrate fluxes must be achieved. As seen in Figure 17.2, however, the relationship between substrate flux and bulk substrate concentration is nonlinear at higher substrate concentrations, meaning that the bulk substrate concentration must be more than doubled to achieve a doubling of the flux. Nevertheless, even though the fractional substrate removal decreases, the mass rate of substrate removal still increases, due to the higher fluxes that can be maintained by the higher concentrations.

It should be noted that the curves for the highest feed substrate concentration do not tell the complete story. This is because the model considers only a single limiting nutrient and does not consider the possibility of oxygen limitations. In reality, the application of such high substrate concentrations would result in oxygen limitations in the upper reaches of the tower, causing poorer performance than the model indicates.[17,22] In that case, substrate removal would be controlled by the rate of oxygen transfer, causing the same mass of substrate to be removed in each successive section of tower depth. This would cause the concentration to decrease linearly with depth with a smaller slope than shown in the figure.

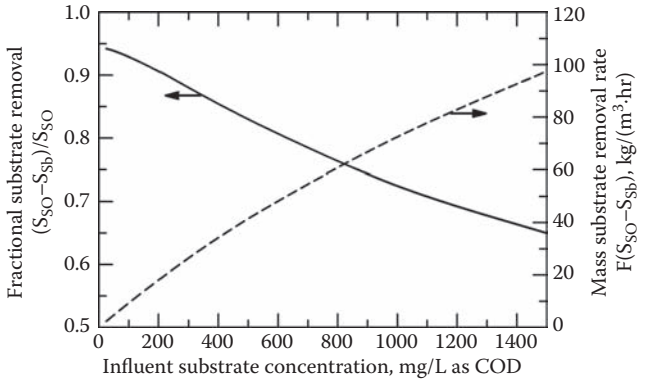

FIGURE 17.8 Effect of influent substrate concentration on the fraction of the influent substrate removed and the mass removal rate of substrate in a packed tower without recirculation. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.1 unless otherwise specified.

17.1.4.4 Effect of Influent Flow Rate on Tower Performance

The effects of influent flow rate are shown in Figure 17.9. As the influent flow rate is increased for a tower of fixed cross-sectional area, the superficial velocity of flow through the tower, which is the applied flow rate, $F(1 + \alpha)$, divided by the cross-sectional area, A_c, is increased. As introduced in Section 14.2.3, this parameter is referred to as the total hydraulic loading, with the acronym THL and the symbol Λ_H. In the general case, it is calculated by

$$\Lambda_H = \frac{F(1+\alpha)}{A_c}. \tag{17.12}$$

Because the THL is the flow parameter of primary importance, Figure 17.9 is presented in terms of it.

The THL is an important design parameter for packed towers. A minimum value must be maintained to keep all of the media wet. Above that minimum THL, as the flow rate is increased the fraction of the applied substrate removed decreases, although the substrate mass removal rate increases, as shown in Figure 17.9. Comparison of Figure 17.9 to Figure 17.8 reveals a distinct similarity in the plots. This is because both represent the response of a packed tower to an increase in loading (i.e., the mass application rate of substrate). In Figure 17.8 that increase was achieved by increasing the influent substrate concentration at fixed flow rate, whereas in Figure 17.9 it was achieved by increasing the applied flow rate at fixed influent substrate concentration. For low flow rates, the packed tower is underloaded, leading to almost complete substrate consumption within the tower. As the flow rate is increased, so is the mass of substrate applied to the tower per unit time. In order to achieve a higher removal rate in response to this increased substrate application rate, the substrate flux into the biofilm at any particular depth must increase and this is achieved by an increase in the bulk substrate concentration at that particular depth. This dependence is not linear because of the nature of Monod kinetics, as shown in Figures 17.2 and 17.3. For low flow rates, the rate of substrate addition to the tower is low, and steady state results in small substrate fluxes across the biofilm. For this condition, low bulk substrate concentrations will result and the substrate flux will change in a linear (first-order) fashion with the substrate concentration. This explains why, initially, the mass rate of substrate removal increases linearly with the THL. As the THL is increased further, causing the substrate concentration in the bulk liquid to increase, Monod kinetics approach zero-order behavior. After that occurs, increases in the substrate concentration do not result in a proportional increase in the specific substrate removal rate and the substrate flux. Consequently, the mass rate

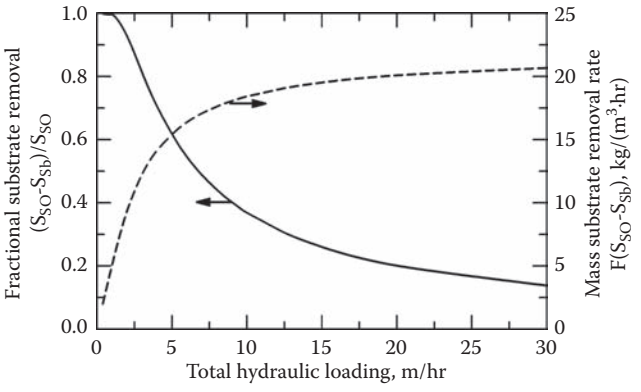

FIGURE 17.9 Effect of the total hydraulic loading on the fraction of the influent substrate removed and the mass removal rate of substrate in a packed tower without recirculation. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.1 unless otherwise specified (flow rates ranged from 20 to 1500 m^3/hr for a fixed cross-sectional area of 25 m^2).

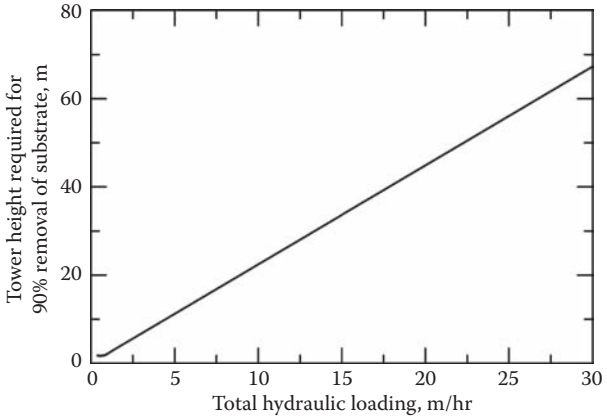

FIGURE 17.10 Effect of the total hydraulic loading on the height of packed tower required to achieve 90% reduction in substrate concentration in the absence of recirculation. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.1 unless otherwise specified (flow rates ranged from 20 to 1500 m^3/hr for a fixed cross-sectional area of 25 m^2).

of substrate removal approaches a maximum value at high THLs. Because of these effects, greater flow rates and their associated THLs require greater tower depths to achieve a fixed effluent concentration, as shown in Figure 17.10.

17.1.5 Performance of a Packed Tower with Flow Recirculation

Recirculation of clarified effluent has a complicated effect on tower performance. First, it reduces the applied substrate concentration by dilution of the feed with the treated effluent, as indicated by Equation 17.1 and illustrated in Figure 17.11a. It also results in flatter substrate concentration profiles (also shown in Figure 17.11a) because a plug-flow reactor behaves more like a CSTR as the recirculation ratio is increased. This latter point is illustrated in Figure 17.12 and follows from the reduction in reaction rate associated with lower substrate concentrations. Recirculation also acts to provide a more uniform biofilm thickness throughout a tower,[31] as shown in Figure 17.11b. Although these findings are consistent with those from other models with different assumptions,[23,31] no generalizations should be made about the magnitude of the effect of recirculation because it depends on the feed flow rate to the tower as well as the mass transfer characteristics of the media. Thus, while recirculation will generally reduce the fractional removal of substrate across a tower, the degree of reduction will be system specific.

Although the model results discussed above show that recirculation of clarified effluent will decrease substrate removal, circumstances exist in which recirculation could increase it. For example, if the feed substrate concentration was so high that oxygen transfer limited substrate removal, recirculation could decrease the problem by reducing the reaction rate and increasing the oxygen transfer rate. Furthermore, the presence of biomass in the recirculated flow can have an impact. The results in Figures 17.11 and 17.12 were obtained by assuming that the settler was perfect so that no biomass was present in the recirculation flow. It is possible, however, that if biomass had been present, the reaction term for substrate removal by suspended organisms would have been large enough to make the effluent substrate concentration lower than it was without recirculation.[22] Thus, while it is true that recirculation generally reduces substrate removal through packed towers, one must not conclude that the effects of recirculation are always negative. Rather, each situation must be evaluated.

During the design of a packed tower for a given feed flow rate, an engineer may choose any cross-sectional area that gives a THL that is acceptable for the media under consideration. Each

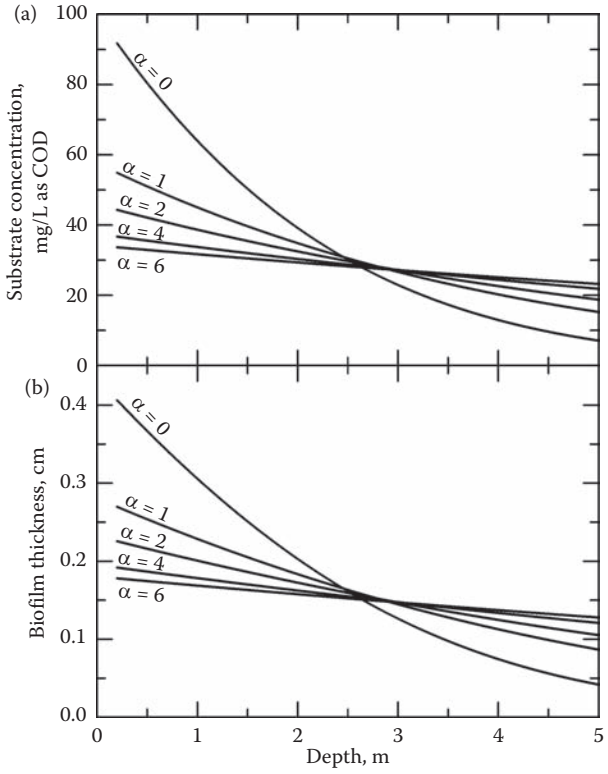

FIGURE 17.11 Effect of recirculation of effluent from a perfect clarifier on the performance of a packed tower of fixed size: (a) effects on substrate concentration profiles, and (b) effects on biofilm thickness. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.1 unless otherwise specified.

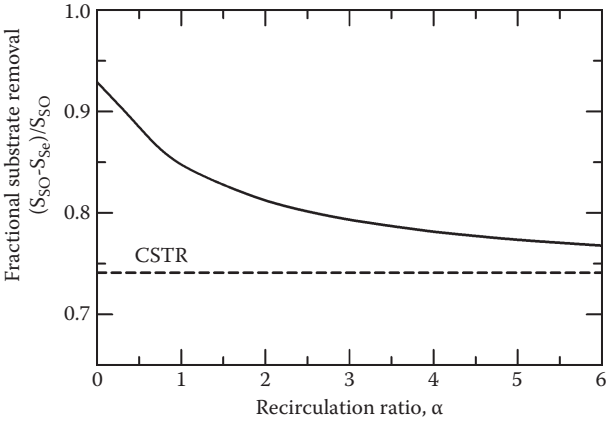

FIGURE 17.12 Effect of recirculation of effluent from a perfect clarifier on the fraction of substrate removed by a packed tower of fixed size. For comparison, the fraction of substrate removed in a single CSTR with the same volume and media surface area is shown. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.1 unless otherwise specified.

type of media has a minimum THL that is required to give a uniform flow distribution across the entire tower cross section and keep all of the media wet. This establishes a lower limit on the THL and a maximum limit on the cross-sectional area to be used. One question that arises during design is whether it is worthwhile to make the cross-sectional area smaller than this value, resulting in a higher THL. To answer that question we need to consider two things.

A tall, narrow tower has a flow pattern that is more likely to conform to plug flow. A short, wide tower deviates more from this pattern and approaches that of a CSTR. The results in Figure 17.12 suggest that the total media volume will be minimized by choosing a tall, narrow tower (PFR) rather than a short, wide one (CSTR). Because a similar conclusion has been reached with other models,[18,23,31] as well as experimentally,[27,34] it appears to be general and thus would be expected to be true for other parameter values as well. It should be recognized, however, that the decrease in tower volume associated with increased tower height will be case specific and may not be significant in some situations.[23]

17.1.6 Factors Not Considered in Model

For a fixed flow rate, changing the cross-sectional area of a packed tower will affect the velocity at which the wastewater travels through it, with smaller areas resulting in higher fluid velocities. The fluid velocity will affect both the external mass transfer coefficient (k_L) and the detachment coefficient (b_D). Both of these parameters will increase if the THL is increased, but they affect the performance of the packed tower in opposite directions. These effects were not included in the model used herein, although they are important. This was done to simplify the structure of the model and to allow us to examine each factor one at a time. There are, however, approaches available in the literature that can be used for including such effects in the model.

17.1.6.1 External Mass Transfer

The flow patterns within a packed tower are very complex, reflecting interactions among fluid elements flowing over different support surfaces, variations in the cross-sectional area available for flow in random packing, irregularities caused by channeling, and short circuiting due to droplets falling from protrusions in the biofilm.[15] Because of these complex flow patterns it has been necessary to develop empirical correlations for k_L within such towers.[26,32] One common form of correlation relates the mass transfer coefficient to the Schmidt number ($Sc = \mu_w/\rho_w D_w$) and the Reynolds number ($Re = v\rho_w d/\mu_w$), where μ_w is the fluid viscosity, ρ_w is its density, v is its bulk fluid velocity past the biofilm, D_w is the diffusivity of the substrate in water, and d is a dimension characterizing the media. Recognizing that the product of the velocity and the density of a fluid is the mass velocity, M, the Reynolds number is often written as Md/μ_w. Using this concept, Wilson and Geankoplis[40] reported the following correlations for mass transfer to liquids in packed beds:

$$k_L = (0.25\ M/\varepsilon\rho_w)Sc^{-0.67}Re^{-0.31} \qquad 55 < Re < 1500 \tag{17.13}$$

and

$$k_L = (1.09\ M/\varepsilon\rho_w)Sc^{-0.67}Re^{-0.67} \qquad 0.0016 < Re < 55. \tag{17.14}$$

These equations are restricted to $0.35 < \varepsilon < 0.75$, where ε is the void space between media elements as a fraction of the total bed volume (i.e., the porosity). Because the Reynolds number is directly proportional to M, Equations 17.13 and 17.14 predict that the mass transfer coefficient increases with $M^{0.69}$, when M is the superficial mass velocity based on the entire bed cross-sectional area normal to the direction of flow. Unfortunately, most plastic media used in packed towers for wastewater treatment have void fractions on the order of 0.95 and thus similar correlations are needed for predicting

k_L in such systems. Dimensional analysis, as well as the above equations, suggests the use of a correlation of the form:

$$k_L = aM^b, \tag{17.15}$$

where a depends on the properties of the fluid and the media while b depends on the flow range employed. The general utility of such correlations is an open question because mass transfer in biological packed towers is influenced strongly by the presence and characteristics of the biofilm,[15,17,22] which will depend to a large degree on the nature and concentrations of the substrates being degraded. Consequently, mass transfer coefficients measured with clean media do not accurately reflect mass transfer in towers containing biofilms. Because the continued refinement of mechanistically based models for attached growth systems requires that accurate mass transfer relationships be available, studies on mass transfer in the presence of biofilm must be done, although they are not likely to be easy to perform. In the mean time, correlations such as those in Equations 17.13 and 17.14 provide a general idea of how external mass transfer coefficients may change when bulk fluid velocities (or superficial mass velocities) are changed.

17.1.6.2 Biomass Detachment

The mechanisms of biofilm detachment are very complex and not fully understood. Erosion plays a part in it[30] and is influenced by at least three factors: high shear on those parts of the biofilm that extend outward from the surface, the nutrient concentration surrounding the biofilm, and loss of cells from the surface of cell clusters in the biofilm.[1] Detachment promoting agents that "dissolve" the extracellular polymeric substances holding the biofilm together also play a role,[42] as does the thickness of the biofilm.[37] Attempts have been made to predict detachment by computationally calculating the hydrodynamic conditions surrounding the biofilm, treating the biofilm as an elastic solid, and quantifying the stress caused by flow through the biofilm.[29] Attempts have also been made to incorporate the other factors into models for detachment.[1,37,42]

Values for the detachment coefficient are system dependent and should be determined by performing a mass balance around a biofilm-containing bioreactor in which suspended growth does not occur to a significant extent. This balance would take the form:

$$F_a X_{total,O} + b_D X_{B,Hf} A_s L_f = F_a X_{total,e}. \tag{17.16}$$

The variables $X_{total,O}$ and $X_{total,e}$ are the influent and effluent total suspended solids concentrations, respectively. They need to be measured, along with the biofilm thickness, to allow determination of the detachment coefficient. The IWA task group on biofilm modeling used a base value of 0.4 d^{-1} for the detachment coefficient in their simpler models,[38] but values as high as 7.43 d^{-1} have been reported for fluidized bed bioreactors.[2] For the conditions employed in the simulations presented in this chapter, the value of the detachment coefficient had little impact on the results.

17.1.6.3 Other Factors Not Considered

The model presented in Section 17.1.2 assumed that a packed tower acts as a plug-flow reactor with no intermixing of fluid elements as the wastewater undergoing treatment flows downward through it. This is a simplification. In reality, two factors act to cause significant degrees of intermixing, the nature of the biofilm and the type of media used for packing.

From observations of bundle media with a clear plastic front section, Hinton and Stensel[15] concluded that while sheet flow is the predominant type in that media, droplet flow is also important and leads to significant fluid intermixing. Droplet flow results from protrusions in the biofilm. Although the cause of protrusions is unknown, they act as sites of droplet formation, allowing droplets to

bypass portions of the media before intersecting it and the sheet flow again. The points where the droplets hit are highly turbulent, resulting in significant intermixing. Thus, while the majority of the flow moves without intermixing, this behavior is frequently interrupted by fluid that has bypassed treatment, causing fluid with a higher substrate concentration to be mixed back in. The net result is less substrate removal than predicted by models like the one used herein.

Logan et al.[24] have attempted to incorporate the effects of media geometry into a model for packed towers. They maintain that different types of media exhibit different flow patterns and that the type of media will influence the degree of intermixing that occurs. Some types of media will establish long flow paths over which the fluid flows in laminar flow. This allows the concentration gradient depicted in Figure 16.4 to be established and maintained over long distances. In contrast, other media have intersections of flow paths. Mixing occurs at those intersections, destroying the concentration gradient and establishing a new average concentration from which a new gradient is established. That gradient will then be maintained until the next mixing point. Modeling results have shown that these effects can cause differences in performance, with the media exhibiting intermixing giving better substrate removal.[24] Nevertheless, it should be pointed out that the improvement shown by intermixing is uncertain, particularly in light of the effects discussed in the preceding paragraph, which are likely to be highly stochastic. Thus, it is apparent that much more study of packed tower hydraulics is required before accurate mechanistic models can be written.

Another factor not considered in the model is oxygen transfer and the potential for oxygen limitations in the biofilm. The implications of this were discussed earlier, but it would be worthwhile considering oxygen transfer further here. Oxygen transfer in biofilm systems cannot be studied accurately in the absence of biological activity and this has hampered development of oxygen transfer relationships.[17,22] There are two reasons for this. First, the biofilm itself influences the fluid flow patterns over the media, thereby preventing studies done on clean media from being representative of media containing biofilms.[15] Second, because the biological activity maintains a low oxygen concentration in the bulk liquid, oxygen transfer continues in situations where the water flowing over clean media would have become saturated. Furthermore, because of the stochastic nature of fluid flow over packed tower media, 10 to 25% of the total biofilm surface area can be devoid of a covering bulk liquid film at any instant.[15] This suggests that the oxygen concentration gradient into the biofilm can change drastically in a random manner, making models that assume a fixed gradient inaccurate. In spite of this complex situation, both theoretical[22] and experimental[17] studies have reached similar conclusions about the maximum oxygen transfer rate for bundle media, which is approximately 650 mg O_2/(m^2·hr). Translation of this into a maximum substrate removal rate without oxygen limitations depends on the true growth yield (see Equation 3.34), the kinetics of biodegradation, and the relative diffusivities of oxygen and the substrate, but it seems unlikely that removal rates in excess of 1400 mg COD/(m^2·hr) could be handled without oxygen limitation and the potential problems associated with it. This suggests that at typical THLs, the maximum applied substrate concentration that would avoid oxygen transfer limitations is around 250 mg COD/L.[23]

Finally, the model was limited to one type of bacteria growing on a single limiting electron donor. However, as we have seen throughout this book, both heterotrophic and autotrophic bacteria will grow simultaneously in the same culture when possible, each with its own electron donor, but competing for the electron acceptor. In addition, in packed towers, as in any biofilm process, the two types of bacteria must compete for space, as discussed in Section 16.4. This competition has a significant effect on packed towers and causes carbon oxidation to occur in the upper portion of the tower, followed by nitrification in the lower portion. Reexamination of Figure 16.20 makes the reason for this behavior clear. For most wastewaters, the relative concentrations of COD and ammonia-N in the feed to a packed tower lie in the region marked H in that figure. Consequently, the heterotrophs displace the autotrophs in the upper portions of the tower and only COD removal will occur. As the fluid passes downward through the tower, the COD concentration is reduced

relative to the ammonia-N concentration, resulting in a zone where both types of bacteria can exist, allowing simultaneous carbon oxidation and nitrification. Finally, further down the tower the concentrations of the two constituents will lie in region A of Figure 16.20 and nitrification will be the predominate reaction. While this sequential pattern is a common observation in practice, its existence can be influenced somewhat by recirculation, which reduces the concentration gradients in the tower, allowing simultaneous carbon oxidation and nitrification to occur over a greater length of the tower in some situations.

17.1.7 Other Packed Tower Models

It is apparent by now that a packed tower is a very involved biochemical operation that is difficult to model theoretically. Consequently, empirical and semiempirical approaches have been used to model it and a number of design equations have been developed. The National Research Council[27] (NRC) and Gallar-Gotaas[6] equations are strictly empirical because they are based only on the application of regression analysis techniques to experimental observations. Consequently, they are valid only in the limited region over which the original data were available and can only be used for interpolation within that region. They cannot be used for extrapolation. Semiempirical models are also primarily interpolative, but can be successfully employed for extrapolation if the extrapolations do not extend them beyond the range over which their simplifying assumptions are valid. Finally, a truly mechanistic model can be used for extrapolation to new conditions, but care must still be exercised to be sure that extrapolations do not violate the basic assumptions of the model. In order to recognize the assumptions inherent in them we shall review some semiempirical and mechanistic models.

17.1.7.1 Grady and Lim Model

The model of Grady and Lim[7] considers the growth of a single organism in the biofilm with no limitation on the amount of terminal electron acceptors available to sustain growth, as was the case with the model described in Section 17.1.2. The bioreactor was assumed to be a perfect PFR and attached growth was described using the effectiveness factor approach (Section 16.2.2). The mass of biomass (suspended and attached) was assumed to be constant and independent of the mass of substrate added to the system, with a constant biofilm thickness across the entire tower depth. Under the assumption of perfect plug flow, a steady-state mass balance on substrate can be written over an infinitesimal volume element of a packed tower similar to the one shown in Figure 17.1. Application of the usual limiting process and use of the overall effectiveness factor (η_{eO}) as presented in Section 16.2.2 to account for simultaneous transport and reaction leads to

$$\frac{dS_{Sb}}{dx} + \eta_{eO}\left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}}\right)\left(\frac{a_s A_c}{F(1+\alpha)}\right) X_{B,Hf} \cdot L_f + \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}}\right)\left(\frac{f_{Ac} A_c}{F(1+\alpha)}\right) Y_{Hobs}(S_{Sa} - S_{Sb}) = 0, \tag{17.17}$$

$$S_{Sb} = S_{Sa} \text{ at } x = 0$$

where a_s is the specific surface area (biofilm surface area per unit volume of tower), and Y_{Hobs} is an observed yield as used in conjunction with Equation 16.19. The second term in the equation represents substrate removal by attached biomass whereas the third term is the removal by suspended biomass. The performance of an ideal packed tower was simulated by numerically integrating Equation 17.17 using the overall effectiveness factor given by Figure 16.9 or an empirical correlation as discussed in Section 16.2.2. Grady and Lim[7] explicitly considered the effect of the bulk fluid velocity (superficial mass velocity) on the external mass transfer coefficient, through the

use of correlations such as those given by Equations 17.13 and 17.14. For their particular system the external mass transfer coefficient was calculated using an expression of the type:

$$k_L = E\left[\frac{F(1+\alpha)}{P \cdot A_c}\right]^{0.69}, \tag{17.18}$$

where E and P are system dependent parameters. Their model gave results that are qualitatively similar to those from the model described in Section 17.1.2.

17.1.7.2 Velz Model

The model of Velz[36] states that the substrate concentration decreases in an exponential manner with depth, which is a characteristic we observed in Figure 17.7:

$$\frac{S_{Se}}{S_{Sa}} = \exp(-KL), \tag{17.19}$$

where K is a constant. Comparison of Equation 17.19 with Equation 17.17 shows that the Velz equation is limited to the situation in which the removal of substrate by suspended microorganisms is negligible, the substrate consumption rate is first-order with respect to the substrate concentration ($S_{Sb} \ll K_S$), and the overall effectiveness factor is constant throughout the tower. Under these circumstances, Equation 17.17 can be integrated to yield:

$$\frac{S_{Se}}{S_{Sa}} = \exp\left\{-\left[\frac{(\hat{q}_H/K_S)a_s A_c \eta_{eO} X_{B,Hf} L_f}{F(1+\alpha)}\right]L\right\}, \tag{17.20}$$

from which we see that

$$K = \frac{(\hat{q}_H/K_S)a_s A_c \eta_{eO} X_{B,Hf} L_f}{F(1+\alpha)} = \frac{\hat{q}_H a_s \eta_{eO} X_{B,Hf} L_f}{K_S \Lambda_H}. \tag{17.21}$$

Thus, the Velz equation is limited to one THL.

17.1.7.3 Eckenfelder Model

Eckenfelder[3] has explicitly accounted for the effects of flow rate:

$$\frac{S_{Se}}{S_{Sa}} = \exp\left\{-K_1 a_s^{1+m}\left[\frac{A_c}{F(1+\alpha)}\right]^n L\right\}, \tag{17.22}$$

where m and n are parameters that are dependent on the media used and K_1 is a rate coefficient. This equation is a modification of the Velz equation in which the effluent concentration is allowed to depend explicitly on F, A_c, α, and a_s. Therefore, we may look upon it as a limiting case of Equation 17.17 in which the removal of substrate by suspended microorganisms is negligible, the substrate consumption rate is first-order with respect to substrate ($S_{Sb} \ll K_S$), and the overall effectiveness factor remains constant throughout the tower but varies with the THL, $F(1 + \alpha)/A_c$. The last point becomes clear if we rearrange Equation 17.22:

$$\frac{S_{Se}}{S_{Sa}} = \exp\left\{-K_1 a_s^{1+m}\left[\frac{A_c}{F(1+\alpha)}\right]\left[\frac{A_c}{F(1+\alpha)}\right]^{n-1} L\right\}, \tag{17.23}$$

and compare it to Equation 17.20. Thus, if η_{eO} is proportional to $[A_c/F(1 + \alpha)]^{n-1}$ they are equivalent for a given media. According to Liptak,[21] most media have an n value of 0.7 to 0.8. Consequently, a 28-fold increase in THL would result in a 1.95 to 2.72-fold increase in $[A_c/F(1 + \alpha)]^{n-1}$. Likewise, for the values of the Biot number, Bi, the Thiele modulus, ϕ^2, and the modified Thiele modulus, ϕ_f, likely to be found in a packed tower, a 28-fold increase in THL will increase η_{eO} by a factor of 2.5. Thus, it appears that the changes in $[A_c/F(1 + \alpha)]^{n-1}$ in the Eckenfelder model are similar to the changes in the overall effectiveness factor caused by flow as presented by Grady and Lim.[7]

17.1.7.4 Kornegay Model

The differential equation used by Kornegay[18] is

$$\frac{dS_{Sb}}{dx} = -K_2\left[\frac{A_c}{F(1+\alpha)}\right]\left(\frac{S_{Sb}}{K_g + S_{Sb}}\right), \qquad S_{Sb} = S_{Sa} \text{ at } x = 0 \tag{17.24}$$

where K_2 depends on the substrate and type of media, and K_g is a pseudo half-saturation coefficient that decreases asymptotically to K_S as the THL becomes very large. Equation 17.24 may be rewritten as

$$\frac{dS_{Sb}}{dx} = -K_2\left[\frac{A_c}{F(1+\alpha)}\right]\left(\frac{S_{Sb}}{K_S + S_{Sb}}\right)\left(\frac{K_S + S_{Sb}}{K_g + S_{Sb}}\right), \tag{17.25}$$

where K_S is the intrinsic half-saturation coefficient that remains invariant for a particular microorganism and substrate. If there is no substrate removal by suspended microorganisms, Equation 17.17 reduces to

$$\frac{dS_{Sb}}{dx} = -K_3\left[\frac{A_c}{F(1+\alpha)}\right]\left(\frac{S_{Sb}}{K_S + S_{Sb}}\right)\eta_{eO}, \qquad S_{Sb} = S_{Sa} \text{ at } x = 0 \tag{17.26}$$

in which K_3 depends on the substrate and the type of media. Comparison of Equation 17.25 with Equation 17.26 reveals that they are equivalent for a given media if

$$\eta_{eO} = \frac{K_S + S_{Sb}}{K_g + S_{Sb}}. \tag{17.27}$$

Let us examine the requirement on η_{eO} in more detail. Inspection of Equation 17.27 reveals that the overall effectiveness factor: approaches unity as the substrate concentration becomes very large; approaches unity when the THL becomes very large so that K_g approaches K_S; decreases as the substrate concentration decreases, approaching an asymptotic value of K_S/K_g as the substrate concentration approaches zero; and decreases as the THL is decreased, causing K_g to increase. Thus, it appears that the Kornegay model can be interpreted as a limiting case of Equation 17.17 in which substrate removal by suspended microorganisms is negligible and the overall effectiveness factor is given by Equation 17.27.

17.1.7.5 Schroeder Model

Schroeder's[33] model states that

$$\frac{dS_{Sb}}{dx} = -\left(\frac{\hat{q}_H S_{Sb}}{K_S + S_{Sb}}\right)\left[\frac{a_s A_c X_{B,Hf} L_f}{F(1+\alpha)}\right]\eta_{eI}, \qquad (17.28)$$

$$S_{Sb} = S_{Sa} \text{ at } x = 0$$

where η_{eI} is the internal effectiveness factor. Comparison of Equation 17.28 with Equation 17.17 reveals that the Schroeder equation is a limiting case in which both the removal of substrate by suspended microorganisms and the external mass transfer resistance are assumed to be negligible. The assumption of negligible external mass transfer resistance may only be valid when the THL is very high. Furthermore, Schroeder assumed that the internal effectiveness factor is directly proportional to the bulk substrate concentration, but simulations done using the Grady and Lim[7] model suggest that this assumption is a poor one.

17.1.7.6 Logan, Hermanowicz, and Parker Model

Logan et al.[24] developed a model that attempts to account for the hydrodynamic properties of different packed tower media. The model assumes that the fluid flows in thin films that are generally laminar, allowing a parabolic velocity profile to be established. Transport of a single limiting substrate is assumed to occur across that liquid film in the absence of reaction in the film; that is, there is no substrate removal by suspended microorganisms. The flux into the biofilm is given by an empirical expression that assumes first-order kinetics and Brownian collisions between the substrate molecules and the cells.[24] Although the rationale for the flux equation is not entirely clear, the model essentially assumes that substrate removal is limited by transport to the biofilm surface and not by the biofilm kinetics.[25] In other words, only external mass transport is assumed to be limiting, which is consistent with the experimental observations of others.[25] The unique feature of this model is that it considers the hydraulic characteristics of the media. In some types of media, the fluid is assumed to flow in thin films throughout the tower depth, allowing stable velocity and concentration profiles to be established. In others, however, intermixing is assumed to occur at regular intervals because of the geometry of the media, disrupting the velocity and concentration profiles at that point, thereby requiring them to be reestablished from the average concentration after intermixing. Although simultaneous transport of substrate and oxygen is not considered and only a single substrate is assumed to limit reaction, the capability is provided to allow computation of the maximum possible oxygen transport rate in the system.

17.1.7.7 Hinton and Stensel Model

Hinton and Stensel[16] developed a model that considers transport of both substrate and oxygen into a biofilm, as well the effects of hydrodynamics in the tower. Consumption of substrate within the biofilm is conceptualized as a one-dimensional diffusion mass transport process with dual substrate limited kinetics. Rather than Monod kinetics, however, the model uses Blackman kinetics, which treats the reaction as being either first-order or zero-order with respect to the substrate and oxygen concentrations. The biofilm is assumed to be deep, with the active depth being determined by exhaustion of either the substrate or oxygen. The hydraulic characteristics of the tower are described as laminar liquid film flow, which is interrupted at regular intervals by falling liquid drops. The interruptions provide mixing so that a uniform average concentration is established at the start of each laminar flow zone. Overall tower performance is determined by repeatedly solving the model equations for the short laminar flow sections and combining the results of many sections. The average section length is similar to the distance between mixing zones in the model of Logan et al.,[24] so although the models are conceptually different, their portrayals of the system hydrodynamics

are similar in effect. Substrate transport is all through the laminar liquid film, but oxygen transport occurs both through the liquid film and through direct contact of the biofilm with the bioreactor gas phase when flow is shifted between paths. Predictions from the model were in agreement with experimental observations in the lab and in pilot plants.[16]

17.2 ROTATING DISC REACTORS

17.2.1 Description and Model Development

17.2.1.1 Description

In rotating disc reactors (RDRs), closely spaced discs are mounted on a common horizontal shaft placed very near to or touching the liquid surface in a long narrow tank. Although called rotating biological contactors (RBCs) in practice, we will refer to them as rotating disc reactors herein to focus on the nature of the reactor employed. The shaft is rotated at constant speed, thereby allowing any point on a disc to be alternately submerged and exposed to the atmosphere. When water containing organic matter, nitrogen, and other nutrients flow through the bioreactor, microorganisms consume the substrates and grow attached to the disc as a biofilm. The rotating action imparts a shear force to the biofilm, keeping its thickness relatively constant by removing the cells generated by consumption of the substrate. The turbulence generated by the rotation transfers oxygen to the bulk liquid and keeps the sloughed microorganisms in suspension so they can be carried out in the effluent. The most common arrangement of the discs is with the shaft perpendicular to the direction of liquid flow, as shown in Figure 17.13. Under those circumstances, the turbulence is sufficient to make the substrate concentration uniform throughout the tank. In other words, for all practical purposes the tank can be considered to be completely mixed and can be modeled as such. We saw in Chapter 7 that bioreactors arranged in series perform better than a single bioreactor of similar total volume. Because of this and because of the modular nature of RDRs, most applications use a series of bioreactors. Consequently, the performance of an RDR system can be modeled as a series of CSTRs containing biofilms, the approach that we used for modeling a packed tower.

Just as in a packed tower, when the concentration of organic matter is high, nitrifying bacteria are unable to compete with the heterotrophic bacteria for space in the biofilm and thus the main reaction is carbon oxidation. However, once the concentration of organic matter has been reduced, nitrifiers can compete effectively for space and nitrification becomes significant. As a consequence, in systems containing a series of RDRs, carbon oxidation is the predominant reaction in the first few bioreactors while nitrification is more important in the later stages. In some cases, wastewater from which the organic matter has been removed in an upstream biochemical operation is applied to an RDR specifically for the oxidation of ammonia-N to nitrate-N, in which case the biofilm is composed almost entirely of nitrifying bacteria. Regardless of the nature of the reactions, however, the configuration of the RDR gives it characteristics that differ from that of a packed tower. In order to gain an appreciation for the differences and similarities of the two attached growth bioreactor types, we will consider only growth of heterotrophic bacteria with a single limiting organic substrate. However, the reader should recognize that nitrifying biofilms will behave in a similar manner.

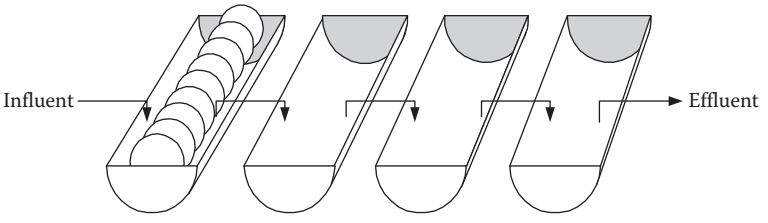

FIGURE 17.13 Schematic diagram of an RDR system.

Because of the complex nature of an RDR, certain simplifying assumptions must be made to model it. The first is that steady-state conditions prevail so that microorganisms are sheared from the surface of the biofilm at a rate equal to their growth. The resulting thickness will be an output of the model. The second assumption is that the turbulence in the bioreactor fluid is sufficient to keep the detached biomass in suspension so that it can be washed out with the effluent. The third is that biomass detachment from the biofilm is not dependent on the rotational speed of the disc. A similar assumption was made in the modeling of packed towers, where the detachment coefficient was assumed to be constant and independent of the THL. In reality, this assumption is not valid, but is made herein because RDRs are typically operated at a constant rotational speed, which will likely result in a constant value for this coefficient. The fourth is that both the attached and detached microorganisms contribute to substrate removal. The fifth is that oxygen and other nutrients are present in excess so that the organic substrate is the growth limiting nutrient. In other words, dual limitation by both the electron donor and the electron acceptor in the biofilm is not considered. The sixth is that the thickness of the liquid film is uniform over the portion of the disc that is not submerged (the aerated sector). The final assumption is that the substrate concentration in the liquid film on the aerated sector depends only on the circumferential angle (θ) and not on the radial position. In other words, the liquid film in the aerated sector is treated as a plug-flow reactor on top of the biofilm. The pseudoanalytical approach will be used for modeling transport and reaction of attached growth.

17.2.1.2 External Mass Transfer

As shown in Figure 17.14, each disc can be divided into two sectors: submerged and aerated. Because the biofilm is attached to the disc, it moves through the bulk fluid in the submerged sector, thereby making the external mass transfer coefficient, k_{Ls}, a function of the rotational speed, ω. As a point on the surface of the disc leaves the submerged sector and enters the aerated sector, a thin film of liquid adheres to it and is carried along with it. Although this film can be assumed to have no motion relative to the biofilm on the disc, its thickness is a function of the rotational speed of the disc.

Mass transfer from a fluid in laminar flow to the surface of a submerged rotating disc was analyzed by von Karman and given by Levich:[20]

$$k_{Ls} = 1.55 \cdot D_W^{0.667} \left(\frac{\mu_W}{\rho_W}\right)^{-0.167} \left(\frac{\omega}{r_o}\right)^{0.5}, \tag{17.29}$$

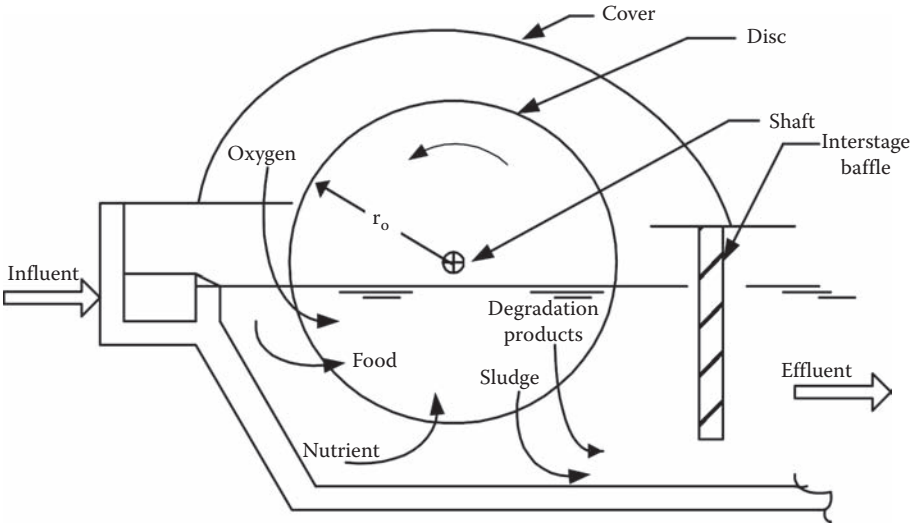

FIGURE 17.14 Schematic diagram of a single disc in an RDR system.

where k_{Ls} is the submerged external mass transfer coefficient, r_o is the outer radius of the disc, D_w is the diffusivity of the substrate in water, ρ_w is the fluid density, μ_w is the fluid viscosity, and ω is the rotational speed of the rotating disc. Equation 17.29 indicates that the external mass transfer coefficient will increase with the square root of the rotational speed. In practice, however, both the proportionality constant and the power on the rotational speed may be different due to deviations from the assumptions made in deriving the equation. Based on Equation 17.29 and substituting for the appropriate parameters, Grady and Lim[7] obtained the following relationship to estimate the external mass transfer coefficient:

$$k_{Ls} = 0.67\left(\frac{\omega}{r_o}\right)^{0.5}. \tag{17.30}$$

The units are m/hr for k_{Ls}, revolutions per minute for ω, and meters for r_o.

In the aerated sector, the entrained fluid forms a stagnant layer on top of the biofilm. The thickness of the liquid film entrained on a flat plate withdrawn vertically from a quiescent liquid has been analyzed by Landau and Levich[19] and found to be

$$L_w = \zeta \cdot v^{0.667}, \tag{17.31}$$

where L_w is the water film thickness, v is the withdrawal velocity, and ζ is a parameter that is dependent on the fluid properties (viscosity, density, surface tension at liquid-air interface). Since the withdrawal velocity of a point on a rotating disc depends on its radial position, some average velocity should be used, such as

$$v = \pi\omega\frac{r_o}{2}. \tag{17.32}$$

This suggests that the average film thickness is given by

$$L_w = \zeta_1 \cdot \omega^{0.667}, \tag{17.33}$$

where ζ_1 is a coefficient whose value depends on the fluid properties, the size of the disc, and its degree of submergence. Hartmann[13] has reported an entrained liquid film thickness of 40 μm for a smooth rotating disc and Equation 17.33 predicts a thickness of 60 μm under the conditions studied. Grieves,[9] on the other hand, reported that the thickness of the entrained stagnant liquid layer on top of a rotating biofilm ranged from 50 to 200 μm and was not reproducible. Such variability is due to the surface film depicted in Figure 16.1. Therefore, it is preferable to add an arbitrary amount to the thickness predicted by Equation 17.33 to account for the entrainment of fluid by the surface biofilm. Hence, a more appropriate form might be

$$L_w = \zeta_2 + \zeta_3 \cdot \omega^{\zeta_4}, \tag{17.34}$$

where ζ_2, ζ_3, and ζ_4 must be determined experimentally. For the purpose of gaining a basic understanding of the behavior of RDR systems a relationship following the form of Equation 17.33 developed by Grady and Lim[7] will be used:

$$L_w = 2.91 \cdot 10^{-4}(\omega \cdot r_o)^{0.667}. \tag{17.35}$$

The units for the thickness and external radius (r_o) are meters, and ω has units of revolutions per minute.

17.2.1.3 Model for the Submerged Sector

Having defined the approach to estimate the mass transfer coefficient in the liquid phase and the thickness of the entrained liquid film in the aerated sector, we will develop a simple model for an RDR using an approach analogous to that used for packed towers. Continuous stirred tank reactors will be used as "building blocks" representing an RDR (or a series of RDRs). The submerged sector will be modeled as a single CSTR containing a biofilm. The aerated sector can be viewed as a plug-flow reactor and its behavior can be approximated by a series of CSTRs.

Assuming steady-state operation and that the submerged sector can be described as a CSTR with a biofilm, and an expression similar to the one obtained in Equation 17.2 can be obtained:

$$F \cdot S_{SO} + F_L \cdot S_{SLR} - F \cdot S_{Sb} - F_L \cdot S_{Sb} - \frac{\hat{\mu}_H}{Y_H}\left(\frac{S_{Sb}}{K_S + S_{Sb}}\right) X_{B,H} \cdot V - J_{S,sub} \cdot f_{sub} \cdot A_s = 0, \quad (17.36)$$

where F_L is the flow rate associated with the liquid film carried from the submerged sector into the aerated sector, S_{SLR} is the substrate concentration in the liquid film on the aerated sector just before the film returns to the submerged sector, f_{sub} is the fraction of the total surface area (A_s) of the discs in the CSTR that is submerged, V is the volume of liquid in the CSTR, and $J_{S,sub}$ is the substrate flux into the biofilm in the submerged sector. Inspection of Equation 17.36 reveals that two liquid streams are considered in the material balance on substrate: the influent coming into the RDR and the liquid coming from the aerated sector.

To understand why the movement of liquid through the aerated sector is included in Equation 17.36, its volumetric flow rate must be compared to the influent flow rate. The volume of liquid present in the aerated sector at any point in time, V_{aer}, is simply the area of the aerated sector multiplied by the thickness of the entrained stagnant liquid film, L_w:

$$V_{aer} = (1 - f_{sub}) A_s \cdot L_w. \quad (17.37)$$

As discussed in Section 20.1.1, a typical RBC shaft with high-density media contains a total surface area of 13,900 m^2 and it is usually placed in a basin with a volume of 45 m^3. The entrained liquid film thickness can range from 50 to 200 μm. Assuming a thickness of 100 μm and a submergence fraction of 50%, the value of V_{aer} is 0.7 m^3 or about 1.5% of the volume of the CSTR in which the discs are submerged. Thus, the volume in the aerated sector at any instant is not large. The important consideration, however, is the flow rate through the aerated sector relative to the influent flow rate entering the RDR.

The flow rate (in m^3/hr) associated with the entrained liquid moving through the aerated sector for a given rotational speed, ω (revolutions/min), is given by

$$F_L = 60 \cdot \omega \cdot A_s \cdot L_w. \quad (17.38)$$

With a typical rotational speed of 1.6 rpm, for the same entrained liquid film thickness, Equation 17.38 gives a flow rate of 3203 m^3/day through the aerated sector. As discussed in Section 20.2.2, the concept of total hydraulic loading is also applicable to RDRs, but in this case it is defined as the ratio of the flow rate entering the RDR divided by the total surface area of the discs. A reasonable value for this parameter is 0.05 m/day. Consequently, for a total disc surface area of 13,900 m^2, 695 m^3/day of wastewater can be treated. Thus, it can be seen that the flow rate through the aerated sector is 4.6 times the influent flow rate to the RDR, which is why it is necessary to consider the entrained flow through the aerated sector in Equation 17.36.

A material balance can be written for the suspended biomass in the submerged sector of the RDR. Assuming that suspended biomass is not carried into the aerated sector and that no suspended biomass enters the submerged sector from the aerated sector leads to

$$F \cdot X_{B,HO} - F \cdot X_{B,H} + \hat{\mu}_H \left(\frac{S_{Sb}}{K_S + S_{Sb}} \right) X_{B,H} \cdot V - b_H \cdot X_{B,H} \cdot V + b_D \cdot X_{B,Hf} \cdot f_{sub} \cdot A_s \cdot L_f = 0. \tag{17.39}$$

Equation 17.39 contains the biofilm thickness, L_f. Just as was done during the development of the model for a packed tower, this variable can be expressed as a function of the substrate flux into the biofilm (see Equation 17.4). In this case, however, two fluxes are involved: that in the submerged sector ($J_{S,sub}$) and that in the aerated sector ($J_{S,aer}$). Consequently:

$$J_{S,sub} \cdot Y_H \cdot f_{sub} \cdot A_s + J_{S,aer} \cdot Y_H \cdot (1 - f_{sub}) A_s - (b_D + b_H) X_{B,Hf} \cdot A_s \cdot L_f = 0. \tag{17.40}$$

Solving for the biofilm thickness, we obtain:

$$L_f = [J_{S,sub} \cdot f_{sub} + J_{S,aer} (1 - f_{sub})] \left[\frac{Y_H}{(b_D + b_H) X_{B,Hf}} \right]. \tag{17.41}$$

Although Equation 17.41 is exact, for the situations considered here, the contribution of substrate utilization in the aerated sector to the biofilm thickness is minor. Consequently, we can approximate Equation 17.41 as

$$L_f \cong \frac{J_{S,sub} \cdot f_{sub} \cdot Y_H}{(b_D + b_H) X_{B,Hf}}. \tag{17.42}$$

This is similar to Equation 16.21. Solving the model for the submerged section follows the same procedure as used for packed towers. It consists of using the pseudoanalytical approach (i.e., finding the solution to Equation 16.38) to find the value of the bulk substrate concentration that results in a substrate flux that satisfies the material balance equations for substrate and biomass (Equations 17.36 and 17.39). MATLAB® was used to implement this scheme numerically.

17.2.1.4 Model for the Aerated Sector

Because the aerated sector behaves as a PFR, it is modeled as a series of 25 CSTRs. The total liquid volume in the aerated sector is given by Equation 17.37; consequently, the volume of each reactor in the series is 1/25th of that value. We saw in Equation 17.42 that the thickness of the biofilm in the RDR is determined primarily by substrate removal in the submerged sector. Therefore, the biofilm thickness in the aerated sector was assumed to be constant and equal to the value given by Equation 17.42. In addition, the biofilm in the aerated sector was assumed to be deep, which means that the substrate is exhausted in the biofilm before it reaches the support surface. This assumption is reasonable because the mass of biomass in the aerated sector is large relative to the mass of substrate applied to it. Nevertheless, the assumption was verified by using Equation 16.35 to calculate the value of ξ, which for deep biofilms has a value equal to one.

Writing a material balance on substrate for the aerated sector results in

$$F_L \cdot S_{Sb} - F_L \cdot S_{SLR} - J_{S,aer} (1 - f_{sub}) A_s = 0. \tag{17.43}$$

As discussed in Section 16.2.3, the flux to a deep biofilm can be calculated using Equation 16.29, which was presented in dimensionless form. Using the appropriate transformations for the dimensionless parameters that equation can be rewritten as

$$J_{S,aer} = \left\{ 2\hat{q}_H \cdot X_{BH,f} \cdot D_e \left[S_{SLR} + K_S \ln\left(\frac{K_S}{K_S + S_{SLR}} \right) \right] \right\}^{0.5}. \quad (17.44)$$

Equations 17.43 and 17.44 can be combined to calculate the effluent substrate concentration from the aerated sector if it were modeled as a single CSTR:

$$S_{Sb} = S_{SLR} + \frac{(1 - f_{sub}) A_s}{F_L} \left\{ 2\hat{q}_H \cdot X_{BH,f} \cdot D_e \left[S_{SLR} + K_S \ln\left(\frac{K_S}{K_S + S_{SLR}} \right) \right] \right\}^{0.5}. \quad (17.45)$$

Because the aerated sector was modeled as a series of 25 CSTRs, Equation 17.45 was used 25 times, with the surface area in each reactor equal to 1/25th of the total surface area of the aerated sector. The influent to the first reactor was S_{Sb}, the substrate concentration in the submerged sector. The effluent from the first reactor was the influent for the second reactor and so forth down the chain. The 25th reactor produced an effluent with concentration S_{SLR}.

17.2.2 Performance of Rotating Disc Reactor Systems

The kinetic parameters, stoichiometric coefficients, and system variables for the RDR considered in this chapter are given in Table 17.2. The rotational speed, total media surface area, percentage of submergence, hydraulic loading, and tank volume RDR are representative of full-scale RBC systems, as discussed in Chapter 20. The kinetic and stoichiometric parameters associated with microbial growth and mass transfer processes in Table 17.2 are the same as those used to simulate the performance of the packed towers as given in Table 17.1. Thus, the relationship between substrate flux and bulk substrate concentration shown in Figure 17.2 is also applicable to the RDR simulations.

The effect of the influent flow rate on the removal of substrate in an RDR with a fixed number of discs rotating at a fixed speed is shown in Figure 17.15. Increasing the flow rate entering the RDR results in an increase in the effluent substrate concentration and a decrease in the percentage of substrate removal. A similar result was obtained when packed towers were considered (Figure 17.9). Increasing the flow rate into the system, while maintaining a constant influent substrate concentration, results in the higher mass application rates of substrate. To attain a steady state, which requires higher reaction rates at higher flow rates, the substrate flux into the biofilm must increase. The only way this can occur is for the bulk substrate concentration to increase. In an RDR the hydraulic loading is a commonly used indicator of performance. It is defined as the flow applied per unit of total (aerated and submerged sector) wetted surface area and is shown across the top of Figure 17.15 to give an indication of its effect on performance.

The effects of influent substrate concentration are also shown in Figure 17.15. Examination of the curves shows that the effluent substrate concentration will increase as the influent substrate concentration is increased. Analogous to increases in flow rate, increases in influent substrate concentration increase the mass application rate of organic matter per unit area of biomass. At the new steady state, an increase in the substrate flux into the biofilm must be achieved, which can only be done by maintaining a higher substrate concentration in the bulk liquid. This dependence is nonlinear, as shown in Figure 17.2, so that larger increases in the influent substrate concentration cause proportionally higher increases in the effluent substrate concentration.

One assumption made in the development of the model is that most of the substrate removal occurs in the submerged sector and this is indeed the case, as shown in Figure 17.16. One implication

TABLE 17.2
Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used to Generate Figures 17.15–17.22

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD·hr)	0.27
$\hat{\mu}_H$	hr^{-1}	0.135
K_S	mg/L as COD	10
Y_H	mg biomass COD/mg substrate COD	0.50
b_D	hr^{-1}	0.0017
b_H	hr^{-1}	0.0025
$X_{B,Hf}$	mg biomass COD/cm^3	40
D_e	m^2/hr	2.667×10^{-6}
k_L	m/hr	$k_{Ls} = 0.67\left(\frac{\omega}{r_o}\right)^{0.5}$
V_L	m^3	45
r_o	m	1.83
L_w	m	$L_W = 2.91 \times 10^{-4}\ (\omega \cdot r_o)^{0.667}$
A_s	m^2	13900
f_{sub}	—	0.40
ω	rpm	1.6
F	m^3/hr	50
S_{SO}	mg/L as COD	100
$X_{B,HO}$	mg/L as COD	0

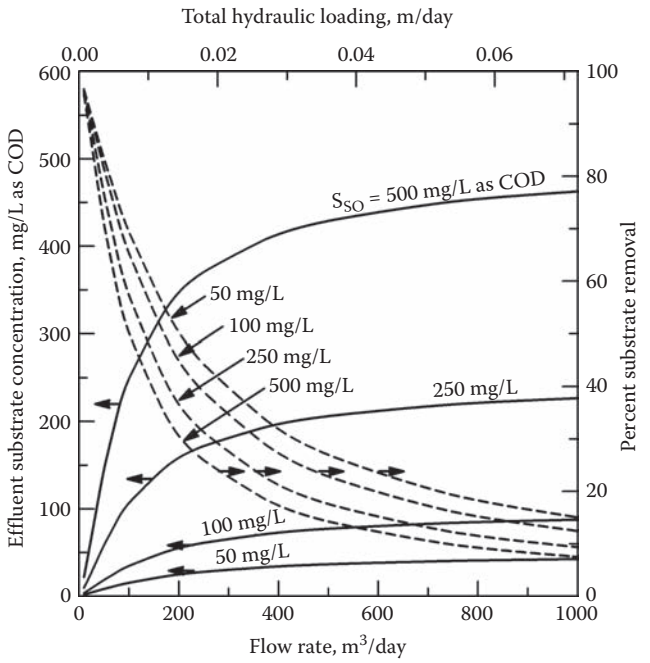

FIGURE 17.15 Effects of influent flow rate and influent substrate concentration on the performance of a single RDR. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.2 unless otherwise specified.

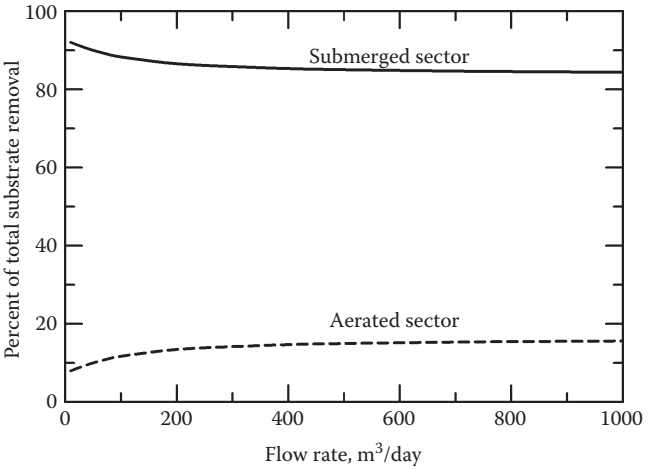

FIGURE 17.16 Effect of flow rate on the fraction of the total substrate removal achieved by the submerged and aerated sectors of a single RDR. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.2 unless otherwise specified.

of that assumption is that most of the biomass growth occurs in the submerged sector, thereby making the thickness of the biofilm a function of the substrate flux occurring in the submerged sector, as indicated by Equation 17.42. Examination of Figure 17.16 shows that as the flow rate to the RDR is increased, the percentage of the total substrate removal associated with the submerged sector declines slightly, while that associated with the aerated sector increases. For higher flow rates, up to 15% of the substrate removal occurs in the aerated sector; hence, the total biofilm thickness calculated using Equation 17.42 will underestimate the biofilm thickness by that percentage. Furthermore, higher influent substrate concentrations and higher rotational speeds result in as much as 25% of the substrate removal occurring in the aerated sector. It is important to recall that all of the simulations resulted in deep biofilms. One characteristic of deep biofilms is that the substrate flux is independent of the biofilm thickness. Hence, the main effect of underestimating the biofilm thickness is an underestimation of the suspended biomass concentration, which contributes very little to the total substrate removal in the system.

Figure 17.17 shows the effect of rotational speed on the performance of an RDR. As the rotational speed is increased, the percentage of substrate removal increases up to an upper limit characterized by the other system parameters. Several factors interact to cause this response. As shown in Equation 17.30, the external mass transfer coefficient in the submerged sector increases as the rotational speed of the discs is increased. Consequently, substrate is consumed more rapidly by the submerged biofilm as the rotational speed is increased. In addition, the coefficient for biomass detachment will increase with the rotational speed because of its impact on shear, but this is not considered in the model.

When considering the effects of rotational speed on RDR performance, the events occurring in the aerated sector also need to be investigated. Figure 17.18 shows the substrate concentration profiles in the aerated sector as a function of the rotational speed. Two interacting events influence those profiles. First, as seen in Equation 17.33, the volume of fluid carried with the discs into the aerated sector increases as the rotational speed is increased. This means that the mass of substrate carried into the aerated sector increases. Second, at higher rotational speeds, it takes less time for a point on the disc to move a given fractional distance around the disc. Following an element of fluid on the disc as it moves from the submerged sector, through the aerated sector, and back to the submerged sector again, we see that at low rotational speeds, the substrate concentration rapidly approaches zero. This is because the large amount of biomass in the biofilm is receiving only a small

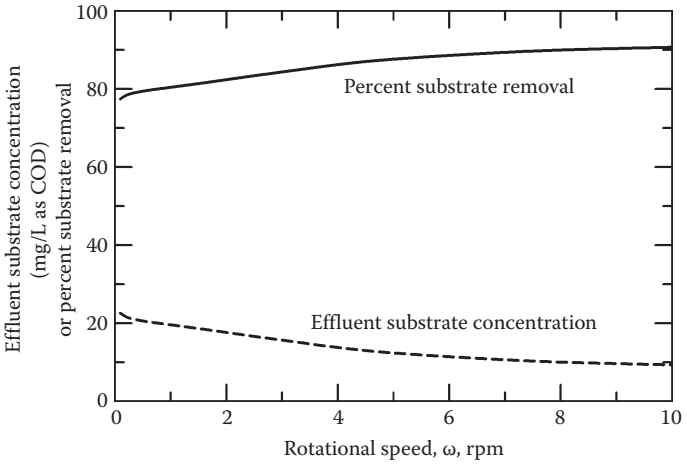

FIGURE 17.17 Effects of rotational speed on the performance of a single RDR. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.2 unless otherwise specified.

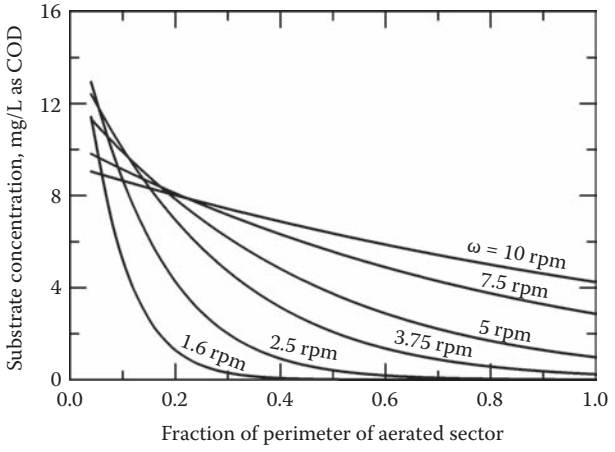

FIGURE 17.18 Effects of rotational speed on the substrate concentration profiles in the aerated sector of a single RDR. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.2 unless otherwise specified.

amount of substrate, and because it takes a point on the disc surface a relatively long time to move through the aerated sector. As the rotational speed increases, the amount of substrate entering the aerated sector increases, but the time required for a point on the disc to move through the aerated sector decreases. These act together to allow substrate to move further around the disc. When the rotational speed is about 3.75 rpm, the substrate concentration reaches zero at the very end of the aerated sector. Further increases in the rotational speed cause the substrate concentration at the end of the aerated sector to increase and for the substrate concentration profile to flatten. At very high rotational speeds, the behavior of the two sectors approaches that of a single CSTR. The typical rotational speed for full-scale RBCs is 1.6 rpm. Although this does not take full advantage of the substrate removal capacity of the aerated sector, it is used because of other factors, such as wear on bearings and other physical features.

The performance of RDR systems is highly dependent on the surface area available for biofilm growth, just as it is for packed towers. The total area was increased by increasing the number of

discs per stage while keeping the diameter of each disc and the basin volume constant. Figure 17.19 shows how the total surface area affects system performance. An increase in the number of discs results in a larger area for both the submerged and aerated sectors (and hence more biomass) and in a greater volume of fluid carried with the discs into the aerated sector. Consequently, the substrate removal rates in both the submerged and the aerated sectors increase with an increase in the number of discs, causing a reduction in the effluent substrate concentration and an increase in the percentage of removal. A similar effect is observed by considering a constant value for the total disc surface area (given in Table 17.2) and changing the fractional submergence of the RDR discs (Figure 17.20). An increase in the fractional submergence increases the total submerged area, which allows more microorganisms to grow on a disc of a fixed size. Consequently, it causes the substrate removal rate in the submerged sector to increase. Although it also decreases the substrate flow through the aerated sector, the net effect is an increase in substrate removal because the submerged sector provides the majority of the total substrate removal. Although not reflected in the model, submergence

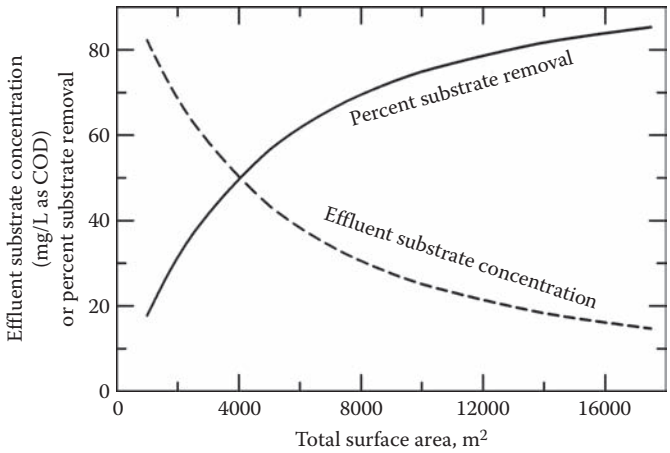

FIGURE 17.19 Effect of total disc surface area on the performance of a single RDR. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.2 unless otherwise specified.

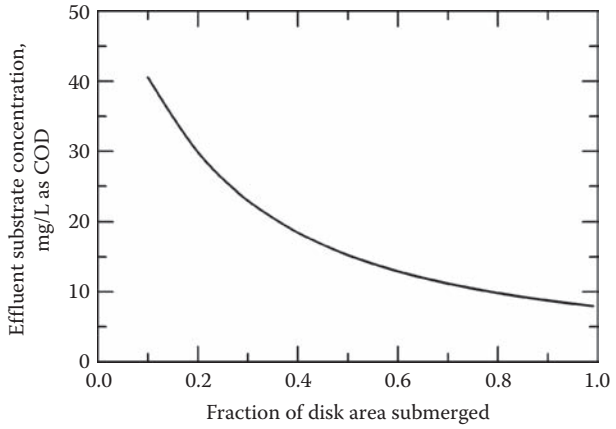

FIGURE 17.20 Effect of the degree of disc submergence on the performance of a single RDR. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.2 unless otherwise specified.

in excess of 0.5 will decrease the rate of oxygen transfer in the system, thereby hurting bioreactor performance.

Scale-up of RDRs is complicated because of the many factors involved. One approach that found use early in the development of RDRs was to maintain a constant peripheral velocity while maintaining the same hydraulic loading, probably as a result of recommendations based on lightly loaded systems.[5] To investigate the efficacy of this practice, simulations were performed to investigate the effect of disc size while maintaining the peripheral velocity and the hydraulic loading constant as the disc diameter was increased, and the results are presented in Figure 17.21. Maintenance of a constant peripheral velocity required that the rotational speed be decreased as the disc diameter was increased (the product $\omega \cdot r_o$ was maintained constant), so the rotational speed corresponding to each disc size is shown on the upper abscissa. Examination of Figure 17.21 shows clearly that the scale-up strategy is not effective since the percentage of substrate removal decreases as the disc size is increased, an observation that has been made in practice as well.[41] This is a result of two effects. First, because the rotational speed is decreased to maintain a constant peripheral velocity as the disc size is increased, the external mass transfer coefficient in the submerged sector is decreased, reducing the substrate removal rate. Second, the decrease in rotational speed reduces the volume of liquid carried through the aerated sector relative to the influent flow rate, which decreases the mass of substrate removed there. Consequently, the loss in performance associated with an increase in disc size is primarily due to the effect of rotational speed on the rate of mass transfer in the submerged sector and the movement of liquid through the aerated sector. In addition to the problems discussed above, several other effects complicate the problem of scale-up.[35] As a result, and because no suitable scale-up strategy has been found, it is recommended that pilot studies be performed with full-scale discs.

It has been demonstrated throughout this book that a reactors-in-series system will outperform a single CSTR of equal volume for a culture that grows according to Monod kinetics. The benefits of staging bioreactors are also applicable to RDR systems. To demonstrate this, simulations were done for six identical RDRs in series, with a total surface area and volume equal to the values from Table 17.2 (previously used for an RDR in a single CSTR). Figure 17.22 shows the substrate concentration profile along the chain and the performance of the staged system is compared to that of the single CSTR. Staging brings great benefits in terms of substrate removal. The main reason for this is that in a single CSTR, the concentration at which substrate is being removed is equal to the effluent substrate concentration. For the reactors-in-series, the first three reactors have a substrate

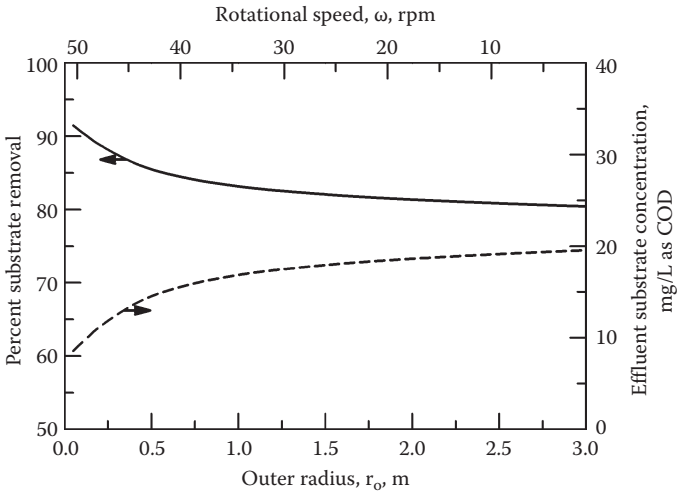

FIGURE 17.21 Effect of disc size on the performance of a single RDR with constant hydraulic loading (0.086 m/day) and peripheral velocity (16.1 m/min). The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.2 unless otherwise specified.

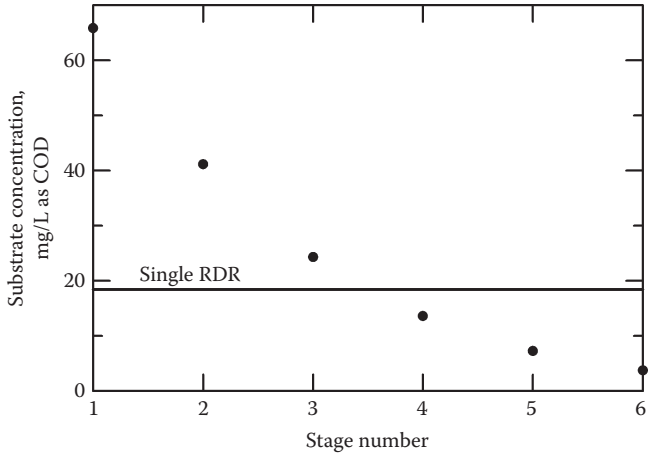

FIGURE 17.22 Effect of staging on the performance of an RDR system. A chain of six RDRs is compared to a single RDR containing the same total biofilm surface area and receiving the same loading. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 17.2.

concentration higher than the effluent substrate concentration from the single CSTR, which means that the substrate flux (substrate removal rate per unit area) will be higher in them than in a single CSTR. The consequence of this is that a large percentage of the substrate is removed at rates higher than that in a CSTR, leading to a higher overall rate of substrate removal and lower effluent substrate concentrations.

17.2.3 Other Rotating Disc Reactor Models

Several other RDR models are available in the literature. Although no model is widely accepted for design purposes, it would be instructive to review briefly the key features of these models in comparison to the model presented above. A detailed analysis of several RDR models has been presented by Spengel and Dzombak[35] as well as by Patwardhan.[28] Readers interested in more information on this subject are urged to consult their work.

17.2.3.1 Grady and Lim Model

The steady-state model developed by Grady and Lim[7,8] has several similarities to the model presented herein. It considers the submerged and aerated sectors distinctly, as well as the effect of the rotational speed on the amount of liquid transferred to the aerated sector and the mass transfer coefficients for the submerged and aerated sectors. The model assumes a constant mass of biomass in the system and this quantity is an input to the model, not an output. Attached growth was modeled using the effectiveness factor approach. The submerged sector was modeled as a CSTR and a material balance on substrate led to

$$F \cdot S_{SO} + F_L \cdot S_{SLR} - F \cdot S_{Sb} - F_L \cdot S_{Sb} - \eta_{eOs} \cdot X_{B,Hf} \cdot L_f \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}} \right) f_{sub} \cdot A_s - \frac{\hat{\mu}_H}{Y_H} \left(\frac{S_{Sb}}{K_S + S_{Sb}} \right) X_{B,H} \cdot V = 0, \tag{17.46}$$

where η_{eOs} is the effectiveness factor for the submerged sector. The aerated sector was modeled as a PFR, by writing a material balance for an infinitesimal section of the sector as a function

of the circumferential angle (θ) and representing the effectiveness factor for the aerated sectors as η_{eOa}:

$$\frac{dS_{SL}}{d\theta} = -\eta_{eOa} \cdot X_{B,Hf} \cdot L_f \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}} \right) \left(\frac{1}{2\pi L_w \omega} \right). \tag{17.47}$$

Because of the assumption made in the model that the substrate concentration in the liquid film was a function of θ alone, the boundary condition was approximated by

$$S_{SL} \approx S_{Sb} \text{ at } \theta = 0, \tag{17.48}$$

and the concentration in the returning liquid (circumferential angle $\theta = \theta_A$) was approximated by

$$S_{SLR} \approx S_{SL} \text{ at } \theta = \theta_A. \tag{17.49}$$

The model does not consider the possibility of the terminal electron acceptor limiting the growth rate, an assumption that was also made in the model presented in Section 17.2.1. The overall response of the model is very similar to that shown in Figures 17.15 to 17.22.

17.2.3.2 Kornegay Model

The Kornegay[18] model ignores the two distinct sectors, aerated and submerged, and assumes that the entire biofilm is exposed to the substrate concentration in the bulk liquid. It also assumes that the rate of substrate removal by suspended cells is negligible. The effects of mass transfer are accounted for by assuming that the Monod half-saturation coefficient is a function of the rotational speed of the discs. This is equivalent to making the overall effectiveness factor dependent on the substrate concentration and the half-saturation coefficient. Thus, Kornegay's[18] model can be viewed as a special case of the Grady and Lim[7,8] model in which $L_w = 0$, η_{eOs} is determined by S_{Sb} and K_S, and the area of the submerged sector is equal to the area of disc covered by biofilm.

17.2.3.3 Model of Hansford, Andrews, Grieves, and Carr

The steady-state model of Hansford et al.[11] has many similarities to the one of Grady and Lim.[7,8] It recognizes that there are two sectors, submerged and aerated, and that a rotating disc carries a liquid film from the tank into the aerated sector as it turns. Their model was developed by writing four substrate mass balance equations: for the liquid film in the aerated sector, for the biofilm in the aerated sector, for the liquid in the tank, and for the biofilm in the submerged sector. The following assumptions were made: the liquid in the tank is completely mixed so that the substrate concentration is uniform; the liquid film and the biofilm in the aerated sector are completely mixed in both the axial and radial directions so that the substrate concentration is a function of only the angular position, θ; the substrate concentration in the biofilm in the submerged sector is uniform with respect to radial, axial, and angular position; the thickness of the liquid film in the aerated sector is constant and independent of the rotational speed; and the suspended biomass in the tank consumes no substrate. The major difference between this model and one of Grady and Lim's[7,8] is that it assumes that the substrate concentration in the biofilm is independent of depth. This is equivalent to saying that there is no mass transfer resistance within the biofilm or that the thickness of the biofilm is so small that the reaction takes place only at its surface. In either case, the internal effectiveness factor is unity. Furthermore, this model assumes that the reaction takes place only in the biofilm rather than in both the biofilm and bulk liquid. Finally, the thickness of the liquid film in the aerated sector is assumed to be constant, whereas the model used by Grady and Lim[7,8] allows variations due to rotational speed. In spite of these differences, the trends predicted by the model of Hansford et al.[11] are qualitatively similar to those of Grady and Lim,[7,8] as well as to trends observed in the field.

17.2.3.4 Model of Famularo, Mueller, and Mulligan

The RDR model of Famularo et al.[4] incorporates the transport of both oxygen and substrate up to and into the biofilm. The basic components of the model (i.e., the mass transfer effects, reaction rates, etc.) are very similar to the ones in the model of Grady and Lim,[7,8] although the effectiveness factor approach is not used. Rather, they use a finite difference procedure to solve the nonlinear coupled differential equations and the trends predicted by the model are in agreement with the trends presented in this chapter. The major benefit of the model of Famularo et al.[4] is that the inclusion of oxygen transfer effects allows prediction of the loading conditions under which oxygen limitations are likely to occur.

17.2.3.5 Model of Watanabe

Watanabe[39] has developed a model for carbon oxidation, nitrification, and denitrification in an RDR that employs a number of empirical relationships. While a direct solution technique is used, the kinetics are simplified, with zero-order intrinsic kinetics being assumed for both nitrification and denitrification, giving overall half-order reaction rates when transport is considered. In addition, an empirical relationship is employed to estimate the distribution of heterotrophic and autotrophic biomass in the biofilm. The model also assumes that the oxygen flux into the biofilm is independent of the composition of the microbial community, so that the use of oxygen by the two bacterial types will depend upon their distribution in the biofilm.

17.2.3.6 Model of Gujer and Boller

Gujer and Boller[10] have also developed an RDR model that incorporates carbon oxidation, nitrification, and denitrification. The reactions included are aerobic growth of heterotrophs, anoxic growth of heterotrophs, and aerobic growth of nitrifiers. Decay of both heterotrophs and autotrophs is by the lysis:regrowth approach, with the lost biomass going directly to readily biodegradable substrate since slowly biodegradable substrate is not included. Surface shear of the biofilm is included, allowing the biofilm thickness to be an output from the model. Dual limitation by both electron donor and acceptor is also included. Because of that, the direct solution approach mentioned in Section 16.3 is employed. The RDR is modeled as a series of CSTRs with reaction occurring only in the submerged sector. The model was able to simulate the effect of organic substrate biodegradation on nitrification and the results were consistent with the concepts presented in Section 16.4, with nitrification occurring in the later stages after the bulk of the organic matter had been utilized. Although the authors caution that the model is not suitable for design, it represents a significant step in the modeling of all possible events in an RDR, although it does not include the effects of disc rotational speed.

17.2.3.7 Model of Spengel and Dzombak

Spengel and Dzombak[35] built on the work of Famularo et al.[4] to develop a very complete model that considers both carbon oxidation and nitrification with competition for oxygen in the biofilm. Their substrate utilization rate terms are double Monod expressions (Equation 3.46), considering both the electron donor and the electron acceptor (oxygen). However, like the model of Grady and Lim,[7,8] the biofilm thickness and density have to be provided as input. Two simplifications are made to facilitate their solution. First, the disc is divided into four sectors, two in the submerged region and two in the aerated region, and the substrate removal rates are calculated for each sector. Second, the biofilm is divided into four layers, with each being considered as a mixed zone. In that way substrate transport into the biofilm can be considered easily. Spengel and Dzombak[35] were able to calibrate the model to mimic the results from pilot tests well, thereby facilitating the interpretation of those tests. Although several issues still need resolution, it is possible that the model will be useful to assist in scale-up.

17.3 KEY POINTS

1. A packed tower may be modeled by assuming plug flow and allowing the substrate to be consumed by both dispersed and attached microorganisms. However, to more accurately reflect reality, intermixing should occur at regular intervals to account for falling liquid droplets and intersecting flows within the media. Both internal and external mass transfer must be accounted for and this may be done by using the pseudoanalytical approach if only a single limiting nutrient is considered.
2. The key relationship in biofilm reactors is the substrate flux into the biofilm versus the bulk substrate concentration. The substrate flux is the result of the combined effect of microbial growth kinetic and mass transfer coefficients.
3. Low substrate fluxes can be maintained with low bulk substrate concentrations. The relationship between these two quantities follows a first-order dependence, because of Monod kinetics being reduced to first-order at low substrate concentrations and mass transfer being a first-order process.
4. To maintain high substrate fluxes into a biofilm, the bulk substrate concentration needs to be high. Under that condition, the dependence of flux on the bulk substrate concentration approaches a half-order relationship.
5. For low and intermediate influent substrate concentrations, the concentration drops exponentially with depth in packed towers, either with or without recirculation of clarified effluent, resembling a first-order process.
6. Biofilm thickness decreases with tower depth because decreasing bulk substrate concentrations lead to lower substrate fluxes.
7. Suspended biomass in packed towers makes a small contribution to total substrate removal in the tower.
8. Larger surface areas in packed towers result in lower effluent substrate concentrations because lower substrate fluxes are needed to maintain steady state.
9. When there is no recirculation and when the feed substrate concentration is high, the substrate concentration drops in a linear manner with depth in a packed tower, resembling a zero-order process.
10. As the feed substrate concentration to a packed tower is increased, the fractional substrate removal decreases, although the mass removal rate increases (but in diminishing increments).
11. Higher recirculation ratios result in flatter substrate profiles in a packed tower. The percentage of substrate removal generally decreases as the recirculation ratio is increased, with the result that larger towers are required to achieve the same fractional substrate removal when recirculation is used.
12. Oxygen transfer can limit substrate removal in a packed tower. The maximum oxygen transfer rate is approximately 650 mg/(m^2·hr) and loadings that require more oxygen than that will not achieve expected levels of substrate removal. At typical total hydraulic loadings, the maximum allowable applied substrate concentration to avoid that limit is around 250 mg COD/L.
13. Because of competition for space and oxygen between heterotrophic and autotrophic bacteria, carbon oxidation generally occurs at the top of packed towers and nitrification near the bottom.
14. A number of semiempirical and mechanistic models are available for modeling the performance of biological packed towers.
15. The performance of a rotating disc reactor (RDR) system can be modeled as a series of continuous stirred tank reactors (CSTRs) containing biofilms.
16. In an RDR the external mass transfer coefficient for the submerged sector increases with the rotational speed of the rotating disc in accordance with a power law.

17. The submerged sector of an RDR can be modeled as a CSTR by writing steady-state mass balances on substrate and suspended biomass around the liquid volume in the tank and using the pseudoanalytical approach to calculate the substrate flux into the biofilm
18. The aerated sector of an RDR can be modeled as a series of CSTRs containing deep biofilms.
19. Even though the aerated sector of an RDR accounts for only a small portion of the total system volume, the flow rate entering it is large compared to the influent flow rate. The flow rate into the aerated sector increases with the rotational speed of the rotating disc in accordance with a power law.
20. Increases in influent flow rate cause increases in the effluent substrate concentration from an RDR.
21. The effluent substrate concentration increases as the influent concentration to an RDR is increased. At low flow rates, the percent substrate removal is relatively independent of the influent substrate concentration, while at high flow rates it decreases as the influent substrate concentration is increased.
22. As the rotational speed of an RDR is increased, the percentage of substrate removal increases up to an upper limit characterized by the system parameters.
23. An increase in the number of discs (total surface area) in an RDR results in an increase in the substrate removal rate and a decrease in the effluent substrate concentration.
24. An increase in the fractional submergence of the discs results in an increase in the substrate removal rate in an RDR.
25. Rotating disc reactors are usually run in series to economically achieve a desired effluent substrate concentration.
26. In a chain of RDRs, the majority of the substrate is removed in the first few reactors and the others contribute relatively little. At higher substrate concentrations, more removal occurs in the later stages.

17.4 STUDY QUESTIONS

1. Describe the assumptions inherent in the packed tower model presented herein, telling how they affect the generalities about packed tower performance arrived at from simulations.
2. Describe the packed tower model used in this chapter, giving its general characteristics.
3. Draw a sketch of the dependence of the substrate flux into a biofilm on the bulk substrate concentration and explain why the curve has the shape that it does.
4. Generate curves such as the one shown in Figure 17.2, using the parameters from Table 17.1 but change each of the following parameters one at a time. Increase k_L by 1000, increase D_e by 1000, and increase $\hat{q}_H$ by 10. Using these curves, explain which factor dominates the dependence of the substrate flux on the bulk substrate concentration.
5. Draw a sketch showing the effect of the influent substrate concentration on the fractional substrate removal and the mass removal rate of substrate in a packed tower. Also explain why the mass removal rate curve has the shape that it does.
6. Draw a sketch showing the effect of the influent flow rate on the fractional substrate removal and the mass removal rate of substrate in a packed tower of fixed size. Also explain why the curves have the shapes that they do.
7. Describe and explain the effects of recirculation of clarified effluent on the fractional substrate removal in a packed tower as predicted by the model presented in this chapter. Also describe situations in which the effects could be different from the model simulations and explain why.
8. Describe the factors that can prevent a packed tower from behaving as a perfect plug-flow reactor and explain their impact on tower performance.

9. Explain why nitrification typically follows carbon oxidation in a packed tower.
10. Explain why and how the external mass transfer coefficient for the liquid in an RDR is influenced by the rotational speed.
11. Describe in general terms the approach that must be taken to model the performance of an RDR.
12. Describe the effects of influent flow rate, influent substrate concentration, and rotational speed on the performance of a single RDR and explain why they occur.
13. Explain why it is not possible to scale-up an RDR by assuming that a full-scale unit will have the same effluent quality as a pilot-scale unit with the same hydraulic loading and peripheral velocity.
14. Explain why RDRs are usually run in series.

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18 Fluidized Bed Biological Reactors

The biofilm reactors that we have considered previously both come under the heading of fixed media bioreactors in which the media stays in one position as the wastewater flows past. As wastewater flows down through a packed tower, the biofilm grows attached to a solid support that is either stacked or dumped into place, removing the soluble substrate as it does so. Careful selection of the hydraulic and organic loadings is required to ensure that sufficient shear occurs relative to growth to prevent the interstitial spaces in the media from plugging, causing the system to fail. In addition, any suspended solids applied to the tower must be sufficiently small to allow them to pass through those spaces without being trapped, because entrapment would lead to plugging. Likewise, in a rotating disk reactor, the rotational speed must be selected to ensure sufficient shear to prevent the biofilm from bridging the spaces between the disks, thereby blocking contact of the wastewater with the biofilm. Furthermore, as with the packed tower, care must be exercised concerning the admission of suspended solids.

In contrast to the fixed media bioreactors, the bioreactors considered in this chapter come under the broad heading of mobile bed bioreactors.[16] Such bioreactors include all biofilm systems with continuously moving media, whether that movement is induced by high air or water velocities or by mechanical stirring. Moving media provide several distinct advantages: they allow better control of biofilm thickness, have superior mass transfer characteristics, are not subject to clogging, and provide very high surface areas for biofilm development while maintaining low pressure drops.[16] As a consequence, they are rapidly gaining acceptance. In this chapter we focus on the theoretical performance of one type of mobile bed bioreactor, the fluidized bed biological reactor (FBBR), because it is one of the more popular ones and has some distinctive characteristics of which environmental engineers should be aware. In Chapter 21 we consider its design, as well as that of some other mobile bed bioreactors.

18.1 DESCRIPTION OF FLUIDIZED BED BIOLOGICAL REACTOR

18.1.1 General Characteristics

A fluidized bed biological reactor (Figure 18.1[30]) is one in which the biofilm grows attached to small carrier particles that remain suspended in the fluid (i.e., fluidized) by the drag forces associated with the upward flow of water. The term bioparticle generally denotes a biofilm covered carrier particle, although in some cases (e.g., granules) bioparticles develop without the presence of a carrier particle. Microbial granulation is commonly associated with anaerobic systems, such as in upflow anaerobic sludge blanket (UASB) reactors. Recently it has been demonstrated that it can also occur under aerobic conditions by manipulation of the growth conditions (organic loading, shear, and hydrodynamics).[6,17] Most FBBRs are two-phase systems, containing only water and bioparticles, and if oxygen is required it is dissolved in the recirculation flow prior to its return to the reactor. However, recent advances in system design have allowed the incorporation of a gas phase, thereby allowing oxygen transfer directly in the bioreactor.[16] Although the popularity of three-phase systems is increasing, they are considerably more complicated to model than two-phase systems, and thus our discussion is limited to the latter. It should be noted that the designation of two-phase and three-phase is made with regard to reactants. Some so-called two-phase systems, such as denitrifying

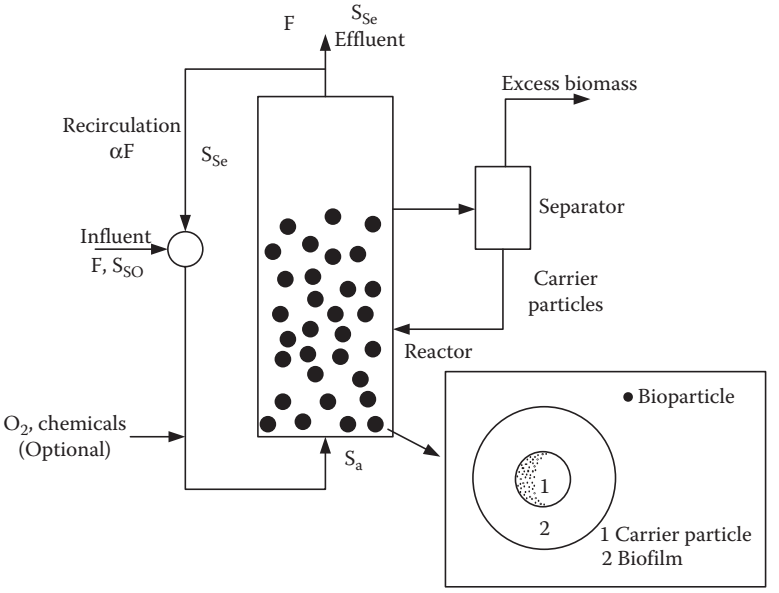

FIGURE 18.1 Schematic diagram of an FBBR.

and methanogenic systems actually have a gas phase in them because of the gas produced by the biological reactions. Nevertheless, they can generally be considered to be two-phase for modeling purposes as long as the gas flow rate is small relative to the liquid flow rate.[13] This restriction may not be met for systems that are very heavily loaded, however, and thus care should be exercised in applying two-phase models to such FBBRs.

Because the bioparticles are retained in the reactor, the effluent from an FBBR often contains a sufficiently low suspended solids concentration to allow its discharge without clarification. Maintenance of the appropriate velocity to achieve the desired degree of suspension usually requires recirculation of bioreactor effluent. As biomass grows, the bioparticles become larger, causing the bed to expand in height. To prevent uncontrolled bed expansion, leading to loss of the bioparticles in the effluent, they are usually removed in a systematic manner to maintain a desired bed height. If the bioparticles contain a carrier particle, the excess biomass is removed in a separator, allowing the carrier particles to be returned. In this way a constant quantity of biomass can be kept in the system while maintaining an effluent low in suspended solids.

In general, FBBRs can be divided into two categories, depending on the nature of the bioparticles.[1] Tower bioreactors are those in which the bioparticles are composed entirely of biomass without a carrier particle at the center (i.e., granules), whereas supported-film bioreactors are those in which the biomass grows as a film on a carrier particle like sand, anthracite, or activated carbon. The sorptive properties of activated carbon provide distinct advantages in some cases, but complicate the analysis of FBBRs that use it as the carrier particles. Consequently, its use is not discussed in this chapter. Rather, the reader should consult other sources for more information.[33] Distinction between the two types of FBBRs is necessary because the presence or absence of a carrier particle has a strong influence on the way the bioparticles behave as they grow larger, as we will see later. The UASB reactor is an important example of a tower bioreactor. In fact, the majority of the models that have been developed for FBBRs originate from studies using these systems.[27] In it the bioparticles grow as small spherical granules containing the complex microbial community associated with methanogenic systems. Most other FBBRs are supported-film bioreactors and, in fact, the two terms are usually used synonymously, as we will do here. Our discussion here is limited to supported-film bioreactors.

The main advantage of FBBRs over other attached growth bioreactors is that the small size of the carrier particles provides a very large specific surface area for biomass growth. Whereas the media in packed towers and rotating disk bioreactors have specific surface areas on the order of 100 m^2/m^3 (see Table 19.2 and Section 20.1.1), the specific surface area provided by typical carrier particles in an FBBR is on the order of 1000 to 3000.[22,30] This allows the maintenance of very high biomass concentrations, ranging from 15,000 mg/L in aerobic FBBRs to 40,000 mg/L in anoxic ones.[22,30] This, in turn, allows very short hydraulic residence times to be used, often on the order of minutes. While packed towers could theoretically contain media of similar size, the downward flow of liquid would cause excessive pressure drop and be prone to clogging as biomass growth occurred. It would also be subject to plugging through entrapment of suspended solids that might enter in the influent. Such solids can pass through FBBRs, however, because of the open structure of the fluidized bed. More information about the characteristics of these unique bioreactors can be found elsewhere.[3,8,19]

18.1.2 Nature of the Biofilm

In Chapter 16 we discussed the concept of a steady-state biofilm. In an FBBR, development of a steady-state biofilm would require the excess biomass to be continually removed from the bioparticle surface and carried away in the effluent. It would also require the bed height to be sufficiently large to accommodate the quantity of biomass associated with the steady-state biofilm. Neither of these requirements is particularly desirable in practice. As noted above, one advantage of the FBBR is that it can be operated in a way that eliminates the need for a final clarifier. This would not be possible if biomass were constantly being sheared from the bioparticles in the FBBR. Furthermore, the thickness associated with a steady-state biofilm is likely to exceed the active thickness of the biofilm[1] (i.e., the depth to which reactants penetrate). This means that the bed height and volume associated with a steady-state biofilm would be greater than that required to achieve the desired effluent quality with a fully active biofilm. Consequently, one characteristic of most FBBRs is the continual wastage of biomass from the top of the bed to maintain a constant bed height less than that associated with a steady-state biofilm. Another complicating factor is that wastage occurs from the top of the bed because that is where the bioparticles with the thickest biofilm reside. After the biomass on the removed bioparticles has been reduced by subjecting them to surface shear forces, the carrier particles are returned to the bed where they again serve as a support for biomass growth. Initially, the carrier particles fall to the bottom of the bed, but they migrate upward as biofilm builds up on them. (The reason for this behavior is explained later.) Consequently, the biofilm thickness on any individual bioparticle is continually changing, which differs from the assumptions associated with steady-state biofilms. Nevertheless, many models of FBBRs assume the existence of a steady-state biofilm to reduce computational complexity.[27] Although such models are very useful for understanding the major factors influencing the behavior of FBBRs, it should be recognized that they differ significantly from the characteristics of most operating FBBRs.[27]

One interesting attribute of FBBRs is that the dry density of the biofilm on a bioparticle depends on the thickness of that biofilm.[10,22,30] Biofilm dry density is defined as the attached dry biomass per unit wet biofilm volume and is the same as the biomass concentration in a biofilm, $X_{B,Hf}$, used in Chapter 16. However, the term dry density, and its associated symbol, ρ_{fd}, is used here to make clear the distinction between the amount of biomass per unit volume of biofilm and the amount per unit volume of bioreactor, X_B. The exact relationship between density and thickness varies from study to study, but Figure 18.2[30] shows one that has been used in FBBR modeling. An important characteristic is that dry densities are very high (on the order of 70 g/L) for thin biofilms (on the order of 200 μm or less), but decrease markedly as the biofilm grows thicker. Although the reason for this behavior is poorly understood, it is probably related to the activity of the biofilm. Thin biofilms tend to be fully penetrated by both electron donor and electron

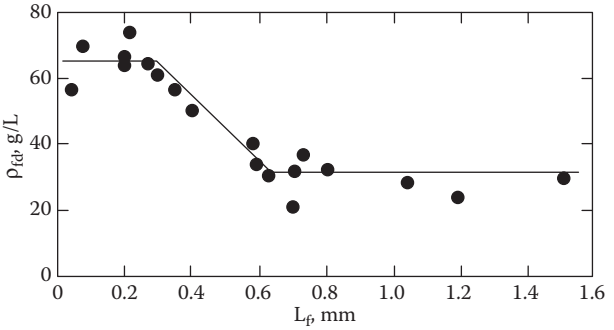

FIGURE 18.2 Effect of biofilm thickness, L_f, on the dry density, ρ_{fd}, of a denitrifying biofilm. (Reprinted from Shieh, W. K. and Keenan, J. D., Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology,* 33:131–69, 1986. Copyright © Springer-Verlag New York, Inc. With permission.)

acceptor, whereas the interior of thick biofilms is devoid of one or both of these substances. The resulting environment leads to microbial reactions that can decrease the quantity of viable microbial cells, thereby decreasing the density. Another possibility is that the hydrodynamic conditions that lead to thin biofilms bring about a morphological change in the biofilm that make it denser.[7] Regardless of the mechanism, however, it is clear from this behavior that thicker biofilms do not necessarily lead to more active biomass. This is one reason that FBBRs are commonly operated to give thin biofilms.

The hydrodynamic conditions required to give a thin biofilm are fairly complex and somewhat counterintuitive. Thinner biofilms develop in systems that have higher first-order detachment rate coefficients, b_D.[15] While at first glance it might appear that b_D will increase whenever the upward velocity of the fluid (superficial velocity) is increased, this is not the case. Generally, the superficial velocities used in FBBRs result in low Reynolds numbers (<10), which means that surface shear is likely to be small.[30] Lower superficial velocities, however, result in a smaller degree of bed expansion, which means that there is a higher probability of collisions among particles. The attrition caused by these collisions has a larger effect on the detachment coefficient than the fluid velocity past the biofilm surface.[13,30] Consequently, higher values of b_D have been observed at lower superficial velocities.[12] Empirical models are available that account for the various factors influencing the detachment rate coefficient.[7]

18.2 FLUIDIZATION

It is apparent from the preceding that the hydrodynamic conditions in a fluidized bed influence the biological characteristics of the bioreactor in important ways. In addition, they also define its physical characteristics, which influence its performance. Consequently, it is important to have a clear understanding of fluidization.

18.2.1 Fluidization of Clean Media[9,18]

Consider a cylindrical vessel like that in Figure 18.1 in which water is flowing upward through a bed of small, biomass-free, spherical carrier particles of equal diameter. As the superficial upflow velocity of the water (equivalent to the total hydraulic loading in a packed tower, defined in Equation 17.12) is increased, the pressure drop through the bed (i.e., the frictional forces acting on the particles) will increase, but the height of the bed will remain constant as shown in Figure 18.3. Ultimately a point will be reached (A) at which the pressure drop through the bed just counterbalances the force of gravity on the particles (taking into consideration the buoyant force of the displaced fluid) and the

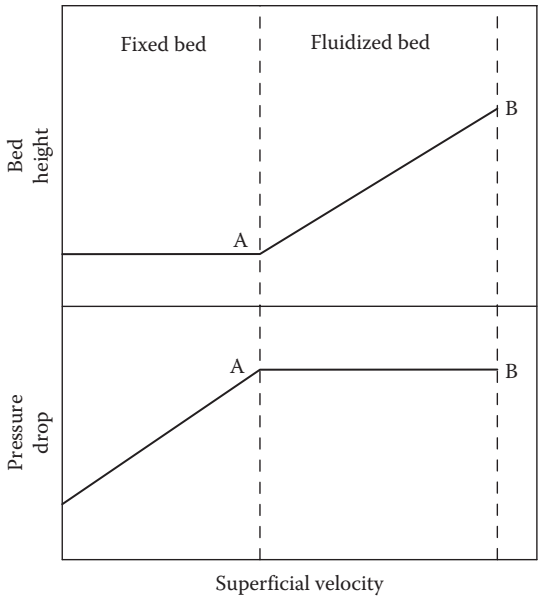

FIGURE 18.3 Idealized effect of superficial velocity on the pressure drop and bed height of a fluidized bed.

particles begin to move. As the superficial velocity is increased further, the particles will move apart so that the frictional forces associated with the local velocity past the particles continue to counterbalance the force of gravity on the particles. As a consequence, the pressure drop through the bed remains constant while the porosity (ε) of the bed increases, causing the bed height to increase. Any particle that rises above the top of the bed due to transient nonuniformities in flow will encounter a local upward velocity that is less than the particle's terminal settling velocity and the particle will fall back into the bed. Ultimately, a point (B) will be reached at which the local fluid velocity around the particles is equal to their terminal settling velocity. If the superficial velocity is increased beyond that point, the upward drag forces on the particles will exceed the downward gravitational forces and the particles will be carried away and the bed will cease to exist. It is then in continuous fluidization.

Consideration of Figure 18.3 suggests that two velocities are critical to defining the operating range for a fluidized bed, those associated with points A and B. The velocity associated with point A is called the minimum fluidization velocity. The formula for its computation can be derived by equating the pressure drop as given by the Ergun equation to the weight of the bed per unit area of cross section, allowing for the buoyant force of the displaced water. This results in a quadratic equation for the minimum fluidization velocity, v_{mf}:

$$\frac{1.75\,\rho_w v_{mf}^2}{d_p}\frac{1}{\varepsilon_M^3}+\frac{150\,\mu_w v_{mf}}{d_p^2}\frac{1-\varepsilon_M}{\varepsilon_M^3}-g(\rho_p-\rho_w)=0, \tag{18.1}$$

in which d_p is the diameter of the particle, ρ_p is its density, ρ_w is the density of water, μ_w is the viscosity of water, g is gravitational acceleration, and ε_M is the minimum porosity at incipient fluidization (i.e., point A). For roughly spherical particles, ε_M is generally between 0.40 and 0.45, decreasing slightly with increasing particle size. The velocity associated with point B is the terminal settling velocity of the carrier particles, v_t. If it is exceeded, the particles are carried away in continuous

fluidization and the bed is destroyed. The equation for its computation is derived by equating the drag force on the particle to the gravitational force minus the buoyant force. The result is

$$v_t = \left[\frac{4g(\rho_p - \rho_w)d_p}{3C_D\rho_w}\right]^{0.5}, \tag{18.2}$$

in which C_D is the drag coefficient. The value of C_D depends on the Reynolds number and correlations are available to relate the two for spherical particles as well as for other shapes. The appropriate velocity to use in computation of the Reynolds number is the terminal settling velocity of the particle. The result is usually called the terminal Reynolds number, Re_t. Because the terminal settling velocity depends on the Reynolds number (through the drag coefficient) and the Reynolds number depends on the velocity, an iterative procedure may be required to compute the terminal settling velocity, depending on the nature of the C_D versus Re_t relationship. Techniques are available for defining ranges of Reynolds numbers over which direct solutions may be possible and the reader should consult other sources to learn more about them.[9,18]

The height that a fluidized bed of clean carrier particles attains (H_{Bp}) depends directly on the porosity that results from the applied superficial velocity. This follows from the fact that the mass of particles in the bed is constant. Therefore:

$$H_{Bp} = H_{Rp}\left(\frac{1-\varepsilon_R}{1-\varepsilon}\right), \tag{18.3}$$

where H_{Rp} is a reference bed height and ε_R is the porosity associated with it. Some use the minimum bed height immediately prior to fluidization as H_{Rp}, in which case, ε_R will be ε_M.[9] Others[23,31] avoid the need to know ε_M by using as H_{Rp} the height that would be occupied by the carrier particles if they formed a solid block with mass equal to the total mass of carrier particles present, in which case ε_R would be zero and H_{Rp} would be given by

$$H_{Rp} = \frac{\mathbf{M}_p}{\rho_p A_c}, \tag{18.4}$$

where $\mathbf{M}_p$ is the mass of carrier particles and A_c is the cross-sectional area of the FBBR. Substitution of Equation 18.4 into Equation 18.3 gives:

$$H_{Bp} = \frac{\mathbf{M}_p}{\rho_p A_c(1-\varepsilon)}, \tag{18.5}$$

which can be used to calculate the height of clean carrier particles in a fluidized bed. However, regardless of which definition of H_{Rp} is used, prediction of the bed height associated with a given superficial velocity requires prediction of the porosity, ε. This can be done by using the Richardson-Zaki equation, which was developed for a bed of uniform size, hard, spherical particles:

$$\frac{v}{v_t} = \varepsilon^n, \tag{18.6}$$

where v is the applied superficial velocity and n is a coefficient. The value of n can be correlated with the Galileo number or with the Reynolds number calculated on the basis of the terminal

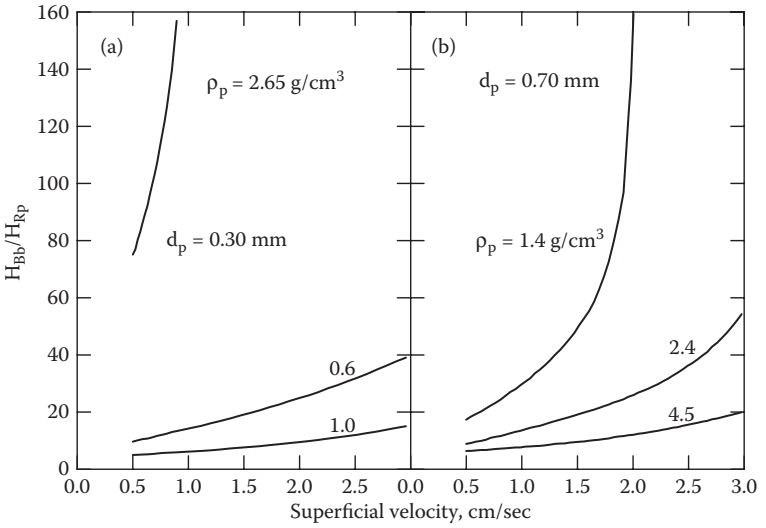

FIGURE 18.4 Effect of superficial velocity, v; carrier particle diameter, d_p; and carrier particle density, ρ_p; on the bed height of a fluidized bed, H_{Bb}; relative to the reference height of the carrier particles, H_{Rp}. The carrier particles are covered with a biofilm with a thickness of 100 μm and a dry density of 65 g/L. (Reprinted from Shieh, W. K. and Keenan, J. D., Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology,* 33:131–69, 1986. Copyright © Springer-Verlag New York, Inc. With permission.)

settling velocity. It typically takes on values between two and five for clean particles, with larger values being associated with smaller Galileo or Reynolds numbers (i.e., smaller particles). Several correlations are available.

Selection of the carrier particle for an FBBR is an important consideration because it has several influences, as we will see shortly. One important consideration is the stability of the bed to fluctuations in flow rate; that is, to fluctuations in superficial velocity. If the expanded bed height is overly sensitive to such fluctuations, small variations in flow rate might carry particles out in the effluent, thereby destroying the bed. Figure 18.4 illustrates how carrier particle size and density affect bed height over the range of superficial velocities commonly encountered.[30] The curves were obtained by simulation and the conditions employed are indicated in the figure. Those curves illustrate that the stability of a bed increases with increasing carrier particle size and density. Common silica sand has a density of around 2.65 g/cm^3, which provides reasonable stability over a broad range of superficial velocities.

18.2.2 Effects of Biomass on Fluidization

The growth of biofilm on the carrier particles changes their fluidization characteristics. This is due to three things. First, growth of the biofilm will change the size of the particle. Second, unless the carrier particle has a density equivalent to the wet density of the biofilm, growth of the biofilm will change the overall effective density of the particle. Third, the surface properties of the biofilm will differ from those of the clean particle, thereby changing the relationship between the drag coefficient, C_D, and the Reynolds number. In addition, growth of the biofilm may change the sphericity of the particle, but that effect has been found to be small,[26] and is not considered further here. Rather, we assume spherical particles.

18.2.2.1 Terminal Settling Velocity

Examination of Equation 18.2 shows that the terminal settling velocity of a particle depends on its diameter, its density, and the drag coefficient. Since all of those characteristics are altered by

growth of a biofilm, it becomes clear that biofilm growth changes the terminal settling velocity. The influence of biofilm thickness, L_f, on bioparticle diameter, d_b, is very straightforward:

$$d_b = d_p + 2L_f. \tag{18.7}$$

The influence of the biofilm growth on the overall effective density of the bioparticle, ρ_b, depends on the density of the carrier particle, ρ_p, and the wet density of the biofilm, ρ_{fw}, as well as the relative volumes occupied by the carrier particle and the biofilm.[13,24,32] Since the volume of a sphere is proportional to its diameter cubed, the relationship is[32]

$$\rho_b = \rho_p \left(\frac{d_p}{d_b}\right)^3 + \rho_{fw}\left[1 - \left(\frac{d_p}{d_b}\right)^3\right]. \tag{18.8}$$

The biofilm wet density is related to its dry density, ρ_{fd}, and the weight fraction moisture content of the biofilm, P':[32]

$$\rho_{fw} = \frac{\rho_{fd}}{1 - P'}. \tag{18.9}$$

The value of the moisture content has been found to be approximately 0.93 over a broad range of biofilm thicknesses.[32] We saw earlier that the biofilm dry density depends on its thickness, and thus the biofilm wet density also varies with biofilm thickness. Therefore, it is not surprising that a number of investigators have reported different values for the wet biofilm density.[24] Nevertheless, a value of 1.1 g/cm^3 has been assumed to be typical of biomass.[24]

The influence of biofilm growth on the drag coefficient has been studied by several investigators.[11,21,24,26] All correlate the drag coefficient to the terminal Reynolds number, Re_t, using an expression of the type:

$$C_D = a\,Re_t^{-b}. \tag{18.10}$$

Figure 18.5[26] shows three relationships and compares them to the relationship of Schiller et al. (see Ref. 26) for clean spherical particles. The equations are given in Table 18.1. Two things are evident from the figure. First, the growth of a biofilm increases the drag coefficient relative to that of a clean particle with equivalent terminal Reynolds number (i.e., equivalent diameter and density). Second, the relationships found by the three studies on biofilms are all different, suggesting that the influence of biofilm growth on C_D may be case specific.

Figure 18.6[26] shows the effect of biofilm growth on the terminal settling velocity of bioparticles in which sand (ρ_p = 2.65 g/cm^3) with a diameter of 0.5 mm serves as the carrier particle. The values were calculated with Equation 18.2 using the bioparticle density, ρ_b, from Equation 18.8 in place of ρ_p and the bioparticle diameter, d_b, from Equation 18.7 in place of d_p. The value of C_D was computed from the correlation of Ro and Neethling[26] shown in Figure 18.5. Three important points are evident in the figure. First, the terminal settling velocity of the bioparticles decreases as the biofilm thickness increases. Since terminal settling velocity is directly proportional to the diameter of a particle (see Equation 18.2), the decrease in terminal settling velocity associated with an increase in biofilm thickness is due to the decrease in the effective density of the bioparticle (see Equation 18.8). Second, as the biofilm thickness increases, a point is eventually reached at which further increases have little effect. In that region the effects of increases in diameter are approximately equal to the effects of decreases in density. Third, the settling velocity of a bioparticle is always lower than that

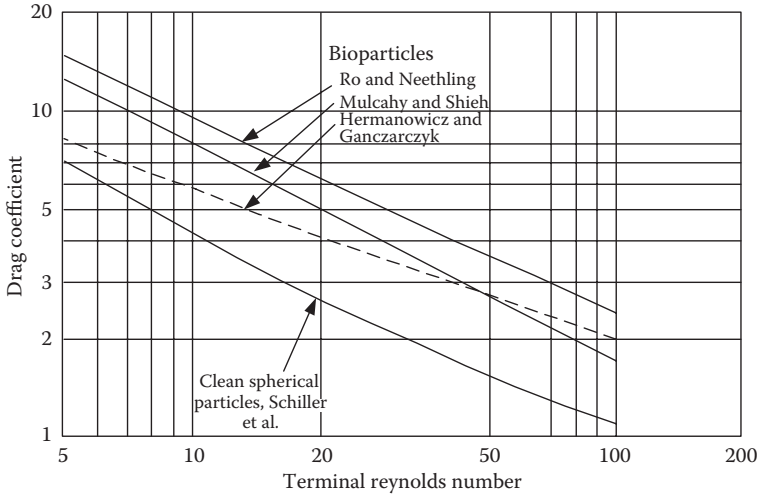

FIGURE 18.5 Effect of terminal Reynolds number, Re_t, on the drag coefficient, C_D, of bioparticles. (Reprinted from Ro, K. S., and Neethling, J. B., Terminal settling characteristics of bioparticles. *Research Journal, Water Pollution Control Federation,* 62:901–6, 1990. Copyright © Water Environment Federation, Alexandria, Virginia. With permission.)

TABLE 18.1
Equations Depicting the Relationships between the Drag Coefficient and the Terminal Settling Velocity of Bioparticles Shown in Figure 18.5

Equation	Applicable Range	Source
$C_D = 17.1\ Re_t^{-0.47}$	$50 < Re_t < 100$	Hermanowicz and Ganczarczyk[11]
$C_D = 36.66\ Re_t^{-0.667}$	$40 < Re_t < 90$	Mulcahy and Shieh[21]
$C_D = 24\ Re_t^{-1.0} + 21.55\ Re_t^{-0.518}$	$15 < Re_t < 87$	Ro and Neethling[26]

of a smooth sphere of equivalent diameter and density. This is due to the effect of the biofilm on the drag coefficient. The latter point is true for a wide range of carrier particle sizes and densities, as well as for a broad range of biofilm thicknesses, with the effect that the settling velocity of a bioparticle is always between 55 and 60% of the velocity of an equivalent density smooth sphere of the same diameter.[26]

The effects of particle density are shown in Figure 18.7[24] for a case in which the growth of the biofilm has no effect on the relationship between the drag coefficient and the terminal Reynolds number. In other words, it assumes that clean carrier particles have the same surface characteristics as those with biofilm. There it can be seen that biofilm growth can increase the settling velocity of carrier particles of low density. In fact, the counteracting effects of the changes in density and diameter can make the settling velocity of bioparticles containing low density carrier particles change in complex ways as they grow larger, particularly when the effects on the drag coefficient are also considered. This can have a significant effect on the migration of bioparticles in FBBRs.

18.2.2.2 Bed Porosity and Expansion

Because growth of a biofilm changes the terminal settling velocity of a particle, it also changes its fluidization properties. One effect is on the porosity associated with a given superficial velocity. According to the Richardson-Zaki equation (Equation 18.6) if the superficial velocity is held constant and the terminal settling velocity of a particle is changed, the porosity of the bed will change. This, in turn will change the height of the fluidized bed, as indicated by Equation 18.3.

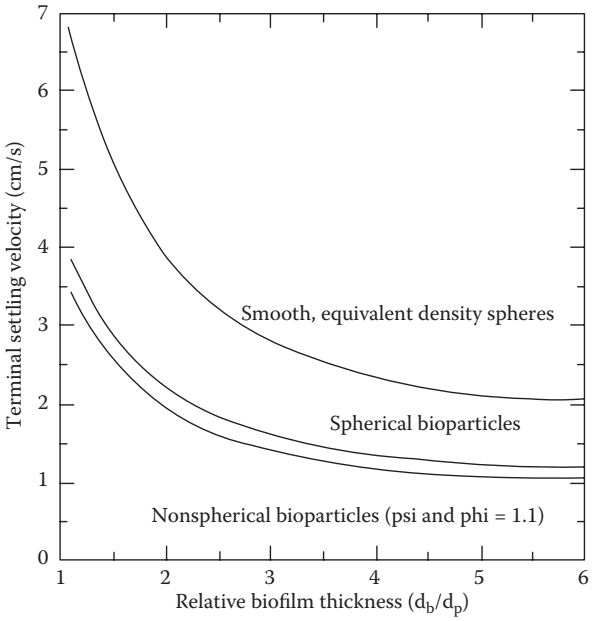

FIGURE 18.6 Effect of relative biofilm thickness, d_b/d_p, on the terminal settling velocity of spherical and nonspherical bioparticles with sand carrier particles with a diameter, d_p, of 0.5 mm. For comparison, the terminal settling velocity of smooth, equivalent density spheres is also shown. (Reprinted from Ro, K. S. and Neethling, J. B., Terminal settling characteristics of bioparticles. *Research Journal, Water Pollution Control Federation,* 62:901–6, 1990. Copyright © Water Environment Federation, Alexandria, Virginia. With permission.)

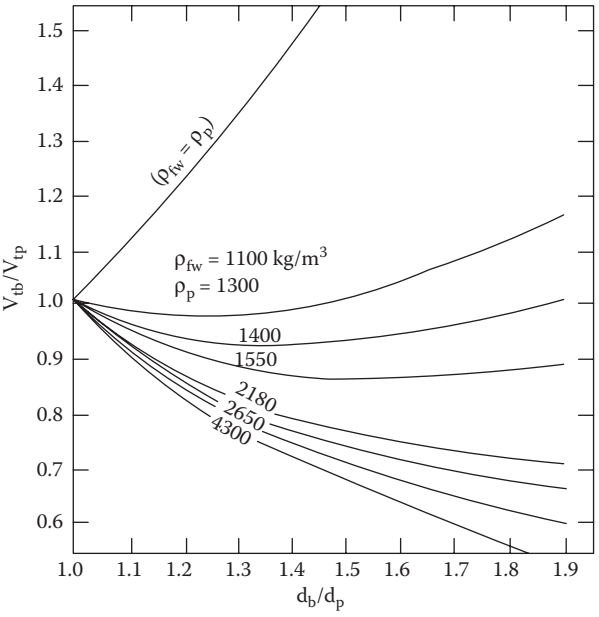

FIGURE 18.7 Effect of relative biofilm thickness, d_b/d_p, and carrier particle density, ρ_p, on the terminal settling velocity of a bioparticle, v_{tb}, relative to the terminal settling velocity of its carrier particle, v_{tp}. (Reprinted from Myška, J. and Švec, J., The distributive properties of a fluidized bed with biomass. *Water Research,* 28:1653–58, 1994. Copyright © Elsevier Science Ltd. With permission.)

Another effect is on the reference bed height. Particles with a larger diameter occupy more space. Thus, the reference bed height will be larger, which will cause the expanded bed height to increase as well. Because the volume of a sphere is proportional to its diameter cubed, the value of the reference bed height for bioparticles, H_{Rb}, can be related to the mass of carrier particles present by[23,31]

$$H_{Rb} = \frac{\mathbf{M}_p}{\rho_p A_c}\left(\frac{d_b}{d_p}\right)^3. \tag{18.11}$$

This equation can be substituted into Equation 18.3 to give the height of a fluidized bed containing bioparticles, H_{Bb}:

$$H_{Bb} = \frac{\mathbf{M}_p}{\rho_p A_c (1-\varepsilon)}\left(\frac{d_b}{d_p}\right)^3. \tag{18.12}$$

The Richardson-Zaki equation (Equation 18.6) has been shown to be applicable to bioparticles, although the correlation between the coefficient n and the Reynolds or Galileo number is different from that for clean particles.[21,29,34] One that works well for a broad range of particle sizes and densities is that of Shieh and Chen,[29] which was developed from the data of Mulcahy and LaMotta:[20]

$$n = 47.36\ Ga^{-0.2576} \qquad 1000 < Ga < 15{,}000, \tag{18.13}$$

where Ga is the Galileo number, given by

$$Ga = \frac{d_b^3 \rho_w (\rho_b - \rho_w) g}{\mu_w^2}. \tag{18.14}$$

The approach above has been used to demonstrate, through modeling, the effect of biofilm thickness on the degree of expansion of an FBBR containing sand with a diameter of 0.4 mm as the carrier particle.[30] The results are shown in Figure 18.8 in which the expanded bed height has been normalized relative to H_{Rp} as computed with Equation 18.4. There it can be seen that even thin biofilms have a strong effect on the height of a fluidized bed. Consequently, during design, careful consideration must be given to the configuration of an FBBR to ensure that it is capable of containing the desired amount of media once a biofilm of the desired thickness has developed.

18.2.2.3 Solids Mixing

The movement of particles in a fluidized bed is a very complex subject that is incompletely understood.[13] In fact, the circumstances and assumptions associated with an analysis of mixing strongly influence the conclusions reached. Nevertheless, it is important to understand the basic forces at work in a fluidized bed as biofilm grows.

Andrews[1] has presented a very thorough analysis of the factors influencing solids mixing. First, it must be recognized that there are two counteracting tendencies affecting particle movement. One is the tendency of fluidized particles to move randomly, which is a disordering tendency. The other is caused by the development of particles of different size due to biofilm growth. If the terminal settling velocities of the various particles are not all the same, the bed tends to stratify, with rapidly settling particles near the bottom and slowly settling ones near the top. This is an ordering tendency, but whether such a tendency is stable depends on the density of the carrier particles.

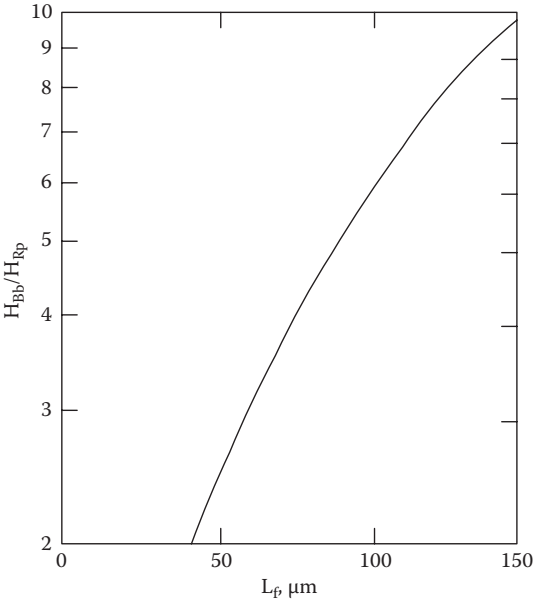

FIGURE 18.8 Effect of biofilm thickness, L_f, on the height of a fluidized bed, H_{Bb}, relative to the reference height of the carrier particles, H_{Rp}. The following conditions were assumed: $d_p = 0.4$ mm, $\rho_p = 2.65$ g/cm^3, $v = 1$ cm/sec, $\rho_{fd} = 65$ g/L, $P' = 0.93$. (Reprinted from Shieh, W. K. and Keenan, J. D., Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology,* 33:131–69, 1986. Copyright © Springer-Verlag New York, Inc. With permission.)

Bioparticles containing carrier particles of low density, similar to the wet density of the biofilm, tend to stratify because the density does not change significantly as the biofilm grows. Only the diameter changes. The same is true for bioparticles without carrier particles, such as UASB granules. In that case, larger particles have a higher settling velocity, causing them to move to the bottom of the bed, where they are exposed to more substrate, causing them to grow even larger. Conversely, smaller particles move to the top, where they are exposed to less substrate, which causes the biofilm to grow more slowly, or even decrease in size because of decay and surface shear. This leads to stratification of the bed, with the possible development of bioparticle sizes well in excess of the optimal, thereby increasing the quantity of inactive biomass in the bed. Under such a situation, biomass wastage should be done from the bottom of the bed.

Bioparticles containing carrier particles of high density, on the other hand, tend to form well-mixed beds, although a degree of stratification can be induced. With high density carrier particles, the settling velocity of the bioparticles decreases as the thickness of the biofilm increases. As a consequence, larger bioparticles tend to move to the top of the bed. Once there, however, they receive less substrate, which causes them to decrease in size, thereby allowing them to move downward into a region of higher substrate concentration. Bioparticles with thin biofilms, on the other hand, move toward the bottom of the bed, where they are exposed to high substrate concentrations, causing more rapid growth and an increase in size. The resulting situation is unstable, inducing motion within the bed, leading ultimately to a relatively uniform bioparticle size throughout the bed.

Control of bed height in an FBBR requires continual wastage of biomass. Otherwise, nonoptimal sized bioparticles develop and the bed height becomes very large.[1] Common practice is to waste biomass from the top of a bed containing high density carrier particles. This induces stratification in the bed because the size of any individual bioparticle is continually changing, preventing the development of a steady-state biofilm. By continually wasting large bioparticles from the top and

returning clean carrier particles, which then migrate to the bottom where they are exposed to high substrate concentrations, the bed is maintained in a dynamic state. Consequently, stratification of such beds is a common occurrence.[2,25] The above analysis is based on the assumption of a uniform carrier particle size. If there are significant differences in carrier particle size, the bed tends to stratify based on carrier particle size rather than bioparticle size.[2] As a consequence, larger support particles tend to stay at the bottom where they accumulate biofilm beyond the optimum thickness, while smaller carrier particles migrate to the top from where they can be ineffectually cycled through the biomass wastage device. Consequently, it is important for FBBRs to have a uniform carrier particle size.

18.2.3 Relationship between Fluidization and Biomass Quantity

It is clear from the preceding that there is a complex relationship between the fluidization regime imposed on an FBBR and the quantity of biomass that may be present. Consequently, it is difficult to intuitively reason out the relationship. Luckily, however, it is a straightforward task to calculate the biofilm thickness that would be associated with a given fluidization regime, provided that sufficient substrate is supplied to maintain that biofilm. If that thickness can be assumed to be representative of the average thickness that could be maintained in an FBBR with a given fluidization regime, then the biomass concentration can be calculated.[23,30–32]

An iterative approach must be used to calculate the biofilm thickness that can be maintained in an FBBR. Figure 18.9 summarizes an approach based on that of Shieh and Keenan.[30] First, the characteristics of the FBBR must be established, including the desired superficial velocity, v; the FBBR cross-sectional area, A_c; the desired fluidized bed height, H_{Bb}; the mass of carrier particles, $\mathbf{M}_p$; their diameter, d_p; and their density, ρ_p. In addition, the properties of the fluid such as its density and viscosity must be established, as should the biofilm moisture content, P′. The computation begins by assuming a biofilm thickness. The assumed value is given the symbol L_{fa} to denote it as an assumed value. The biofilm dry density associated with the assumed biofilm thickness can be determined from information such as that in Figure 18.2 or an appropriate empirical equation, allowing the biofilm wet density to be calculated with Equation 18.9. The bioparticle diameter can be calculated with Equation 18.7 and that, in turn, can be used to calculate the bioparticle density with Equation 18.8. The terminal settling velocity of the bioparticle can then be calculated with Equation 18.2 (substituting d_b for d_p and ρ_b for ρ_p) using a relationship for C_D such as one of the ones in Figure 18.5 as expressed with Equation 18.10. The coefficient n in the Richardson-Zaki equation can then be estimated with Equation 18.13, allowing the porosity of the fluidized bed to be calculated with Equation 18.6. It can then be used to calculate the bed height. The calculated value is denoted as H_{Bbc}. The value of H_{Bbc} is then compared to the desired bed height used to begin the computations. If they are equal, then the assumed biofilm thickness is correct and can be taken as the true thickness, L_f. If $H_{Bbc} > H_{Bb}$, then the assumed biofilm thickness is too large and a new smaller value should be assumed for repeating the computations. Conversely, if $H_{Bbc} < H_{Bb}$, a larger biofilm thickness should be assumed. Finally, once the correct biofilm thickness has been found, the concentration of biomass per unit volume of fluidized bed, X_B, can be calculated with

$$X_B = \frac{\rho_{fd}}{\rho_p} \frac{\mathbf{M}_p}{A_c H_{Bb}} \left[\left(\frac{d_b}{d_p} \right)^3 - 1 \right]. \tag{18.15}$$

The procedure illustrated in Figure 18.9 can be used to investigate the effect of biofilm thickness on the biomass concentration in an FBBR. The result of such an exercise is shown in Figure 18.10[30] for sand as the carrier particle with a diameter of 0.4 mm. The biomass dry density was assumed to be

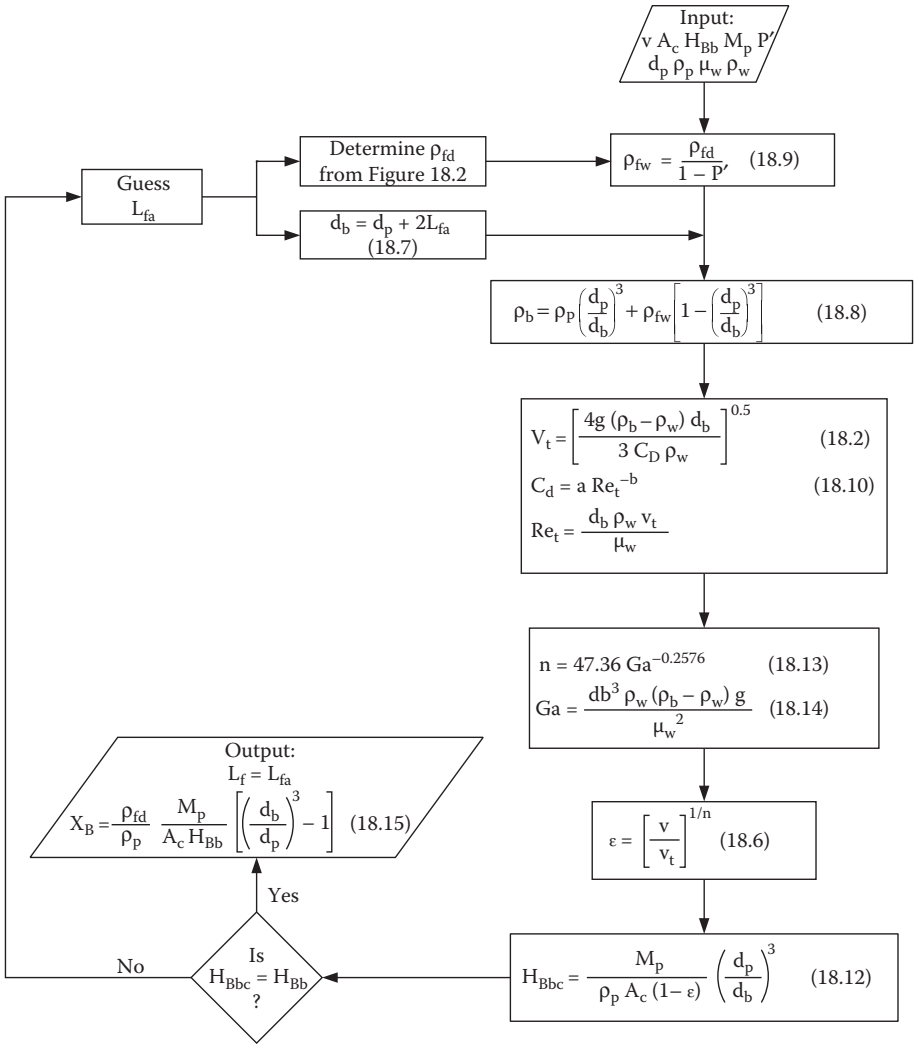

FIGURE 18.9 Algorithm for computation of the biofilm thickness associated with a fluidized bed.

constant with a value of 65 g/L, which is consistent with Figure 18.2 over the range of biofilm thicknesses considered. Examination of the figure reveals that a biofilm thickness of 100 μm maximizes the concentration of biomass. Beyond that thickness, the increase in biomass associated with a thicker biofilm is offset by the reduction in the number of particles per unit volume due to increased bed expansion. If the objective was to maximize biomass concentration, then a biofilm thickness of 100 μm should be chosen. However, it should be recognized that such a strategy may not optimize overall FBBR performance. To do that, consideration must also be given to the effectiveness of the biofilm.[30] We examine that question below.

It should also be recognized that Equation 18.15 simply determines the biomass concentration and biofilm thickness that could be maintained by the hydrodynamic conditions in the bed. It does not tell whether they can be supported by the substrate loading on the bioreactor or whether a desired effluent substrate concentration can be achieved. One way to estimate whether a desired film thickness can be supported is to compare it to a steady-state biofilm. Because we wish to maintain the biofilm in a dynamic state, the biofilm thickness in the FBBR must be equal to or less than the steady-state biofilm thickness. To calculate the steady-state biofilm thickness, assume that the

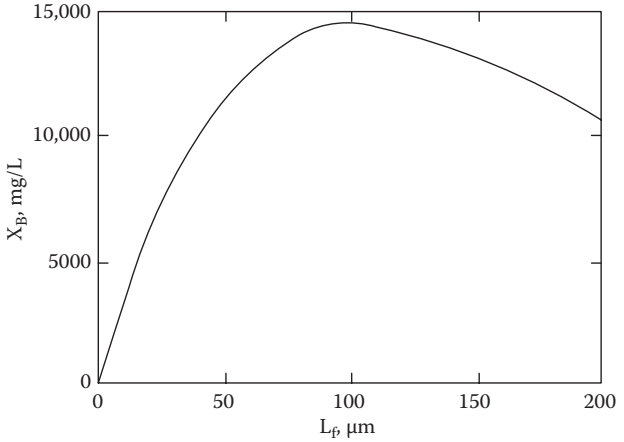

FIGURE 18.10 Effect of biofilm thickness, L_f, on the biomass concentration, X_B, in an FBBR. The following conditions were assumed: $d_p = 0.4$ mm, $\rho_p = 2.65$ g/cm^3, v = 1 cm/sec, $\rho_{fd} = 65$ g/L, $P' = 0.93$. (Reprinted from Shieh, W. K. and Keenan, J. D., Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology,* 33:131–69, 1986. Copyright © Springer-Verlag New York, Inc. With permission.)

FBBR is completely mixed and that all bioparticles are exposed to a substrate concentration that is the average of the applied (considering recirculation) and the desired effluent substrate concentrations. Knowing that concentration, the steady-state biofilm thickness can be determined using the pseudoanalytical approach, as illustrated in Example 16.2.3.1. As long as the calculated biofilm thickness is less than the steady-state value, it is possible to support that thickness. Whether the computed biomass concentration can achieve the desired effluent concentration requires application of an FBBR model incorporating the assumed mass of carrier particles supporting the desired biofilm thickness.

18.3 MODELING FLUIDIZED BED BIOLOGICAL REACTORS

The modeling of FBBR performance requires the integration of information from several sources.[23] First, a biofilm model must be available that gives the rate of substrate conversion by individual bioparticles. Such a model can be developed using one of the approaches presented in Chapter 16. The only major difference is the incorporation of spherical coordinates to account for the shape of the bioparticles. Second, a model must be available that accounts for the effects of fluidization on the biofilm thickness and the number of bioparticles per unit of fluidized bed volume that can be maintained by the hydrodynamic conditions imposed. Such a model was presented in Figure 18.9. Finally, one must have available an overall bed model that links the biofilm and fluidization models to yield substrate concentration as a function of axial position within the FBBR. The structure of this component model depends on the fluid regime within the FBBR.

Several FBBR models are available in the literature,[1,2,13,14,22,23,25,28,30,31] although many are variations of the same basic model. Most use an effectiveness factor approach for modeling transport and reaction within the biofilm and thus consider only a single limiting nutrient. In addition, most assume a uniform carrier particle size and consider the biofilm thickness to be uniform throughout the bed. In spite of these simplifications, the modeling of FBBRs is the most complex of all of the biochemical unit operations because of the interactions between the biofilm and fluidization submodels. Space does not allow us to consider all of the features of the models. Rather, only the major points are given and the reader is encouraged to consult the references for more details.

18.3.1 Biofilm Submodel

The biofilm submodel must consider simultaneous reaction and transport. Consequently, correlations must be available for relating the liquid phase mass transfer coefficient, k_L, to the hydraulic conditions in the bioreactor. Several have been proposed.[2,13,30] Shieh and Keenan[30] recommend the use of a correlation developed for fluidized beds from experimental data collected at Reynolds numbers within the range common to FBBR operation:

$$k_L = \frac{0.81}{\varepsilon}\left[\frac{D_w^{1.333} v \rho_w^{0.333}}{\mu_w^{0.333} d_b}\right], \tag{18.16}$$

where D_w is the diffusivity of the substrate in water and all other terms have been defined previously. This correlation reveals that a typical value of k_L is 0.01 cm/s for FBBR conditions, which some believe is sufficiently high to allow external mass transfer resistance to be ignored.[23] Consequently, to simplify computations, this is frequently done.[1,14,23,30] The most exact approach, however, would be to consider both internal and external mass transfer resistance as was done in Chapter 16.

The effectiveness factor approach is the most common method of handling simultaneous reaction and transport in the biofilm, although all of the approaches in Chapter 16 have been used. While models[13] are available that use effectiveness factors for intrinsic Monod kinetics through the use of relationships like that in Figure 16.9, more assume the limiting case of either zero-order ($S_{Sb} \gg K_S$) or first-order ($S_{Sb} < K_S$) intrinsic kinetics, where S_{Sb} is the bulk liquid phase substrate concentration and K_S is the Monod half-saturation coefficient. For zero-order kinetics in the absence of external mass transfer resistance, the effectiveness factor is defined as the biofilm volume containing substrate divided by the total biofilm volume.[23] Consequently, when the bioparticle is fully penetrated with substrate, the effectiveness factor has a value of 1.0. For first-order kinetics in the absence of external mass transfer resistance, the effectiveness factor has been defined as the substrate flux into the spherical bioparticle divided by the intrinsic rate when all of the biofilm is surrounded by substrate at the bulk substrate concentration.[23]

For both first- and zero-order kinetics in the absence of external mass transfer resistance, the effectiveness factor can be correlated with an appropriate Thiele modulus. Furthermore, the correlations can be reduced to single curves for each type of kinetics by using an appropriate characteristic biofilm thickness, L_{fc}, which is defined as

$$L_{fc} = \frac{\text{biofilm volume}}{\text{biofilm exterior surface area}} \tag{18.17}$$

and

$$L_{fc} = \frac{d_b^3 - d_p^3}{6d_b^2}. \tag{18.18}$$

For zero-order kinetics, the zero-order effectiveness factor, η_{eZ}, can be correlated with a modified zero-order Thiele modulus, ϕ_{zm}, defined as[30]

$$\phi_{Zm} = L_{fc}\left(\frac{\rho_{fd}\hat{q}_H}{D_e S_{Sb}}\right)^{0.5}, \tag{18.19}$$

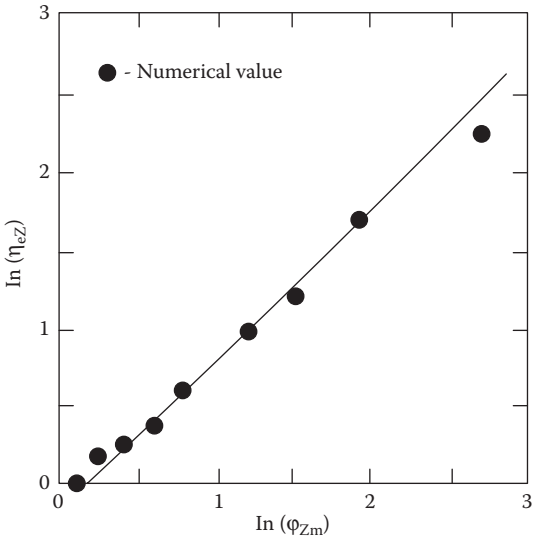

FIGURE 18.11 Relationship between the bioparticle zero-order effectiveness factor, η_{eZ}, and the modified zero-order Thiele modulus, ϕ_{zm}. (Reprinted from Shieh, W. K. and Keenan, J. D., Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology,* 33:131–69, 1986. Copyright © Springer-Verlag New York, Inc. With permission.)

where $\hat{q}_H$ is the maximum specific substrate removal rate and D_e is the effective diffusivity. The correlation is shown in Figure 18.11, and is described well by the following empirical equation:[30]

$$\eta_{eZ} = \frac{1.2712}{\phi_{Zm}}. \tag{18.20}$$

For first-order kinetics, the first-order effectiveness factor, η_{e1}, can be correlated with a modified first-order Thiele modulus, ϕ_{1m}, defined by Equation 16.12 except for the use of the characteristic film thickness, $L_{fc:}$[30]

$$\phi_{1m} = L_{fc}\left(\frac{\rho_{fd}\hat{q}_H}{K_S D_e}\right)^{0.5}. \tag{18.21}$$

Recall that ρ_{fd} and $X_{B,Hf}$ are the same. It should be noted that $\hat{q}_H/K_S$ is equivalent to k_e, the mean reaction rate coefficient defined by Equation 3.45. The correlation between η_{e1} and ϕ_{1m} is shown in Figure 18.12, and is described well by the first-order, nonspherical form for homogeneous reaction media proposed by Aris:[5]

$$\eta_{e1} = \frac{\coth(3\phi_{1m})}{\phi_{1m}} - \frac{1}{3\,\phi_{1m}^2}. \tag{18.22}$$

The points in the figures were computed for a variety of bioparticle and carrier particle sizes, thereby demonstrating that the characteristic film thickness works well as a normalizing factor.

Because the characteristic biofilm thickness serves as a normalizing factor that allows effectiveness factor correlations developed for planar coordinates to be used with spherical particles, it should be possible to use the general correlation for Monod kinetics shown in Figure 16.9 by using an appropriately modified Thiele modulus. This would allow external mass transfer resistance to be

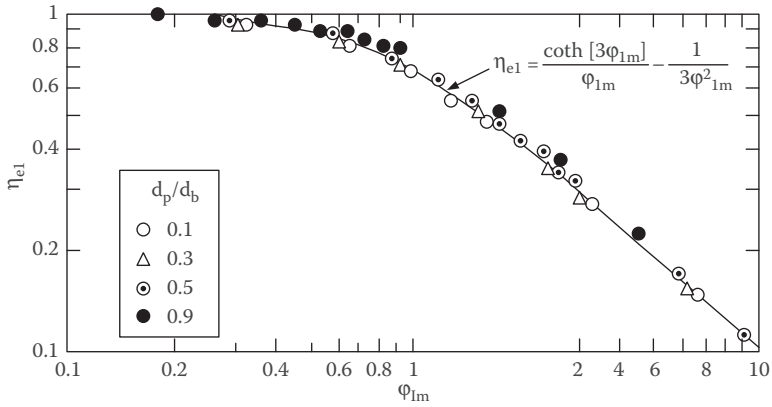

FIGURE 18.12 Relationship between the bioparticle first-order effectiveness factor, η_{e1}, and the modified first-order Thiele modulus, ϕ_{1m}. (Reprinted from Shieh, W. K. and Keenan, J. D., Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology,* 33:131–69, 1986. Copyright © Springer-Verlag New York, Inc. With permission.)

handled with little additional effort. Therefore, depending on the kinetic and mass transfer characteristics of the system, Equation 18.20, Equation 18.22, or Figure 16.9 can be used to calculate the substrate removal rate by bioparticles surrounded by substrate at a given concentration. This information can then be used to calculate system performance in the same manner as used in Chapters 16 and 17.

Another approach used in Chapter 16 for determining substrate removal rates by biofilms is the pseudoanalytical approach with a steady-state biofilm. The assumption of a steady-state biofilm is consistent with the situation encountered in packed towers and rotating disk reactors, but is inconsistent with FBBRs from which bioparticles are constantly wasted, as discussed previously. Thus, while FBBR models are available that assume steady-state biofilms, one must question their relevance to most operating FBBRs.[27] As a consequence, steady-state biofilm FBBR models are not discussed here.

18.3.2 Fluidization Submodel

The effect of the fluidization regime on the thickness of biofilm that can be maintained on carrier particles of a given size and density constrained within a bed of fixed height was discussed in Section 18.2.3. Figure 18.9, shown there, presents the algorithm for calculating that biofilm thickness, and as such, represents a fluidization submodel that can be used. No further discussion of it is needed here. However, it should be noted that such a model assumes a uniform biofilm thickness throughout the FBBR, which may not conform to reality for some FBBRs, as pointed out previously. Other models[1,25] are capable of handling variations in particle size within the bed, but because of space constraints and their added complexity, they are not discussed here. Rather, the reader is referred to the cited papers.

18.3.3 Reactor Flow Submodel

The reactor flow submodel must link the biofilm and fluidization submodels to allow computation of the performance of an FBBR. The type of model that should be used depends on the hydraulic regime in the FBBR. The nature of that regime is determined primarily by the degree of substrate utilization across the bed and the amount of effluent recirculated to maintain the appropriate fluidization velocity and to provide the required amount of electron acceptor. If the recirculation ratio is high and the influent substrate concentration is low so that the change in substrate concentration

across the bed (after dilution of the influent by the recirculation flow) is small, then the bed can be considered to behave as if it were a completely mixed reactor. The validity of this approach can be checked easily by comparing the diluted influent concentration as calculated with Equation 17.1 to the assumed effluent concentration. On the other hand, if the recirculation ratio is low (e.g., <2) or the degree of dilution is small, the FBBR must be treated as a plug-flow reactor, either with or without axial dispersion, or as a series of continuous stirred tank reactors (CSTRs). For cases where the recirculation ratio approaches zero, as is usually the case in UASB bioreactors, the FBBR may have to be treated as a combination of CSTR(s) and plug-flow reactors to account for nonideal behavior,[35] such as the existence of dead zones and short-circuiting. However, consideration need not be given to changes in the degree of axial dispersion from point to point in an FBBR due to differences in the porosity because that level of complexity cannot be justified.[1] Examples of all of these approaches can be found in the literature, depending on the situation being modeled. The equations used are typical of these various flow regimes as discussed in previous chapters and thus are not presented here.

Basically the approach to FBBR modeling is iterative, with the number of loops depending on the reactor flow submodel. Only a completely mixed FBBR (both bioparticles and liquid) is considered to describe the procedure, but the concepts can be extended to other flow regimes, which usually require more iterative loops. First, the characteristics of the FBBR must be established, including the desired superficial velocity, v; the FBBR cross-sectional area, A_c; the desired fluidized bed height, H_{Bb}; the mass of carrier particles, $\mathbf{M}_p$; their diameter, d_p; and their density, ρ_p. The biofilm thickness that can be maintained by these conditions can then be computed using the procedure in Figure 18.9. That thickness determines the bioparticle diameter, which defines the characteristic biofilm thickness, L_{fc}, which is used to determine the effectiveness factor in the biofilm submodel. If the effectiveness factor expression includes the bulk substrate concentration, then one must be assumed. It is equivalent to the effluent substrate concentration for a completely mixed FBBR. The biofilm submodel is then used in the reactor flow submodel to compute the output substrate concentration. This is a direct computation for a completely mixed FBBR, but an iterative procedure is required for a plug-flow or tanks-in-series flow regime. If the computed concentration is different from the assumed value, then a new value must be assumed and the procedure repeated until the computed effluent concentration agrees with the value assumed. This is the effluent substrate concentration from the FBBR. The entire procedure can be repeated for different initial conditions (i.e., v, $\mathbf{M}_p$, d_p, or ρ_p) thereby relating performance to those conditions. This allows identification of the conditions giving an effluent concentration equal to or less than some desired value.

18.4 THEORETICAL PERFORMANCE OF FLUIDIZED BED BIOLOGICAL REACTORS

We saw in Section 18.2 that for a given superficial velocity the expansion of a fluidized bed depends on the size and density of the carrier particles, as well as on the thickness of the biofilm. In addition, for a given degree of expansion (porosity), the number of particles per unit bed volume also depends on those factors. Consequently, the biomass concentration in the bed is influenced by them as well. The effect of biofilm thickness on the biomass concentration was illustrated in Figure 18.10, and similar figures could be generated illustrating that values for the carrier particle size and density that maximize the biomass concentrations also exist.[30,31] Thus, from consideration of the effects of fluidization alone, it can be seen that there are complex interactions among the factors that influence FBBR performance. Fluidization effects do not tell the whole story, however. Because of the need for transport of reactants into the biofilm, not all of the biomass has the same activity. That is why the effectiveness factor is less than 1.0. Furthermore, the effectiveness factor depends on the size of the bioparticle and the thickness of the biofilm, as reflected in the modified Thiele moduli as used in Figures 18.11 and 18.12. This

suggests that the combination of bioparticle characteristics that maximizes the ability of the FBBR to remove substrate is different from that which maximizes biomass concentration.[30,31] Because each situation is unique and complex, mathematical models are required for their analysis and several have been developed that integrate submodels of the type discussed in the preceding section.[1,2,4,13,23,25,28,30,31]

The theoretical performance of FBBRs can be examined with those models. The result from one such exercise is shown in Figure 18.13.[31] It shows the effect of particle diameter and biofilm thickness on the time required for 90% removal of substrate by biomass with an intrinsic zero-order reaction in an FBBR containing a fixed mass of carrier particles operated with a fixed superficial velocity. The flow regime in the FBBR was characterized as plug flow. Thus, the required reaction time corresponds to the fractional height in the bed at which 90% of the substrate is removed. In other words, it corresponds to a required bed height and media mass. Examination of the figure reveals that the optimal (smallest) reaction time (and therefore bed size) is associated with moderately thin biofilms growing on small carrier particles. In fact, others have shown that for a given carrier particle size, the bioparticle effectiveness factor is maximized when the biofilm thickness is slightly less than the thickness at which all substrate would be exhausted.[2,4] Consequently, the optimum biofilm thickness depends on the diffusivity of the substrate in the biofilm and the biodegradation kinetics. The benefit of small carrier particles derives directly from the fact that for a given mass of carrier particles, the surface area for biofilm growth increases as the carrier particle diameter decreases. Nevertheless, the curvature associated with the optimal region in Figure 18.13 is relatively shallow in both dimensions, suggesting that the designer has some latitude in selecting a carrier particle size and the desired biofilm thickness (i.e., fluidization conditions). Similar conclusions regarding the relative effects of carrier particle size and biofilm thickness have also been reached with a model assuming first-order intrinsic kinetics.[2,4] Thus, they can be considered to be general.

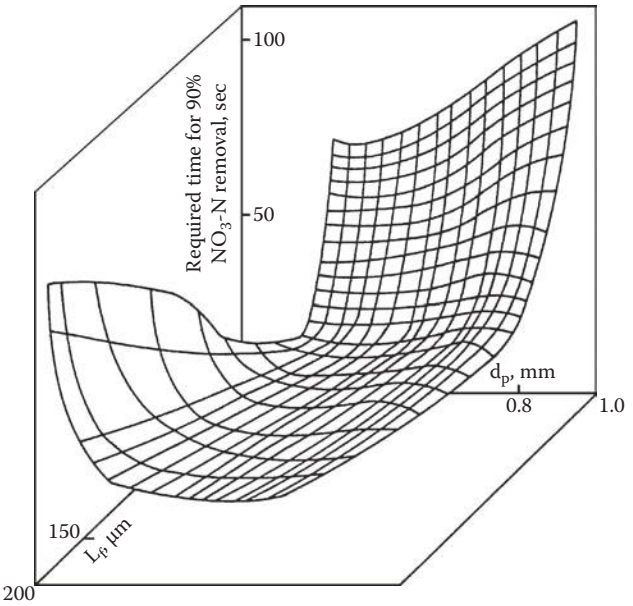

FIGURE 18.13 Combined effects of carrier particle diameter, d_p, and biofilm thickness, L_f, on the time required to remove 90% of the influent nitrate in a denitrifying FBBR. (Reprinted from Shieh, W. K. Mulcahy, L. T., and LaMotta, E. J., Mathematical model of the fluidized bed biofilm reactor. *Enzyme and Microbial Technology,* 4:269–75, 1982. Copyright © Elsevier Science Ltd. With permission.)

The above information considers selection of the optimal carrier particle size and biofilm thickness. Once they have been fixed and the mass of carrier particles and porosity have been selected, then the required superficial velocity (and associated bed diameter) and bed height become fixed. A question then arises about the performance of the FBBR if the influent flow rate or substrate concentration changes. Some understanding of the response can be obtained by considering what would happen to the quantity of biomass in the system, which depends on how the FBBR is operated. Consider the case in which biomass wastage is practiced to maintain a fixed bed height and the recirculation rate is adjusted to maintain a constant superficial velocity through the bed. If the influent flow rate or concentration was increased while maintaining the same mass input rate of substrate and the same superficial velocity, then the impact on system performance would be minimal because the mass input of substrate per unit of biomass would stay about the same. On the other hand, if the mass input rate of substrate increased, the output substrate concentration would increase by a proportionally greater amount. Because of the increased input, the biomass would be exposed to higher substrate concentrations, which would cause it to grow faster, leading to thicker biofilms. This would cause the degree of expansion in the bed to increase, which would reduce the number of carrier particles associated with the fixed bed height. Even though each carrier particle left in the system would have a thicker biofilm, the mass of biomass in the system would decrease because the increased biomass on each carrier particle would not compensate for the loss of carrier particles. Furthermore, the effectiveness of each carrier particle would be decreased because of the increased film thickness. On the other hand, if the recirculation rate was decreased to maintain the same mass of carrier particles within the prescribed bed height, the increase in the output substrate concentration would not be as great because more biomass could be maintained in the system. Likewise, if the bed height was allowed to expand to accommodate the increased mass input rate of substrate, there would be little impact on performance.

The above suggests that an FBBR can be thought of somewhat like a suspended growth system. If a suspended growth CSTR is operated at a fixed solids retention time (SRT), the mass of biomass in the system increases in proportion to an increase in the influent mass flow rate of substrate, and the effluent concentration remains the same. On the other hand, if it is operated at a fixed mixed liquor suspended solids concentration, the process loading factor increases (the SRT decreases) and the effluent concentration increases. The FBBR acts similarly. If it is operated in a manner that allows the mass of biomass to increase, the impact of an increase in the mass input rate is minimal. Conversely, if the operational practice results in the same or less biomass, performance suffers.

The concept of SRT in a fluidized bed is a helpful one, but one must recognize that the situation is more complex than in a suspended growth bioreactor because of the mass transfer limitations in biofilms.[10] Long SRTs can lead to thick biofilms, which have a lower effectiveness factor. Thus, FBBRs with long SRTs can have lower volumetric removal rates. Conversely, at short SRTs, even though the biofilms are thin, the amount of biomass may be insufficient to get good removal. In other words, the fact that there is an optimal biofilm thickness associated with a given particle size means that there is also an optimum SRT for a given situation.

18.5 SIZING A FLUIDIZED BED BIOLOGICAL REACTOR

The sizing of an FBBR proceeds in a logical and straightforward manner, utilizing the information presented earlier in this chapter. As with all other biological processes, the parameters in the model must be specified, as must the influent flow rate and concentration, and the desired substrate removal across the system. Shieh and Keenan[30] have presented procedures for estimating the needed parameters. The sizing of the FBBR entails choosing a porosity, the carrier particle, the optimal biofilm thickness, the superficial velocity and the associated recirculation and bioreactor cross section, and the bed height.[2]

Andrews[2] advocates using the smallest porosity that prevents the particles from agglomerating or having a collision frequency that would cause excessive shear. He states that a porosity of 0.60 is a reasonable compromise for the minimum porosity in the fluidized bed, which would be at the base. After the porosity is fixed, the next decision is to select the carrier particles. This is an important decision because everything else follows from it. We saw in Figure 18.4 that carrier particle size and density have an important impact on the stability of the bed. Light, small particles form a bed that is susceptible to large fluctuations in expanded bed height by small variations in superficial velocity. Sand ($\rho_p = 2.65$) with a particle diameter of around 0.5–0.6 mm is commonly used because it is readily available and offers good stability.

Having chosen the porosity and the carrier particle, it is now possible to choose the biofilm thickness that maximizes the average volumetric reaction rate in the system, which is equivalent to the product of the effectiveness factor times the biomass concentration. Using the average substrate concentration across the tower, the effectiveness factor can be calculated as a function of biofilm thickness using the characteristic biofilm thickness, L_{fc}, given by Equation 18.18 and the appropriate modified Thiele modulus. If it is necessary to consider external mass transfer resistance in determining the effectiveness factor, then the superficial velocity associated with each biofilm thickness has to be computed for use in the appropriate mass transfer coefficient correlation. The biomass concentration can be calculated with Equation 18.23, which was derived by substituting Equation 18.12 into Equation 18.15:

$$X_B = \rho_{fd}(1-\varepsilon)\left[1-\left(\frac{d_p}{d_b}\right)^3\right]. \tag{18.23}$$

It should be recalled that the dry density of the biofilm is a function of its thickness, as shown in Figure 18.2. Consequently, an appropriate correlation should be used with Equation 18.23. The optimal biofilm thickness is obtained by plotting the product of the effectiveness factor times X_B as a function of the biofilm thickness and selecting the value that maximizes the product.

Once the optimum biofilm thickness has been chosen, the superficial velocity required to achieve the desired porosity can be calculated with Equation 18.6. The terminal settling velocity of the bioparticle can be calculated with Equation 18.2 after replacement of ρ_p with ρ_b and d_p with d_b. The value of the coefficient n can be calculated with Equations 18.13 and 18.14. At this point a check should be made to ensure that the required superficial velocity does not exceed the terminal settling velocity of the clean carrier particle or any bioparticle desired in the bioreactor. If it does and there is no error in the computations, then the chosen carrier particle is not feasible for the desired biofilm thickness and another must be selected.

Having selected the superficial velocity, it is now possible to determine the required cross-sectional area and associated height for the FBBR. The superficial velocity in an FBBR is equivalent to the total hydraulic loading in a packed tower, Λ_H, which was given by Equation 17.12. Thus,

$$v = \Lambda_H = \frac{F(1+\alpha)}{A_c}. \tag{18.24}$$

Since the superficial velocity is known, A_c can be calculated after the recirculation ratio, α, has been chosen. Several factors go into the selection of α.[30] When an aerobic two-phase FBBR is being used, oxygen is provided by dissolving it in the recirculation stream. When high purity oxygen is used for the supply, approximately 60 mg/L of oxygen can be dissolved in wastewater and used in the FBBR with less than 1% loss.[30] Thus, the amount of recirculation required can be calculated from a mass balance on chemical oxygen demand across the FBBR. Recirculation can also be provided to maintain a constant superficial velocity across the FBBR when the influent flow is variable. Often,

however, it is best to set the influent equal to the highest expected flow rate across the system and to add to it the amount of recirculation required to transfer the needed oxygen, and to use those values in Equation 18.24 to calculate the cross-sectional area.

Finally, after selection of α and A_c, the tower height, H_{Bb}, can be calculated. This requires use of the appropriate reactor flow submodel as described in Section 18.3.3. If the recirculation flow is small, then a model for plug flow or plug flow with dispersion would be appropriate. Alternatively, a tanks-in-series model could be used as well. On the other hand, if the recirculation rate is high, it might be possible to treat the entire bioreactor as a CSTR. In addition, if the bed is likely to be stratified with a variety of bioparticle sizes, then that expectation can be incorporated into the reactor flow submodel. The computational procedure involved depends on the type of model employed. The goal, however, is to determine the residence time or expanded bed height required to achieve the required effluent substrate concentration. Once H_{Bb} is known, then the mass of carrier particles, $\mathbf{M}_p$, can be calculated with Equation 18.12.

18.6 KEY POINTS

1. A fluidized bed biological reactor (FBBR) is one in which the biofilm grows attached to small carrier particles that remain suspended in the fluid (i.e., fluidized) by the drag forces associated with the upward flow of water. It has several advantages over other attached growth bioreactors: better control of biofilm thickness, superior mass transfer characteristics, less tendency to clog, very high surface areas for biofilm development, and low pressure drops.
2. One attribute of FBBRs is that the dry density of the biofilm that develops on a bioparticle depends on the thickness of the biofilm, with thinner films exhibiting higher density. Consequently, thicker biofilms do not necessarily lead to more active biomass.
3. Two superficial velocities are critical to defining the operating range for a fluidized bed, those associated with points A and B in Figure 18.3. The velocity associated with point A is called the minimum fluidization velocity and is the velocity at which the particles just begin to move apart. The velocity associated with point B is the terminal settling velocity of the carrier particles. If it is exceeded, the particles are carried away in continuous fluidization and the bed is destroyed.
4. The growth of biofilm on carrier particles changes their terminal settling velocity. This is due to three things. First, growth of the biofilm changes the size of the particle. Second, unless the carrier particle has a density equivalent to the wet density of the biofilm, growth of the biofilm changes the overall effective density of the particle. Third, the surface properties of the biofilm differ from those of the clean particle, thereby changing the relationship between the drag coefficient, C_D, and the Reynolds number.
5. Because growth of a biofilm changes the terminal settling velocity of a particle, it also changes its fluidization properties. As a consequence, the height of a fluidized bed increases as the thickness of the biofilm increases.
6. Bioparticles containing uniform diameter carrier particles of low density, similar to the wet density of the biofilm, tend to stratify because the bioparticle density does not change as the biofilm grows. Under such a situation, biomass wastage should be done from the bottom of the bed. Conversely, bioparticles containing uniform diameter carrier particles of high density tend to form well-mixed beds because the density of the bioparticle decreases as the biofilm grows. Biomass wastage from these beds should be done from the top.
7. The biofilm thickness that can be maintained by the hydrodynamic conditions in an FBBR is controlled by the superficial upflow velocity imposed on the bioreactor and the desired fluidized bed height. It must be calculated iteratively by the procedure illustrated in Figure 18.9.

8. Modeling the performance of an FBBR requires linkage of a biofilm submodel with fluidization and reactor flow submodels. The procedure is iterative for most situations.
9. For a given mass of carrier particles with a given density, there is a combination of biofilm thickness and carrier particle diameter that maximizes the conversion rate of substrate per unit bioreactor volume. The biofilm thickness associated with that optimum is usually slightly less than the biofilm depth at which all substrate would be exhausted and, thus, is usually small.
10. When sizing an FBBR, a porosity around 0.60 should be chosen because it prevents the particles from agglomerating or having a collision frequency that would cause excessive shear. Sand ($\rho_p = 2.65$) with a particle diameter of around 0.5–0.6 mm is commonly used as a carrier particle because it is readily available and offers good stability of bed height against changes in superficial velocity.

18.7 STUDY QUESTIONS

1. Describe the general characteristics of an FBBR, including its advantages over other attached growth processes, and differentiate between a tower bioreactor and a supported-film bioreactor.
2. Explain why thin biofilms are often more desirable than thick ones.
3. Describe what happens to the pressure drop and the porosity of a bed of small, spherical particles as the superficial upflow velocity through it is increased and explain why those events occur.
4. Spherical sand ($\rho_p = 2.65$ g/cm^3) particles with a diameter of 0.6 mm are to be fluidized by water at a temperature of 20°C. Determine the minimum velocity for fluidization and the terminal settling velocity of the particles. Assume that the minimum porosity at fluidization is 0.45. State the relationship that you chose to use to determine the drag coefficient from the terminal Reynolds number and justify your choice.
5. Describe how growth of a biofilm influences the terminal settling velocity of a bioparticle containing sand as the carrier particle. Also explain why the effect occurs.
6. Rework Study Question 4, but assume that a biofilm with a thickness of 0.10 mm has grown on the carrier particle. Assume that the weight fraction moisture content of the biofilm is 0.93 and that its dry density varies with the biofilm thickness as shown in Figure 18.2.
7. A FBBR has a diameter of 7.5 cm and contains 1.0 kg of spherical sand ($\rho_p = 2.65$ g/cm^3) carrier particles with a diameter of 0.6 mm. Determine the bed height when a biofilm with a thickness of 0.10 mm has grown on the particles and they are fluidized with a superficial velocity of 2.5 cm/sec. Assume that the weight fraction moisture content of the biofilm is 0.93 and that its dry density varies with the biofilm thickness as shown in Figure 18.2.
8. Explain how and why the density of the carrier particle influences the degree of solids mixing that occurs in an FBBR.
9. An FBBR has a diameter of 7.5 cm and contains 1.0 kg of spherical sand ($\rho_p = 2.65$ g/cm^3) carrier particles with a diameter of 0.6 mm. Determine the biofilm thickness that could be carried on the bioparticles if the bed height is maintained at 1.0 m while the bed is being fluidized with a superficial velocity of 1.5 cm/sec. Assume that the weight fraction moisture content of the biofilm is 0.93 and that its dry density varies with the biofilm thickness as shown in Figure 18.2.
10. Explain the significance of the characteristic biofilm thickness as defined by Equations 18.17 and 18.18.
11. Prepare a flow diagram showing the steps that must be followed to calculate the effluent substrate concentration from a completely mixed FBBR with intrinsic zero-order kinetics.

12. Repeat Study Question 11 for intrinsic first-order kinetics.
13. Prepare a flow diagram showing the steps that must be followed to calculate the effluent substrate concentration from an FBBR that has a plug-flow liquid phase but a uniform bioparticle size. Assume intrinsic first-order kinetics.
14. Explain why the biofilm thickness that maximizes the quantity of biomass in an FBBR may not be the same as the biofilm thickness that maximizes the conversion rate of substrate per unit bioreactor volume.
15. Prepare a flow diagram showing the steps that must be followed in sizing an FBBR to achieve a desired effluent concentration.

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Part V

Applications: Attached Growth Reactors

Part IV presents the fundamental principles of ideal attached growth reactors and their application to packed towers, rotating disc reactors, and fluidized bed reactors. In Part V, those principles are applied to the practical design and operation of a variety of attached growth reactors. Chapter 19 addresses the design of trickling filters, which historically were the principal packed towers used in practice. Chapter 20 addresses rotating biological contactors, the main application of rotating disc reactors. Both have been widely used in practice for removal of biodegradable organic matter, combined carbon oxidation and nitrification, and separate stage nitrification. Although the current emphasis on biological nutrient removal, for which trickling filters and rotating biological contactors are not well suited, has decreased the number of new installations, many are still in service. Because of the need to keep them performing well throughout their design lifetimes, it is important that their design and operational characteristics be presented. Finally, Chapter 21 addresses a variety of submerged fixed film reactors that have undergone various degrees of development and application in practice. These include downflow and upflow packed bed reactors, fluidized and expanded bed reactors, moving bed reactors, and integrated fixed film activated sludge systems. They have been used for removal of biodegradable organic matter, combined carbon oxidation and nitrification, separate stage nitrification, and denitrification using both the carbon in the wastewater itself and supplemental carbon (such as methanol). As in Part III, several process design approaches are presented, representing diverse degrees of sophistication and information requirements. The reader is referred to Chapter 10 for discussions of the iterative nature of biological process design and the need for a variety of design procedures, both of which are equally applicable to attached growth systems.

19 Trickling Filter

The term trickling filter represents an array of attached growth biochemical operations in which wastewater is applied to fixed media in an air filled packed tower. Treatment of the wastewater is accomplished by microorganisms growing attached to the media, of which there are several types. Trickling filters are aerobic and are used to oxidize biodegradable organic matter, forming biomass. The produced biomass sloughs from the media and is separated from the treated wastewater in a downstream clarifier. Trickling filters are also used to oxidize ammonia-N to nitrate-N. Nitrification can either occur in a trickling filter that is being used for oxidation of organic matter, a process called combined carbon oxidation and nitrification, or it can occur in a trickling filter receiving wastewater that has previously been treated to remove organic matter, a process called separate stage nitrification. Both are discussed in this chapter; the theoretical performance of packed towers is discussed in Chapter 17.

19.1 PROCESS DESCRIPTION

19.1.1 General Description

Figure 19.1 presents a schematic diagram of a trickling filter. A typical trickling filter consists of five major components: the media bed, the containment structure, the wastewater application (or dosing) system, the underdrain system, and the ventilation system.[65,66] The media bed provides the surface upon which the microorganisms grow. Media options include rock, wood, and synthetic plastic of various types and configurations. Rock is seldom used in new trickling filters today, although many older facilities contain rock media.

The containment structure retains the media and applied wastewater and controls the effects of wind. Some media, such as rock and random plastic, are not self-supporting and, in these instances, the containment structure must also support the media. The containment structure is often constructed of concrete, either poured in place or precast panels. Other materials such as wood, fiberglass, and coated steel have also been used, particularly when the media is self-supporting.

The application system uniformly applies the wastewater to the media bed. Uniform application is necessary to ensure wetting all of the media. The application system is also used to control dosing frequency, which affects process performance.

The underdrain system has two functions. One is to collect the treated effluent for conveyance to further treatment or to discharge. The second is to provide a plenum to allow air passage through the open media bed, thereby providing the oxygen required for aerobic metabolism. Clay or concrete underdrain blocks are often used for rock media trickling filters because of the weight that must be supported. Many types of underdrain systems, such as concrete piers, wood stringers, and reinforced fiberglass grating, are used with other media.

Oxygen required to meet the metabolic needs of the microorganisms is provided by the vertical flow of air through the media. As discussed in Section 19.2.5, ventilation to provide that air can be by either natural draft or mechanical means. In natural draft systems the difference in density between air inside and outside the trickling filter causes air within the trickling filter to either rise or sink. This results in a continuous flow of air through the media. Density differences arise because air within the trickling filter quickly becomes saturated with water vapor and reaches the temperature of the applied wastewater. Consequently, the magnitude of the density difference depends on the temperature and humidity of the ambient air. One disadvantage of natural draft ventilation is

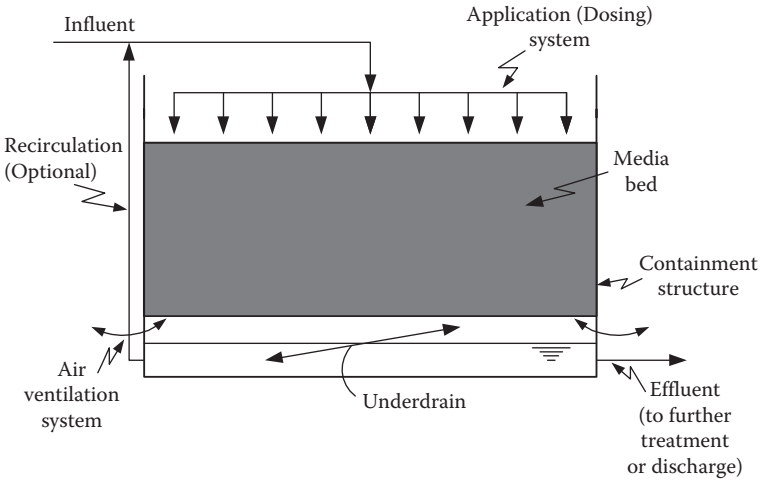

FIGURE 19.1 Schematic diagram of a trickling filter.

that neutral density conditions can occur, resulting in the absence of air flow through the trickling filter and the development of anaerobic conditions. In forced draft ventilation systems the air is applied to the trickling filter by mechanical means. In all cases, air must be uniformly distributed across the media to ensure that oxygen is provided to the entire bioreactor.

As indicated in Figure 19.1, trickling filter effluent may be recirculated and mixed with the influent wastewater prior to its application to the trickling filter. Recirculation dilutes the influent wastewater and also allows separation of the hydraulic and organic loadings to the unit. Recirculation is an essential process component in some applications. The need for recirculation and the various recirculation configurations are discussed later.

The influent to a trickling filter must generally be pretreated to remove nonbiodegradable particulate matter such as plastics, rags, and stringy material. Materials of this type can easily plug the distributor and the media, leading to unequal flow distribution and poor performance. Debris removal by coarse screens is not generally acceptable for trickling filter applications, but adequate removal can be accomplished using fine screens (generally 1 mm opening or less) or primary clarifiers. Primary clarifiers are used most often.

The media provides a surface for the growth of microorganisms and the mechanism for retaining the microorganisms in the unit. Organic matter removal and nitrification occur by the same mechanisms as in any other aerobic biochemical operation. Soluble organic matter diffuses into the biofilm located on the media surface and is used as a carbon and energy source by heterotrophic bacteria. Colloidal and particulate organic matter are removed by sorption and entrapment. They are subsequently hydrolyzed into soluble organic matter by the action of extracellular enzymes. The soluble organic matter is then metabolized by the heterotrophic bacteria contained within the biofilm. Ammonia-N also diffuses into the biofilm where part is used by the heterotrophs for biomass synthesis and the remainder is oxidized to nitrate-N by nitrifying bacteria. The removal of organic matter and nitrification result in the production of additional biomass and increased biofilm thickness. When the biofilm reaches a thickness that can no longer be supported on the media, the excess sloughs off and passes into the treated effluent. Chapter 16 describes the role of diffusion in controlling the metabolic processes occurring within the biofilm.

Trickling filter effluents are usually treated in clarifiers to remove the produced biomass, although this may not be necessary in some separate stage nitrification applications because of the low yield of nitrifying bacteria. For example, consider a separate stage nitrification application in which 20 mg/L of ammonia-N is being oxidized. Since the yield coefficient for nitrifying bacteria is approximately 0.15 mg total suspended solids (TSS)/mg ammonia-N oxidized, only about 3 mg/L

of nitrifying bacteria will be produced. This increase in suspended solids concentration may not be significant in relation to plant effluent suspended solids limits, and thus a liquid-solids separation device may not be required downstream of the trickling filter.[61]

As discussed in Chapter 17, the liquid flow pattern through a trickling filter may generally be thought of as plug flow with dispersion. Because of this flow pattern and because the microorganisms are fixed on the media, variations in the composition of the biomass often exist along the depth of the trickling filter. This is in contrast to the activated sludge process where biomass recycle results in a uniform biomass composition throughout the bioreactor. The variation in biomass composition through the depth of a trickling filter can have significant impacts on process performance. For example, carbon oxidation typically occurs in the upper portion of combined carbon oxidation and nitrification systems, while nitrification occurs in the lower portion,[63] as illustrated in Figure 19.2. This is due to competition between heterotrophic and autotrophic bacteria for space within the biofilm, as discussed in Chapter 16. In the upper levels of a trickling filter both organic matter and ammonia-N concentrations are relatively high and will, generally, not limit the specific growth rate of either the heterotrophic or nitrifying bacteria. Under these conditions the heterotrophic bacteria can grow faster than the nitrifying bacteria and outcompete them for space within the biofilm. As the wastewater flows down through the trickling filter, organic matter is removed, ultimately causing its concentration to limit the specific growth rate of the heterotrophic bacteria. Because the ammonia-N concentration is still high, a point is reached at which the specific growth rate of the nitrifying bacteria exceeds the specific growth rate of the heterotrophic bacteria. Under these conditions the nitrifying bacteria can effectively compete with the heterotrophic bacteria for space and will become established in the biofilm. As discussed in Chapter 16, the soluble biodegradable organic matter concentration must be reduced to about 20 mg/L as chemical oxygen demand (COD) before this can occur.[45,63]

The plug-flow nature of the trickling filter can also result in reduced growth and biomass accumulation in the lower portion of the tower, leading to patchy growth as indicated in Figure 19.2.[7,48] This occurs because of the low yield of nitrifying bacteria and the presence of predators that consume trickling filter biomass. The reduced biofilm thickness can result in a diminished wastewater treatment capacity in the lower portion of the tower, which can be particularly important when the loading increases, such as during diurnal high flow events.

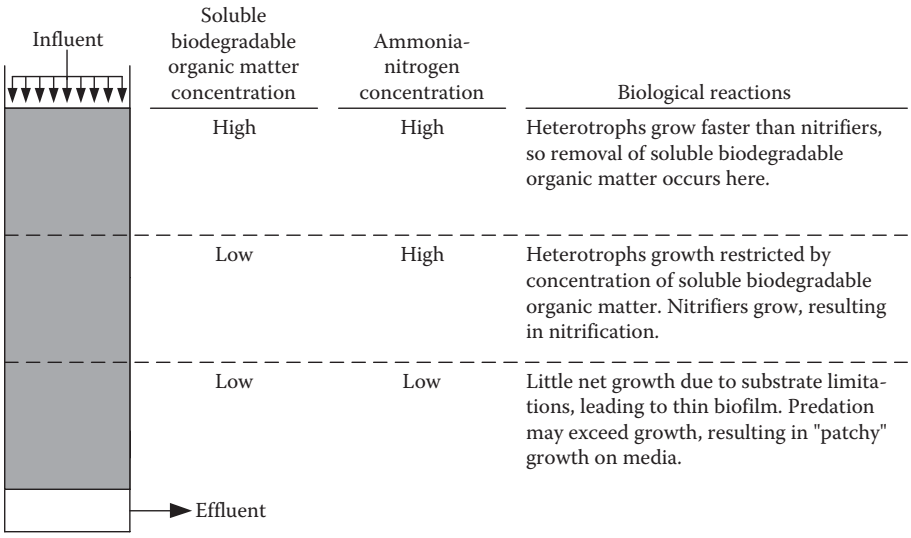

FIGURE 19.2 Representation of the biological reactions occurring at various depths in a trickling filter accomplishing combined carbon oxidation and nitrification.

Although trickling filters are generally thought of as aerobic processes, in most cases the biofilm is relatively thick and exceeds the depth of oxygen penetration.[68] Consequently, the biofilm consists of an outer aerobic layer and an inner anoxic/anaerobic layer. This affects process performance in many significant ways. For example, the occurrence of a zone of low dissolved oxygen concentration within the biofilm allows denitrification to occur, although the extent is limited. The removal of organic matter typically occurs in the upper levels of the trickling filter, whereas nitrification occurs in the lower levels. As a consequence, the concentration of organic matter is low in the region where nitrate-N is produced. However, some denitrification can occur when treated effluent containing nitrate-N is recirculated to the process influent.[45]

19.1.2 Process Options

Trickling filter process options vary with the treatment objective, the media type, and the nature of the other unit operations in the process train.

19.1.2.1 Treatment Objectives

Trickling filters are used to treat a wide variety of wastewaters to achieve various treatment objectives. Consequently, as indicated in Table 19.1, those two factors can be used to characterize the trickling filter process. Because the degree of treatment is often determined by the process organic loading rate, it can be used as a quantitative indicator of the degree of treatment. The total organic loading is the mass flow rate of biodegradable organic matter in the influent wastewater (excluding recirculation) divided by the media volume, V_M. It is given the acronym TOL and the symbol Λ_S, and is calculated as

$$\Lambda_S = \frac{F(S_{SO} + X_{SO})}{V_M}, \tag{19.1}$$

where S_{SO} and X_{SO} are the concentrations of readily and slowly biodegradable substrate in the influent, respectively, and F is the influent flow rate. The concentration of biodegradable organic matter is typically expressed as either five-day biochemical oxygen demand (BOD_5) or as COD, making the units for TOL kg $BOD_5/(m^3 \cdot day)$ or kg $COD/(m^3 \cdot day)$.

Influent wastewater containing both organic matter and ammonia-N can be treated in roughing, carbon oxidation, or combined carbon oxidation and nitrification applications. The difference between them is the degree of treatment, as indicated by the TOL.[65,66] A relatively high TOL is used in a roughing filter, resulting in residual organic matter and suspended solids concentrations that generally exceed the conventional definition of secondary treatment (30 mg/L each of BOD_5 and TSS). Roughing applications are generally used to lower the concentration of organic matter prior to further biological treatment. A somewhat lower TOL is used in a carbon oxidation application so that relatively complete removal of organic matter (approaching the definition of secondary

TABLE 19.1
Trickling Filter Process Applications

Application/Objective	Influent Wastewater	TOL kg $BOD_5/(m^3 \cdot Day)$
Roughing	Screened wastewater or primary effluent	1.5–3.5
Carbon oxidation	Screened wastewater or primary effluent	0.7–1.5
Combined carbon oxidation and nitrification	Screened wastewater or primary effluent	<1.0
Separate stage nitrification	Secondary effluent	NA[a]

[a] Not applicable.

treatment) is achieved. Finally, an even lower TOL is used to achieve combined carbon oxidation and nitrification. Carbon oxidation is relatively complete in the upper portion of the trickling filter, thereby allowing the growth of nitrifying bacteria in the lower portion, as described above.

Separate stage nitrification differs in that a stream that is relatively low in biodegradable organic matter and suspended solids (BOD_5 and TSS generally less than 30 mg/L), but with a significant ammonia-N concentration, is applied to the trickling filter.[7,18,48,61,65,66] As a consequence, the microbial community that develops in the biofilm is enriched in nitrifying bacteria. In this instance the TOL is not an issue. Instead, the ammonia loading and other operational factors determine the degree of ammonia removal. For example, the nitrification efficiency might be correlated with the total ammonia-N loading (TAL) with units of kg NH_3-N/($m^3 \cdot$ day) and the symbol Λ_{NH}. Alternatively, there may be instances in which it is advantageous to express nitrification performance in terms of the loading of total Kjeldahl nitrogen (TKN) on the process. In that case, one might speak of a total nitrogen loading (TNL) with units of kg N/($m^3 \cdot$ day) and symbol Λ_N.

19.1.2.2 Media Type

Many types of media have been used in trickling filters but, in general, they can be divided into rock media and high-rate media.[14,65,66] Table 19.2 summarizes the characteristics of various rock and high-rate media, while Figure 19.3 presents photographs of several typical media. Although rock media trickling filters are seldom built today, the reader should be familiar with their characteristics because many are still in use.

An ideal rock media consists of rounded river rock of relatively uniform size and shape.[38,65,66] Typical rock media are approximately 5 cm in diameter, although larger and smaller rock can be used. Rounded river rock is durable, and its rounded nature and uniform size minimize bed consolidation and plugging. Irregular materials such as slag have also been used, but they exhibit a greater potential for plugging due to the entrapment of biomass by the irregular surfaces and openings. Media attrition will also occur due to freeze-thaw cycles, resulting in the production and accumulation of fines that contribute to plugging. This is particularly true for slag media due to their rough surfaces that allow water to intrude into the media.

As indicated in Table 19.2, the primary characteristics of rock trickling filter media are high unit weight, low specific surface area, and low void space. These impose significant design and operational constraints. The higher the unit weight, the greater the structural requirements and the shallower the media depth must be to avoid media crushing. For rock media trickling filters, media depths are typically on the order of 2 m, which restricts treatment efficiency to some extent. Some have been constructed to greater depths using special construction techniques, but this is not conventional practice. The relatively low specific surface area and void space restrict the organic loadings that can be applied. The low specific surface area limits the capacity for biofilm growth and consequently, the treatment capability. Similarly, the low void volume limits the space available for the passage of air, water, and sloughed biomass through the media, thereby increasing the potential for media plugging at higher organic loadings. As a consequence, rock media have typically been used at low to moderate TOLs, generally in the range of 0.5–1.5 kg BOD_5/($m^3 \cdot$ day). More significantly, the shallow media depth and low organic loadings result in low wastewater hydraulic application rates. Although recirculation can be used to provide increased total hydraulic loadings (THLs, see Equation 17.12), typical design practice has been to limit recirculation ratios to minimize energy requirements. As a consequence, the THL to rock media is typically on the order of 0.5 m/hr, which is significantly less than that used with high-rate media. The adverse impacts of these low THLs on process performance are discussed later.

As illustrated in Table 19.2, high-rate media are characterized by significantly lower unit weight, higher specific surface area, and greater void space than rock media. As a consequence, media depths are typically greater (generally 5–7 m), and a wider range of TOLs can be used [up to 3.5 kg BOD_5/($m^3 \cdot$ day)]. Because of the greater media depth, the cross-sectional area of a high-rate media trickling filter will be significantly less than that of a comparably sized rock media trickling filter.

TABLE 19.2
Characteristics of Trickling Filter Media

Media Type	Description	Size cm	Unit Weight kg/m³	Specific Surface Area m²/m³	Void Space %
Rock	Rounded river rock of uniform size (desirable) or more irregular slag (less desirable)	2.5–7.5 5–10	1,500 1,600	60 45	50 60
High-rate					
Bundle	Sheets of PVC formed in various configurations and fastened together to form bundles with openings of various sizes and orientations	61 × 61 × 122	30–80	88–105	>95
Vertical, semicorrugated (VSC)		61 × 61 × 122	65–95	140–150	>95
Vertical, fully corrugated (VFC)					
60° cross flow (XF)					
Random (RA)	Irregular shapes manufactured of extruded plastic elements	2–10 (varies for individual products)	30–65 50–80	80–115 140–165	>94 >94
Horizontal (HO)	Wooden pallets, typically constructed of redwood or pressure treated lumber	122 × 122 × 4.8	165	46	>90

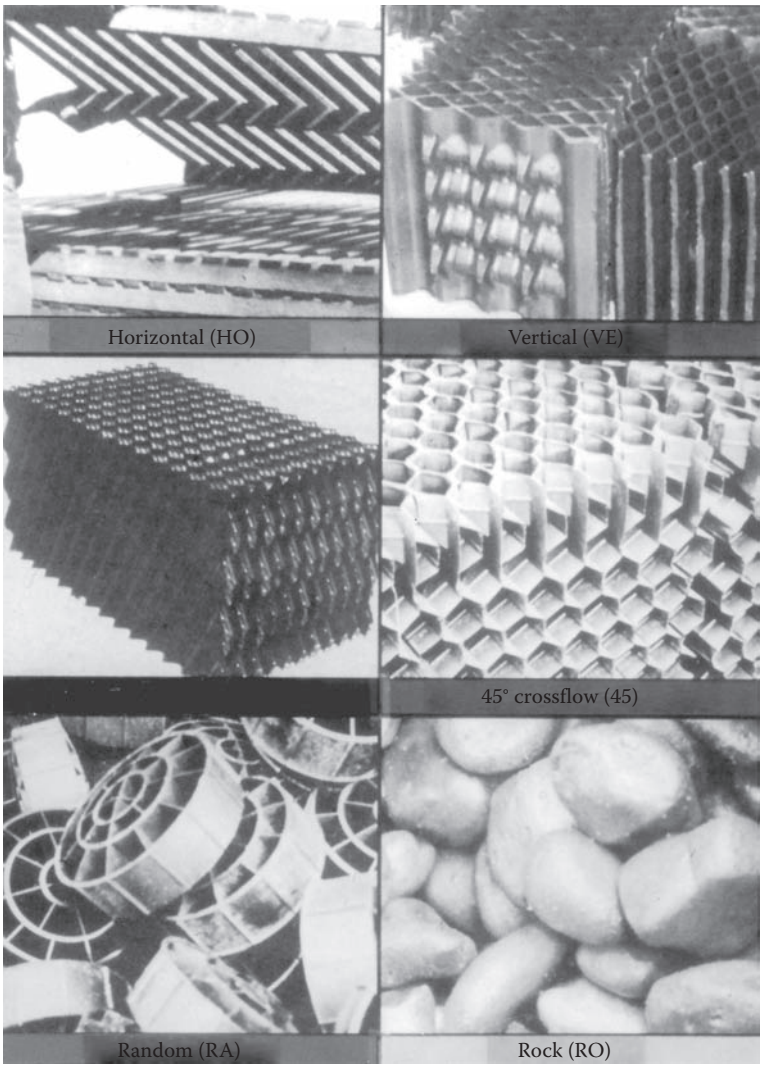

FIGURE 19.3 Typical trickling filter media. (From Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992. Copyright © Water Environment Federation. Reprinted with permission.)

A THL of 1.8 m/hr is typically considered necessary to fully wet and completely utilize high-rate trickling filter media.

The effects of media type on trickling filter performance are discussed more fully in Section 19.2.6, but a brief summary is provided here. Data suggest that the performance of rock and high-rate media trickling filters are similar when they are loaded at relatively low organic loading rates, less than about 1 kg $BOD_5/(m^3 \cdot day)$. In contrast, at higher TOLs the performance of trickling filters using high-rate media is superior. Other operating characteristics of rock and high-rate media trickling filters tend to be somewhat different. For example, excess biomass production rates may be somewhat less in rock media trickling filters than in high-rate media trickling filters.[24] This is thought to result from the greater biomass retention characteristics of rock media, resulting in increased degradation (probably anaerobic) of the retained biomass. The biomass retention characteristics may also result in variations in the settling characteristics of the produced biosolids and in the clarity of the settled effluent.

Plastic high-rate trickling filter media are manufactured with a variety of specific surface areas. Variations in the specific surface area are accomplished by modifications in the media configuration and dimensions. Consider plastic sheet (or bundle) media as an example. Increased specific surface area is accomplished by manufacturing media sheets with slightly smaller indentations, which allows media sheets to be placed closer together. Media with a specific surface area of approximately 100 m^2/m^3 is typically used when screened wastewater or primary clarifier effluent is being treated, whereas media with a specific surface area of approximately 140 m^2/m^3 is typically used when secondary effluent is being treated. The more open, low specific surface area media is needed to avoid plugging problems associated with the relatively high biomass production rates experienced when screened wastewater or primary clarifier effluent is being treated. Since the biomass production rate is lower when secondary effluent is being treated, media with a higher specific surface area media can be used in that application.

As indicated in Table 19.2, the specific surface area of wood media is relatively low. However, experience indicates that the application of return activated sludge to a trickling filter containing this media results in significant interstitial biomass growth, which increases its effective treatment capacity.[13,25] Side-by-side testing indicates that its efficiency will be equivalent to that of bundle media when return activated sludge (RAS) is recycled to the wood media. No similar improvement in performance is observed for the other media. Process configurations that recycle RAS to an upstream trickling filter are discussed in the next section.

19.1.2.3 Coupled Trickling Filter/Activated Sludge Systems

Coupled trickling filter/activated sludge (TF/AS) systems use an upstream trickling filter combined with a downstream suspended growth biochemical operation to accomplish overall wastewater treatment.[13,25] Figure 19.4 illustrates a typical system. Such systems are referred to as coupled processes because no liquid-solids separation device is provided between the trickling filter and the suspended growth bioreactor. As a consequence, biomass that grows on the trickling filter sloughs off and passes directly into the suspended growth bioreactor where it is enmeshed into and becomes part of the suspended biomass. A significant portion of the biomass contained in the suspended growth bioreactor (generally 60–90%) is originally grown in the trickling filter.

Trade-offs exist relative to the sizes of the two biochemical operations. If a relatively small trickling filter is used, then a larger suspended growth bioreactor must be used to accomplish a specified treatment goal. Conversely, if a relatively large trickling filter is used, the treatment goal can be accomplished using a smaller suspended growth bioreactor. This trade-off provides one of the

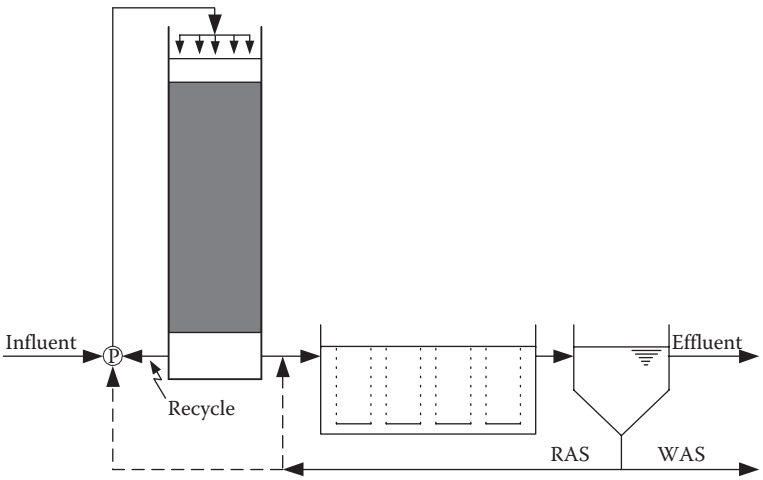

FIGURE 19.4 Schematic diagram of the coupled trickling filter/activated sludge (TF/AS) process.

TABLE 19.3
Coupled TF/AS System Options

	Unit Size		
Process Option	Trickling Filter	Suspended Growth Bioreactor	RAS Recycle Destination
Trickling filter/solids contact (TF/SC)	Large	Small	Suspended growth bioreactor
Activated biofilter (ABF)	Large	None	Trickling filter
Roughing filter/activated sludge (RF/AS)	Small	Large	Suspended growth bioreactor
Biofilter/activated sludge (BF/AS)	Small	Large	Trickling filter

criteria for classifying coupled TF/AS systems. Another criterion is the location for the RAS; either around the entire system (i.e., to the trickling filter) or around only the suspended growth bioreactor. Table 19.3 uses these two criteria to classify coupled TF/AS systems.

Both the trickling filter/solids contact (TF/SC) and the activated biofilter (ABF) processes use relatively large trickling filters and small suspended growth bioreactors. In both of these processes, removal of organic matter is achieved in the trickling filter. The suspended growth bioreactor is used primarily to flocculate and enmesh fine suspended solids contained in the trickling filter effluent and to prepare the biosolids for efficient removal in the secondary clarifier. In the TF/SC process the RAS is recycled to the suspended growth bioreactor, which is referred to as a solids contact basin. In the ABF process, no distinct suspended growth bioreactor is provided. Rather, underflow from the clarifier is recycled to the trickling filter and the suspended biomass is developed and contained only in the recirculating fluid mass.

The roughing filter/activated sludge (RF/AS) and biofilter/activated sludge (BF/AS) processes use small trickling filters. As a consequence, the removal of organic matter in the trickling filter is incomplete and the suspended growth bioreactor removes the remaining organic matter. In the RF/AS process the RAS is recycled only to the suspended growth bioreactor, whereas in the BF/AS process it is recycled to the trickling filter. In some instances (particularly the treatment of high concentration, readily biodegradable wastewaters such as from the food processing industry), recycle of the RAS to the trickling filter can improve sludge settling characteristics. This is thought to be related to a "selector effect" that is achieved by contacting the RAS with high concentrations of organic matter in the highly aerated biofilter environment.[24,25] The use of selectors to control sludge settleability in activated sludge systems is discussed in Section 11.2.1.

19.1.3 Comparison of Process Options

The trickling filter process is a stable, reliable process that is capable of providing economical wastewater treatment. Energy requirements are typically lower than in suspended growth systems, which was one reason for the renewed popularity of trickling filters in the 1980s, a time when energy costs were escalating rapidly. The process is also relatively simple to operate. Capital costs can be high compared to other biochemical operations, and process operation cannot be adjusted as easily in response to loading and/or performance variations. The biggest drawback of trickling filters is that their performance may not meet current discharge standards. However, the performance of coupled TF/AS systems will generally equal that of suspended growth systems. The simplicity, stability, and low energy requirements for the trickling filter process have made it a popular option for carbon oxidation and nitrification. Primary clarification is typically provided prior to a trickling filter to minimize the debris loading.

Table 19.4 summarizes the benefits and drawbacks of the various trickling filter options. Roughing applications can provide very economical removal of organic matter, particularly from

TABLE 19.4
Trickling Filter Process Comparison

Process	Benefits	Drawbacks
Treatment Objectives		
Roughing	Economical, particularly for high strength wastewaters Simple to design and operate Process and facility design well known	Further treatment typically required prior to discharge Generally requires secondary clarification
Carbon oxidation	Economical Simple to design and operate Process and facility design well known	Performance is consistent, but may not reliably meet stringent performance standards Generally requires secondary clarification Limited operational flexibility
Combined carbon oxidation and nitrification	Simple to design and operate	Performance is consistent, but may not reliably meet stringent performance standards Generally requires secondary clarification Limited operational flexibility
Separate stage nitrification	Simple to design and operate	Performance is consistent, but may not reliably meet stringent performance standards Limited operational flexibility
Media Type		
Rock	Large number of existing applications Quite effective at low to moderate organic loading rates	Relatively expensive due to structural constraints Not applicable for high loading applications Odor potential
High-rate	Economical Applicable to a wide range of process loadings and applications Process and facility design well known	Media collapses have occurred due to improper application and/or manufacturing

Coupled Trickling Filter/Activated Sludge (TF/AS)		
Trickling filter/solids contact (TF/SC)	Stable, reliable performance Simple to design and operate Low energy Process and facility design well known	Moderate capital cost
Activated biofilter (ABF)	Simple to design and operate Low energy Low capital cost Process and facility design well known	Process performance variable, except at low loading
Roughing filter/activated sludge (RF/AS)	Stable, reliable performance Simple to design and operate Low capital cost Process and facility design well known	Moderate energy cost
Biofilter/activated sludge (BF/AS)	Stable, reliable performance Simple to design and operate Low capital cost Process and facility design well known Improved sludge settling characteristics in some applications	Moderate energy cost

high strength wastewaters. Further treatment of the roughing filter effluent (in addition to secondary clarification) is typically required prior to final discharge, but the size of the downstream treatment system is reduced.

Carbon oxidation applications offer the advantages of favorable economics, simple design and operation, and well-known process and facility design procedures. Process performance is consistent and reliable, but may not meet the stringent standards now typically required.

Combined carbon oxidation and nitrification and separate stage nitrification applications are simple, both to design and to operate. Like carbon oxidation processes, process performance is also consistent and reliable but with recognized performance limitations.

A comparison of rock and high-rate media indicates substantial advantages for high-rate media. The only drawback associated with high-rate media is that media collapses can occur as a result of improper application or manufacturing. This suggests that experience is necessary in its application. Due to the associated benefits, nearly all new trickling filters are constructed using high-rate media. However, a large number of rock media trickling filters exist and many are providing effective and economical service.

The coupled TF/AS process was developed to take advantage of the energy efficiency and stability of the trickling filter while also achieving the excellent effluent quality obtained with the activated sludge process. Experience indicates that this objective is typically achieved. Performance differences between the four primary TF/AS options are relatively minor. An economic trade-off exists between the more energy-efficient TF/SC and ABF processes and the less capital intensive RF/AS and BF/AS processes. Multiple modes of operation are often incorporated into full-scale facilities so that selection of a single coupled process option is not necessary.

19.1.4 Typical Applications

The trickling filter process is widely accepted and used for the aerobic biological treatment of wastewaters. It is used for both the removal of organic matter and nitrification. Its long history has resulted in a large number of operating installations. Many of the older installations use rock media, and a large number of successful installations of various sizes exist. Installations constructed in the last 30 years generally use high rate media, either plastic sheet, random, or horizontal. Plastic sheet media is currently the most popular due to its availability, cost, and good performance characteristics. Horizontal media is seldom used in new installations today due to its higher costs relative to plastic media. However, many existing installations exist because of its popularity during the 1970s.

Roughing filters are often used to pretreat industrial wastewaters containing high concentrations of readily biodegradable organic matter. The roughing filter effluent (after secondary clarification) typically receives further treatment, either in another biological treatment system located at the industrial site or in a municipal wastewater treatment system. Roughing filters have also been used to pretreat mixtures of municipal and industrial wastewater containing high concentrations of organic matter.

Use of the trickling filter for carbon oxidation has declined since the 1950s due to its general inability to meet stringent discharge standards. The coupled TF/AS systems were developed in response to this performance shortfall and have been quite successful in meeting stringent discharge standards.[13,25,37,40] Such systems can be designed for carbon removal applications alone or to accomplish combined carbon oxidation and nitrification. They have generally proven effective in achieving this goal while also retaining the basic operating characteristics of the trickling filter process.

The use of high-rate media trickling filters for separate stage nitrification has been actively considered for more than two decades, and some facilities have been in operation for a number of years demonstrating successful performance.[51,59,61,65,66] Interest in this application has increased as practitioners have identified potential cost and operational advantages.

While the trickling filter process has been used widely for many decades, more recent process and mechanical improvements have led to increased performance and more favorable economics. Continued development of high-rate trickling filter media has resulted in good performance and very competitive media costs. Improved understanding of trickling filter hydraulics has led to revised hydraulic loading and wetting regimes that provide better control of biofilm thickness, thereby producing a biofilm that is thinner, more aerobic, and more active. Experience with these improved facilities continues to accumulate. Experience with nitrification in trickling filter applications has caused increased interest in these options. The ongoing development of coupled TF/AS process options has resulted in improved control over effluent suspended solids concentrations, thereby allowing trickling filter based processes to produce effluent quality that rivals that of the activated sludge process.

The RF/AS and BF/AS processes have proven to be quite effective for treating higher strength industrial or industrial/municipal wastewaters. The trickling filter provides process stability and control of filamentous microorganisms, while the suspended growth bioreactor allows an effluent of excellent quality to be produced. In addition, the use of relatively small trickling filters results in a system with moderate capital and operating costs. As noted above, experience indicates that recycle of the RAS to the trickling filter (thereby converting it to a biofilter) generally results in the most complete control of solids settling characteristics. The ABF process has received relatively little use by itself, but many systems incorporate the flexibility to operate in this mode and it is used effectively to reduce energy costs during periods of lower process loading.

While basic information on the combined carbon oxidation/nitrification and separate stage nitrification processes has been available for nearly 30 years, significant interest in these processes has developed more recently, leading to successful applications.[17,45,56,60] They offer the potential for reduced energy costs, good process stability, and favorable economics.

Trickling filters have not proven to be as adaptable to nutrient removal as suspended growth bioreactors. While phosphorus removal in coupled TF/AS processes has been demonstrated, their removal capability is hindered by the oxidation of organic matter in the trickling filter.[55] Biological nitrogen removal is also problematic for the same reason. This must be considered when selecting a trickling filter based option for a new or expanded wastewater treatment facility.

19.2 FACTORS AFFECTING PERFORMANCE

Over its long history of use, a large database has been assembled describing the factors affecting the performance of the trickling filter process. Unfortunately, in many cases the data are contradictory and incomplete. This arises largely because of the interrelation between trickling filter design and operational parameters. It also arises because our understanding of the trickling filter has evolved throughout its history and continues to evolve today. Our current understanding allows recognition that effects once thought to be significant are really artifacts of past design and operational practices and are not fundamental process variables. This section discusses the primary factors that affect trickling filter performance.

19.2.1 Process Loading

The performance of any biochemical operation is affected by the process loading (i.e., the amount of substrate applied per unit time per unit mass of biomass). In suspended growth systems the process loading is expressed as the solids retention time (SRT) or the process loading factor (i.e., F/M ratio). Both have physical meaning in a suspended growth system because the biomass is well mixed and can be characterized by a single parameter. In addition, the operational parameters necessary to calculate them are easily measured. This is not true for attached growth systems such as trickling filters. The biomass cannot be characterized by a single parameter because it is not uniformly distributed through the bioreactor. Furthermore, it is not possible to easily determine the biomass

concentration within a trickling filter, thereby making it impossible to calculate either an SRT or a process loading factor. While some values for the biomass concentration in a trickling filter have been reported,[24] no consensus exists as to the appropriateness of this approach. As a consequence, other measures of process loading must be used.

As discussed in Chapter 16, substrate removal in biofilm processes is expressed by the substrate flux, J_S, which is the mass of substrate per unit time transported into and consumed by the biofilm per unit biofilm planar surface area. Logically, the biofilm process loading can be expressed in the same fashion, as the mass of substrate applied per unit time per unit of total planar biofilm area. For organic substrate, the result, referred to as the surface organic loading (SOL), λ_S, is expressed in units of kg substrate/($m^2 \cdot$ day). The SOL is also sometimes referred to as the applied flux. Another approach to expressing the loading on a trickling filter is the TOL, defined by Equation 19.1 and typically expressed in units of kg substrate/($m^3 \cdot$ day). The SOL and the TOL are related by the specific surface area of the media, a_s:

$$\lambda_S = \frac{\Lambda_S}{a_s} = \frac{F(S_{SO} + X_{SO})}{A_s}, \tag{19.2}$$

where A_s is the wetted surface area of media available for biofilm growth. Since the quantity of biomass in a trickling filter is proportional to both the media area and the media volume, both the TOL and the SOL are analogous to the process loading factor for suspended growth bioreactors. This analogy is incomplete, however, due to the fact that the biomass composition is uniform in a suspended growth system while it varies with bioreactor depth in a trickling filter, as discussed above. However, it is a useful analogy if not carried too far.

Theoretically the SOL is superior to the TOL as an expression of biofilm process loading. This is because, as discussed in Chapter 16, biofilm reactions are generally limited by mass transfer and the overall reaction rate is a function of the biofilm surface area. An inherent assumption in the concept of the SOL is that the biofilm surface area is proportional to the media surface area. A second assumption is that the substrate is dissolved and diffuses into the biofilm. While these assumptions are good ones for many biofilm processes, for some trickling filter applications they are not. This occurs for several reasons, including media plugging and inadequate wetting, the presence of colloidal organic matter in wastewater, and the occurrence of multiple substrate limitations. The result is that, for some trickling filter applications, performance correlates better with the TOL than with the SOL.

The concept of surface loading is not limited to the removal of organic substrate. Rather, it can also be applied to nitrification. For example, in a trickling filter performing only nitrification, the nitrification efficiency might be correlated with the surface ammonia-N loading (SAL) with units of kg NH_3-N/($m^2 \cdot$ day) and symbol λ_{NH}. Alternatively, one may wish to express nitrification performance in terms of the loading of TKN on the process. In that case, one might speak of a surface nitrogen loading (SNL) with units of kg N/($m^2 \cdot$ day) and symbol λ_N. These surface loadings are related to the corresponding total volumetric loadings in the same way that the SOL is related to the TOL (i.e., by Equation 19.2).

Plugging and channeling of flow in a trickling filter results in incomplete wetting and utilization of the media surface area provided, thereby making the biofilm surface area less than the media surface area. Plugging and channeling especially occur when a trickling filter is used for carbon oxidation. This is because of high biomass production rates, which cause excess biomass to accumulate in the media. The resulting incomplete media wetting and partial media plugging produce variable and incomplete utilization of the media area provided.[30] Although media with higher specific surface areas are theoretically better, beyond a point they will not result in greater biofilm surface area because the smaller openings associated with the media will cause increased plugging and reduced wetting of the available media surface area. In short, more media surface area does not necessarily

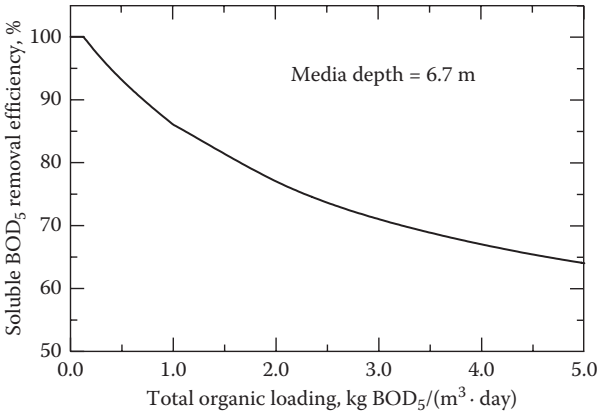

FIGURE 19.5 Typical relationship between total organic loading (TOL) and soluble BOD_5 removal efficiency for a trickling filter containing high-rate media.

mean more treatment capacity. The situation is further complicated by the fact that, in many wastewaters, a significant portion of the biodegradable organic matter is present in either a suspended or colloidal form. This organic matter is removed initially by flocculation and entrapment, just as in suspended growth systems. It is then hydrolyzed and the soluble organic matter degraded.[8,9] It is not yet clear what controls the removal rate of such organic matter in a biofilm reactor. Moreover, its retention within the trickling filter can result in additional channeling of flow.

As discussed in Section 19.1.2 and illustrated in Table 19.1, the trickling filter TOL can be correlated with its treatment objectives. Performance relationships can also be presented graphically, as illustrated in Figure 19.5 where a typical relationship between the soluble BOD_5 removal efficiency of a trickling filter and its TOL is presented. The nitrification performance of trickling filters accomplishing combined carbon oxidation and nitrification can also be correlated with the TOL, as illustrated in Figure 19.6.[17] The performance data presented in this figure are also correlated with the SOL in the upper axis (called areal loading).

In trickling filters accomplishing either carbon oxidation or combined carbon oxidation and nitrification, it is likely that oxygen is the limiting substance. As discussed in Sections 16.4 and 19.1.1, when both biodegradable organic matter and ammonia-N are present, heterotrophs and autotrophs compete for space in the aerobic portion of the biofilm, with autotrophs being excluded until the concentration of biodegradable organic matter drops below about 20 mg/L as COD. Therefore, if the rate of oxygen transfer into the biofilm is limiting, in the upper portion of the trickling filter, the oxidation rate of biodegradable organic matter will be limited by the rate of oxygen transfer, whereas in the lower portion of the trickling filter oxygen transfer will limit the oxidation rate of ammonia-N. If both substrates are expressed in oxygen units, it should be possible to correlate the rate at which oxygen demand is being satisfied with the trickling filter process loading. This has been done and the rate per unit of media volume is called the oxidation rate, Λ_{OR}.[16] It is calculated using the following equation:

$$\Lambda_{OR} = \frac{F(S_{SO} + 4.57\ S_{NO})}{V_M}, \tag{19.3}$$

where S_{SO} is the influent organic matter concentration expressed in BOD_5 units and S_{NO} is the nitrate-N concentration in the trickling filter effluent. At typical process loadings, the BOD_5 is generally equal to the oxygen required to oxidize the biodegradable organic matter applied, which is one reason that S_{SO} is expressed in those units. For trickling filters accomplishing combined carbon

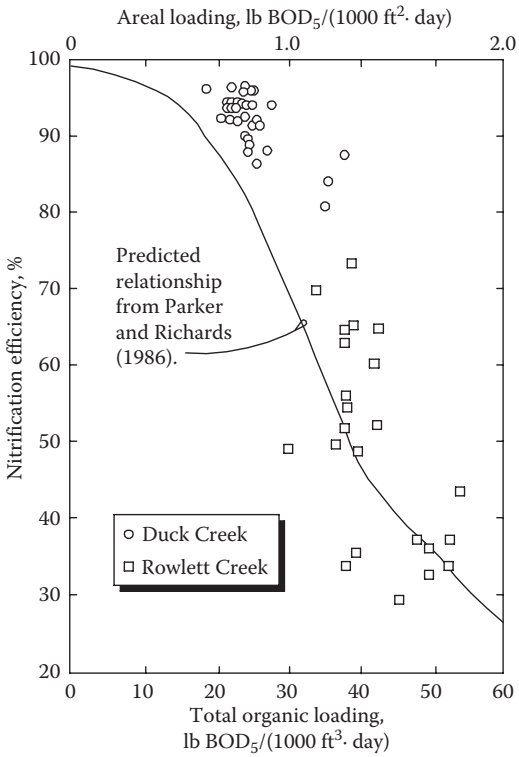

FIGURE 19.6 Effect of total organic loading on the nitrification efficiency in a trickling filter accomplishing combined carbon oxidation and nitrification. (From Daigger, G. T., Norton, L. E., Watson, R. S., Crawford, D., and Sieger, R. B., Process and kinetic analysis of nitrification in coupled trickling filter/activated sludge processes. *Water Environment Research,* 65:750–58, 1993. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

oxidation and nitrification, the oxidation rate has been observed to be relatively constant at a value between 0.5 and 1.0 kg O_2 demand/(m^3 · day) until the ammonia-N concentration drops below about 3–5 mg/L, at which point the ammonia-N concentration becomes rate limiting, causing the rate to drop.[30]

The nature of the biofilm that develops in separate stage nitrification applications is different from that which develops in carbon oxidation applications. This is because of the low yields associated with nitrifying bacteria. In fact, the biofilm that develops in these applications conforms much more closely to the theoretical assumptions upon which the models presented in Chapter 16 and Section 17.1.2 are based. Consequently, for these applications media plugging and flow channeling are less of an issue and the SOL can be used effectively to characterize the process loading. Although a number of relationships are available in the literature,[65,66] the one developed by Parker and coworkers[3,47,48] conforms most closely to the theory of Chapter 16 and will be presented in the Section 19.3.2.

Hydraulic loading rates can also affect trickling filter performance. However, the theoretical performance of a trickling filter is relatively insensitive to the THL as long as the organic loading is fixed. Practical experience confirms this prediction, once a minimum THL is achieved.[65,66] These results have been interpreted as indicating that a minimum THL is necessary to achieve complete wetting and utilization of the trickling filter media. Once this minimum THL has been achieved, further performance improvements are not observed when the hydraulic loading is increased further since the media is fully wet and active. For trickling filters using high-rate media, the minimum THL is approximately 1.8 m/hr. The minimum THL for trickling filters using rock media

is not as well defined. However, THLs used in practice typically range from 0.35 to 0.75 m/hr. As discussed below, for economic reasons, most trickling filters are operated with THLs near the minimum for the media used, and thus little information is available about maximum allowable THLs. For high-rate media, the maximum THL is most likely caused by excessive removal of the biofilm. Estimates of such limits can be obtained from the concept of Spülkraft, which is discussed in Section 19.2.7.

19.2.2 Recirculation

The term recirculation refers to the return of trickling filter effluent, either prior to or following secondary clarification, to the trickling filter influent. The primary purpose of recirculation is to uncouple the hydraulic and organic loading rates. The trickling filter media volume is selected to give the desired organic loading rate, expressed as either the SOL or the TOL. A depth is then selected, fixing the cross-sectional area of the media. The THL that will be achieved is then calculated and compared to the required minimum. If the THL is less than the minimum, either recirculation flow must be provided to make the THL equal or exceed the minimum, or a greater depth must be selected, thereby decreasing the cross-sectional area. The effects of media depth are discussed in the next section. In most cases increased recirculation will require increased pumping, thereby increasing operating costs. For this reason, recirculation in excess of that required to achieve the required minimum THL is not desirable. Moreover, as discussed in Sections 17.1.5 and 19.2.1, increased recirculation beyond the minimum required to achieve good wetting generally does not result in improved treatment. Thus, recirculation beyond that required to attain the minimum THL is not generally used unless other factors require it.

Another purpose for recirculation applies when a high-strength wastewater is being treated. As discussed in Section 17.1.6, oxygen limitations may be expected in trickling filters whenever the concentration of biodegradable organic matter exceeds about 250 mg/L as COD. This prediction is generally confirmed by practice.[65,66] Treatment performance may not necessarily be adversely impacted, but odors can be produced by the development of anaerobic conditions within the biofilm. When the concentration of biodegradable organic matter in the influent wastewater exceeds 250 mg/L as COD, it can be diluted by recirculating treated effluent. The quantity of recirculation required to dilute the applied biodegradable COD to 250 mg/L is calculated and compared to the recirculation required to achieve the minimum THL. The greater of the two is then provided.

Several recirculation options have been developed and tested over the long history of the trickling filter process.[65,66] They vary primarily in the source of the recirculation flow, the location to which it is recirculated, and, if trickling filters in series are used, whether recirculation occurs around each individually or around both. Sources of recirculation flow include both clarified and unclarified trickling filter effluent. Recirculation destinations include the trickling filter influent and the influent to an upstream primary clarifier. Recirculation of final clarifier effluent to the influent of a primary clarifier adds treated, oxygenated flow to it, which can reduce odor emissions. However, the cost of such a practice is high since the sizes of both the primary and secondary clarifiers must be increased to accommodate the recirculated flow. Recirculation of clarified effluent directly to the trickling filter has a similar drawback since the secondary clarifier must be sized to accommodate the recirculation flow, but with no compensating benefits. Consequently, the recirculation option illustrated in Figure 19.1, direct recirculation of unclarified trickling filter effluent to the trickling filter influent, is the one most frequently used today. In some instances this approach also results in a simplified pumping system as influent wastewater and recycled effluent can be combined in a single pumping system for application to the trickling filter. This option also allows sloughed biomass to be recirculated to the trickling filter, although the resulting impact on treatment efficiency is small due to the short hydraulic residence time (HRT) of the trickling filter.

19.2.3 Media Depth

Trickling filters have been constructed with a wide range of media depths. Depths are generally around 2 m for rock media, with a typical range of 1–2.5 m. Depths up to 12 m have been used for high-rate media, but a maximum depth of about 6.7 m is often used because of media structural considerations. Like recirculation, differences of opinion have existed over the years concerning the impact of media depth on trickling filter performance.[65,66] For example, a strict interpretation of the Velz, Eckenfelder, and Kornegay models discussed in Section 17.1.7 indicate that, for a constant media volume, trickling filter performance improves as the media depth is increased. Models such as those have provided the basis for constructing trickling filters with relatively large media depths (over 10 m). However, these predictions have been demonstrated to be incorrect. In contrast, the model of Logan et al.,[34] also discussed in Section 17.1.7, indicates that, for a constant media volume, trickling filter media depth has only a modest affect on trickling filter performance.

The available data are generally consistent with the predictions of the model of Logan et al.[34] and indicate that the impact of media depth on performance is relatively small for a constant TOL. Modest performance improvement is obtained by increasing media depths up to about 4 m, but performance improvements for further increases are negligible. However, maintenance of the minimum THL, as discussed in the previous two sections, is critical to maintaining consistent trickling filter performance over a wide range of media depths. This issue arises because of practical considerations. For a constant trickling filter media volume, as the media depth is decreased the media cross-sectional area (A_c) increases. Thus to maintain a constant THL as the cross-sectional area is increased, the recirculation flow rate must be increased. The interactions between TOL, media depth, and THL are illustrated by the following example.

Example 19.2.3.1

A trickling filter utilizing high-rate media is to be sized with a TOL of 1 kg $BOD_5/(m^3 \cdot day)$ to treat a wastewater with a flow rate of 5000 m^3/day and a BOD_5 concentration of 150 mg/L. The minimum acceptable THL for the media is 1.8 m/hr.

a. What media volume is required?
The required media volume can be calculated with a rearranged form of Equation 19.1:

$$V_M = \frac{(5000)(150)}{(1.0)(1000)} = 750 \text{ m}^3.$$

b. What would the THL be for a media depth of 5 m if no recirculation is used?
At a media depth of 5 m, the trickling filter cross-sectional area is

$$A_c = \frac{750}{5} = 150 \text{ m}^2.$$

The THL can be calculated with Equation 17.12 using a recirculation ratio of zero:

$$\Lambda_H = \frac{5000(1.0+0)}{150} = 33.3 \text{ m/day} = 1.39 \text{ m/hr}.$$

This is less than the minimum value of 1.8 m/hr. Consequently, the media provided will not be fully utilized and such a design would not be acceptable.

c. What recirculation rate would be required to attain the minimum THL for a media depth of 5 m?
The required THL is 1.8 m/hr = 43.2 m/day. The recirculation ratio can be calculated with a rearranged form of Equation 17.12:

$$\alpha = \frac{(43.2)(150)}{5000} - 1 = 0.30.$$

Thus, the required recirculation rate is (0.3)(5000) = 1500 m^3/day.

d. What media depth would be required to achieve a THL of 1.8 m/hr (=43.2 m/day) without recirculation?
The required cross-sectional area can be calculated with a rearranged form of Equation 17.12:

$$A_c = \frac{5000(1+0)}{43.2} = 116 \text{ m}^2.$$

To provide a total media volume of 750 m^3, the depth must be

$$L = \frac{750}{116} = 6.5 \text{ m}.$$

The choice between the two alternative designs would have to be made on the basis of economics and other such factors since they should both give similar performance.

This example illustrates the relationship between THL and media depth for a fixed media volume. Although it is recognized today that a minimum THL must be maintained to fully utilize the media and that recirculation can be used to achieve that THL, analysis of historical studies suggests that this requirement was not widely recognized in the past.[34] As a consequence, data were collected from trickling filters with various media depths and with various THLs that were below the minimum value.[10,11,31,65,66] As expected, these data indicated that trickling filters with greater media depths, and consequently greater THLs, performed significantly better than trickling filters with smaller depths. Today we recognize that this effect was largely attributable to the variation in THL and that media depth itself exerted only a minimal influence.

19.2.4 Temperature

Temperature is another factor whose impact on trickling filter performance has historically been poorly understood. Ample full-scale evidence exists demonstrating that the performance of a trickling filter can decline significantly during periods of cold weather.[5] Based on such observations, it was concluded long ago that the trickling filter process is relatively temperature sensitive. With an improved understanding of biofilm processes, however, it was realized that, in many instances, substrate removal is controlled more by mass transfer than by biological reaction.[42] Moreover, since temperature effects on mass transfer are often modest, the effect of temperature on trickling filter performance should also be relatively modest. Further analysis suggests that the observed significant effect of cold weather operation on trickling filter performance is often a result of severe temperature drops due to the physical configuration of the system. When such drops occur, the biological reaction is severely retarded, and performance does indeed suffer. Therefore, the key to good performance is to limit temperature changes to the range over which temperature has little effect and improved design concepts can be used

to do that. Of particular importance are physical design concepts to minimize heat loss during cold weather operation.

Heat loss occurs from trickling filters during cold weather by a variety of mechanisms. Conductive heat losses through the walls of the trickling filter are generally minor because the residence time of the fluid within the trickling filter is so short. Rather, the two most important heat loss mechanisms are wind effects and increased ventilation during cold weather. Experience indicates that significant heat loss can occur as the influent wastewater flows out of the distribution system, through the air, and onto the media. Several methods are available to reduce heat loss by this mechanism. They include:

- Constructing deep trickling filters with small cross-sectional areas. The reduced cross-sectional area increases the THL and reduces the need for recirculation, which increases the number of exposures of the wastewater to the cooling influence of flowing through the distributor.
- Extending the side wall upward so that it is 1.5–2 m above the distributor. This reduces the cooling effect of wind on the flow being applied to the trickling filter.
- Covering the trickling filter.

Several full-scale installations exist in which retrofitting wind screens to the top of the trickling filters or adding covers has significantly improved cold weather performance.

Ventilation control is also quite important during cold weather to minimize heat loss.[65,66] As discussed above, temperature differences between the wastewater and the ambient air cause the density of the air inside the trickling filter to differ from the density of the air outside, inducing air to flow through the trickling filter. Although wastewater temperatures vary seasonally, their variation is typically much less than the variation in ambient air temperatures, resulting in much greater temperature differentials in the winter. For example, wastewater temperatures during the summer may reach 25–30°C, which are comparable to the air temperature. In contrast, in colder climates influent wastewater temperatures seldom drop below 10°C in winter while ambient air temperatures can reach –10 to –20°C. Consequently, there is an increased density difference between the air inside and outside the trickling filter in winter, causing increased air flow through the trickling filter. This tends to increase the cooling effect as the ambient air temperature decreases.

For trickling filters with natural draft ventilation, the effect of the increased temperature difference during winter operation can be controlled by providing adjustable dampers on the air inlets. During cold weather operation the dampers are throttled to restrict air flow through the trickling filter, thereby reducing cooling affects. During warm weather the dampers are opened to increase the air flow. Covering the trickling filter provides an even greater opportunity to control air flow, as long as adjustable dampers are included in the design. Yet another approach is forced draft ventilation, which makes the air flow independent of the ambient air temperature.

A detailed discussion of current knowledge concerning the inherent affects of temperature on trickling filter performance is provided elsewhere.[66] The results are best quantified for separate stage nitrification and suggest either no noticeable affect or a modest affect represented by a temperature coefficient (θ) of at most 1.02. This affect is similar to that for rotating biological contactors, as described in Section 20.2.4. With proper temperature control, as described above, such modest affects of temperature on performance may be expected.

19.2.5 Ventilation

The resistance to air flow through a properly designed trickling filter is quite low, and thus only a small motive force is required to induce it. Natural draft ventilation operates very effectively

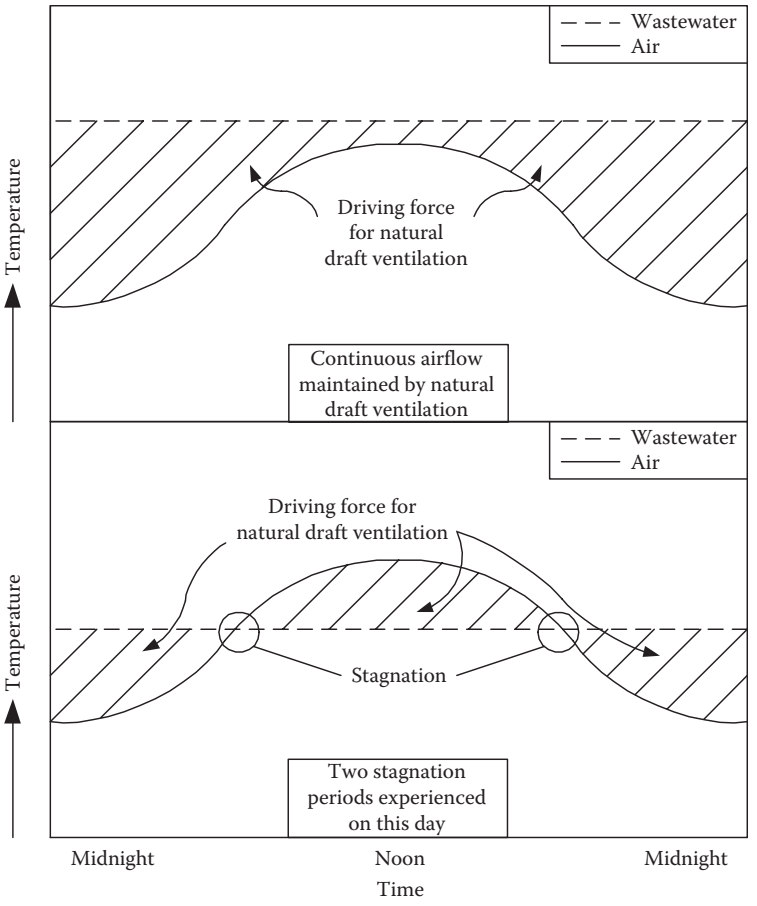

FIGURE 19.7 Effect of relative temperatures of air and wastewater on natural draft ventilation.

as long as the temperature and humidity differences between the air inside and outside the trickling filter are sufficiently large to generate the needed force. Unfortunately, instances occur in which the densities of air inside and outside the trickling filter are the same.[5,54,57] Neglect for a minute the effect of humidity differences on air density and consider only temperature differences. Wastewater temperatures are relatively constant throughout a typical day, but ambient air temperatures vary significantly. Consequently, during many days it is quite likely that the ambient air temperature will equal the wastewater temperature at least twice, as illustrated in Figure 19.7. This is particularly true during periods of moderate to warm air temperature. When this occurs, no density difference exists between the air inside and outside the trickling filter and no motive force exists to move air. The fact that both temperature and humidity affect air density complicates the phenomenon somewhat, but does not change its basic nature. As a result, during certain periods little or no air will flow through tricking filters using natural draft ventilation, negatively impacting aerobic treatment. This situation can be tolerated in many instances, particularly when incomplete treatment can be accepted on a temporary basis and when odor production is not a significant problem. In other instances, more positive control of air flow is required. This requires forced draft ventilation.

Natural draft ventilation requires the provision of sufficient open area to allow air to flow into the tricking filter, through the media, and out. For trickling filters without covers, little restricts the flow of air at the top. In contrast, air will not enter or exit the bottom of the trickling filter

unless openings with sufficient area are provided to allow air to enter the underdrain system. The clear opening between the top of the water surface and the bottom of the trickling filter media must also be sufficient. Consequently, the design of a natural draft ventilation system involves providing sufficient vent area and free distance between the water surface in the underdrain and the bottom of the media. Because the draft is generated uniformly across the entire cross-sectional area of the trickling filter by the difference in air density, uniform distribution of the air occurs naturally.

Forced draft ventilation systems overcome the difficulties associated with natural draft systems by providing positive movement of air through the trickling filter with a fan. In contrast to natural draft ventilation systems, forced draft systems require a means to distribute the air uniformly across the cross-sectional area of the trickling filter. This is usually accomplished by connecting a piping network with multiple openings to the ventilation fan, as described in Section 19.3.5.

19.2.6 Media Type

The types of trickling filter media in common use today are discussed in Section 19.1.2, compared in Table 19.2, and illustrated in Figure 19.3. There are several opinions about the relative merits of these media.[14,43,50,52] Daigger and Harrison[14] conducted a comprehensive investigation of several of the media operated over a wide range of organic loading rates. Their results indicated that all of the tested media gave significant removal of organic matter, but that some did better, both statistically and practically. Interestingly, some media provided superior performance at low TOLs, whereas others provided superior performance at high TOLs. Good performance was provided by rock media at low TOLs but not at high TOLs. At low TOLs the best performance was provided by 60° cross-flow (XF) media. In contrast, fully corrugated vertical media (VFC) provided the best performance at high TOLs. Performance of both the random (RA) and horizontal (HO) media was inferior to that of the bundle media at all TOLs. The superior performance of XF media at low TOLs has been confirmed by several other researchers, but the superior performance of VFC media at high organic loading rates has been contested by others. While results such as these are significant, it must be recognized that nearly all media testing has been conducted at pilot scale. It has been suggested that pilot-scale results may not accurately reflect full-scale performance due to the difficulty of simulating full-scale wastewater distribution systems in pilot-scale facilities.[65,66]

Operational differences have also been noted among the various trickling filter media. Some of the more significant ones are as follows:[4,14,65,66]

- A relatively thick and biologically diverse biofilm typically develops on rock media. It can be several mm thick, and it consists of a rich array of bacteria, Eucarya such as protozoa and rotifers, and macroinvertebrates such as worms and fly larvae. These thick biofilms result in a high degree of anaerobic activity. It is thought by some that this causes lower net biomass production rates from rock media trickling filters. The low void volume and irregular pathways through rock media may make it more susceptible to plugging than high-rate media.
- The 60° cross-flow media possesses good flow redistribution characteristics, particularly in comparison to vertical media. This may be one of the reasons that superior performance is observed for cross-flow media at low TOLs. It may also result in increased plugging potential, particularly at high TOLs.
- Random media possesses good flow redistribution characteristics. However, localized dry spots and areas of ponding have been observed, suggesting areas of incomplete utilization. Anaerobic activity may develop within the ponded areas, contributing to odors.

The dry areas also provide locations for the growth of nuisance organisms, particularly Psychoda flies.

- Relatively thin biofilms develop on horizontal wood media when applied in conventional trickling filter applications. When RAS is applied in an ABF or BF/AS application, however, a significant amount of interstitial growth develops, which also contributes to organic matter removal.

These differences in operational characteristics should be considered when selecting a media for a particular application.

19.2.7 Distributor Configuration

Several wastewater distribution systems have been used, including fixed nozzles with or without periodic dosing and rotary distributors with or without speed control. Experience indicates that the distributor type significantly affects the hydraulic flow pattern and the biofilm thickness within the trickling filter. Both of these factors significantly affect trickling filter performance.

Figure 19.8 illustrates the two types of distributors typically used in trickling filters, rotary and fixed nozzle. Generally, rotary distributors give better performance than fixed nozzle distributors.[65,66] This occurs for a variety of reasons. One is more uniform flow distribution. Flow is applied quite uniformly to the section of a trickling filter over which a rotary distributor is passing. Since the distributor passes repetitively over the trickling filter, on an average basis wastewater is applied uniformly over its surface. Theoretically, uniform distribution of wastewater can also be obtained with a fixed nozzle distributor. However, experience indicates that uniform distribution is very difficult, if not impossible, to achieve. Fixed nozzles must emit a fine spray to provide uniform application over the trickling filter. This generally requires relatively small openings. However, since most wastewaters contain some particulate matter, small openings are prone to plugging, which disrupts flow distribution. Consequently, fixed nozzle designs generally represent a compromise between flow distribution and plugging potential. The poor operating characteristics of fixed nozzle distributors can be mitigated to a certain extent by using higher recirculation flows to increase the applied THL. Hydraulic loading rates used with fixed nozzle distributors are often two to three times higher than those used with rotary distributors. The minimum THL of 1.8 m/hr discussed above for high-rate media is that required for rotary distributors. For fixed nozzle

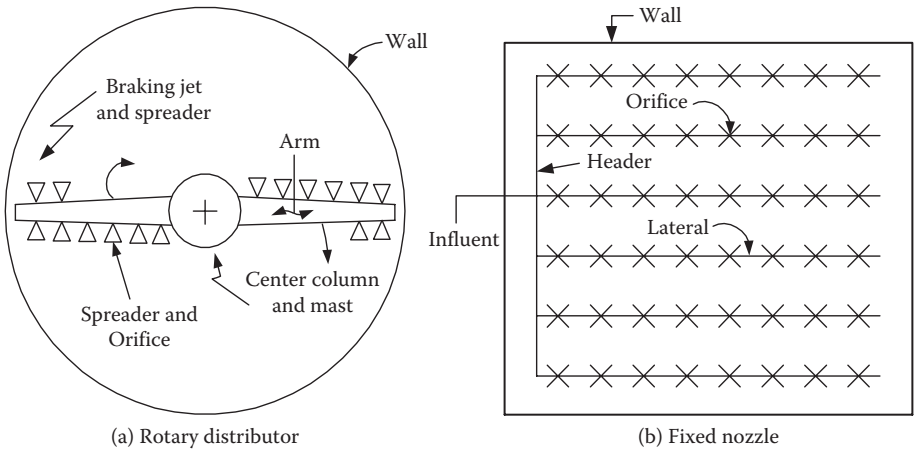

FIGURE 19.8 Schematic diagrams of rotary and fixed nozzle distributors.

distributors, the minimum THL is on the order of 4–6 m/hr. Even with these provisions, significant plugging potential still exists for many fixed nozzle distributor designs.

Another advantage of rotary distributors is that the instantaneous hydraulic loading rate that occurs as the distributor passes over a section of the trickling filter is significantly higher than the average THL. This high instantaneous loading rate can produce a flushing effect that removes excess biomass from the trickling filter, thereby maintaining a thinner, more active biofilm. This does not occur when fixed nozzle distributors are used.

The advantages of periodic dosing achieved with rotary distributors in circular trickling filters can also be attained with reciprocating distributors in square or rectangular trickling filters. While rotary distributors have proven to be quite reliable from a mechanical perspective, much less experience exists with reciprocating distributors. As a consequence, most trickling filters are circular units with rotary distributors.

A concept being tested and applied in full-scale applications is Spülkraft, SK, which is the depth of water in mm applied to the trickling filter in one passage of the distributor arm.[49,65,66] For a rotary distributor, the Spülkraft is calculated as follows:

$$SK = \frac{\Lambda_H (1000)}{N_a \cdot \omega_d (60)}, \tag{19.4}$$

where Λ_H is the THL in m/hr, N_a is the number of arms on the distributor, and ω_d is the distributor rotational speed in rev/min. A low SK is generally used for normal operation and a high value is used for periodic flushing to remove excess biomass and maintain a thin, active biofilm.[1,2,26–28,36,58,65,66] Table 19.5 presents SK values that are suggested based on current experience and judgment.[65,66]

Most rotary distributors use hydraulic propulsion. The energy of the influent flow is discharged in a horizontal direction, causing the arm to rotate in the opposite direction. The rotational speed can be controlled by discharging some of the influent from the opposite side of the distributor arm, but the range that can be achieved is limited. An alternative approach is the power driven distributor, which uses an electric motor to control rotational speed.[65,66] Significantly greater

TABLE 19.5
Suggested Spülkraft Values

TOL	Spülkraft, SK, mm/pass	
kg BOD_5/(m^3 · Day)	Design	Flushing
0.25	10–100	≥200
0.50	15–150	≥200
1.00	30–200	≥300
2.00	40–250	≥400
3.00	60–300	≥600
4.00	80–400	≥800

Note: Adapted from Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, *Manual of Practice No. 8*, 4th ed., Water Environment Federation, Alexandria, Virginia, 1998; Water Environment Federation, *Aerobic Fixed-Growth Reactors*, Water Environment Federation, Alexandria, Virginia, 2000.

control of rotational speed, and consequently greater control of Spülkraft, is provided by this system.

19.2.8 Wastewater Characteristics

As with all biochemical operations, the characteristics of the wastewater being treated affect the performance of a trickling filter. The more easily biodegradable the wastewater, the higher the organic loading that can be applied while still achieving acceptable effluent quality. There are limits of course, because ultimately, the rate of oxygen transfer will control the rate at which organic matter can be removed and oxidized.

Just as in the activated sludge process, various wastewater components are removed by a variety of mechanisms. Readily biodegradable substrate is removed by diffusion through the biofilm to microorganisms that biodegrade it. Slowly biodegradable substrates are initially removed by flocculation and entrapment mechanisms, just like in the activated sludge process. They are then hydrolyzed by extracellular enzymes before they are biodegraded. The presence of particulate material will adversely impact the capacity of a trickling filter to remove soluble organic matter because when particulate matter is incorporated into the biofilm, it displaces active biomass from the aerobic active biofilm layer.[53] This is also analogous to the activated sludge process where particulate organic matter accumulates in the mixed liquor suspended solids (MLSS).

In contrast to the activated sludge process, the COD/TKN ratio of the wastewater significantly affects tricking filter nitrification efficiency.[65,66] This occurs because of the spatial distribution of microorganisms within the trickling filter, as discussed in Section 19.1.1 and illustrated in Figure 19.2.

19.2.9 Effluent Total Suspended Solids

The overall performance of a trickling filter system depends not only on the removal of soluble and particulate organic matter, but also on the concentration of suspended solids in the system effluent, and this has been the factor that has historically limited system performance. Relatively complete removal of soluble organic matter can be obtained at appropriate loadings, but the concentration of suspended solids may not be low enough to meet performance objectives. In many instances, trickling filter effluents contain finely divided, colloidal suspended solids that settle poorly in secondary clarifiers. This characteristic led to development of coupled TF/AS processes. Coupling a suspended growth bioreactor with a trickling filter introduced a suspended biomass capable of flocculating and removing the colloidal suspended solids contained in the trickling filter effluent.

The concentration of suspended solids in the effluent from a trickling filter system is dependent on the settleability of the suspended solids. Many factors affect that settleability, including the TOL, the characteristics of the wastewater (particularly the proportion of suspended and/or colloidal matter), and the THL. Biological flocculation of suspended solids in the trickling filter effluent generally improves as the TOL is reduced. Poorer flocculation is generally observed from roughing filter applications than for carbon oxidation applications. The best flocculation is obtained with combined carbon oxidation and nitrification applications. This is consistent with experience with the activated sludge process wherein reduced process loading (i.e., higher SRT) results in increased flocculation, as discussed in Section 11.2.1. Improved flocculation can be obtained by using hydraulic loadings that maintain a thin, active biofilm. A thin biofilm can be maintained by proper control of the SK factor, as discussed in Section 19.2.7.

The biodegradable organic matter in the effluent from a trickling filter system is made up of both soluble and particulate material. The concentration of particulate biodegradable organic matter, in turn, is a function of the suspended solids concentration and its biodegradability. The biodegradability of trickling filter effluent suspended solids, as reflected by the BOD_5, is a function of the

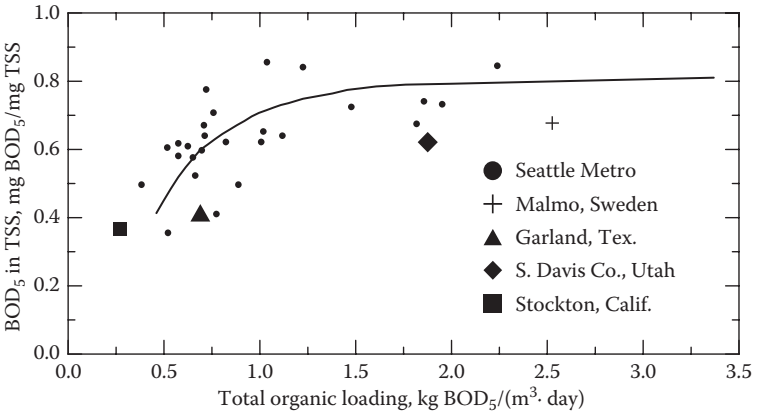

FIGURE 19.9 Effect of total organic loading (TOL) on the BOD_5 of suspended solids leaving a trickling filter. (From Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

TOL, as illustrated in Figure 19.9.[65,66] Consequently, the higher the TOL, the lower the suspended solids concentration must be in the system effluent to meet treatment criteria.

Example 19.2.9.1

A trickling filter receiving a wastewater with a total BOD_5 of 150 mg/L achieves 90% removal of soluble BOD_5 by being operated at a TOL of 0.75 kg $BOD_5/(m^3 \cdot day)$. What will be the BOD_5 of the clarified effluent if it contains 30 mg/L of suspended solids? Assume that the soluble BOD_5 concentration of the influent wastewater is 65% of the total BOD_5.

a. First calculate the soluble BOD_5 in the effluent.
   The soluble BOD_5 in the influent wastewater is (0.65)(150) = 97.5 mg/L. Since 90% of it is removed, the soluble BOD_5 concentration in the trickling filter effluent is 9.75 mg/L.
b. Next estimate the BOD_5 due to particulate matter in the effluent.
   Entering Figure 19.9 with a TOL of 0.75 kg $BOD_5/(m^3 \cdot day)$, it can be seen that the BOD_5/TSS ratio is 0.5. Since the suspended solids concentration in the effluent is 30 mg/L, the particulate BOD_5 is (0.5)(30) = 15 mg/L.
c. The total BOD_5 is just the sum of the soluble and particulate BOD_5 values.
   Total BOD_5 = 9.75 + 15 = 24.75 mg/L.

19.3 PROCESS DESIGN

Although considerable progress has been made in our understanding of substrate transport and microbial growth in biofilms, as discussed in Chapter 16, the application of that understanding to trickling filter design has been slow. This is primarily because trickling filters are considerably more complicated than the conceptual models used to describe biofilms. For example, we saw earlier in this chapter that flow through a trickling filter is both intermittent and highly irregular, whereas biofilm models commonly assume steady flow at a constant rate. Thus, while mechanistic models such as those in Section 17.1.2 have helped us develop a better understanding of trickling filter performance, they have not found application in practice. Rather, the design of trickling filters is based primarily on empirical correlations and simplified models depicting pilot- and full-scale system performance. Furthermore, because BOD_5 has been used historically to measure the concentration of biodegradable organic matter in those systems, many design relationships are typically presented

in terms of it without regard to whether the organic matter is soluble or particulate. Finally, many relationships express the performance of the trickling filter system (i.e., the bioreactor plus the settler that follows it). Although this approach has had a negative impact on our ability to isolate and understand the fundamental processes involved, it has worked satisfactorily when used by experienced designers. Thus, it represents the current state-of-the-art and is the approach that we use here. However, it is likely that as more basic data are collected from large-scale systems that better approaches will evolve.

This section reviews several procedures for sizing trickling filters for a variety of applications, criteria for sizing the ventilation system, and approaches for sizing coupled TF/AS processes. When sizing a trickling filter, consideration must be given to the nature of the media being used. Trickling filters containing rock and random pack high-rate media can be built to any depth consistent with the structural constraints of the media, as discussed earlier in this chapter. Bundle high-rate media, on the other hand, is modular, and economics dictate that the depth conform to an integer number of modules. One module of such media typically is 0.61 m high, so the depth should be in increments of 0.61 m. Furthermore, as discussed in Section 19.2.3, the maximum unsupported depth to which bundle media can be stacked is about 6.7 m, or 11 modules high. These constraints must be considered when selecting the depth of a trickling filter. Table 19.6 summarizes the general approach to sizing a trickling filter process. Specific applications of this approach are illustrated below.

19.3.1 Sizing Trickling Filters with Black-Box Correlations

Historically, "black-box" correlations (i.e., those simply correlating output with input) of performance data from full-scale trickling filter applications have been used to size rock trickling filters. As discussed in Section 17.1.7, two such correlations are those developed by the National Research Council (NRC)[41] and by Galler and Gotaas.[21] As also discussed in that section, great care should be exercised when black-box correlations are used. The data set upon which they are based should be reviewed carefully by the design engineer to ensure that it is similar to the proposed application. If significant differences exist, then the equation should not be used.

TABLE 19.6
Summary of Trickling Filter Process Design Procedure

1. Summarize process design and loading conditions including maximum, minimum, and average sustained temperature; maximum, minimum, average, and design process influent flows and pollutant loadings; and desired effluent quality.
2. Where necessary, convert process loadings, expressed in conventional parameters (such as BOD_5) into the units used in the process design (such as biodegradable COD).
3. Select an effluent quality goal. As discussed in Section 10.4.4, selection of that goal should consider uncertainty and variability in process performance.
4. Select a design approach appropriate for the selected effluent quality goal (roughing, carbon oxidation, combined carbon oxidation and nitrification, separate stage nitrification) and media type.
5. Use the selected design approach to calculate the required media volume.
6. Calculate media depth to maintain necessary minimum THL based on influent flow.
7. Check media depth to determine whether calculated value is within the acceptable range. If not, adjust media depth and determine the need for recirculation to maintain minimum acceptable THL.
8. If a coupled trickling filter/activated sludge process is to be used, select the appropriate SRT.
9. Calculate the waste sludge production rate and, based on the selected SRT and the desired MLSS concentration, calculate the suspended growth basin volume.
10. Calculate the suspended growth oxygen requirement.
11. Summarize the results in tabular form.

The NRC correlation was based on performance data from 34 rock media trickling filters operating at military installations.[41] It reflects the impact of TOL and recirculation on performance. The influent BOD_5 concentrations at the plants were relatively high (generally 200 mg/L or greater). Consequently, the correlation may not accurately reflect the performance of rock media trickling filters treating lower strength wastewater. The Galler-Gotaas[21] correlation is based on 322 observations at typical municipal wastewater treatment plants. It recognizes the importance of organic loading, hydraulic loading, recirculation, media depth, and temperature on system performance. Both the NRC and the Galler-Gotaas correlation equations were developed on the basis of the BOD_5 concentration in the process influent and the secondary clarifier effluent. Consequently, they implicitly incorporate the impact of both the trickling filter and the secondary clarifier. When using either it must be assumed that the secondary clarifier is properly sized and operated. Because both black-box correlations are used for sizing rock media trickling filters and few are being designed today, neither is presented here. However, they are presented in standard design manuals.[65,66]

19.3.2 Sizing Trickling Filters with Loading Factor Relationships

The process loading factor approach to trickling filter design reflects the fact that trickling filter performance is generally correlated with the TOL or the SOL, as discussed in Section 19.2.1. The steps required when using this approach include: selection of an appropriate TOL or SOL based on experience and/or on performance correlations such as those shown in Figures 19.5 and 19.6; use of the selected relationship to calculate the required media volume; and use of typical media depths and hydraulic loadings, as described in Sections 19.2.1, 19.2.2, and 19.2.3, to dimension the trickling filter. The following examples illustrate the use of the loading factor relationships to design a roughing filter, a carbon oxidation application, and a trickling filter achieving combined carbon oxidation and nitrification. The first example considers a roughing filter.

Example 19.3.2.1

Size a trickling filter for a roughing filter application in which 70% of the soluble BOD_5 is to be removed. The wastewater flow rate is 5000 m^3/day, and the total BOD_5 concentration is 150 mg/L. Bundle media is to be used with a minimum THL of 1.8 m/hr. Assume that the relationship between soluble BOD_5 removal and TOL shown in Figure 19.5 is applicable.

a. What TOL should be used to achieve the treatment objective?
   From Figure 19.5, a TOL of 2.5 kg $BOD_5/(m^3 \cdot day)$ will achieve the treatment objective.
b. What media volume is required?
   The media volume can be calculated with a rearranged form of Equation 19.1:

$$V_M = \frac{(5000)(150)}{(2.5)(1000)} = 300 \text{ m}^3.$$

c. What media depth is required to maintain the minimum allowable THL of 1.8 m/hr with no recirculation?
   The media depth is just the volume divided by the cross-sectional area. The area can be obtained from the definition of the THL, given by Equation 17.12. For $\alpha = 0$ and $\Lambda_H = 1.8$ m/hr (=43.2 m/day):

$$A_c = \frac{5000}{43.2} = 115.7 \text{ m}^2.$$

$$L = \frac{300}{115.7} = 2.6 \text{ m}$$

While this is acceptable, as discussed in Section 19.2.3, performance is generally reduced for media depths less than about 4 m. Thus, a greater depth should be used, which is acceptable since that will give a smaller cross-sectional area, thereby giving a greater THL. The depth of a typical sheet media bundle is 0.61 m. If we take 4 m as being the shortest acceptable depth and recognize that an integer number of bundles must be used, then the media depth should be at least 4.27 m, corresponding to seven layers of media.

d. What THL would result if the trickling filter were 4.27 m high?

$$A_c = \frac{300}{4.27} = 70.4 \text{ m}^2$$

$$\Lambda_H = \frac{5000}{70.4} = 71 \text{ m/day} = 2.96 \text{ m/hr}.$$

This is an acceptable THL.

This example illustrates that, for roughing filter applications, the media depth may be established at minimum design values. It also illustrates that relatively high THL rates are sometimes used.

The following example uses the process loading factor approach to size a trickling filter for relatively complete removal of biodegradable organic matter, such as for secondary treatment.

Example 19.3.2.2

Size a trickling filter for 90% removal of soluble BOD_5. As in Example 19.3.2.1, the wastewater flow rate is 5000 m^3/day, the total BOD_5 concentration is 150 mg/L, and bundle media with a minimum THL of 1.8 m/hr is to be used. Assume that the relationship between soluble BOD_5 removal and TOL shown in Figure 19.5 is applicable.

a. What media volume is required?
From Figure 19.5 it can be seen that a TOL of 0.75 kg $BOD_5/(m^3 \cdot \text{day})$ must be used to achieve 90% removal of soluble BOD_5. The media volume can be calculated with a rearranged form of Equation 19.1:

$$V_M = \frac{(5000)(150)}{(0.75)(1000)} = 1000 \text{ m}^3.$$

b. If a media depth of 6.7 m is used, which is the maximum unsupported depth of plastic sheet bundle media and corresponds to 11 layers of bundles, what recirculation flow rate is required to maintain a THL of 1.8 m/hr (43.2 m/day)?
The recirculation ratio, α, can be calculated with Equation 17.12 after calculating the cross-sectional area of the trickling filter.

$$A_c = \frac{1000}{6.7} = 149.2 \text{ m}^2$$

$$\alpha = \frac{(43.2)(149.2)}{5000} - 1 = 0.29.$$

Thus, the recirculation flow rate is (0.29)(5000) = 1445 m^3/day, making the total flow rate to the trickling filter 6445 m^3/day.

This example illustrates that even though the trickling filter media depth is established at the maximum safe media depth from a structural perspective, a small amount of recirculation may still be required to maintain the minimum required THL. This often occurs when relatively high removal of biodegradable organic matter is needed.

The next example uses the surface loading approach to size a trickling filter for combined carbon oxidation and nitrification.

Example 19.3.2.3

Size a trickling filter to accomplish combined carbon oxidation and nitrification of the wastewater used in the preceding two examples. The treatment goal is to remove more than 85% of the influent ammonia-N. Bundle media with a specific surface area of 100 m^2/m^3, a minimum THL of 1.8 m/hr, and a maximum depth of 6.7 m is to be used.

a. What would an appropriate TOL be?
   Figure 19.6 shows that an SOL of 0.002 kg $BOD_5/(m^2 \cdot day)$ is required to achieve a nitrification efficiency of greater than 85%. Equation 19.2 can be used to convert this to a TOL for media with a specific surface area of 100 m^2/m^3:

$$\Lambda_S = (100)(0.002) = 0.2 \text{ kg } BOD_5/(m^3 \cdot day).$$

b. Calculate the trickling filter media volume.

$$V_M = \frac{(5000)(150)}{(0.2)(1000)} = 3750 \text{ m}^3.$$

   The media volume could also have been calculated directly from the SOL and the media specific area. At a BOD_5 loading of 750 kg BOD_5/day and an SOL of 0.002 kg $BOD_5/(m^2 \cdot day)$, a total media surface area of 375,000 m^2 would be needed. Since the specific surface area of the media is 100 m^2/m^3, the total media volume is 3750 m^3.

c. What recirculation flow rate is required to maintain a THL of 1.8 m/hr (43.2 m/day) at a media depth of 6.7 m?
   For this application, the cross-sectional area is

$$A_c = \frac{3750}{6.7} = 560 \text{ m}^2.$$

   This can be used in Equation 17.12 to calculate the recirculation ratio, α:

$$\alpha = \frac{(43.2)(560)}{5000} - 1 = 3.84.$$

   Thus, the recirculation flow rate must be (3.84)(5000) = 19,190 m^3/day.

This example illustrates that a large media volume may be needed when a high level of treatment is required. This can result in a high recirculation requirement.

Table E19.1 compares the trickling filter sizes required for the applications considered in Examples 19.3.2.1 through 19.3.2.3. The roughing application (Example 19.3.2.1) achieves partial removal of biodegradable organic matter and requires a relatively small trickling filter of moderate depth, operating with a relatively high THL (3.0 m/hr) without recirculation. Removal of

TABLE E19.1
Comparison of Results for Examples 19.3.2.1–19.3.2.3

Application	TOL $\frac{kg\ BOD_5}{m^3 \cdot Day}$	Depth m	Surface Area m^2	Flow Rate m^3/Day	
				Influent	Recirculation
Roughing	2.50	4.27	70.4[a]	5000	0
Secondary treatment	0.75	6.70	149.2[b]	5000	1445
Carbon oxidation and nitrification	0.20	6.70	560.0[b]	5000	19,190

[a] THL = 3.0 m/hr
[b] THL = 1.8 m/hr

biodegradable organic matter to the level of secondary treatment (Example 19.3.2.2) requires a larger trickling filter with a greater depth, operating at the minimum THL (1.8 m/hr) with a modest amount of recirculation. Finally, combined carbon oxidation and nitrification (Example 19.3.2.3) requires an even larger trickling filter, although at the same depth, operating at the minimum THL (1.8 m/hr) with a large amount of recirculation. Capital costs increase as the degree of treatment is increased because larger trickling filters must be constructed. Operating costs also increase because the influent must be pumped to a higher elevation and because more recirculation flow must be pumped.

Next consider the design of a trickling filter for separate stage nitrification. As discussed in Section 19.2.1, separate stage nitrification conforms more closely than other trickling filter applications to the theoretical models of Chapter 16 and Section 17.1.2. Consequently, it is best characterized by the surface loading approach. The design methodology proposed by Parker and coworkers[3,47,48] provides the basis for the approach discussed here.

In the upper regions of a trickling filter accomplishing separate stage nitrification the rate of ammonia-N oxidation will be controlled by the rate of oxygen transfer into the biofilm. This occurs because the ammonia-N concentration in the liquid phase is relatively high whereas the dissolved oxygen concentration is limited to the solubility of oxygen in water (in the range of 7–12 mg/L for typical wastewater temperatures). Because approximately 4.3 mg of O_2 are required for the oxidation of 1 mg of ammonia-N, and because the half-saturation coefficient for oxygen is relatively high, oxygen will be the limiting reactant for ammonia-N concentrations above about 3 to 5 mg-N/L. However, in the lower regions of the trickling filter the rate of ammonia-N oxidation will become limited by the ammonia-N concentration. This situation can be characterized by the following expression:

$$J_{NH} = J_{NH,max}\left(\frac{S_{NH}}{K_{g,NH} + S_{NH}}\right), \tag{19.5}$$

where J_{NH} is the ammonia-N flux into the biofilm [g NH_3-N/($m^2 \cdot$ day)], $J_{NH,max}$ is the maximum flux when ammonia-N is not limiting, S_{NH} is the ammonia-N concentration in the bulk liquid, and $K_{g,NH}$ is a pseudo half-saturation coefficient similar to that employed by Kornegay[32] to account for both mass transport and the degree of saturation on the reaction rate in a trickling filter (see Section 17.1.7). The value of the maximum flux is determined by the rate of oxygen transfer into the biofilm.

Several methods are available for obtaining the parameters in Equation 19.5. One is through the analysis of operating data from similar, full-scale applications. Another is to operate a pilot plant and to analyze the resulting performance data. Parker and coworkers[3,47,48] have proposed the use of

the trickling filter oxygen transfer model of Logan et al.,[35] along with corrections for temperature, the suspended solids concentration in the trickling filter influent, and the THL. With this approach $K_{g,NH}$ must still be selected independently.

Use of the surface loading approach is simply an extension of the approach presented above, but with consideration of the effects of the bulk liquid ammonia-N concentration on the ammonia-N flux into the biofilm. Because the flux equation is approximate, it is typically applied by dividing the trickling filter into a series of incremental elements within which the ammonia-N concentration, and hence the flux, can be considered to be constant. First, an ammonia-N concentration is selected above which the flux can be considered to be constant. The media volume required to reduce the influent concentration to that value is then determined by assuming that the flux is $J_{NH,max}$ and using the approach illustrated in Example 19.3.2.3. The remainder of the trickling filter is then broken into elements within which the concentration change is small, allowing the volume of each increment to be calculated for the flux corresponding to the average concentration within the increment. This is continued down the trickling filter until the desired effluent concentration is attained. The total media volume required is just the sum of the incremental media volumes. The media depth and THL are then specified and the trickling filter dimensions are selected using the same approaches used in Examples 19.3.2.1 through 19.3.2.3. As discussed in Section 19.1.2, higher density media (specific surface area typically 140 m^2/m^3) are usually used for separate stage nitrification applications to increase the volumetric treatment efficiency. This can be done because the risk of plugging and channeling is less with this application.

Example 19.3.2.4

A separate stage nitrification trickling filter is to be sized to treat a high quality secondary effluent with a flow rate of 5000 m^3/day and an ammonia-N concentration of 20 mg-N/L. The required effluent ammonia-N concentration is 1 mg-N/L. Pilot testing resulted in the selection of a value for $J_{NH,max}$ of 1.5 g NH_3-N/(m^2 · day) and a value for $K_{g,NH}$ of 1.5 mg-N/L for design purposes. Since most of the biodegradable organic matter has been removed, thereby reducing the chances of plugging, media with a specific surface area of 140 m^2/m^3 will be used to reduce the size of the unit. In addition, a minimum THL of 1.8 m/hr will be maintained.

a. Assuming that the flux will be constant at $J_{NH,max}$ for ammonia-N concentrations of 5 mg-N/L or higher, what media volume is needed to reduce the concentration from 20 mg-N/L to 5 mg-N/L?
   The mass removal rate of ammonia-N in the first increment is just the flow rate times the required concentration change, or (5000)(20–5) = 75,000 g NH_3-N/day. Since the maximum flux is 1.5 g NH_3-N/(m^2 · day), the required media surface area is 75,000 ÷ 1.5 = 50,000 m^2. If the specific surface area of the media is 140 m^2/m^3, the required media volume is 50,000 ÷ 140 = 357 m^3.
b. What additional media volume is required to reduce the ammonia-N concentration to 1 mg-N/L, the desired effluent value?
   For this analysis, divide the remainder of the trickling filter into zones within which the bulk liquid ammonia-N concentration is reduced by 0.5 mg-N/L (i.e., from 5 to 4.5 mg-N/L, and so on). Use the average ammonia-N concentration within each zone to calculate the flux of ammonia-N into the biofilm in that zone, and then use that flux to calculate the required media volume for the zone. For illustrative purposes consider a reduction in the ammonia-N concentration from 5 to 4.5 mg-N/L. The average ammonia-N concentration for this zone will be 4.75 mg-N/L, and the average flux of ammonia-N into the biofilm will be (Equation 19.5):

$$J_{NH} = 1.5\left(\frac{4.75}{1.5+4.75}\right) = 1.14 \text{ g } NH_3\text{-N/}(m^2 \cdot \text{day}).$$

TABLE E19.2
Computation of the Media Volume Required to Reduce the Concentration of Ammonia-N from 5.0 to 1.0 mg/L in Example 19.3.2.4

Zone mg-N/L	Average S_{NH} mg-N/L	Average J_{NH} g NH_3-N/(m^2 · Day)	Media Volume m^3
4.5–5.0	4.75	1.14	15.7
4.0–4.5	4.25	1.11	16.1
3.5–4.0	3.75	1.07	16.7
3.0–3.5	3.25	1.03	17.3
2.5–3.0	2.75	0.97	18.4
2.0–2.5	2.25	0.90	19.8
1.5–2.0	1.75	0.81	22.0
1.0–1.5	1.25	0.68	26.3

Then, the media volume will be

$$V_M = \frac{(5000)(5-4.5)}{(1.14)(140)} = 15.7 \text{ m}^3.$$

Continue this calculation for each of the zones and tabulate the results. They are shown in Table E19.2. Adding the volumes of the individual zones gives the total media volume required to reduce the ammonia-N concentration from 5 to 1 mg-N/L. It is 152.3 m^3. This compares to a volume of 357 m^3 to reduce the ammonia-N concentration by 15 mg-N/L, from 20 to 5 mg-N/L.

c. What total media volume would be required to reduce the ammonia-N concentration from 20 to 1 mg-N/L?
It is the sum of the two media volumes calculated above, or 357 + 152 = 509 m^3.

d. What media depth would be required if no recirculation was used?
The media depth is just the volume divided by the cross-sectional area. The area can be obtained from the definition of the THL, given by Equation 17.12. For $\alpha = 0$ and $\Lambda_H = 1.8$ m/hr (=43.2 m/day):

$$A_c = \frac{5000}{43.2} = 115.7 \text{ m}^2$$

$$L = \frac{509}{115.7} = 4.4 \text{ m}.$$

At 0.61 m/module, this depth is equivalent to 7.2 modules. Since whole modules must be used, eight will be required. This corresponds to a total media depth of 4.9 m, which gives a corresponding cross-sectional area of 104 m^2. The slightly smaller area will result in a modest increase in the THL, which is acceptable. The diameter of the trickling filter would be 11.5 m.

19.3.3 Sizing Trickling Filters with the Modified Velz/Germain Equation

An empirical model that has been referred to in the literature as either the modified Velz or the Germain equation has been used extensively to size trickling filters for removal of organic matter.[22,62,65,66] The Velz equation is described in Section 17.1.7. The development of the Velz/

Germain equation assumes first-order removal of substrate with respect to both the substrate concentration and the biomass concentration. It also assumes that the biomass concentration is constant throughout the trickling filter and that the trickling filter behaves as a plug-flow reactor with a residence time, τ. Based on these assumptions, the concentration of soluble substrate in the effluent, S_{Se}, is related to the concentration of soluble substrate actually applied to the top of the trickling filter, S_{Sa}, by

$$\frac{S_{Se}}{S_{Sa}} = \exp(-k \cdot X_{B,Hf} \cdot \tau), \tag{19.6}$$

where k is a reaction rate coefficient and $X_{B,Hf}$ is the biomass concentration. However, empirically it was found that the HRT is related to the trickling filter THL and the media depth, L, according to

$$\tau = \frac{C \cdot L}{\Lambda_H^n}, \tag{19.7}$$

where C and n are empirical coefficients. Combining these two equations and combining k, C, and $X_{B,Hf}$ into a new coefficient K, called the treatability coefficient, gives:

$$\frac{S_{Se}}{S_{Sa}} = \exp\left(\frac{-K \cdot L}{\Lambda_H^n}\right). \tag{19.8}$$

When recirculation is not used, the applied soluble substrate concentration is just the concentration in the process influent wastewater. However, when recirculation is used, the influent is diluted by the recirculated effluent, making the applied substrate concentration less than the concentration in the untreated wastewater. In that case, the applied substrate concentration is given by Equation 17.1. Combining Equations 19.8, 17.1, and 17.12, which defines the THL, gives the most common form of the modified Velz/Germain equation:

$$\frac{S_{Se}}{S_{Sa}} = \left\{(1+\alpha)\exp\left[\frac{K \cdot L}{\left(\frac{F(1+\alpha)}{A_c}\right)^n}\right] - \alpha\right\}^{-1}. \tag{19.9}$$

This equation is dimensional, with the value of K depending on the value of n and the units used for flow rate, F, cross-sectional area, A_c, and depth, L. Consequently, when a value of K is taken from the literature, particular attention must be paid to the units used and those units must be retained. Since K is a reaction rate coefficient, its value is temperature dependent. That dependency is typically expressed by Equation 3.99, with the temperature coefficient θ set to a value of 1.035. Sometimes Equation 19.9 is made explicitly dependent on temperature by substituting Equation 3.99 directly into it. Finally, in many applications of the Velz/Germain equation, the term F/A_c is referred to as the unit wastewater application rate.

As discussed in Section 17.1.7, in the Eckenfelder[19,20] model the treatability coefficient, K, is expressed as $K_1 a_s$ where K_1 is another treatability coefficient and a_s is the media specific surface area. As discussed previously in this chapter, this may be appropriate for applications where

channeling of flow and media plugging are not likely to be important, but is inappropriate when they are. The Eckenfelder equation has received less use than the modified Velz/Germain equation and, consequently, the empirical data base from which to select appropriate model coefficients is much more limited. This reduces its practical utility. However, when sufficient data are available it may be a superior model with which to correlate full-scale or pilot-scale data.

Although the modified Velz/Germain and the Eckenfelder equations were developed theoretically, it is now recognized that their theoretical basis is flawed. Consequently, they must be viewed as empirical models that are best used to correlate performance data. Nevertheless, because the modified Velz/Germain equation has been widely used, a significant empirical database exists, and it has been used to estimate a variety of trickling filter performance characteristics.[65,66] Like the NRC[41] and Galler–Gotaas[21] equations, it has been used to characterize the removal of organic matter by a trickling filter and secondary clarifier system. In these instances the performance of the secondary clarifier is implicitly incorporated into the treatability coefficient. It has also been used to characterize the removal of organic matter across just the trickling filter itself. Furthermore, in some of those cases the relationship between the concentration of soluble organic substrate in the influent and effluent from the trickling filter is characterized, whereas in others the relationship between the total concentration of biodegradable organic matter in the influent and the concentration of soluble organic matter in the effluent is characterized. Due to this variation in usage, it is important that the basis for treatability coefficients taken from the literature be identified.

The available data[65,66] indicate that the value of n in the modified Velz/Germain equation varies with the trickling filter media type and application, ranging from 0.3 to 0.7. However, it is a standard practice by some to use a value of 0.5 for all situations so that the relative treatability of various wastewaters can be assessed by the numerical value of the treatability coefficient. Although this approach results in some sacrifice in precision, it does allow easy comparison of the relative performance of various trickling filter applications.

Experience indicates that a number of adjustments must be made to allow the modified Velz/Germain equation to accurately characterize the performance of full-scale trickling filter applications. Adjustments must be made for the depth of the trickling filter media and for the influent waste strength. A correction can also be made if the THL is below the minimum required value of 1.8 m/hr. The need for these adjustments is related to the fact that this model is not fundamental in nature. As discussed above, the best use of the modified Velz/Germain equation is to correlate full-scale or pilot-scale performance data. It can be used to interpolate within a particular data set, but caution should be exercised if performance estimates are to be extrapolated beyond the existing data base. Detailed descriptions of the use of the modified Velz/Germain equation to predict the performance of full-scale trickling filter applications are presented elsewhere.[65,66]

Examination of Equation 19.9 reveals that it contains eight terms: the wastewater flow rate, F; the wastewater soluble substrate concentration, S_{SO}; the desired effluent soluble substrate concentration, S_{Se}; the treatability coefficient, K; the exponent, n; the media depth, L; the cross-sectional area, A_c; and the recirculation ratio, α. In a design setting F, S_{SO}, S_{Se}, K, and n are all known, leaving three unknowns. Since there is one equation and three unknowns, two will be free design variables and can be chosen at will, allowing the other to be calculated. One approach is to choose the recirculation ratio and the cross-sectional area to set the THL within the allowable range, allowing the media depth, L, to be calculated. Another is to specify the depth, typically at the maximum unsupported depth for the particular media, and then investigate the effects of the recirculation ratio on the area required to maintain the minimum THL. Since total systems costs depend on the media volume used, the amount of recirculation pumped, and the height to which the wastewater and recirculation flow must be pumped, the opportunity exists for the designer to seek an optimal design that minimizes the system costs.

Example 19.3.3.1 illustrates use of the modified Velz/Germain equation to design a trickling filter for the removal of biodegradable organic matter.

Example 19.3.3.1

Consider the wastewater that was used in the examples of Section 19.3.2, for which 65% of the BOD_5 is soluble. Use the modified Velz/Germain equation to size a trickling filter to produce an effluent with a soluble BOD_5 concentration of 10 mg/L, which is essentially the same effluent quality achieved in Example 19.3.2.2. Assume a value for K of 0.4 $(m/hr)^{0.5}/m$ (which is equivalent to the value of 0.075 $gpm^{0.5}/ft$ often reported in the literature[65,66]) and a value for n of 0.5. The temperature is 20°C. The characteristics of the media are such that the THL must be maintained at a value of 1.8 m/hr or greater, the depth of one bundle is 0.61 m, and the unsupported depth must be no greater than 6.7 m (11 bundles).

a. What is the influent soluble BOD_5 concentration?
   Since the total BOD_5 in the influent is 150 mg/L and 65% is soluble, the soluble BOD_5 concentration is 97.5 mg/L.
b. What media volume is required if no recirculation is used and the cross-sectional area is selected to maintain a THL of 1.8 m/hr (43.2 m/day)?
   The required cross-sectional area can be calculated from the definition of THL, given by Equation 17.12:

$$A_c = \frac{5000}{43.2} = 115.7 \text{ m}^2.$$

   The media depth can be calculated with Equation 19.9 after substituting all of the known values. To be consistent with the units of K, F should be expressed in units of m^3/hr. Thus, $F = 208 \text{ m}^3/\text{hr}$:

$$\frac{10}{97.5} = \left\{(1+0)\exp\left[\frac{(0.4)L}{\left(\frac{208(1+0)}{115.7}\right)^{0.5}}\right] - 0\right\}^{-1}.$$

   Taking the inverse of each side gives:

$$97.5 = \exp\,[0.298\ L]$$

   and

$$L = \frac{\ln(9.75)}{0.298} = 7.64 \text{ m}.$$

   This exceeds the typical maximum value of 6.7 m. Furthermore, this represents 12.5 bundles, suggesting that the depth would have to be increased to 13 bundles (7.93 m), exceeding the maximum by two bundles. Thus, such a design would be questionable and the designer would have to carefully investigate the media characteristics before adopting it. If it were used, the total media volume would be (115.7)(7.93) = 917 m^3.
c. If the depth is limited to 6.7 m and the THL is maintained at the minimum value of 1.8 m/hr, what recirculation rate and cross-sectional area are required?
   Again, Equation 19.9 must be used. Recognizing that the term $F(1 + \alpha)/A_c$ is just the THL, which has a value of 1.8 m/hr, gives:

$$\frac{10}{97.5} = \left\{(1+\alpha)\exp\left[\frac{(0.4)(6.7)}{(1.8)^{0.5}}\right] - \alpha\right\}^{-1}.$$

Simplifying and taking the inverse of both sides gives:

$$9.75 = (1 + \alpha)(7.37) - \alpha.$$

Solving for α gives:

$$\alpha = 0.37.$$

Thus, the recirculation rate is (0.37)(5000) = 1850 m^3/day.
For this α, the trickling filter cross-sectional area is

$$A_c = \frac{(5000)(1+0.37)}{43.2} = 159 \text{ m}^2.$$

The total media volume is (159)(6.7) = 1065 m^3, which is quite similar to the volume for Example 19.3.2.2, which had similar design criteria. Comparing this result to the deeper trickling filter sized in part b, this represents an increase in volume of 1065 – 917 = 148 m^3, or 16%.

The above example illustrates one of the controversial aspects of the modified Velz/Germain equation. It predicts that the required media volume decreases as the media depth increases, which is in contrast with actual experience as discussed in Section 19.2.3. As discussed previously in this section, the value of K is often adjusted for media depth. The correction typically used is[65,66]

$$K_2 = K_1\left(\frac{L_1}{L_2}\right)^{0.5}, \tag{19.10}$$

where the subscripts represent two different depths. This further illustrates the empirical nature of the modified Velz/Germain equation and the fact that care must be exercised in its use. Nevertheless, the extensive database available makes the modified Velz/Germain equation useful for design purposes.

19.3.4 The Model of Logan, Hermanowicz and Parker

The model of Logan et al.,[34,35] discussed in Section 17.1.7, represents a new generation of more fundamental trickling filter models. It is based on mass transfer principles for soluble biodegradable organic matter and oxygen. It also accounts for the multicomponent nature of most wastewaters by dividing the organic material into five categories, each with different diffusional characteristics. The removal of each category is modeled by assuming the presence of a thick biofilm (i.e., one where the biofilm thickness exceeds that required for aerobic metabolism) and laminar flow over an inclined surface. Substrate removal occurs by transport in the thin liquid film that flows over the media and into the biofilm. The effects of media configuration on performance are predicted based on differences in flow patterns through the various media and their impact on substrate transport in the liquid film. This model provides no coupling between the removal of organic material in the trickling filter and oxygen transfer. Rather, it limits the maximum substrate flux to a value consistent with the oxygen transfer capability of the system.

The model of Logan et al.[34,35] is mentioned to alert the reader to its existence and to the fact that research is ongoing to develop more fundamental trickling filter design models. As discussed previously, this model has been used successfully to predict the oxygen transfer rate, and hence the zero-order flux of ammonia-N, in separate stage nitrifying trickling filters. Although some day

fundamental models will be used routinely to characterize the performance of full-scale trickling filters, insufficient experience currently exists with the model of Logan et al.[34,35] to conclude that it can be broadly used for the design of full-scale trickling filters.[29]

19.3.5 Ventilation System

Proper design of the ventilation system is necessary to maintain aerobic conditions within a trickling filter. The pressure drop for air flowing through a trickling filter is typically low, often less than 1 mm of water per m of media depth. Due to this low pressure drop, only a small motive force is necessary to cause air to flow. For forced draft ventilation systems, the low pressure drop results in relatively low power requirements for the air supply fans, generally on the order of a few kW even for large trickling filters. The head on the ventilation fan will generally be less than 1.5 cm of water. The relatively low pressure drop for air flowing through a trickling filter also affects the design of the air distribution system.

Adequately sized underdrains are required to allow air to flow into or out of the bottom of the trickling filter. To ensure that air can move to and from the center of the trickling filter, underdrains are sized so that they are submerged no more than 50% at the peak hydraulic flow.[65,66] In addition, natural draft ventilation systems need a sufficiently large vent area to allow the air to flow to or from the underdrains. Typical criteria for trickling filters using plastic sheet media are the vent area per unit of media volume or the vent area per unit of peripheral length. Values such as 1–2 m^2/1000 m^3 of media or 0.022–0.033 m^2/m of tower periphery are often used.[65,66] Another criterion for rock media trickling filters is a vent area equal to 15% of the trickling filter cross-sectional area.[65,66]

The considerations for sizing a forced draft ventilation system are somewhat different. First, air must be distributed equally across the entire cross section of the trickling filter. This differs from a natural draft ventilation system where the air flow will naturally distribute itself, as discussed in Section 19.2.5. Natural drafting will tend to occur in a trickling filter using forced draft ventilation, resulting in short-circuiting, if the distribution of air across the trickling filter cross section is not uniform. In fact, the natural draft tendency in a trickling filter with forced draft ventilation can result in air flowing up in one portion of the trickling filter and down in another. The air flow distribution system for a forced draft ventilation system consists of a piping network with a series of openings located across the cross section of the trickling filter. These openings are sized so that the air flow rate through each is the same, thereby resulting in uniform air flow distribution. A velocity of 0.3–0.6 m/sec is also typically maintained as another measure to provide good air flow distribution. Air flow requirements are generally calculated based on process oxygen requirements and the oxygen transfer efficiency, which are typically on the order of 2%–10%. Forced draft ventilation systems can be designed as either up flow or down flow systems. Down flow systems can be designed without covers, while covers are generally required for up flow systems.

19.3.6 Coupled Trickling Filter/Activated Sludge Processes

As discussed in Section 19.1, coupled TF/AS processes consist of three components: a trickling filter, a suspended growth bioreactor, and a secondary clarifier. The trickling filter is sized using the procedures for sizing a stand-alone trickling filter. The secondary clarifier is sized in the same manner as one used with the activated sludge process. Since coupled TF/AS processes typically provide good control over the growth of filamentous microorganisms, the secondary clarifier will produce a thick RAS for recycle to the suspended growth bioreactor. A further consideration arises with TF/SC systems, however, because the primary function of the suspended growth bioreactor in them is to flocculate the poorly settleable suspended solids leaving the trickling filter. Experience indicates that the biological floc produced in some of these systems is relatively weak, possibly because few filamentous bacteria are present to serve as a backbone, as discussed in Section 11.2.1.

TABLE 19.7
Combinations of TOL and SRT Used to Achieve Good Bioflocculation in Coupled TF/AS Systems

System	TOL kg $BOD_5/(m^3 \cdot Day)$	SRT Days
RF/AS and BF/AS	3–4	3.0
RF/AS and BF/AS	2–2.5	2.0
TF/AS	0.6–1.0	1.0
ABF	<0.6	<1.0

Consequently, flocculating inlet wells are often provided in the secondary clarifiers used with TF/SC processes.[46] Such inlet wells can also be provided with other coupled TF/AS processes, but they are generally considered to be less necessary.

The design of the suspended growth bioreactor requires consideration of both the removal of biodegradable organic matter and the establishment of conditions necessary for good flocculation. As with activated sludge systems, the latter often controls the design.[39] Nevertheless, the trickling filter and the suspended growth bioreactor must also be viewed as a system and designed accordingly. A portion of the organic matter contained in the influent wastewater will be metabolized in the trickling filter, and the remainder must be metabolized in the suspended growth bioreactor. Likewise, a certain degree of flocculation will occur in the trickling filter, and the suspended growth bioreactor must be sized to achieve the remaining flocculation required. In general, the lower the TOL on the trickling filter, the greater the metabolism of organic matter and the higher the degree of flocculation that occurs there. This results in reduced requirements for organic matter metabolism and flocculation in the suspended growth bioreactor. Likewise, application of a higher TOL to the trickling filter results in a greater need for organic matter stabilization and flocculation in the suspended growth bioreactor. Table 19.7 provides general guidance concerning the relationship between the trickling filter TOL and the suspended growth bioreactor SRT required to achieve good flocculation in coupled TF/AS processes.

In coupled TF/AS processes the SRT is calculated in the same fashion as for the activated sludge process. It is the mass of suspended solids in the suspended growth bioreactor divided by the rate at which suspended solids are wasted from the system, either intentionally or unintentionally. At steady state, of course, the rate at which suspended solids are wasted from the process must equal the waste solids production rate, which is determined by the biological reactions occurring in both the trickling filter and the suspended growth bioreactor. Consequently, waste solids production calculations must consider the pollutant loadings placed on the entire system, not just the organic matter contained in the trickling filter effluent.[13,25] Biomass grown in the trickling filter will slough off and pass into the suspended growth bioreactor. Such growth must be accounted for in the total process waste solids production calculation. It is common to estimate solids production using a net process yield, $Y_{n,T}$, as discussed in Section 10.4.1. Furthermore, even though soluble substrate may remain in the trickling filter effluent, it is all typically removed in the suspended growth bioreactor. Consequently, the mass of MLSS in the suspended growth bioreactor can be calculated with Equation 10.2 and the solids wastage rate can be calculated with Equation 10.3. In both cases, the term $(S_{SO} + X_{SO})$ should reflect the concentrations of organic matter entering the system. Less information is available about net process yields in coupled TF/AS systems than is available for activated sludge systems. However, because the retention of biomass in a trickling filter increases as the TOL is decreased, the value of $Y_{n,T}$ typically is influenced more by the TOL on the trickling filter than by the SRT of the suspended growth bioreactor and decreases as the TOL is decreased. Furthermore, values are typically on the order of 0.7–0.9 mg TSS/mg BOD_5. The following example illustrates this approach for the design of a TF/SC system.

Example 19.3.6.1

A high quality effluent in terms of effluent BOD_5 and suspended solids is desired for the trickling filter sized in Example 19.3.2.2. Size a solids contact unit to achieve this goal, thereby converting the system into a coupled TF/SC system. Assume that the net process yield, $Y_{n,T}$, has a value of 0.70 mg TSS/mg BOD_5 and that the MLSS concentration in the suspended growth bioreactor is 2500 mg/L.

a. What SRT value would be appropriate for this application?
   The TOL used for the trickling filter in Example 19.3.2.2 is 0.75 kg $BOD_5/(m^3 \cdot day)$. Thus, from Table 19.7, a suspended growth bioreactor with an SRT of 1.0 day would be appropriate.
b. What should the volume of the suspended growth bioreactor be?
   From Example 19.3.2.2, the flow rate is 5000 m^3/day and the total BOD_5 concentration is 150 mg/L (150 g/m^3). The required reactor volume can be calculated with Equation 10.2 by making use of the fact that the desired MLSS concentration is 2500 mg/L (2500 g/m^3):

$$V = \frac{(1.0)(0.70)(5000)(150)}{2500} = 210 \text{ m}^3.$$

   This gives an HRT of 1.0 hr. The acceptability of this size from a mixing energy and oxygen transfer perspective would have to be verified using the procedures discussed in Section 11.2.5.
c. What is the excess solids production rate?
   This can be calculated with Equation 10.3:

$$W_M = (5000)(0.70)(150) = 525{,}000 \text{ g/day} = 525 \text{ kg/day}.$$

A similar approach can be used to design a coupled RF/AS system, as illustrated in the following example.

Example 19.3.6.2

The roughing filter sized in Example 19.3.2.1 is to be used in a coupled RF/AS process. What size suspended growth bioreactor is required? Assume that the net process yield, $Y_{n,T}$, has a value of 0.80 mg TSS/mg BOD_5 and that the MLSS concentration in the suspended growth bioreactor is 2500 mg/L. The net process yield is higher than in the preceding example because the TOL on the trickling filter is higher. Also assume that because the subject wastewater is readily biodegradable, flocculation will control the design.

a. What SRT value would be appropriate for this application?
   From Example 19.3.2.1, the TOL is 2.5 kg $BOD_5/(m^3 \cdot day)$. Consequently, from Table 19.7, an SRT of two days is necessary to obtain good flocculation.
b. What should the volume of the suspended growth bioreactor be?
   From Example 19.3.2.1, the flow rate is 5000 m^3/day and the total BOD_5 concentration is 150 mg/L (150 g/m^3). The required reactor volume can be calculated with Equation 10.2 by making use of the fact that the desired MLSS concentration is 2500 mg/L (2500 g/m^3):

$$V = \frac{(2.0)(0.80)(5000)(150)}{2500} = 480 \text{ m}^3.$$

   This gives an HRT of 2.3 hr. The acceptability of this size from a mixing energy and oxygen transfer perspective would have to be verified using the procedures discussed in Section 11.2.5.

c. What is the excess solids production rate?
This can be calculated with Equation 10.3:

$$W_M = (5000)(0.80)(150) = 600{,}000 \text{ g/day} = 600 \text{ kg/day}.$$

More solids are produced than in the TF/SC system because the TOL on the roughing filter is much higher, thereby making $Y_{n,T}$ higher and the SRT of the suspended growth bioreactor is not large enough to reduce it.

While both process kinetics and flocculation must be considered when sizing a coupled TF/AS process, the criteria presented in Table 19.7 will typically control the process size since flocculation is generally the governing event. The concentration of soluble, biodegradable organic matter in the coupled TF/AS process effluent can be calculated by using the specific growth rate of the biomass in the suspended growth bioreactor, just as with other suspended growth bioreactors. However, for coupled TF/AS processes a significant fraction of the biodegradable organic matter contained in the process influent wastewater will be removed in the trickling filter, and this will result in a significant input of microorganisms into the suspended growth bioreactor. This must be considered in the calculation of the specific growth rate if accurate predictions of the effluent soluble substrate concentration are to be made. The effect of influent biomass on the specific growth rate in a suspended growth bioreactor is discussed in Section 5.2.3, and the results are presented as Equation 5.62. It is repeated here, with the only modification being that the source of the heterotrophic biomass in the suspended growth bioreactor influent is identified:

$$\mu_H = \frac{1}{\Theta_c} + b_H - \frac{X_{B,H,T,TFE}}{\tau \cdot X_{B,H,T}}, \tag{19.11}$$

where $X_{B,H,T,TFE}$ is the heterotrophic biomass concentration, in TSS units, in the trickling filter effluent resulting from a single pass of the wastewater over the trickling filter. In other words, it is the concentration that would result if the trickling filter were the only biochemical operation being used. All other symbols refer to the suspended growth bioreactor. As illustrated, the feed of microorganisms into the suspended growth bioreactor reduces the specific growth rate, thereby reducing the effluent substrate concentration below the value that would be obtained in a system with the same SRT but with no biomass input.

If the characteristics of the trickling filter effluent (i.e., S_{Se} and $X_{B,H,T,TFE}$) could be defined, the relationships presented in Section 5.2.3 could be used to size the suspended growth bioreactor of a coupled TF/AS system. Unfortunately this cannot be done easily. First, we saw in Section 19.3.1 that trickling filter design procedures focus on the removal of substrate rather than on the growth of biomass. Thus, while S_{Se} may be well defined, $X_{B,H,T,TFE}$ is not. If an attempt was made to estimate biomass growth by using the true growth yield alone and neglecting cell decay, $X_{B,H,T,TFE}$ would be overestimated, thereby causing the substrate removal capability of the suspended growth bioreactor to be overestimated. Second, the suspended solids in the effluent from the trickling filter consist of heterotrophic biomass, cell debris, inert influent suspended solids, and unmetabolized substrate, with their relative quantities depending on the characteristics of the influent wastewater, as well as on the TOL and THL of the trickling filter. Precise prediction of the concentrations of these constituents, which are required for use of the equations in Section 5.2.3, is not currently possible. Third, both aerobic and anaerobic metabolism can occur within the biofilm of a trickling filter. Although the outer portion of the biofilm is aerobic, the inner portion may be anaerobic. Moreover, the relative importance of aerobic and anaerobic metabolism will vary depending on the nature and concentration of the biodegradable organic matter in the influent wastewater, the availability of oxygen, and hydraulic conditions affecting biofilm thickness. Since yields are quite different

under aerobic and anaerobic conditions, biomass production can vary significantly. This further complicates prediction of the concentrations of various types of suspended solids in the trickling filter effluent. Finally, biomass can accumulate within a trickling filter and be sloughed periodically, as discussed in Section 19.4. This results in time-variant concentrations of biomass and other particulate constituents in the trickling filter effluent. In many instances sloughing cycles occur over the course of several days, or even several weeks, a time interval that can significantly exceed the suspended growth bioreactor SRT. In fact, significant variations in suspended growth bioreactor MLSS concentrations have been observed as a result of trickling filter sloughing cycles.[13,25] These variations can affect the performance of the suspended growth bioreactor in a significant manner. Consequently, the suspended growth bioreactor is typically sized by using the net process yield approach as illustrated in Examples 19.3.6.1 and 19.3.6.2.

Some of the difficulties discussed above can be avoided when the focus is on nitrification because then only the autotrophic biomass concentration need be known. As a result, the relationships presented in Section 5.2.3 have been used successfully to characterize the removal of ammonia-N in a coupled TF/AS process accomplishing combined carbon oxidation and nitrification.[6,15] The concentration of nitrifiers in the trickling filter effluent was estimated as the concentration of ammonia-N nitrified in the trickling filter multiplied by the nitrifier true growth yield. This was permissible because there is little difference between the true growth yield and the observed yield for autotrophic bacteria. This concentration was used, along with Equation 19.11, to predict the nitrifier specific growth rate in the suspended growth bioreactor, allowing estimation of the effluent ammonia-N concentration. The performance relationship developed is presented in Figure 19.10

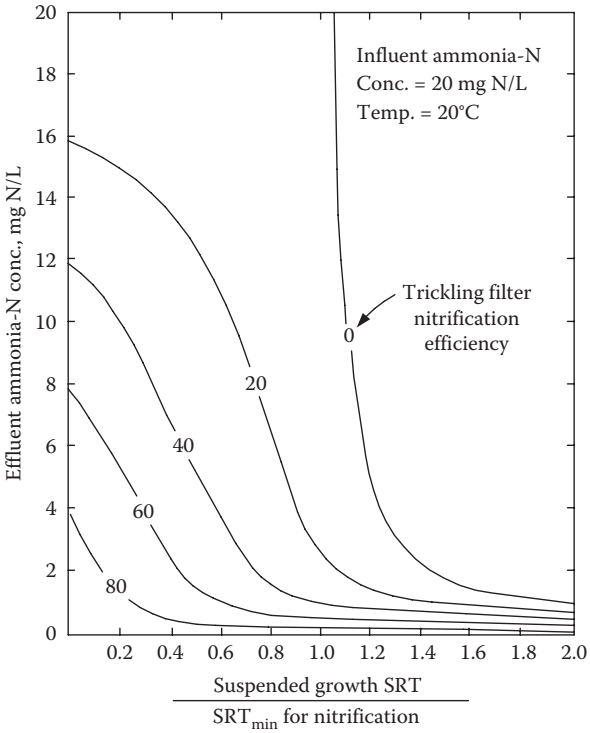

FIGURE 19.10 Effect of the SRT in a suspended growth bioreactor and the nitrification efficiency in an upstream trickling filter on the effluent ammonia-N concentration from a coupled TF/AS system. (From Daigger, G. T., Norton, L. E., Watson, R. S., Crawford, D., and Sieger, R. B., Process and kinetic analysis of nitrification in coupled trickling filter/activated sludge processes. *Water Environment Research,* 65:750–58, 1993. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

where the effluent ammonia-N concentration is plotted as a function of the suspended growth bioreactor SRT divided by the nitrifier minimum SRT. Several curves are presented corresponding to different trickling filter nitrification efficiencies. As seen, the sloughing of nitrifiers from the trickling filter allows the suspended growth bioreactor to maintain nitrification even when operating at a nitrification design factor that would otherwise cause washout of the nitrifiers. Furthermore, the greater the seeding effect (i.e., the greater the ammonia-N conversion in the trickling filter), the lower the effluent ammonia-N concentration. The potential adverse effects of sloughing from the upstream trickling filter on the validity of this approach have been discussed in the literature.[12,44] Nevertheless, the results suggest that more fundamental procedures may be developed in the future for the design of suspended growth bioreactors in coupled TF/AS processes.

If the characteristics of the trickling filter effluent are characterized sufficiently well, the procedures of Section 5.2.3 can also be used to calculate the oxygen requirement in the suspended growth bioreactor of a coupled TF/AS process. Just as with any suspended growth process, the oxygen requirements must be compared with the energy input required for mixing and the larger of the two selected. The relationship between oxygen and mixing requirements differs among the various coupled TF/AS processes.

The primary function of the suspended growth bioreactor in a TF/SC process is flocculation. Consequently, the TF/SC suspended growth bioreactor will generally be mixing limited. In contrast, substantial stabilization of biodegradable organic matter occurs in the suspended growth bioreactor of BF/AS or RF/AS systems so that the energy input may be determined either by oxygen or mixing requirements. The difficulty, in any case, is in determining the degree of stabilization of biodegradable organic matter in the upstream trickling filter.

If the trickling filter effluent cannot be characterized sufficiently well to allow the relationships of Section 5.2.3 to be used to calculate the suspended growth bioreactor oxygen requirements, empirical correlations can be used. It will be recalled from Section 10.4.1 that the process oxygen stoichiometric coefficient, Y_{O_2}, is often used with Equation 10.4 to estimate the oxygen requirement for the activated sludge process in the absence of other data. Figure 10.8 shows how the value of that parameter varies with the SRT. A similar approach can be used to estimate the oxygen requirement in coupled TF/AS systems, except that an equation expresses the effect of SRT (or F/M ratio, U) on the process oxygen stoichiometric coefficient, rather than a figure. Based on studies of several full-scale coupled TF/AS processes, Harrison[23] used the following equation to estimate the overall oxygen stoichiometric coefficient that would occur in the suspended growth bioreactor in the absence of the trickling filter:

$$Y_{O_2} = Y_{O_2,s} + \frac{Y_{O_2,d} \cdot b_H}{U}, \tag{19.12}$$

where $Y_{O_2,s}$ is the oxygen stoichiometric coefficient for synthesis, taken equal to 0.6 mg O_2/mg BOD_5; $Y_{O_2,d}$ is the oxygen stoichiometric coefficient for decay, taken equal to 1.2 mg O_2/mg volatile suspended solids (VSS); b_H is the decay coefficient, taken equal to $(0.115)(1.025)^{T-20}$ day^{-1}, where T is the temperature of the mixed liquor in the suspended growth bioreactor in °C; and U is the F/M ratio, based on the process influent BOD_5 loading and the mixed liquor volatile suspended solids (MLVSS) inventory in the suspended growth bioreactor. In addition, Harrison[23] developed the following equation relating the oxygen stoichiometric coefficient for the trickling filter, $Y_{O_2,TF}$, to its TOL, Λ_S, expressed in units of kg $BOD_5/(m^3 \cdot day)$:

$$Y_{O_2,TF} = \frac{0.25}{\Lambda_S} + 0.254. \tag{19.13}$$

The oxygen stoichiometric coefficient for the suspended growth bioreactor, $Y_{O2,SG}$, in a coupled TF/AS system can be estimated as the difference between Y_{O2}, as calculated with Equation 19.12, and $Y_{O2,TF}$, as calculated with Equation 19.13:

$$Y_{O_2,SG} = Y_{O_2} - Y_{O_2,TF} \tag{19.14}$$

The oxygen requirement in the suspended growth bioreactor can then be estimated by multiplying $Y_{O_2,SG}$ by the mass of BOD_5 entering the coupled TF/AS system per unit time, as indicated by Equation 10.4. The following example illustrates the technique.

Example 19.3.6.3

What is the oxygen requirement for the suspended growth bioreactor in the RF/AS system sized in Example 19.3.6.2, as estimated using the procedure of Harrison? Assume that the temperature is 25°C and that the MLSS is 75% volatile.

a. What is the F/M ratio for the process based on the organic matter entering the system and the mass of MLSS in the suspended growth bioreactor?
The F/M ratio can be calculated with Equation 5.48 by extending it to account for the particulate contribution to the total BOD_5. From Example 19.3.6.2, the MLSS concentration is 2500 mg/L and the bioreactor volume is 480 m^3. Because the MLSS is 75% volatile, the MLVSS concentration is (0.75)(2500) = 1875 mg/L = 1875 g/m^3. The wastewater flow rate is 5000 m^3/day and the influent BOD_5 concentration is 150 mg/L = 150 g/m^3. Therefore,

$$U = \frac{(5000)(150)}{(1875)(480)} = 0.83 \text{ kg } BOD_5/\text{kg VSS}.$$

b. What would the overall process oxygen stoichiometric coefficient be if no trickling filter was present?
This can be calculated with Equation 19.12 after substituting the appropriate values for the oxygen stoichiometric coefficients for synthesis and decay:

$$Y_{O_2} = 0.6 + \frac{(1.2)\left[(0.115)(1.025)^{T-20}\right]}{0.83} = 0.788.$$

c. What is the oxygen stoichiometric coefficient for the trickling filter?
This can be calculated with Equation 19.13. The roughing filter TOL is 2.5 kg $BOD_5/(m^3 \cdot \text{day})$. Therefore:

$$Y_{O_2,TF} = \frac{0.25}{2.5} + 0.254 = 0.354.$$

d. What is the oxygen stoichiometric coefficient for the suspended growth bioreactor?
The oxygen stoichiometric coefficient for the suspended growth bioreactor is just the difference between the overall and trickling filter oxygen stoichiometric coefficients, as reflected by Equation 19.14:

$$Y_{O_2,SG} = 0.788 - 0.354 = 0.434.$$

e. What is the oxygen requirement for the suspended growth bioreactor?
The oxygen requirement for the suspended growth bioreactor can be calculated with Equation 10.4 using the value of $Y_{O_2,SG}$ calculated in part c above and the influent flow rate and BOD_5 concentration given in part a:

$$RO_{SG} = (5000)(0.434)(150) = 326{,}000 \text{ g } O_2/\text{day} = 326 \text{ kg } O_2/\text{day}.$$

One final point. Since flocculation rather than the removal of organic matter often controls the SRT required for the suspended growth bioreactor, a contact and sludge reaeration configuration similar to the contact/stabilization process can often be used.[46] This approach reduces the total volume of the suspended growth bioreactor.

19.4 PROCESS OPERATION

One of the often stated advantages of the trickling filter process is that it provides stable, reliable performance with relatively little operator attention. Operation with little attention is possible because treatment capacity is determined primarily by the volume of media provided and its configuration. Consequently, daily operation typically consists of maintaining pumps and equipment. As a result, the perception exists that little can be done from an operational perspective if the desired degree of treatment is not being achieved; rather, the only solution is to add additional media volume. To a certain extent, this is true. However, in recent years it has been discovered that a greater degree of operational control is afforded by the process than originally thought and that proper operation is required to achieve optimum performance. This is particularly true for the TF/AS processes. This section provides an overview of trickling filter operation. Additional, detailed information is provided for the interested reader in various operational manuals of practice.[64,67]

19.4.1 Typical Operation

Experience indicates that the performance of a trickling filter can be improved significantly by proper control of the hydraulic regime applied to it. Section 19.2.7 discussed the results obtained with the use of mechanically driven distributors. For routine operation, a relatively rapid speed gives optimum treatment. The speed can be reduced periodically, however, causing increased hydraulic shear, with the associated sloughing of excess biomass from the media. This concept has been used to control trickling filter biofilm thickness.

Several approaches are available for temporarily increasing the hydraulic loading on a trickling filter. One that has worked well with a variety of high-rate media is to turn on all of the influent pumps while directing the flow to only one of the trickling filters. Increasing the hydraulic loading by a factor of 2.5–3 times for a period of about an hour is generally sufficient to slough off excess biomass. The frequency with which this should be done can be determined by periodically increasing the hydraulic loading and monitoring the trickling filter effluent. Samples can be collected every few minutes and examined visually. If dark colored (generally black) suspended solids are discharged from the trickling filter during the flushing event, the flushing frequency should be increased. On the other hand, if dark solids are not discharged, the flushing frequency can be decreased. A flushing frequency on the order of once per week is often found to be appropriate.

Another approach has been used with rock media trickling filters. A rope is tied to the end of the rotary distributor and used to restrain it. This reduces the rotational speed, causing the same effect as reducing the speed of a mechanically driven distributor. This practice has been referred to as “walking” the trickling filter. The flushing of dark, anaerobic solids from the trickling filter indicates the need for an increased flushing frequency, just as described above.

Proper control of the ventilation system is important to maintain aerobic conditions at all times and minimize heat loss during cold weather operation. As discussed in Section 19.2.4, the openings on natural draft ventilation systems must be reduced during cold weather operation to reduce air flow and the resulting heat loss. In contrast, openings must be increased as much as possible during more temperate conditions to encourage adequate air flow.

Odors have frequently been detected from trickling filters. The reasons for this are many and varied, and may be complex for any particular installation. Good ventilation is one factor in odor control, along with control of biofilm thickness to minimize anaerobic activity in it. Odors will also occur when septic wastewater is treated in a trickling filter. Volatile materials will be stripped from the wastewater when it is applied to the trickling filter, and airflow through the trickling filter will disperse the stripped compounds. In such cases, pretreatment of the wastewater to remove odorous materials may be necessary.[13,25]

19.4.2 Coupled Processes

The operation of coupled TF/AS processes incorporates features of the operation of both components. As discussed above, preceding a suspended growth bioreactor by a trickling filter typically results in good control over the growth of filamentous microorganisms in the suspended growth bioreactor. This is generally a benefit and results in a process that is stable and reliable. However, it can result in a turbid effluent when insufficient filaments grow. The TF/SC process is particularly susceptible to this because most substrate is removed in the trickling filter, leaving little to be removed in the suspended growth bioreactor where filamentous microbial growth occurs. As a consequence, the mixed liquor from a TF/SC process is fragile and must be handled gently to avoid floc shear and elevated effluent suspended solids concentrations. Growth of insufficient filaments is less of a problem for processes such as RF/AS and BF/AS because only a portion of the organic matter is removed in the trickling filter, leaving some to be metabolized in the suspended growth bioreactor. Consequently, a moderate population of filamentous microorganisms can be maintained in such processes, strengthening the floc.

A well-designed coupled TF/AS process will provide the flexibility for operation in several modes. This is desirable because wastewater treatment facilities are not typically loaded to their design values on a consistent basis. During periods of low loading, the TOL on the trickling filter will be less than the design value, giving a higher degree of treatment there. Since less substrate removal will be required in the suspended growth bioreactor, the flexibility should be provided to reduce its SRT to lower aeration power requirements. This may also be necessary to avoid unwanted nitrification. Plant operations personnel must effectively use the operational flexibility provided in the design to optimize plant performance and minimize treatment costs. For example, during periods of reduced loading, many BF/AS facilities are operated either with less suspended growth bioreactor volume in service or in the ABF mode. Likewise, facilities designed as RF/AS processes can be operated as TF/SC processes during periods of reduced loading. Finally, a TF/SC system can be operated simply as a trickling filter during periods of reduced loading.

Periodic sloughing of biomass occurs in many trickling filter installations. In coupled TF/AS systems this periodic sloughing can lead to a sudden increase in the mass of MLSS in the suspended growth bioreactor.[13,25,46] Because of the good sludge settling properties of the suspended biomass, such sudden increases can often be tolerated with little or no adverse impact on effluent quality. However, if the process is loaded to its maximum treatment capacity, such sloughing events can lead to significant loss of suspended solids from the final clarifier, with the associated deterioration in effluent quality. This can be avoided by maintaining sufficient hydraulic loading on the trickling filter to periodically slough excess biomass, demonstrating the importance of this practice. In systems where controlled sloughing is not practiced, the mass of MLSS in the suspended growth bioreactor must be maintained at reduced levels in anticipation of periodic sloughing events.

19.4.3 Nuisance Organisms

An extremely diverse biota can develop in a trickling filter. It contains a variety of organisms, ranging from bacteria to protozoa, worms, adult and larval filter flies (often of the genus Psychoda), and snails. This diverse biota contributes to stabilization of the organic matter applied to the trickling filter by increasing the length of the food chain. For example, protozoa feed on bacteria, worms feed on protozoa, and so on, with a yield associated with each trophic level, thereby reducing net biomass production. This is particularly important in rock media trickling filters where an extremely diverse biota can develop. However, in some instances some of these organisms can cause operational problems.

Filter flies can be a nuisance, particularly if a large hatch results in sufficient numbers to impact plant personnel and neighbors. Snails cause problems because their shells act like grit, accumulating in downstream unit operations, such as the suspended growth bioreactor of a coupled TF/AS process. They can also accumulate in the solids handling system, such as in anaerobic digesters, resulting in significant volume losses. Filter fly larvae, worms, and snails can cause performance problems in nitrifying trickling filters if they consume biomass faster than it is being produced, as has been observed in a number of trickling filters. For these reasons, it is important to control nuisance organisms.

In spite of the need, few techniques have been developed to control nuisance organisms in trickling filters. In some cases, chlorination (for all media), flooding (especially rock media trickling filters), and hydraulic flushing have been successful. Treatment of nitrifying trickling filters with ammonia at a sufficient concentration and pH to produce a nonionized ammonia concentration of 1 mg/L as N has proven effective and represents an emerging opportunity.[33] Media selection also affects the growth of nuisance organisms to some extent. Media with poor wetting characteristics are prone to the growth of nuisance organisms because they utilize the wet and dry areas of the media for different portions of their life cycle.[13,14,25,65,66]

19.5 KEY POINTS

1. A trickling filter is an attached growth, aerobic biochemical operation consisting of five major components: (1) the media bed, (2) the containment structure, (3) the wastewater application (or dosing) system, (4) the underdrain, and (5) the ventilation system.
2. Oxygen is provided either by natural draft or forced draft ventilation. Natural draft ventilation occurs because of differences in the density of air inside and outside the trickling filter. Forced draft ventilation requires a motive force for air, such as a fan.
3. Trickling filter systems often include a clarifier to separate produced biomass, although it may not be needed in separate stage nitrification applications because of the low growth yield.
4. The liquid flow pattern through a trickling can be characterized as plug flow with dispersion. Consequently, a spatial distribution of microorganisms will develop with heterotrophic bacteria in the upper portion of the trickling filter and, if the organic loading is sufficiently low, autotrophic nitrifying bacteria in the lower portion.
5. Trickling filters are classified by their treatment objective: partial removal of organic matter (referred to as a roughing trickling filter), relatively complete removal of organic matter (carbon oxidation), combined carbon oxidation and nitrification, and separate stage nitrification. These applications are defined by the total organic loading (TOL) applied to the trickling filter and the characteristics of the wastewater.
6. Due to significant differences in performance and operational characteristics, trickling filter processes are also classified by media type.
7. Adequate pretreatment must be provided before wastewater is applied to a trickling filter. Most installations include primary clarifiers for this purpose.

8. Coupled trickling filter/activated sludge (TF/AS) systems consist of a trickling filter, a suspended growth bioreactor, and a clarifier. The trickling filter effluent passes directly into the suspended growth bioreactor without clarification. As a consequence, the biologies of the two biochemical operations interact directly.
9. Trickling filters can be used to treat a wide range of wastewaters. They are often used to treat high strength, readily biodegradable wastewaters where they provide preliminary or roughing treatment prior to a suspended growth system. They can also be used for municipal wastewaters, although in some cases they must be coupled with a suspended growth bioreactor to produce the high quality effluent required. Combined carbon oxidation and nitrification and separate stage nitrification can also be readily accomplished.
10. Trickling filter process loadings can be expressed as either the total organic loading (TOL) or the surface organic loading (SOL). The TOL is expressed per unit of media volume (e.g., kg COD or $BOD_5/(m^3 \cdot day)$), while the SOL is expressed per unit of media surface area (e.g., kg COD or $BOD_5/(m^2 \cdot day)$). Similar loadings can be defined for ammonia-N and total Kjeldahl nitrogen (TKN) for nitrification applications.
11. The total hydraulic loading (THL) is the applied flow rate per unit of cross-sectional area, and typically has units of m/hr. Application of a minimum THL is required to achieve effective use of all of the media. In some cases this requires recirculation of treated effluent. However, the recirculation rate is not a fundamental design parameter. Increased recirculation flow beyond the amount required to achieve the minimum THL will not improve performance.
12. For a constant TOL, trickling filter performance is affected by media depth only over a relatively narrow range. Performance improves as the depth is increased up to about 3 to 4 meters, but little improvement occurs as the depth is increased further.
13. Trickling filter performance is affected by the temperature of the wastewater flowing over the media. Because trickling filters are effective heat transfer devices, steps must be taken during winter operation to mitigate the cooling effects of the ambient air temperatures.
14. Trickling filters can use either rotary or fixed nozzle distributors. Rotary distributors possess significant performance and operational advantages, and they are used more frequently.
15. Natural draft ventilation systems are designed to provide sufficient vent and underdrain areas to allow adequate convective air movement for oxygen transfer. Forced draft ventilation systems utilize fans and duct systems to evenly distribute the needed air.
16. Trickling filter effluents may contain fine, colloidal suspended solids that settle poorly in conventional clarifiers. Coupled TF/AS systems provide greater control over these suspended solids by flocculating them for removal in the clarifier.
17. Process loading factors, such as the TOL and SOL, can be used to size trickling filters. Process performance data from either comparable full-scale applications or a pilot plant is used to select the appropriate loading factor.
18. The Velz/Germain[22,62] equation is an empirical model that has been used frequently to size trickling filters. Because of its frequent use, a significant database exists to allow selection of appropriate model coefficients.
19. The model of Logan et al.[34,35] is a more fundamental trickling filter model. Experience with it is currently limited, and this must be considered when using it to size full-scale applications.
20. A trade-off is inherent in the design of a coupled TF/AS system. Use of a large trickling filter allows use of a small suspended growth bioreactor, and vice versa. Consequently, the trickling filter TOL and suspended growth bioreactor solids retention time must be selected together.
21. Oxygen requirements in coupled TF/AS systems can be determined by either fundamental process calculations or empirical correlations. In either case, the trickling filter reduces the oxygen requirement in the suspended growth bioreactor.

22. Improved trickling filter performance can be obtained by proper control of the biofilm thickness through periodic flushing. Flushing can be accomplished in numerous ways.
23. Significant operating economies can be achieved when a trickling filter system is designed with the flexibility to be operated in more than one mode.
24. Nuisance organisms that can proliferate in trickling filters include flies, snails, and worms. Problems caused by the growth of these organisms range from simple nuisance to reduced performance. Techniques available to control their growth include control of the THL, flushing, chlorination, treatment with elevated nonionized ammonia, and flooding.

19.6 STUDY QUESTIONS

1. Describe the functions of the five major components of a trickling filter.
2. Describe the mechanisms responsible for natural draft ventilation in a trickling filter and define the conditions under which stagnation is likely to occur. How does the configuration of the air distribution system in a trickling filter with natural draft ventilation differ from that in a trickling filter with forced draft ventilation? Why?
3. Describe the types of microorganisms present in the upper, middle, and lower portions of a trickling filter accomplishing combined carbon oxidation and nitrification and explain why the distribution develops?
4. Prepare a table that describes the various trickling filter process options, concentrating on differences associated with treatment objectives and media type. The table should define the essential features of each option.
5. Prepare a table that describes the various coupled TF/AS process options. What are the principal differences between the processes? Provide a quantitative definition of each process option.
6. Develop a table listing the benefits, drawbacks, and typical applications of the various trickling filter process options.
7. A wastewater with a flow rate of 3500 m^3/day and a BOD_5 concentration of 120 mg/L is to be treated using a trickling filter sized at a TOL of 0.6 kg $BOD_5/(m^3 \cdot day)$. Develop a curve showing the recirculation ratio required to maintain a minimum THL of 1.8 m/hr as a function of trickling filter media depth. Define the practical range of media depths. Select a design media depth for this application and present the rationale for your selection.
8. Discuss the techniques that can be used to minimize the heat loss from a trickling filter during cold weather operation.
9. Prepare a table contrasting the characteristics of the various trickling filter media. What are the benefits and drawbacks associated with each media type and when would each typically be used?
10. Prepare a table contrasting the relative benefits and drawbacks of rotary and fixed nozzle distributors.
11. Consider the trickling filter application discussed in Study Question 7. Using the recommended SK values presented in Table 19.5 calculate the necessary rotary distributor rotational speed for typical operation and for flushing. Assume that a four arm distributor is used.
12. Use the process loading factor approach to size a plastic sheet media trickling filter to accomplish tertiary nitrification. The secondary effluent to be nitrified has a low BOD_5 concentration, an ammonia-N concentration of 25 mg-N/L, and a flow rate of 15,000 m^3/day. The desired effluent ammonia-N concentration is 2 mg-N/L. Document and justify any assumptions necessary to complete these calculations.
13. Using the coefficients in Example 19.3.3.1 and Equation 19.9, develop a curve demonstrating the impact of the TOL on the effluent soluble BOD_5 concentration from a trickling

filter treating a wastewater with a soluble BOD_5 concentration of 125 mg/L and a flow rate of 4000 m^3/day. The subject trickling filter has a media depth of 6.7 m. Maintain a THL of 1.8 m/hr.

14. Consider a wastewater with a flow of 15,000 m^3/day and a BOD_5 concentration of 175 mg/L. Size both the trickling filter and the suspended growth bioreactor for each of the four coupled TF/AS processes. Define the rationale for selection of the specific process loadings. For the BF/AS and RF/AS processes, use the empirical relationships of Harrison[23] to calculate the suspended growth bioreactor oxygen requirement.
15. Make a list of the critical considerations in the operation a trickling filter. What additional considerations are involved in the operation of a coupled TF/AS process?

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20 Rotating Biological Contactor

The term rotating biological contactor (RBC) refers to a class of aerobic attached growth bioreactors containing circular shaped corrugated plastic media that are mounted on a horizontal shaft, partially submerged (typically 40%) in the wastewater, and rotated at a speed of one to two revolutions per minute to alternately expose them to the wastewater and to the atmosphere. Figure 20.1 provides a schematic diagram. A number of manufacturers have produced RBC equipment, but they are all similar and produce similar results.

Microorganisms grow on the media and metabolize biodegradable organic matter and nitrogen-containing compounds in the wastewater. Just as with a trickling filter, the produced biomass will slough off of the media and be transported by the wastewater to a clarifier where it is separated from the treated effluent. When the loading rate of biodegradable organic matter is sufficiently low, nitrifying bacteria will grow on the media and convert ammonia-N to nitrate-N. Wastewater that has been partially treated in another biochemical operation can be applied to an RBC system, where it will be nitrified.

20.1 PROCESS DESCRIPTION

Many RBC facilities were installed in the United States in the 1970s, with over 600 installations at the peak use of this technology. It is estimated that approximately 3000 installations have existed worldwide. Many of these facilities have not performed as expected, and the process has fallen into disfavor with many plant owners and operators. However, experience with existing installations has established the appropriate range of applications for the RBC process, as well as the range of operating conditions resulting in acceptable performance. Consequently, practitioners can upgrade and expand existing RBC facilities with confidence and properly evaluate RBCs as an alternative for new applications. A number of publications detail the development of RBC technology and document its appropriate operating range.[3,25,27] In addition, design manuals have summarized the current state of the art.[26,28,31] In spite of increased certainty in its application, few new RBC installations have been implemented in recent years as greater focus has been placed on nutrient removal applications, for which RBCs are not well suited.

20.1.1 General Description

A description of the RBC process is provided in Chapter 17 and immediately above. As indicated in Figure 20.1, a cover (typically fiberglass) is provided over each individual RBC unit for protection against inclement weather, freezing, and sunlight, which accelerates media deterioration. Covers also reduce heat loss, allow the off-gas to be collected for odor control, and minimize algae growth. Alternatively, entire installations can be placed in buildings, but this can result in a humid, corrosive atmosphere leading to accelerated corrosion.

Nearly all manufacturers produce individual RBC units to standard dimensions. A typical media bundle is 3.66 m in diameter and 7.62 m long, on a shaft that is 8.23 m long. Consequently, at a typical rotational velocity of 1.6 rpm the peripheral velocity of the disc is 18.3 m/min. The media is manufactured from high density polyethylene containing UV inhibitors. The individual sheets are corrugated in much the same fashion as plastic sheet trickling filter media. Corrugations increase the stiffness of each disc, increase the available surface area, improve mass transfer, and serve to define the spacing between individual disks. Just as with plastic sheet trickling filter media, the

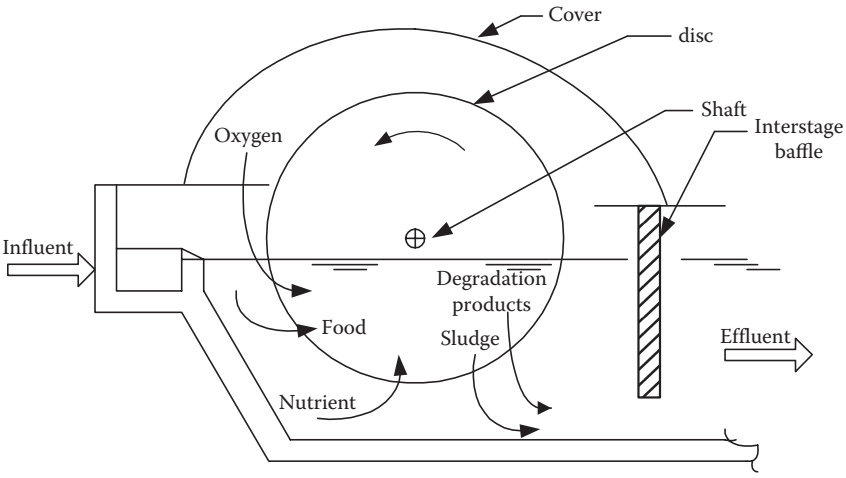

FIGURE 20.1 Schematic diagram of an RBC.

size of the corrugations defines how closely together the individual sheets of media can be placed, thereby determining the media density. Standard density media has a specific surface area of about 115 m^2/m^3, so each standard shaft provides 9300 m^2 of media surface area. High density media has a specific surface area of about 175 m^2/m^3, providing 13,900 m^2 of media surface area per shaft. Minor density differences occur from one manufacturer to another, so it is possible to purchase media with slightly larger or smaller surface areas per shaft. Like trickling filter installations, the media density used for a particular application is determined by the characteristics of the wastewater being treated and by the treatment objectives.

Individual RBC units (called shafts) generally are arranged in series to maximize capacity and treatment efficiency. Baffles are used to separate the RBC shafts into a series of completely mixed bioreactors, each referred to as a stage. The effects of staging on system performance are discussed in Sections 17.2.2 and 20.2.3. A single stage may contain more than one shaft, but because each stage is completely mixed, all shafts in a stage behave in the same manner. A volume of 45 m^3 per shaft is typically used to size the bioreactor. As illustrated in Figure 20.2a, a typical system for removal of biodegradable organic matter (i.e., secondary treatment) might use three stages in series, with the first stage containing two shafts. Such a series of stages is referred to as a treatment train. A mixture of standard and high density RBC shafts can be used in a single train, although the initial shaft will generally contain standard density media. A larger number of RBC shafts in series will typically be used for advanced treatment applications (i.e., both carbon oxidation and nitrification), as illustrated in Figure 20.2b.

The axis of each individual RBC shaft is typically placed perpendicular to the direction of flow through the train. As indicated in Figure 20.1, the RBC shafts generally rotate in the direction that causes the top of the media to move opposite to the direction of flow. This minimizes short-circuiting.

The baffles used to define the individual stages in a treatment train are typically not load bearing and thus are not capable of isolating an individual RBC shaft. Rather, they are often moveable to allow the number of stages and their sizes to be adjusted in response to long-term variations in process loadings. A typical interstage baffle is illustrated in Figure 20.1.

Staging can also be accomplished using a single RBC shaft, as illustrated in Figure 20.2c. The single shaft is placed in a bioreactor with a volume of 45 m^3, with the shaft parallel to the long dimension of the bioreactor. Flow is parallel to the shaft, and interstage baffles are placed at various points along the shaft to provide the necessary staging. This arrangement is used in small wastewater treatment plants where only a small number of RBC units is needed.

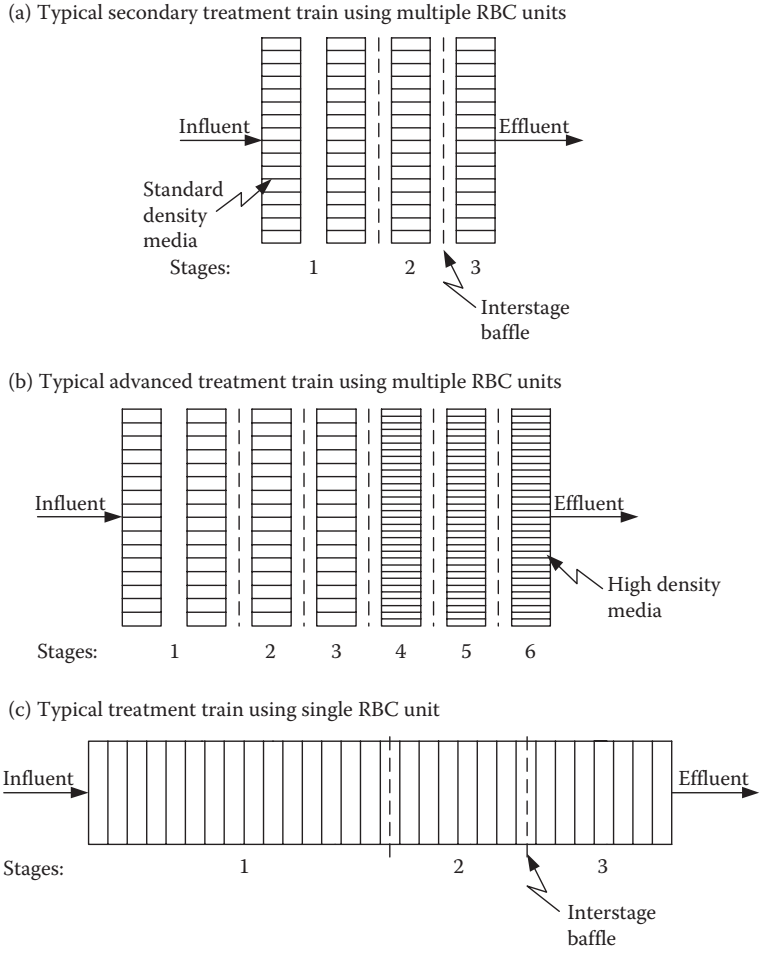

FIGURE 20.2 Examples of RBC trains.

Individual RBC trains are arranged in parallel with flow split equally to each train, as illustrated in Figure 20.3. Larger wastewater treatment plants will use several trains of parallel shafts. In smaller facilities the "end flow" configuration illustrated in Figure 20.2c is used for each train. For systems removing organic matter, the effluent from the RBC trains will typically be combined and conveyed to secondary clarifiers for the removal of sloughed biomass. Clarification of the RBC effluent may not be necessary for tertiary nitrification applications.

20.1.2 Process Options

20.1.2.1 Treatment Objectives

Rotating biological contactors are used to remove biodegradable organic matter and convert ammonia-N and organic-N to nitrate-N. As discussed in Section 20.2.1, operational problems caused by high unit organic loading rates restrict the use of RBCs for partial removal of organic matter (i.e., for "roughing" treatment). However, they can be used quite effectively for substantial removal of organic matter. Process effluent (i.e., clarified) five-day biochemical oxygen demand (BOD_5) and total suspended solids (TSS) concentrations can easily be reduced to less than 30 mg/L each, and even lower concentrations can be obtained in some instances. This degree of treatment can be accomplished by applying proper organic and hydraulic loading rates, as discussed below.

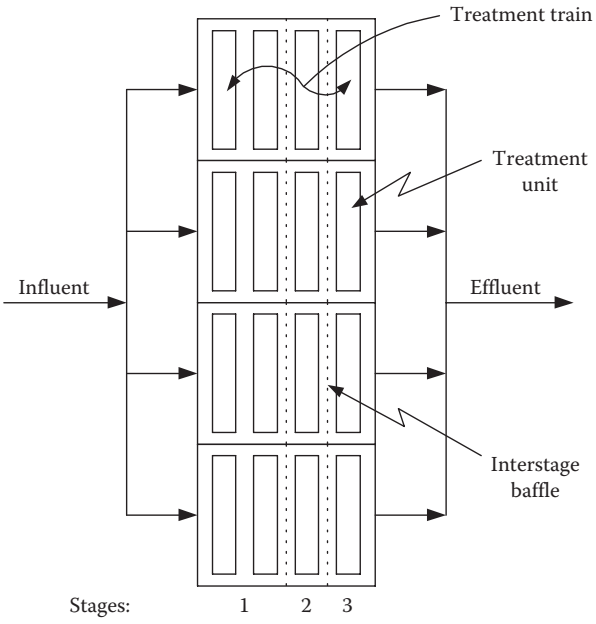

FIGURE 20.3 Typical configuration of an RBC treatment facility.

Combined carbon oxidation and nitrification can also be accomplished in an RBC system. As discussed in Section 16.4, heterotrophic and autotrophic bacteria compete for space within the aerobic portion of a biofilm, causing heterotrophs to predominate when both organic substrate and ammonia-N concentrations are high. Consequently, the oxidation of organic matter will generally occur in the initial stages of the RBC train, just as it occurs in the top of a trickling filter. However, if the organic loading on the train is sufficiently low, the organic substrate concentration will be reduced sufficiently so that autotrophs will be able to compete in the latter stages. As with other aerobic fixed film processes, this occurs when the soluble substrate concentration is reduced to about 20 mg/L as chemical oxygen demand (COD) (15 mg/L as BOD_5).[25,26,31] Thus, the primary distinction between a secondary treatment application (the removal of organic substrate alone) and a combined carbon oxidation and nitrification application (the removal of organic substrate and the oxidation of ammonia-N to nitrate-N) is the organic loading. A larger number of stages may be used for combined carbon oxidation and nitrification to increase the degree of staging and separate the heterotrophic and autotrophic bacteria.

Rotating biological contactors can also be used for separate stage nitrification; that is, to nitrify streams containing relatively high concentrations of ammonia-N and low concentrations of organic matter. Such applications may not require downstream clarification because of the low biomass production rates associated with nitrification. Separate stage nitrification applications are distinguished from combined carbon oxidation and nitrification applications by the characteristics of the wastewater being treated. If the concentration of organic substrate in the influent wastewater is low and the concentration of ammonia-N is substantial, the biofilm that develops will be enriched in nitrifiers and the impact of the organic matter on process sizing will be negligible. Benchmarks for distinguishing a separate stage nitrification application are an influent BOD_5/TKN (total Kjeldehl nitrogen) ratio less than about 1.0 and/or an influent soluble BOD_5 concentration less than about 15 mg/L.

Rotating biological contactors have also been used to accomplish denitrification. In these applications the RBC unit is entirely submerged and covers are provided to exclude air. The influent is generally a nitrified secondary effluent, so an electron donor must be provided. These applications are quite limited and will not be discussed further.

20.1.2.2 Equipment Type

A motive force is necessary to rotate the RBC shaft. Two general approaches are used: mechanical drives and air drives. Mechanical drive systems consist of an electric motor, a speed reducer, and a belt or chain drive for each shaft. The electric motors are typically 3.7 or 5.6 kW and the RBC rotational speed is typically 1.2 to 1.6 rpm. The capability to adjust speed and rotational direction can be provided by using speed reducers with multiple pulley or sprocket ratios or through a variable speed drive. These features can be used to control the buildup of excess biomass.

Air drive units increase the oxygen transfer capacity of an individual RBC unit and reduce the number of electric motors required. Cups are added to the periphery of the media and oriented to collect air injected under the RBC shaft. The cups are either 10 or 15 cm long, depending on the organic loading to the unit. The airflow per shaft ranges from 4.2 to 11.3 m^3/min under standard conditions, which is typically sufficient to provide rotational speeds of 1.0 to 1.4 rpm. Air is generally provided to all shafts by a centralized blower system. The quantity of air required varies depending on the specific configuration and operating conditions.

Mechanically driven systems provide reliable, consistent rotation of the RBC shaft and media. However, they are susceptible to biomass buildup when they are organically overloaded or subjected to high sulfide loading, as discussed in Section 20.2.6. Air drive systems provide enhanced oxygen transfer, and the injected air can assist with the removal of excess biomass. Both of these effects can be beneficial in a heavily loaded unit. The primary disadvantage of air drive systems is that they are more susceptible to loping, which is uneven rotation caused by the development of nonuniform biomass growth around the circumference of the RBC media. Uneven rotational speed results as the heavier portion of the disc is lifted out of the liquid, rotated to the top, and allowed to descend by gravity back into the liquid.

A more recent innovation is the submerged biological contactor, within which 70 to 90% of the media is submerged. They are generally aerated. Claimed advantages include reduced structural loadings on the shaft and bearings, improved biomass control, the ability to use larger media bundles, and increased treatment capacity. To date these systems have received limited use.

20.1.3 Comparison of Process Options

Table 20.1 summarizes the primary benefits and drawbacks of the RBC process. It is mechanically simple, which simplifies normal equipment maintenance. It is also an uncomplicated process, thereby lessening the need for intensive daily process control actions. The energy requirements are relatively low, being only those required to rotate the media. Finally, it is modular in nature, which simplifies design and construction.

Its principal drawbacks are that process performance is sensitive to wastewater characteristics and loadings, and that it possesses little operational flexibility to adjust to varying loading and operating conditions. As discussed in Section 20.2.1, high organic loadings can result in excessive biomass growth, which structurally overloads the media and shaft. This problem is exacerbated by

TABLE 20.1
RBC Process Benefits and Drawbacks

Benefits	Drawbacks
Mechanically simple	Performance susceptible to wastewater characteristics
Simple process, easy to operate	Limited process flexibility
Low energy requirements	Limited ability to scale-up
Modular configuration allows easy construction and expansion	Adequate pretreatment required

elevated levels of hydrogen sulfide in the influent wastewater. Although significant deterioration in treatment capacity and performance result from excessive biomass, the steps that can be taken to minimize its impact are relatively limited. Fortunately, the conditions leading to cataclysmic declines in performance are now relatively well defined and can generally be avoided if the operating conditions for the facility are well defined.

An early claimed benefit of the RBC process was its ability to respond successfully to shock loads.[2] However, subsequent experience has demonstrated that its capability to respond to shock loads is much like that of the trickling filter process. Shock loads will not cause complete process failure, but the process does not generally possess sufficient reserve capacity to successfully treat them.[9,11] Consequently, equalization should be provided upstream of an RBC process if the ratio of peak to average loading exceeds 2.5.[26,31]

The size of an individual RBC shaft limits the maximum plant size for which the RBC process is practical. As discussed above, each shaft can contain a media surface area of 9300 or 13,900 m^2. Consequently, only for small to medium wastewater flow rates can sufficient media be provided by a reasonable number of RBC units. For example, a total media surface area of about 280,000 m^2 might be required to treat a typical municipal wastewater with a design flow of 15,000 m^3/day. This could be provided by 24 RBC shafts configured in six trains consisting of four shafts each. An equivalent media surface area would be provided by two trickling filters, each 16.4 m in diameter and 6.7 m deep. To treat a flow rate 10 times as large, the number of RBC units required would increase 10-fold; in this case to 60 trains, each with four individual RBC shafts. In contrast, the equivalent trickling filter installation would still require only two trickling filters, although each would have to be 51.9 m in diameter and 6.7 m deep. Alternatively, four trickling filters, each 36.7 m in diameter and 6.7 m deep, could be used. From a cost, construction, and operational perspective, the smaller number of trickling filters would be more desirable. This factor tends to limit the use of RBCs to smaller wastewater treatment plants.

A final drawback of RBCs is the need for adequate preliminary treatment. Debris such as rags, plastics, and fibrous material can clog the RBC media if present in sufficient quantities, and grit will settle in the RBC bioreactor due to the relatively low level of turbulence provided. In general, the minimum degree of preliminary treatment required is fine screening (less than 1 mm opening) and excellent grit removal. Primary clarification is provided in many instances. The cost of the necessary preliminary treatment facilities must be included in any cost evaluation of the RBC process.

The biomass produced in the RBC process generally settles and thickens readily. Consequently, the waste solids stream leaving the final clarifier can either be recycled to the primary clarifier to be settled and thickened with the primary solids, or it can be thickened separately. Other solids thickening options can also be applied successfully.

20.1.4 Typical Applications

Rotating biological contactors have typically been used to provide secondary treatment to municipal wastewater. They have also been used to nitrify municipal wastewaters, either in combined carbon oxidation and nitrification applications or in separate stage nitrification applications. Approximately 70% of the applications in the United States have been for removal of organic matter, 25% for combined carbon oxidation and nitrification, and 5% for separate stage nitrification.[13,32] Due to the poor economy of scale for this technology, it has been used most frequently for wastewater treatment plants with flows below about 40,000 m^3/day. It has also been used successfully in a number of industrial applications, particularly those involving wastewaters of moderate to low strength and with low concentrations of hydrogen sulfide.

The reliability of RBCs improved considerably in the 1980s and 1990s. Nevertheless, while many RBC installations have provided acceptable performance, many others have not met performance expectations. For example, a survey indicated that over 80% of the RBCs designed before 1980 have experienced operational problems.[32] Many of these problems have been solved by

improved construction techniques and the use of appropriate organic and hydraulic loading rates. Consequently, RBC technology is now sufficiently well defined so that it is possible to clearly evaluate existing facilities for upgrades and new applications.

20.2 FACTORS AFFECTING PERFORMANCE

Many factors affect the performance of RBC wastewater treatment systems. This section emphasizes those factors that are particularly significant and/or are relatively unique to the RBC process.

20.2.1 Organic Loading

The performance of an RBC facility is significantly affected by the organic loading.[10,18,20,21,33] The organic loading on an RBC is typically expressed on the basis of the total media surface area, A_s, not just the wetted or submerged areas. Consequently, it is equivalent to the surface organic loading (SOL) as used in trickling filter design. The definition of the SOL is given by Equation 19.2:

$$\lambda_S = \frac{F(S_{SO} + X_{SO})}{A_s}. \tag{19.2}$$

All terms in the equation are as previously defined. Typical units for SOL are g COD/(m^2·day) or g BOD_5/(m^2·day). Figure 20.4 illustrates the typical relationship between the SOL and the removal rate for readily biodegradable organic matter for a variety of full-scale RBC installations. The removal rate generally increases as the SOL increases but at a decreasing rate. As a result, the substrate removal efficiency decreases as the SOL increases. The SOL values can be calculated for an individual RBC stage or for the entire RBC treatment system. In the former case, the organic matter concentration is that entering the particular stage and A_s is the area of media in the stage. In the latter case, the organic matter concentration is that in the process influent and A_s is the media surface area for the entire system.

The organic loading cannot be increased indefinitely, as might be suggested by Figure 20.4. Rather, it is limited by the maximum oxygen transfer capacity of an individual RBC shaft. Analysis of data from full-scale RBC facilities indicates that oxygen transfer limitations occur at SOL values to individual RBC shafts of about 32 g BOD_5/(m^2 · day). This value is commonly taken to correspond to a soluble BOD_5 SOL of 12 to 20 g/(m^2 · day), which corresponds to a soluble biodegradable COD SOL of 20 to 35 g COD/(m^2 · day).[19,25–27,31] Because of oxygen limitations, excessive growth

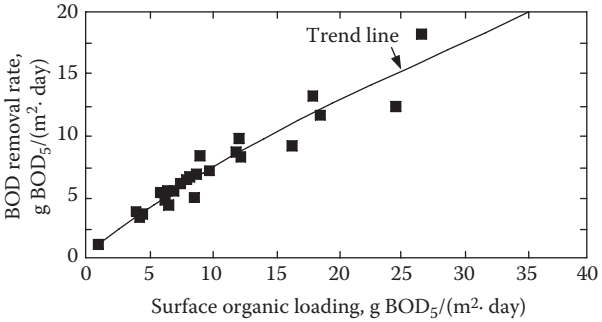

FIGURE 20.4 Effect of surface organic loading (SOL) on the surficial BOD_5 removal rate (flux) for full-scale RBC facilities treating domestic wastewater. (From Water Environment Federation, *Wastewater Treatment Plant Design*, 4th Edition Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1998. Copyright © Water Environment Federation. Reprinted with permission.)

of the nuisance organism *Beggiatoa* occurs at SOLs in excess of these values. The mechanisms for *Beggiatoa* growth and its impact on the performance of RBC systems are discussed in Section 20.2.6. Standard density RBC media (9300 m^2/shaft) should be used in RBC stages that are highly loaded or where *Beggiatoa* growth is possible. Excess biological growths are more difficult to remove from high density media and, consequently, use of such media can further exacerbate operational problems in highly loaded stages.

The organic loading affects nitrification in an RBC system. As discussed above, nitrifying bacteria can effectively compete for space in a biofilm once the concentration of soluble organic substrate is reduced below 20 mg COD/L (15 mg BOD_5/L). The SOL, among other factors, determines whether that occurs. An empirical relationship has been developed to show the effect of SOL (expressed as g total biodegradable COD/(m^2·day)) on the nitrification rate that can develop in an RBC unit performing combined carbon oxidation and nitrification:[17]

$$f_{NH} = 1.43 - 0.1\lambda_S; \quad 4.3 < \lambda_S < 14.3, \tag{20.1}$$

where f_{NH} is the fraction of the rate that would occur in the absence of simultaneous carbon oxidation. The limits on f_{NH} are one and zero. This equation indicates that no nitrification will occur when the SOL ≥ 14.3 g total biodegradable COD/(m^2·day) and that unrestricted nitrification will occur at SOLs of 4.3 g total biodegradable COD/(m^2·day) or less. At all SOL values between those extremes, significant competition occurs between heterotrophs and autotrophs, resulting in reduced nitrification rates.

For separate stage nitrification systems the classical relationship between the growth limiting ammonia-N concentration and the flux into the biofilm, J_{NH}, as described by Equation 19.5, is observed. According to Equation 20.1 this will occur when the loading of biodegradable COD is below 4.3 g COD/(m^2·day). Figure 20.5 presents the relationship. As indicated there, at ammonia-N concentrations above about 5 mg/L nitrification proceeds in RBC units at a rate of about 1.5 g N/(m^2·day). This represents a zero-order biofilm in which the nitrification rate is not limited by the bulk ammonia-N concentration. Rather, it is limited by the rate of oxygen transfer to the biofilm. As the bulk ammonia-N concentration decreases below 5 mg/L, it begins to become rate limiting. Examination of Figure 20.5 suggests that the pseudo half-saturation coefficient for ammonia-N for RBC biofilms, $K_{g,NH}$, is approximately 2 mg N/L.

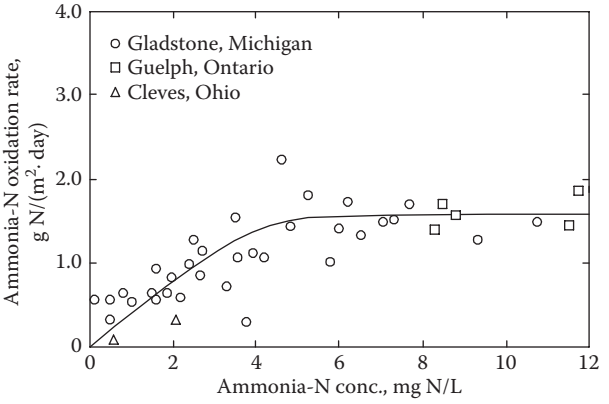

FIGURE 20.5 Effect of the ammonia-N concentration in an RBC stage on the ammonia-N oxidation rate in the absence of significant carbon oxidation. (Adapted from U.S. Environmental Protection Agency, *Design Information on Rotating Biological Contactors*, EPA/600/2-84/106, U.S. Environmental Protection Agency, Cincinnati, OH, 1984.)

During combined carbon oxidation and nitrification, the rate of nitrification must be obtained by multiplying the rate associated with the ammonia-N concentration, as given by Figure 20.5 or Equation 19.5, by the value of f_{NH} obtained with Equation 20.1. The use of these relationships is illustrated in the example that follows.

Example 20.2.1.1

One stage of an RBC system treating domestic wastewater contains a single shaft of high density media ($A_s = 13{,}900$ m^2) and receives 2000 m^3/day of flow containing 50 mg/L of biodegradable COD and 10 mg/L of ammonia-N. Determine the concentration of ammonia-N in the effluent from the stage. Assume that the curve in Figure 20.5 represents the observed effect of the ammonia-N concentration on the nitrification rate.

a. Will nitrification occur in the stage?
   Two factors determine whether nitrification will occur in an RBC, the soluble COD in the stage and the SOL applied to it. The SOL must be less than 14.3 g COD/(m^2·day). The SOL can be calculated with Equation 19.2:

$$\lambda_S = \frac{(2000)(50)}{13{,}900} = 7.19 \text{ g COD/(m}^2 \cdot \text{day)}.$$

   Since the SOL is less than 14.3 g COD/(m^2·day), nitrification will occur, provided the soluble COD in the stage is below 20 mg/L. Examination of Figure 20.4 shows the relationship between the SOL and the organic matter removal rate per unit area in an RBC. In that figure, the SOL is expressed in terms of BOD_5. Equation 9.31 shows that the biodegradable COD in domestic wastewater is 1.71 times the BOD_5. Thus an SOL of 14.3 g COD/(m^2·day) is equivalent to an SOL of 8.36 g BOD_5/(m^2·day). From Figure 20.4 it can be seen that an RBC receiving such a loading will have a organic removal rate around 6 g BOD_5/(m^2·day), for a removal efficiency of approximately 73%. This suggests that the total biodegradable COD leaving the stage will be less than 15 mg/L, making the soluble COD well below that value. Thus, we can expect nitrification to occur in the stage.

b. At what fraction of the unrestricted nitrification rate will nitrification occur in the RBC?
   The fractional nitrification rate can be estimated with Equation 20.1 using the value of the SOL calculated in part a:

$$f_{NH} = 1.43 - (0.1)(7.19) = 0.71.$$

   Thus, nitrification will occur in the RBC at 71% of the rate at which it would occur in an RBC performing only nitrification.

c. What is the concentration of ammonia-N leaving the RBC stage?
   The effect of the ammonia-N concentration on the nitrification rate is shown in Figure 20.5. However, since the ammonia-N concentration in the RBC stage depends on the removal rate in the stage, and no analytical expression is available relating the rate and the concentration, an iterative procedure must be used. Start by assuming that the concentration is above 5 mg/L so that the rate is independent of the concentration. In that case, the rate given by the figure is 1.5 g NH_3-N/(m^2·day). The rate in the RBC will be only 71% of that value, however, because of the carbon oxidation occurring. Thus, the rate is 1.07 g NH_3-N/(m^2·day). The effluent concentration resulting from that rate must be calculated from a mass balance on the stage and compared against the assumed value. If they do not agree, a new value must be assumed and the process repeated. Performing the mass balance:

$$\text{Input rate} = (2000)(10) = 20{,}000 \text{ g/day}$$

$$\text{Removal rate} = (1.07)(13{,}900) = 14{,}800 \text{ g/day}$$

Output rate = 20,000 − 14,800 = 5200 g/day

Effluent concentration = 5200 ÷ 2000 = 2.6 mg N/L

This concentration is lower than the assumed value. Thus, the removal rate will be lower than was assumed, making the concentration higher than 2.6 mg/L. For the second iteration, assume a concentration between the two previous values. Use the average of the two, which is 3.8 mg/L. Entering Figure 20.5 with that concentration, the unrestricted rate is 1.24 g NH_3-N/(m^2·day). The rate in the presence of carbon oxidation is 71% of that value or 0.88 g NH_3-N/(m^2·day). Repeating the mass balance results in an effluent ammonia-N concentration of 3.9 mg/L. This is sufficiently close to the assumed value, so the effluent ammonia-N concentration can be considered to be around 3.8 to 3.9 mg/L.

20.2.2 Hydraulic Loading

The total hydraulic loading (THL) for an RBC system, $\Lambda_{H,RBC}$, is defined slightly differently than for a trickling filter. It is defined with respect to the media surface area and is calculated as follows:

$$\Lambda_{H,RBC} = \frac{F}{A_s}. \tag{20.2}$$

Comparison of Equations 19.2 and 20.2 indicates that the SOL and the THL are related to one another by the pollutant concentration in the influent wastewater.

The performance of RBC systems has historically been correlated with the THL. Manufacturers of RBC equipment have developed and published performance curves for domestic wastewater in which effluent quality is plotted as a function of the THL. The general shape of the effluent quality versus a THL relationship is independent of the influent wastewater strength, but the quantitative relationship varies with it. Figure 20.6 presents a typical relationship; this particular one was developed by the Autotrol Corporation (Crystal Lake, Illinois), which popularized the RBC process in the United States. Experience has indicated that the relationships developed by some of the RBC manufacturers accurately predict the performance trends observed by full-scale RBC facilities.[28,31] Thus, this general approach is a useful one for characterizing full-scale RBC facility performance. However, this same experience also indicates that the performance relationships published by many of the RBC equipment manufacturers provide optimistic estimates of facility performance.[28,31] Consequently, care should be exercised in the selection and application of such empirical relationships.

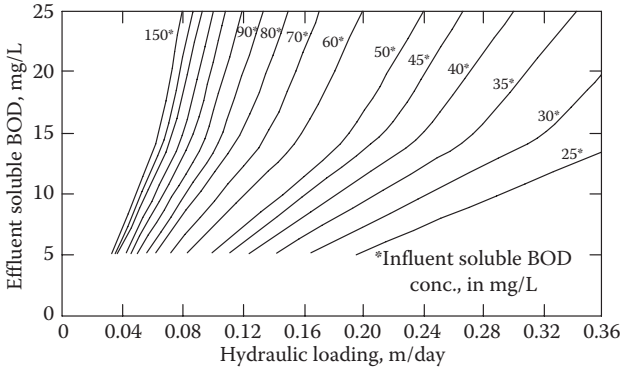

FIGURE 20.6 Typical design correlation for soluble BOD_5 removal in municipal wastewater treatment.

20.2.3 Staging

As discussed above and illustrated in Section 17.2.2, RBC facilities are typically staged to improve overall performance. In a staged system, the SOL for the initial stage is higher than the value for the entire system. As indicated in Figure 20.4, increased SOL values result in an increased removal rate. Consequently, by staging the bioreactor, the overall average removal rate can be increased, even though the SOL on the last stage is relatively low. Since the process effluent quality is determined by the SOL on the last stage, a good quality effluent can be produced even with a relatively high overall average SOL. The principal constraint is that the SOL in any stage must not exceed the oxygen transfer capacity of a shaft, 32 g $BOD_5/(m^2 \cdot day)$, as discussed in Section 20.2.1. The performance benefits of staging RBCs were demonstrated early in the development of this technology.[4]

Staging is particularly effective in systems that are required to achieve combined carbon oxidation and nitrification. Staging in the upstream portion of the train produces a high SOL, resulting in a high rate of organic matter removal. This results in reduced SOLs in subsequent stages and concentrations of soluble biodegradable organic matter less than 20 mg/L as COD. This, in turn, allows nitrifying bacteria to become established in the later stages of the RBC train. As indicated in Figure 20.5, nitrification is first-order for low ammonia-N concentrations, but it is zero-order for ammonia-N concentrations greater than 5 mg/L. Staging does not increase the ammonia-N flux into the biofilm when the ammonia-N concentration exceeds 5 mg/L, but it does increase the overall flux when the ammonia-N concentration is less than 5 mg/L. Table 20.2 summarizes recommendations from several sources concerning the number of stages for particular applications.[26,31,32]

Staging also affects the nature of the biomass that develops on individual RBC shafts and their resulting pollutant removal capability. This is illustrated most graphically in an RBC system that is used for combined carbon oxidation and nitrification. As indicated in the preceding paragraph, in the initial stages the biofilm will consist primarily of heterotrophs, making it very active and capable of high organic matter fluxes. The later stages, however, will contain biofilms with increasing proportions of nitrifying bacteria. As a result, the organic removal capability of these later stages will be reduced, although the nitrification capability will be enhanced.

20.2.4 Temperature

The effect of temperature on the performance of RBC systems is similar to that of other attached growth processes. Since the reaction rate is influenced strongly by diffusion, wastewater temperature has little effect on process performance over a wide range of temperatures. The effect of temperature is generally neglected for temperatures over about 15°C, although removal rates generally decline for lower temperatures.[3,26,31] Temperature relationships have been presented in two ways. Figure 20.7 provides one relationship. To use it, first determine the media surface area required for the particular application at a temperature above 55°F (13°C). Then enter Figure 20.7 and select the multiplier for the actual operating temperature. Finally, multiply the area by the multiplier to

TABLE 20.2
Summary of Staging Recommendations

Carbon Oxidation		Nitrification	
Effluent Soluble BOD_5 Conc., mg/L	Number of Stages	Effluent NH_3-N Conc., mg/L	Number of Stages
>25	1	>5	1
15–25	1 or 2	<5	Based on kinetics
10–15	2 or 3		
<10	3 or 4		

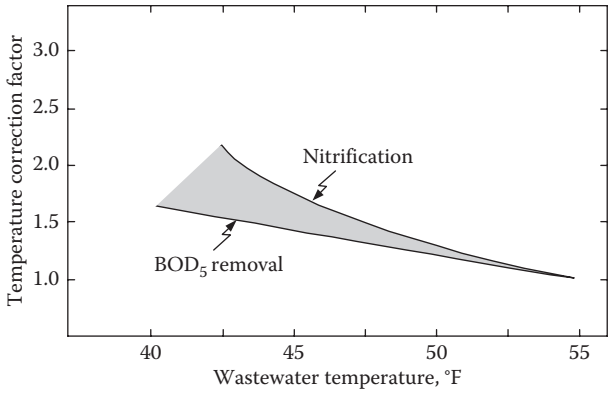

FIGURE 20.7 Effect of temperature on the area requirements for RBCs. (Adapted from U.S. Environmental Protection Agency, *Nitrogen Control*, EPA/625/R-93/010, U.S. Environmental Protection Agency, Washington, DC, 1993.)

determine the appropriate area at the actual operating temperature. Table 20.3 provides three other relationships from the literature.[28] To use Table 20.3, determine the expected biofilm flux, neglecting the impacts of temperature. Then enter the table to determine the correction coefficient for the expected temperature. Multiply the expected flux by that coefficient to obtain the design value, which can then be used to determine the required media area.

20.2.5 Wastewater Characteristics

The characteristics of the wastewater being treated will affect the performance of an RBC system in several ways, just as it will any other biochemical operation. For example, the flux into the biofilm may be smaller for large, slowly biodegradable compounds than for small, readily biodegradable

TABLE 20.3
Summary of Manufacturers' Temperature Correction Recommendations

Temp	Manufacturer		
°C	Envirex	LYCO	Walker Process
Temperature Correction Factor: BOD_5 Removal			
17.8	1.00	1.00	1.00
12.8	1.00	1.00	1.00
10.0	0.87	0.83	0.87
7.2	0.76	0.71	0.73
5.5	0.67	0.66	0.65
4.4	0.65	0.62	—
Temperature Correction Factor: Nitrification			
17.8	1.00	1.00	1.39
12.8	1.00	1.00	1.00
10.0	0.78	0.78	0.78
7.2	0.48	0.57	0.56
5.5	0.42	—	0.43
4.4	—	—	—

compounds. The presence of particulate organic matter may reduce the flux of soluble substrate since the particulate matter occupies space within the biofilm, reducing the biomass concentration within the biofilm, which decreases the rate of biodegradation. Hydrolysis of entrapped particulate organic matter releases soluble organic matter, which reduces the diffusion of soluble organic matter from the bulk liquid into the biofilm.

Another wastewater constituent that has a particularly important impact on the performance of RBC systems is hydrogen sulfide. As discussed in the next section, growth of the sulfide oxidizing bacterium *Beggiatoa* can cause operating problems in an RBC system.

20.2.6 Biofilm Characteristics

The character of the biofilm that develops on an RBC can significantly affect its performance. This, in turn, is affected by the operating conditions imposed on each RBC shaft. In the initial stages of an RBC train, where the removal of organic matter occurs, a normal biofilm is grey-brown in color and of a stable and controllable thickness.[3] It is composed primarily of aerobic heterotrophic bacteria, entrapped particulate organic matter, and Eucarya. If the overall process organic loading is sufficiently low, a nitrifying biofilm will develop in the later stages. A nitrifying biofilm is generally thinner than a heterotrophic biofilm and reddish-brown in color.[22] In separate stage nitrification systems, the biofilm will be highly enriched in nitrifiers and may be tan or bronze in color.

In some instances a tenacious whitish biofilm develops that will not slough off of the RBC media.[8,25,26,31] Since it does not slough off, it continues to buildup, which can lead to structural overload and physical failure of the RBC unit. Failure occurs either in the media or the shaft. Media can fail directly or it can detach from the rotating shaft. The result, in any case, is an inoperative unit. The tenacious biofilm develops from growth of the filamentous bacterium *Beggiatoa*. The filamentous nature of this microorganism reinforces the RBC biofilm, thereby increasing its resistance to hydraulic shear and decreasing sloughing rates. The whitish color of the biofilm is a result of the color of the *Beggiatoa*.

Beggiatoa obtains energy by oxidizing hydrogen sulfide and other compounds containing reduced sulfur. When sufficient quantities of hydrogen sulfide are present in the system, either from the influent wastewater or produced internally, *Beggiatoa* can compete effectively with heterotrophic bacteria for space within the aerobic biofilm. Hydrogen sulfide will be produced within the biofilm when the organic loading on individual RBC shafts exceeds 32 g BOD_5/(m^2·day), which corresponds to the maximum oxygen transfer capacity, as described previously. Anaerobic conditions in the interior of the biofilm allow sulfate reducing bacteria to use sulfate as a terminal electron acceptor, resulting in hydrogen sulfide production. The hydrogen sulfide diffuses outward into the overlying aerobic layer where it is used as an energy source by *Beggiatoa*.

The nature of the wastewater collection system can aggravate *Beggiatoa* problems. For example, if the collection system consists of a series of force mains, septic conditions can result in both hydrogen sulfide production and solubilization of biodegradable organic matter. The resulting increased concentration of readily biodegradable organic matter can elevate the SOL in the initial stage of the RBC train, resulting in anaerobiosis and even more hydrogen sulfide production. The location of such a system in a warm climate will make the problem worse.

Several approaches are available to deal with excessive *Beggiatoa* growth. One is to eliminate the conditions that cause it. The influent hydrogen sulfide concentration can be reduced by chemical oxidation with chlorine, hydrogen peroxide, or dissolved oxygen; by precipitation with iron; or by alteration of the operation of the collection system. The reader is referred to a Manual of Practice[29] for more information. Another is to reduce the SOL on an overloaded stage, either by reducing the entire process SOL or by redistributing the influent load. Methods for reducing the overall SOL include placing more RBC trains in service and increasing the removal of organic matter in upstream unit operations, such as by chemically enhancing the performance of a primary clarifier. Methods for redistributing the influent loading include changing the process staging, using effluent

recirculation, or step feeding load to downstream stages. Finally, several approaches are available for physically or chemically removing excess biomass from individual RBC shafts. Examples include oxidizing the biomass with chlorine, elevating the pH with caustic, scouring with air, changing the shear characteristics by reversing the rotational direction, recirculating effluent, and turning off and drying out individual shafts.[23,25,26,31]

Several approaches are available for monitoring the buildup of excessive biomass on individual RBC units. Visual inspection can be effective because of the distinctive appearance of *Beggiatoa*. Individual RBC shafts can also be equipped with load cells to periodically weigh them to detect increases in biofilm mass. This latter approach is particularly useful when excessive *Beggiatoa* growth occurs on a persistent basis. Regular measurements can be made, allowing periodic operational adjustments. Several references provide further details.[23,25,26,31]

20.3 PROCESS DESIGN

The complex nature of the events occurring in RBC systems was discussed in Section 17.2. Because of that complexity, fundamental design models for the RBC process have not been developed. Consequently, empirical approaches are used for design. Most of those approaches express the SOL in terms of BOD_5 so we will do the same herein. This section describes the design procedures for RBCs that remove organic matter and nitrify, both separately and together. The use of pilot plants to develop site-specific design data is also discussed. Some general comments on RBC design procedures are also provided. Table 20.4 summarizes the general approach to designing an RBC. Application of this general approach is illustrated in the following sections.

20.3.1 Removal of Biodegradable Organic Matter

20.3.1.1 General Approach

Several empirical design approaches have been presented in design manuals.[26,31] This section presents two equations that have been found to most accurately characterize the performance of full-scale systems treating domestic wastewaters, the first-order and the second-order models. Both can be used to estimate the total media area required for domestic wastewaters. The design of RBC systems to treat industrial wastewaters generally requires full-scale experience with the same or a similar wastewater or a pilot study.

TABLE 20.4
Summary of RBC Process Design Procedure

1. Summarize process design and loading conditions including maximum, minimum, and average sustained temperature; maximum, minimum, average, and design process influent flows and pollutant loadings; and desired effluent quality.
2. Where necessary, convert process loadings, expressed in conventional parameters (such as BOD_5) into the units used in the process design (such as biodegradable COD).
3. Select a design expression appropriate for the particular application.
4. Select an effluent quality goal. As discussed in Section 10.4.4, selection of that goal should consider uncertainty and variability in process performance.
5. Use the design expression and the effluent quality goal to calculate the total media surface area required.
6. Determine the media surface areas required in the first stage to keep the stage surface organic loading (SOL) below 32 g $BOD_5/(m^2 \cdot day)$ to prevent excessive growth of *Beggiatoa*.
7. Select the number of trains to be used, the number of stages in each train, and the number of shafts in each stage.
8. Summarize the results in tabular form.

After the total media surface area and the area in the first stage have been determined, engineering experience and judgment must be used to configure the system. Generally, a minimum of four trains is desirable from an operational perspective because when one train is out of service for maintenance, three-quarters of the total media volume will remain in service. This will generally be sufficient to produce an acceptable quality effluent in the short term. Guidance concerning the number of stages is provided in Table 20.2. Even though the total media volume may have been selected to give the desired effluent substrate concentration, staging that media as recommended in the table will provide a factor of safety in the design. Finally, the number of shafts in the first stage is determined by the minimum area required to prevent oxygen limitations. The number of shafts in the remaining stages can be selected based on the experience of the designer and other considerations. However, at no time should the loading on any stage exceed 32 g $BOD_5/(m^2 \cdot day)$.

20.3.1.2 First-Order Model

The first-order model is analogous to the Velz/Germain equation used to design trickling filters. It was first presented by Benjes[5,24] as follows:

$$\frac{S_{Se}}{S_{SO}} = \exp\left[-k_1\left(\frac{V_M}{F}\right)^{0.5}\right], \tag{20.3}$$

where S_{Se} and S_{SO} are the concentrations of total biodegradable organic matter in the clarified process effluent and influent, respectively, V_M is the media volume, F is the influent flow rate, and k_1 is a first-order reaction rate coefficient. It should be noted that even though the first-order model is similar to the Velz/Germain equation, it is based on concentration of total organic matter entering and leaving the process rather than on the soluble organic matter as was done in Equation 19.6. Based on a review of operating data from 27 full-scale municipal wastewater treatment plants, a value for k_1 of 0.3 was selected when S_{Se} and S_{SO} are measured as BOD_5, V_M is expressed in ft^3, and F is expressed in gallons per minute.[5,24] Figure 20.8[31] compares the predictions of this equation with the results from the plants and indicates a generally good fit. Note that the plot was prepared with $100[(S_{SO} - S_{Se})/S_{SO}]$ as the ordinate and the SOL as the abscissa. Because of the nature of the abscissa, a separate curve results from Equation 20.3 for each influent substrate concentration and the plant data have been grouped into three sets to show that effect. Nevertheless, the data scatter indicates that some facilities may perform less efficiently than indicated by the equation. As a

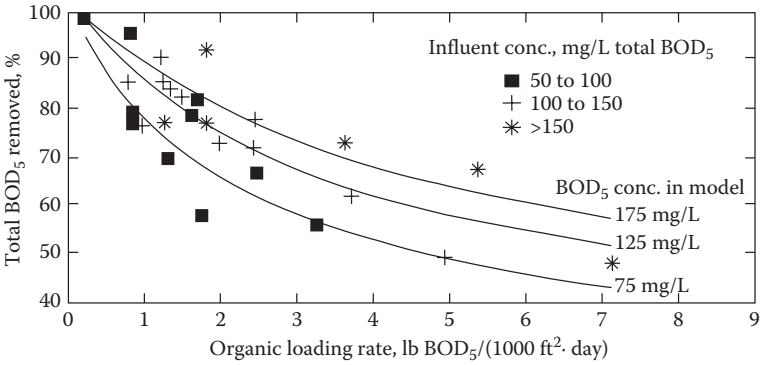

FIGURE 20.8 Comparison of predictions from first-order model (Equation 20.3, $k_1 = 0.3$) with data from 27 full-scale RBC installations. The curves are the model predictions and the points represent the data. (From Water Environment Federation, *Wastewater Treatment Plant Design*, 4th Edition Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1998. Copyright © Water Environment Federation. Reprinted with permission.)

consequence, some design manuals suggest the use of a more conservative k_1 value in the range of 0.2 to 0.25.[28,31]

Because Equation 20.3 was developed using standard density RBC shafts that contain media with a specific surface area of 35 ft^2/ft^3 (115 m^2/m^3), it can be converted into an expression based on media surface area. Using metric units and a value for k_1 of 0.3, the expression becomes:

$$\frac{S_{Se}}{S_{SO}} = \exp\left[-0.4\left(\frac{A_s}{F}\right)^{0.5}\right], \tag{20.4}$$

where A_s is the total media surface area expressed in m^2 and F is the wastewater flow rate in m^3/day. Even though Equation 20.4 was developed for standard density media, the important outcome from its use is the total media surface area required. Consequently, once the required surface area has been determined, it can be achieved with a combination of standard and high density media, as long as standard density media is used in the most heavily loaded stages. Use of the first-order model, as expressed in Equation 20.4, to design an RBC for the removal of biodegradable organic matter is illustrated in the following example.

Example 20.3.1.1

Use the first-order equation of Benjes[5,24] to design an RBC to treat a wastewater with an average flow of 8000 m^3/day and a BOD_5 concentration (after primary clarification) of 120 mg/L. The wastewater is domestic in origin. Secondary treatment is required, so the BOD_5 and TSS concentrations should not exceed 30 mg/L on a monthly average basis. A robust system capable of reliably meeting this requirement is desired.

a. What design effluent quality should be used?
   Using the procedure of Roper et al., as discussed in Section 10.4.4, a design BOD_5 concentration of 20 mg/L is appropriate to allow a facility to reliably comply with a monthly average effluent BOD_5 limit of 30 mg/L.
b. What media surface area is required to produce the selected effluent quality?
   The required media surface area can be calculated with a rearranged form of Equation 20.4:

$$A_s = F\left[\frac{-\ln\left(S_{Se}/S_{SO}\right)}{0.4}\right]^2,$$

$$A_s = 8000\left[\frac{-\ln\left(20/120\right)}{0.4}\right]^2 = 160{,}500\ m^2.$$

c. What is the minimum media area required in the first stage to avoid oxygen transfer limitations?
   The limiting SOL is 32 g $BOD_5/(m^2 \cdot day)$. Using this value, the minimum media area required in the first stage can be calculated with Equation 19.2:

$$A_{s,1} = \frac{(8000)(120)}{32} = 30{,}000\ m^2.$$

   Standard density media must be used in the first stage because of the amount of biomass growth that will occur there. Since that media has a surface area of 9300 m^2/shaft, 3.22 shafts would be required to obtain the needed minimum area. However, since whole shafts must be used, at least four shafts are needed, thereby providing 37,200 m^2.

d. How many trains should the facility have?
   A minimum of four trains is desirable from an operational perspective. Since at least four shafts are required in the first stage, it is logical to have four trains, with at least one shaft in the first stage of each train.
e. How many stages should be used in each train?
   From Table 20.2, two or three stages are needed to produce an effluent soluble BOD_5 of 10 to 15 mg/L (required to meet a total BOD_5 of 30 mg/L) and three or four stages are needed to produce an effluent soluble BOD_5 less than 10 mg/L (required to meet a total BOD_5 of 20 mg/L). Since the required effluent BOD_5 concentration is 30 mg/L and the design concentration is 20 mg/L, the use of three stages is reasonable.
f. How should the trains be configured?
   From part b above, the total media area for the system must be at least 160,500 m^2. Since four trains will be used, each train should have a total media area of at least 40,125 m^2. The simplest system would contain three shafts per train, one in each stage. As mentioned above, for carbon oxidation, standard density media (9300 m^2/shaft) must be used in the first stage. However, high density media (13,900 m^2/shaft) can be used in later stages if sufficient organic matter has been removed to prevent the growth of thick biofilms. With a total media area of 40,125 m^2 per train, if three stages were used with one shaft per stage, the average media area per shaft would have to be 13,375 m^2. There is no combination of standard and high density media that can provide this average area with three shafts. Only all high density media can do so, which is unacceptable. Consequently, more than three shafts must be used in each train.

Four shafts per train requires an average area per shaft of about 10,000 m^2. This requirement cannot be met with standard density media alone, but can easily be met with a combination of standard and high density media. Trying different combinations reveals that three standard density shafts and one high density shaft per train provides a total area of 41,800 m^2 per train, which is adequate. The total media area in the system would be 167,200 m^2. Two standard density media and two high density media shafts will provide 46,400 m^2 per train, or 185,600 m^2 for the system, which is approximately 15% greater than the needed area. There is an additional cost associated with the use of high density media, so the choice between the two options is an economic one in which the additional cost must be justified on the basis of improved performance or reliability.

The desired three stages might typically be achieved by using two standard density shafts as the first stage, with one other shaft in each subsequent stage. The use of two shafts in the first stage ensures that the SOL on that stage is well below the limit based on oxygen transfer. If only one high density shaft is used in the system, it should be in the last stage. Regardless of which choice is made, each train should be configured with moveable baffles so that staging can be adjusted if necessary.

20.3.1.3 Second-Order Model

The second-order model, first proposed by Opatken,[15,16] assumes that the flux of soluble, biodegradable organic matter into a biofilm is second-order with respect to its bulk liquid phase concentration. The second-order flux equation was incorporated into the mass balance equation for soluble substrate in a single completely mixed bioreactor containing a biofilm. That equation was then solved and generalized to estimate the concentration of soluble, biodegradable organic matter in stage N of a multistage train, $S_{S,n}$:

$$S_{S,n} = \frac{-1 + \sqrt{1 + 4k_2 \cdot \tau_n \cdot S_{S,n-1}}}{2k_2 \cdot \tau_n}, \tag{20.5}$$

where $S_{S,n-1}$ is the concentration of soluble substrate entering stage N from the preceding stage, k_2 is the second-order reaction rate coefficient, and τ_n is the hydraulic residence time of stage N. When $S_{S,n-1}$ and $S_{S,n}$ are expressed in units of g/m^3 (or mg/L) of soluble BOD_5 and τ_n is in units of hours, k_2 has been found to have a value of 0.083 m^3/(g·hr) for domestic wastewater when the tank volume

to media surface area ratio is 4.89×10^{-3} m^3/m^2. This ratio is equivalent to 45 m^3/shaft for standard density media.

As with the first-order model, Equation 20.5 can be converted into one based on media surface area and wastewater flow rate. Using appropriate units conversions, and a value for k_2 of 0.083 $m^3/(g \cdot hr)$, the term $k_2 \cdot \tau_n$ becomes 0.00974 $A_{s,n}/F$, where $A_{s,n}$ is the media area (in m^2) in stage N and F is the flow to that stage, expressed in m^3/day. The units for this term are m^3/g of BOD_5. The complete expression is

$$S_{S,n} = \frac{-1 + \sqrt{1 + (4)(0.00974)(A_{s,n}/F)(S_{S,n-1})}}{(2)(0.00974)(A_{s,n}/F)}. \tag{20.6}$$

The variables $S_{S,n}$ and $S_{S,n-1}$ must be expressed as soluble BOD_5 when using this expression.

The second-order model is used repetitively to estimate the effluent quality from one stage to another in an RBC train. Consequently, unlike the first-order model, it directly considers the benefits to be gained by staging. Thus, it is not uncommon for the second-order model to result in smaller systems than the first-order model for the same design situation. Because it minimizes the amount of media required, the media area in the first stage is typically set to give an SOL on that stage of 12 to 20 g soluble $BOD_5/(m^2 \cdot day)$—equivalent to 32 g $BOD_5/(m^2 \cdot day)$ based on total BOD_5—the maximum that can be applied without oxygen transfer limitations. If no data are available about the soluble BOD_5 of the influent wastewater, it is often assumed that one-half of the total BOD_5 in the untreated wastewater is soluble. The influent soluble BOD_5 concentration is then used in conjunction with Equation 20.6 to estimate the effluent soluble BOD_5 concentration from the first stage. Various configurations for the remaining stages are then evaluated to select the one that most economically meets the effluent quality goal, bearing in mind that the SOL on any stage must not exceed 12 to 20 g $BOD_5/(m^2 \cdot day)$ based on the soluble BOD_5. When the last stage is reached, since both the influent and the desired effluent concentrations are known, the required media area can be calculated directly. Rearrangement of Equation 20.6 gives an equation for $A_{s,N}$, where N represents the last stage and $S_{S,N-1}$ is the soluble BOD_5 concentration entering the last stage:

$$A_{s,N} = (102.8)(F)\left(\frac{S_{S,N-1} - S_{Se}}{S_{Se}^2}\right). \tag{20.7}$$

If the calculated area is less than the area provided by a single shaft, the area of a single shaft must be used. This will produce a better effluent quality, which can be calculated with Equation 20.6. Of course, the loading on the stage must be sufficiently low to prevent an oxygen limitation, but this will generally not be a problem for low effluent concentrations.

Equation 20.6 estimates the effluent soluble BOD_5 concentration. The total BOD_5 concentration in the process effluent must be estimated by adding the BOD_5 associated with the effluent suspended solids to the soluble BOD_5 concentration calculated with Equation 20.6 for the last stage. One would expect the BOD_5 of the suspended solids to depend on the loading on the RBC system, just as the BOD_5 of suspended solids from trickling filters depends on the loading, as illustrated in Figure 19.9. Unfortunately, comparable data are not available for RBCs. Consequently, many designers assume that the soluble BOD_5 of a clarified RBC effluent is about one-half of the total BOD_5.

Example 20.3.1.2

Use the second-order model to size the RBC system considered in Example 20.3.1.1. Assume that four trains will be provided, just as in that example. Also assume that the primary clarifier removes 30% of the BOD_5 applied to it.

a. What must the design effluent soluble BOD_5 be?
In part a of Example 20.3.1.1, it was determined that the design total BOD_5 must be 20 mg/L to meet the effluent criteria reliably. If we assume that the effluent soluble BOD_5 will be one-half of the total BOD_5, then the design effluent soluble BOD_5 must be 10 mg/L.

b. What is the soluble BOD_5 of the wastewater to be treated?
As discussed above, when no other information is available, designers typically assume that 50% of the BOD_5 in unsettled domestic wastewater is soluble. Because the primary clarifier removes 30% of the BOD_5 and the clarified wastewater has a total BOD_5 of 120 mg/L, the total BOD_5 of the untreated wastewater is

$$\frac{120}{(1-0.3)} = 171 \text{ mg/L}.$$

Assuming that 50% of that BOD_5 is soluble, the soluble BOD_5 is

$$S_{SO} = (171)(0.5) = 86 \text{ mg/L}.$$

c. What is the soluble BOD_5 concentration in the effluent from the first stage if it is made as small as possible without risk of oxygen limitation?
In part c of Example 20.3.1.1, the minimum size of the first stage to avoid oxygen limitations was determined to be four standard density media shafts, with one in each of four trains. Use Equation 20.6 to determine the effluent soluble BOD_5 from this stage. The most straightforward way to approach the problem is to work with one train, since they will all be equal. Since the total flow to the system is 8000 m^3/day, the flow per train is 2000 m^3/day. Furthermore, since the first stage of each train contains one standard density shaft, the media area per stage is 9300 m^2. Therefore, application of Equation 20.6 gives:

$$S_{S,1} = \frac{-1+\sqrt{1+(4)(0.00974)(9300/2000)(86)}}{(2)(0.00974)(9300/2000)} = 33.9 \text{ mg/L}.$$

d. What is the soluble BOD_5 concentration in the effluent from the second stage if it is also made as small as possible without risk of oxygen limitation?
To avoid oxygen limitations, no stage should have an SOL that exceeds 12 to 20 g soluble BOD_5/(m^2·day). Using a conservative value of 12 g soluble BOD_5/(m^2·day) for the second stage, and recognizing that the flow rate per train is 2000 m^3/day, the minimum media surface area required in the second stage of each train can be calculated with Equation 19.2:

$$A_{s,2} = \frac{(2000)(33.9)}{12} = 5600 \text{ m}^2.$$

This is less than the media surface area provided by one standard density shaft, so the minimum acceptable media area is that associated with one standard density shaft, or 9300 m^2. Using this area, the effluent soluble BOD_5 concentration from stage two can be calculated using Equation 20.6:

$$S_{S,2} = \frac{-1+\sqrt{1+(4)(0.00974)(9300/2000)(33.9)}}{(2)(0.00974)(9300/2000)} = 18.5 \text{ mg/L}.$$

e. What size must the third stage be to produce an effluent with a soluble BOD_5 concentration of 10 mg/L?
Since the influent and the desired effluent BOD_5 concentrations are known, the required media area can be calculated with Equation 20.7:

$$A_{s,3} = (102.8)(2000)\left(\frac{18.5-10}{10^2}\right) = 17{,}500 \text{ m}^2.$$

This is larger than the area provided by either a standard or a high density shaft. Thus, the desired effluent quality cannot be met with a single shaft stage. While two standard media shafts could be used in the third stage to provide the needed area, this would not be an economic design. It would be better to add more media area to the second stage by changing it from the standard density to high density media, thereby reducing the BOD_5 concentration entering the third stage. This is possible because of the low BOD_5 concentration entering the second stage and the low SOL. Using Equation 20.6 for a second stage with an area of 13,900 m^2 reveals that its effluent substrate concentration will be 16.2 mg/L. Using Equation 20.7 with this value as the influent soluble BOD_5 into stage three reveals that the required media area is 12,750 m^2, which can be met with one high density shaft.

f. Summarize the design.
The design consists of four trains, each with three stages consisting of one shaft each. The first stage uses a standard density shaft, while the following two stages use high density shafts. The total media surface area for each train is therefore 37,100 m^2, giving 148,400 m^2 for the entire system.

Comparison of Examples 20.3.1.1 and 20.3.1.2 indicates that application of the first-order model results in a more conservative design than the use of the second-order model. This is because the second-order model calculates the media area needed in each stage, thereby directly considering the benefits of staging. The first-order model, on the other hand, calculates the total media area needed in the entire system, and then divides it into stages with the empirical recommendations of Table 20.2. Nevertheless, both approaches resulted in a four train system containing three stages per train. The total media area calculated with the first-order model was 13% greater, however. Considering the range of assumptions required to complete the two examples, a 13% difference is not large and the similarity in the two resulting designs is encouraging. These two equations and a variety of other process design approaches, including the relationships used by various manufacturers of RBC equipment, have been compared in design manuals.[28,31] Those comparisons indicate that the first and second-order models provide relatively conservative and consistent estimates of RBC process performance that agree reasonably well with the performance of full-scale plants treating domestic wastewater. Consequently, use of those models has been recommended for the design of municipal systems in the absence of relevant pilot- and/or full-scale data.[28,31]

20.3.2 Separate Stage Nitrification

The approach used to design RBC systems for separate stage nitrification is the same as that outlined in Section 20.3.1, except that Figure 20.5 provides the performance relationship generally used. In order to accomplish separate stage nitrification, the concentration of soluble, biodegradable COD in the wastewater must be less than 20 mg/L (soluble BOD_5 less than 15 mg/L). Because of the low concentration of biodegradable organic matter, the first stage SOL will generally be less than 32 g $BOD_5/(m^2 \cdot day)$ and thus will not need to be checked. The low first stage SOL will allow high density media to be used throughout the process train.

Inspection of Figure 20.5 indicates that the flux of ammonia-N into the biofilm will be sustained at a value of 1.5 g NH_3-N/($m^2 \cdot$day) until the ammonia-N concentration is lowered to about 5 mg/L. Thus, ammonia-N removal behaves in a zero-order manner at concentrations above that value, allowing the RBC media area required to lower the ammonia-N concentration from the influent concentration to 5 mg/L to be calculated directly using that flux. Then, a trial and error procedure based on the curve in Figure 20.5 must be used to determine the additional media area required to lower the ammonia-N concentration from 5 mg/L to the desired effluent concentration. Alternatively, the graphical procedure mentioned in Section 20.2.2 and presented in Section 20.3.4 could also be used. The computational procedure is illustrated in the next example.

Some researchers have advocated that the flow direction through separate stage nitrification trains be reversed periodically (say weekly) to maintain complete and effective nitrifying biofilm

development along the entire treatment train.[6] Such an approach will affect facility layout and configuration, but it may result in better system performance due to the maintenance of a more active nitrifying population.

Example 20.3.2.1

A wastewater treatment plant treating a flow of 8000 m^3/day is achieving reliable secondary treatment (soluble BOD_5 concentration < 15 mg/L) but must be upgraded to provide nitrification. The ammonia-N concentration in the current effluent is 25 mg/L, and the effluent ammonia-N goal is 2.5 mg/L. Size a tertiary RBC system to accomplish this goal.

a. What surface area of media is required to lower the ammonia-N concentration from 25 mg/L to 5 mg/L?
From Figure 20.5, as long as the ammonia-N concentration is above 5 mg/L and the SOL is less than 4.3 g COD/(m^2·day), the flux of ammonia-N into the biofilm is 1.5 g NH_3-N/(m^2·day). The mass of ammonia-N to be removed is just the flow times the required concentration change of 20 mg/L, or 160,000 g N/day. From the definition of flux, the required RBC surface area is

$$A_s = \frac{160{,}000}{1.5} = 106{,}700 \text{ m}^2.$$

At 13,900 m^2/shaft for high density media, this would require 7.7 shafts. Since only whole shafts can be used, use eight for an area of 111,200 m^2.

Check the SOL to ensure that the assumption of a nitrification rate that is unrestricted by carbon oxidation is correct. As discussed in Section 20.2.1, nitrification will proceed at its maximum rate as long as the SOL is less than 4.3 g COD/(m^2·day). The SOL can be calculated with Equation 19.2, but only the soluble BOD_5 is specified in the problem statement. Thus, an estimate must be made of the COD concentration. As indicated earlier, the soluble BOD_5 is often assumed to be one-half of the total BOD_5. Furthermore, Equation 9.31 states that the biodegradable COD can be estimated as 1.71 times the BOD_5. Using these conversion factors, a soluble BOD_5 of 15 mg/L is equivalent to a COD of 51 mg/L. Consequently, if all of the media is placed into a single stage, the SOL is

$$\lambda_S = \frac{(8000)(51)}{111{,}200} = 3.67 \text{ g COD/(m}^2 \cdot \text{day)}.$$

Since this value is less than 4.3 g COD/(m^2·day), nitrification would proceed at the unrestricted rate.

b. What media area is required to lower the ammonia-N concentration to 2.5 mg/L in a single stage?
Since a single stage is to be used, the ammonia-N concentration in the bioreactor will be 2.5 mg/L. From Figure 20.5, the ammonia-N flux at a concentration of 2.5 mg/L is approximately 0.75 g NH_3-N/(m^2·day). Again, the mass of ammonia-N to be removed is just the flow times the required concentration change. Since the concentration change is 2.5 mg/L, the mass of ammonia-N to be removed is 20,000 g/day. From the definition of flux, the required RBC surface area is

$$A_s = \frac{20{,}000}{0.75} = 26{,}700 \text{ m}^2.$$

This area can be provided by two high density shafts.

c. How should the system be configured?
Since a total of 10 shafts are required and an integer number of shafts must be used in each train, configure the system as two trains containing five shafts each. Since no benefits are gained by staging as long as the ammonia-N removal rate is zero-order (i.e., as long as the ammonia-N concentration is 5 mg/L or above), the first four shafts in each train can be placed into a single stage. This will ensure that the SOL on that stage stays low enough so that carbon oxidation does not interfere with nitrification. The remaining fifth shaft per train should be placed in a separate stage to achieve the desired effluent ammonia-N concentration of 2.5 mg/L. Consequently, each train should have two stages with four shafts in the first stage and one in the second. Movable baffles should be used to achieve the staging. Furthermore, provisions could be made to allow the direction of flow to be reversed to maintain a more active biofilm in the last stage. If this were done, the baffle between the first and second stage would be moved to maintain four shafts in the first stage and one in the last.

20.3.3 Combined Carbon Oxidation and Nitrification

The design of combined carbon oxidation and nitrification systems incorporates the principles presented in the two previous sections. The major difference is that the effects of carbon oxidation on the nitrification rate must be considered through the use of Equation 20.1, as was done in Example 20.2.1.1. The procedure is illustrated in the following example.

Example 20.3.3.1

Reconsider the wastewater for which the carbon oxidation designs were performed in Examples 20.3.1.1 and 20.3.1.2, and for which the separate stage nitrification system was designed in Example 20.3.2.1. In this case, however, design a single RBC system to remove the biodegradable organic matter and to lower the effluent ammonia-N concentration to 2.5 mg/L. The concentration of ammonia-N available to be nitrified (after hydrolysis of organic nitrogen and uptake of nitrogen for cell synthesis) is 25 mg/L. Since four trains are used in Examples 20.3.1.1 and 20.3.1.2, configure the system as four trains in parallel.

a. How much media should the system contain?
Because of the competition for space between heterotrophs and autotrophs in the biofilm, nitrification will follow carbon oxidation to a considerable degree, even in systems in which both events are occurring. Consequently, we can obtain a conservative estimate of the media area required by summing the values determined in the preceding examples. From Example 20.3.1.1, a conservative design for the removal of biodegradable organic matter would require 160,500 m^2. From Example 20.3.2.1, a media surface area of 106,700 m^2 is required to reduce the ammonia-N concentration from 25 to 5 mg/L and 26,700 m^2 to reduce it from 5 to 2.5 mg/L in the absence of carbon oxidation. Use of the sum of these areas as an initial estimate of the required media area:

$$A_s = 160{,}500 + 106{,}700 + 26{,}700 = 293{,}900 \text{ m}^2.$$

b. How should the media be configured in each of the four trains?
The surface area required for each train is one-fourth of 293,900 m^2 or 73,500 m^2. Consider a layout consisting of two standard density and four high density shafts. The standard density shafts will be placed in a single stage to ensure that the first stage SOL will be well below the value that will cause *Beggiatoa* growth. The remainder of the shafts will be high density to minimize the total number of shafts. Using this configuration provides a total surface area per train of

$$A_s = (2)(9300) + (4)(13{,}900) = 74{,}200 \text{ m}^2.$$

Consider a five stage system using the two standard density shafts as the first stage and each of the high density shafts as an individual stage. Staging of the high density shafts will maximize treatment capacity. As in the other designs, moveable baffles should be provided to maximize flexibility.

c. What is the BOD_5 removal profile through each train?
This information is needed to estimate where nitrification will occur. Use the second-order model, Equation 20.7, to estimate the profile by calculating sequentially from stage to stage. The media surface area, A_s, and the value of A_s/F in each stage, calculated with a flow rate of 2000 m^3/day for each train, are summarized in Table E20.1, along with the soluble BOD_5 concentration in each stage. The concentration of 4.6 mg/L in stage five indicates that the effluent total BOD_5 concentration will be low, provided that good removal of suspended solids occurs in the final clarifier.

d. What is the zero-order ammonia-N removal rate in each stage?
Two factors influence nitrification. One is the SOL, and its effects on the nitrification rate can be estimated by using Equation 20.1. The second is the ammonia-N concentration, and its effect can be quantified using Figure 20.5. Furthermore, nitrification will be initiated only when the soluble BOD_5 concentration drops below about 15 mg/L.

Consider the impact of the SOL through the use of Equation 20.1. The SOL in that equation is expressed in units of g COD/(m^2·day), but the substrate concentration leaving each stage in Table E20.1 is expressed as soluble BOD_5. Consequently, the COD concentration must be approximated from the BOD_5 information. This can be done as it was in part a of Example 20.3.2.1, making the COD approximately 3.4 times the soluble BOD_5. The concentrations of soluble BOD_5 and the equivalent total substrate COD entering each stage are tabulated in Table E20.2. The SOL on each stage can be calculated with Equation 19.2 using the appropriate media area in the stage and the flow rate of 2,000 m^3/day to each train. The SOL values are also tabulated in the table. Finally, once the SOL is known, the fractional nitrification rate, f_{NH}, can be calculated with Equation 20.1. It should be recalled that no nitrification will occur when the SOL exceeds 14.3 g COD/(m^2·day) and that nitrification will not be influenced by organic substrate removal when the SOL is less than 4.3 g COD/(m^2·day). Furthermore, little if any nitrification will occur when the soluble BOD_5 concentration in a stage exceeds 15 mg/L. It is clear that no nitrification will occur in stage one. For stage two, the SOL is 12.7 g COD/(m^2·day) and the soluble BOD_5 concentration in the stage is 13.5 mg/L. Thus, a small amount of nitrification may occur. However, it is not likely to be significant, so to be conservative f_{NH} was taken as zero. Examination of the SOL into stages four and five suggests that organic substrate removal will not restrict nitrification in them, making f_{NH} equal to 1.0. Only in stage three will significant competition occur between heterotrophs and autotrophs, giving an f_{NH} value of 0.77. The f_{NH} values are summarized in Table E20.2.

TABLE E20.1
Soluble BOD_5 Concentrations through the RBC Stages in Example 20.3.3.1

Stage	Surface Area* m^2	A_s/F day/m	Soluble BOD_5 mg/L
0	—	—	86.0
1	18,600	9.30	25.8
2	13,900	6.95	13.5
3	13,900	6.95	8.6
4	13,900	6.95	6.1
5	13,900	6.95	4.6

**Per train.*

TABLE E20.2
Ammonia-N Concentrations through the RBC Stages in Example 20.3.3.1

Stage	Influent Soluble BOD_5 mg/L	Influent COD mg/L	SOL g COD/ m^2.day	f_{NH}	NH_3-N mg/L
1	86.0	292	31.4	0.0	25.0
2	25.8	88	12.7	0.0	25.0
3	13.5	46	6.6	0.77	17.0
4	8.6	29	4.2	1.0	6.5
5	6.1	21	3.0	1.0	2.0

e. What is the ammonia-N profile through a train?
As long as the ammonia-N concentration is above 5 mg/L, calculation of the ammonia-N profile is straightforward because the nitrification rate is not influenced by the ammonia-N concentration. However, once the concentration drops below 5 mg/L, the effect of the concentration must be considered, which requires an iterative procedure, as illustrated in Example 20.2.1.1.

As indicated in Table E20.2, no nitrification will occur in stages one and two, and thus the ammonia-N concentration in them is the same as the influent concentration, 25 mg/L. It is likely that the ammonia-N concentration will be high enough in stage three to allow nitrification to occur at a rate that is unaffected by the concentration. From Figure 20.5 the maximum rate in the absence of carbon oxidation is found to be 1.5 g NH_3-N/(m^2·day). However, from Table E20.2 the value of f_{NH} is seen to be 0.77. Therefore, the ammonia-N removal rate is 1.16 g NH_3-N/(m^2·day). Using this, the effluent ammonia-N concentration can be calculated from a mass balance on the stage.

$$\text{Input rate} = (2000)(25) = 50{,}000 \text{ g/day}$$

$$\text{Removal rate} = (1.16)(13{,}900) = 16{,}100 \text{ g/day}$$

$$\text{Output rate} = 50{,}000 - 16{,}100 = 33{,}900 \text{ g/day}$$

$$\text{Effluent concentration} = 33{,}900 \div 2000 = 17 \text{ mg N/L}$$

The concentration is well above 5 mg/L, so the assumption of a zero-order rate is justified. Examination of the mass removal rate in stage three and comparison of it to the mass input rate into stage four suggests that the concentration in stage four will also be high enough to allow nitrification to occur at the zero-order rate. In this case, however, carbon oxidation has no effect, as indicated in Table E20.2. Thus, the rate will be 1.5 g NH_3-N/(m^2·day). Repeating the procedure above for stage four reveals that the effluent ammonia-N concentration will be 6.5 mg/L. Since this is above 5 mg/L, the assumption that nitrification will occur at the zero-order rate is justified.

Since the ammonia-N concentration entering stage five is close to 5 mg/L, the effluent concentration from stage five will be less than 5 mg/L, causing the ammonia-N flux to be limited by the ammonia-N concentration. This requires use of an iterative procedure to estimate the effluent ammonia-N concentration, just as was done in Example 20.2.1.1. Application of that procedure reveals that the effluent ammonia-N concentration from stage five is 2.0 mg/L.

Although the effluent ammonia-N concentration is slightly lower than the desired effluent concentration, the necessity to use full shafts of fixed area makes it unlikely that the media area could be reduced while still meeting the effluent requirement. Consequently, this is considered to be an acceptable design.

20.3.4 Pilot Plants

Pilot studies and/or the analysis of full-scale data from similar systems can provide the basis for the design of systems treating a wide variety of wastewaters, such as industrial wastewater, contaminated groundwater, and so on. Such data can also be used to optimize the expansion of an existing system. Pilot studies can be conducted with either a single stage or a multistage system. The surface loading rates on the pilot unit are varied and the pollutant concentrations into and out of each stage are measured, thereby allowing the fluxes to be determined for the various loadings. For existing full-scale systems, similar information can be obtained by collecting interstage data at the operating loading rate and then increasing the loading to individual units to determine performance limits. Regardless of whether the studies are performed on pilot- or full-scale systems, the process loading rate should be maintained at a constant value until steady-state operating conditions are achieved before collecting data. This may require two to three weeks or more for each loading rate. Either a single stage or a multistage pilot system may be rented or purchased. A multistage pilot unit requires fewer steady-state operating periods because of the unique loading to each stage, and thus need not be operated as long but costs more. Care must be exercised when analyzing the data from a multistage system because different microbial communities may develop on the individual stages. If this occurs, separate performance relationships must be developed for each stage.

Pilot studies should be conducted using full-scale (i.e., 3.6 m diameter) units rotating at 1.4 to 1.6 rpm. This is because of the effects of disc diameter and rotational speed on process performance, as discussed in Section 17.2.2. In the past, small diameter pilot-scale units have been used without consideration for the differences in mass transfer characteristics between pilot-scale and full-scale units.[14] The result has been full-scale units that did not perform as expected. Furthermore, as illustrated in Figure 17.21, scale-up by maintaining the same THL and peripheral velocity in the pilot unit as in full-scale units does not work either.[12,33] Consequently, the only safe approach is to use pilot units with full diameter discs. The shaft length is simply shorter.

Data from full-scale or pilot-scale units can be analyzed using the models and empirical relationships presented above. For the removal of biodegradable organic matter this consists of the first-order and second-order models. For nitrification this involves comparison of the measured ammonia-N flux with the typical values presented in Figure 20.5. In some instances, a Monod-type model has been fit to data from pilot units,[27] although it should be recognized that in such cases the half-saturation coefficient incorporates the mass transfer effects, as indicated in Equation 19.5 for trickling filters. Any result that indicates zero-order organic substrate or ammonia-N fluxes greater than those presented in this chapter should be viewed with skepticism. They indicate that, for some reason, the pilot unit is achieving a greater oxygen transfer rate than is typical in full-scale units.

As mentioned in Section 20.2.2, graphical procedures provide an alternative approach for using pilot-scale data to size an RBC system and they have been used successfully in several instances.[1,7,9] They are based on a mass balance across an RBC stage. The basic procedure will be illustrated for organic matter removal, but applies equally well for nitrification. To begin the procedure, pilot-plant data are used to calculate the removal rate per unit area (i.e., the flux) of organic matter, J_S, for each stage and/or each loading. The rates are then plotted as a function of the residual concentration in the stage and a smooth curve is drawn through the data. Such a curve will generally have a shape like that illustrated in Figure 20.9. Since each stage in an RBC can be considered to be completely mixed, a steady-state mass balance on soluble substrate across stage N is

$$F \cdot S_{S,n-1} - F \cdot S_{S,n} - J_{S,n} \cdot A_{s,n} = 0, \tag{20.8}$$

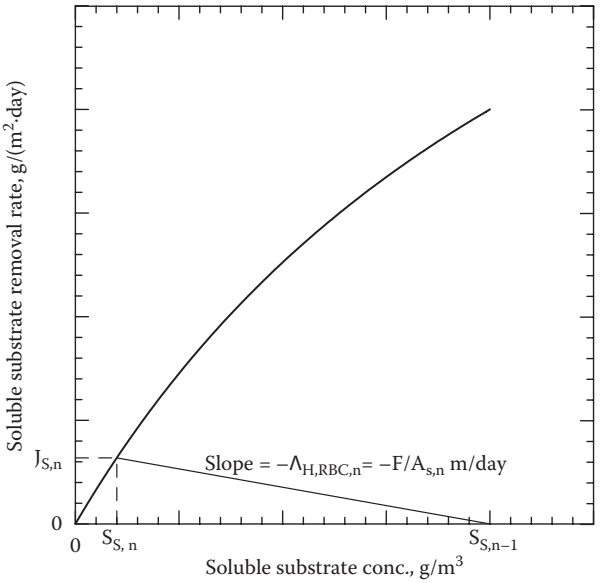

FIGURE 20.9 Illustration of the graphical procedure for determining the hydraulic loading required to reduce the soluble substrate concentration from S_{SO} to S_S in a single stage RBC.

where $J_{S,n}$ is the flux in stage N, which is related to the substrate concentration $S_{S,n}$ in that stage as illustrated in Figure 20.9. All other terms have been defined previously. Rearrangement of this equation provides the rationale for the graphical approach:

$$\frac{J_{S,n}}{S_{S,n} - S_{S,n-1}} = \frac{-F}{A_{s,n}} = -\Lambda_{H,RBC,n}. \tag{20.9}$$

As depicted by Equation 20.9 and illustrated in Figure 20.9, a line drawn from $J_{S,n}$ to the input substrate concentration, $S_{S,n-1}$, on the abscissa will have a slope of $-F/A_{s,n}$, which is just the THL on the stage times −1.0. This line is called the operating line. Once the plot of flux versus substrate concentration has been drawn from the pilot-plant data, the THL required to decrease the substrate concentration from $S_{S,n-1}$ to $S_{S,n}$ in a single stage can be determined by drawing an operating line from $S_{S,n-1}$ on the abscissa to the flux associated with $S_{S,n}$ and measuring the slope. Since the flow rate is known, the required stage surface area can be determined. Furthermore, by recognizing the definition of the SOL, it can be seen that the intersection of the operating line with the ordinate is that loading. This allows the graphical procedure to also be used to determine the output concentration from any stage in an RBC train. The SOL on the stage is calculated with Equation 19.2 and an operating line is drawn from the SOL on the ordinate to the influent substrate concentration on the abscissa. The intersection of that operating line with the curve gives the flux in that stage and the output substrate concentration, $S_{S,n}$.

The graphical procedure can also be used to determine the output concentration from a staged RBC system or to determine the size system required to achieve a desired effluent concentration. The former can be done directly whereas the latter requires an iterative approach. For an existing system, the THL is calculated for each stage. The procedure, illustrated in Figure 20.10, is initiated by plotting an operating line with slope $-F/A_{s,1}$ from the system influent substrate concentration, $S_{S,O}$ on the abscissa to the rate curve. The intersection of that operating line with the curve gives the output substrate concentration from stage one, $S_{S,1}$. Since that concentration is also the influent concentration to stage two, the procedure can be repeated to determine the output concentration

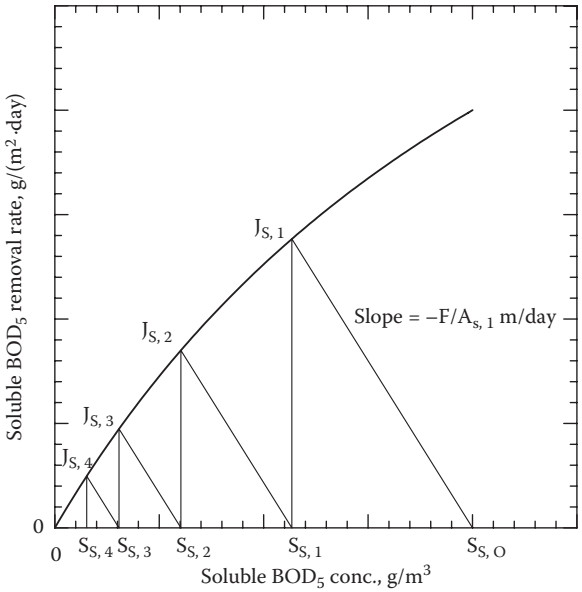

FIGURE 20.10 Illustration of the graphical procedure for determining the hydraulic loading required on each stage of a multistage RBC to reduce the soluble substrate concentration from S_{SO} to a desired effluent concentration, S_{Se}.

from stage two, and so on, until the final stage is reached, as shown in Figure 20.10. If all stages have the same hydraulic loading, then all of the operating lines will be parallel, as illustrated in the figure.

For design of a new system, an iterative procedure must be used. Generally, the smallest system will result when the first stage is loaded with an SOL of 32 g $BOD_5/(m^2 \cdot day)$ because that will minimize its size. Thus, to start the procedure, an operating line is drawn from the system influent substrate concentration on the abscissa to the limiting SOL on the ordinate. The intersection of that operating line with the rate curve gives the output substrate concentration from stage one. A tentative decision is then made about the number of additional stages needed and their size. If they are all the same size they will all have the same THL, making the slopes of the operating lines the same for all stages. The graphical procedure is then employed to determine if the tentative design will achieve the effluent quality goal. If it does not, then a new size is selected and the procedure is repeated. The only constraint is that each stage must consist of at least one complete shaft of standard (9300 m^2) or high density (13,900 m^2) media. The advantage of the graphical procedure is that it is rapid and allows the designer to visualize easily the impact of decisions. The procedure is illustrated in the following example.

Example 20.3.4.1

Data were collected with a single stage RBC pilot plant to determine the treatability of a soluble industrial wastewater. They were used to determine the relationship between the concentration of soluble BOD_5 and the related substrate flux into the biofilm, J_S, which is presented in Figure E20.1. The plotted points represent the data, whereas the curve represents the general relationship between the soluble BOD_5 flux and the soluble BOD_5 concentration. The wastewater to be treated has a design flow rate of 1500 m^3/day and a soluble BOD_5 concentration of 200 mg/L. Size both a single stage and a multistage system to reduce the soluble BOD_5 concentration to 20 mg/L or less.

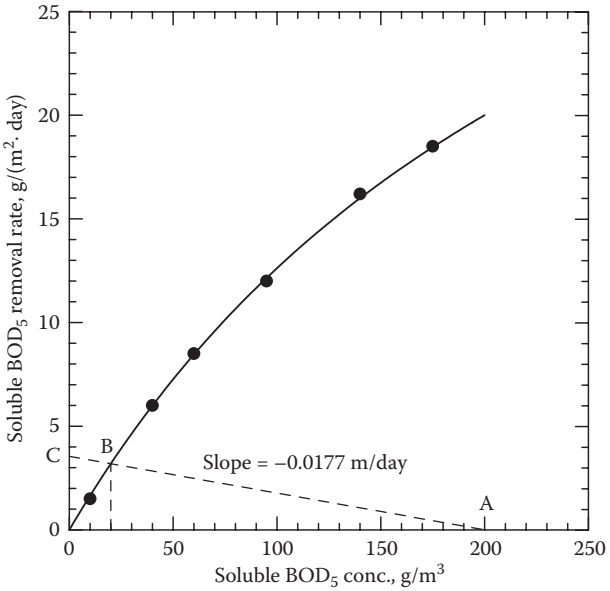

FIGURE E20.1 Application of the graphical procedure to determine the required hydraulic loading on the single stage RBC in Example 20.3.4.1.

a. How large would a single stage system have to be?
Two points must be located on the graph in Figure E20.1 to determine the size of a single stage system. The first is point A, the location of the influent substrate concentration on the abscissa. The second is point B, corresponding to the substrate flux associated with the desired effluent substrate concentration of 20 mg/L. The value of the flux at point B is 3.19 g $BOD_5/(m^2 \cdot day)$. An operating line is then drawn between A and B. Its slope gives the THL, which can be estimated by applying Equation 20.9:

$$-\Lambda_{H,RBC,n} = \frac{3.19}{20-200} = -0.0177 \text{ m/day}.$$

Since the flow rate is 1500 m^3/day, the required surface area of media is 84,600 m^2. Because of the high influent substrate concentration, standard density media (9300 m^2) should be used, requiring 9.10 shafts. Since whole shafts must be used, 10 shafts are required.

Given the low flux, it is unlikely that the SOL will be high enough to cause an oxygen limitation, but this can be easily checked by noting point C, the point at which the operating line intersects the ordinate. There it can be seen that the SOL is only 3.5 g $BOD_5/(m^2 \cdot day)$, which is well below the maximum allowable.

b. How large would a multistage system have to be?
This analysis is presented in Figure E20.2, which is drawn with an expanded ordinate. The SOL on the initial stage is limited to 32 g $BOD_5/(m^2 \cdot day)$ to avoid oxygen deficiencies and excessive *Beggiatoa* growth. The operating line for the first stage is constructed as line A-B, beginning at the influent BOD_5 concentration of 200 mg/L on the abscissa (point A) and terminating on the ordinate at the maximum allowable SOL of 32 g $BOD_5/(m^2 \cdot day$; point B). The slope of this line is –0.16 m/day, giving the THL on the first stage. The substrate flux into the biofilm on the first stage is 13.8 g $BOD_5/(m^2 \cdot day)$, giving an effluent soluble BOD_5 concentration of 113 mg/L, as indicated by points C and D in the figure. Since the flow rate is 1500 m^3/day and the THL is 0.16 m/day, the required surface area in the first stage is 9375 m^2. The difference between the required area and the area provided by a standard density shaft (9300 m^2) is within the error of the analysis. Consequently, a single standard density shaft can be used.

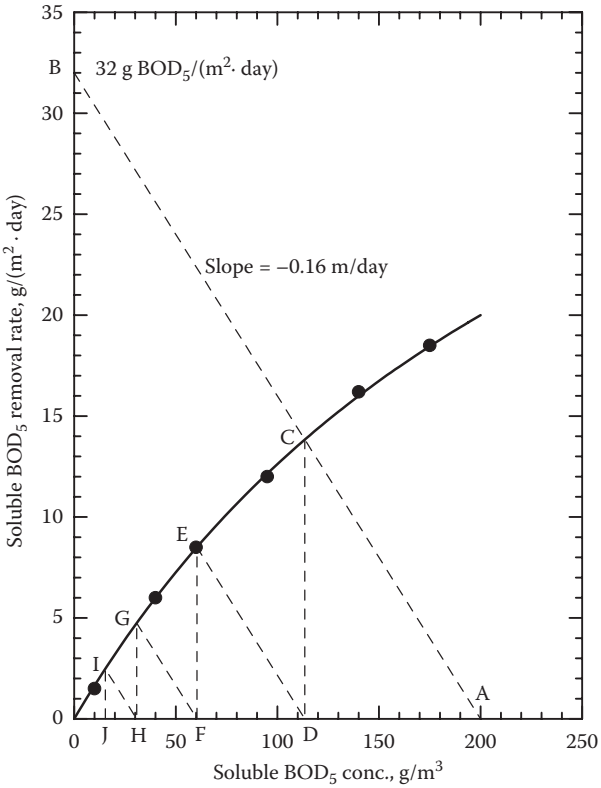

FIGURE E20.2 Application of the graphical procedure to determine the required hydraulic loading on each stage of the multistage RBC in Example 20.3.4.1.

As a first trial, assume that all stages will be equal sized, so that the THL on each stage will be the same. Consequently, the substrate flux into the biofilm and the output soluble BOD_5 concentration from each subsequent stage can be determined by using operating lines of equal slope, which will be parallel to the operating line for the first stage. From Figure E20.2 it can be seen that four stages are required to reduce the effluent soluble BOD_5 concentration to less than 20 mg/L. In fact, the effluent $SBOD_5$ will be about 15 mg/L. As in the first stage, each stage would contain a single, standard density media shaft.

The multistage system requires four standard density media shafts, whereas the single stage system requires 10. This illustrates the benefits of staging in RBC systems.

20.3.5 General Comments

The design of RBC systems must consider not only their normal operation, but also operation under unusual conditions. Consequently, operational flexibility must be provided. The following is a list of factors that should be considered and/or provided:[31]

- The need for supplemental aeration for mechanical drive systems. Supplemental aeration can help control biomass growth on the discs and allow the first stage to better accommodate SOLs approaching the oxygen transfer limit. Aeration is particularly needed when the influent hydrogen sulfide concentration is elevated.
- A means for removing excess biomass, such as by air or water stripping, chemical addition, or rotational speed control and/or reversal.

- Multiple treatment trains. Multiple trains allow a portion of the system to be taken off-line for maintenance while minimizing the impact on effluent quality. They also allow the number of trains in service to be adjusted in proportion to the actual long-term plant load.
- Removable baffles between all stages. These allow the degree of staging to be altered.
- Positive influent flow control to each train.
- Alternate operating modes such as step feeding. These allow the loading to a particular stage to be changed.
- Positive air flow metering to each shaft when supplemental aeration or air drive units are used.
- Recirculation of treated effluent to the first stage. This reduces influent concentrations and uses the dissolved oxygen in the recirculation to oxidize hydrogen sulfide in the influent.
- Methods for removing hydrogen sulfide from the influent (e.g., by oxidation or precipitation) or for reducing the influent organic loading (e.g., by chemical addition to the upstream primary clarifier).
- The provision of hydraulic or electronic load cells on selected shafts if excess biomass buildup is a possibility.

20.4 PROCESS OPERATION

Efficient and effective operation of an RBC system is achieved by systematic application of the principles described above. Appropriate overall and individual stage organic and ammonia loadings must be maintained. This may require periodic removal of RBC trains from service or their return to service in response to long-term changes in process loadings. Interstage baffles may also be relocated as process loadings change. For carbon oxidation applications, it may be possible to save energy by removing trains from service when the long-term loading is less than the design loading. Furthermore, this may be necessary when it is desirable to minimize nitrification, which will tend to occur in an underloaded system. On the other hand, if nitrification is required, then a sufficient number of trains should always be in service both to remove the influent organic matter and to nitrify.

Achieving the objectives described above will require active management of the number of trains in service. If an RBC train is to be taken out of service for a short period of time, it can be idled by discontinuing its feed but continuing to rotate the media to keep the biofilm wet. The activity of the biomass will slowly drop as it undergoes decay, but the biofilm will remain attached to the media. In this way, when the unit is brought back on line, only a few days will be required to achieve full treatment capacity, with the number depending on the length of time the unit has been out of service. If the unit will be out of service so long that essentially all activity is lost, the unit should be drained and flushed. If this is not done and rotation is stopped, the biomass exposed to the air will dry out and partially detach from the media. This biomass will slough off of the media when rotation is resumed, and new growth will have to be established. Furthermore, odors will develop because the water in the bioreactor will become anaerobic due to lack of aeration. Once an RBC is devoid of biofilm, 7 to 10 days are required for establishment of a significant biofilm when it is placed back in service, and 20 to 30 days may be required before its full treatment capacity is achieved.

Like all biochemical operations, effective operation of an RBC system requires routine sampling and analysis, as well as the performance of preventative maintenance. These topics are covered well in a variety of sources.[30,32]

20.5 KEY POINTS

1. Rotating biological contactors (RBCs) contain vertical discs of corrugated plastic media mounted on a horizontal shaft, submerged approximately 40% in the process flow, and rotated at 1.6 rpm. Several manufactures produce RBC equipment, but they all use standard

dimensions and similar configurations. A typical media bundle (usually referred to as a shaft) is 3.66 m in diameter and 7.62 m long. Standard density media provides around 9300 m^2 of surface area per shaft while high density media provides 13,900 m^2.

2. Individual RBC units are arranged into treatment trains generally containing 4 to 8 shafts each. The treatment trains operate in parallel and independently of each other. Within a treatment train, wastewater flows in series from one RBC shaft to the next. Moveable baffles are used to establish physically and functionally distinct zones called stages. Each stage contains one or more RBC shafts.
3. Rotating biological contactor installations are classified using two criteria: the treatment objective and the equipment used. Treatment objectives include the removal of biodegradable organic matter, combined carbon oxidation and nitrification, and separate stage nitrification. Equipment types include mechanically driven and air driven units.
4. Many RBC systems designed and installed prior to 1980 experienced significant equipment and/or performance problems. However, based on today's understanding, RBC facilities can be properly designed and operated.
5. The surface organic loading (SOL), typically expressed as the g $BOD_5/(m^2 \cdot day)$, is one of the principle factors determining the overall performance of an RBC. It determines the extent to which biodegradable organic matter is oxidized and whether nitrification will occur.
6. The SOL on any RBC stage must not exceed a limiting value, generally taken to be 32 g $BOD_5/(m^2 \cdot day)$, which represents the maximum rate at which organic matter can be applied while still maintaining aerobic conditions within the RBC biofilm.
7. Nitrification occurs in RBC systems when the concentration of soluble, biodegradable organic matter surrounding the biofilm is reduced below about 20 mg COD/L (15 mg BOD_5/L). Equation 20.1 can be used to adjust nitrification rates for the presence of heterotrophs in the biofilm when both carbon oxidation and nitrification are occurring in the same RBC stage.
8. The zero-order ammonia-N flux in separate stage nitrification systems is approximately 1.5 g $N/(m^2 \cdot day)$ and is limited primarily by the maximum rate of oxygen transfer into the biofilm. This flux is observed when ammonia-N concentrations exceed 5 mg/L. Lower fluxes are observed at lower ammonia-N concentrations.
9. As in most biochemical operations, staging allows advantage to be taken of the effect of bulk substrate concentration on the reaction rates in RBC systems, thereby reducing the required media volume. The higher SOLs in the initial stages of the RBC cause higher bulk substrate concentrations, which allow high reaction rates, while the lower SOLs on the downstream stages allow the desired effluent quality to be achieved.
10. The effect of temperature on RBC reaction rates is relatively small down to a temperature of about 15°C. This occurs because the flux to the biofilm is generally controlled by diffusion rather than by microbial reactions down to that temperature. Below 15°C, however, biological reaction rates slow sufficiently to make the flux dependent on temperature. Relationships are available to characterize the effects.
11. Excessive growths of the sulfur oxidizing bacteria *Beggiatoa* occur when hydrogen sulfide and biodegradable organic matter are present in sufficient concentrations. Hydrogen sulfide can be present in the influent wastewater or it can be produced in the biofilm by anaerobic conditions caused by organic overloading.
12. The conditions leading to excessive *Beggiatoa* growth are well defined and techniques are available for controlling it. One important condition is to keep the SOL below 32 g $BOD_5/(m^2 \cdot day)$.
13. Two models frequently used to characterize the flux of biodegradable organic matter into an RBC biofilm are the first-order model, which is similar to the Velz-Germain equation used to size trickling filters, and the second-order model. Typical values of the model

parameters have been developed for domestic wastewaters. Site-specific values must be developed for industrial wastewaters.

14. Care must be exercised when interpreting data from RBC pilot plants. Since maximum fluxes in RBC systems are generally controlled by oxygen transfer rates, and the oxygen transfer characteristics of smaller pilot-scale units may exceed those of full-scale units, the pilot plant may overestimate full-scale RBC treatment capacity. The safest approach is to use pilot plants containing full diameter discs running at typical rotational speeds.
15. Design of an RBC system requires determination of the total media surface area required and distribution of that media into parallel treatment trains containing individual shafts in series. The designer has considerable latitude in selection of the system configuration.
16. Several features, listed in Section 20.3.5, should be considered for incorporation into any RBC facility. These features provide the flexibility required to allow the system to be adjusted to meet changing operational needs.
17. Operation of an RBC system requires maintenance of proper loadings on the treatment units. This will require units to periodically be taken out of service or returned to service.

20.6 STUDY QUESTIONS

1. Summarize the current developmental status and typical applications for RBC wastewater treatment technology.
2. Describe the relative advantages and disadvantages of mechanical versus air drives for RBC units.
3. Prepare a table summarizing the primary benefits and drawbacks of RBCs.
4. Describe the sequence of events that occurs when the SOL on the first stage of an RBC system exceeds 32 g $BOD_5/(m^2 \cdot day)$. Why do those events occur?
5. Explain why the concentration of soluble, biodegradable organic matter must be lowered to about 20 mg/L as COD before nitrification can occur in an RBC.
6. Describe why the substrate removal rate in RBCs is generally not affected by temperature until it is relatively low. At what temperature is an effect generally observed?
7. Describe the mechanism for growth of the nuisance microorganism *Beggiatoa* in RBC systems.
8. Using the graphical procedure illustrated in Section 20.3.4, explain why staging generally improves the performance of an RBC system. What constraints exist on the use of staging to improve RBC performance?
9. What wastewater characteristics are particularly important when an RBC is being considered for a specific application? How are these characteristics related to RBC performance?
10. Summarize the factors that should be considered in the physical design of an RBC system.
11. Domestic wastewater with a design flow rate of 15,000 m^3/day, a BOD_5 concentration of 90 mg/L, a soluble BOD_5 concentration of 60 mg/L, and an ammonia-N concentration of 20 mg/L must be treated.
    a. Use the first-order model to size an RBC system to reduce the BOD_5 to 20 mg/L. In your design, select the total number of shafts required and the manner in which they should be arranged.
    b. Check the proposed design using the second-order model and explain any differences that might exist between the two designs.
    c. Design a combined carbon oxidation and nitrification system to reduce the ammonia-nitrogen concentration to 2 mg/L. In your design, select the total number of shafts required and the manner in which they should be arranged.

12. A wastewater with a flow of 20,000 m^3/day, a BOD_5 of 15 mg/L, and an ammonia-N concentration of 30 mg/L must be treated to reduce the effluent ammonia-N concentration to less than 0.5 mg/L. Size an RBC system to accomplish this treatment goal. Assume that Figure 20.5 represents the effect of the ammonia-N concentration on the nitrification rate.
13. An industrial wastewater with a flow rate of 10,000 m^3/day and a soluble BOD_5 concentration of 150 mg/L must be treated to a soluble BOD_5 concentration of 10 mg/L. A pilot-plant study provided the data in Figure E20.2. Use the graphical procedure with that data to design a multistage RBC system to meet the treatment objective. Justify all decisions made for the design.

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21 Submerged Attached Growth Bioreactors

Attached growth processes have been used for the aerobic treatment of wastewaters for over a century, primarily in the form of trickling filters and rotating biological contactors. Nevertheless, during the past three decades much has been learned about the mechanisms by which such processes operate, leading to the development of new bioreactor configurations, such as the fluidized bed biological reactor (FBBR) discussed in Chapter 18. Like the FBBR, this new generation of attached growth bioreactors tends to have media that is submerged in the process flow; hence the name submerged attached growth bioreactor (SAGB). This chapter introduces some of the more promising SAGBs.

21.1 PROCESS DESCRIPTION

Many SAGB configurations have been conceived of and evaluated. Some were specifically developed for and applied to anaerobic systems. (See Chapter 14 for more detail on anaerobic systems.) In general, interest in submerged attached growth systems stems from the high biomass concentrations that can be achieved, resulting in short hydraulic residence times (HRTs) in comparison to suspended growth systems with equivalent solids retention times (SRTs).[68] This results from the use of media with high specific surface areas, as discussed in Section 18.1.1. Short HRTs result in compact systems, which can be quite advantageous when land area is limited. These highly compact systems generally use submerged media operated in either an upflow or a downflow mode. Integrated fixed film activated sludge (IFAS) systems, in which submerged media is added to a suspended growth system, are also receiving interest.

21.1.1 General Description

Figure 1.32 provides a schematic diagram of a typical SAGB. Its primary components are a reactor vessel, support media for biofilm growth, influent distribution system, and effluent withdrawal system. An oxygen transfer system may also be provided. Influent wastewater is added to the bioreactor and microorganisms grow attached to the submerged media, thereby accomplishing biochemical conversions. The types of microorganisms that grow depend on the constituents in the wastewater and the environmental conditions provided in the bioreactor. Flow may be upward, downward, or horizontally across the media, and a wide variety of media types can be used. Table 21.1 summarizes the media characteristics and operating conditions for the SAGBs considered in this chapter. As indicated there, the media may be packed, fluidized, or free floating. Effluent is recirculated in some instances to maintain required fluid velocities through the system, as described in Chapter 18. Significantly, SAGBs use media with a high specific surface area, about an order of magnitude higher than the specific surface areas of trickling filter or rotating biological contactor (RBC) media. This suggests that SAGBs can contain much more biomass, thereby allowing smaller HRTs to be used and, indeed, this has been their main appeal.

Many SAGBs are operated without a solids separation and recycle system. In such bioreactors, the hydraulic residence time (HRT) is less than the minimum solids retention time (SRT) required for microbial growth on the substrates provided. As a result, the growth of suspended microorganisms

TABLE 21.1
Packed and Fluidized Bed Reactors

Process	Media Characteristics			Typical Hydraulic Loading Rate, m/hr		Porosity %	Comments
	Type	Size mm	Specific Surface Area m^2/m^3	Average	Peak		
Packed bed							
Downflow	Fired clay	2–6	1000–1500	2–3	5–6	40	Hydraulic loading rates vary with application. Specific surface area excludes internal micropores.
	Rounded sand	2–3	1500	2–3	5–6	40	Hydraulic loading rates vary with application.
Upflow	Fired clay	2–6	1000–1500	2–3	5–6	40	Hydraulic loading rates vary with application. Specific surface area excludes internal micropores.
	Polystyrene beads	2–3	1000–1100	2–3	5–6	40	Hydraulic loading rates vary with application.
Fluidized bed	Sand	0.3–0.7	2600–3900	24 (min)	36 (max)	60	Hydraulic loading rate dictated by media size and density.
	GAC	0.6–1.4	2300–3300	27 (min)	37 (max)	60	Hydraulic loading rate dictated by media size and density. Specific surface area excludes internal micropores.
Free floating	Plastic-K1	9 × 7[a]	500	N/A[b]	N/A[b]	70[c]	Horizontal flow through media suspended by aeration or mixing in conventional bioreactor
	Plastic-K2	15 × 15[a]	350	N/A[b]	N/A[b]	70[c]	
	Plastic-K3	25 × 12[a]	500	N/A[b]	N/A[b]	70[c]	
	Plastic-Biochip	48 × 2.2[a]	1200	N/A[b]	N/A[b]	70[c]	

[a] Diameter × length for cylindrical shape.
[b] Not applicable. Horizontal flow through media.
[c] Maximum fill fraction.

is minimized and the growth of attached microorganisms is maximized. For these systems, a means for controlled removal of the attached biomass must be provided. The means depends on the system configuration and is discussed in conjunction with each system.

Other SAGBs use both suspended and attached growth biomass. These IFAS systems are illustrated schematically in Figure 1.33, and the characteristics of the media used in them are summarized in Table 21.2. The suspended component of the biomass is maintained by passing the bioreactor effluent through a liquid-solids separation device, such as a clarifier, and recycling the separated biomass to the bioreactor in the same manner as in the activated sludge process. Biomass that grows on the fixed media will occasionally slough off and be incorporated into the suspended biomass. The excess biomass produced is removed as waste solids, just like in the activated sludge process.

Heterotrophic microorganisms grow in these systems if biodegradable organic matter is present in the influent wastewater, but the type of metabolism they exhibit depends on the terminal electron acceptor provided. Aerobic metabolism predominates if oxygen is supplied to the system, either dissolved in the influent wastewater or transferred to the bioreactor contents by the addition of air or pure oxygen. Denitrification will occur if dissolved oxygen (DO) is not supplied but nitrate-N is provided. Anaerobic processes will occur if both oxygen and nitrate-N are excluded.

Nitrogenous compounds in the wastewater can also be transformed if appropriate environmental conditions are established. Influent ammonia-N can be nitrified if sufficient oxygen is supplied and the total organic loading (TOL) is low enough for nitrifying bacteria to grow. Alternatively, influent nitrate-N can be denitrified if oxygen is excluded and organic matter is available. The development of a microbial community capable of achieving biological phosphorus removal has also been demonstrated for at least one bioreactor configuration.[13] Thus, a wide range of biological conversions can be accomplished in these reactors.

21.1.2 Downflow Packed Bed Bioreactors

Figure 21.1 provides a schematic diagram of a downflow packed bed (DFPB) bioreactor. Wastewater is applied to the top and passes downward through the media to an effluent collection system. Relatively small diameter media and modest superficial velocities are used, as indicated in Table 21.1, which prevents biomass from being detached and transported through the media bed. These characteristics allow DFPB bioreactors to function as filters and to remove particulate matter contained in the wastewater. As a result, these bioreactors provide both biological treatment and filtration. Because suspended solids contained in the wastewater will accumulate in the bioreactor, they must periodically be removed by backwashing. Backwashing also removes microorganisms that grow within the media, thereby controlling the biomass inventory. As a result of the need for backwashing, the effluent collection system is designed much like the underdrain system of conventional granular media filters. This means that it must both collect effluent and distribute backwash water uniformly over the bottom of the media bed. Backwash collection facilities are also required.

Downflow packed bed bioreactors have been used for carbon oxidation, combined carbon oxidation and nitrification, and separate stage nitrification.[40,54,65,68] Oxygen must be supplied to accomplish nitrification and is often supplied in carbon oxidation systems. This is done by adding air through a distribution system located about two-thirds of the way from the top of the bioreactor. Oxygen transfer can be quite efficient due to the tortuous paths that air bubbles must follow as they pass upward through the packed bed. The addition of air to the bioreactor results in its division into two zones. The upper two-thirds is the aerobic bioreactor section, while the lower one-third functions as a filter. One proprietary version of DFPB bioreactors uses media with an effective size of 2 to 6 mm that is manufactured from a fired clay.[38] Experience indicates that the porous media formed in the firing process provides internal surface area that retains bacteria in the packed bed during the backwashing process. Activated carbon has also been used as media. Aerated DFPB bioreactors are sometimes referred to as biological aerated filters (BAFs).

TABLE 21.2
Example Media for Integrated Fixed Film Activated Sludge (IFAS) Systems

Media	Media Dimensions	Volume % of Aeration Basin	Biomass/Activity	Comments
Reticulated polyurethane foam pads		10–30	5–20 g/L of pad volume; 5 mg NH_3-N/(day·pad) for nitrification.	Higher biomass concentrations lead to anaerobic conditions.
Captor	1.3 cm × 2.5 cm × 2.5 cm			
Linpor	1.0 –1.3 cm cubes			
Ringlace	Looped PVC about 5 mm in diameter, placed 4–10 cm apart. Density of 120–500 m/m^3	25–35	3–12 g/m of media length; up to 1.75 kg NH_3-N/(day·1000 m) for nitrification.	Low organic loadings lead to worm growth. Worms controlled by anoxia.
Trickling filter media	PVC vertical trickling filter media, 90–165 m^2/m^3	25–75 (with or without RAS recycle)	1500–6000 g/m^3 of media. See Figure 21.17 for activity.	Can use with or without suspended growth produced by RAS recycle from secondary clarifier.

Note: The free floating plastic media in Table 21.1 are also used, but the information about them is not repeated here.

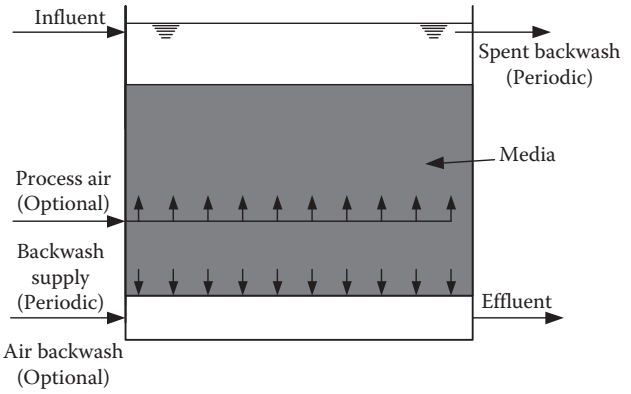

FIGURE 21.1 Downflow packed bed (DFPB) bioreactor.

Downflow packed bed bioreactors can also be used for denitrification. One proprietary device containing 2 mm diameter rounded sand has proven effective for combined denitrification and suspended solids removal.[64,70] Because both biodegradable organic matter and nitrate-N must be present in the influent wastewater for denitrification to occur, organic matter must be added when the wastewater has been nitrified in an upstream treatment system, as illustrated in Figure 21.2a. Materials such as methanol,[64,70] fermented wastewater, and fermented solids[1,22] can be used for this purpose. Alternately, effluent from a downstream nitrification process can be recirculated, mixed with influent wastewater, and applied to a DFPB denitrification bioreactor,[11,31,38,68] much in the manner of the modified Ludzak-Ettinger (MLE) suspended growth system, as illustrated in Figure 21.2b. In this case, however, a distinct microbial community will develop in each bioreactor. Nitrogen gas will accumulate in a DFPB bioreactor when it is used for denitrification. This gas must be purged periodically from the bioreactor or it will displace fluid from the interstitial spaces and restrict liquid flow. Purging is often accomplished by temporarily stopping the influent flow and using an air-wash cycle.

21.1.3 Upflow Packed Bed Bioreactors

The upflow packed bed (UFPB) bioreactor, illustrated in Figure 21.3, is a more recent development. It is similar to the DFPB units described above, except that flow is upward rather than downward. Similar media sizes and superficial velocities are used, as indicated in Table 21.1. Bed depths tend to be between 2 and 4 m. Aeration is provided in the same manner as in DFPB bioreactors, thereby creating an aerobic bioreaction zone in the top and a filtration zone in the bottom. Alternatively, aeration can be provided from the bottom, if the entire bed is to be aerobic and used for bioreaction. Upflow packed bed bioreactors are used for carbon oxidation; combined carbon oxidation and nitrification; separate stage nitrification; separate stage denitrification; and combined carbon oxidation, nitrification, and denitrification.[38,61,66,68] The last operational mode is achieved by sizing the aerated zone so that nitrification is achieved and recirculating nitrified effluent to the bioreactor influent to supply nitrate-N to the lower portion of the bioreactor, thereby converting it into an anoxic zone.

One commercial version of the UFPB system uses the fired clay media described above.[38] This media is heavier than water and rests on an underdrain system. Another commercial system uses 2 to 5 mm plastic beads, which are lighter than water and tend to float.[31] As a result, a grid is required at the top of the bioreactor to retain the media, but an underdrain system is not required. Like aerated DFPB bioreactors, when UFPB bioreactors are operated in an aerated mode they are commonly referred to as biological aerated filters (BAF).

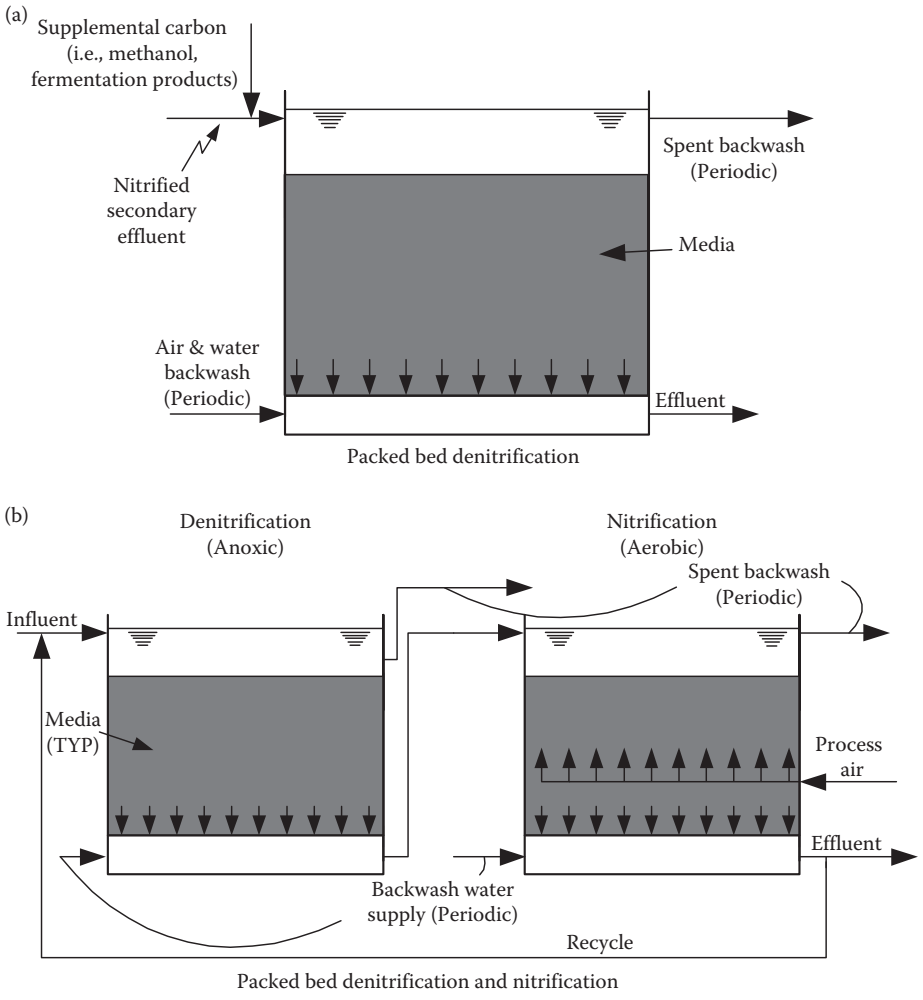

FIGURE 21.2 Downflow packed bed systems for denitrification.

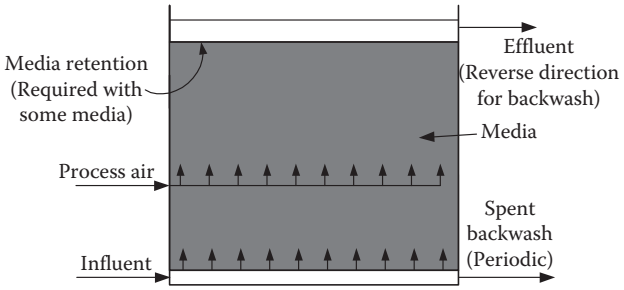

FIGURE 21.3 Upflow packed bed (UFPB) bioreactor.

21.1.4 Fluidized and Expanded Bed Biological Reactors

Chapter 18 provides a detailed description of fluidized bed biological reactors (FBBRs) and a theoretical analysis of their performance. This section summarizes information about FBBRs to allow their comparison with the other SAGBs.

Figure 1.29 provides a schematic diagram of the FBBR process. Process influent flow, consisting of a mixture of wastewater and recirculated effluent, is added to the bottom of the bioreactor through a distribution system and flows upward to a collection system, thereby fluidizing and expanding the media bed.[7] A variety of media can be used, but, as indicated in Table 21.1, silica sand with a diameter of 0.3 to 0.7 mm and granular activated carbon (GAC) with a diameter of 0.6 to 1.4 mm are used most often. The small media diameter provides a large specific surface area for biofilm growth. As discussed in Section 18.2.2, the diameter of a bioparticle increases, but its density decreases, as biofilm growth develops on it. The net result is a decrease in its settling velocity, resulting in the migration of particles with the greatest amount of biofilm to the top of the fluidized bed. The mass of biomass in the bioreactor is controlled by periodically removing bioparticles from the top of the bed and processing them through a media cleaning system. The media cleaning system typically consists of a pump (where turbulence shears the attached biofilm) and a liquid-solids separation device (such as a cyclone) where the media is separated from the biomass. The removed biomass is directed to solids processing, while the cleaned media is returned to the bioreactor.

Fluidized bed bioreactors can be applied to aerobic carbon oxidation,[7] combined carbon oxidation and nitrification,[23] separate stage nitrification,[8,14] denitrification,[64] and anaerobic treatment.[56,58] For aerobic applications, a portion of the process influent flows through a pressurized oxygen transfer vessel where it is saturated with pure oxygen, as illustrated in Figure 1.29. No oxygen is supplied for anoxic applications. However, an electron donor must be supplied either by the influent wastewater or by addition of a supplement like methanol. For a wastewater containing both biodegradable organic matter and ammonia-N, combined nitrification and denitrification using the wastewater organic matter requires two FBBRs operating in series in the same manner as described for DFPB bioreactors and illustrated in Figure 21.2b. The first FBBR is anoxic and receives influent wastewater and recirculation from the downstream aerobic FBBR. Anaerobic operation of an FBBR requires the exclusion of both oxygen and nitrate-N.[15,21] Somewhat smaller sand media has been used for anaerobic applications, ranging from 0.2 to 0.5 mm. The units can be operated in either an expanded or a fluidized mode, depending on the upflow velocity. In the expanded bed biological reactor (EBBR) mode the velocity is sufficient to expand the bed by 15 to 30%. Particles in the bed are partially supported by the upflowing fluid and partially by contact with adjacent particles and, consequently, they tend to remain in the same relative position within the bed. In the FBBR mode a higher upflow velocity is used, suitable to expand the bed between 25 and 300% so that the particles are fully supported by upflowing fluid, allowing free movement of particles within the bed. A facility designed as an FBBR can also be operated as an EBBR, but a facility designed as an EBBR may not have sufficient height to be operated as an FBBR. The two operating modes result in different biomass densities in the bed and different mass transfer characteristics. Higher reaction rates can be achieved in EBBRs but control of biomass is more effective in the FBBR. Mixing of the bed is created by both the upflow of process flow and gas produced by the anaerobic biological reactions. Effective biomass concentrations on the order of 15 to 35 g/L as volatile suspended solids (VSS) can be accumulated, thereby allowing relatively low HRTs (0.2 to 2 days) and high volumetric organic loadings with values over 20 kg COD/(m^3·day) common.[15,21]

21.1.5 Moving Bed Biological Reactors

The moving bed biological reactor (MBBR) uses free floating plastic media as characterized in Table 21.1. Initially developed in Norway and marketed as Kaldnes media,[36] but with competitive products now available in the marketplace, this media is essentially neutrally buoyant when biofilm

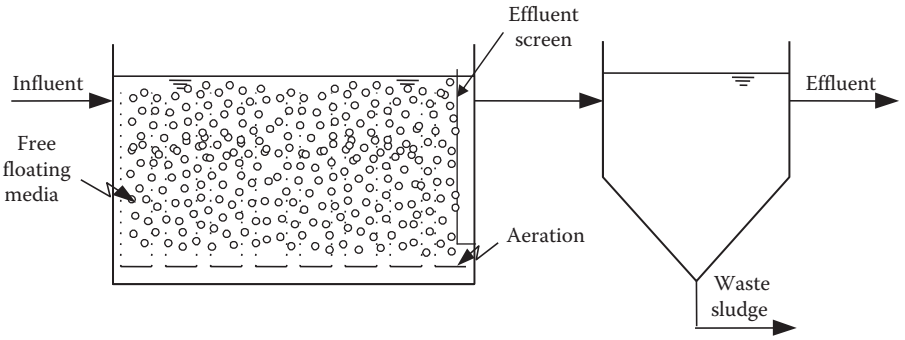

FIGURE 21.4 Moving bed biological reactor (MBBR).

is attached to it. Similar in configuration to random trickling filter media (illustrated in Figure 19.3) but smaller in size, these media are placed in conventional suspended growth bioreactors, as illustrated in Figure 21.4. Media circulation within the bioreactor is provided either by the aeration system (when aerobic conditions are maintained—as illustrated in Figure 21.4) or by mixers (when oxygen is excluded). Media is retained in the bioreactor using effluent screens of various configurations. Excess biomass sloughs off of the media and passes into the process effluent where it must be separated in a downstream liquid-solids separation system. Clarifiers can be used (as illustrated in Figure 21.4), as can alternative systems such as dissolved air flotation.

If oxygen is provided, the MBBR process can be used for carbon oxidation, combined carbon oxidation and nitrification, or separate stage nitrification. If oxygen is excluded, it can be used for denitrification, if nitrate and an electron donor are present and mixing is provided.[35,36,42,43] Baffles can be used to create various zones within the bioreactor and either aeration or mixing can be provided to allow the desired biochemical environment to be created. Thus, separate zones can be created for carbon oxidation, nitrification, and denitrification. Nitrified process flow can be recirculated from a downstream aerobic nitrification zone to provide nitrate to an upstream denitrification zone, allowing organic matter from the influent wastewater to be used as an electron donor, in a manner similar to the MLE process. Separate stage denitrification can occur when an electron donor is added to nitrified process flow in a mixed zone. Excess biological phosphorus removal is not easily accomplished in an MBBR because cycling of biomass through anaerobic and aerobic zones is necessary for the requisite biomass to develop. The free floating media is also used in IFAS systems.

21.1.6 Integrated Fixed Film Activated Sludge

Integrated fixed film activated sludge systems, illustrated in Figure 1.33, consist of an activated sludge bioreactor with attached growth media in it.[29] Suspended biomass is removed in the secondary clarifier and recycled to the bioreactor to maintain a desired suspended biomass inventory, just as in the activated sludge process. Biomass also accumulates on the media and provides additional biomass inventory. Excess attached biomass periodically sloughs from the media and is incorporated into the suspended biomass. Suspended biomass is wasted from the system to maintain a desired suspended growth SRT. Just as with the MBBR process, various anoxic and aerobic zones can be created, thereby allowing carbon oxidation, combined carbon oxidation and nitrification, and combined carbon oxidation, nitrification, and denitrification to occur. In addition, recirculation of suspended biomass from an aerobic zone to an anaerobic zone where volatile fatty acids are present can allow biological phosphorus removal to occur. Essentially all of the biological nutrient removal processes described in Chapter 12 can be implemented using IFAS.

The attached biomass contained in IFAS systems contributes to carbon oxidation, nitrification, and denitrification, and to the production of suspended biomass with improved settling

characteristics, allowing sludge volume indexes as low as 50 mL/g to be routinely observed in some systems.[30] Consequently, the addition of media increases the total biomass in the system in two ways: by providing sites for attached growth and by allowing an increased mixed liquor suspended solids (MLSS) concentration to be maintained through improved settling properties.

A variety of media have been used in IFAS systems, as listed in Table 21.2. Historically plastic sheet trickling filter media (see Figure 19.3), Ringlace, and porous polyurethane foam pads (in processes called Captor and Linpor) have been used. Currently the floating plastic media used in the MBBR process is gaining popularity.

Ringlace is a flexible, looped, rope-like material, constructed of polyvinyl chloride woven into strands, that provides a high surface area for biological growth.[5,29,63,68] It is hung on racks suspended over the air diffusers in the bioreactor. This placement results in circulation of mixed liquor past the media, thereby transporting substrate and DO to the attached growth. A similar media placement is used for plastic trickling filter media.[5,63,68] Figure 21.5 illustrates the placement of Ringlace or trickling filter media in IFAS systems.

In the Captor and Linpor processes, polyurethane foam pads are placed in the bioreactor in a free floating fashion and retained there by effluent screens, as illustrated in Figure 21.6.[5,12,63,68] Even when biomass accumulates inside them, the pads may float on the liquid surface due to their low density. The action of the diffused aeration system causes the pads to circulate within the bioreactor, although there may be a tendency for them to accumulate in the effluent end of the bioreactor. Air lift pumps are often used to recirculate the foam pads to counter this tendency and to maintain a uniform distribution throughout the bioreactor. Other mechanisms, such as a stream of air, are used to remove pads from effluent screens. A similar configuration is used with free floating plastic media.

The oxygen requirements of both the suspended and attached biomass must be met by the oxygen transfer system installed in the bioreactor. As illustrated in Figure 21.5, a spiral roll diffuser layout is required to ensure recirculation of mixed liquor through fixed media. This layout may also be needed with some free floating media to mix them vertically in the bioreactor and prevent their accumulation at the liquid surface. Spiral roll aeration systems have slightly lower oxygen transfer

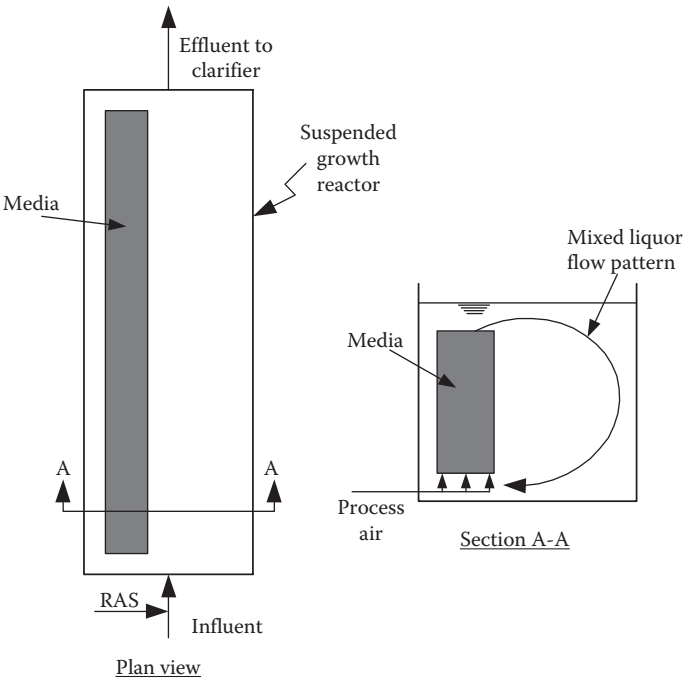

FIGURE 21.5 Placement of fixed media in integrated fixed film activated sludge systems.

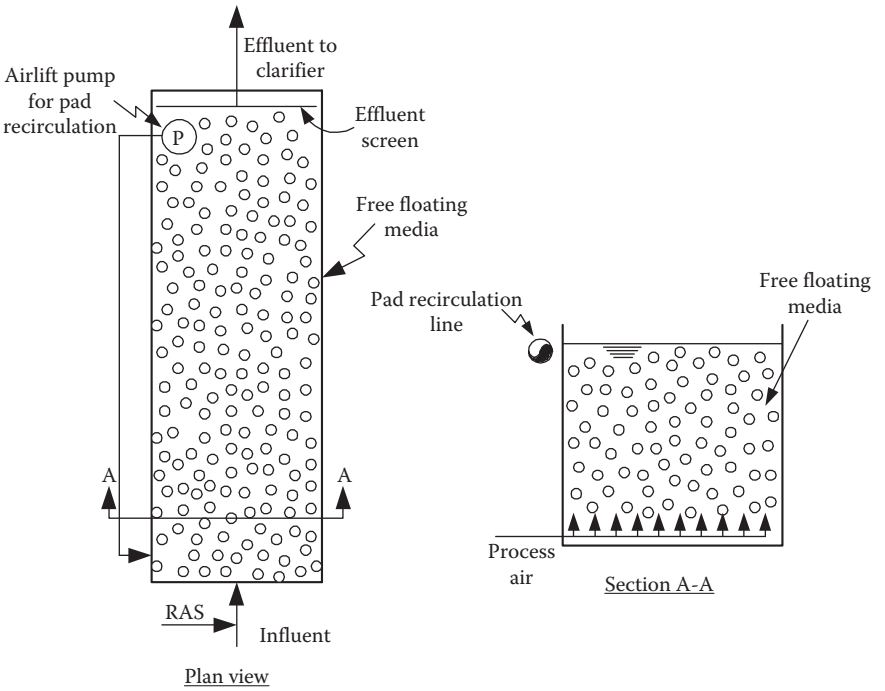

FIGURE 21.6 Configuration of integrated fixed film activated sludge systems using free floating media.

efficiencies than full floor coverage systems, suggesting that more energy may be needed with IFAS systems.[62] As discussed in Section 21.2.6, an increased DO concentration must also be maintained in these systems. The reduced basin volumes made possible by the presence of both suspended and attached biomass results in increased volumetric oxygen transfer requirements, which can exceed the values typically possible with conventional oxygen transfer systems in purely suspended growth systems (discussed in Sections 10.3.4 and 11.2.5). Fortunately, the presence of media in some IFAS systems apparently causes increased bubble hold up, thereby increasing the volumetric oxygen transfer rates and allowing smaller bioreactors to be used.

The location of the media can significantly impact its effectiveness. Early experience with highly loaded suspended growth systems indicated that the media could be placed uniformly throughout the bioreactor.[29,30] However, subsequent experience with more lowly loaded processes accomplishing nitrogen removal suggested that media placement is a critical factor.[50,53] For these systems, a mixed heterotrophic and autotrophic biofilm must develop because heterotrophic growth helps nitrifiers attach to the media. Consequently, the media should be placed in a location where the concentration of biodegradable organic matter has been lowered, but not completely removed, and where ammonia-N is still available. Experience also suggests that simultaneous nitrification and denitrification can occur inside the biofilm of some media.

21.1.7 Other Process Options

Many other attached growth bioreactor options have been evaluated and some are in commercial development. Some consist of filtration systems that are operated as bioreactors, such as when methanol is fed for denitrification to an upflow moving bed filter.[20,27] Others involve media used in other attached growth systems, such as the submerged biofilter that uses trickling filter media (either rock or plastic) in a liquid filled bioreactor.[16] Another example is the Surfpac process in which RBCs are added to activated sludge bioreactors.[68] A wide variety of moving bed bioreactors using a number of carrier media have also been considered.[28] The air-lift bioreactor is an FBBR with excellent mass and oxygen transfer.[59]

Much work is ongoing in this interesting and important area and further significant developments are likely to occur. The interested reader is referred to the literature for ongoing developments.

21.1.8 Comparison of Process Options

Table 21.3 summarizes the benefits and drawbacks of the various SAGBs. Both upflow and downflow packed bed bioreactors accomplish biological treatment and filtration, which can be advantageous

TABLE 21.3
Comparison of Submerged Attached Growth Processes

Process	Benefits	Drawbacks
Downflow packed bed (DFPB)	• Efficient biological oxidation • Efficient oxygen transfer • Filtration capability • Denitrification possible • Separate liquid-solids separation not required • Process design basis well established • Significant full-scale experience	• Filtration less efficient than in UFPB • More complex backwash than UFPB • Less efficient mass transfer than FBBR • Lower volumetric loadings than FBBR • Uses conventional equipment (filtration), but in unconventional mode.
Upflow packed bed (UFPB)	• Efficient biological oxidation • Efficient oxygen transfer • Filtration capability • Filtration more efficient than DFPB • Backwash simpler than DFPB • Denitrification possible • Separate liquid-solids separation not required • Process design basis well established • Significant full-scale experience	• Less efficient mass transfer than FBBR • Lower volumetric loadings than FBBR • Uses conventional equipment (filtration), but in unconventional mode.
Fluidized bed (FBBR)	• Efficient biological oxidation • Efficient mass transfer • Highest volumetric loading rates • Denitrification possible • Anaerobic operation possible • Separate liquid-solids separation not required • Process design basis well established • Successful full-scale installations	• Complex oxygen transfer system required for aerobic applications. • Essentially no filtration capability • Poor removal of particulate substrates • Long-term operating history somewhat limited
Moving bed bioreactor (MBBR)	• Efficient biological oxidation • Denitrification possible • Uses conventional wastewater treatment equipment • A variety of liquid-solids separation approaches can be used	• Requires separate liquid-solids separation step • Flocculation of influent particulate matter may be poor • No filtration capability • Volumetric loadings higher than purely suspended growth systems but lower than other attached growth systems
Integrated fixed film activated sludge (IFAS)	• Efficient biological oxidation • Can provide stable nitrification, even with limited suspended growth SRT • Denitrification possible • Uses conventional wastewater treatment equipment • Presence of suspended biomass provides excellent removal of colloidal and particulate substrate	• Requires separate liquid-solids separation step • No filtration capability • Volumetric loadings higher than purely suspended growth systems but lower than other attached growth systems • Process design basis not well established • Limited full-scale application

when a stream containing biodegradable organic matter and suspended solids is being treated. Efficient oxygen transfer occurs because low pressure air flows through tortuous paths within the packed bed. Furthermore, combined carbon oxidation, nitrification, and denitrification are possible, and biological phosphorus removal has also been achieved, demonstrating the versatility of these bioreactors. Some differences exist between the upflow and downflow options. Manufacturers who produce both typically promote the upflow option due to superior operating and loading characteristics.[38,60]

Fluidized bed biological reactors provide superior mass transfer characteristics and improved reaction rates for the removal of soluble materials. Oxygenation is more problematic than for packed bed bioreactors because all oxygen must be transferred to the process influent flow, which may be difficult to do for high-strength wastewaters. Combined carbon oxidation, nitrification, and denitrification can be obtained, but this requires the use of bioreactors in series with recirculation from the downstream aerobic FBBR to the upstream anoxic FBBR. Anaerobic treatment can also be successfully accomplished. Influent suspended solids are not efficiently removed in FBBRs.

Both packed and fluidized bed bioreactors provide the benefits of short HRTs, resulting in small bioreactor sizes. This can be quite important when space is limited and compact bioreactors are needed to achieve treatment goals.

The MBBR process is a highly flexible system for accomplishing carbon oxidation; combined carbon oxidation and nitrification; combined carbon oxidation, nitrification, and denitrification; separate stage nitrification; and separate stage denitrification. A downstream liquid-solids separation system is needed to remove produced biomass, and a variety of options can be used. However, flocculation of influent particulate matter may be poor, leading to poor performance of the liquid-solids separation system unless a chemical coagulant is used. Although specialized media and media retention equipment are used, they are installed in conventional bioreactors that use conventional oxygen transfer and liquid-solids separation equipment. The use of conventional facilities and equipment facilitates implementation of the process. Work is ongoing to develop enhanced biological phosphorus removal capabilities with these systems.[17,33]

Integrated fixed film activated sludge systems incorporate many of the benefits of the activated sludge process along with the shorter HRTs possible with attached growth bioreactors. These benefits include flexibility, high efficiency, and the capability to achieve a high quality effluent. The shorter HRTs result in smaller bioreactor sizes than in the activated sludge process. The attached growth can significantly enhance the settling and thickening characteristics of the suspended biomass.[30] In addition, the slower growing nitrifiers may grow selectively on the media, thereby stabilizing nitrification.[26,48,51] A drawback of this technology is that separate clarifiers and solids recycle are required. Another drawback is that performance data and operating experience are limited, thereby restricting the range of applications in which it can be used. The number of installations is increasing, however, thereby increasing the relevant knowledge base.

21.1.9 Typical Applications

Industrial applications of FBBRs often focus on the removal of organic matter from industrial wastewaters or contaminated groundwater. They have also been used to pretreat wastewaters prior to treatment at an industrial facility or in a municipal wastewater treatment plant. Both aerobic and anaerobic pretreatment applications have been successful. Operational problems with early municipal carbon removal applications have restricted the use of FBBRs for that application.[56] However, they are an accepted technology for the denitrification of nitrified municipal wastewaters.[64,70] The use of GAC as the media is well demonstrated at both bench and pilot-scales, and several full-scale applications are now in place.[58] This modification combines the potential for adsorption of slowly biodegradable, recalcitrant, or inhibitory organic compounds with the benefits of FBBRs for treatment of biodegradable organic matter. In summary, after an extended development period, the use of FBBRs is now rapidly expanding.[57]

Submerged packed bed bioreactors have been used successfully in the United States for carbon oxidation, combined carbon oxidation and nitrification, separate state nitrification, and separate stage denitrification at municipal wastewater treatment plants.[64,70] They provide the added function of filtration, thereby producing an effluent that meets stringent water reuse requirements. Aerobic packed bed bioreactor technology is now well proven and widely used in Europe, and new developments are focusing on the use of this technology for nutrient removal. Over 500 full-scale applications exist[55] and more are being added. Several commercial embodiments of the process are currently available.

Relatively few full-scale IFAS applications currently exist, but there is significant interest in its use for upgrading existing suspended growth processes to increase capacity, improve performance, and achieve nutrient removal. Moving bed bioreactor technology has found significant application in Europe, where more than 300 installations exist[35] and interest in North America is growing.[51] Significant experience is expected to develop in the next few years.

21.2 FACTORS AFFECTING PERFORMANCE

Factors affecting the performance of submerged attached growth bioreactors are similar to those affecting the performance of other biological treatment processes. Those particularly important to these bioreactors include total organic loading, substrate flux and surface loading, hydraulic loading, SRT, and DO concentration.

21.2.1 Total Volumetric Loading

As with other attached growth bioreactors, the total volumetric loading is calculated as the mass of substrate applied to the bioreactor per unit time divided by the media volume. When the substrate is organic matter, the total volumetric loading is the TOL. Consequently, it is calculated with Equation 19.1.

The performance of packed and fluidized bed bioreactors has traditionally been characterized by graphical relationships between the TOL and the removal efficiency. Figures 21.7 through 21.9 provide examples of such relationships for packed bed bioreactors treating municipal wastewater to achieve a variety of treatment goals. Figure 21.7 illustrates the relationship between the TOL (based on total chemical oxygen demand, COD) and the effluent total COD concentration for 12 full-scale facilities and demonstrates the typical decrease in effluent quality (increased effluent

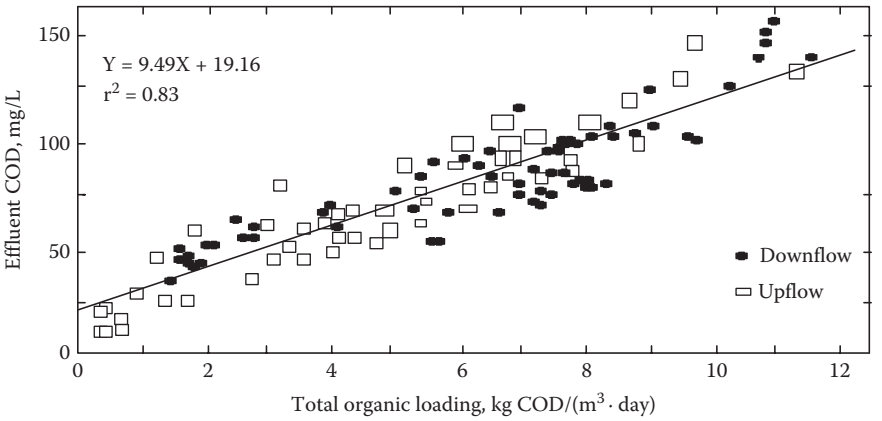

FIGURE 21.7 Effect of total organic loading (TOL) on effluent COD concentration. The data are from 12 full-scale DFPB and UFPB bioreactors containing fired clay media. (From Canler, J. P. and Perret, J. M., Biological aerated filter: Assessment of the process based on 12 sewage treatment plants. *Water Science and Technolo*gy, 29 (10/11): 13–22, 1994. Copyright © IWA Publishing. Reprinted with permission.)

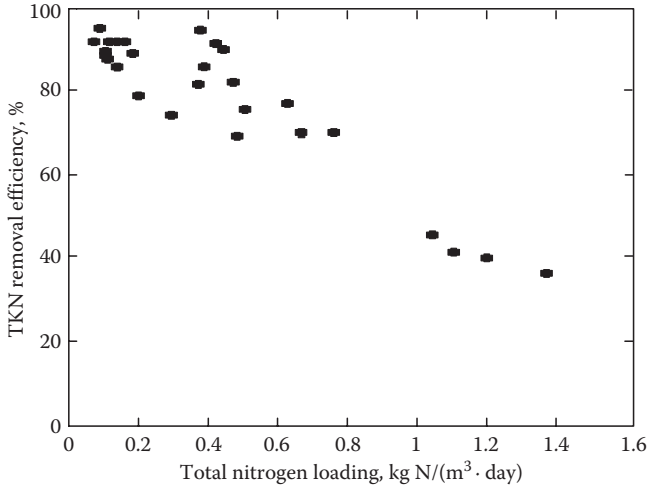

FIGURE 21.8 Effect of total nitrogen loading (TNL) on TKN removal efficiency during combined carbon oxidation and nitrification (COD/TKN ≈ 10). The data are from 12 full-scale DFPB and UFPB bioreactors containing fired clay media. (From Canler, J. P. and Perret, J. M., Biological aerated filter: Assessment of the process based on 12 sewage treatment plants. *Water Science and Technolo*gy, 29 (10/11): 13–22, 1994. Copyright © IWA Publishing. Reprinted with permission.)

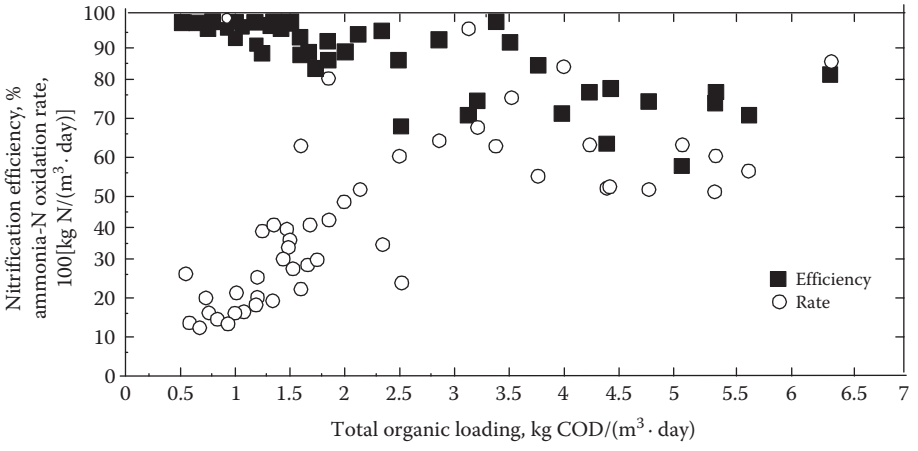

FIGURE 21.9 Effect of total organic loading (TOL) on nitrification efficiency and volumetric ammonia-N oxidation rate in packed bed bioreactors performing combined carbon oxidation and nitrification. (From Rogalla, R., Payraudeau, M., Bacquet, G., Bourbigot, M., Sibony, J., and Filles, P., Nitrification and phosphorus precipitation with biological aerated filters. *Research Journal, Water Pollution Control Federati*on, 62:169–76, 1990. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

total COD) observed with increasing TOL.[6] Performance could also be characterized by using the removal efficiency, where a decrease in efficiency would be observed as the TOL was increased, or by the volumetric removal rate expressed as kg/(m^3·day). In the last case, a linear increase in the volumetric removal rate would initially be observed as the TOL was increased, much as in Figure 20.4. With further increases in TOL, however, the rate of increase in the volumetric removal rate would decrease until a relatively constant plateau value was reached.

The total nitrogen loading (TNL) is calculated in the same manner as the TOL, except that the influent total Kjeldahl nitrogen (TKN) concentration is used instead of the influent COD or five-day biochemical oxygen demand (BOD_5) concentration:

$$\Lambda_N = \frac{F(S_{NHO} + S_{NSO} + X_{NSO})}{V_M}, \tag{21.1}$$

where Λ_N is the TNL and the sum of S_{NHO}, S_{NSO}, and X_{NSO} is the TKN. Figure 21.8 illustrates the relationship between the TNL and the TKN removal efficiency for the same 12 full-scale facilities considered in Figure 21.7.[6] In this instance, data from periods when the facilities were accomplishing combined carbon oxidation and nitrification are presented. Notice that the TKN removal efficiency is high at low values of the TNL, but begins to deteriorate as the TNL increases. This is exactly as expected when one considers that the TOL was high when the TNL was high. Figure 21.9[41] presents nitrification performance data for another facility accomplishing combined carbon oxidation and nitrification, but in this case ammonia-N removal is presented as a function of the TOL expressed on a COD basis. Both the ammonia-N removal efficiency and the ammonia-N volumetric oxidation (removal) rate are presented. Initially, the volumetric removal rate increases with increasing TOL because as the TOL is increased, so is the TNL. Thus, as long as the TOL is low enough to allow nitrifiers to compete for space in the biofilms, more nitrification will occur as more nitrogen is added. Ultimately, however, as the TOL is increased further the ammonia-N oxidation rate plateaus and declines as the nitrifiers have more difficulty existing in the system. The ammonia-N removal efficiency parallels the trends in volumetric removal rate. Note that, for a combined carbon oxidation and nitrification application using a particular wastewater, the TOL and TNL are directly related by the COD/TKN ratio of the influent wastewater. Consequently, only one of the two loading rates needs to be specified to define the performance of the bioreactor.

For tertiary nitrification applications, in which the TOL is low enough to allow full growth of nitrifiers in the biofilm and most nitrogen is in the form of ammonia-N, nitrification performance can be correlated with the ammonia-N loading, TAL:

$$\Lambda_{NH} = \frac{F \cdot S_{NHO}}{V_M}, \tag{21.2}$$

where Λ_{NH} is the TAL. Figure 21.10[38] illustrates the relationship between the TAL and the volumetric ammonia-N oxidation rate for 10 full-scale packed bed nitrification applications. Most were tertiary nitrification applications. The range of TAL values applied in this instance was relatively limited. As a consequence, a linear increase in the volumetric oxidation rate with an increase in the TAL was observed. If a broader range of TALs had been used, the volumetric oxidation rate would have reached a plateau, just as the rate of nitrification did in RBCs (see Figure 20.5).

The relationship between loading rate and process performance will vary with both the application and the bioreactor type. For example, Pujol et al.[38] indicate the following volumetric loading limitations for UFPB bioreactors using expanded clay media: organic matter removal, 10 kg COD/(m^3·day); nitrification, 1.5 kg NH_3-N/(m^3·day); and denitrification using methanol: >4 kg NO_3-N/(m^3·day). These are maximum loading rates, and lower values may be required depending on effluent requirements. By comparison, a loading rate of 6.4 kg NO_3-N/(m^3·day) or greater can be achieved for denitrification by FBBRs using sand media.[64,70] Anaerobic FBBRs (and EBBRs) are typically sized based on the TOL.[15,21] However, Ratcliff and Heath[39] demonstrated that the TAL can vary by a factor of 4 between various upflow SAGBs,[39] emphasizing the importance of understanding the performance characteristics of the specific SAGB being considered.

The TOL concept can be applied to MBBR systems, but experience demonstrates that the surface loading, as described in Section 21.2.2 is more appropriate. Likewise, the TOL can be applied to

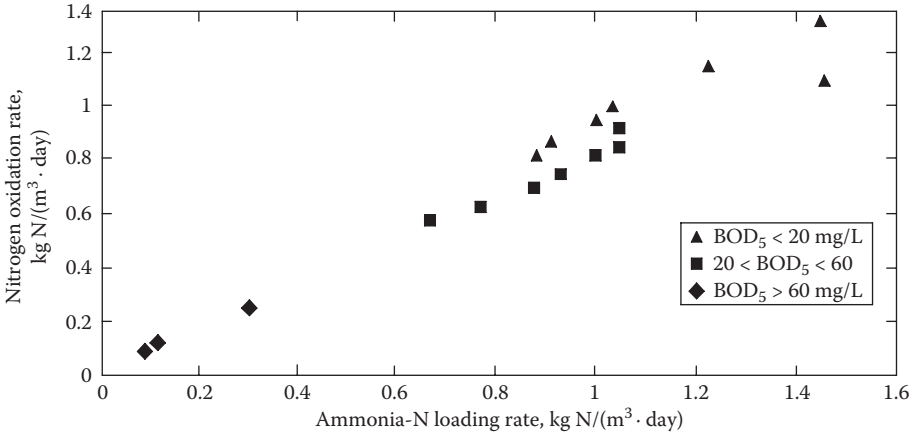

FIGURE 21.10 Effect of ammonia-N loading (TAL) on the volumetric ammonia-N oxidation rate in packed bed bioreactors. The data are from 10 full-scale nitrification plants. (From Pujol, R., Hamon, M., Kandel, X., and Lemmel, H., Biofilters: Flexible, reliable biological reactors. *Water Science and Technolog*y, 29 (10/11): 33–38, 1994. Copyright © IWA Publishing. Reprinted with permission.)

IFAS bioreactors, but it is less useful than for bioreactors containing only attached growth. This is because some means is required to equate the attached and suspended biomass within an IFAS bioreactor. Side-by-side comparisons between bioreactors with and without attached growth and/or with different types of media can be made based on the relative TOLs that can be achieved. However, the actual performance is a function of the relative amounts of attached and suspended biomass. When the total biomass concentration and its activity can be quantified, approaches typically applied to suspended growth bioreactors, such as the SRT, may be more useful process loading measures.

21.2.2 Substrate Flux and Surface Loading

Substrate flux and surface loading also influence the performance of a SAGB. Their relevance for characterizing the performance of attached growth bioreactors is discussed in Section 19.2.1, and the relationship between SOL and TOL is presented by Equation 19.2. Similar arguments and expressions apply for the fluxes and surface loadings of total nitrogen (SNL) and ammonia-N (SAL) in nitrifying SAGBs. One of the principal values of fluxes is that they allow comparisons to be made between different types of attached growth bioreactors and media. An example of the use of the ammonia-N flux for this purpose is provided by Figure 21.11 where fluxes for three packed bed bioreactors using three types of media are compared as a function of the residual ammonia-N concentration.[2] Similar fluxes were obtained in two bioreactors but substantially lower fluxes were observed with the other. Just as with trickling filters (as discussed in Section 19.2.1), these differences likely reflect differences in the hydrodynamic efficiency of the particular reactors considered.

Surface organic loading is more appropriate than TOL for some SAGBs, as illustrated in Figure 21.12.[34] The filtered COD removal rate (COD passing a 1.2 μm filter), expressed as g SCOD/(m^2·day), for MBBRs containing three types of media, correlated well with the SOL, expressed as g SCOD/(m^2·day), regardless of the media used. A maximum removal rate of approximately 27.5 g SCOD/(m^2·day) was reached at SOL values greater than about 60 g SCOD/(m^2·day), which corresponded to a SCOD concentration in the reactor greater than 200 mg/L (g/m^3). A residual DO concentration of 6.2 to 6.4 mg/L was maintained. The maximum removal rate in Figure 21.12 is similar to the maximum allowable loading rate on the first stage of an RBC (20 to 35 g COD/(m^2·day)), as discussed in Section 20.2. Furthermore, the shape of the curve in Figure 21.12b is similar to that for organic matter removal in RBCs (Figure 20.4).

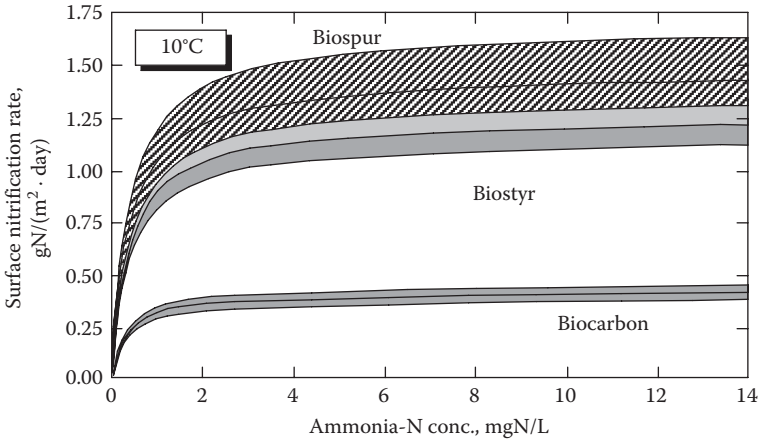

FIGURE 21.11 Effect of ammonia-N concentration on the surface nitrification rate (ammonia-N biofilm flux) in three types of SAGBs accomplishing tertiary nitrification. The range shown is for aeration rates of 15, 20, and 25 m/hr. (From Boller, M., Gujer, W., and Tschui, M., Parameters affecting nitrifying biofilm reactors. *Water Science and Technolo*gy, 29 (10/11): 1–11, 1994. Copyright © IWA Publishing. Reprinted with permission.)

It can be seen in Figure 21.13 that a maximum nitrification rate of about 1.5 g NH_3-N/(m^2·day) was achieved in a full-scale MBBR plant. This is similar to that achieved for separate stage nitrification in RBCs, as illustrated in Figure 20.5, and in trickling filters, as illustrated in Example 19.3.2.4. A more complete analysis of nitrification in MBBRs is presented by Hem et al.[18]

21.2.3 Total Hydraulic Loading

The total hydraulic loading (THL) for a SAGB, also referred to as the superficial velocity, is calculated in exactly the same manner as the THL for a trickling filter, as given by Equation 17.12, or the superficial velocity of an FBBR, as given by Equation 18.24. However, the THL affects the operation and performance of different SAGBs in different ways. For packed bed bioreactors it represents a constraint since an increase in the THL causes increased headloss, more frequent backwashing, and a larger total volume of backwash water. The specific upper limit depends on the media type, the influent suspended solids concentration, and the desired degree of treatment. For example, Pujol et al.[38] have suggested the following maximum THLs for UFPB bioreactors using expanded clay media: organic matter removal, 6 m/hr; tertiary nitrification, 10 m/hr; and denitrification using methanol, 14 m/hr. In contrast, for FBBRs a minimum upflow superficial velocity must be maintained to achieve the necessary fluidization of the media, as discussed in Section 18.2.1. This value is generally on the order of 35 m/hr for typical sand media. The THL is not a factor for MBBR and IFAS processes.

21.2.4 Solids Retention Time

It is possible to measure the SRT for some attached growth bioreactors, such as FBBRs using sand media. In such instances, the SRT can be used to quantify and control process performance, just as with a suspended growth process, although it must be recognized that the thickness of the biofilm will determine the effectiveness of the biomass as discussed in Section 18.2.3. For an FBBR, samples of bioparticles can be collected and composited to produce a representative sample of the entire bioreactor. The VSS concentration in this sample can then be measured and, from it, the total biomass inventory determined. The biomass wasted from the system in both the effluent and by operation of the growth control system can also be measured, and the SRT calculated in the same fashion as for a suspended growth system. When this is done, bioreactor performance can be correlated with the SRT, as illustrated in Figure 21.14.[56] If data on the effect of SRT on effluent

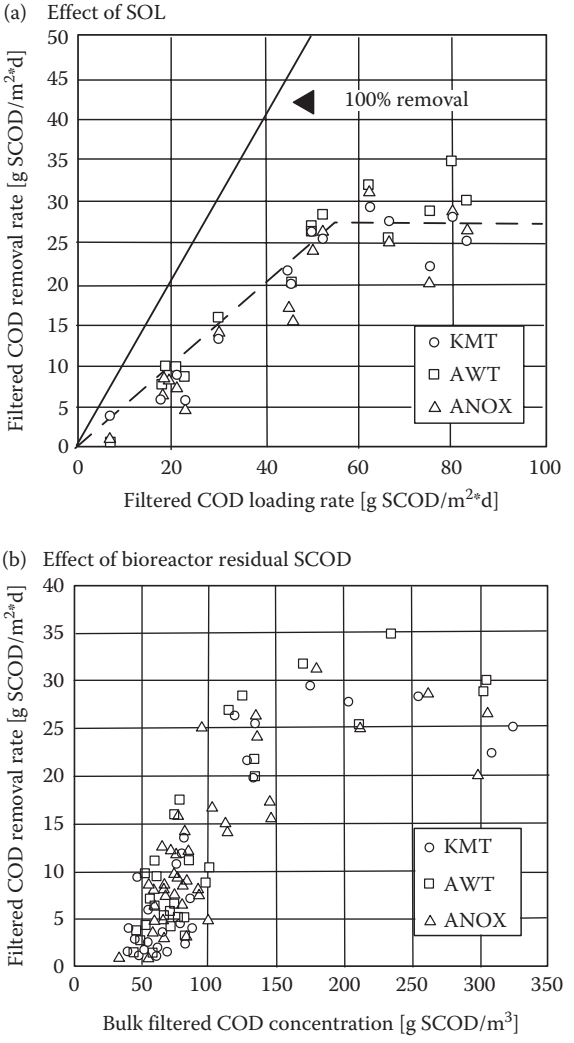

FIGURE 21.12 Surface removal rate [g SCOD/(m²·day)] for filtered COD (1.2 μm filter) for three types of free floating plastic media (KMT, AWT, and ANOX) with different specific surface areas. (From Ødegaard, H., Glsvold, B., and Strickland, J., The influence of carrier size and shape in the moving bed biofilm process. *Water Science and Technolo*gy, 41 (4/5): 383–91, 2000. Copyright © IWA Publishing. Reprinted with permission.)

quality can be collected for a SAGB, kinetic parameters can be determined for the particular wastewater and bioreactor configuration. However, it must be borne in mind that the parameters are not intrinsic, but implicitly incorporate the mass transfer characteristics for the particular system. Consequently, they cannot be extrapolated to other systems. Procedures for completing this analysis for FBBRs are conceptually the same as those used for suspended growth bioreactors discussed in Chapter 9.

The SRT can also be used to characterize the performance of IFAS processes. Two values of the SRT can be calculated, one based only on the suspended biomass (suspended growth SRT) and the other based on the suspended plus the attached biomass (total SRT). However, it must be recognized that the nature of the suspended and attached biomass may be different and, consequently, they may not be easily equated. Experience with IFAS systems indicates that the presence of the attached biomass allows them to achieve effective nitrification with suspended growth SRTs equal to or less than the minimum SRT for the growth of nitrifying bacteria.[12,30,51,53] The suspended biomass is

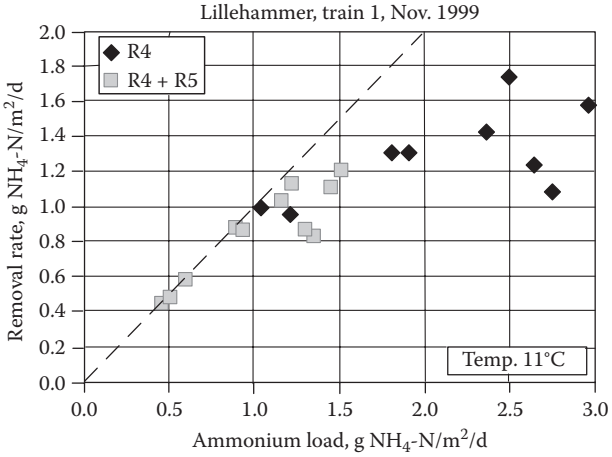

FIGURE 21.13 Effect of SAL on surface ammonia removal rate for free floating plastic media in the MBBR at Lillehammer, Norway. (From Ødegaard, H., Rusten, B., and Wessman, W., State of the art in Europe of the moving bed biofilm reactor (MBBR) proces*s, Proceedings of the Water Environment Federation 77th Annual Technical Exhibition and Confere*nce, October 2–6, 2004, New Orleans, LA, CD-ROM, Session 4, pp 348–54, Water Environment Federation, Alexandria, Virginia, 2004. Copyright © Water Environment Federation. Reprinted with permission.)

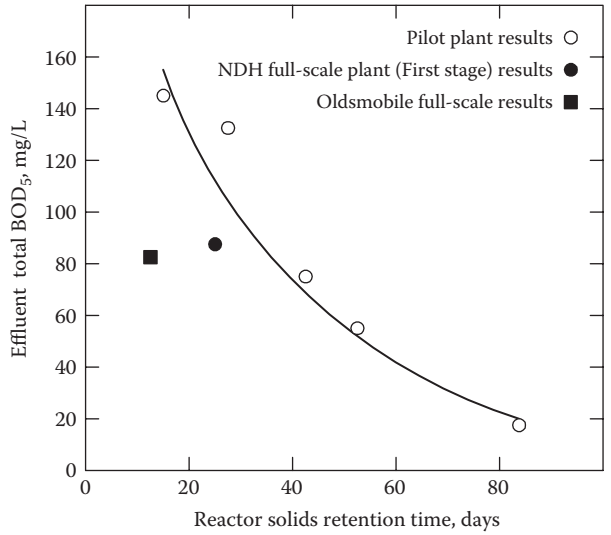

FIGURE 21.14 Effect of SRT on the total BOD_5 in the effluent from an FBBR treating automotive metal working wastewater. (From Sutton, P. M., Biological fluidized beds for water and wastewater treatment: A user's forum, *Proceedings of the seminar held at the Sheraton University Inn*, Ann Arbor, Michigan, P. M. Sutton & Associates, Inc, 1990. Reprinted with permission.)

predominantly heterotrophic and maintains a low bulk liquid phase organic substrate concentration. Since the suspended growth SRT is low, nitrifiers cannot grow in suspension, but they can form effective biofilms because the low organic substrate concentration allows them to compete effectively for space in the biofilm. Consequently, the biofilm provides the nitrifiers required to achieve process stability and to allow the system to successfully process peak ammonia-N loadings. Experience also indicates that significant denitrification can occur in the biofilms of some systems, depending on the thickness of the biofilm that develops.

21.2.5 Hydraulic Residence Time

In general, the performance of an attached growth process is not affected by the HRT, as long as the TOL and THL are satisfactory. This is because the amount of biomass in the system is determined by the surface area of media provided, the TOL, and other operating factors, not the length of time that the flow remains in the bioreactor. However, if the HRT exceeds the minimum SRT for biomass growth on the substrate provided, biomass will grow in suspension, which is undesirable in packed beds, FBBRs, and MBBRs because the suspended growth will pass into the effluent, increasing its organic matter content and reducing effluent quality. The growth of suspended biomass will also interfere with biofilm development because it reduces the amount of substrate available to the attached biomass. This can be particularly troublesome during start-up. This is because low organic loading rates may be needed initially to avoid process overloading as the biofilm develops. If the low TOL is achieved by applying only a portion of the wastewater flow, the resulting HRT may allow significant suspended growth to develop, hampering biofilm development and slowing start-up. In such cases, the wastewater should be diluted to allow operation at a reduced TOL while maintaining a short HRT to avoid the growth of suspended biomass.

21.2.6 Dissolved Oxygen Concentration

The DO concentration in aerobic SAGBs must generally be maintained at higher concentrations than in comparable suspended growth processes to achieve the needed oxygen mass transfer rate into the biofilm. In SAGBs oxygen must first be transferred into the bulk fluid by the oxygen transfer system before it can be transferred into the biofilm. The effect of DO concentration on the volumetric nitrification rate is illustrated in Figure 21.15 for nitrification in free floating plastic media.[44] Figure 21.16a further illustrates this for the same media and demonstrates the interacting effects of DO concentration and organic matter SOL on surface nitrification rates.[18,35] These results suggest that the DO concentration can be controlled to save energy by operating at lower DO concentrations during periods of lower loading when maximum biofilm removal rates are not needed, and increasing the DO concentration only during higher loading periods when high removal rates are needed. Dissolved oxygen concentrations can be increased by either increasing the aeration rate or by enriching the feed gas with pure oxygen.

It is important to recognize that the conditions inside the biofilm will be quite different from those in the bulk fluid. The concentrations of electron donors and electron acceptors will change

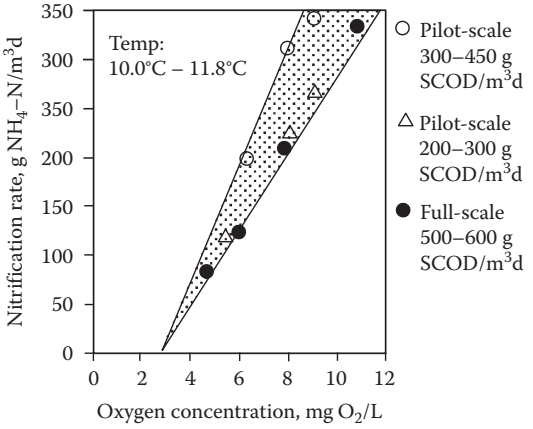

FIGURE 21.15 Effect of DO concentration on the volumetric nitrification rate for free floating plastic media. (From Rusten, B., Siljudalen, J. G., and Nordeidet, B., Upgrading to nitrogen removal with the KMT moving bed biofilm process. *Water Science and Technolog*y, 29 (12): 185–95, 1994. Copyright © IWA Publishing. Reprinted with permission.)

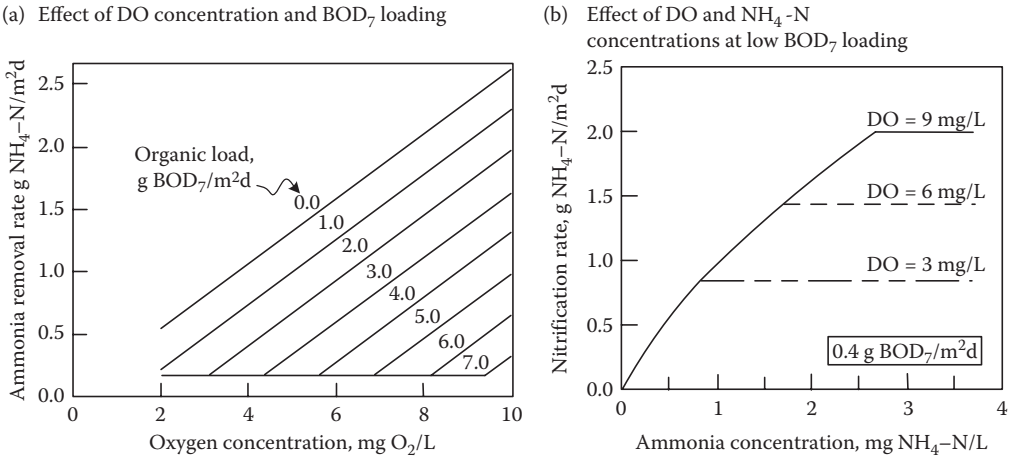

FIGURE 21.16 Effect of DO concentration on the surface nitrification rate for free floating plastic media accomplishing both combined carbon oxidation and nitrification and separate stage nitrification. (From Ødegaard, H., Rusten, B., and Wessman, F., State of the art in Europe of the moving bed biofilm reactor (MBBR) process, *Proceedings of the Water Environment Federation 77th Annual Technical Exhibition and Confere*nce, October 2– 6, 2004, New Orleans, LA, CD-ROM, Session 4, pp 348– 54, Water Environment Federation, Alexandria, Virginia, 2004. Copyright © Water Environment Federation. Reprinted with permission.)

with depth, as illustrated in Figure 16.19. This means that denitrification can occur in the biofilm even though DO may be present in the bulk liquid. The only way to fully analyze these effects is through the use of generalized models, such as those discussed in Part IV, but such models are not yet commonly used in design. Consequently, care must be exercised to ensure that the imposed operational conditions will achieve the desired result. This can be done most effectively with pilot studies and/or through relevant experience.

21.2.7 Other Factors

Like other biological processes, the performance of SAGBs is influenced by a variety of environmental conditions, such as temperature, pH, nutrient supply, and so on. As with trickling filters and RBCs, the importance of mass transport into the biofilms means that the impact of temperature is less significant than for suspended growth systems, until it is sufficiently low (~15°C) to make the biological reaction rates controlling. Relatively little information is available in the literature on the effects of temperature on substrate removal rates, but they can generally be assumed to be similar to those observed for RBCs (see Section 20.2.4). The pH must be in the physiological range, so sufficient buffering must be present in the influent wastewater to accommodate process alkalinity consumption (e.g., by nitrification) and production (e.g., by denitrification).

21.3 PROCESS DESIGN

21.3.1 General Design Procedures

The design of SAGBs is largely empirical in nature. Loading versus performance correlations, such as those illustrated in Figures 21.7 through 21.9 and Figure 21.12 are used to select a volumetric loading (TOL, Equation 19.1; TNL, Equation 21.1; or TAL, Equation 21.2) or a surface loading (SOL, SNL, or SAL), thereby providing the required bioreactor media volume. The basic relationship between volumetric loading and surface loading is given by Equation 19.2. The bioreactor configuration is then selected based on minimum and/or maximum hydraulic loading rates and typical media bed

TABLE 21.4
Summary of Submerged Attached Growth Reactor Design Procedure

1. Summarize process design and loading conditions, including maximum, minimum, and average sustained temperature; maximum, minimum, average, and design process influent flows and pollutant loadings; and desired effluent quality.
2. Where necessary, convert process loadings, expressed in conventional parameters (such as BOD_5) into the units used in the process design (such as biodegradable COD).
3. Select an effluent quality goal. As discussed in Section 10.4.4, selection of that goal should consider uncertainty and variability in process performance.
4. Select a design approach appropriate for the selected effluent quality goal (roughing, carbon oxidation, combined carbon oxidation and nitrification, separate stage nitrification), and reactor type.
5. Use the selected design approach to calculate the required media volume.
6. For packed and fluidized beds, select media depth to maintain acceptable minimum and maximum THLs based on influent flow. For fluidized beds, include recirculation if necessary.
7. Select reactor configuration. For MBBR and IFAS systems this involves selection of the number of reactor stages and their relative sizes.
8. Calculate process oxygen requirements and use to size the oxygen transfer system. Consider oxygen requirements for both carbon oxidation and nitrification, along with impacts of denitrification, as appropriate.
9. Calculate the waste sludge mass.
10. If separate stage denitrification is to be accomplished, calculate the supplemental carbon requirement.
11. Summarize the results in tabular form.

depths. This procedure is similar to that used in the design of trickling filters (see Section 19.3). For packed bed bioreactors one constraint is the maximum THL since excessive headloss will occur if the bioreactor is operated at higher THLs. This sets the minimum cross-sectional area and the maximum media depth. For FBBRs a minimum THL must be maintained to fluidize the bed while a maximum THL must not be exceeded to prevent media loss. When coupled with recirculation requirements, these set the maximum and minimum cross-sectional areas, respectively. Procedures for determining the minimum and maximum THLs for FBBRs are presented in Section 18.2. Moving bed biological reactors are typically sized based on the surface loading. For IFAS systems, the relative amounts of suspended and attached growth must be determined. This determination is generally based on experience with the specific system and application. Volume requirements are then determined in the same manner as for an activated sludge system, as discussed in Section 11.3. The SAGB design procedure is summarized in Table 21.4 and described further below.

Excess biomass production rates and heterotrophic oxygen requirements are calculated based on net process yields, $Y_{n,T}$, and process oxygen stoichiometric coefficients, Y_{O_2}, just as is done in Equations 10.3 and 10.4 for the preliminary design of activated sludge systems. However, in some cases significant organic matter will be in the process effluent, requiring those equations to be corrected for its presence. In addition, if nitrification is occurring, the oxygen requirement for nitrification, RO_A, is calculated with a modified form of Equation 11.16. Since the observed yield of nitrifiers is quite low, the right term within the brackets of that equation is typically ignored, allowing it to be simplified to

$$RO_A = 4.57F(S_{N,a} - S_{NH}). \tag{21.3}$$

The concentration of nitrogen available to the nitrifiers, $S_{N,a}$, is calculated with Equation 11.17. Use of that equation requires knowledge of the heterotrophic nitrogen requirement, NR, which for activated sludge systems is calculated with Equation 5.47. However, in this case, only the net process yield, $Y_{n,T}$, is known so NR is estimated with

$$NR = 0.10Y_{n,T}. \tag{21.4}$$

Some denitrification applications require the addition of an electron donor such as methanol. For these systems, the concentration of methanol, S_{MeOH}, that must be added to remove the total amount of electron acceptor present is calculated from basic stoichiometry:

$$S_{MeOH} = 2.47S_{NOO} + 1.53S_{NO_2O} + 0.87S_{OO}, \tag{21.5}$$

where S_{NOO} is the influent nitrate-N concentration in mg/L, S_{NO_2O} is the influent nitrite-N concentration in mg/L, and S_{OO} is the influent DO concentration in mg/L. The stoichiometric coefficients in Equation 21.5 are based on the net biomass yield for growth on methanol as the electron donor and carbon source. They can be calculated using the procedures presented in Chapters 3, 5, and 6. When exact information on the nature of the influent is not available, a methanol dose of 3 mg MeOH/mg NO_x-N is typically used, where NO_x-N represents the sum of the influent nitrate-N and nitrite-N concentrations. The design of DFPB bioreactors must also accommodate frequent air purge cycles to release nitrogen gas that accumulates in the bioreactor. Procedures for accomplishing these computations are presented elsewhere.[64]

Bioreactor geometry is an important factor for several of the SAGBs, and the resulting physical constraints must be considered during design. For example, packed and fluidized bed bioreactors require even flow distribution over the entire cross-sectional area, which is accomplished by nozzles located in either the influent or effluent regions. The physical constraints in accomplishing this distribution are very much like those involved in the design of a granular media filter. One result is a maximum allowable cross-sectional area corresponding to the area over which adequate flow distribution can be achieved. The depth of an aerobic packed bed bioreactor must be adequate to allow the aeration system to meet the oxygen requirements. Furthermore, the depths of both packed and fluidized bed bioreactors must also be sufficient to produce the required effluent quality, which they should do if the loading was appropriately chosen. Ongoing developments by system developers and suppliers are refining these restrictions.[66] Bioreactor geometry is generally less restrictive for the MBBR and IFAS systems.

21.3.2 Packed Bed Bioreactors

The design of packed bed bioreactors is based on the selection of an appropriate TOL and bioreactor configuration consistent with constraints on the maximum THL that can be applied to the bioreactor to avoid excessive headloss and backwash recycle volumes. These principles are illustrated in the following example where the technology is applied for a carbon removal application.

Example 21.3.2.1

Size a UFPB bioreactor using fired clay media to treat a wastewater with a flow rate of 10,000 m^3/day. The wastewater has been pretreated by chemically enhanced primary treatment, removing all suspended solids and leaving a COD concentration of 240 mg/L. The treatment objective is partial removal of organic matter without nitrification. For this application, Pujol et al.[38] recommend a maximum THL of 6 m/hr for the media to be used.

a. What TOL should be used?
   Figures 21.7 and 21.9, and the recommendations of Pujol et al.[38] in Section 21.2.1, suggest that a TOL of 10 kg COD/(m^3·day) will allow significant COD removal with minimal nitrification. From Figure 21.7, the effluent COD will be approximately 115 mg/L.

b. What bioreactor volume (media volume) is required?
The media volume can be calculated from the definition of TOL, given by Equation 19.1, after noting that a TOL of 10 kg COD/(m^3·day) is equivalent to 10,000 g COD/(m^3·day):

$$V_M = \frac{(10{,}000)(240)}{10{,}000} = 240 \text{ m}^3.$$

c. Will the THL be acceptable if the bioreactor bed depth is 2 m?
For a 2 m depth, the bed cross-sectional area will be

$$A_c = \frac{240}{2} = 120 \text{ m}^2.$$

The THL is given by Equation 17.12. Recirculation is not typically employed with a UFPB bioreactor, so:

$$\Lambda_H = \frac{10{,}000}{120} = 83.3 \text{ m/day} = 3.5 \text{ m/hr}.$$

This is less than 6 m/hr, so it is acceptable.

d. How much oxygen must be provided?
The oxygen requirement can be estimated with Equation 10.4. For a high-rate application such as this, synthesis oxygen requirements would be exerted but little decay will occur. Therefore, assume a value of 0.5 mg O_2/mg COD removed for Y_{O_2}. Because the effluent COD is approximately 115 mg/L, which is a significant fraction of the influent COD, its presence must be accounted for. Therefore:

$$RO = (10{,}000)(0.5)(240 - 115) = 625{,}000 \text{ g/day} = 625 \text{ kg } O_2\text{/day}.$$

e. What will the excess biomass production rate be?
From a COD balance, if Y_{O_2} has a value of 0.5 g O_2/g COD removed, $Y_{n,T} \cdot i_{O/XB,T}$ must have a value of 0.5 g biomass COD/g COD removed, giving $Y_{n,T}$ a value of 0.42 g biomass total suspended solids (TSS)/g COD removed if $i_{O/XB,T}$ has a value of 1.2 mg COD/mg TSS. Therefore, from Equation 10.3 the excess biomass production rate, W_M, will be

$$W_M = (10{,}000)(0.42)(240 - 115) = 525{,}000$$
$$= 525 \text{ kg biomass TSS/day}.$$

Note that a high TOL is required to avoid nitrification for this application. This high TOL limits the peak application rate to the bioreactor because substrate is already being removed rapidly. Consequently, if the TOL is increased transiently during peak loadings, the effluent substrate concentration will increase proportionally, as indicated in Figure 21.7. In addition, peak loadings to this type of bioreactor will also be limited by the maximum allowable THL of 6 m/hr. Since the design THL was 3.5 m/hr, the maximum flow that the system could handle from a hydraulic perspective would be 6.0/3.5 = 1.71 times the average flow. Thus, it can be seen that designing a system with a high TOL and a high THL limits its ability to handle peak loads.

A lower TOL is required to reliably accomplish nitrification. The following example illustrates the use of a UFPB bioreactor for such an application.

Example 21.3.2.2

Repeat Example 21.3.2.1, but size the bioreactor for nearly complete nitrification. The TKN concentration of the pretreated wastewater is 35 mg/L.

a. What TOL is required in this application?
   Examination of Figure 21.9 reveals that the volumetric ammonia-N oxidation rate increases linearly with the TOL up to TOLs of about 3 to 3.5 kg COD/(m^3·day). The nitrification efficiency also remains above 80% up to a TOL of about 2.5 kg COD/(m^3·day). Consequently, select a TOL of 2.5 kg COD/(m^3·day), giving a nitrification rate of 0.6 kg NH_3-N/(m^3·day). From Figure 21.7, this TOL will produce an effluent COD of approximately 40 mg/L.
b. What bioreactor volume (media volume) is required?
   The media volume can be calculated from the definition of TOL, given by Equation 19.1, after noting that a TOL of 2.5 kg COD/(m^3·day) is equivalent to 2500 g COD/(m^3·day):

$$V_M = \frac{(10{,}000)(240)}{2500} = 960 \text{ m}^3.$$

c. What media depth would be required if a THL of 1.5 m/hr were chosen for this application?
   The bed cross-sectional area can be calculated from the definition of THL, as given by Equation 17.12, after noting that 1.5 m/hr is equivalent to 36 m/day:

$$A_c = \frac{10{,}000}{36} = 278 \text{ m}^2.$$

   The media depth is just the volume divided by the cross-sectional area:

$$L = \frac{960}{278} = 3.45 \text{ m}.$$

   This is an acceptable depth and will help ensure adequate oxygen transfer.
d. What is the process oxygen requirement?
   For a low rate process such as this, the process oxygen stoichiometric coefficient will be much greater than in Example 21.3.2.1 because more biomass decay will occur. Thus, to be conservative, use a value of 0.7 mg O_2/mg COD removed for Y_{O_2}. The heterotrophic oxygen requirement can be calculated with Equation 10.4. Because the effluent COD is approximately 40 mg/L, which is a significant fraction of the influent COD, its presence must be accounted for. Therefore:

$$RO_H = (10{,}000)(0.7)(240 - 40) = 1{,}400{,}000 \text{ g/day}$$
$$= 1400 \text{ kg } O_2\text{/day}.$$

   To this we must add the oxygen required for nitrification. This can be estimated with Equation 21.3, which requires estimation of $S_{N,a}$, the nitrogen available to the nitrifiers, and S_{NH}, the effluent ammonia-N concentration. The nitrogen available to the nitrifiers can be calculated with Equation 11.17, with the use of Equation 21.4 for NR. To use Equation 21.4 we must know $Y_{n,T}$. From a COD balance, since Y_{O_2} is 0.7 g O_2/g COD removed and $i_{O/XB,T}$

is 1.2 mg COD/mg TSS, $Y_{n,T}$ must be 0.25 g biomass TSS/g COD removed. The TKN in the influent is 35 mg/L. Therefore, from Equation 11.17:

$$S_{N,a} = 35 - (0.10)(0.25)(240 - 40) = 30.0 \text{ mg/L}.$$

The effluent ammonia-N concentration can be estimated from a mass balance on nitrogen using the media volume and the volumetric nitrification rate of 0.6 kg NH_3-N/(m^3·day):

$$\text{Input rate} = (10{,}000)(30) = 300{,}000 \text{ g/day}$$

$$\text{Removal rate} = (0.6)(960) = 576 \text{ kg/day} = 576{,}000 \text{ g/day}$$

Since the removal rate exceeds the application rate, nitrification will occur until the ammonia-N concentration becomes rate limiting. Thus, the requirement for almost complete nitrification should be met. To be conservative in calculating the autotrophic oxygen requirement, assume that the effluent ammonia-N concentration is zero. Then the autotrophic oxygen requirement can be calculated with Equation 21.3:

$$RO_A = (4.57)(10{,}000)(30) = 1{,}370{,}000 \text{ g/day} = 1370 \text{ kg/day}.$$

So, the total process oxygen requirement is

$$RO = 1400 + 1370 = 2770 \text{ kg } O_2\text{/day}.$$

e. What is the excess biomass production rate?
This can be calculated with Equation 10.3 using the $Y_{n,T}$ value of 0.25 g biomass TSS/g COD removed that was used above

$$W_M = (10{,}000)(0.25)(240 - 40) = 500{,}000 \text{ g TSS/day}$$

$$= 500 \text{ kg biomass TSS/day}.$$

This calculation assumes that the contribution of autotrophic biomass to the excess biomass production rate is small, as discussed in Section 6.3.2.

Note the much larger bioreactor volume required and the greater media depth for this application in comparison to the carbon oxidation application. A substantial peak flow could be processed through this bioreactor without experiencing excessive headloss and backwash recycle volumes because the THL is also lower than in the previous case. The oxygen requirement is increased significantly, not only due to the oxygen requirement for nitrification but because of the increased oxygen requirement for the oxidation of organic matter. Excess biomass production, on the other hand, is not significantly affected. This arises because of the higher degree of stabilization of the removed organic matter.

Designs that accomplish an intermediate level of organic matter removal are also possible. However, it must be recognized that they will also generally achieve at least partial nitrification.

Packed bed bioreactors can also be used for denitrification. The reader is referred to the U.S. EPA *Nitrogen Control Manual*[64] and other publications on this topic[9,10] for a complete discussion of the design of such systems.

21.3.3 Fluidized and Expanded Bed Biological Reactors

Although as presented in Section 18.5, procedures have been developed to optimize the size of an FBBR or EBBR, many engineers approach their design in a manner similar to that used to design packed bed bioreactors. The first step is to select a design TOL or SOL and use it to calculate the bioreactor bed volume. The primary difference between the design of FBBRs or EBBRs and packed bed bioreactors is that both minimum and maximum values of the THL must be maintained for FBBRs and EBBRs whereas only the maximum THL was constrained in the design of packed bed bioreactors. For FBBRs and EBBRs a minimum THL must be maintained to fluidize the bed. For FBBRs the maximum THL must be set to avoid bed washout, whereas for EBBRs it is set so as not to exceed a desired degree of bed expansion. Table 21.1 lists typical minimum (in the column labeled "Average") and maximum THLs for FBBRs using sand and GAC media. The values differ only by a factor of 1.5 for sand media, suggesting that they represent the range of reasonable operating conditions, rather than extreme values. Indeed, examination of Figure 18.4 shows that these THLs will give reasonable fluidization of typical particle sizes when they carry a biofilm of about 100 μm thickness. Since this thickness often maximizes the effective biomass concentration, typical operating conditions for FBBRs are close to optimum, thereby justifying the simplified design approach commonly used.

The need to maintain specified THLs requires a trade-off between the FBBR or EBBR bed depth and the recirculation ratio that is directly analogous to that for trickling filters, as discussed in Section 19.3. For a fixed bioreactor volume, the greater the bed depth, the smaller the cross-sectional area and the less the need for recirculation to maintain a desired THL. Unlike trickling filters, FBBR and EBBR bed depths are constrained primarily by economic considerations rather than by structural constraints. Bed depths in the range of 2 to 6 m are typical, although greater depths can be used. In some cases, maximum bed depths will be used to minimize recirculation. In other cases, reduced bed depths will be required to avoid excessive THLs. The following two examples illustrate these two different cases.

Example 21.3.3.1

An industrial wastewater with a flow rate of 7500 m^3/day and a nitrate-N concentration of 100 mg/L is to be denitrified in a FBBR. The wastewater is devoid of organic matter so methanol will be added as an electron donor. Pilot studies have suggested that a nitrate loading rate of 6 kg NO_3-N/(m^3·day) will allow complete denitrification to be achieved with methanol in a bed of 0.6 mm sand particles. The THL is to be kept between 24 and 36 m/hr to achieve appropriate fluidization of the media with an optimum biofilm thickness.

a. What fluidized media volume is required?
   The nitrate loading rate is equivalent to a TOL. Consequently, the fluidized media volume can be calculated with Equation 19.1 after recognizing that 6 kg NO_3-N/(m^3·day) is equivalent to 6000 g NO_3-N/(m^3·day):

$$V_M = \frac{(7500)(100)}{6000} = 125 \text{ m}^3.$$

b. Can a bed depth of 6 m be used?
   A bed depth of 6 m is permissible as long as it does not result in a THL greater than 36 m/hr without recirculation. Thus, the THL must be calculated with Equation 17.12 to check it. This requires knowledge of the cross-sectional area. For the 6 m depth:

$$A_c = \frac{125}{6} = 20.8 \text{ m}^2.$$

Therefore, the THL is

$$\Lambda_H = \frac{(7500)(1+0)}{20.8} = 360 \text{ m/day} = 15 \text{ m/hr.}$$

This is less than 36 m/hr, which suggests that a bed depth of 6 m is permissible. However, because the THL is less than the minimum THL (24 m/hr), recirculation will be required.

c. What recirculation flow rate is needed to reach the minimum THL?
A THL of 24 m/hr is equivalent to 576 m/day. Using this in a rearranged form of Equation 17.12 gives:

$$\alpha = \frac{(576)(20.8)}{7500} - 1 = 0.60.$$

Since the flow rate is 7500 m^3/day, the recirculation flow rate required to maintain the minimum THL is 4500 m^3/day.

d. How much methanol must be provided daily?
The required methanol concentration can be calculated with Equation 21.5:

$$S_{MeOH} = (2.47)(100) = 247 \text{ mg/L.}$$

The mass of methanol required daily is just the concentration times the flow rate or

$$\text{Methanol requirement} = (7500)(247) = 1{,}850{,}000 \text{ g/day} = 1850 \text{ kg/day.}$$

In the previous example recirculation was required even when the tallest feasible FBBR was used. This was due in part to the high concentration of nitrate-N that had to be removed. Treatment of a wastewater with a lower concentration will require a smaller total volume, which will result in a reduced cross-sectional area for a given height. The THL will increase as the cross-sectional area is reduced, and a point can be reached at which the THL exceeds the maximum allowable value. At this point the FBBR cross-sectional area reaches a minimum value and further reductions in bioreactor volume can be achieved only by reducing the bed depth. The following example illustrates the design of an FBBR where the bed depth must be less than the maximum value to keep the THL below the maximum value.

Example 21.3.3.2

Rework Example 21.3.3.1, but with an influent nitrate-N concentration of 25 mg/L.

a. What fluidized media volume is required?
As before, the fluidized media volume can be calculated with Equation 19.1 after recognizing that 6 kg NO_3-N/(m^3·day) is equivalent to 6000 g NO_3-N/(m^3·day):

$$V_M = \frac{(7500)(25)}{6000} = 31.25 \text{ m}^3.$$

b. Can a bed depth of 6 m be used?
A bed depth of 6 m is permissible as long as it does not result in a THL greater than 36 m/hr without recirculation. Thus, the THL must be calculated with Equation 17.12 to check it. This requires knowledge of the cross-sectional area. For the 6 m depth:

$$A_c = \frac{31.25}{6} = 5.21 \text{ m}^2.$$

Therefore, the THL is

$$\Lambda_H = \frac{(7500)(1+0)}{5.21} = 1440 \text{ m/day} = 60 \text{ m/hr.}$$

This exceeds the maximum allowable THL and thus an FBBR with a depth of 6 m cannot be used.

c. What bed depth is required if the THL is restricted to 36 m/hr without recirculation?
The FBBR cross-sectional area can be calculated with Equation 17.12 by using $\alpha = 0$ and recognizing that 36 m/hr = 864 m/day:

$$A_c = \frac{(7500)(1+0)}{864} = 8.68 \text{ m}^2.$$

The height can be calculated as

$$L = \frac{31.25}{8.68} = 3.6 \text{ m.}$$

The height exceeds 2 m and is acceptable. If it had been too small it would have been necessary to reduce the TOL to achieve an acceptable size FBBR.

Further details on the design of fluidized bed reactors are presented elsewhere.[56,64,70]

21.3.4 Moving Bed Biological Reactors

Moving bed biological reactors are generally sized using surface loadings (SOL, SNL, and SAL, as appropriate). The use of surface loadings to size fixed film bioreactors was illustrated for nitrifying trickling filters in Example 19.3.2.4 and for RBCs in Sections 20.3.2 and 20.3.3. Similar approaches are used with MBBRs. The only significant difference is that the DO can be adjusted based on the design of the oxygen transfer system, which affects the maximum surface loading for aerobic processes, as discussed in Section 21.2.6. Relationships between surface loadings and effluent quality, such as those presented in Figures 21.12, 21.13, and 21.16 can be determined from pilot studies and full-scale systems, and such data are accumulating for a variety of wastewaters. The literature, which is growing rapidly, should be consulted to determine current information. An analysis of issues associated with the design of MBBRs is available.[51]

21.3.5 Integrated Fixed Film Activated Sludge Systems

Procedures for the design of IFAS systems are evolving. Consequently, the reader is advised to consult the most recent literature for current information and procedures.[45,46,48,51,68,69] Because IFAS systems incorporate suspended growth, their design makes significant use of the suspended growth design procedures presented in Chapters 10, 11, and 12. The procedure can be viewed as designing a suspended growth process with consideration of the impacts of the attached growth. Three procedures are generally used to account for the impacts of attached growth: (1) equivalent biomass, (2) kinetic, and (3) mathematical modeling. For the equivalent biomass approach, the biomass accumulating as attached growth is determined and is assumed to contribute to the overall system inventory in the same fashion as the suspended biomass. Thus, the process design is essentially the same as for suspended growth systems, except that part of the biomass inventory is suspended and the remainder is attached. Information on equivalent biomass for various IFAS media is presented in Table 21.2 and elsewhere.[51] The kinetic approach also builds upon existing suspended growth

TABLE 21.5
Comparison of IFAS Design Approaches

Approach	Benefits	Drawbacks
Equivalent biomass	• Based on well-established and understood suspended growth design approaches. • Simple computations.	• Does not explicitly address differences in composition of the suspended and attached biomass. • Accurate results possible only with site-specific (full- or pilot-scale) data or results from comparable applications.
Kinetic	• Builds upon the well-established and understood suspended growth design approaches. • Empirically addresses differences in suspended and attached biomass composition.	• Requires relevant and/or site-specific data to account for suspended and attached biomass composition. • More complex computations.
Mathematical modeling	• Builds upon the well-established and understood suspended and attached growth design approaches. • Fundamentally addresses differences in suspended and attached biomass composition. • Can be applied with only modest site-specific data.	• More complex computations.

design procedures but explicitly reflects differences in biomass composition and reaction kinetics for the suspended and attached biomass.[45–47,49] Experience with the kinetic approach is growing and suggests that use of the approach can result in successful applications. Mathematical modeling incorporates the fundamental principles of suspended and attached biomass growth and is generally based on the International Water Association activated sludge and biofilm models.[19,67] Modeling offers the potential to calculate the distribution and activity of the suspended and attached biomass fractions and, consequently, can predict rather than just characterize performance (which is what the kinetic approach does). Initial development and application of the mathematical modeling approach is encouraging.[3,4,37] Both the kinetic and the mathematical modeling approaches are also applicable to MBBRs. While use of the equivalent biomass approach, when coupled with experience in selection of the attached biomass contribution, can result in an appropriate and effective design, it is essentially empirical in nature. It is expected that its use will be replaced with the kinetic approach and eventually with the mathematical modeling approach as knowledge and experience evolves. Table 21.5 contrasts these approaches.

Physical design considerations are important. For example, the design of the oxygen transfer system must be integrated with the placement of the attached growth media. The system must be configured to recirculate mixed liquor through fixed media so that DO, soluble organic matter, and ammonia-N are transported to the attached biomass. To date, diffused air systems in spiral roll configurations have been used to accomplish this, as illustrated in Figure 21.5. Baffles may be needed to ensure an appropriate quantity and distribution of mixed liquor flow through the media. The configuration of the oxygen transfer system can also affect the distribution of free floating media, such as free floating plastic media and polyurethane pads, as discussed in Section 21.1.6. The oxygen transfer system must be designed to meet the oxygen requirement for both the suspended and the attached biomass. Sufficient bioreactor volume must also be available to contain the added media. The velocity of mixed liquor approaching the media retention screens must be controlled to prevent the buildup of media adjacent to the screens, which can affect the configuration and, consequently, the hydraulic flow pattern through the media zone. Specific considerations and criteria are evolving and depend on the media type. As an example, however, it is suggested that the approach velocity for some free floating media be limited to no more than 10 to 15 m/hr. Guidelines on the implementation of IFAS systems are available.[24,25,52].

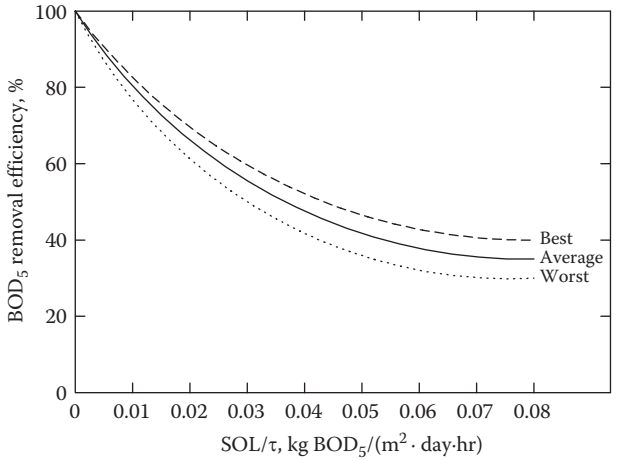

FIGURE 21.17 Effect of SOL per unit of residence time on the percentage of BOD_5 applied that is removed by attached biomass growing on vertical plate media in IFAS systems. (From U.S. Environmental Protection Agency, *Inert Biomass Support Structures in Aerated Suspended Growth Systems: An Innovative/Alternative Technology Assessment,* EPA/600/X-87/078a&b, U.S. Environmental Protection Agency, Cincinnati, OH, 1986.)

The following example illustrates the addition of fixed media to upgrade an overloaded activated sludge system. It illustrates the types of impacts that have been observed with these systems and the design approaches that have been suggested to deal with them. As discussed in Section 21.1.6, the attached biomass contributes both to the removal of organic matter and to the production of suspended biomass with improved settling characteristics. Consequently, the addition of fixed media results in an increase in total biomass in the system by two mechanisms: directly by the attached biomass and indirectly by allowing an increased MLSS concentration to be handled by the final settler. The impact of improved solids settling characteristics on treatment capacity is determined first. Then, the media volume required to treat the remainder of the design load is determined. Finally, the required media volume is compared to the bioreactor volume to ensure that it can be placed in the existing bioreactor. For this example, a relationship from the literature is used to characterize the contribution of the attached biomass to organic matter removal, as presented in Figure 21.17.[63]

Example 21.3.5.1

The activated sludge system in a municipal wastewater treatment plant receives a flow of 10,000 m^3/day containing a BOD_5 concentration of 150 mg/L. The bioreactor volume is 1000 m^3, giving an HRT of 2.4 hr. The system is overloaded and producing an effluent of poor quality. Furthermore, the overloading has resulted in an activated sludge with very poor settling characteristics, thereby limiting the MLSS concentration that can be reliably maintained to 1000 mg/L. This, in turn, has contributed to the problem because the low MLSS concentration results in a process loading factor (F/M ratio) of 1.5 kg BOD_5/(kg MLSS·day) and an SRT of about one day. Fixed media is to be installed in the activated sludge bioreactor to increase the biomass in the process and improve effluent quality. The plan is to use modular plastic trickling filter media with a specific surface area of 100 m^2/m^3. Experience indicates that this will allow a higher MLSS concentration to be maintained; a value of 3000 mg/L is anticipated. The target F/M ratio for suspended biomass to achieve improved solids settleability and effluent quality is 0.3 kg BOD_5/(kg MLSS·day). How much attached growth media will be required to accomplish the goal?

a. How much BOD_5 could the suspended biomass remove at the revised operating condition?
   At an MLSS concentration of 3000 mg/L (3000 g/m^3) the mass of suspended biomass in the system would be (1000)(3000) = 3,000,000 g = 3000 kg. If that suspended biomass were removing organic matter at a rate of 0.3 kg BOD_5/(kg MLSS·day), it could remove (0.3)(3000) = 900 kg BOD_5/day.
b. How much BOD_5 must the attached biomass remove?
   The attached biomass must remove everything not removed by the suspended biomass. The total BOD_5 loading on the process is

$$(10{,}000)(150) = 1{,}500{,}000 \text{ g/day} = 1500 \text{ kg } BOD_5\text{/day}.$$

   Therefore, the BOD_5 to be treated by the attached growth is

$$1500 - 900 = 600 \text{ kg } BOD_5\text{/day}.$$

c. How much attached growth media is required to remove this mass of BOD_5?
   Figure 21.17 presents performance curves for attached growth media in an activated sludge bioreactor. They represent the percentage of BOD_5 removal that can be attributed to the attached growth. Since the attached growth in the system under consideration must remove 600 kg/day of the 1500 kg BOD_5/day applied to the system, the removal efficiency for the attached growth must be 40%. Entering Figure 21.17, with this figure reveals that the SOL per unit of residence time must be 0.06 kg BOD_5/(m^2·day) per hour of aeration.
   For this application, the BOD_5 loading is 1500 kg BOD_5/day and the HRT is 2.4 hr. Thus, the required media surface area is

$$A_s = \frac{1500}{(0.06)(2.4)} = 10{,}400 \text{ m}^2.$$

   If trickling filter media with a density of 100 m^2/m^3 is used, this corresponds to 104 m^3 of media. This is about 10% of the total bioreactor volume, which is acceptable.

Research is ongoing to develop procedures for using IFAS systems to upgrade suspended growth bioreactors to achieve nitrification and denitrification. Current knowledge is incorporated into the design approaches discussed above and contrasted in Table 21.5. Preliminary results suggest that the presence of the attached biomass allows the system to be operated at a suspended biomass SRT that would, by itself, provide only partial nitrification. The presence of the attached biomass allows sufficient nitrifiers to grow to achieve nearly complete nitrification. Thus, the evolving design procedure will likely involve the selection of a suspended growth SRT to achieve partial nitrification and the selection of a fixed media surface area to maintain an adequate additional mass of nitrifiers to achieve complete nitrification. As discussed in Section 21.1.6, preliminary experience indicates that the fixed media must be placed in a location where soluble, biodegradable organic matter is present to allow heterotrophs to grow on the media as a flocculating agent. Ammonia-N must also be present to allow nitrifiers to grow. While current design procedures are largely empirical in nature, the mathematical modeling approach, which can incorporate the competition between heterotrophic and autotrophic biomass within both the suspended and attached biomass components, will be further developed in the future. With this more fundamental approach it will be possible to predict suspended and attached biomass composition, thereby placing the design of IFAS systems on a firmer basis. The reader is encouraged to consult both existing and evolving information on this topic.

21.3.6 General Design Experience

Experience with the application and physical design of SAGBs is accumulating rapidly. This experience indicates that a significant degree of pretreatment is necessary to remove debris and fibrous material from the wastewater prior to applying it to the bioreactor. These materials can clog the inlet or outlet structures or can become enmeshed in the media. The specific problems associated with each bioreactor configuration and media type differ slightly, but they all suggest the need for good preliminary treatment. For example, clogging of inlet or outlet structures is particularly an issue for packed and fluidized bed bioreactors. Media clogging and/or clogging of media retention screens, on the other hand, is a particular concern in MBBR and IFAS systems. In the latter case, fibrous material can form into long ropes that can become entangled with the media and/or its retention screens. Experience indicates that at least primary clarification should be provided for all SAGBs. Furthermore, some applications may require a higher level of pretreatment, such as chemically enhanced primary clarification or high-rate activated sludge. The use of screens in place of primary clarifiers should be regarded with caution due to their poor capture of fibrous materials. If they are used, at least two levels should be provided, with the second level containing relatively small openings.

Packed bed and fluidized bed bioreactors have different abilities for removing particulate organic matter. Packed bed bioreactors can remove these materials through filtration and subsequent biodegradation, whereas particulate organic material will generally pass through an FBBR.

Recent experience has revealed the potential for excessive growths of difficult to remove biomass in some packed bed bioreactor applications. This has occurred in applications with high loading rates, such as the one described in Example 21.3.2.1. Developments in this area should be followed carefully as this technology is applied.

21.4 PROCESS OPERATION

One of the frequently reported advantages of attached growth processes is that they are simpler to operate and require less attention than comparable suspended growth bioreactors. This suggests that operation of an attached growth process will be more focused on mechanical aspects than on process control. Experience with SAGBs confirms this, especially for the packed and fluidized bed options. Process operation generally concentrates on the maintenance of appropriate organic and hydraulic loading rates, which is accomplished by keeping an adequate number of bioreactors in service. Other process requirements must also be met, such as an adequate supply of oxygen for aerobic applications and sufficient quantities of an electron donor for denitrification applications. The bioreactor must also be monitored to ensure that it is maintained in proper operating condition. This can be accomplished by monitoring process performance and by visual observation of the bioreactor and its contents. Increased process control flexibility can be built into MBBRs and made available to operations personnel. Media can be moved from one zone to another (especially if the design does not require complete filling of all zones with media) to adjust process performance capabilities, zones can be converted from aerobic to anoxic operation if zones with both mixing and aeration equipment are provided, and the DO within aerobic zones can be adjusted. Integrated fixed film activated sludge systems may be viewed as suspended growth processes that also incorporate attached growth and, consequently, they provide all of the operational flexibility of suspended growth facilities along with the added flexibility provided by attached biomass. The reader is referred to Sections 11.4 and 12.4 for additional details concerning suspended growth operation.

An issue of particular concern is the control of biomass within a SAGB. Each type of SAGB incorporates mechanisms for maintaining the quantity and activity of attached biomass at appropriate levels. For packed beds, this is accomplished by backwashing. In addition, for denitrification in packed beds, a purge cycle may be required to remove nitrogen gas. For FBBRs, biomass control is accomplished by operation of the growth control system. A particular concern for these systems arises in connection with

separate stage nitrification systems operated at less than design loading conditions. The low growth yield of nitrifiers can lead to diminished nitrifier accumulation when these units are loaded at less than their full design loading condition, leading to inadequate performance under peak loading conditions. This can be addressed by maintaining a reduced number of units in direct service to maintain loading conditions more closely approaching design values, thereby maintaining a healthy population of nitrifiers, and by rotating the units in and out of service so that all units are "fed" at regular intervals.

Biomass control in MBBR and IFAS processes is accomplished by adjusting aeration rates to control fluid circulation and media sloughing rates or by passing polyurethane pad media through a cleaning device. Feeding patterns and zone operation can be adjusted to manage the growth of nitrifiers on the media. A healthy nitrifier biomass must be maintained in MBBR media intended for nitrification, and this requires regular feeding, as described above for packed and fluidized beds. Attached nitrifier growth can be diminished significantly in IFAS systems if operating conditions allow significant suspended growth nitrification. Regular monitoring of the biomass concentration and activity is necessary to determine how the biomass control system should be operated. This requires continual evaluation of process efficiency. In addition, for packed bed bioreactors, it requires monitoring hydraulic capacity and headloss; for FBBRs, bed height and biomass development on the carrier particles; and for MBBR and IFAS processes, observation of the media. Recent information on the operation of packed bed SAGBs is available.[32]

One problem observed with the media in some IFAS systems (particularly with rope media like Ringlace) is the growth of worms on the media, which can consume the biofilm and decrease process efficiency.[30,50,53] Problems associated with worm growth can be minimized by maintaining a sufficiently high TOL to sustain active heterotrophic growth. Infestations can also be controlled by periods of anoxia since the worms are obligate aerobes. An anoxic period of 12 hours has proven effective in at least one application.[53]

Information on the operation of these systems is often available from the system suppliers.

21.5 KEY POINTS

1. Submerged attached growth bioreactors (SAGBs) provide high biomass densities, resulting in compact bioreactors that require significantly less land area than traditional suspended and attached growth systems.
2. Submerged attached growth bioreactors can be used for aerobic and anaerobic removal of biodegradable organic matter, for nitrification, and for denitrification. Some SAGBs can also be operated to allow phosphorus accumulating microorganisms to grow.
3. Flow through the media in an SAGB can be upward, downward, or horizontal. Furthermore, granular media can be either packed or fluidized. Most SAGBs rely primarily on the attached biomass and are designed and operated with hydraulic residence times (HRTs) sufficiently short to minimize the growth of suspended biomass. However, integrated fixed film activated sludge (IFAS) processes use clarifiers to recycle suspended biomass, so that both suspended and attached biomass are utilized for treatment.
4. Packed bed bioreactors are usually used to treat wastewaters containing both soluble and particulate organic matter, such as domestic wastewaters. Fluidized bed biological reactors (FBBRs) are usually used to treat wastewaters containing mostly soluble organic matter, such as industrial wastewaters and contaminated groundwaters. Moving bed biological reactors (MBBRs) can treat a variety of wastewaters, but coagulation during primary treatment may be required if the wastewater contains significant quantities of particulate matter. Integrated fixed film activated sludge processes are usually used to upgrade existing suspended growth bioreactors.
5. The performance of SAGBs is often characterized by correlating it with the total organic loading (TOL), which is the mass flow rate of biodegradable organic matter divided by the bioreactor media volume.

6. The surface loading can also be used to characterize the performance of SAGBs. This parameter is particularly useful for characterizing the performance of attached media in MBBR and IFAS processes.
7. The total hydraulic loading (THL) for many SAGBs must be controlled within specified limits. For packed beds, the THL must not exceed certain maximum values to avoid excessive headloss and unreasonable backwash volumes. For FBBRs, the THL must be sufficiently large to maintain the media in a fluidized condition, yet not so large as to wash out the bioparticles.
8. The solids retention time (SRT) can be used to characterize the performance of FBBRs using inert media such as sand, and the suspended growth component of an IFAS.
9. Many factors, such as pH, temperature, and dissolved oxygen (DO) concentration affect the performance of SAGBs. Their effects are similar to those observed with other attached growth processes. In contrast to other fixed film processes, the DO in many SAGB processes can be adjusted, thereby changing the maximum removal rate.
10. Design procedures for SAGBs use the performance correlations described in items 5 through 8, above. Excess biomass production rates and oxygen requirements are estimated using net process yield factors and oxygen stoichiometric coefficients, similar to those used in the initial design of suspended growth systems.
11. Appropriate pretreatment is required to ensure long-term performance of SAGBs. Clogging of media and/or nozzles can result from debris or fibrous materials present in the wastewater. Pretreatment of municipal wastewater generally consists of primary clarification, either with or without chemical enhancement. A high-rate biological process may also be used.
12. Experience with IFAS processes indicates that placement of the media can be important in determining the effectiveness of the biomass that develops on it.

21.6 STUDY QUESTIONS

1. Prepare a table summarizing the characteristics of the submerged attached growth bioreactors discussed in this chapter. The table should summarize the differences in biomass concentration that can be achieved with the various bioreactors and the resulting differences in allowable organic loadings. Where possible, provide independent confirmation of the loading differences.
2. Prepare a table summarizing the benefits and drawbacks of SAGBs.
3. Describe how the TOL affects the removal of organic material and nitrification in an aerobic SAGB. Prepare a graph depicting the growth of heterotrophic and autotrophic bacteria in such a bioreactor, along with the removal of biodegradable organic matter and ammonia-N.
4. A wastewater with a flow of 25,000 m^3/day and a biodegradable COD concentration of 300 mg/L must be treated to reduce it to 60 mg/L. A UFPB bioreactor is to be used. Select the appropriate TOL and size the bioreactor. Determine the peak flow rate that can be processed. Determine the oxygen requirement and the excess biomass production rate. Will nitrification occur in this bioreactor?
5. The wastewater described in Study Question 4 must be treated to reduce the ammonia-N concentration to less than 5 mg/L. The initial TKN concentration is 40 mg/L. Evaluate two stage versus single stage bioreactor configurations. Compare the total bioreactor volume, the oxygen required, and the excess biomass produced for the two options. Summarize the relative advantages and disadvantages of each option.
6. A submerged packed bed system is to be designed to accomplish combined nitrification and denitrification of the wastewater described in Study Questions 4 and 5. Describe the procedure that you would use to size this system.

7. An industrial wastewater with a flow rate of 7500 m^3/day, a nitrate-N concentration of 150 mg/L, and no significant organic matter must be treated to remove the nitrate-N prior to discharge to a municipal wastewater treatment plant. Size and configure an FBBR for this purpose. The maximum and minimum practical bed depths are 7 m and 5 m, respectively.
8. Repeat Study Question 7, but with an influent nitrate-N concentration of 50 mg/L.
9. An IFAS bioreactor is to be designed to treat a wastewater with a flow of 25,000 m^3/day and a BOD_5 concentration of 175 mg/L. Plastic trickling filter media with a specific surface area of 140 m^2/m^3 will be used as the attached growth media, and will occupy a volume equal to 30% of the suspended growth bioreactor volume. The suspended growth process loading factor is to be maintained at less than 0.3 kg BOD_5/(kg MLSS·day), and the MLSS concentration is to be no more than 2500 mg/L. The performance of the attached growth media can be characterized using Figure 21.17. Size the IFAS bioreactor.
10. Based on your knowledge of the factors that affect the behavior of biofilms, describe those factors that limit the performance of SAGBs. Based on this analysis, how could the performance of these systems be improved?

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Part VI

Future Challenges

The history of wastewater treatment has been one of responding to need. The first concern of environmental engineers was to protect human health, and early wastewater treatment systems focused primarily on that objective, with some consideration for aesthetics. Next came the additional concern for the quality of the receiving water, with emphasis on the maintenance of adequate oxygen concentrations. This led to the development of biological treatment systems for the removal of oxygen demanding material (e.g., organic matter and ammonia-N). Then, as eutrophication became a problem, it became necessary to remove nutrients. The first application of biological systems for this purpose was denitrification. Then, as biological processes for the removal of phosphorus became available they began to displace physical/chemical systems because of their advantages. The next challenge was the need to remove individual organic compounds to low levels, rather than just removing organic matter in general. This has been due to the production and use of xenobiotic organic chemicals (XOCs) and to the discovery in our surface waters of trace quantities of organic chemicals from personal care and household products. Although XOCs are foreign to the biosphere, they have many beneficial uses in society, as do the consumer products leading to trace contaminants. Consequently, just as environmental engineers were given the responsibility of protecting human health and the aquatic environment from conventional pollutants, they have also been given the task of ensuring that XOCs and the residues from consumer products do not enter the environment in excessive quantities through treated wastewaters. This has proven to be a complicated assignment. Part of the difficulty is a result of the very nature of XOCs and other trace contaminants; because they are foreign to the biosphere they are often hard to biodegrade. Additional difficulties arise because of the low concentrations in which they are present in wastewater and the impacts of chemical/physical mechanisms on them. As a way of introducing this topic, Chapter 22 will address the factors influencing the fate and effects of XOCs and trace contaminants in bioreactors. Another important challenge facing environmental engineers, both now and in the future, is the design and operation of sustainable systems in an increasingly resource-limited world. Biochemical operations can play an important role in meeting this challenge because they can be employed to achieve greater availability of water resources, lower consumption of energy and chemicals, and greater resource recovery. The final chapter in this book addresses this significant topic.

22 Fate and Effects of Xenobiotic Organic Chemicals

It is clear from the material presented in the preceding chapters that our ability to design bioreactors for oxidation of biogenic organic matter, nitrification, and denitrification is well established. Furthermore, our understanding of biological phosphorus removal has advanced significantly, leading to better models, which have resulted in improved design procedures. Less well established, however, is our ability to design biological treatment systems for the biodegradation of xenobiotic organic chemicals (XOCs). The term xenobiotic means "foreign to the biosphere" and xenobiotic chemicals are those that have been produced by humankind through our chemical and pharmaceutical industries. While there is no question that these industries have had a large beneficial effect on society, there have also been negative effects associated with the release of XOCs to the environment. One job of the environmental engineer is to minimize those negative effects by destroying chemical residues using technologies that result in the minimum overall environmental impact.

In most cases, biological treatment processes were not designed with removal of XOCs in mind; nevertheless, these treatment systems are important tools in that effort. Consequently, in this chapter we will explore some of the factors that must be considered when applying biological treatment systems for the destruction of waterborne XOCs. We will focus primarily on XOCs from industrial operations where the contaminants are present at mg/L to μg/L concentrations. We know much less about trace XOCs that have been the focus of recent studies. They typically include pharmaceuticals and industrial products associated with personal care (e.g., shampoos and lotions) and the home (e.g., flame retardants). These tend to be present at μg/L to ng/L concentrations in wastewaters. Although these concentrations are lower than typical industrial XOC concentrations, many of the chemicals involved are biologically active at these levels, making their removal during wastewater treatment important. Ironically, these low concentrations make trace XOCs difficult to study.

22.1 BIODEGRADATION

At first glance it might seem strange to try and use biological treatment systems to destroy XOCs. After all, the destruction of organic matter in such systems is caused by microorganisms that use the organic matter as a carbon and energy source. The pathways employed for degrading biogenic organic matter have developed over long periods of time and are quite effective. Since XOCs have only been on the earth for a very short time period (most for 60 years or less), how is it possible for microorganisms to possess enzymatic pathways that can act on them? The answer to that question is twofold. First, most XOCs have some structural similarity to biogenic materials, allowing them to fit into active sites on enzymes that did not evolve with them as substrates. Second, the specificity of enzymes is not exact.[1] While the type of reaction catalyzed by a given enzyme is very specific, the range of substrates upon which it can act is not. In other words, biodegradation of XOCs is largely fortuitous. With this in mind, we will first consider the requirements for biodegradation of an XOC.

22.1.1 Requirements for Biodegradation

Because initiation of biodegradation of an XOC requires the presence of an enzyme that is able to perform a transformation reaction, it follows that biodegradation of the XOC requires the presence of a microorganism with the genetic capability to synthesize that enzyme. Furthermore, if mineralization of the XOC is to occur, the transformation product from the first reaction must serve as the substrate for another transforming enzyme, and so on, until ultimately a biogenic product is formed that will funnel into common metabolic pathways. The enzymes catalyzing reactions subsequent to the initial reactions may not reside in the same microorganism as the ones producing the compounds upon which they act; consequently, mineralization may require the concerted action of a community, rather than a single type of organism.

Regardless of where the genes reside, maintenance of the appropriate genetic capability is an absolute requirement for biodegradation to occur. This means that the environmental conditions must be appropriate for growth of the required microorganisms. Not only must the required nutrients and electron acceptor be provided, but the system solids retention time (SRT) must be sufficiently long to maintain all of the microorganisms in the system. If the system SRT is less than that value, the required microorganisms will be lost and biodegradation will cease.[13] This is an important concept because many XOCs are biodegraded slowly, making the minimum SRT relatively long, as illustrated in Figure 10.3. Furthermore, the biodegradation kinetics of some XOCs are particularly temperature sensitive, causing the minimum SRT to rise rapidly as the temperature drops.[30]

The requirement to maintain the appropriate microbial community can cause serious problems if the XOC is only intermittently discharged to a bioreactor. The capability to degrade any given organic chemical is not necessarily widely distributed in the microbial world. While the ability to use common sugars like glucose and fructose is widespread,[11] other biogenic organic compounds are utilized by only a fraction of all isolates tested.[11,22] Thus, one would anticipate that the ability to use an XOC would be even less widely distributed. When discharge of an XOC to a bioreactor stops, the niche by which the capable microorganisms are maintained may be eliminated, causing them to be lost from the system. The time constant for their loss is the SRT of the system, with the result that about 95% of them will be lost in three SRTs. Thus, it can be seen that most of the microorganisms capable of degrading an XOC can be lost if the XOC is absent from the bioreactor feed for only a few weeks, unless there is some other means of preserving their niche. This means that the capable population must be rebuilt upon reintroduction of the XOC in the influent, causing time lags during which release of the XOC in the effluent occurs.

Even if the capable microorganisms are retained, chances are that the required enzymes will not be synthesized in the absence of the XOC. This follows from the fact that most organic compounds are degraded by inducible enzymes; that is, those that are synthesized only when they are needed. When the inducer is no longer present, their synthesis is stopped and the unused enzymes are degraded, freeing their amino acids for use in the synthesis of new enzymes. Because proteins have a relatively high turnover rate in bacteria, this loss can be quite rapid, causing a culture to lose the ability to degrade an XOC much faster than would be predicted from simple washout of the enzyme.[3] Then when the inducer is reintroduced into the system, new enzyme synthesis must occur, leading to another type of lag. While this lag is considerably shorter than the lag associated with the regrowth of a population, it can still be sufficiently long to allow discharge of an XOC following its reintroduction in the feed to a bioreactor. All of this suggests that the very nature of biological systems dictates that they will be most effective in destroying XOCs when the XOCs are continuously discharged to treatment systems.

Trace XOCs may serve as the exception to the concept that enzymes that facilitate the biodegradation of XOCs must be induced by the presence of the XOC or a structurally similar compound. Rather, basal (noninduced) enzyme levels may provide sufficient activity. For example, in the case of 17α-ethinylestradiol, an oral contraceptive hormone, the oxidation potential provided by basal oxygenase enzyme activity in a noninduced, mixed heterotrophic culture exceeded the amount needed

to oxidize and cleave the aromatic rings present in the 17α-ethinylestradiol, thereby initiating its mineralization.[40] As we learn more about biodegradable trace XOCs, mechanisms of biodegradation will be better understood, leading to improved approaches to their biodegradation.

22.1.2 Factors Influencing Biodegradation

The primary factor determining the ability of microorganisms to degrade an XOC, as well as the kinetics of that biodegradation, is its molecular structure. The closer that structure is to the structure of a biogenic compound, the easier the XOC will be to biodegrade because the more readily it will fit into common metabolic pathways. Xenophores are substituents on organic molecules that are physiologically uncommon or entirely nonphysiological,[1] and their presence is one factor that can make a compound xenobiotic. Because they are alien to most organisms, they hinder the functioning of many enzymes and it is only the nonspecificity of those enzymes that allows them to function in the presence of xenophores. The nature, number, and position of xenophores all influence the biodegradability of an XOC. Halogens, nitro groups, and cyanide groups are typical xenophores and all reduce biodegradability in comparison to the unsubstituted compound. Furthermore, the greater the number of xenophores, the less susceptible the XOC is to biodegradation. However, it is difficult to generalize about the effect of the position of a xenophore. A given xenophore in one position may have little impact, whereas in another its effect may be large. Similarly, one xenophore in a given position may have a strong effect, whereas another xenophore in the same position may have none. Because of these widely diverse effects, there is strong interest in the development of structure-biodegradability relationships that can be used to deduce biodegradability from a compound's molecular structure.[1] However, it must be recognized that the knowledge base for such relationships is still limited, thereby restricting their utility. Prediction of biodegradability is an evolving science and much is still to be learned.

With a few exceptions, microorganisms are thought to degrade only organic compounds that are dissolved in the aqueous phase.[1] Furthermore, as we saw in Chapter 3, the rate at which bacteria grow on a substrate is a function of its liquid phase concentration. This means that solubility has a profound effect on the biodegradability of any organic compound, whether xenobiotic or biogenic. Many xenophores reduce aqueous solubility, thereby reducing bioavailability and the rate of biodegradation. Furthermore, if a compound has very low solubility, it may be difficult to induce the enzymes required for its biodegradation.[1]

Finally, the environment has a strong effect on biodegradability. In addition to the usual effects of pH, temperature, and the availability of nutrients and electron acceptor, the presence or absence of molecular oxygen can have a strong effect. For example, some enzymatic steps, such as those carried out by oxygenase enzymes, require the presence of molecular oxygen, whereas others, such as reductive dehalogenation, require its absence. The engineer must be knowledgeable about the nature of the potential pathways for biodegradation of a given XOC so that the appropriate environment can be provided. Failure to do so will result in an inadequate system that cannot meet effluent goals.

22.1.3 Classes of Biodegradation and Their Models

Engineers need to quantify biodegradation rates in order to design a biological process capable of achieving a desired effluent concentration of a given XOC. This requires the use of models and the evaluation of the parameters in them. For modeling purposes, biodegradation has been divided into two broad categories, growth-linked and cometabolic.

22.1.3.1 Growth-Linked Biodegradation

Most biodegradation occurs by growth-linked metabolism. By that we mean that the microorganisms performing the biodegradation receive their carbon and energy from degradation of the

XOC and that the XOC can serve as the sole carbon and energy source for microbial growth. As a consequence, the transformation products from the initial biodegradative reactions with the XOC ultimately enter into the normal metabolic pathways wherein some of the carbon is incorporated into new cell material and the remainder is released as carbon dioxide. The result is mineralization of the XOC because it has been converted into those innocuous products. Engineered systems almost always employ natural microbial communities, rather than pure microbial cultures, and so it is not unusual for biodegradation of XOCs to occur by the coordinated action of several species within that community.[64] Such assemblages are called consortia. The models employed are similar to those used elsewhere in this text, which also reflect the activity of microbial communities.

When biodegradation occurs by growth-linked metabolism, the reactions involved may be depicted by the chemical oxygen demand (COD)-based stoichiometric equation for biomass growth depicted by Equation 3.8. Consequently, the rates of biomass growth, substrate removal, and oxygen utilization are all linked as expressed in Equation 3.34. Furthermore, the rate of biomass growth is first-order with respect to the concentration of biomass actually involved in the biodegradation, as depicted in Equation 3.35. In other words, everything is the same as growth on soluble COD, except that the heterotrophic biomass concentration must reflect only the concentration of the capable biomass, which will be only a fraction of the total heterotrophic biomass concentration as discussed in Section 22.1.1. The specific growth rate coefficient, μ, is related to the concentration of the XOC through the Monod equation (Equation 3.36) if the XOC is not inhibitory to its own biodegradation and by the Andrews equation (Equation 3.39) if it is. The parameters in these models may be evaluated in batch reactors using the techniques discussed in Section 9.4.

As discussed in Section 9.4.1, two types of kinetic parameter estimates, intrinsic and extant, may be obtained, depending on the type of assay employed and its effect on the physiological state of the bacteria.[29] Intrinsic parameter values are those that reflect the ability of the bacteria to grow in unrestricted growth on an XOC as sole carbon and energy source. They are unaffected by the growth history of the biomass, except as it determines the composition of the microbial community, and thus are useful for comparing the relative biodegradability of organic compounds. Extant parameter values, on the other hand, reflect the physiological state of the biomass in the bioreactor from which it was obtained. They are most useful for predicting process performance,[48] but their estimation requires knowledge of the concentration of capable biomass.[47] Because it is sometimes difficult to estimate the capable biomass in absolute terms, a calibration procedure has been developed for use with existing bioreactors.[20] Furthermore, even though wastewaters always contain complex mixtures of organic compounds, the extant parameter values describing the biodegradation of an XOC can be determined from tests employing the XOC as the sole substrate. This follows from the observation that parameter values measured in single-substrate tests are the same as those measured in tests in which the biomass is simultaneously using a complex mixture of biogenic substrates at the same specific rate that it is using them in the bioreactor from which the biomass came.[21]

In general, extant parameter estimates are characterized by lower $\hat{\mu}$ and K_S values than intrinsic ones.[48] For example, intrinsic K_S values tend to be on the order of 1 to 10 mg/L whereas extant K_S values generally lie between 0.1 and 1.0 mg/L.[48] Extant estimates generally predict lower effluent XOC concentrations from completely mixed activated sludge (CMAS) systems at typical SRTs than do intrinsic parameters and are better predictors of bioreactor performance.[48]

22.1.3.2 Cometabolic Biodegradation

Cometabolism is the transformation of an organic compound by microorganisms that are unable to use the compound or its transformation products as a source of carbon or energy.[36] Consequently, the microorganisms derive no nutritional benefit from a substrate that they cometabolize and, in

fact, the transformation may be detrimental to the cells.[12,16] Cometabolism results from the lack of specificity of enzymes discussed previously.[17,36] In this case, however, the product of the reaction is a dead-end product for the microorganisms carrying out the reaction. As a consequence, the microorganisms must have another substrate that they use for energy and growth. Cometabolism is easy to demonstrate in pure culture because the transformation product accumulates. In mixed microbial communities, on the other hand, the transformation product may serve as a growth substrate for other types of microorganisms, leading ultimately to the mineralization of the original substrate. Because the microorganism carrying out the cometabolic transformation derives no benefit from it, there is no selective pressure to foster its growth or to retain it in the system. As a consequence, if it is necessary to depend on a cometabolic transformation, an environment must be created that will retain the desired microbe and foster its growth. This can be a major challenge for a design engineer.

Cometabolism of a large number of XOCs has been demonstrated in pure culture.[1,16] An example of a major engineering application of cometabolism, however, has been in the transformation of trichloroethylene and other halogenated aliphatic compounds by methanotrophs.[16] This transformation is particularly interesting because it requires reducing power in the form of NADH and also leads to inactivation of the microorganisms. As a consequence, an engineered system must provide an electron donor and/or a growth substrate to replenish the reducing power consumed in the cometabolic reaction and to continually provide new biomass to replace that lost to inactivation. Understanding the reactions involved and expression of them in forms that can be used in conceptual and mathematical models has been a major challenge involving a number of researchers in both engineering and microbiology.

The modeling of cometabolic biodegradation is more complicated than the modeling of growth-linked biodegradation, but is built on the same concepts. Two separate reactions must be modeled, growth on the carbon and energy source and the cometabolic transformation. Growth on the carbon and energy source is modeled in exactly the same way as any other growth-linked process, using the Monod equation as discussed throughout this text. The cometabolic transformation is usually modeled as an enzymatic reaction using either the Michaelis-Menten equation (analogous in form to the Monod equation) for a noninhibitory substrate or the Haldane equation (analogous in form to the Andrews equation) for an inhibitory one. Competition may occur between the growth substrate and the cometabolic substrate for the shared enzyme, and thus terms for this competition must be provided in the reaction rate expressions for both processes. In addition, provision must be made for utilization of the growth substrate to provide reducing power if it is required. This has been handled as an additional decay term. Finally, inactivation of the cells must be modeled if it occurs, and this, too, has been handled as another decay term. Details about this modeling approach are provided elsewhere.[12,16]

22.2 ABIOTIC REMOVAL MECHANISMS

Biodegradation may not be the only mechanism contributing to the loss of an XOC from a bioreactor. Because of the physicochemical properties of some XOCs, abiotic removal mechanisms such as volatilization to the atmosphere and sorption onto solids may contribute to their loss. Quantification of those losses is important for a variety of reasons. First, they represent ways in which the XOCs can enter the environment without alteration. Since XOCs may have negative environmental effects, such losses should be minimized. Second, depending on the configuration of the bioreactor, abiotic losses may influence the concentrations of the XOCs in the effluent and thus accurate prediction of those concentrations requires that the magnitude of the abiotic losses be known. In this section we will review briefly techniques for quantifying such losses. The rate expressions presented can be used in mass balance equations for the XOCs along with the rate expression for biodegradation in order to obtain a complete picture of the fate of the XOCs.

22.2.1 Volatilization

22.2.1.1 Models for Volatilization

Volatilization is an interphase mass transfer process in which a constituent in the liquid phase is transferred to the gas phase and it may contribute to the loss of an XOC from an activated sludge bioreactor during the transfer of oxygen to the system. Models for interphase mass transfer are of the form

$$r_{v,XOC} = -K_{L,XOC}a(S_{XOC} - S^*_{XOC})V, \tag{22.1}$$

where $r_{v,XOC}$ is the rate of loss by volatilization of the XOC from the control volume V (mg/hr), $K_{L,XOC}a$ is the overall liquid phase mass transfer coefficient for the XOC (hr^{-1}), S_{XOC} is the liquid phase concentration of the XOC in the control volume (mg/L), and S^*_{XOC} is the liquid phase concentration that would exist if the liquid were in equilibrium with the gas phase. Equilibrium between the gas and liquid phases is assumed to be governed by Henry's Law for dilute solutions[46,60]

$$S^*_{XOC} = \frac{C_{G,XOC}}{H_{c,XOC}}, \tag{22.2}$$

in which $C_{G,XOC}$ is the gas phase concentration of the XOC (mg/L) and $H_{c,XOC}$ is the dimensionless Henry's Law coefficient for the XOC. The Henry's Law coefficient in this context is the mass per volume-based dimensionless coefficient:

$$[H_{c,XOC}] = \frac{\dfrac{\text{mg of XOC in gas phase}}{\text{L of gas phase}}}{\dfrac{\text{mg of XOC in liquid phase}}{\text{L of liquid phase}}} = \frac{\dfrac{\text{moles of XOC in gas phase}}{\text{L of gas phase}}}{\dfrac{\text{moles of XOC in liquid phase}}{\text{L of liquid phase}}}. \tag{22.3}$$

It is related to the conventional Henry's Law coefficient, H_{XOC}, which has units of (atmospheres of XOC in gas phase·m_L^3)/(moles of XOC in liquid phase) by[52]

$$H_{c,XOC} = \frac{H_{XOC}}{RT}, \tag{22.4}$$

where R is the universal gas constant ($= 82.06 \times 10^{-6}$ atm·m^3/mole·K) and T is the absolute temperature (K).

The equation for the mass rate of loss of an XOC from an activated sludge bioreactor by volatilization depends on the type of oxygen transfer system employed. If mechanical surface aeration is used, the value of S^*_{XOC} can be considered to be zero because the atmosphere acts as an infinite sink for the XOC, making $C_{G,XOC}$ equal to zero.[38,52] Consequently, Equation 22.1 can be simplified to

$$r_{v,XOC} = -K_{L,XOC}a \cdot S_{XOC} \cdot V, \tag{22.5}$$

and the mass loss rate can be considered to be first-order with respect to the liquid phase concentration of the XOC. When oxygen is transferred to an activated sludge bioreactor by diffused aeration,

the situation is more complicated.[37,58] As an air bubble rises to the surface from the diffuser, the concentration of the XOC in it continually increases, thereby reducing the driving gradient for transfer. Consequently, the use of Equation 22.5 would overestimate the loss of the XOC by volatilization. Under that circumstance, the mass removal rate by volatilization is given by

$$r_{v,XOC} = C_{G,XOC} \cdot Q, \tag{22.6}$$

where Q is the volumetric air flow rate through a bioreactor of volume V and

$$C_{G,XOC} = S_{XOC} \cdot H_{c,XOC}\left[1 - \exp\left(-\frac{K_{L,XOC}a \cdot V}{Q \cdot H_{c,XOC}}\right)\right]. \tag{22.7}$$

22.2.1.2 Estimation of Coefficients

Estimation of the volatile losses of an XOC from an activated sludge bioreactor by either type of oxygen transfer system requires knowledge of the mass transfer coefficient $K_{L,XOC}a$. Measurement of a mass transfer coefficient is time consuming and expensive. Since information is generally available about the overall mass transfer coefficient for oxygen to the bioreactor, $K_{L,O2}a$, it would be advantageous to be able to estimate the mass transfer coefficient for the XOC from the oxygen mass transfer coefficient, and this is what is commonly done, using an expression of the form:[37,38,52,58]

$$K_{L,XOC}a = \Phi_{XOC} \cdot K_{L,O2}a, \tag{22.8}$$

in which Φ_{XOC} is a proportionality factor for a given XOC.

According to the two-resistance theory of Lewis and Whitman,[44] the overall liquid phase mass transfer coefficient is the result of two resistances in series due to the liquid and gas boundary layers:

$$\frac{1}{K_{L,XOC}a} = \frac{1}{k_{L,XOC}a} + \frac{1}{k_{G,XOC}a \cdot H_{c,XOC}}. \tag{22.9}$$

Many authors have assumed that the overall liquid phase mass transfer coefficient for any XOC can be approximated by $k_{L,XOC}a$, the liquid film coefficient, as can be done for oxygen, which is a slightly soluble gas. This cannot be done for all XOCs, however, because of their physicochemical characteristics,[37,38,52,58] and thus consideration must also be given to the gas phase film coefficient, $k_{G,XOC}a$. The liquid film coefficients for two solutes may be related to each other by the ratio of their diffusivities in water raised to the n power. The relationship between the overall liquid phase coefficients, however, will also depend on the fraction of the mass transfer resistance for the XOC that can be attributed to the liquid phase.[38,52] As a consequence, Φ_{XOC} can be represented by[38,52]

$$\Phi_{XOC} = \frac{\left[\frac{D_{w,XOC}}{D_{w,O_2}}\right]^n}{\left[1 + \frac{1}{\left(\frac{k_{G,XOC}}{k_{L,XOC}}\right)H_{c,XOC}}\right]}, \tag{22.10}$$

in which $D_{w,XOC}$ and D_{w,O_2} are the diffusivities of the XOC and oxygen in water, respectively. Caution must be used in the application of Equation 22.10 because considerable variability exists in the literature regarding the values of both the diffusivity and the Henry's Law coefficient for XOCs. Furthermore, there is uncertainty about the value of n, although it appears to be approximately 0.5, which is in accordance with both the penetration and the surface renewal theories of mass transfer.[52]

The major difficulty in the application of Equation 22.10 is the lack of a broad base of information about the ratio of the gas to liquid phase film coefficients, $k_{G,XOC}/k_{L,XOC}$. Munz and Roberts[52] and Hsieh et al.[38] have both studied the effect of the power input per unit volume (P/V) for lab-scale mechanical surface aeration systems. Munz and Roberts[52] studied seven XOCs over a power range from 10 to 320 W/m^3 and found that:

$$\text{Log}_{10}\left(\frac{k_{G,XOC}}{k_{L,XOC}}\right) = -0.376\ \text{Log}_{10}\left(\frac{P}{V}\right) + 2.389. \tag{22.11}$$

Hsieh et al.[38] studied 20 XOCs over a power range from 30 to 500 W/m^3 and found that:

$$\text{Log}_{10}\left(\frac{k_{G,XOC}}{k_{L,XOC}}\right) = -1.85\ \text{Log}_{10}\left(\frac{P}{V}\right) + 2.43. \tag{22.12}$$

The difference in the correlations obtained by the two groups of researchers suggests that the system configuration influences the effect of the power density of mechanical surface aeration on the value of $k_{G,XOC}/k_{L,XOC}$. Nevertheless, both correlations suggest that it is possible to obtain a rough estimate of the overall liquid phase mass transfer coefficient for an XOC from knowledge of the oxygen transfer coefficient, and that may be all that is needed to establish the relative importance of volatilization as a removal mechanism. Hsieh et al.[37] have also studied the effect of the air flow rate per unit volume (Q/V) in a diffused aeration system on the value of $k_{G,XOC}/k_{L,XOC}$. The range of $k_{G,XOC}/k_{L,XOC}$ values was relatively small (2.2–3.6) over a broad range of Q/V, suggesting that a fixed value of 2.6, which was the mean of their observations, could be used for all systems with little error.[37] The small values of $k_{G,XOC}/k_{L,XOC}$ for these systems, however, suggests that gas phase resistance is very important in them and should not be neglected in estimating overall liquid phase mass transfer coefficients for XOCs.

22.2.2 Sorption

Any XOC that is sorbed onto the biomass and other solids in a bioreactor will be removed from the system with the waste biomass. Since the sorbed XOC will not be chemically altered, the potential exists for it to desorb from the waste biomass during its handling, processing, and disposal, potentially leading to a release of the XOC to the environment. Because of the potential of such releases, it is important to quantify them so that they may be controlled or eliminated if necessary.

22.2.2.1 Mechanisms and Models

Sorption of an XOC onto biomass is a complex process, involving both adsorption to the surface of the solids and absorption into cellular components, particularly the lipids. Because the exact mechanism is seldom known, the term "sorption" is typically used to describe the phenomenon and the determination of sorption coefficients is accomplished empirically. Unlike sorption to soil and sediments, which typically display two distinctly different rates with each contributing significantly to the removal,[39] sorption onto biomass is very rapid, with the vast majority of

the sorption occurring in a matter of minutes, followed by slow sorption of an additional small amount over a period of hours.[18,71] For example, one study[76] found that the liquid phase concentration of di-*n*-butyl phthalate was essentially the same after two minutes of contact with biomass as it was after 72 hours of contact, whereas others have found that equilibrium was approached within an hour.[19,75]

Desorption is the release of a sorbed chemical from the sorbent and, as such, is the opposite of sorption. In some cases, sorption is fully reversible, with the desorption relationship being the same as the sorption relationship. For example, the sorption of di-*n*-butyl phthalate,[76] lindane,[71] diazinon,[5,6] and 2-chlorobiphenyl[5,6] were all found to be fully reversible. In other cases, sorption may be irreversible, suggesting that chemical reactions are involved,[70] or the reversibility may change over time,[75] suggesting a shift in the relative importance of adsorption and absorption. All of this suggests that the reversibility of sorption will depend on the nature of the chemical and biosolids, and must be determined on a case-by-case basis. In spite of this, reversibility appears to be common and is often assumed.

A number of models are available to express the equilibrium relationship between the concentration of an XOC in the liquid phase and the quantity of the XOC on the solid phase.[46,60] All are referred to as isotherms because the equilibrium relationship is influenced by temperature, requiring it to be quantified for a fixed temperature. For most purposes, the Freundlich isotherm is adequate:

$$C_{S,XOC} = k_{s,XOC} \cdot S_{XOC}^{n}, \tag{22.13}$$

where $C_{S,XOC}$ is the concentration of the XOC on the solid phase, $k_{s,XOC}$ is the sorption coefficient, and n is an empirical coefficient. The units of $k_{s,XOC}$ depend on the value of n and the units of $C_{S,XOC}$ and S_{XOC}. Typically, $C_{S,XOC}$ has units of mg/g and S_{XOC} has units of mg/L. Thus, when n has a value of 1.0, $k_{s,XOC}$ has units of L/g. At the low concentrations at which XOCs are typically present in wastewaters, isotherms are often linear, allowing n to be taken as 1.0.[60] Such a value should not be assumed for all cases, however, particularly for higher XOC concentrations. The best policy is to determine the values of $k_{s,XOC}$ and n experimentally.

The major loss of an XOC due to sorption onto biomass in the activated sludge process comes from biomass wastage. Consequently, the rate of loss by sorption, $r_{s,XOC}$, is given by

$$r_{s,XOC} = -F_W \cdot X_{M,T,w} \cdot k_{s,XOC} \cdot S_{XOC}^{n}, \tag{22.14}$$

where F_w and $X_{M,T,w}$ are the flow rate and concentration of the wasted mixed liquor suspended solids from the bioreactor.

22.2.2.2 Estimation of Coefficients

Although it is preferable to evaluate the coefficients $k_{s,XOC}$ and n experimentally for a given biomass, the procedure is tedious and expensive. Thus, it would be desirable to have a way to estimate them from the literature, particularly for preliminary engineering studies in which only an estimate of the relative importance of sorption as a removal mechanism is needed. Two methods exist. One allows transfer of information on a given XOC from one biomass to another, whereas the other allows extrapolation of information on one XOC to another XOC of the same type for the same biomass.

The vast majority of sorption to the biomass in an activated sludge system is to the organic fraction. Since the materials constituting the sorptive organic fractions of various sludges are similar, if a sorption coefficient is expressed per unit of organic carbon it can be used for any type of wastewater solids, even those from different plants.[19] Furthermore, since the mixed liquor volatile suspended solids (MLVSS) concentration is proportional to the organic carbon content of an

activated sludge, the same should be true when the sorption coefficient is expressed per unit of MLVSS. Thus, a literature value for the sorption coefficient for a given XOC on activated sludge can be used to approximate the sorption coefficient on biomass from another plant, provided that the coefficients are expressed per unit of organic carbon or MLVSS.[19]

The octanol:water partition coefficient, k_{OW}, is a commonly reported characteristic of organic chemicals and is representative of their hydrophobicity, with larger k_{OW} values indicating more hydrophobic compounds.[60] Because sorption is related to the tendency of an XOC to leave the water phase, it is related to its hydrophobicity. Consequently, the sorption coefficient, $k_{s,XOC}$, is related to the octanol:water partition coefficient. A number of researchers have developed correlations of the type

$$\text{Log}_{10}\ k_{s,XOC} = a\ \text{Log}_{10}\ k_{OW,XOC} + b, \tag{22.15}$$

where $k_{s,XOC}$ is expressed on a per unit carbon basis. Schwarzenbach et al.[60] summarized values for the coefficients a and b for a number of types of XOCs likely to be found in wastewater. For several types of compounds, the values of a were around 1.0. For classes of XOCs for which that is true, it is possible to estimate the sorption coefficient for one XOC from a measured sorption coefficient for another on the same biomass. If a can be set equal to 1.0, then it follows from Equation 22.15 that

$$k_{s,XOC2} = k_{s,XOC1}\left(\frac{k_{OW,XOC2}}{k_{OW,XOC1}}\right). \tag{22.16}$$

Thus, the sorption coefficient for XOC #2 can be estimated from a measured sorption coefficient for XOC #1 using handbook values of the octanol:water partition coefficients for the two XOCs.

22.3 RELATIVE IMPORTANCE OF BIOTIC AND ABIOTIC REMOVAL

The relative contributions of biotic and abiotic removal mechanisms toward the removal of an XOC from a bioreactor can be obtained by substitution of the rate expressions for biodegradation, volatilization, and sorption into the mass balance equations for the XOC in the particular bioreactor type and solving those equations as discussed elsewhere in this text. The resulting output will depend on the configuration of the bioreactor and the method of oxygen transfer employed, and thus will be specific to each system. Often, however, during preliminary engineering studies it would be advantageous to have a rough estimate of the relative importance of the three removal mechanisms. Such an estimate can be obtained easily for a CMAS system operated with the Garrett flow scheme and being oxygenated by mechanical surface aeration by combining the rate expressions for the two abiotic removal mechanisms with biodegradation as depicted by the simple, traditional model of Chapter 5 and making a few simplifying assumptions.[28]

In most situations, the concentration of a particular XOC in the influent to an activated sludge system will be only a small fraction of the total biodegradable COD entering the plant. Consequently, a loss of some of the XOC due to the abiotic mechanisms will have little, if any, effect on the mixed liquor suspended solids (MLSS) concentration, $X_{M,T}$. This is important because sorption of the XOC will occur on the entire MLSS, not just the capable biomass. Furthermore, based on the discussion in Section 22.1, it is likely that only a small fraction of the biomass in the MLSS will be involved in the biodegradation of the XOC.[9] By assuming that this capable biomass arises only from degradation of the XOC, its concentration, $X_{B,XOC,T}$, can be calculated as a separate entity. A steady-state mass balance on the capable biomass leads to Equation 5.21, the familiar expression linking the specific growth rate to the SRT. Substitution of this expression into the Monod

equation (Equation 3.36) leads to the familiar equation for the concentration of a soluble substrate in a CMAS bioreactor:

$$S_{XOC} = \frac{K_{S,XOC}\left(\frac{1}{\Theta_c} + b_{XOC}\right)}{\hat{\mu}_{XOC} - \left(\frac{1}{\Theta_c} + b_{XOC}\right)}. \tag{5.22a}$$

In this case, however, the kinetic coefficients, $\hat{\mu}_{XOC}$ and $K_{S,XOC}$, are specific to the XOC. The important point about Equation 5.22a in this context is that the effluent concentration of an XOC from a CMAS system is determined solely by the system SRT and is independent of the abiotic removal mechanisms. This is an important concept.

The impact of the abiotic mechanisms in a CMAS bioreactor is to reduce the concentration of the biomass capable of degrading the XOC. In this case the mass balance equation on that XOC must contain three loss terms, rather than the single loss term used in Equation 5.26. Loss of the XOC by biodegradation is still given by Equation 3.43, while loss by volatilization and sorption are given by Equations 22.5 and 22.14, respectively. Furthermore, since the Garrett flow scheme is being used, the concentration of the waste biomass, $X_{M,T,w}$, is equal to the MLSS concentration, $X_{M,T}$. Substitution of these loss terms into the mass balance equation for the XOC and simplification lead to the equation for the capable biomass concentration, $X_{B,XOC,T}$ in the presence of abiotic losses:

$$X_{B,XOC,T} = \frac{Y_{XOC,T}}{\left(\frac{1}{\Theta_c} + b_{XOC}\right)}\left[\frac{1}{\tau}(S_{XOC,O} - S_{XOC}) - K_{L,XOC}a \cdot S_{XOC} - \frac{X_{M,T} \cdot k_{s,XOC} \cdot S_{XOC}}{\Theta_c}\right]. \tag{22.17}$$

In the absence of abiotic removal mechanisms, both $K_{L,XOC}a$ and $k_{s,XOC}$ are equal to zero, causing Equation 22.17 to simplify to Equation 5.28a, the familiar equation for the biomass concentration in a CMAS bioreactor:

$$X'_{B,XOC,T} = \frac{Y_{XOC,T}}{\left(\frac{1}{\Theta_c} + b_{XOC}\right)}\left[\frac{1}{\tau}(S_{XOC,O} - S_{XOC})\right], \tag{5.28a}$$

where the prime on $X_{B,XOC,T}$ denotes that it represents the concentration of capable biomass that would result from the biodegradation of the XOC in the absence of abiotic removal mechanisms. Both Equation 22.17 and Equation 5.28a are based on the assumption that the only source of capable biomass is the degradation of the XOC. Comparison of those equations makes it clear that the role of abiotic removal mechanisms in a CMAS bioreactor is to reduce the concentration of the capable biomass.

Equations 22.17 and 5.28a can be used to estimate the contribution of abiotic mechanisms to the overall removal of an XOC by a CMAS system. Examination of Equation 5.28a reveals that in the absence of abiotic removal mechanisms, the capable biomass concentration, $X'_{B,XOC,T}$, is directly proportional to the removal of the XOC by the bioreactor ($S_{XOC,O} - S_{XOC}$). Furthermore, since the concentration of the XOC in the bioreactor, S_{XOC}, is not affected by abiotic mechanisms and is determined solely by the SRT as denoted in Equation 5.22a, the decrease in the capable biomass concentration associated with the action of the abiotic mechanisms is directly proportional to the amount of the XOC removed by those mechanisms. Thus, if we let $\Delta X_{B,XOC,T}$ represent the decrease in the

capable biomass concentration as a result of the action of the abiotic removal mechanisms, then γ, the fraction of the XOC removal attributable to abiotic removal mechanisms can be calculated as

$$\gamma = \frac{\Delta X_{B,XOC,T}}{X'_{B,XOC,T}}. \tag{22.18}$$

The value of $\Delta X_{B,XOC,T}$ can be obtained by subtracting Equation 22.17 from Equation 5.28a. Substitution of it into Equation 22.18 and rearrangement leads to

$$\gamma = \frac{\alpha_a \left(\frac{S_{XOC}}{S_{XOC,O}} \right)}{\left(1 - \frac{S_{XOC}}{S_{XOC,O}} \right)}, \tag{22.19}$$

in which α_a is a dimensionless abiotic loss coefficient. It is made up of the dimensionless volatilization loss coefficient, α_v, and the dimensionless sorption coefficient, α_s:

$$\alpha_a = \alpha_v + \alpha_s, \tag{22.20}$$

where

$$\alpha_v = \tau \cdot K_{L,XOC} a \tag{22.21}$$

and

$$\alpha_s = \frac{\tau \cdot k_{s,XOC} \cdot X_{M,T}}{\Theta_c}. \tag{22.22}$$

Equation 22.19 lends itself to graphical presentation, as shown in Figure 22.1 where the fraction of XOC removal due to abiotic mechanisms, γ, is plotted versus the fraction of the XOC remaining in the effluent, $S_{XOC}/S_{XOC,O}$, with the abiotic loss coefficient, α_a, as a parameter. Values of α_a from 0.001 to 1000 are provided to cover a broad range of conditions. Examination of Figure 22.1 makes it clear that the smaller the fraction of the XOC remaining in the CMAS effluent, the less important abiotic removal mechanisms are. This follows directly from the fact that the concentration of the XOC in the CMAS effluent is controlled by the SRT as expressed in Equation 5.22, and from the fact that both volatilization and sorption are directly proportional to the XOC concentration in the CMAS bioreactor. However, Equation 5.22 also tells us that the concentration of the XOC in the CMAS system is independent of the influent XOC concentration. As a consequence, at a fixed SRT, the lower the influent XOC concentration, $S_{XOC,O}$, the greater the contribution of abiotic removal mechanisms to the overall removal of the XOC, although the total mass of the XOC removed by those mechanisms will be constant.

Equation 22.20 shows that the effects of the two abiotic removal mechanisms are additive. Consequently, the relative importance of each is simply α_v/α_a or α_s/α_a. This can be very helpful in assessing their relative contributions in a given system. For example, *m*-xylene is a volatile XOC that is not very sorptive. In a lab-scale CMAS system, the value of α_v was 76.8 whereas the value of α_s was only 0.007.[28] Consequently, only 0.009% of the abiotic losses were due to sorption. Conversely,

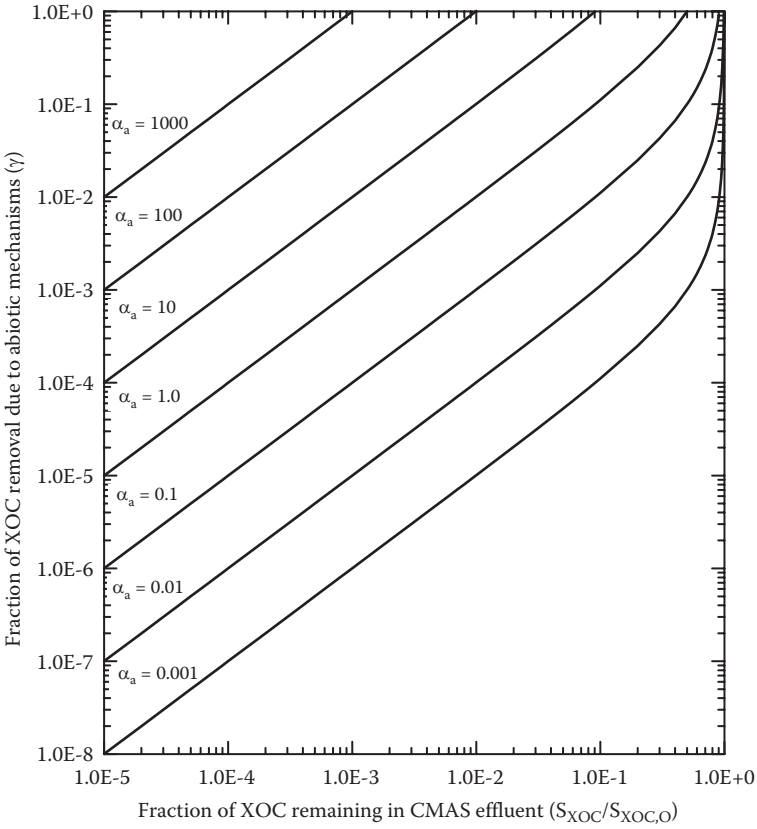

FIGURE 22.1 Graphical presentation of Equation 22.19 showing the fraction of XOC removal due to abiotic mechanisms as a function of the fraction of the XOC remaining in the effluent from a CMAS bioreactor. The abiotic loss coefficient, α_a, is defined in Equations 22.20–22.22. (From Grady Jr., C. P. L., Magbanua, B. S., Brau, S., and Sanders II, R. W., A simple technique for estimating the contribution of abiotic mechanisms to the removal of SOCs by completely mixed activated sludge. *Water Environment Research,* 69:1232–37, 1997. Copyright © Water Environment Federation, Alexandria, Virginia. Reprinted with permission.)

di-*n*-butyl phthalate is highly sorptive, but not very volatile. In the same lab-scale CMAS system, the value of α_v was 0.048, whereas the value of α_s was 0.229.[28] Thus, in that case, 82.7% of the abiotic losses were due to sorption. Thus, it can be seen that the relative importance of the abiotic mechanisms depends strongly on the physicochemical properties of the XOC as well as on the nature of the CMAS system.

22.4 EFFECTS OF XENOBIOTIC ORGANIC CHEMICALS

So far we have focused on the fate of XOCs in bioreactors. However, XOCs can also affect the performance of a bioreactor by retarding the removal of biogenic organic matter and nitrification. This can occur through inhibition and toxicity. Inhibition is when the presence of a chemical reduces the rate of a microbial process, such as growth and substrate utilization. Toxicity is when the presence of a chemical causes all microbial activity to cease. Inhibition increases as the concentration of an inhibitor increases and toxicity occurs when the concentration of the inhibitor becomes sufficiently high to stop microbial activity. Currently, inhibition is treated by environmental engineers as if it were fully reversible, but whether this is generally true is unclear. Toxicity, on the other hand, is usually considered to be irreversible, although little evidence exists concerning this either.

22.4.1 Mechanisms and Models for Inhibition and Toxicity

The mechanisms by which XOCs inhibit microbial growth and substrate removal are not well defined and have only recently become of interest. Consequently, our knowledge of the subject is rather limited. Nevertheless, one would think that the mechanisms are probably the same for inhibition and toxicity, with reaction rates becoming progressively slower as more damage occurs until ultimately the cumulative damage is sufficient to disrupt all activity and we say that toxicity has occurred.

Some XOCs have very specific effects on microbial cells, whereas others have a more general or nonspecific effect. Among the specific effects are radical forming reactions by transition metals, forming hydroxyl radicals that react with a broad range of macromolecules; reactions of both organic compounds and metals with thiol groups in enzymes, changing their conformation, and thus their reaction rates; formation of covalent bonds with amino acid side chains in enzymes, altering the conformation of the active site; and interference with protein synthesis. In addition, XOCs that are analogs of biogenic organic compounds can bind irreversibly with the active site on an enzyme, blocking its activity.

A good example of a nonspecific effect is provided by the action of hydrocarbons.[62] Many hydrocarbons are amphiphilic; that is, they contain both hydrophobic and hydrophilic moieties. As such they behave like the phospholipids that form cellular membranes, making them soluble in those membranes. As hydrocarbons dissolve in the membrane, they disrupt its structural integrity, thereby interfering with its major functions, which are to serve as a barrier separating the cytoplasm from the environment, provide for energy transduction, and provide spatial organization for certain enzymes. At low concentrations, an alteration of these functions merely reduces the activity of the cell. However, as the concentration of the hydrocarbon in the environment is increased its concentration in the membrane also increases, thereby causing greater disruption and a greater effect. Ultimately, when the concentration is sufficiently high, the cell stops functioning.

Models for inhibition quantify the effects by altering the values of $\hat{\mu}$ and K_S associated with carbon oxidation or nitrification. Although several have been proposed, the most general is that of Han and Levenspiel,[32] which was given as Equation 3.42:

$$\mu = \hat{\mu}\left(1-\frac{S_i}{S_i^*}\right)^n\left[\frac{S_S}{S_S+K_S\left(1-\frac{S_i}{S_i^*}\right)^m}\right]. \tag{3.42}$$

In this equation, S_i^* is the concentration of the inhibitor (i.e., the XOC) causing all microbial activity to cease. In other words, it is the toxic concentration. The magnitudes of the coefficients m and n can be altered to represent the type of inhibition occurring. Four types are generally modeled,[34,56,72] and they are defined in Table 22.1 in terms of their effects on $\hat{\mu}$ and K_S. Also shown are the characteristics m and n must assume to model the effects. Identification of the inhibitor type is important

TABLE 22.1
Types of Inhibition

Inhibition Type	Effect on $\hat{\mu}$	Value of n	Effect on K_S	Value of m
Competitive	None	0	Increase	<0
Noncompetitive	Decrease	>0	None	0
Uncompetitive	Decrease	>0	Decrease	>0
Mixed	Decrease	>0	Increase	<0

because it determines the manner in which the substrate and inhibitor concentrations interact in regulating substrate removal and biomass growth, thereby governing the response of a bioreactor to an inhibitory shock load.[59] For example, if an inhibitor acts in a competitive manner, it will have no effect when the substrate concentration is high because it does not effect $\hat{\mu}$. On the other hand, a mixed inhibitor is the worst type because it will have a negative effect regardless of the substrate concentration.

22.4.2 Effects of Xenobiotic Organic Chemicals on Carbon Oxidation and Nitrification

Many studies have been conducted in which the concentration of an XOC causing a 50% reduction in microbial activity (IC_{50}) was quantified and these have been tabulated for a variety of microbial groups, including aerobic heterotrophs, nitrifiers, and methanogenic bacteria.[10] Tables of IC_{50} values are very helpful for comparing the relative inhibition caused by different XOCs in a semiquantitative manner and they have demonstrated that nitrifying bacteria are much more sensitive than either of the other groups to a broad range of XOCs, with IC_{50} values being as much as two-orders of magnitude lower. Furthermore, they have also shown that methanogens are sometimes more sensitive than aerobic heterotrophs but not always. The disadvantage of IC_{50} values is that they cannot be translated directly into a quantitative effect on the kinetic parameters in the Monod equation.[72] Thus, they cannot be used to predict the effect of an inhibitor on the performance of a bioreactor. Unfortunately, relatively few studies have been done to quantify the effects of XOCs on the kinetic parameters in the Monod equation. Procedures are now available by which this can be done.[43,73]

Volskay et al.[74] studied the effects of 14 XOCs on the kinetics of carbon oxidation by aerobic heterotrophic bacteria. Three of them were found to be uncompetitive inhibitors, one to be a noncompetitive inhibitor, and 10 to be mixed inhibitors. Mixed inhibitors are the worst type, because they decrease the value of $\hat{\mu}$ and increase the value of K_S, both of which cause the substrate concentration associated with a given SRT in an activated sludge bioreactor to increase. The values of the toxic XOC concentration, S_i^*, were relatively high, however, with the smallest measured value being that for tetrachloroethylene at 126 mg/L. This suggests that XOCs must be present at relatively high concentration to have a deleterious effect on aerobic heterotrophs, although mixtures of XOCs will have an additive effect so that the total concentration is what is important.[31] Volskay et al.[74] also found that the values of m and n in Equation 3.42 were not equal to 1.0, and thus the effects of the XOC on the kinetics of carbon oxidation were nonlinear. Furthermore, in some cases the absolute values of m and n were greater than 1.0, suggesting that the effect increases more and more as the XOC concentration is increased.

Kim et al.[41] studied the effects of three chlorinated phenols and three chlorinated anilines on aceticlastic methanogens and their results were similar to those in the preceding paragraph. All six compounds acted as mixed inhibitors, all had nonlinear effects, and all had relatively high toxic concentrations. The major difference was that the absolute values of m and n were larger (generally greater than 1.0), meaning that the effects increased drastically as the concentration of the XOC was increased.

Most studies on inhibition of nitrification have resulted in IC_{50} values rather than in specific effects on $\hat{\mu}$ and K_S. Nevertheless, there is a general perception in the environmental engineering literature that XOCs inhibit nitrification in a noncompetitive manner. This perception arises in part from studies that reached that conclusion even though the manner in which the experiments were run did not provide the type of data required to justify it.[54,69] Although others have found that a variety of XOCs do impact nitrification in a noncompetitive manner,[4,24,45] the amount of data available do not justify a general conclusion about the nature of the inhibition associated with XOCs. However, it is clear that the oxidation of ammonia-N is much more sensitive to XOCs than is oxidation of nitrite-N.[35] Thus ammonia oxidation must be considered to be the weak link in nitrification.

22.5 EXPERIENCE WITH XENOBIOTIC ORGANIC CHEMICALS

We have about a hundred years experience in the design and operation of biological systems for the removal of organic matter from domestic wastewater. As a result, we have had the opportunity to try many things, some of which worked and some of which did not. Nevertheless, through that experience and the research that it fostered, we have been able to establish the many fundamental principles that are the basis for this book. Consequently, a designer who applies them can be confident that a proposed design will succeed. Our experience with biological nutrient removal systems, on the other hand, dates back almost 50 years, to the early 1960s. Since then we have learned much about them, allowing generalizations to be made in the form of the International Water Association activated sludge models, which are facilitating further research and development. While our experience with nutrient removal systems is not as old as that for systems focused on removal of organic matter, designers can still have reasonable confidence that a proposed system will function as planned, particularly if its design has been based on pilot studies. In contrast to the above, our experience with XOCs in biological wastewater treatment systems is relatively recent, dating from the late 1970s and early 1980s. This is not to suggest that such chemicals were not present in wastewaters prior to that date. Rather, that was when we began to be concerned with the discharge of individual compounds in treated effluents rather than just with chemical or biochemical oxygen demand. More recently, we have begun to study the fate of trace XOCs in biological processes, and have started to elucidate operating conditions that enhance their overall removal. Nevertheless, our knowledge of how to design a biological process to enhance trace XOC removal is largely undeveloped. Thus, even though considerable effort has been expended by many researchers, it is not yet possible to reach broad generalizations like those presented in the preceding chapters.[26,27]

Publicly owned treatment works normally receive industrial XOCs at low concentrations, typically in the μg/L range. Industrial wastewater treatment facilities, on the other hand, often receive them at high concentrations, reaching to hundreds of mg/L. In spite of this broad range, biological treatment systems have been found to be very robust and to do a very good job removing many types of XOCs to low levels.[26,27] Beyond that, it is difficult to generalize. This difficulty stems from the fact that the ability of a particular bioreactor system to remove a given XOC depends strongly on the configuration of the system, the physicochemical characteristics of the XOC, and the XOC's biodegradation kinetics. In spite of the complexity this suggests, our experience with XOCs is consistent with the principles set forth in the preceding sections of this chapter. For example, compounds like chloroform, trichloroethane, dichlorobenzene, toluene, xylene, and ethylbenzene have high Henry's Law coefficients and low octanol:water partition coefficients. Consequently, volatilization is a more important removal mechanism in the activated sludge process than sorption.[55] Furthermore, chlorine is a more powerful xenophore than methyl and hydroxyl groups, and thus the halogenated volatile compounds are removed more by volatilization than by biodegradation, whereas the reverse is true for the nonhalogenated ones.[55] On the other hand, polycyclic aromatic hydrocarbons have low Henry's Law coefficients, high octanol:water partition coefficients, and low rates of biodegradation. Thus, most of their removal is due to sorption onto solids.[49] Finally, pesticides like diazinone, 2,4-dichlorophenoxyacetic acid, and 2,4,5-trichlorophenoxyacetic acid tend to be very resistant to biodegradation, and have low Henry's Law coefficients and low octanol:water partition coefficients. Consequently, they pass through activated sludge systems with little removal.[51] These examples make it clear that the biodegradability and physicochemical characteristics of an XOC determine its fate. For simple CMAS systems, Equation 22.19 and Figure 22.1 can assist an engineer in determining what that fate is likely to be. More complex systems require the use of models.[25,50]

The recent emphasis on trace XOCs, comprising pharmaceuticals, personal care products, and household chemicals, came from their broad presence in surface[42] and drinking waters.[68] Indeed, this experience has emphasized that wastewater treatment is the first step in drinking water treatment.

Since wastewater treatment plants have not been designed with trace XOC removal as a goal, most research to date has focused on fate studies that show which compounds are removed through bioreactors, and what bioreactor conditions correlate with higher levels of removal. While many trace XOCs are biodegradable, especially under aerobic conditions, few studies have shown if loss of the parent compound correlates with its mineralization or whether degradation products expressing biological activity remain. Some compounds, like carbamazepine and chlorinated organophosphorus flame retardants, are highly resistant to biotransformation and are routinely found in the effluents of biological treatment processes.[8,15,57] Of those trace XOCs that have been shown to biodegrade during biological treatment, the best removals tend to correlate with nitrifying activated sludge[14,67] and biological nitrogen removal systems[2] that have sequential aerobic and anoxic zones. Nitrifiers have been shown to transform some trace XOCs,[61,63,77] but the relative roles of autotrophic and heterotrophic bacteria in trace XOC removal is currently uncertain.[40] Many trace XOCs are polar, although not highly so. Some that are less polar, such as brominated flame retardants and musk fragrances, sorb to biomass.[53,66] Unlike industrial XOCs, however, octanol:water partition coefficients are not good predictors of trace XOC sorption to biomass,[66] suggesting that their sorption is influenced by other physicochemical features. Some trace XOCs, such as musk fragrances and organophosphorus flame retardants,[57] have moderate Henry's Law coefficients and can be stripped by aeration. However, most trace XOCs have very low Henry's Law coefficients and are not removed by volatilization.

The type of bioreactor employed will also influence the fate of XOCs. For example, trickling filters are not as capable of removing nonvolatile XOCs as activated sludge systems, although they have equal capability for the removal of volatile XOCs.[33] This suggests that volatilization may be an important removal mechanism in trickling filters, which one would expect given their configuration. Although experimental studies on large-scale facilities have not confirmed it, modeling studies have suggested that attached growth bioreactors should be more resistant to inhibitory, but biodegradable, XOCs than suspended growth systems.[23,65] This is because the inner layers in a biofilm experience a lower substrate concentration than the outer layers, which gives a higher reaction rate for an inhibitory substrate. Thus, the overall reaction rate will be higher than the bulk substrate concentration would suggest.

Even within a given type of named biochemical operation, bioreactor configuration can have important effects. Tanks-in-series or plug-flow activated sludge systems have concentration gradients in them whereas CMAS systems are uniform throughout. As a consequence, volatilization is less important as a removal mechanism in the latter.[7]

In summary, although we have not been concerned about individual XOCs for very long, our knowledge base is growing rapidly. Furthermore, the things that we are learning are consistent with the principles of bioreactors presented throughout this text. By applying those principles in practice, progress in this important area should be rapid and we should soon be able to design biological wastewater treatment systems for XOC removal with much more confidence.

22.6 KEY POINTS

1. The ability of microorganisms to biodegrade a xenobiotic organic chemical (XOC) is the result of two things. First, most XOCs have some structural similarity to biogenic materials, allowing them to fit into active sites on enzymes that did not evolve with them as substrates. Second, the specificity of enzymes is not exact.
2. The molecular structure of an XOC is the primary determinant of its biodegradability, but factors such as solubility, the presence of molecular oxygen, and the availability of appropriate nutrients and electron acceptors have strong effects on whether biodegradation can occur.
3. When biodegradation occurs by growth-linked metabolism, the rates of biomass growth, substrate removal, and oxygen utilization are all linked and are all first-order with respect

to the concentration of biomass actually involved in the biodegradation. In other words, everything is the same as growth on soluble chemical oxygen demand (COD) as depicted throughout this text.

4. The modeling of cometabolic biodegradation requires two separate reactions, growth on the carbon and energy source and the cometabolic transformation. Growth on the carbon and energy source is done in exactly the same way as any other growth-linked process. The cometabolic transformation is usually modeled as an enzymatic reaction using either the Michaelis-Menten or Haldane equation. It may also be necessary to consider competition between the growth substrate and the cometabolic substrate for the shared enzyme, utilization of the growth substrate to provide reducing power, and inactivation of the cells.
5. Because of the physicochemical properties of some XOCs, abiotic removal mechanisms such as volatilization to the atmosphere and sorption onto solids may contribute to their loss from a bioreactor. They permit the XOCs to enter the environment without alteration and thus their contribution should be minimized.
6. The mass transfer coefficient for loss of an XOC from a bioreactor by volatilization can be estimated from knowledge of the mass transfer coefficient for oxygen, the diffusivity of the XOC, and the power per unit volume expended for oxygen transfer.
7. The sorption coefficient for loss of an XOC from a bioreactor through biomass wastage can be estimated from knowledge of the sorption coefficient for a similar XOC on the same biomass by using the octanol:water partition coefficients for the two XOCs.
8. The effluent concentration of an XOC from a completely mixed activated sludge (CMAS) system is determined solely by the system solids retention time (SRT) and is independent of the abiotic removal mechanisms. The impact of the abiotic mechanisms is to reduce the concentration of the biomass capable of degrading the XOC. Consequently, the fraction of XOC removal due to abiotic mechanisms is equal to the fractional reduction in capable biomass concentration due to them.
9. Abiotic losses of an XOC from a CMAS system can be characterized by a dimensionless abiotic loss coefficient, α_a, which is the sum of the dimensionless abiotic loss coefficients for sorption and volatilization. Consequently, the effects of the two abiotic removal mechanisms are additive.
10. Xenobiotic organic chemicals can affect the performance of a bioreactor by retarding the removal of biogenic organic matter and nitrification. This can occur through inhibition and toxicity. Inhibition is when the presence of a chemical reduces the rate of a microbial process, such as growth and substrate utilization. Toxicity is when the presence of a chemical causes all microbial activity to cease.
11. The kinetic effects of inhibitors may be classified as competitive, noncompetitive, uncompetitive, or mixed, depending on how the inhibitors affect the $\hat{\mu}$ and K_S values describing carbon oxidation or nitrification. Identification of the inhibitor type is important because it determines the response of a bioreactor to an inhibitory shock load. The model of Han and Levenspiel (Equation 3.42) may be used as a general model for inhibition and toxicity.
12. Nitrifiers are much more sensitive to inhibition than are aerobic heterotrophic and methanogenic bacteria. Furthermore, ammonia oxidation is considerably more sensitive than nitrite oxidation.
13. Biological treatment systems have been found to be very robust and to do a very good job removing many types of XOCs to low levels. However, because of our relatively limited experience with XOCs and because the ability of a particular bioreactor system to remove a given XOC depends strongly on the configuration of the system, the physicochemical characteristics of the XOC, and the XOC's biodegradation kinetics, it is difficult to generalize about their fate and effects in bioreactors.

22.7 STUDY QUESTIONS

1. List and discuss the requirements for biodegradation of an XOC.
2. List and discuss the factors determining whether biodegradation of a given XOC will occur.
3. Explain the differences between growth-linked and cometabolic biodegradation and the implications of those differences to the modeling of each.
4. Explain why the rate expression for loss of an XOC by volatilization from a bioreactor using mechanical surface aeration for oxygen transfer is different from the rate expression for volatile losses from a bioreactor using diffused aeration.
5. Describe how to estimate the mass transfer coefficient for removal of an XOC by volatilization from knowledge of the mass transfer coefficient for oxygen.
6. Describe the two methods for estimating the coefficient describing sorption of an XOC onto the biomass in a bioreactor.
7. Explain why abiotic removal mechanisms become less important as the fraction of an XOC remaining in a CMAS effluent becomes smaller.
8. A CMAS bioreactor has a volume of 6667 m^3, receives wastewater at a flow rate of 40,000 m^3/day, and operates with a MLSS concentration of 2750 mg/L when the SRT is four days. Oxygen is transferred to it by mechanical surface aeration with a power input of 350 kW, which achieves an oxygen transfer coefficient of 10 h^{-1}. Two of the constituents in the wastewater are dichloromethane and diethyl phthalate, both of which are present in the influent at a concentration of 1.0 mg/L. The concentration of dichloromethane in the effluent is 5.0 μg/L while the concentration of diethyl phthalate is 15.0 μg/L. Sorption studies on the same MLSS with di-*n*-butyl phthalate gave a linear isotherm with a sorption coefficient of 0.98 L/g. Using physicochemical characteristics available in the literature, estimate the contributions of volatilization and sorption to the removal of dichloromethane and diethyl phthalate.
9. Define the four types of inhibition in terms of their effects on $\hat{\mu}$ and K_S and then describe the characteristics of the coefficients m and n in the Han and Levenspiel model required to model each type. What is the significance of a large absolute value of m or n?

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23 Designing Systems for Sustainability

The sustainability of our water and wastewater systems is becoming an important topic.[3,5] Fortunately the environmental engineering profession is developing processes, many of which are biological, that offer the potential for increased sustainability. These processes operate by the principles presented in previous chapters and, consequently, can be characterized and analyzed using those principles. In this chapter a number of biological processes will be discussed in terms of their relationship to sustainability. Before doing this, however, a brief background on sustainability will be provided.

23.1 DEFINING SUSTAINABILITY

23.1.1 The Context for Improved Sustainability

Why is sustainability an important issue? The simple answer is that growth of the human population, coupled with increased per-capita resource consumption, is stressing the natural environment and reducing the ability of Earth to provide the materials necessary for human life.[22,31]

23.1.1.1 Demographic Trends

The current human population is about 6.2 billion and its doubling time is about 40 years.[17,28] Fortunately, the growth rate is decreasing and the human population is projected to reach about 9 billion by 2050 and remain essentially constant for the rest of the century (low-range estimate peaking at 7 billion and then declining; high-range estimate increasing to 14 billion by the end of the century). Although plots of human population appear similar to bacterial growth curves, the underlying causes are different, with human population being governed by social and technological factors rather than resource depletion. This difference is critical because a stationary (or even declining) phase for the human population based on resource limitations would represent a miserable existence! An important point is that the projected leveling of human population is expected to result from declining birth rates in developing countries resulting from increases in the standard of living of the poor. However, an increased standard of living inevitably results in an increase in per-capita resource consumption. Therefore, the availability of sufficient resources is necessary for the development of a stable human population at a level that avoids significant human misery.

Population growth is occurring principally in developing countries and this trend is expected to continue. In contrast, the combined population of the developed countries is relatively constant. The native population growth rate of many developed countries is less than the replacement rate (i.e., less than 2.1 children for every couple of childbearing age), leading to a declining and aging population. This is the situation in Japan and some countries in Western Europe. In other developed countries immigration is leading to net population growth. This is the situation in the United States, where the population is expected to grow from a current population of 300 million to about 450 million by 2050. The net result is that, on average, the additional 3 billion people expected on Earth by 2050 will reside in developing countries.

Another important trend is urbanization.[17] The twentieth century can be viewed as the transition from a principally rural existence for humans to a principally urban existence. Significantly, the population growth throughout the twenty-first century is expected to reside in urban areas.

To summarize, 3 billion additional people are expected on Earth by about 2050 and they will reside in new or growing urban areas in developing countries. The population of some developed countries will increase, principally as a result of immigration, and these additional people will also reside in urban areas. The population of other developed countries will decline.

23.1.1.2 Resource Consumption

Increasing human population and improved standards of living will result in greater pressure on the resources provided by Earth. In the absence of changed behavior, demand will increase much more than the 50% increase in population. Already, evidence of emerging resource constraints, resource depletion, and human impacts is apparent and growing. Issues include deforestation, declining fish populations, climate change, and depletion of natural resources such as fossil fuel and other necessary materials (such as phosphate). Of particular importance to water professionals is the fact that currently about 0.5 billion people live under conditions of water stress. Furthermore, by 2025 it is expected that 2 billion people will live under conditions of water scarcity, and 4 billion will live under conditions of water stress.[3] This illustrates the task before us!

23.1.1.3 Sustainable Development

Infrastructure must be constructed to meet the basic needs of a growing population—development will occur. However, to prevent further environmental degradation it must occur in a fashion that preserves, and ideally enhances, the ability of Earth to provide the resources necessary for life, both human and nonhuman. This forms the basis for the concept of sustainable development. It is a positive approach that suggests that, if development is going to occur, it can be used to enhance human existence.

23.1.2 The Triple Bottom Line: Social, Economic, Environmental

Understanding the concept of sustainable development is one thing. It is quite another to be able to use it to evaluate development options. One view of sustainability, referred to as absolute sustainability, focuses on developing recycling systems, resulting in a constant and sustaining stock of all materials necessary for human life. This view of Earth as a "spaceship," where no net waste can be produced, has several flaws. First, it is focused on ensuring that "things do not get any worse." Second, it is a conservative view, inherently assuming that future scenarios will differ little from current ones. Third, by focusing on technological solutions it neglects the importance of social systems in achieving sustainability. Fourth, there are no accepted measures of absolute sustainability, and work on the development of such measures suggests that a true measure will be quite difficult to implement. This greatly restricts the applicability of this type of analysis for practical engineering applications. Another concept is relative sustainability, which compares options and seeks to implement the one that is the most sustainable. It is viewed by some as simply slowing the rate of decline—allowing things to get slowly worse. However, its application can result in consistent improvement as options that improve sustainability are selected and implemented. For this approach to be useful, however, the relative sustainability of various options must be measurable in a useful fashion. This is possible, as will be described below.

This brings us to the concept of the triple bottom line (TBL). This phrase implies that business must focus not only on the conventional bottom line of profitability, but also on a broader set of objectives encompassing social and environmental issues.[31] Another useful concept from the business literature is the "license to operate," which states that in order for a business to have a "social license" to function, it must not only be profitable but also contribute to society and environmental protection. These concepts have been adopted by public and private entities, both for-profit and not-for-profit. Increasingly, all organizations must consider their triple bottom line to earn their license to operate.

The TBL concept is useful for developing measures of relative sustainability. The most sustainable option is the one that best balances (optimizes the mix of) the relevant social, economic, and environmental considerations, as illustrated in Figure 23.1. The term relevant applies to a specific

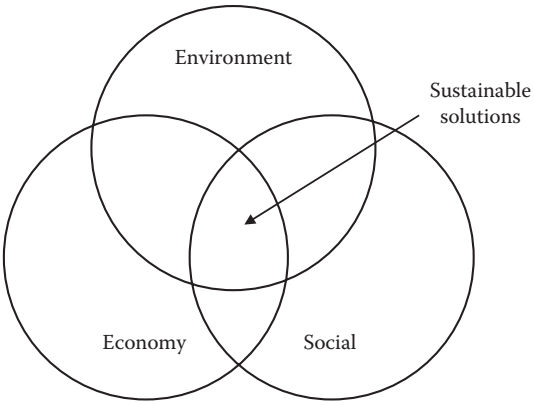

FIGURE 23.1 Triple bottom line assessment of relative sustainability.

situation; that is, those factors that are specifically affected by the given decision to be made. Only those factors contributing to the relative sustainability are included in the analysis. This is much easier than measuring absolute sustainability, allowing the analysis procedure to be applied to a much wider range of decisions. But, difficulties remain. Social and broad environmental impacts of alternative decisions must be assessed in some relevant fashion, even if only on a relative basis. Therefore, tools for making such assessments must be available. Environmental engineers are well equipped to determine the costs and performance capabilities of particular environmental control measures, but are less capable of assessing the broader economic and environmental impacts, much less the social impacts. Thus, environmental engineers must engage other environmental, economic, and social professionals in their work. Procedures using semiquantitative measurement (utility) scales have been developed and are beginning to be used routinely. Referred to by various terms, such as multicriteria analysis and decision analysis, they use quantitative and semiquantitative utility scales coupled with site-specific weighting factors to reflect the relative importance of each individual criterion to overall sustainability. Table 23.1 provides some factors considered in these procedures, along with measures that can be applied. These factors are subsequently weighted based on the values of those affected by the decision to allow an overall assessment to be computed. Specific details on the application of these approaches are provided elsewhere.[6]

One further comment about sustainability concerns the importance of systems rather than individual technologies. Systems, not individual technologies, produce sustainable results. Therefore, when considering the impacts on sustainability of a particular environmental control option, the impacts of the entire set of actions required to implement the option (i.e., the system) must be considered, not just an individual technical component in it. Use of a highly efficient and effective technical component does not make an entire system sustainable, especially if the use of that component also requires the use of another, highly polluting, component. Likewise, the broader impacts of a particular option must be considered. An option that produces a high quality aqueous effluent may not achieve a sustainable result if it transfers contaminants to the air or soil. The composite performance of systems and their broad and systematic impacts must be considered when developing and subsequently evaluating their sustainability.

In summary, the following factors are important considerations for the measurement of relative sustainability:

- Relative sustainability is achieved by measuring the relative contribution of a particular action or option to the triple bottom line consisting of social, economic, and environmental impacts.
- The relative sustainability of a variety of options can be measured using weighted semiquantitative utility scales.

TABLE 23.1
Example Factors That Can Be Used to Quantify Sustainability and Specific Measures of Them

Factor	Quantification Method
Social	
Public health protection	Extent of exposure to specific public health hazard
Creation of public amenities	Number of visitors or monetary value of amenity created
Job creation	Number of local jobs created
Economic	
Ability to pay	Average cost compared to median income
Water supply created	Volume or economic value (based on available alternatives)
Economic development	Value in dollars
Environmental	
Carbon footprint	CO_2 equivalents
Water resource left in environment	Volume and some measure of value in terms of enhanced habitat (e.g., increased fishery productivity)
Resource protection	Amount of resource protected

- The impacts (both positive and negative) of the entire system must be considered. Sustainability is about assessing the impacts of entire systems.
- The relative importance of social, economic, and environmental impacts (both positive and negative) is determined by the relative values of those impacted by the system to be implemented. This means that the selection of alternatives is site-specific. There is no "universal" solution to a particular problem—different solutions may be appropriate to essentially similar problems implemented in different localities.

23.1.3 Technical Objectives for More Sustainable Systems

A key role that environmental engineers play in sustainable development is to conceive of water management systems (both domestic and industrial) with superior performance characteristics. While such systems must be subjected to TBL analysis before their overall contribution to sustainability can be determined, the characteristics of systems with the potential to contribute to increased sustainability can be identified. These characteristics can then be used to identify and develop individual technologies that can be part of such systems. Three characteristics are readily recognized: (1) achieving greater water resources availability, (2) lowering energy and chemical consumption, and (3) recovering resources. They form the basis for the discussion of relevant and evolving biological treatment technologies in the remainder of this chapter. Before beginning this discussion, the efficiency and effectiveness of product replacement, source separation, and source control should be emphasized. If contaminants are never added to wastewater to begin with, control technologies for their removal are not needed, decreasing the resources needed for treatment, and reducing water use. Such approaches can be applied in both industrial and domestic settings. While product replacement, source separation, and source control are not the subject of this book, the reader should be aware of them and use them in preference to treatment wherever possible.

23.1.3.1 Greater Water Resource Availability

Population growth and increased living standards result in increased water demands, not only for domestic consumption but also for industrial and agricultural use. Unless innovative solutions are developed, significant portions of the human population will be faced with water scarcity in

the foreseeable future. Unfortunately, although water is a renewable resource, increased water scarcity demonstrates that the natural hydrologic cycle is insufficient to meet the increasing needs imposed by human activities. Therefore, human intervention is needed to increase the availability of water supplies. Water conservation, along with more efficient water management systems, is one component of the solution to the growing water resource challenge, but it alone is insufficient. Fortunately, water reclamation and reuse are available to recycle water for various purposes.

Water reclamation and reuse is an established practice in a wide variety of venues, such as in industry where it is coupled with water conservation.[1,9] Likewise, domestic wastewater is reclaimed for a variety of uses, including industry, agriculture, energy production, urban irrigation, and water supply supplementation. Reclamation and reuse of potable water can be classified as either planned or unplanned, depending on how it has been developed and implemented. It can also be classified as direct or indirect, depending on whether the reclaimed water is introduced into a water distribution system or into a water supply reservoir or drinking water aquifer. Policies and procedures for these reuse options are well established and treatment technologies are available. Biological treatment is a key component of essentially any reclamation and reuse system for wastewaters containing biodegradable organic matter.

23.1.3.2 Lowering Energy and Chemical Consumption

As increased levels of wastewater treatment have been applied to promote human health, protect the environment, and facilitate water reclamation and reuse, the associated consumption of energy and chemicals has increased. This increased consumption competes with consumption of resources for all other human uses. Therefore, the development of wastewater treatment systems that consume less energy and chemicals is essential. Fortunately, relevant biological treatment technologies are available and in development.

23.1.3.3 Recovering Resources

Resource recovery represents a further response to increased resource consumption. In our case, necessary resources are recovered from wastewater streams. A wide variety of useful resources can be recovered from wastewater streams, including energy and nutrients. The opportunity to extract biodegradable polymers from wastewater is also being evaluated.

23.2 TECHNOLOGIES TO ACHIEVE GREATER WATER RESOURCE AVAILABILITY

Greater water resource availability is achieved whenever we avoid contamination of surface or groundwater that can be used as an agricultural, industrial, or domestic water supply. Thus, wastewater treatment itself is a key element in achieving greater water resource availability. Water reclamation and reuse is a further practice, with well-developed and demonstrated technologies. Three technologies are increasingly contributing to water resource protection and water reclamation and reuse: (1) membrane bioreactors, (2) biological nutrient removal, and (3) advanced oxidation coupled with biological activated carbon. Key features of these technologies and their potential to contribute to increased sustainability are summarized in Table 23.2.

23.2.1 Membrane Bioreactors

Membrane bioreactor (MBR) technology is increasingly being used to treat wastewater and as a key component of water reclamation and reuse systems.

23.2.1.1 Technology Description

Membrane bioreactor activated sludge (MBRAS) is a variation of the activated sludge process in which a membrane system is used for liquid-solids separation. As discussed in Chapters 10 and 11, the design of MBRAS systems is based on the same principles as any other activated sludge process. Biological nutrient removal (BNR) can also be incorporated into MBRAS processes, as discussed

TABLE 23.2
Key Features and Contributions to Sustainability of Technologies to Achieve Greater Water Resource Availability

Technology	Key Features	Contribution to Sustainability
Membrane bioreactors	• Long SRT • High reliability • Inherent automation • Superior removal of particulate and colloidal matter	• Production of high quality effluent • Excellent pretreatment if further treatment required • Facilitates distributed systems
Biological nutrient removal	• Efficient and cost-effective removal of nitrogen and phosphorus	• Efficient protection of surface waters from eutrophication
Advanced oxidation	• Direct oxidation of some recalcitrant organic compounds • Increased biodegradability of recalcitrant organic compounds	• Mineralization of organic compounds of concern • Chemical alteration of organic compounds to products of reduced concern

in Chapter 12. With a bioreactor sized to meet process volumetric oxygen transfer requirements and with mixed liquor suspended solids (MLSS) concentrations greater than 6 g/L (often in the 8 to 10 g/L range), solids retention times (SRTs) in excess of 10 days can easily be maintained in MBRAS systems. Long SRT values also contribute to good sludge filterability, further resulting in relatively complete biodegradation of organic material in MBRAS.

23.2.1.2 Contribution to Sustainability

The features of MBRAS allow it to be incorporated into treatment systems that contribute to increased water resource availability. Long SRTs, along with the superior control of particulate matter resulting from the use of membrane filtration for liquid-solids separation, result in a process that reliably produces excellent effluent quality. The long SRT generally results in near complete metabolism of biodegradable organic matter, including slowly biodegradable trace contaminants, and nitrification.[27] The benefits of longer SRTs to the removal of a wide variety of xenobiotic compounds (including endocrine disruptors and pharmaceuticals/personal care products) are discussed in Section 22.1.1. The superior control of particulate and colloidal matter by MBRAS, even in comparison to long SRT activated sludge processes followed by granular media filtration, contributes to the removal of adsorbable trace constituents associated with effluent particulate and colloidal matter.[10]

The superior removal of particulate and colloidal matter by the membrane filters used in MBRAS systems contributes to superior control of pathogens, both by direct filtration and because many pathogens (especially viruses) adsorb onto the MLSS, which is subsequently removed by the membrane filters. The low colloidal matter content of the effluent further facilitates disinfection by conventional approaches such as chlorination, ultraviolet (UV) light, and ozonation. Likewise, membrane filtration is a superior pretreatment technique when the biological process effluent is to be subjected to further treatment, such as by reverse osmosis.[25,26] Such downstream treatment is often needed when a high quality, near-potable quality water is to be produced.

Another important feature of MBRAS relates to the ease with which it can be automated. This feature, coupled with the highly reliable performance it provides, results in a biological treatment process that can be remotely deployed. This has led to increased use of MBRAS in satellite water reclamation and reuse systems. Various system configurations can be used. In one, commonly referred to as “sewer mining,” an MBRAS process is located adjacent to a wastewater interceptor

at a location where uses for reclaimed water exist. Wastewater is withdrawn from the interceptor as needed to meet the reclaimed water demand and treated in the MBRAS. Solids produced in the treatment process are returned to the interceptor for conveyance to the downstream main wastewater treatment plant.

23.2.2 Biological Nutrient Removal

Biological nutrient removal contributes to increasing water resource availability by protecting surface waters.

23.2.2.1 Technology Description

Biological nutrient removal processes, discussed extensively in Chapter 12, incorporate anoxic and anaerobic zones into aerobic suspended growth biological wastewater treatment processes to remove nitrogen and phosphorus. Submerged attached growth bioreactors (SAGBs) can also be used for nitrogen removal, as discussed in Chapter 21. Packed bed SAGBs can contribute to phosphorus removal through increased removal of particulate matter and the phosphorus associated with it.

23.2.2.2 Contribution to Sustainability

Historically the focus of surface water quality protection was on removal of biodegradable organic matter and the oxidation of ammonia. By reducing the discharge of these oxygen demanding materials, oxygen depletion in the receiving water was minimized. However, for many receiving waters this was not sufficient to prevent water quality deterioration because the discharge of nitrogen and phosphorus resulted in excess algal growth. Algae produce oxygen by photosynthesis during the day, but they respire at night causing oxygen depletion. Algal photosynthesis also fixes carbon dioxide and produces biodegradable organic matter (i.e., the algae), which exerts an oxygen demand when the algae subsequently die and degrade. This oxygen demand can be substantial. For example, we saw in Table 3.1 that the oxygen demand associated with the biodegradable organic matter in domestic wastewater is on the order of 2 kg O_2/kg organic matter, whereas that associated with ammonia-N is 4.6 kg O_2/kg NH_3-N. In contrast, the algae produced from the discharge of 1.0 kg of nitrogen can exert 14 kg of oxygen demand, while that produced from the discharge of 1.0 kg of phosphorus can exert 100 kg of oxygen demand. No wonder nutrient removal is being required in so many water bodies! Algae also produce a variety of products that adversely affect the suitability of water for consumption.

The development of BNR processes has made the removal of nitrogen and phosphorus from wastewaters more feasible and economical, thereby allowing its widespread application. Biological nutrient removal has become a standard treatment technology for wastewaters discharging to sensitive receiving streams, many of which are also used as water supplies.

23.2.3 Advanced Treatment Coupled with Biodegradation

While most of the organic contaminants in wastewaters can be removed by biological processes, some cannot because they are not biodegradable. Nevertheless, these nonbiodegradable contaminants must often be removed before the treated effluent can be used to supplement available water supplies. A variety of physical-chemical processes are available for doing this, including: activated carbon adsorption; advanced oxidation using ozone, ozone/hydrogen peroxide, Fenton's reagent (iron with either hydrogen peroxide or ozone), and UV light with either ozone or hydrogen peroxide; and membrane processes such as nanofiltration or reverse osmosis. The reader is referred elsewhere for a further discussion of the direct physical-chemical treatment of biological effluents.[1,27] However, it should be noted that advanced oxidation processes need not completely destroy organic contaminants to be beneficial; rather, they can be used to increase their biodegradability by partial oxidation.[11,12]

23.2.3.1 Technology Description

If a wastewater contains a high proportion of nonbiodegradable organic matter, advanced oxidation can be used prior to a biological process to increase its effectiveness. This is most commonly done with specific wastewater streams that can be segregated at complex industrial facilities. More commonly, advanced oxidation is applied downstream of biological treatment to oxidize recalcitrant organic compounds and/or to increase their biodegradability. In such situations the effluent from the advanced oxidation process must be further treated to remove the biodegradable organic compounds produced. Biological activated carbon (BAC) is often used for this purpose.[27] Biological activated carbon is a conventional activated carbon system in which backwashing and carbon regeneration are managed to allow a biofilm to develop on the carbon. Organic compounds are removed by adsorption, but the biofilm oxidizes the biodegradable ones, thereby regenerating the activated carbon in-situ. The combination of advanced oxidation (often using ozone) followed by BAC is being used more frequently in advanced water treatment applications to remove residual organic matter.

23.2.3.2 Contribution to Sustainability

Advanced oxidation processes contribute to sustainability by allowing recalcitrant organic compounds to be mineralized, either by complete oxidation or by modification of their structure so that they become biodegradable. The organic compounds that are the targets for advanced oxidation are often toxic, carcinogenic, endocrine disrupting, or pharmaceutically active. Consequently, their mineralization or conversion to more benign forms eliminates significant environmental and public health concerns. For many compounds, advanced oxidation coupled with biodegradation is superior from both a cost and a resource consumption perspective.

23.3 TECHNOLOGIES TO ACHIEVE LOWER ENERGY AND CHEMICAL CONSUMPTION

Reduction of the energy requirements for wastewater treatment is desirable both because global energy resources are limited and because energy production from fossil fuels releases carbon dioxide, a greenhouse gas (GHG), which contributes to climate change. Energy consumption can be reduced by examining the entire water/wastewater system. For example, distributed water reclamation and reuse systems reduce pumping costs compared to conventional centralized systems because neither the wastewater nor the reclaimed water needs to be conveyed as far.[4,5] Numerous technologies to reduce energy and chemical consumption are in development. Microbial fuel cells represent a potentially transformational technology whereby the organic matter in wastewater is used to produce electrical energy.[16] Although this technology is in the early research phase, the reader is urged to follow its development. Reduced chemical consumption is desirable because most treatment chemicals are nonrenewable. This section discusses four technologies that can lower energy and chemical consumption: (1) anaerobic treatment, (2) biological nutrient removal, (3) nitritation and denitritation, and (4) biological air treatment. Key features of these technologies and their potential to contribute to increased sustainability are summarized in Table 23.3.

23.3.1 Anaerobic Treatment

Anaerobic treatment is an established technology for industrial wastewater treatment and it is attracting increasing interest for municipal wastewater treatment, especially in tropical climates. Anaerobic digestion is well established for stabilizing residual sludges and advanced digestion technologies are being developed to achieve increased volatile solids destruction, resulting in increased methane production. These systems also offer increased pathogen inactivation. Cell lysis technologies are also being investigated to increase volatile solids destruction in digestion systems.[21]

TABLE 23.3
Key Features and Contributions to Sustainability of Technologies to Achieve Lower Energy and Chemical Consumption

Technology	Key Features	Contribution to Sustainability
Anaerobic treatment	• Biological stabilization of organic matter to CH_4 rather than CO_2 • Enhanced biodegradability of particulate matter (for advanced digestion and cell lysis technologies)	• Energy required to transfer oxygen in aerobic process is avoided • Production of energy source (CH_4) that can replace fossil fuel
Nitritation and denitritation	• Conversion of ammonia-N to nitrite-N, rather than nitrate-N • Autotrophic denitrification with ammonia-N serving as electron donor and nitrite-N as electron acceptor • Short SRT for nitritation, approaching the minimum aerobic SRT • Very long SRT for autotrophic denitritation • Generally applies to streams with high ammonia concentration	• Process oxygen requirements reduced 25% for nitritation compared to nitrification • Carbon requirements for heterotrophic denitritation reduced by 40% compared to heterotrophic denitrification • Process oxygen requirements reduced 62.5% for nitritation and autotrophic denitritation compared to conventional nitrification and heterotrophic denitrification; carbon requirements eliminated
Biological nutrient removal	• Efficient and cost-effective removal of nitrogen and phosphorus	• Reduced or eliminated carbon requirement for nitrogen removal • Reduced process oxygen and alkalinity requirements for predenitrification • Reduced or eliminated chemical requirements for phosphorus removal
Biological air treatment	• Biological removal of a wide variety of constituents from air streams	• Cost-effective and efficient alternative to chemically intensive systems • Facilitates location of treatment facilities in optimal locations by addressing air emission issues

23.3.1.1 Technology Description

The principles of anaerobic treatment are presented in Chapters 8 and 14, and commonly used systems are described in Chapters 14 and 15. Anaerobic treatment is a well-accepted and commonly used process for the treatment of many industrial wastewaters. Anaerobic lagoons are also used occasionally to treat municipal wastewater, especially for small flows prior to facultative or aerated ponds. Upflow anaerobic sludge blanket (UASB) technology has been applied to the treatment of municipal wastewater, especially in tropical climates where wastewater temperatures are seldom less than about 20°C.[7,13,14,29] Application of this approach is well documented in Central and South America (especially Brazil) and is growing in Southern Asian locations like India. Although design criteria for the treatment of municipal wastewater are evolving, currently accepted criteria are listed in Table 23.4.

The application of staged anaerobic digestion systems is growing due to increased volatile solids destruction, which results in increased methane production. Enhanced volatile solids destruction also reduces digested sludge volume, thereby reducing the energy required to transport the residual biosolids to ultimate reuse or disposal. These processes are described in Section 14.1.2. As discussed in Section 14.2.9, waste activated sludge (WAS) hydrolyzes more slowly and to a lesser extent than primary sludge because the cell biomass (both active biomass and cell debris) hydrolyzes very slowly in an anaerobic environment. Advanced digestion processes are thought to result in increased hydrolysis of cell biomass, giving faster and

TABLE 23.4
Design Criteria for UASB Treatment of Municipal Wastewater

Criteria	Value
Temperature	20°C–30°C
Organic loading rate	2–4 kg $BOD_5/(m^3 \cdot day)$
Upflow velocity	0.5–1.0 m/hr
Depth	4–6 m
HRT	4–8 hr
SRT	30–50 days
COD removal efficiency	55%–75%
BOD_5 removal efficiency	55%–75%
TSS removal efficiency	65%–80%
Sludge production	0.4–0.6 kg TSS/kg TSS_{IN}
Methane production	0.35 m^3/kg COD_{Rem}

more complete volatile solids destruction. Cell lysis technologies represent another approach to achieving this objective. Approaches include ultrasound, chemical-mechanical, electrical, and thermo-mechanical systems. Ultrasound involves exposure of thickened waste activated sludge to sound waves, creating cavitation, which produces locally intense conditions that can rupture microbial cells. A typical chemical-mechanical method involves modestly increasing the pH of the thickened WAS and then pumping it through a homogenizer, like that used for milk. Pulsed electrical discharges can also be used to disrupt and lyse microbial cells. Thermo-mechanical treatment involves heating dewatered sludge to temperatures of about 170°C under elevated pressure (6 bar) and then suddenly releasing the pressure. All of these treatments have been demonstrated to be effective in rupturing microbial cells, thereby resulting in an increased rate and extent of volatile solids destruction in subsequent anaerobic digestion. Their cost-effectiveness is under investigation.

23.3.1.2 Contribution to Sustainability

As discussed in Section 10.3.1, anaerobic treatment results in the conversion of biodegradable organic matter into methane, carbon dioxide, and water. The chemical oxygen demand (COD) of the biodegradable organic matter converted to methane is conserved, so the conversion of 1 kg of COD results in the production of 0.35 m^3 of CH_4 under standard conditions. Alternatively, in an aerobic process, between 0.4 and 0.7 kg of O_2 would be required to stabilize the organic matter and 0.3 to 0.6 kg of biomass COD would be produced. Assuming an oxygen transfer system with an efficiency of 1.2 kg O_2/(kW·hr), between 0.3 and 0.6 kW·hr of energy would be required to supply the needed oxygen. In addition, the biomass formed would have to be processed and disposed of requiring further energy. On the other hand, the methane produced in an anaerobic process can be used as an energy source; the energy equivalent of the methane produced from the kg of biodegradable COD is approximately 3.9 kW·hr. Thus, not only is it unnecessary to expend energy to supply oxygen to an anaerobic process, the production of methane provides a significant quantity of energy that can be used as needed. Furthermore, anaerobic processes eliminate the GHG emissions associated with aerobic processes when the energy required for oxygen transfer is produced from fossil fuels. However, it is important to recognize that capture and combustion of the produced methane is critical because the GHG potential of methane is 21 times that of carbon dioxide.[2] One source of methane emissions that is difficult to control is that dissolved in the liquid effluent from the process. This emission can be a significant fraction of the methane produced when the wastewater is dilute (biodegradable COD less than about 2000 mg/L).

23.3.2 Biological Nutrient Removal

Biological nutrient removal reduces energy and chemical consumption associated with the removal of nitrogen and phosphorus.

23.3.2.1 Technology Description

Biological nutrient removal technology is described in Chapter 12 and Section 23.2.2.

23.3.2.2 Contribution to Sustainability

Biological nutrient removal contributes to lower energy and chemical consumption by reducing the energy and chemicals required to remove the nutrients nitrogen and phosphorus. By using the organic matter in the influent wastewater as the electron donor for denitrification, the need to add an electron donor is reduced or eliminated. Denitrification also reduces process oxygen requirements because some of the oxygen demand contained in the influent wastewater is satisfied using nitrate-N produced in the process rather than with oxygen. As mentioned several times, 2.86 kg of oxygen demand are satisfied for each kg of NO_3-N denitrified. Net alkalinity requirements for nitrification are also reduced due to the production of alkalinity by denitrification, as long as denitrification occurs prior to nitrification. For each mg of NO_3-N denitrified, 3.6 mg of alkalinity as $CaCO_3$ are produced. Biological phosphorus removal reduces chemical requirements and sludge production by reducing or eliminating the need to add metal salts to precipitate phosphorus.

23.3.3 Nitritation and Denitritation

With certain high strength wastewaters it is possible to truncate the nitrification and denitrification processes, thereby achieving conversion of ammonia-N to nitrogen gas with less oxygen and less organic electron donor than required in conventional BNR processes. This is accomplished by stopping nitrification and initiating denitrification at nitrite, rather than nitrate, a scheme known as nitritation and denitritation. Nitritation refers to the conversion of ammonia-N to nitrite-N, while denitritation refers to the conversion of nitrite-N to N_2.

23.3.3.1 Technology Description

As discussed in Chapter 12, conventional nitrogen removal processes first require the conversion of ammonia-N to nitrate-N. Although often modeled as a single step process, this conversion is actually accomplished by two separate populations of bacteria. One population, the ammonia oxidizing bacteria (AOB), converts ammonia-N to nitrite-N, whereas the second, known as nitrite oxidizing bacteria (NOB), converts nitrite-N to nitrate-N. Since the valence of nitrogen in ammonia is –3 and the valences of nitrogen in nitrite and nitrate are +3 and +5, respectively, oxidation of ammonia-N to nitrite-N requires the transfer of six electrons, whereas the oxidation of ammonia-N to nitrate-N requires the transfer of eight. That is, oxidation of ammonia-N to nitrite-N requires 25% less oxygen than its oxidation to nitrate-N. Likewise, since the valence of nitrogen gas is zero, reduction of nitrate-N to N_2 requires five electrons, whereas the reduction of nitrite-N to N_2 requires only three, for a 40% saving in the electron donor requirement. Given the savings in both oxygen and electron donor requirements associated with nitritation coupled with denitritation, one might wonder why processes exploiting them are not the norm. The reason is that, under most conditions, AOB generally grow more slowly than NOB so that, whenever conditions are created that allow the growth of AOB, NOB also grow.

It has been discovered that the growth of NOB can be inhibited to a greater extent than AOB under certain conditions, thereby allowing ammonia-N to be converted to nitrite-N, rather than nitrate-N.[30] This has been done in a variety of environments involving wastewater streams with

relatively high ammonia-N concentrations (in excess of about 100 mg-N/L), such as those produced when anaerobically digested sludge is dewatered, those containing landfill leachate, and those from some industries. Work is ongoing to better characterize the conditions that favor the growth of AOB and discourage NOB, but those identified to date include elevated temperature coupled with short aerobic SRT and elevated free ammonia and nitrous acid concentrations. Ammonia is a weak base with a pK_a of approximately 9.3. Thus, the nonionized (or free) ammonia concentration increases with an increase in total ammonia concentration and pH. Nitrous acid is a strong acid with a pK_a of approximately 3.4. Thus, higher total nitrite concentrations and lower pH increase the nonionized (or free) nitrous acid concentration. These factors work together to create conditions that can inhibit NOB when treating high ammonia-N process streams. The influent stream will generally have a moderately elevated pH due to the basic nature of ammonia that, coupled with the high ammonia concentration, leads to elevated free ammonia concentrations. Nitritation decreases the pH by the consumption of alkalinity associated with the conversion of a base (ammonia) to an acid (nitrous acid). Assuming that ammonia is present as ammonium bicarbonate, the stoichiometry for the inorganic constituents involved in nitritation is as follows:

$$NH_4^+ + 1.5\ O_2 + 2\ HCO_3^- \rightarrow NO_2^- + 3\ H_2O + 2\ CO_2. \tag{23.1}$$

Thus, as the nitrite-N concentration increases, the pH will generally decrease, leading to elevated free nitrous acid concentrations. Consequently, nitrogen species that can inhibit NOB are consistently present as influent ammonia-N is converted to nitrite-N and pH is depressed due to alkalinity consumption by nitritation.

Neglecting heterotrophic growth and using methanol as the electron donor, the reaction for denitritation is

$$NO_2^- + 0.5\ CH_3OH + 0.5\ CO_2 \rightarrow 0.5\ N_2 + HCO_3^- + 0.5\ H_2O. \tag{23.2}$$

Combining Equations 23.1 and 23.2 and expressing the influent ammonia-N as ammonium bicarbonate gives the overall reaction for nitritation and denitritation using methanol as the electron donor:

$$NH_4HCO_3 + 1.5\ O_2 + 0.5\ CH_3OH \rightarrow 0.5\ N_2 + 3.5\ H_2O + 1.5\ CO_2. \tag{23.3}$$

A variety of process configurations can be used to accomplish nitritation/denitritation. The most common couple nitritation with heterotrophic denitritation. One is similar to the modified Ludzack-Ettinger process for nitrification and partial denitrification discussed in Sections 7.5, 12.1.2, and 12.3.1. As illustrated in Figure 23.2a, the high ammonia wastewater and a supplemental organic electron donor enter an initial anoxic zone along with mixed liquor containing nitrite-N recirculated from a downstream aerobic zone. Denitritation occurs in the upstream anoxic zone and nitritation occurs in the downstream aerobic zone. Alternatively, cyclic aeration can be used with supplemental organic electron donor addition during the anoxic period to allow denitritation to occur, as illustrated in Figure 23.2b. Systems both with and without solids separation and recycle have been implemented, depending on the process temperature and influent wastewater strength. Aerobic SRT values must be controlled to allow AOB to grow while minimizing NOB growth. Figure 10.4 can be used as a general guide to the required aerobic SRT, which should be maintained close to, but not less than, the minimum SRT values shown. Aeration and electron donor addition must also be controlled. Online control using pH, oxidation-reduction potential, and nutrient analyses are all options that have been used successfully in bench- and pilot-scale studies. The relative advantages and disadvantages of these options will likely be demonstrated as full-scale systems are implemented.

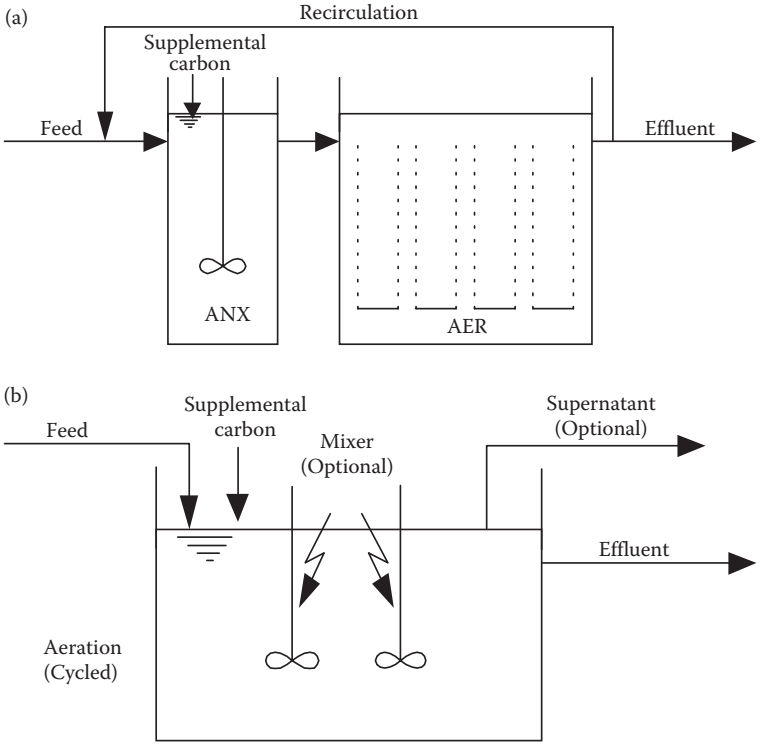

FIGURE 23.2 Nitritation and denitritation processes using heterotrophic denitrification. (a) Option with separate anoxic and aerobic stages. (b) Option with cyclic aeration.

A second scheme for accomplishing nitritation and denitritation utilizes the autotrophic reduction of nitrite-N to nitrogen gas using ammonia-N as the electron donor. A simplified version of this reaction follows (a small amount of nitrate-N is also formed but is neglected here):

$$NH_4^+ + NO_2^- \rightarrow N_2 + 2\ H_2O. \tag{23.4}$$

Referred to as the anammox reaction (Section 2.3.3), bacteria capable of accomplishing it were first characterized in natural environments and then applied in processes.[24] As noted in Equation 23.4, both ammonia-N and nitrite-N must be present in near equimolar concentrations. This can be accomplished if nitritation of a high ammonia stream occurs without heterotrophic denitritation. Typically, such streams contain about one mole of alkalinity for each mole of ammonia-N, whereas two moles are required for aerobic nitritation as indicated in Equation 23.1. Thus, sufficient alkalinity is available to convert half of the influent ammonia-N to nitrite-N, leaving the other half as ammonia-N:

$$2\ NH_4HCO_3 + 1.5\ O_2 \rightarrow NH_4^+ + NO_2^- + 3\ H_2O + 2\ CO_2. \tag{23.5}$$

This is exactly the mix of ammonia-N and nitrite-N required for the anammox reaction, as indicated by Equation 23.4. The nitrite-N formed is subsequently reduced autotrophically under anoxic conditions using ammonia-N as the electron donor via the anammox reaction (Equation 23.4). The overall reaction for nitritation and autotrophic denitritation using anammox is

$$2\ NH_4HCO_3 + 1.5\ O_2 \rightarrow N_2 + 5\ H_2O + 2\ CO_2. \tag{23.6}$$

Comparing Equation 23.3 to Equation 23.6 shows that nitritation with autotrophic denitritation cuts the oxygen requirement in half compared to nitritation with heterotrophic denitritation and eliminates the need for an organic electron donor.

The anammox bacteria responsible for autotrophic denitritation grow very slowly.[8,23] Doubling times at 20°C are on the order of 20 days, compared to 2 days for autotrophic nitrifiers. Work is ongoing to further define the growth requirements for the anammox bacteria, which will lead to better bioreactor configurations. A variety of suspended and attached growth systems have demonstrated nitritation and autotrophic denitritation and further advances are expected.

23.3.3.2 Contribution to Sustainability

The stoichiometry for the nitritation and denitritation reactions demonstrates the decreased energy and chemical consumption they provide. Nitritation reduces process oxygen requirements by 25% relative to conventional nitrification. If nitritation is coupled with heterotrophic denitritation, electron donor requirements are reduced by 40% compared to conventional denitrification of nitrate-N. Alkalinity requirements are also eliminated. If nitritation is coupled with autotrophic denitritation, oxygen requirements are further reduced to half those associated with nitritation and heterotrophic denitritation since only half of the ammonia-N needs to be oxidized to nitrite-N, and organic electron donor requirements are eliminated. Thus, compared to conventional nitrification and denitrification, nitritation and autotrophic denitritation reduces oxygen requirements by 62.5%. To date, these benefits have been demonstrated only for process streams with high ammonia concentrations.

23.3.4 Biological Air Treatment

By its very nature, this book has focused on biological processes used to treat liquid streams and slurries. However, biological processes are also used to treat the air streams emitted from wastewater treatment plants.[32] The use of biological systems for this application is a relatively new development as, historically, chemical systems have been used. Biological systems have not only proven to be robust and effective for these applications, they have also resulted in significant chemical savings.

23.3.4.1 Technology Description

While a wide variety of biological processes have been developed to treat contaminated air streams, the vast majority of them are attached growth systems. The classical example is the biofilter, as illustrated in Figure 23.3. These are shallow beds into the bottom of which the contaminated air stream is introduced. The porous bed supports a biofilm that metabolizes organic and inorganic compounds in the air stream passing through. Organic materials such as wood chips and compost

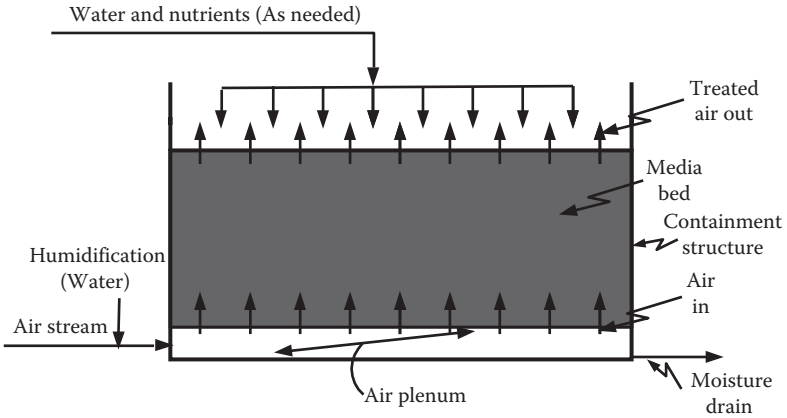

FIGURE 23.3 Biological air treatment process.

have historically served as the media to which the biofilm attaches. These materials provide nutrients needed for biological growth and, consequently, deteriorate over time, necessitating their periodic replacement. To avoid this requirement, inert media of a variety of descriptions, including rounded river rock, lava rock, and a variety of proprietary media have been used.

All of the principles applicable to the treatment of liquid streams also apply to the treatment of contaminated air streams. Sufficient biomass must be retained to metabolize the target compounds, nutrients must be provided, and temperature and pH must be maintained in the physiological range for the target organisms. In addition, sufficient moisture must be present to allow microbial growth. Because contaminated air streams often contain insufficient moisture, they must be humidified. This requires care because excessive moisture must be avoided to prevent channeling and plugging. Required nutrients can be provided using raw or partially treated wastewater, as well as process effluent. These features are illustrated in Figure 23.3.

Contaminated air streams at wastewater treatment plants can contain hydrogen sulfide, reduced sulfur compounds, ammonia, nitrogen containing compounds such as amines, and a wide variety of organic compounds. Hydrogen sulfide can be detected by humans at extremely low concentrations (approximately 0.4 ppb by volume) and, consequently, its removal is a key objective in many air treatment systems. Oxidation of hydrogen sulfide to sulfate produces sulfuric acid, which lowers the pH. Because many species of sulfide oxidizing bacteria are acidophilic, successful oxidation of hydrogen sulfide can be accomplished even at low pH. However, reduced pH can contribute to deterioration of support media and other portions of the system. This has contributed to the increased popularity of inert support media, which are more acid resistant than organic media. Reduced pH can also inhibit the bacteria responsible for the metabolism of other constituents in the air stream, leading to the use of two biofilters in series in many cases—the first operating at low pH to remove hydrogen sulfide and the second operating at more neutral pH to remove other constituents.

While biofilters have traditionally been the system of choice for biological air treatment, newer systems incorporating alternate media and recirculating microbial slurries have been developed that have greater volumetric mass transfer and reaction rates and greater depths, thereby allowing configurations with smaller footprints to be used. These can be quite competitive with biofilters, especially on constrained sites where sufficient space may not be available for a conventional biofilter and where higher performance is needed. Detailed information on biological air treatment systems is available elsewhere.[32]

23.3.4.2 Contribution to Sustainability

Biological air treatment systems contribute to sustainability in two important ways. First, they provide an alternative to chemical air treatment systems, thereby reducing or eliminating chemical requirements for this purpose. This can be significant at facilities receiving wastewaters that are odorous or with industrial components containing volatile compounds, or at plants with extensive solids processing because the ventilation air from these processes is often quite odorous. Second, they make air treatment less expensive. Public acceptability of wastewater treatment facilities depends not only on their effectiveness but also on their impacts on the local community. Air emissions are one of those impacts. While odors have historically been viewed as an inevitable result of wastewater treatment, today they are becoming unacceptable to the public. Thus, the availability of efficient and cost-effective air treatment technology facilitates the siting of treatment facilities in their most logical location.

23.4 TECHNOLOGIES TO ACHIEVE RESOURCE RECOVERY

Human activity results in the consumption of a wide variety of resources. In a fully sustainable system, resources would be used and recycled or renewable sources would be used. The two previous sections have discussed several resource recovery opportunities. Water reclamation and reuse, as discussed in Section 23.2, helps to close the water cycle and makes increasing volumes of water available for human use. Likewise, energy recovery through the production of methane gas results

TABLE 23.5
Key Features and Contributions to Sustainability of Technologies to Achieve Resource Recovery

Technology	Key Features	Contribution to Sustainability
Biological nutrient removal and recovery	• Concentration and separation of phosphorus, ammonia, and magnesium • Precipitation as calcium phosphate [$Ca_3(PO_4)_2$] or struvite ($MgNH_4PO_4$)	• Recovery of nitrogen and phosphorus in reusable forms • Reduced nitrogen and phosphorus discharges
Land application of biosolids	• Conversion of wastewater residuals to usable form • Application to land as fertilizer	• Recycle of nutrient and organic matter content of biosolids for beneficial purposes

in the production of a renewable energy source. Moreover, other useful products can be extracted from wastewater streams. For example, research is ongoing at the bench-scale to produce usable biopolymers from wastewater.[15] As with microbial fuel cells, the reader is urged to follow development of this technology. This section discusses one evolving and one widely used resource recovery technology: (1) the recovery of nutrients and (2) reuse of biosolids. Key features of these technologies and their potential to contribute to increased sustainability are summarized in Table 23.5.

23.4.1 Biological Nutrient Removal and Recovery

As discussed in Section 2.4.6, the phosphate accumulating organisms (PAOs) active in biological phosphorus removal systems can accumulate high concentrations of phosphorus in the form of polyphosphate granules. Under anaerobic conditions this accumulated phosphorus is released, leading to elevated phosphorus concentrations. The cations magnesium and potassium are also taken up and released, also leading to increased concentrations. Ammonia concentrations are also increased if the sludge undergoes digestion. Elevated concentrations of these inorganic species form the basis for cost-effective nutrient recovery systems.[33]

23.4.1.1 Technology Description

Two products are receiving principal focus: (1) calcium phosphate and (2) struvite. Calcium phosphate is produced by adding slaked lime (calcium hydroxide) to a concentrated phosphorus-containing stream:

$$3\ Ca(OH)_2 + 2\ PO_4^{3-} \rightarrow Ca_3(PO_4)_2 \downarrow + 6\ OH^- . \tag{23.7}$$

This is often done in a fluidized bed reactor so that the precipitate forms in a crystalline fashion, thereby providing calcium phosphate with essentially the same composition and properties as the phosphate rock mined and used as a raw material by the phosphate industry.

Struvite is the common name for magnesium ammonia phosphate ($MgNH_4PO_4$), which readily forms in many sludge processing systems, such as in anaerobic digesters or when anaerobically digested sludge is dewatered.[18–20] Anaerobic digestion of sludges from biological phosphorus removal plants can result in quite high concentrations of all three ions needed to form struvite. Both phosphate and magnesium are released from the PAOs, and ammonia (along with some phosphate and magnesium) is released during the destruction of biomass. Struvite is relatively soluble, but its solubility decreases as the pH increases. Consequently, struvite formation is often observed at points of turbulence where carbon dioxide is stripped from the sludge, resulting in localized increases in pH. This can result in severe scaling, which can be an operational problem, as discussed in Section 14.4.2. However, if the precipitation of struvite can be controlled, it can be

recovered, providing material that can serve either as a phosphate source or as a high quality, slow release fertilizer.

Several process options have been investigated to produce a stream with relatively low suspended solids but elevated phosphate concentrations. One early version was the Phostrip biological phosphorus removal process in which return activated sludge (RAS) is held under anaerobic conditions to select for PAOs and to release phosphorus. The released phosphorus is subsequently elutriated from the RAS and precipitated as calcium phosphate using lime. A more recent variation of this process decants the effluent from the anaerobic zone of a conventional biological phosphorus removal process to obtain a low volume, high phosphate concentration stream that can be chemically treated. Process streams from the anaerobic digestion of waste biological phosphorus removal sludge can also be processed to produce struvite. Struvite formation is often limited by the available magnesium, which can be added to increase the precipitated struvite.

23.4.1.2 Contribution to Sustainability

The recovery of nutrients from wastewater allows them to be recycled back into food production. Phosphorus and nitrogen are necessary nutrients for plant growth and are typical constituents of fertilizers added to soils to increase their agricultural productivity. Ultimately some of the added nutrients end up in the food consumed by human beings. A small portion of those nutrients is retained for our growth, but most is excreted and ends up in our wastewater. If nutrient removal is practiced at wastewater treatment plants, most of the removed nitrogen is returned to the atmosphere, but the removed phosphorus remains with the removed biosolids. If the biosolids are recycled to agriculture, then the cycle of phosphorus uptake by crops and excretion by humans is closed, but if they are not, the phosphorus ends up in our waterways and, eventually, the ocean.

The recovery of phosphorus from wastewater allows phosphate to be recycled to agriculture independent of the recycle of biosolids to agriculture. Two factors are important in this regard. First, phosphate rock is a limited resource that is expected to last no more than about 150 to 200 years at current consumption rates. Second, prior to the implementation of phosphorus removal at wastewater treatment plants, the land application of biosolids was limited by the nitrogen content of the biosolids produced because the agronomic requirement for nitrogen was reached at lower biosolids loading rates than the agronomic requirement for phosphorus. However, when phosphorus removal is practiced, the phosphorus content of biosolids is increased sufficiently that it will control the application rate of the biosolids, thereby lowering the mass of biosolids that can be applied per acre. This can adversely impact the economics of biosolids land application because more land is required. Greenhouse gas production is also increased because the longer haul distances associated with the greater land requirements result in greater fossil fuel consumption to transport the biosolids. Because fewer biosolids can be applied per acre, the agricultural value of biosolids is reduced, thereby decreasing the number of farmers who are willing to accept biosolids on their lands. On the other hand, the separate recovery of phosphorus (e.g., as struvite) eliminates this issue by producing a biosolids that can be applied at the higher loading rates associated with the nitrogen requirement. It also allows phosphate to be recycled into the fertilizer industry, thereby extending the life of limited phosphate resources.

Nitrogen used in agriculture is generally produced by extracting it from air using the Haber-Bosch process. The removal of nitrogen from wastewater by nitrification and denitrification simply returns nitrogen back into the atmosphere. However, fertilizer production by the Haber-Bosch process is very energy-intensive, and this energy consumption could be avoided if nitrogen was recovered directly from wastewater. The formation of struvite is one approach. The land application of biosolids is another.

23.4.2 Land Application of Biosolids

The land application of biosolids produced at both municipal and industrial wastewater treatment plants is a well-established practice. It is environmentally beneficial because it recycles the nutrient

content of the produced biosolids. As discussed in Sections 13.1.2 and 14.1.2, the pathogen content of biosolids must be controlled to protect public health. Likewise, the content of other potentially harmful constituents, such as heavy metals and toxic organic compounds, must be controlled to acceptable levels. Biosolids application rates must provide nutrients at agronomic rates so that they are fully utilized and their presence in run-off is minimized. Historically, conventional mesophilic anaerobic digestion has been the principal approach to producing Class B biosolids (Table 14.1) for land application, although aerobic digestion has frequently been used in smaller treatment plants. More recently, advanced anaerobic digestion technologies have been developed to allow production of higher quality biosolids.

23.4.2.1 Technology Description

As discussed in Section 14.1.6, the term advanced anaerobic digestion refers to a range of technologies that use staged reactors, potentially separating solids hydrolysis and acetogenesis from methanogenesis and also possibly including a mix of mesophilic and thermophilic operating temperatures. These modifications can improve the quality of biosolids from anaerobic digestion in several ways. First, they can significantly increase the inactivation of pathogens, resulting in biosolids that can be more widely distributed (i.e., meet "Class A" pathogen reduction requirements, Table 14.1). Secondly, they can achieve greater destruction of biodegradable organic matter, giving a more stable product with less potential for producing odors during application. Increased destruction of organic matter also results in increased methane production, which increases energy recovery. Currently, significant work is ongoing to quantify the potential of various advanced digestion processes.

23.4.2.2 Contribution to Sustainability

Land application represents one of the best approaches available for recycling and reusing the nutrients and organic matter in biosolids, especially for agricultural purposes. Not only is a full range of nutrients provided (the macronutrients nitrogen and phosphorus as well as an array of micronutrients), but the organic content of the soil is increased, which improves its water holding capacity. Furthermore, carbon can be retained in the soil matrix, providing for its long-term sequestration to reduce greenhouse gas emissions. One of the historical concerns associated with the reuse of biosolids has been the presence of contaminants, such as heavy metals and persistent organic compounds. Fortunately, the levels of these contaminants in biosolids have been declining, primarily as a result of source control programs. Other concerns associated with biosolids application include the potential for disease transmission and odor production. As noted above, advanced digestion technologies address these concerns directly, resulting in biosolids with more acceptable characteristics for land application.

23.5 CLOSING COMMENTS

This chapter addresses how sustainability objectives can be achieved with a variety of biological treatment technologies. As discussed in Section 23.1.2, sustainability is achieved by systems, not by individual technologies. The information in Sections 23.2, 23.3, and 23.4 can be used to identify technologies that can be components of more sustainable systems, but it is the performance of the system that must be measured (by TBL analysis) to determine which achieves greater sustainability. The discussion above also illustrates that a particular technology may contribute to sustainability in several ways. Biological nutrient removal offers an excellent example because it provides both water resource availability and lower energy and chemical consumption, while forming the basis for resource recovery. Likewise, advanced digestion systems, perhaps coupled with cell lysis and nutrient recovery technologies, can be used to produce systems that recover more energy and nutrients for reuse. Finally, this chapter demonstrates the complexity of any sustainability analysis. For example, a particular component may appear to decrease sustainability because of its energy requirements, whereas, in reality, its use causes an overall net reduction in energy requirements for the entire system. As engineers become more aware of the importance of sustainable systems, it is almost certain that improvements will be

made in existing biological processes and that new biological processes will be developed to increase the contribution of these important technologies to sustainable systems.

23.6 KEY POINTS

1. The concern for increased sustainability is created by population growth and increased per capita resource consumption.
2. Sustainability must be defined and made measurable before more sustainable environmental solutions can be selected and implemented. In this regard, sustainable development is the provision of infrastructure that enhances the ability of the planet to sustain human life.
3. Two views of sustainability exist: (1) absolute and (2) relative. Relative sustainability is a more useful concept upon which to develop measurement tools because it can be more easily applied and can result in the selection of options that enhance sustainability.
4. Sustainability is characterized by the triple bottom line (TBL) that considers social, economic, and environmental factors. Considering social and environmental factors, along with economic ones, leads to retention of the social "license to operate" for both private and public entities.
5. The TBL forms the basis for techniques to measure the relative sustainability of options. These procedures use utility scales to measure the relative contribution of individual social, economic, and environmental considerations relevant to a specific decision. The values of the stakeholders form the basis for establishing the weighting factors that quantify the relative importance of these individual considerations.
6. Sustainability must be measured for systems, not for individual system components. This is to avoid simple transfers of adverse impacts and to ensure that the broad impacts of a particular action are fully assessed.
7. System configuration can lead to reduced resource requirements. For example, distributed water reclamation and reuse systems reduce energy required for pumping because both collected wastewater and reclaimed water need not be conveyed as far.
8. The membrane bioreactor process contributes to greater water resource availability because of its long solids retention time (SRT), which allows for complete nitrification and improved removal of biodegradable organic matter (including trace constituents); its superior removal of particulate and colloidal matter, which facilitates disinfection and further treatment; and its suitability for automation, which facilitates remote operation.
9. Biological nutrient removal contributes to greater water resource availability by protecting surface water from the adverse impacts of eutrophication, by reducing energy and chemical use, and by providing the basis for nutrient recovery.
10. Advanced oxidation coupled with biodegradation contributes to greater water resource availability by increasing the removal and biodegradability of recalcitrant organic compounds.
11. Anaerobic treatment of industrial wastewater is an established process that is being extended to the treatment of municipal wastewater, especially in tropical locations where wastewater temperatures are seldom less than 20°C. Anaerobic treatment reduces energy requirements and sludge production and produces methane, which can serve as an energy source. The methane must be captured and used or combusted to avoid significant greenhouse gas impacts (CH_4 greenhouse gas potential = 21 × CO_2 impacts).
12. Advanced digestion and cell lysis technologies are increasingly being applied to stabilize wastewater sludges. Benefits include increased methane production, reduced biosolids volumes, and improved pathogen destruction.
13. Nitritation is the conversion to ammonia-N to nitrite-N, rather than to nitrate-N (nitrification). Nitritation reduces process energy requirements by 25% compared to nitrification, and carbon requirements for heterotrophic denitrification are reduced by 40%. The reduction

of nitrite-N to nitrogen gas is known as denitritation. Nitritation occurs when the growth of ammonia oxidizing bacteria is encouraged, but the growth of nitrite oxidizing bacteria is inhibited. This can result from elevated temperature, short SRT, and elevated nonionized ammonia or free nitrous acid concentrations.

14. In the anammox process, ammonia-N serves as the electron donor and nitrite-N as the electron acceptor, resulting in the conversion of the nitrogen in both the ammonia and the nitrite to nitrogen gas. Coupled with nitritation, anammox eliminates the need for organic carbon for nitrogen removal and reduces process oxygen requirements to half those required for nitritation and to only 37.5% of those required for conventional nitrification and denitrification. The anammox bacteria grow quite slowly, with a doubling time of about 20 days at 20°C.
15. Biological nitrogen removal reduces energy and chemical consumption by reducing process oxygen requirements (2.86 kg oxygen demand satisfied per kg NO_3-N denitrified) and producing alkalinity that can offset that consumed by nitrification (3.6 kg alkalinity as $CaCO_3$ produced per kg NO_3-N denitrified).
16. Biological air treatment reduces chemical consumption compared to conventional physical-chemical treatment technologies. Several process configurations are available, including shallow packed beds using organic or inorganic media and fixed film and slurry biotowers. Moisture control is important—sufficient moisture must be available for biological growth, but excess moisture must be avoided to prevent plugging and channeling in the bed. Sulfuric acid production during hydrogen sulfide oxidation can cause pH depression and the need for two treatment stages in some cases.
17. Biological nutrient removal systems can concentrate phosphorus, magnesium, and ammonia to levels that allow either calcium phosphate or magnesium ammonium phosphate (struvite) to be extracted from wastewater. Calcium phosphate of the same quality as the phosphate rock that is mined and used as a raw material by the fertilizer industry can be produced. Phosphate is precipitated by adding lime, generally in a fluidized bed reactor. Struvite formation requires equimolar concentrations of magnesium, ammonia, and phosphate. For many process streams, this requires the addition of magnesium. Struvite is an excellent slow release fertilizer.
18. Nutrient recovery contributes to increased sustainability because recoverable phosphate rock resources may only last 150 to 200 years at current consumption rates. Recovery of nutrients from wastewater helps to close the nutrient cycle for human food production.
19. Land application allows the nutrients and organic matter in biosolids to be utilized in agriculture. Advanced digestion technologies can contribute to improved biosolids quality by reducing pathogen levels (approaching or achieving Class A quality) and odor potential.

23.7 STUDY QUESTIONS

1. Prepare a table summarizing the factors leading to the need for increased sustainability and how they are creating that need.
2. Compose your own definition of sustainability. Base it upon your assessment of the current and future world situation and your hopes and dreams for you and your descendents.
3. Develop a list of factors that can be used to quantify the social contribution of a sustainable wastewater management option, along with quantitative measures.
4. Develop a list of factors that can be used to quantify the economic contribution of a sustainable wastewater management option, along with quantitative measures.
5. Develop a list of factors that can be used to quantify the environmental contribution of a sustainable wastewater management option, along with quantitative measures.
6. Develop a potential sustainable urban water management option, including the wastewater management component.

7. Formulate a set of triple bottom line evaluation criteria for the potentially sustainable urban water management option formulated in Study Question 6.
8. Develop a potential urban water management option, including the wastewater management component, which minimizes net energy consumption.
9. Formulate a set of triple bottom line evaluation criteria for the potentially sustainable, low energy urban water management option formulated in Study Question 8.
10. Develop a potential urban water management option, including the wastewater management component, which incorporates resource recovery.
11. Formulate a set of triple bottom line evaluation criteria for the potentially sustainable urban water management option that incorporates resource recovery formulated in Study Question 10.
12. Summarize lessons learned from the development and evaluation of potentially sustainable urban water management options.

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Appendix A: Acronyms

Acronym	Definition	Section Where First Used
1-D	One-dimensional	16.5
2-D	Two-dimensional	16.5
3-D	Three-dimensional	16.5
A/AD	Anoxic/aerobic digestion	1.3.1
AAD	Advanced anaerobic digestion	14.1.2
ABF	Activated biofilter	19.1.2
AD	Anaerobic digestion	1.3.1
ADM	Anaerobic digestion model	6.0
ADP	Adenosine diphosphate	2.4.1
AEL	Aerobic lagoon	15.1.2
AER	Aerobic	Figure 1.10
AF	Anaerobic filter	1.3.1
ANA	Anaerobic	Figure 1.10
ANL	Anaerobic lagoon	1.3.1
ANX	Anoxic	Figure 1.10
AOB	Ammonia oxidizing bacteria	2.3.2
AOL	Aerial organic loading	15.2.3
A/O	Anaerobic/oxic	1.3.1
A^2/O	Anaerobic/anoxic/oxic	1.3.1
AR	Anoxic recirculation	12.1.2
ASM	Activated sludge model	6.0
ATAD	Autothermal thermophilic aerobic digestion	1.3.1
ATP	Adenosine triphosphate	2.4.1
BAC	Biological activated carbon	23.2.3
BAF	Biological aerated filter	21.1.2
BF/AS	Biofilter/activated sludge	19.1.2
BNR	Biological nutrient removal	1.3.1
BOD	Biochemical oxygen demand	9.6
BOD_5	Five-day biochemical oxygen demand	3.2.10
BPR	Biological phosphorus removal	2.2.1
BSM	Benchmark system model	8.3.2
CAD	Conventional aerobic digestion	1.3.1
CAS	Conventional activated sludge	1.3.1
CFSTR	Continuous-flow stirred tank reactor	4.3.1
CMAL	Completely mixed aerated lagoon	1.3.1
CMAS	Completely mixed activated sludge	1.3.1
COD	Chemical oxygen demand	1.2.1
CSAS	Contact stabilization activated sludge	1.3.1
CST	Capillary suction time	11.4.2
CSTR	Continuous stirred tank reactor	1.2.3

(*Continued*)

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Acronym	Definition	Section Where First Used
DFBP	Downflow packed bed	Figure 1.32
DNA	Deoxyribonucleic acid	9.2.1
DO	Dissolved oxygen	1.0
DSVI	Diluted sludge volume index	2.3.1
EAAS	Extended aeration activated sludge	1.3.1
EBBR	Expanded bed biological reactor	21.1.4
ED	Entner-Doudoroff	2.4.6
EGSB	Expanded granular sludge blanket	14.1.3
EMP	Embden-Meyerhof-Parnas	2.4.6
EPS	Extracellular polymeric substance	2.3.1
F/AL	Facultative/aerated lagoons	1.3.1
FA	Free ammonia	3.2.10
FBBR	Fluidized bed biological reactor	1.2.3
FB/EB	Fluidized bed/expanded bed	14.1.3
FISH	Fluorescence in situ hybridization	16.5
F/M	Food to microorganism ratio	5.1.7
FNA	Free nitrous acid	3.2.10
FSS	Fixed suspended solids	5.2
GAC	Granular activated carbon	21.1.4
GAO	Glycogen accumulating organism	2.2.1
GHG	Greenhouse gas	23.3
HO	Horizontal	19.2.6
HPOAS	High purity oxygen activated sludge	1.3.1
HRT	Hydraulic residence time	4.3.2
IC_{50}	Inhibitor concentration causing a 50% reduction in activity	22.4.2
IFAS	Integrated fixed film activated sludge	1.3.2
IIM	Insoluble inorganic matter	1.1
IOM	Insoluble organic matter	1.1
IWA	International Water Association	6.0
JHB	Johannesburg	12.1.2
LCFA	Long chain fatty acid	8.1
MBBR	Moving bed biological reactor	1.3.2
MBR	Membrane bioreactor	1.3.1
MBRAS	Membrane bioreactor activated sludge	1.3.1
MLE	Modified Ludzak-Ettinger	1.3.1
MLR	Mixed liquor recirculation	Figure 1.10
MLSS	Mixed liquor suspended solids	1.3.1
MLVSS	Mixed liquor volatile suspended solids	11.4.3
MOP	Manual of practice	11.3.2
MPN	Most probable number	Table 13.2
MW	Molecular weight	3.1.1
NAD	Nicotinamide adenine dinucleotide	2.4.1

(CONTINUED)

Acronym	Definition	Section Where First Used
NADH	Reduced nicotinamide adenine dinucleotide	2.4.1
NADP	Nicotinamide adenine dinucleotide phosphate	2.4.1
NADPH	Reduced nicotinamide adenine dinucleotide phosphate	2.4.1
NOB	Nitrite oxidizing bacteria	2.3.2
NR	Nitrified mixed liquor recirculation	12.3.3
NRC	National Research Council	17.1.6
NUR	Nitrate utilization rate	9.5.3
NUR_g	Nitrate utilization rate associated with the utilization of readily biodegradable substrate	9.5.3
NUR_h	Nitrate utilization rate associated with the utilization of slowly biodegradable substrate	9.5.3
NVSS	Nonvolatile suspended solids	13.1.1
O&M	Operations and maintenance	Table 15.1
OHO	Ordinary heterotrophic organisms	7.7.2
OUR	Oxygen uptake rate	9.3.1
OUR_g	Oxygen uptake rate associated with the utilization of readily biodegradable substrate	9.5.3
OUR_h	Oxygen uptake rate associated with the utilization of slowly biodegradable substrate	9.5.3
PAO	Phosphate accumulating organism	2.2.1
PFR	Plug-flow reactor	1.2.3
PHA	Poly-β-hydroxyalkanoate	2.4.6
PHB	Poly-β-hydroxybutyrate	2.4.2
PHV	Poly-β-hydroxyvalerate	2.4.6
Poly-P	Polyphosphate	2.4.6
RA	Random	19.2.6
RAS	Return activated sludge	1.3.1
RBC	Rotating biological contactor	1.3.2
RDR	Rotating disc reactor	1.2.3
RF/AS	Roughing filter/activated sludge	19.1.2
RNA	Ribonucleic acid	2.4.2
RTD	Residence time distribution	4.3.2
SAGB	Submerged attached growth bioreactor	Figure 1.32
SAL	Surface ammonia-N loading	19.2.1
SAS	Selector activated sludge	1.3.1
SBR	Sequencing batch reactor	1.2.3
SBRAS	Sequencing batch reactor activated sludge	1.3.1
SCOD	Soluble chemical oxygen demand	21.2.2

(*Continued*)

(CONTINUED)

Acronym	Definition	Section Where First Used
SFAS	Step feed activated sludge	1.3.1
SIM	Soluble inorganic matter	1.1
SK	Spülkraft	19.2.7
SMP	Soluble microbial product	2.4.3
SNL	Surface nitrogen loading	19.2.1
SOL	Surface organic loading	19.2.1
SOM	Soluble organic matter	1.1
SOUR	Specific oxygen uptake rate	11.2.4
SP	Soluble phosphorus	12.2.2
SRT	Solids retention time	1.3.1
$SSVI_{3.5}$	Stirred sludge volume index at 3.5 g/L	2.3.1
SVI	Sludge volume index	2.3.1
SVI_m	Mallory sludge volume index	Table 2.3
TAL	Total ammonia-N loading	19.1.2
TBL	Triple bottom line	23.1.2
TCA	Tricarboxylic acid	2.4.6
TF	Trickling filter	1.3.2
TF/AS	Trickling filter/activated sludge	19.1.2
TF/SC	Trickling filter/solids contact	19.1.2
THL	Total hydraulic loading	14.2.3
TKN	Total Kjeldahl nitrogen	9.6
TNL	Total nitrogen loading	19.1.2
TO	Transfer rate of oxygen	12.3.1
TOL	Total organic loading	19.1.2
TP	Total phosphorus	12.2.2
TPAD	Temperature phased anaerobic digestion	14.1.6
TS	Total solids	Table 13.2
TSS	Total suspended solids	2.4.1
UASB	Upflow anaerobic sludge blanket	1.3.1
UCT	University of Cape Town	1.3.1
UFPB	Upflow packed bed	Figure 1.32
UV	Ultraviolet	23.2.1
VFA	Volatile fatty acid	1.3.1
VFC	Vertical fully corrugated	19.2.6
VIP	Virginia Initiative Plant	12.1.2
VOL	Volumetric organic loading	14.1.3
VS	Volatile solids	Table 13.2
VSS	Volatile suspended solids	2.4.1
WAS	Waste activated sludge	1.3.1
XF	Cross flow	19.2.6
XOC	Xenobiotic organic chemical	22.0

Appendix B: Symbols

Symbol	Definition	Units	Place or Equation Where First Used
a	Empirical coefficient	—	17.15; 22.14
a_i	Molar stoichiometric coefficient for reactant A_i	mole/mole	3.1
a_s	Specific surface area	L^{-1}	16.42
A	Constant in the Arrhenius equation	Varies	3.91
A_c	Cross-sectional area	L^2	4.7
A_i	Constituent i in a stoichiometric equation	mole or M	3.1
A_L	Surface area of a lagoon	L^2	15.1
A_s	Planar biofilm surface area normal to the direction of diffusion; Total media surface area in a packed tower or an RDR	L^2	16.7; Section 17.1.2
A_{si}	Planar surface area of biofilm in the i^{th} CSTR in a chain of N CSTRs	L^2	17.2
$A_{s,n}$	Total media surface area in stage N of an RBC system	L^2	20.6
$A_{s,N}$	Total media surface area in the last stage of an RBC system	L^2	20.7
b	Decay coefficient, traditional approach	T^{-1}	3.56
b	Empirical temperature coefficient	$T^{-1}(°C)^{-1}$	3.102
b	Empirical coefficient	—	17.15; 22.15
b_A	Decay coefficient for autotrophs	T^{-1}	Section 3.9.2
b_C	Competition coefficient for a biofilm	T^{-1}	16.59
b_D	Detachment coefficient	T^{-1}	16.21
b_H	Decay coefficient for heterotrophs, traditional approach	T^{-1}	Equation 3.56a in Section 5.1.3
b_i	Decay coefficient (traditional approach) for species i in a biofilm	T^{-1}	16.53
b_L	Decay coefficient, lysis:regrowth approach	T^{-1}	3.63
$b_{L,A}$	Decay coefficient for autotrophs, lysis:regrowth approach	T^{-1}	Table 6.1
$b_{L,H}$	Decay coefficient for heterotrophs, lysis:regrowth approach	T^{-1}	Table 6.1
$b_{L,PAO}$	Decay coefficient for PAOs, lysis:regrowth approach	T^{-1}	Sec. 7.7.1
$b_{L,PHA}$	Decay coefficient for PHA, lysis:regrowth approach	T^{-1}	Sec. 7.7.1
$b_{L,PP}$	Decay coefficient for polyphosphate, lysis:regrowth approach	T^{-1}	Sec. 7.7.1
b_{MV}	Decay coefficient for MLSS, VSS units	T^{-1}	13.2
b_{XOC}	Decay coefficient for biomass grown on an XOC	T^{-1}	5.22 in Chapter 22
Bi	Biot number	—	16.11
BOD_5	5-Day biochemical oxygen demand concentration	ML^{-3}	9.26
BOD_5/P	BOD_5 to phosphorus removal ratio	$M_{BOD5}M_P^{-1}$	12.4
BOD_u	Ultimate biochemical oxygen demand concentration	ML^{-3}	9.27
c	Empirical temperature coefficient	$°C^{-1}$	3.102
C	Temperature coefficient	$°C^{-1}$	3.97
C	Empirical coefficient in the modified Velz/Germain equation for a trickling filter	Varies	19.7
C_A	Concentration of reactant A	ML^{-3}	4.3
C_{AO}	Concentration of reactant A in the influent	ML^{-3}	4.3

(*Continued*)

(CONTINUED)

Symbol	Definition	Units	Place or Equation Where First Used
C'_A	Concentration of reactant A in the stream passing through the biomass separator	ML^{-3}	5.4
C_D	Drag coefficient	—	18.2
$C_{G,i}$	Concentration of constituent i in the gas phase	Atm	8.4
$C_{G,XOC}$	Concentration of an XOC in the gas phase	ML^{-3}	22.2
COD_{BO}	Influent biodegradable COD concentration	ML^{-3}	9.29
COD_{IO}	Influent inert COD concentration	ML^{-3}	9.32
COD_{TO}	Influent total COD concentration	ML^{-3}	9.21
$C_{S,XOC}$	Solid phase concentration of an XOC	MM^{-1}	22.13
d	Characteristic dimension in the Reynolds number	L	Section 16.2.1
d_b	Diameter of a bioparticle	L	18.7
d_p	Diameter of a carrier particle	L	18.1
D	Batch holding time for pathogen destruction	T	Table 14.1
D_e	Effective diffusivity in a biofilm	L^2T^{-1}	16.5
D_L	Axial dispersion coefficient	L^2T^{-1}	4.22
D_L/vL	Dispersion number	—	4.23
D_w	Diffusivity in water	L^2T^{-1}	16.1
D_{w,O_2}	Diffusivity of oxygen in water	L^2T^{-1}	22.10
$D_{w,XOC}$	Diffusivity of an XOC in water	L^2T^{-1}	22.10
E	Empirical coefficient	LT^{-1}	17.18
E'_0	Standard oxidation-reduction potential	mV	Table 2.6
$E(t)$	Fraction of fluid elements leaving a reactor that have residence times between t and t + dt	—	4.16
E_{XMV}	Efficiency of VSS destruction	%	13.4
f_A	Active fraction of the biomass	—	5.36
f_{Ac}	Fraction of packed tower cross-sectional area occupied by liquid film	—	Section 17.1.2
f_D	Fraction of active biomass contributing to biomass debris, traditional approach	—	3.53
f'_D	Fraction of active biomass contributing to biomass debris, lysis:regrowth approach	—	3.60
f_e	Fraction of electron donor used for energy	—	3.14
f_{MP}	Fraction of active biomass contributing to biomass associated products, traditional approach	—	3.74
f'_{MP}	Fraction of active biomass contributing to biomass associated products, lysis:regrowth approach	—	below 3.75
f_{NH}	Fraction of the nitrification rate that would occur in an RBC biofilm in the absence of carbon oxidation	—	20.1
$f_{NO,D}$	Fraction of nitrate-N denitrified in an initial anoxic zone	—	12.7
$f_{NO,R}$	Fraction of nitrate-N recirculated to an initial anoxic zone	—	12.8
f_s	Fraction of electron donor captured through synthesis	—	3.14
f_{sub}	Fraction of the total disc surface area in an RDR that is submerged	—	17.36
$f_{V,A}$	Fraction of system volume in which nitrification occurs	—	11.36
$f_{V,ANX}$	Fraction of system volume that is anoxic	—	12.17
$f_{V,SS}$	Fraction of system volume in which biomass synthesis occurs on readily biodegradable substrate	—	11.32

(CONTINUED)

Symbol	Definition	Units	Place or Equation Where First Used
$f_{V,XS}$	Fraction of system volume in which biomass synthesis occurs on slowly biodegradable substrate	—	11.28
$f_{XA,C}$	Fraction of autotrophic biomass in the contact tank of a CSAS system	—	11.49
f_{XBi}	Fraction of biomass density in a biofilm due to species i	—	Section 16.5
f_{XBias}	Fraction of biomass density at the attachment surface of a biofilm due to species i	—	16.59
$f_{XM,AER}$	Fraction of MLSS maintained under aerobic conditions	—	12.1
$f_{XM,ANA}$	Fraction of MLSS maintained under anaerobic conditions	—	12.3
$f_{XM,ANX}$	Fraction of MLSS maintained under anoxic conditions	—	12.2
$f_{XM,C}$	Fraction of MLSS in the contact tank of a CSAS system	—	11.46
$f_{XM,F}$	Fraction of the system MLSS in the fictitious bioreactor	—	11.31
$f_{XM,i}$	Fraction of MLSS in tank i of a chain of CSTRs	—	11.43
$f_{XM,N}$	Fraction of MLSS in the last tank of a chain of CSTRs	—	11.42
$f_{XS,H}$	Fraction of additional slowly biodegradable substrate oxidized during a transient load	—	11.12
F	Volumetric flow rate	L^3T^{-1}	4.3
F_a	Total flow rate applied to a packed tower	L^3T^{-1}	Figure 17.1
F_c	Volume applied to an SBR each cycle	L^3	7.11
F_L	Effective flow rate of liquid film on the aerated sector of an RDR	L^3T^{-1}	17.36
F_O	Influent volumetric flow rate	L^3T^{-1}	4.3
F_r	Volumetric flow rate of biomass recycle (RAS) stream	L^3T^{-1}	5.81
F_R	Volumetric flow rate of recirculation stream	L^3T^{-1}	14.5
F_w	Volumetric flow rate of biomass wastage stream	L^3T^{-1}	Figure 5.1 and Equation 5.1
$F_{w,i}$	Volumetric flow rate of biomass wastage stream from bioreactor i	L^3T^{-1}	7.1
$F(t)$	Fraction of fluid elements leaving a reactor that have residence times less than t	—	4.16
g	Gravitational acceleration	LT^{-2}	18.1
G	Root-mean-square velocity gradient	T^{-1}	11.6
Ga	Galileo number	—	18.13
h	Depth of liquid above diffuser	L	11.6
H	Depth of a suspended growth bioreactor	L	11.37
H′	Heat transfer coefficient	$MT^{-3}{}^{\circ}F^{-1}$	15.3
H_{Bb}	Height of a fluidized bed containing bioparticles	L	18.12
H_{Bbc}	Calculated height of a fluidized bed containing bioparticles	L	Figure 18.9
H_{Bp}	Height of a fluidized bed containing clean carrier particles	L	18.3
$H_{c,XOC}$	Dimensionless Henry's law coefficient for an XOC	—	22.2
H_i	Henry's law coefficient for constituent i	Atm L^3Mol^{-1}	8.4
H_{XOC}	Henry's law coefficient for an XOC	Atm L^3Mol^{-1}	22.4
H_{Rb}	Reference bed height of a fluidized bed containing bioparticles	L	18.11
H_{Rp}	Reference bed height of a fluidized bed containing clean carrier particles	L	18.3
$i_{N/XB}$	Mass of nitrogen per mass of COD in active biomass	$M_N M_{COD}^{-1}$	3.55
$i_{N/XD}$	Mass of nitrogen per mass of COD in biomass debris	$M_N M_{COD}^{-1}$	3.55

(*Continued*)

(CONTINUED)

Symbol	Definition	Units	Place or Equation Where First Used
$i_{O/EA}$	Mass of COD per mass of electron acceptor	$M_{COD}M_{EA}^{-1}$	10.5
$i_{O/XB,T}$	Mass of COD per mass of biomass measured as TSS	$M_{COD}M_{TSS}^{-1}$	5.10
$i_{O/XB,V}$	Mass of COD per mass of biomass measured as VSS	$M_{COD}M_{VSS}^{-1}$	5.11
$i_{O/XM,T}$	Mass of COD per mass of MLSS measured as TSS	$M_{COD}M_{TSS}^{-1}$	Section 9.5.1
$i_{O/XM,V}$	Mass of COD per mass of MLSS measured as VSS	$M_{COD}M_{VSS}^{-1}$	Section 9.5.1
$i_{P/XB}$	Mass of phosphorus per mass of COD in active biomass	$M_PM_{COD}^{-1}$	3.89
$I_{H_2,kinetic}$	Hydrogen switching function	—	8.7
$I_{H_2,thermo}$	Thermodynamic feasibility switch	—	8.10
I_{pH}	pH inhibition function	—	8.11
J_{NH}	Flux of ammonia-N	$ML^{-2}T^{-1}$	19.5
$J_{NH,max}$	Maximum flux of ammonia-N	$ML^{-2}T^{-1}$	19.5
J_S	Flux of substrate	$ML^{-2}T^{-1}$	16.1
$J_{S,aer}$	Flux of substrate into the biofilm in the aerated sector of an RDR	$ML^{-2}T^{-1}$	17.40
$J_{S,sub}$	Flux of substrate into the biofilm in the submerged sector of an RDR	$ML^{-2}T^{-1}$	17.36
J_{Si}	Flux of substrate into the biofilm in the i^{th} CSTR in a chain of N CSTRs	$ML^{-2}T^{-1}$	17.2
J_{SR}	Reference flux for normalized loading curves	$ML^{-2}T^{-1}$	Section 16.2.3
J_S^*	Dimensionless substrate flux	—	16.31
J_{SR}^*	Dimensionless reference flux for normalized loading curves	—	Section 16.2.3
$J_{S,deep}^*$	Dimensionless substrate flux into a deep biofilm	—	16.29
$J_{S,first}^*$	Dimensionless substrate flux into a first-order biofilm	—	16.48
$J_{S,fp}^*$	Dimensionless substrate flux into a fully penetrated biofilm	—	16.43
$J_{S,zero}^*$	Dimensionless substrate flux into a zero-order biofilm	—	16.50
J_X	Flux of biomass in a biofilm	$ML^{-2}T^{-1}$	16.54
k	Rate coefficient	Varies	3.95
k	BOD first-order rate coefficient	T^{-1}	9.27
k	Empirical coefficient in the modified Velz/Germain equation for a trickling filter	$M^{-1}L^3T^{-1}$	19.6
k_1	First-order reaction rate coefficient	T^{-1}	20.3
k_2	Second-order reaction rate coefficient	$M^{-1}L^3T^{-1}$	20.5
k_a	Ammonification rate coefficient	$M^{-1}LT^{-1}$	3.79
k_{dis}	Disintegration rate coefficient	$M^{-1}L^3T^{-1}$	8.5
k_e	Mean reaction rate coefficient in COD units	$M^{-1}L^3T^{-1}$	3.45
$k_{e,T}$	Mean reaction rate coefficient in TSS units	$M^{-1}L^3T^{-1}$	5.51
$k_{G,XOC}a$	Gas film mass transfer coefficient for an XOC	T^{-1}	22.9
k_h	Hydrolysis coefficient	T^{-1}	3.77
$k_{h,comp}$	Hydrolysis coefficient for a given biochemical component in anaerobic systems	T^{-1}	8.6
$k_{h,overall}$	Combined overall coefficient for disintegration and hydrolysis	T^{-1}	Section 8.2.1
k_L	Liquid phase mass transfer coefficient	LT^{-1}	16.2
k_L^*	Dimensionless liquid phase mass transfer coefficient	—	16.40
k_{Ls}	Liquid phase mass transfer coefficient in the submerged sector of an RDR	LT^{-1}	17.30
$k_{L,XOC}a$	Liquid film mass transfer coefficient for an XOC	T^{-1}	22.9

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Symbol	Definition	Units	Place or Equation Where First Used
k_{OW}	Octanol:water partition coefficient	—	22.15
$k_{s,XOC}$	Sorption coefficient for an XOC	L^3M^{-1}	22.13
K	Equilibrium constant	—	8.8
K	Empirical coefficient in Velz model for a packed tower	—	17.19
K	Empirical coefficient in the modified Velz/Germain equation for a trickling filter	Varies	19.8
K_1	Empirical coefficient in the Eckenfelder model for a packed tower	—	17.22
K_2	Empirical coefficient in the Kornegay model for a packed tower	—	17.24
K_3	Empirical coefficient	—	17.26
K_A	Half-saturation coefficient for acetate	ML^{-3}	3.82
K_{AA}	Half-saturation coefficient for amino acids	ML^{-3}	8.11
K_g	Pseudo half-saturation coefficient in the Kornegay model	ML^{-3}	17.24
$K_{g,NH}$	Pseudo half-saturation coefficient for ammonia-N	ML^{-3}	19.5
K_I	Inhibition coefficient	ML^{-3}	3.39
K_{I,H_2}	Inhibition coefficient for H_2 against anaerobic oxidation	ML^{-3}	8.7
K_{IO}	Inhibition coefficient for oxygen against denitrification	ML^{-3}	3.48
K_{IPP}	Inhibition coefficient for Poly-P storage	$M_PM_{COD}^{-1}$	3.87
$K_{L,O_2}a$	Overall liquid phase mass transfer coefficient for oxygen	T^{-1}	22.8
$K_{L,XOC}a$	Overall liquid phase mass transfer coefficient for removal of an XOC by volatilization	T^{-1}	22.1
K_{NH}	Half-saturation coefficient for ammonia-N	ML^{-3}	3.51
K_{NO}	Half-saturation coefficient for nitrate-N	ML^{-3}	3.48
K_O	Half-saturation coefficient for oxygen	ML^{-3}	Section 3.2.9
$K_{O,A}$	Half-saturation coefficient for oxygen for autotrophs	ML^{-3}	Section 3.2.9
$K_{O,H}$	Half-saturation coefficient for oxygen for heterotrophs	ML^{-3}	Section 3.2.9
K_{PO_4}	Half-saturation coefficient for soluble phosphate	ML^{-3}	3.85
K_{PHA}	Half-saturation coefficient for PHA	MM^{-1}	3.85
K_{PMAX}	Maximum mass of Poly-P per unit mass of PAO	$M_PM_{COD}^{-1}$	3.87
K_{PP}	Half-saturation coefficient for Poly-P	ML^{-3}	3.82
K_{RQ}	Reaction quotient	—	8.9
K_S	Half-saturation coefficient for substrate	ML^{-3}	3.36
K_{S1}	Half-saturation coefficient for complementary nutrient No. 1	ML^{-3}	3.46
K_{S2}	Half-saturation coefficient for complementary nutrient No. 2	ML^{-3}	3.46
$K_{S,XOC}$	Half-saturation coefficient for growth on an XOC	ML^{-3}	5.22 in Chapter 22
K_X	Half-saturation coefficient for hydrolysis of slowly biodegradable substrate	MM^{-1}	3.77
L	Length of a plug-flow reactor; depth of a packed tower	L	4.12
L_f	Biofilm thickness	L	16.6
L_{fc}	Characteristic biofilm thickness	L	18.17
L_{fa}	Assumed biofilm thickness	L	Figure 18.9
L_f^*	Dimensionless biofilm thickness	—	16.43
L_w	Stagnant liquid film thickness	L	16.1
m	Exponent depicting the effect of an inhibitor on K_S	—	3.42
m	Empirical exponent in the Eckenfelder model	—	17.22

(*Continued*)

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Symbol	Definition	Units	Place or Equation Where First Used
M	Mass velocity	$ML^{-2}T^{-1}$	17.13
M_{O_2}	Mass of oxygen consumed per liter of batch reactor volume	ML^{-3}	9.20
$\mathbf{M}_p$	Mass of carrier particles in a fluidized bed	M	18.4
n	Exponent depicting the effect of an inhibitor on $\hat{\mu}$	—	3.42
n	Empirical exponent in the Eckenfelder model	—	17.22
n	Empirical exponent in the Richardson-Zaki equation	—	18.6
n	Empirical coefficient in the modified Velz/Germain equation for a trickling filter	—	19.7
n	Empirical exponent in Equation 22.10	—	22.10
n	Empirical exponent in the Freundlich equation	—	22.13
n	Number of stages in an RBC system	—	Section 20.3.1
N	Number of CSTRs in series	—	4.21
N_a	Number of arms on a trickling filter rotary distributor	—	19.4
N_{AA}	Nitrogen content of amino acids	$Mol \cdot M^{-1}$	8.11
N_{bac}	Nitrogen content of bacteria	$Mol \cdot M^{-1}$	8.11
N_c	Number of cycles per day applied to an SBR	T^{-1}	7.11
NR	Nitrogen requirement per unit of COD removed	$M_N M_{COD}^{-1}$	5.46
ON_{TO}	Influent total organic nitrogen concentration	ML^{-3}	9.35
$\overline{p}_{CO_2}$	Partial pressure of carbon dioxide	Atm	14.7
P	Power input	ML^2T^{-3}	11.1
P	Empirical coefficient	$L^{-1}T$	17.18
P′	Weight fraction moisture content of a biofilm	—	18.9
q	Specific substrate removal rate with biomass in COD units	T^{-1}	3.43
q_H	Specific substrate removal rate of heterotrophs with biomass in COD units	T^{-1}	16.17
$q_{NO/XB}$	Specific nitrate-N utilization rate associated with biomass decay	T^{-1}	12.12
$q_{NO/XS}$	Specific nitrate-N utilization rate associated with slowly biodegradable substrate utilization	T^{-1}	12.11
$\hat{q}$	Maximum specific substrate removal rate with biomass in COD units	T^{-1}	3.44
$\hat{q}_{AA}$	Maximum specific amino acid uptake rate with biomass in COD units	T^{-1}	8.11
$\hat{q}_H$	Maximum specific substrate removal rate of heterotrophs with biomass in COD units	T^{-1}	16.6
$\hat{q}_{PHA}$	Maximum specific rate of PHA formation with biomass and PHA expressed in COD units	T^{-1}	3.82
$\hat{q}_{PP}$	Maximum specific Poly-P storage rate with biomass in COD units	$M_P M_{COD}^{-1} T^{-1}$	3.87
Q	Air flow rate	L^3T^{-1}	11.2
r	Generalized reaction rate	$ML^{-3}T^{-1}$	3.10
r_A	Reaction rate for reactant A	$ML^{-3}T^{-1}$	4.3
r_i	Reaction rate for reactant A_i	$ML^{-3}T^{-1}$	3.10
r_j	Generalized reaction rate for reaction j	$ML^{-3}T^{-1}$	3.12
r_o	Outer radius of the submerged sector of an RDR	L	17.29

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Symbol	Definition	Units	Place or Equation Where First Used
$r_{s,XOC}$	Rate of loss of an XOC from an activated sludge system by sorption	MT^{-1}	22.14
r_{SA}	Reaction rate for acetate	$ML^{-3}T^{-1}$	3.83
r_{SMP}	Reaction rate for soluble microbial products	$ML^{-3}T^{-1}$	3.72
r_{SNH}	Reaction rate for ammonia-N	$ML^{-3}T^{-1}$	3.59
r_{SNS}	Reaction rate for soluble organic-N	$ML^{-3}T^{-1}$	3.78
r_{SO}	Reaction rate for dissolved oxygen	$ML^{-3}T^{-1}$	3.34
$r_{SO,i}$	Volumetric oxygen transfer rate in tank i	$ML^{-3}T^{-1}$	11.38
r_{SNO}	Reaction rate for nitrate-N	$ML^{-1}T^{-1}$	3.91
r_{SPO_4}	Reaction rate for soluble phosphate	$ML^{-3}T^{-1}$	3.84
$r_{SPO_4,anx}$	Anoxic reaction rate for soluble phosphate	$ML^{-3}T^{-1}$	3.93
r_{SS}	Reaction rate for readily biodegradable substrate	$ML^{-3}T^{-1}$	3.34
$r_{v,XOC}$	Rate of loss of an XOC by volatilization	$ML^{-3}T^{-1}$	22.1
r_{XB}	Reaction rate for active biomass	$ML^{-3}T^{-1}$	3.34
r_{XBPAO}	Reaction rate for PAO biomass	$ML^{-3}T^{-1}$	3.85
$r_{XBPAO,anx}$	Anoxic reaction rate for PAO biomass	$ML^{-3}T^{-1}$	3.90
r_{XC}	Reaction rate for composite particulate material	$ML^{-3}T^{-1}$	8.5
r_{Xcomp}	Reaction rate for a given biochemical component of particulate material	$ML^{-3}T^{-1}$	8.6
r_{XD}	Reaction rate for biomass debris	$ML^{-3}T^{-1}$	3.57
r_{XNS}	Reaction rate for particulate organic nitrogen	$ML^{-3}T^{-1}$	3.66
r_{XPHA}	Reaction rate for PHA	$ML^{-3}T^{-1}$	3.83
$r_{XPHA,anx}$	Anoxic reaction rate for PHA	$ML^{-3}T^{-1}$	3.91
r_{XPP}	Reaction rate for stored polyphosphate	$ML^{-3}T^{-1}$	3.84
$r_{XPP,anx}$	Anoxic reaction rate for stored polyphosphate	$ML^{-3}T^{-1}$	3.92
r_{XS}	Reaction rate for slowly biodegradable substrate	$ML^{-3}T^{-1}$	3.65
R	Gas constant	$kJMole^{-1}K^{-1}$	3.95
R	Overall stoichiometric equation	—	3.14
R_a	Half reaction for the electron acceptor	—	3.14
R_c	Half reaction for cell material	—	3.14
R_d	Half reaction for the electron donor	—	3.14
Re	Reynolds number	—	17.13
Re_t	Terminal Reynolds number	—	18.10
REA	Mass rate of electron acceptor utilization	MT^{-1}	10.5
Ri	Rittmann number	—	16.34
$RN_{A,max}$	Maximum mass nitrification rate	MT^{-1}	11.35
RO	Mass rate of oxygen utilization	MT^{-1}	5.42
RO_A	Steady-state oxygen requirement for autotrophs	MT^{-1}	6.2
$RO_{A,C}$	Autotrophic oxygen requirement in the contact tank of a CSAS system	MT^{-1}	Example 11.3.5.3
$RO_{A,D}$	Autotrophic oxygen requirement for biomass decay	MT^{-1}	11.34
$RO_{A,D,i}$	Autotrophic oxygen requirement for biomass decay in tank i	MT^{-1}	11.39
$RO_{A,D,s}$	Autotrophic oxygen requirement for biomass decay in a selector	MT^{-1}	Example 11.3.4.3
$RO_{A,max}$	Maximum mass rate of oxygen utilization by autotrophs	MT^{-1}	11.15
$RO_{A,S}$	Autotrophic oxygen requirement in the stabilization basin of a CSAS system	MT^{-1}	Example 11.3.5.3
$RO_{A,SN}$	Autotrophic oxygen requirement for biomass synthesis	MT^{-1}	11.33

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Symbol	Definition	Units	Place or Equation Where First Used
$RO_{A,SN,C}$	Autotrophic oxygen requirement for biomass synthesis in the contact tank of a CSAS system	MT^{-1}	11.52
$RO_{A,SN,i}$	Autotrophic oxygen requirement for biomass synthesis in tank i	MT^{-1}	11.39
$RO_{A,SN,s}$	Autotrophic oxygen requirement for biomass synthesis in a selector	MT^{-1}	Example 10.3.4.3
$RO_{A,SN,S}$	Autotrophic oxygen requirement for biomass synthesis in the stabilization basin of a CSAS system	MT^{-1}	11.53
$RO_{A,TS}$	Transient-state oxygen requirement for autotrophs	MT^{-1}	11.13
$RO_{A,XNS}$	Autotrophic oxygen requirement for biomass synthesis from particulate organic nitrogen	MT^{-1}	11.54
$RO_{A,XNS,C}$	Autotrophic oxygen requirement for biomass synthesis from particulate organic nitrogen in the contact tank of a CSAS system	MT^{-1}	Example 11.3.5.3
$RO_{A,XNS,S}$	Autotrophic oxygen requirement for biomass synthesis from particulate organic nitrogen in the stabilization basin of a CSAS system	MT^{-1}	Example 11.3.5.3
RO_{C}	Oxygen requirement in the contact tank of a CSAS system	MT^{-1}	Example 11.3.5.3
RO_{H}	Steady-state oxygen requirement for heterotrophs	MT^{-1}	10.10
$RO_{H,C}$	Heterotrophic oxygen requirement in the contact tank of a CSAS system	MT^{-1}	Example 11.3.5.3
$RO_{H,D}$	Heterotrophic oxygen requirement for biomass decay	MT^{-1}	11.26
$RO_{H,D,C}$	Heterotrophic oxygen requirement for biomass decay in the contact tank of a CSAS system	MT^{-1}	Example 11.3.5.3
$RO_{H,D,i}$	Heterotrophic oxygen requirement for biomass decay in tank i	MT^{-1}	11.27
$RO_{H,D,s}$	Heterotrophic oxygen requirement for biomass decay in a selector	MT^{-1}	Example 11.3.4.3
$RO_{H,D,S}$	Heterotrophic oxygen requirement for biomass decay in the stabilization basin of a CSAS system	MT^{-1}	Example 11.3.5.3
$RO_{H,S}$	Heterotrophic oxygen requirement in the stabilization basin of a CSAS system	MT^{-1}	Example 11.3.5.3
$RO_{H,SS}$	Heterotrophic oxygen requirement for biomass synthesis from readily biodegradable substrate	MT^{-1}	11.24
$RO_{H,SS,i}$	Heterotrophic oxygen requirement for biomass synthesis from readily biodegradable substrate in tank i	MT^{-1}	11.39
$RO_{H,TS}$	Transient state oxygen requirement for heterotrophs	MT^{-1}	11.12
$RO_{H,XS}$	Heterotrophic oxygen requirement for biomass synthesis from slowly biodegradable substrate	MT^{-1}	11.25
$RO_{H,XS,C}$	Heterotrophic oxygen requirement for biomass synthesis from slowly biodegradable substrate in the contact tank of a CSAS system	MT^{-1}	Example 11.3.5.3
$RO_{H,XS,i}$	Heterotrophic oxygen requirement for biomass synthesis from slowly biodegradable substrate in tank i	MT^{-1}	11.39
$RO_{H,XS,s}$	Heterotrophic oxygen requirement for biomass synthesis from slowly biodegradable substrate in a selector	MT^{-1}	Example 11.3.4.3
$RO_{H,XS,S}$	Heterotrophic oxygen requirement for biomass synthesis from slowly biodegradable substrate in the stabilization basin of a CSAS system	MT^{-1}	Example 11.3.5.3
RO_{i}	Oxygen requirement in tank i	MT^{-1}	11.38
RO_{P}	Peak oxygen requirement for a system	MT^{-1}	11.20
RO_{s}	Oxygen requirement in a selector	MT^{-1}	Example 11.3.4.3
RO_{S}	Oxygen requirement in the stabilization basin of a CSAS system	MT^{-1}	Example 11.3.5.3
$RX_{I,T}$	Input rate of inert organic solids	MT^{-1}	Equation 15.13

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Symbol	Definition	Units	Place or Equation Where First Used
S	Soluble constituent concentration	ML^{-3}	Section 5.1.1
S_A	Acetate concentration	ML^{-3}	3.82
S_{AA}	Amino acid concentration	ML^{-3}	8.11
S_{ALK}	Alkalinity	Mole	Table 6.1
S_{BAlk}	Bicarbonate alkalinity concentration	ML^{-3}	14.7
Sc	Schmidt number	—	17.13
S_C	Soluble COD concentration	ML^{-3}	Section 9.2.1
S_{CO}	Influent soluble COD concentration	ML^{-3}	Section 9.2.1
S_D	Biomass associated product concentration	ML^{-3}	Section 2.4.7
S_{EPS}	Extracellular polymeric substance	ML^{-3}	Section 2.4.7
S_{H_2}	Concentration of H_2 in the liquid phase	ML^{-3}	8.7
S_i	Concentration of constituent i	Mol L^{-3}	8.4
S_i	Inhibitory chemical concentration	ML^{-3}	3.42
S_i^*	Inhibitor concentration that causes all microbial activity to cease	ML^{-3}	3.42
S_I	Soluble inert organic matter concentration	ML^{-3}	5.54
S_{IO}	Influent soluble inert organic matter concentration	ML^{-3}	5.54
S_{MeOH}	Methanol concentration	ML^{-3}	21.5
S_{MP}	Soluble microbial product concentration	ML^{-3}	Section 2.4.7
$S_{N,a}$	Influent nitrogen concentration available to the nitrifying bacteria	ML^{-3}	11.16
$S'_{N,a}$	Influent soluble nitrogen concentration available to the nitrifying bacteria	ML^{-3}	11.51
S_{NH}	Ammonia-N concentration	ML^{-3}	3.51
S_{NHC}	Ammonia-N concentration in the contact tank of a CSAS system	ML^{-3}	11.49
S_{NHO}	Influent ammonia-N concentration	ML^{-3}	6.2
S_{NIO}	Influent soluble inert organic-N concentration	ML^{-3}	9.36
S_{NO}	Nitrate-N concentration	ML^{-3}	3.48
S_{NOO}	Influent nitrate-N concentration	ML^{-3}	6.3
S_{NO_2O}	Influent nitrite-N concentration	ML^{-3}	21.5
S_{NS}	Soluble organic-N concentration	ML^{-3}	3.78
S_{NSO}	Influent soluble organic-N concentration	ML^{-3}	9.23
S_O	Dissolved oxygen concentration	ML^{-3}	3.31
S_{OO}	Influent dissolved oxygen concentration	ML^{-3}	21.5
S_S	Readily biodegradable (soluble) substrate concentration	ML^{-3}	Section 2.4.7
S_S^*	Concentration of an inhibitory substrate giving μ^*	ML^{-3}	3.41
S_{S1}	Complementary substrate No. 1	ML^{-3}	3.46
S_{S2}	Complementary substrate No. 2	ML^{-3}	3.46
S_{Sa}	Applied substrate concentration	ML^{-3}	17.1
S_{Sb}	Bulk liquid phase substrate concentration	ML^{-3}	16.1
S_{Sb}^*	Dimensionless bulk liquid phase substrate concentration	—	16.39
S_{Sbi}	Bulk liquid phase concentration of the substrate being used by species i in a biofilm; Bulk liquid phase concentration of substrate in the i^{th} CSTR in a chain of N CSTRs	ML^{-3}	Section 16.4; 17.2
S_{Sbi}^*	Dimensionless bulk liquid phase concentration of substrate in the i^{th} CSTR in a chain of N CSTRs	—	16.39a in Section 17.1.2

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Symbol	Definition	Units	Place or Equation Where First Used
S_{Sbi-1}	Bulk liquid phase concentration of substrate entering the i^{th} CSTR in a chain of N CSTRs	ML^{-3}	17.2
S_{Sbmin}	Minimum bulk substrate concentration required to maintain a steady-state biofilm	ML^{-3}	16.22
S^*_{Sbmin}	Dimensionless minimum bulk substrate concentration required to maintain a steady-state biofilm	—	16.33
$S_{Sbmin2C}$	Minimum bulk liquid concentration of substrate 2 required to allow coexistence of two species in a biofilm	ML^{-3}	16.61
S_{SCMAL}	Soluble substrate concentration leaving a CMAL	ML^{-3}	15.15
S_{Se}	Effluent substrate concentration	ML^{-3}	17.1
S_{Sf}	Substrate concentration at the liquid-biofilm interface	ML^{-3}	Figure 16.4
S_{SF}	Readily biodegradable substrate concentration in a fictitious CSTR	ML^{-3}	11.29
S_{Si}	Readily biodegradable (soluble) substrate concentration in bioreactor i; Electron donor used by microbial species i in a biofilm	ML^{-3}	7.5; Section 16.4
S_{SLR}	Soluble substrate concentration in the liquid film on the aerated sector of an RDR at its point of return to the tank	ML^{-3}	17.36
S_{Smin}	Minimum attainable soluble substrate concentration from a CSTR	ML^{-3}	5.23
$S_{S,n}$	Soluble substrate concentration in stage N of a multistage RBC	ML^{-3}	20.5
S_{SN}	Soluble substrate concentration in the N^{th} cell of a multicell benthal stabilization basin	ML^{-3}	15.15
S_{SO}	Influent readily biodegradable substrate concentration	ML^{-3}	Figure 5.1 and Equation 5.24
S_{SOi}	Concentration of readily biodegradable substrate entering bioreactor i	ML^{-3}	7.5
S_{Ss}	Substrate concentration at the liquid-biofilm interface	ML^{-3}	16.1
S^*_{Ss}	Dimensionless substrate concentration at the liquid-biofilm interface	—	16.29
S_{Ssi}	Substrate concentration at the liquid-biofilm interface in the i^{th} CSTR of a chain of N CSTRs	ML^{-3}	16.30a in Section 17.1.2
S^*_{Ssi}	Dimensionless substrate concentration at the liquid-biofilm interface in the i^{th} CSTR of a chain of N CSTRs	—	16.38a in Section 17.1.2
S_T	Tracer concentration	ML^{-3}	4.18
S_{TAlk}	Total alkalinity concentration	ML^{-3}	14.9
S_{TO}	Influent tracer concentration	ML^{-3}	4.18
S_{TN}	Tracer concentration leaving the N^{th} reactor	ML^{-3}	4.21
S_{VFA}	Volatile fatty acid concentration	ML^{-3}	14.9
S_{XOC}	Liquid phase concentration of an XOC	ML^{-3}	22.1
$S_{XOC,O}$	Influent liquid phase concentration of an XOC	ML^{-3}	22.17
S^*_{XOC}	Concentration of an XOC that would exist in the liquid phase if it were in equilibrium with the gas phase	ML^{-3}	22.1
SK	Spülkraft	L	19.4
t	Time	T	4.3
T	Temperature	°C, K	3.95
T_{max}	Empirical temperature coefficient	°C	3.102
T_{min}	Empirical temperature coefficient	°C	3.102
T_A	Weekly average air temperature	°F	15.3

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Symbol	Definition	Units	Place or Equation Where First Used
T_i	Weekly average influent wastewater temperature	°F	15.3
T_L	Weekly average lagoon temperature	°F	15.3
TO	Rate of oxygen transfer	MT^{-1}	12.15
u	Temperature coefficient in the Arrhenius equation	$kJMole^{-1}$	3.95
U	Process loading factor	T^{-1}	5.48
U_{ANX}	Process loading factor for the anoxic zone	T^{-1}	12.11
U_i	Process loading factor in bioreactor i	T^{-1}	7.6
U_s	Process loading factor in a selector	T^{-1}	11.40
v	Velocity	LT^{-1}	4.23
v_{mf}	Minimum fluidization velocity	LT^{-1}	18.1
v_t	Terminal settling velocity	LT^{-1}	18.2
v_{tb}	Terminal settling velocity of a bioparticle	LT^{-1}	Figure 18.7
v_{tp}	Terminal settling velocity of a carrier particle	LT^{-1}	Figure 18.7
V	Volume	L^3	4.3
V_{aer}	Volume of liquid in the aerated sector of an RDR	L^3	17.37
V_{AER}	Volume of aerobic zone in MLE system	L^3	Example 12.3.1.2
V_{ANX}	Volume of anoxic zone in MLE system	L^3	Example 12.3.1.2
$V_{ANX,2}$	Volume of second anoxic zone in a four-stage Bardenpho system	L^3	Example 12.3.1.4
V_{br}	Volume of biomass retained per cycle in an SBR	L^3	7.12
$V_{br,min}$	Minimum possible volume of biomass retained per cycle in an SBR	L^3	11.59
V_C	Volume of contact tank in CSAS system	L^3	11.45
V_F	Volume of a fictitious CSTR	L^3	11.30
V_i	Volume of bioreactor i	L^3	7.1
V_L	Lower feasible bioreactor volume; Total liquid volume in a packed tower	L^3	Figure 10.6; Section 17.1.2
V_{Li}	Liquid volume in the i^{th} CSTR of a chain of N CSTRs used to represent a packed tower	L^3	17.2
$V_{L,FS}$	Lower feasible bioreactor volume based on floc shear	L^3	11.4
$V_{L,OT}$	Lower feasible bioreactor volume based on oxygen transfer	L^3	11.5
V_M	Volume of media in a packed tower or trickling filter	L^3	Section 17.1.2
V_{MLSS}	Volume of MLSS used in a batch test	L^3	9.20
V_{nr}	Volume of nitrate containing fluid retained per cycle in an SBR	L^3	7.13
V_N	Volume of the last tank in a chain of CSTRs	L^3	11.41
V_s	Volume of a selector	L^3	11.40
V_S	Volume of the stabilization basin in a CSAS system	L^3	11.48
V_T	Total volume of a chain of CSTRs	L^3	7.2
V_U	Upper feasible bioreactor volume	L^3	Figure 10.6
V_{WW}	Volume of wastewater used in a batch test	L^3	9.20
W	Width of a suspended growth bioreactor	L	11.37
$W_{M,T}$	Mass wastage rate of MLSS in TSS units	MT^{-1}	5.60
W_{total}	Mass wastage rate of total biomass in COD units	MT^{-1}	Example 5.2.2.1
$W_{total,T}$	Mass wastage rate of total biomass in TSS units	MT^{-1}	5.39

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Symbol	Definition	Units	Place or Equation Where First Used
x	Distance from a reference point	L	4.7
X	Particulate constituent concentration	ML^{-3}	Section 5.1.1
X_B	Active biomass concentration	ML^{-3}	3.35
$X_{B,A}$	Active autotrophic biomass concentration in COD units	ML^{-3}	Table 6.1
$X_{B,AA}$	Amino acid consuming biomass concentration in COD units	ML^{-3}	8.11
$X_{B,A,T}$	Active autotrophic biomass concentration in TSS units	ML^{-3}	11.15
$X_{B,A,V}$	Active autotrophic biomass concentration in VSS units	ML^{-3}	13.3
$X_{B,H}$	Active heterotrophic biomass concentration in COD units	ML^{-3}	3.31
$X_{B,Hb}$	Active heterotrophic biomass concentration (in COD units) in the bulk liquid phase of a biofilm reactor	ML^{-3}	16.17
$X_{B,He}$	Active heterotrophic biomass concentration (in COD units) in the effluent from a bioreactor	ML^{-3}	Figure 17.1
$X_{B,Hf}$	Active heterotrophic biomass concentration (in COD units) in a biofilm	ML^{-3}	16.6
$X_{B,Hi}$	Active heterotrophic biomass concentration (in COD units) in suspension in the i^{th} reactor of a chain of N CSTRs	ML^{-3}	17.2
$X_{B,HO}$	Initial or influent active heterotrophic biomass concentration in COD units	ML^{-3}	Section 9.4.2
$X_{B,H,T}$	Active heterotrophic biomass concentration in TSS units	ML^{-3}	5.10
$X_{B,H,T,TFE}$	Active heterotrophic biomass concentration in trickling filter effluent in TSS units	ML^{-3}	19.11
$X_{B,H,TO}$	Influent active heterotrophic biomass concentration in TSS units	ML^{-3}	5.62
$X_{B,H,V}$	Active heterotrophic biomass concentration in VSS units	ML^{-3}	5.11
$X_{B,H,VO}$	Influent active heterotrophic biomass concentration in VSS units	ML^{-3}	13.21
$X_{B,if}$	Density of species i (in COD units) at some point in a biofilm	ML^{-3}	16.51
$X_{B,PAO}$	PAO biomass concentration	ML^{-3}	3.82
$X_{B,XOC,T}$	Concentration of biomass capable of degrading an XOC	ML^{-3}	22.17
$X'_{B,XOC,T}$	Concentration of biomass capable of degrading an XOC that would be present in the absence of abiotic removal mechanisms	ML^{-3}	5.28 in Chapter 22
X_{Be}	Biomass concentration in the effluent from a packed tower	ML^{-3}	Figure 17.1
X_{Bf}	Density of total biomass (in COD units) in a biofilm	ML^{-3}	16.51
X_C	Composite particulate materials	ML^{-3}	Section 8.1.1
X_{comp}	A specific biochemical component of particulate material	ML^{-3}	8.6
X_D	Biomass debris concentration in COD units	ML^{-3}	Section 2.4.7
$X_{D,T}$	Biomass debris concentration in TSS units	ML^{-3}	Section 5.1.4
$X_{D,TO}$	Influent biomass debris concentration in TSS units	ML^{-3}	5.65
$X_{D,V}$	Biomass debris concentration in VSS units	ML^{-3}	Section 5.1.4
$X_{D,VO}$	Influent biomass debris concentration in VSS units	ML^{-3}	13.21
X_{Df}	Density of biomass debris in a biofilm	ML^{-3}	Section 16.4
X_{FO}	Influent fixed solids concentration	ML^{-3}	Section 9.6
X_I	Particulate inert organic matter concentration in COD units	ML^{-3}	Section 5.2.2
$X_{I,T}$	Particulate inert material concentration in TSS units	ML^{-3}	5.55
$X_{I,V}$	Particulate, inert organic matter concentration in VSS units	ML^{-3}	13.3
$X_{I,VO}$	Influent particulate, inert organic matter concentration in VSS units	ML^{-3}	13.21
X_{IO}	Influent particulate inert organic matter concentration in COD units	ML^{-3}	Section 5.2.2
$X_{IO,T}$	Influent particulate inert material concentration in TSS units	ML^{-3}	5.55

(CONTINUED)

Symbol	Definition	Units	Place or Equation Where First Used
X_M	MLSS concentration in COD units	ML^{-3}	Section 5.2.2
$X_{M,F}$	FSS concentration	ML^{-3}	13.15
$X_{M,FO}$	Influent concentration of FSS	ML^{-3}	13.16
$X_{M,T}$	MLSS concentration in TSS units	ML^{-3}	5.58
$X_{M,T,ANA}$	MLSS concentration in TSS units in the anaerobic zone of a BNR system	ML^{-3}	12.19
$X_{M,T,ANX}$	MLSS concentration in TSS units in the anoxic zone of a BNR system	ML^{-3}	12.19
$X_{M,T,C}$	MLSS concentration in TSS units in the contact tank of a CSAS system	ML^{-3}	11.45
$X_{M,T,e}$	MLSS concentration in effluent stream in TSS units	ML^{-3}	9.1
$X_{M,T,F}$	MLSS concentration in TSS units in a fictitious CSTR	ML^{-3}	11.31
$X_{M,T,i}$	MLSS concentration in TSS units in tank i of a chain of CSTRs	ML^{-3}	7.1
$X_{M,T,L}$	Lower limit on feasible MLSS concentrations in TSS units	ML^{-3}	Example 10.3.3.3
$X_{M,T,N}$	MLSS concentration in TSS units in the last tank of a chain of CSTRs	ML^{-3}	11.41
$X_{M,T,r}$	MLSS concentration in biomass recycle stream in TSS units	ML^{-3}	5.81
$X_{M,T,r,max}$	Maximum attainable MLSS concentration in biomass recycle stream in TSS units	ML^{-3}	11.58
$X_{M,T,U}$	Upper limit on feasible MLSS concentrations in TSS units	ML^{-3}	Example 10.3.3.3
$X_{M,T,w}$	MLSS concentration in biomass wastage stream in TSS units	ML^{-3}	5.81
$X_{M,V}$	MLSS concentration in VSS units	ML^{-3}	13.1
$X_{M,VO}$	Influent MLSS concentration in VSS units	ML^{-3}	13.4
$X_{M,V,b}$	Biodegradable MLSS concentration in VSS units	ML^{-3}	13.1
$X_{M,V,bO}$	Influent biodegradable MLSS concentration in VSS units	ML^{-3}	13.5
$X_{M,V,n}$	Nonbiodegradable MLSS concentration in VSS units	ML^{-3}	13.1
$X_{M,V,nO}$	Influent nonbiodegradable MLSS concentration in VSS units	ML^{-3}	13.4
X_{NIO}	Influent inert organic-N concentration in COD units	ML^{-3}	9.36
X_{NS}	Particulate organic-N concentration	ML^{-3}	3.66
X_{NSO}	Particulate organic-N concentration in wastewater	ML^{-3}	9.23
X_{PN}	Photosynthetic microorganism concentration in N^{th} tank	ML^{-3}	15.14
X_{PO}	Photosynthetic microorganism concentration in influent	ML^{-3}	15.14
X_{PP}	Polyphosphate concentration in biomass	ML^{-3}	3.82
X_{PHA}	PHA concentration in biomass	ML^{-3}	3.85
X_S	Slowly biodegradable (particulate) substrate concentration in COD units	ML^{-3}	Section 2.4.7
X_{SO}	Slowly biodegradable (particulate) substrate concentration in wastewater in COD units	ML^{-3}	9.21
$X_{S,V}$	Slowly biodegradable (particulate) substrate concentration in VSS units	ML^{-3}	13.3
X_{total}	Total biomass concentration in COD units	ML^{-3}	Section 5.1.5
$X_{total,e}$	Total biomass concentration in COD units in effluent	ML^{-3}	17.16
$X_{total,O}$	Total biomass concentration in COD units in influent	ML^{-3}	17.16
$X_{total,T}$	Total biomass concentration in TSS units	ML^{-3}	5.34
$X_{total,Te}$	Total biomass concentration in TSS units in effluent	ML^{-3}	Section 9.2.1

(*Continued*)

(CONTINUED)

Symbol	Definition	Units	Place or Equation Where First Used
$X_{total,Tw}$	Total biomass concentration in TSS units in wastage stream	ML^{-3}	Section 9.2.1
X_w	Particulate constituent concentration in the wastage stream	ML^{-3}	5.1
Y	True growth yield	$M_{COD}M_{COD}^{-1}$	Section 2.4.1
Y_A	True growth yield for autotrophs (AOB + NOB) on COD/N basis	$M_{COD}M_N^{-1}$	3.27
Y_{AA}	True growth yield of amino acid consuming bacteria on COD basis	$M_{COD}M_{COD}^{-1}$	8.11
Y_{AOB}	True growth yield for ammonia oxidizing bacteria (AOB) on COD/N basis	$M_{COD}M_N^{-1}$	3.25
$Y_{A,T}$	True growth yield for autotrophs on TSS/N basis	$M_{TSS}M_N^{-1}$	6.2
Y_H	True growth yield for heterotrophs on COD/COD basis	$M_{COD}M_{COD}^{-1}$	3.16
$Y_{H,T}$	True growth yield for heterotrophs on TSS/COD basis	$M_{TSS}M_{COD}^{-1}$	5.14
Y_{Hobs}	Observed yield for heterotrophs on COD/COD basis	$M_{COD}M_{COD}^{-1}$	16.6
$Y_{Hobs,T}$	Observed yield for heterotrophs on TSS/COD basis	$M_{TSS}M_{COD}^{-1}$	5.37
$Y_{H,V}$	True growth yield for heterotrophs on VSS/COD basis	$M_{VSS}M_{COD}^{-1}$	5.14
Y_{MP}	Microbial product yield	$M_{COD}M_{COD}^{-1}$	Section 2.4.3
Y_{NOB}	True growth yield for nitrite oxidizing bacteria on COD/N basis	$M_{COD}M_N^{-1}$	3.26
$Y_{n,T}$	Net process yield on TSS/COD basis	$M_{TSS}M_{COD}^{-1}$	10.2
$Y_{n,V}$	Net process yield on VSS/COD basis	$M_{VSS}M_{COD}^{-1}$	Section 10.4.1
Y_{obs}	Observed yield	$M_{COD}M_{COD}^{-1}$	Section 2.4.2
Y_{O_2}	Process oxygen stoichiometric coefficient	$M_{O_2}M_{BOD5}^{-1}$	10.4
$Y_{O_2,d}$	Oxygen stoichiometric coefficient for decay	$M_{O_2}M_{VSS}^{-1}$	19.12
$Y_{O_2,s}$	Oxygen stoichiometric coefficient for synthesis	$M_{O_2}M_{BOD5}^{-1}$	19.12
$Y_{O_2,SG}$	Oxygen stoichiometric coefficient for the suspended growth bioreactor in a TF/AS system	$M_{O_2}M_{BOD5}^{-1}$	19.14
$Y_{O_2,TF}$	Oxygen stoichiometric coefficient for a trickling filter	$M_{O_2}M_{BOD5}^{-1}$	19.13
Y_{PAO}	True growth yield for PAOs	$M_{COD}M_{COD}^{-1}$	3.86
Y_{PHA}	PHA requirement for Poly-P storage	$M_{COD}M_P^{-1}$	3.88
Y_{PO_4}	Poly-P requirement (S_{PO4} release) for PHA storage	$M_PM_{COD}^{-1}$	3.84
$Y_{XOC,T}$	True growth yield for biomass degrading an XOC	$M_{TSS}M_{COD}^{-1}$	22.17
z	Dimensionless length	—	4.23

Symbol	Definition	Units	Section or Equation Where First Used
α	Biomass recycle or recirculation ratio	—	Figure 5.2
α'	Empirical parameter in pseudoanalytical approach to biofilms	—	16.35
α_a	Dimensionless abiotic loss coefficient for an XOC	—	22.19
α_N	Coefficient in COD mass balance to account for the type of nitrogen source	—	3.94
α_s	Dimensionless sorption coefficient for an XOC	—	22.20
α_v	Dimensionless volatilization coefficient for an XOC	—	22.20
β	Mixed liquor recirculation ratio	—	7.13
β'	Empirical parameter in pseudoanalytical approach to biofilms	—	16.35
γ	Liquid specific weight	$ML^{-2}T^{-2}$	11.6
γ	Fraction of XOC removal from a CMAS system attributable to abiotic removal mechanisms	—	22.18
γ_i	COD-based stoichiometric coefficient for reactant A_i	MM^{-1}	3.4
$\gamma_{i,j}$	COD-based stoichiometric coefficient for reactant A_i in reaction j	MM^{-1}	Section 3.1.3
$\Gamma_{A,S}$	Areal organic loading rate	$ML^{-2}T^{-1}$	15.1
$\Gamma_{A,XB}$	Areal loading rate for biodegradable suspended solids	$ML^{-2}T^{-1}$	15.12
$\Gamma_{V,S}$	Volumetric organic loading rate	$ML^{-3}T^{-1}$	14.1
δ	Anoxic mixed liquor recirculation ratio	—	12.19
$\Delta E_O'$	Standard oxidation reduction potential	mV	Section 2.4.1
$\Delta(F{\cdot}S_N)_{a,TS}$	Transient increase in available ammonia-N	MT^{-1}	11.13
$\Delta(F{\cdot}S_{NHO})$	Transient increase in ammonia-N loading	MT^{-1}	11.14
$\Delta(F{\cdot}S_{NSO})$	Transient increase in soluble organic-N loading	MT^{-1}	11.14
$\Delta(F{\cdot}S_{SO})$	Transient increase in the loading of readily biodegradable substrate	MT^{-1}	11.12
$\Delta(F{\cdot}X_{NSO})$	Transient increase in particulate organic-N loading	MT^{-1}	11.14
$\Delta(F{\cdot}X_{SO})$	Transient increase in the loading of slowly biodegradable substrate	MT^{-1}	11.12
$\Delta G^{0'}$	Gibbs free energy change	kJ	Section 2.4.1
ΔN	Mass rate of nitrate-N removal by denitrification	MT^{-1}	Section 6.4.2
ΔN_{SS}	Mass rate of nitrate-N removal associated with biomass synthesis from readily biodegradable substrate	MT^{-1}	12.10
ΔN_{XB}	Mass rate of nitrate-N removal associated with biomass decay	MT^{-1}	Example 12.3.1.4
ΔN_{XS}	Mass rate of nitrate-N removal associated with slowly biodegradable substrate utilization and decay	MT^{-1}	Example 12.3.1.3
$\Delta N/\Delta S$	Amount of nitrate-N required to serve as electron acceptor for a unit of substrate COD	$M_N M_{COD}^{-1}$	12.6
ΔP	Phosphorus removed in a BPR system	$M_P M_{BOD5}^{-1}$	12.4
ΔS	Mass rate of COD removal	MT^{-1}	Section 6.4.2
$\Delta S/\Delta N$	Amount of substrate COD required to remove a unit of nitrate-N by denitrification	$M_{COD} M_N^{-1}$	6.4
ΔV	Infinitesimal volume	L^3	Figure 4.2
Δx	Length of infinitesimal volume	L	4.7
$\Delta X_{B,XOC,T}$	Decrease in the concentration of biomass degrading an XOC	ML^{-3}	22.18

(*Continued*)

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Symbol	Definition	Units	Section or Equation Where First Used
ε	Porosity	—	17.13
ε_M	Porosity at incipient fluidization	—	18.1
ε_R	Porosity associated with the reference bed height	—	18.3
E_{XMV}	VSS destruction efficiency	%	13.4
ζ	Fraction of a cycle in an SBR devoted to fill plus react	—	7.8
ζ	Empirical coefficient	$L^{0.333}T^{0.667}$	17.31
ζ_1	Empirical coefficient	$LT^{0.667}$	17.33
ζ_2	Empirical coefficient	L	17.34
ζ_3	Empirical coefficient	$LT^{\zeta 4}$	17.34
ζ_4	Empirical exponent	—	17.34
η_e	Effectiveness factor for biofilms	—	16.6
η_{eE}	External effectiveness factor	—	Section 16.2.2
η_{eI}	Internal effectiveness factor	—	Section 16.2.2
η_{eO}	Overall effectiveness factor	—	Section 16.2.2
η_{eOa}	Overall effectiveness factor in the aerated sector of an RDR	—	17.47
η_{eOs}	Overall effectiveness factor in the submerged sector of an RDR	—	17.46
η_{eZ}	Zero-order effectiveness factor	—	18.20
η_{e1}	First-order effectiveness factor	—	18.22
η_g	Anoxic growth factor	—	Table 6.1
$\eta_{g,POA}$	Anoxic PAO growth factor	—	3.90
η_h	Anoxic hydrolysis factor	—	Table 6.1
η_P	In-process energy efficiency for mechanical aeration systems	T^2L^{-2}	11.1
η_Q	Field oxygen transfer efficiency for diffused aeration systems	—	11.2
θ	Temperature coefficient	—	3.99
θ	Dimensionless time	—	4.22
θ	Circumferential angle in an RDR	°	Section 17.2.1
θ_A	Angle transcribed by the aerated sector of an RDR	°	17.49
Θ_c	Solids retention time	T	5.1
$\Theta_{c,AER}$	Aerobic solids retention time	T	12.1
$\Theta_{c,AER,eq}$	Equivalent aerobic solids retention time	T	12.18
$\Theta_{c,ANA}$	Anaerobic solids retention time	T	12.3
$\Theta_{c,ANX}$	Anoxic solids retention time	T	12.2
Θ_{ce}	Effective solids retention time in an SBR	T	7.9
Θ_{cmin}	Minimum solids retention time at which biomass can grow on a given influent substrate concentration	T	5.25
$\Theta_{c,r}$	Solids retention time required to obtain a desired effluent quality	T	11.10
$\Theta_{c/XS}$	Solids retention time required to degrade slowly biodegradable substrate	T	11.28
κ	Parameter depicting the deviation of the Thiele modulus from first-order kinetics	—	16.13
λ_N	Surface total nitrogen loading to an attached growth bioreactor	$ML^{-2}T^{-1}$	Section 19.2.1
λ_{NH}	Surface ammonia-N loading to an attached growth bioreactor	$ML^{-2}T^{-1}$	Section 19.2.1

(CONTINUED)

Symbol	Definition	Units	Section or Equation Where First Used
λ_S	Surface organic loading to an attached growth bioreactor	$ML^{-2}T^{-1}$	19.2
Λ_H	Total hydraulic loading to a bioreactor; superficial velocity	LT^{-1}	14.5
$\Lambda_{H,RBC}$	Total hydraulic loading on an RBC	LT^{-1}	20.2
Λ_N	Total nitrogen loading to an attached growth bioreactor	$ML^{-3}T^{-1}$	Section 19.1.2
Λ_{NH}	Total ammonia-N loading to an attached growth bioreactor	$ML^{-3}T^{-1}$	Section 19.1.2
Λ_{OR}	Oxidation rate in a trickling filter	$ML^{-3}T^{-1}$	19.3
Λ_S	Total organic loading to an attached growth bioreactor	$ML^{-3}T^{-1}$	19.1
μ	Specific growth rate coefficient	T^{-1}	3.35
μ_A	Specific growth rate coefficient for autotrophs	T^{-1}	Section 9.5.3
$\mu_{A,C}$	Specific growth rate coefficient for autotrophs in the contact tank of a CSAS system	T^{-1}	11.49
μ_H	Specific growth rate coefficient for heterotrophs	T^{-1}	3.35a in Section 5.1.3
$\mu_{H,C}$	Specific growth rate coefficient for heterotrophs in the contact tank of a CSAS system	T^{-1}	11.45
$\mu_{H,F}$	Specific growth rate coefficient for heterotrophs in a fictitious CSTR	T^{-1}	11.29
$\mu_{H,i}$	Specific growth rate coefficient for heterotrophs in bioreactor i	T^{-1}	7.5
$\mu_{H,N}$	Specific growth rate coefficient for heterotrophs in the last equivalent tank of an SFAS system	T^{-1}	11.41
μ_{Hmax}	Maximum specific heterotrophic growth rate associated with a given influent substrate concentration	T^{-1}	5.24
μ_i	Specific growth rate of species i in a biofilm	T^{-1}	16.53
μ_{ias}	Specific growth rate of species i at the attachment surface of a biofilm	T^{-1}	16.59
μ_P	Specific growth rate coefficient for photosynthetic microorganisms	T^{-1}	15.14
μ_w	Absolute viscosity of water	$MT^{-1}\,L^{-1}$	11.6
μ_{wc}	Absolute viscosity of water in centipoise	$MT^{-1}\,L^{-1}$	13.10
$\hat{\mu}$	Maximum specific growth rate coefficient	T^{-1}	3.36
$\hat{\mu}_A$	Maximum specific growth rate coefficient for autotrophs	T^{-1}	3.50
$\hat{\mu}_{Am}$	Maximum specific growth rate coefficient for autotrophs at optimum pH	T^{-1}	3.50
$\hat{\mu}_H$	Maximum specific growth rate coefficient for heterotrophs	T^{-1}	Section 3.2.10
$\hat{\mu}_{PAO}$	Maximum specific growth rate coefficient for PAOs	T^{-1}	3.85
$\hat{\mu}_{XOC}$	Maximum specific growth rate coefficient for degradation of an XOC	T^{-1}	5.22 in Chapter 22
μ^*	Maximum observed specific growth rate	T^{-1}	3.40
ξ	Parameter relating the dimensionless substrate flux to the dimensionless flux to a deep biofilm	—	16.31
Ξ	Fraction of substrate aerobically stabilized	—	15.10
Π	Volumetric power input	T^{-1} or $ML^{-1}T^{-3}$	Section 11.2.5
Π_L	Lower limit on volumetric power input	T^{-1} or $ML^{-1}T^{-3}$	Figure 10.6

(*Continued*)

(CONTINUED)

Symbol	Definition	Units	Section or Equation Where First Used
$\Pi_{L,P}$	Lower limit on volumetric power input for mechanical aeration systems	$ML^{-1}T^{-3}$	11.3
$\Pi_{L,Q}$	Lower limit on volumetric power input for diffused aeration systems	T^{-1} or LT^{-1}	11.3
Π_{U}	Upper limit on volumetric power input	T^{-1} or $ML^{-1}T^{-3}$	Figure 10.6
$\Pi_{U,P}$	Upper limit on volumetric power input for mechanical aeration systems	$ML^{-1}T^{-3}$	11.4
$\Pi_{U,Q}$	Upper limit on volumetric power input for diffused aeration systems	T^{-1} or LT^{-1}	11.4
ρ_b	Density of a bioparticle	ML^{-3}	18.8
ρ_{fd}	Dry density of the biofilm on an FBBR bioparticle	ML^{-3}	18.9
ρ_{fw}	Wet density of the biofilm on an FBBR bioparticle	ML^{-3}	18.8
ρ_p	Density of a carrier particle	ML^{-3}	18.1
ρ_w	Density of water	ML^{-3}	Section 16.2.1
ς_{DO}	Dissolved oxygen safety factor	—	11.9
ς_{PL}	Peak load safety factor	—	11.8
ς_U	Safety factor for uncertainty	—	11.10
τ	Hydraulic residence time	T	4.15
τ_e	Effective hydraulic residence time in an SBR	T	7.8
τ_{min}	Minimum allowable hydraulic residence time	T	Example 5.1.3.1
τ_n	Hydraulic residence time in stage N of an RBC system	T	20.5
υ	Fraction of CSAS system volume in the contact tank	—	11.47
ϕ	Thiele modulus	—	16.12
ϕ_f	Modified Thiele modulus	—	16.14
ϕ_{Zm}	Modified zero-order Thiele modulus	—	18.19
ϕ_{lm}	Modified first-order Thiele modulus	—	18.21
Φ_{XOC}	Proportionality factor relating the mass transfer coefficient for an XOC to the mass transfer coefficient for oxygen	—	22.8
Ψ_i	Mass-based stoichiometric coefficient for reactant A_i	MM^{-1}	3.2
$\Psi_{i,j}$	Mass-based stoichiometric coefficient for reactant A_i in reaction j	MM^{-1}	3.11
ω	Rotational speed of an RDR	$RevT^{-1}$	17.29
ω_d	Rotational speed of a rotary distributor on a trickling filter	$RevT^{-1}$	19.4
Ω	S_{Smin} in a CSTR receiving active biomass in the influent expressed as a fraction of S_{Smin} in absence of such biomass	—	5.68

Appendix C: Unit Conversions

U.S. Units to Metric Units

Multiply U.S. Units	By	To Obtain Metric Units
ac	4.047×10^{3}	m^2
ac	0.4047	ha
BTU	0.2520	kcal
BTU	1.055	kJ
BTU	2.931×10^{-4}	kW·hr
degrees F	0.5556 (°F − 32)	degrees C
ft	0.3048	m
ft/hp	0.4087	m/kW
ft^2	9.290×10^{-2}	m^2
ft^2/ft^3	3.281	m^2/m^3
ft^3	2.832×10^{-2}	m^3
$ft^3/(min \cdot ft^2)$	0.3048	$m^3/(min \cdot m^2)$
gal	3.785×10^{-3}	m^3
gal/(day · ac)	9.357×10^{-7}	$m^3/(day \cdot m^2)$
gal/(day · ft^2)	4.074×10^{-2}	$m^3/(day \cdot m^2)$
gal/(min · ft^2)	58.674	$m^3/(day \cdot m^2)$
gal/(min · ft^2)	6.791×10^{-4}	$m^3/(sec \cdot m^2)$
gal/min	5.451	m^3/day
gal/min	6.308×10^{-5}	m^3/sec
gal/(min · hp)	8.460×10^{-5}	$m^3/(sec \cdot kW)$
hp	0.7457	kW
hp/(1000 ft^3)	26.334	kW/(1000 m^3)
hp/(10^6 gal)	0.1973	kW/(1000 m^3)
lb (mass)	0.4536	kg
lb/(ac · day)	1.121	kg/(ha · day)
lb/(ac · day)	1.121×10^{-4}	kg/(m^2 · day)
lb/(1000 ft^2 · day)	4.882	g/(m^2 · day)
lb/(1000 ft^2 · day)	4.882×10^{-3}	kg/(m^2 · day)
lb/(1000 ft^3 · day)	1.602×10^{-2}	kg/(m^3 · day)
lb/(hp · hr)	0.6083	kg/(kW · hr)
Mgd	3.785×10^{3}	m^3/day
Mgd	4.381×10^{-2}	m^3/sec

Metric Units to U.S. Units

Multiply Metric Units	By	To Obtain U.S. Units
degrees C	1.800 (°C + 32)	degrees F
g/(m^2 · day)	0.2048	lb/(1000 ft^2 · day)
ha	2.471	ac
kcal	3.968	BTU
kg	2.205	lb (mass)
kg/(ha · day)	0.8922	lb/(ac · day)
kg/(kW · hr)	1.644	lb/(hp · hr)
kg/(m^2 · day)	8922	lb/(ac · day)
kg/(m^2 · day)	204.8	lb/(1000 ft^2 · day)
kg/(m^3 · day)	62.428	lb/(1000 ft^3 · day)
kJ	0.9478	BTU
kW	1.341	hp
kW · hr	3412	BTU
kW/(1000 m^3)	3.797×10^{-2}	hp/(1000 ft^3)
kW/(1000 m^3)	5.068	hp/(10^6 gal)
m	3.281	ft
m/kW	2.447	ft/hp
m^2	2.471×10^{-4}	ac
m^2	10.76	ft^2
m^2/m^3	0.3048	ft^2/ft^3
m^3	35.31	ft^3
m^3	264.2	gal
m^3/day	0.1835	gal/min
m^3/day	2.642×10^{-4}	Mgd
m^3/(day · m^2)	1.069×10^6	gal/(day · ac)
m^3/(day · m^2)	24.55	gal/(day · ft^2)
m^3/(day · m^2)	1.704×10^{-2}	gal/(min · ft^2)
m^3/(min · m^2)	3.281	ft^3/(min · ft^2)
m^3/(sec · m^2)	1472	gal/(min · ft^2)
m^3/sec	1.585×10^4	gal/min
m^3/sec	22.83	Mgd
m^3/(sec · kW)	1.182×10^4	gal/(min · hp)

ENVIRONMENTAL ENGINEERING

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