

# Biological Wastewater Treatment Processes

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## 11.1 Introduction

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National Pollutant Discharge Elimination System (NPDES) permits limit the final effluent's total suspended solids (SS) and 5-day biochemical oxygen demand (BOD<sub>5</sub>). Other quality parameters and sampling requirements are included as needed. However, most biological processes produce effluents that contain only a few mg/L of soluble BOD<sub>5</sub>, or less, and the BOD<sub>5</sub> of the final effluent is mostly biomass that has escaped capture by the final clarifier. Thus, the NPDES permits control final clarifier design and operation, not the biological process itself. Some biological processes, like completely mixed activated sludge, produce flocs that settle slowly, and this should be a consideration in their selection and operation. If effluents containing much less than 10 mg/L of SS or BOD<sub>5</sub> are required, then additional effluent treatment like coagulation, settling, and filtration should be considered.

Because the permitted effluent BOD<sub>5</sub> does not control the biological process, the design procedure focuses on other considerations like waste sludge production, oxygen utilization, nitrification, and biological nutrient removal. The most common choice is whether or not to nitrify, and this choice determines the solids' retention time (SRT) in activated sludge plants and the hydraulic loading in trickling filter plants.

Most aerobic biological processes are capable of similar carbon removal efficiencies, and the criteria for choosing among them are largely economic and operational. Activated sludge plants tend to be capital-, labor-, and power-intensive but compact. They are usually adopted in urban areas. Ponds and irrigation schemes require little capital, labor, or power but need large land areas per caput. They are usually adopted in rural areas. Trickling filters and other fixed film processes fall between activated sludge and ponds and irrigation in their requirements.

Biological nutrient removal (BNR) is most developed and best understood in the activated sludge process. Therefore, most BNR facilities are modifications of the activated sludge.

The jargon of the profession now distinguishes between *aerobic*, *anoxic*, and *anaerobic* conditions. Aerobic means that dissolved oxygen is present (and nonlimiting). Both anoxic and anaerobic mean that dissolved oxygen is absent. However, anaerobic also means that there is no other electron acceptor present, especially nitrite or nitrate, whereas anoxic means that other electron acceptors are present, usually nitrate and sometimes sulfate. Most engineers continue to classify methanogenesis from hydrogen and carbon dioxide as an anaerobic process, but in the new jargon, it is better classified as an anoxic process, because carbon dioxide is the electron acceptor, and because energy is captured from proton fluxes across the cell membrane.

The following descriptions use the recommended notations of the International Water Association (Grau et al., 1982, 1987) and the International Union of Pure and Applied Chemistry (Mills et al., 1993). The *Système International d'Unités* (Bureau International, 1991) is strictly followed, except where cited authors use another. In those cases, the cited author's units are quoted.

## References

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## 11.2 Activated Sludge

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The principal wastewater treatment scheme is the activated sludge process, which was developed by Ardern and Lockett in 1914. Its various modifications are capable of removing and oxidizing organic matter, of oxidizing ammonia to nitrate, of reducing nitrate to nitrogen gas, and of achieving high removals of phosphorus via incorporation into biomass as volutin crystals.

### Biokinetics of Carbonaceous BOD Removal

Designs can be based on calibrated and verified process models, pilot-plant data, or the traditional rules of thumb. The traditional rules of thumb are acceptable only for municipal wastewaters that consist primarily of domestic wastes. The design of industrial treatment facilities requires pilot testing and careful wastewater characterization. Process models require calibration and verification on the intended wastewater, although municipal wastewaters are similar enough that calibration data for one facility is often useful at others.

Most of the current process models are based on Pearson's (1968) simplification of Gram's (1956) model with additional processes and variables proposed by McKinney (1962). Pirt's (1965) maintenance concept is also used below.

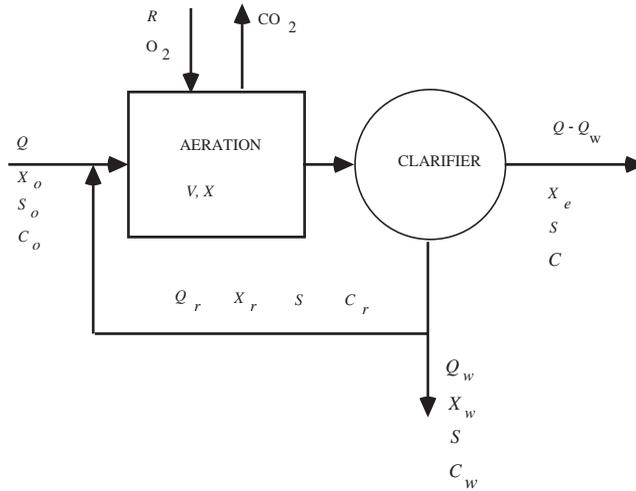


FIGURE 11.1 Generic activated sludge process.

### State Variables and Kinetic Relations

A wide variety of state variables has been and is being used to describe the activated sludge process. The minimal required set is probably that proposed by McKinney (1962).

Refer to Fig. 11.1. A steady state mass balance on the secondary clarifier for any component of the suspended solids leads to Eq. (11.1):

$$Q_w X_w + (Q - Q_w) X_e = (Q + Q_r) X - Q_r X_r \quad (11.1)$$

- where
- $Q$  = the raw or settled wastewater flow rate ( $\text{m}^3/\text{s}$ )
  - $Q_r$  = the recycle (return) activated sludge flow rate ( $\text{m}^3/\text{s}$ )
  - $Q_w$  = the waste-activated sludge flow rate ( $\text{m}^3/\text{s}$ )
  - $X$  = the suspended solids' concentration in the aeration tank, analytical method and model variable unspecified ( $\text{kg}/\text{m}^3$ )
  - $X_e$  = the suspended solids' concentration in the final effluent, analytical method and model variable unspecified ( $\text{kg}/\text{m}^3$ )
  - $X_r$  = the suspended solids' concentration in the return activated sludge, analytical method and model variable unspecified ( $\text{kg}/\text{m}^3$ )
  - $X_w$  = the suspended solids' concentration in the waste-activated sludge, analytical method and model variable unspecified ( $\text{kg}/\text{m}^3$ )

Eq. (11.1) is used below to eliminate the inflow and outflow terms in the aeration tank mass balances.

The left-hand side of Eq. (11.1) represents the net suspended solids production rate of the activated sludge process. It also appears in the definition of the *solids' retention (detention, residence) time* (SRT):

$$\Theta_X = \frac{VX}{Q_w X_w + (Q - Q_w) X_e} \quad (11.2)$$

- where
- $V$  = the aeration tank volume ( $\text{m}^3$ )
  - $\Theta_X$  = the solids' retention time, SRT (s)

Synonyms for SRT are biological solids' residence time, cell residence time (CRT), mean cell age, mean (reactor) cell residence time (MCRT), organism residence time, sludge age, sludge turnover time, and

solids' age. Synonyms for the reciprocal SRT ( $1/\Theta_x$ ) are cell dilution rate, fraction sludge lost per day, fraction rate of removal of sludge solids, net growth rate, and VSS wasting rate.

The SRT is the average time solids spend in the activated sludge system, and it is analogous to the hydraulic retention time (HRT). The solids in the secondary clarifier are often ignored on the grounds that they are a negligible portion of the total system biomass and are inactive, lacking external substrate. However, predation, endogenous respiration, cell lysis, and released nutrient uptake and denitrification may occur in the clarifier. So, if it holds a great deal of solids (as in the sequencing batch reactor process), they should be included in the numerator of Eq. (11.2). The amount of solids in the clarifier is under operator control, and the operator can force Eq. (11.2) to be true as written.

If the suspended solids have the same composition everywhere in the system, then each component has the same SRT.

Sludge age has been used to mean several different things: the ratio of aeration tank biomass to influent suspended solids loading (Torpey, 1948), the reciprocal food-to-microorganism ratio (Heukelekian, Orford, and Manganeli, 1951), the reciprocal specific uptake rate (Fair and Thomas, 1950), the aeration period (Keefer and Meisel, 1950), and some undefined relationship between system biomass and the influent BOD<sub>5</sub> and suspended solids loading (Eckenfelder, 1956).

The dynamic mass balances for McKinney's variables in the aeration tank of a completely mixed activated sludge system and their steady state solutions are given below. The resulting formulas were simplified using Eq. (11.1) above:

**Active Biomass,  $X_{va}$**

accumulation in aeration = inflow – outflow + reproduction – "decay"

$$\frac{d(VX_{va})}{dt} = Q_r X_{var} - (Q + Q_r) X_{va} + \mu V X_{va} - k_d V X_{va} \quad (11.3)$$

$$\frac{1}{\Theta_x} = \mu - k_d$$

where  $k_d$  = the "decay" rate (per s)

$t$  = clock time (s)

$X_{va}$  = the active biomass concentration in the aeration tank (kg VSS/m<sup>3</sup>)

$X_{var}$  = the active biomass in the recycle activated sludge flow (kg VSS/m<sup>3</sup>)

$\mu$  = the specific growth rate of the active biomass (per s)

Eq. (11.3) is true for all organisms in every biological process. However, in some processes, e.g., trickling filters, the system biomass cannot be determined without destroying the facility, and Eq. (11.3) is replaced by other measurable parameters.

The "active biomass" is a model variable defined by the model equations. It is not the actual biomass of the microbes and metazoans in the sludge.

The "decay" rate replaces McKinney's original endogenous respiration concept; and it is more general. The endogenous respiration rate of the sludge organisms is determined by measuring the oxygen consumption rate of sludge solids suspended in a solution of mineral salts without organic substrate. The measured rate includes the respiration of algae, bacteria, and fungi oxidizing intracellular food reserves (true endogenous respiration) and the respiration of predators feeding on their prey (technically exogenous respiration). Wuhrmann (1968) has shown that the endogenous respiration rate declines as the solids' retention time increases.

The decay rate is determined by regression of the biokinetic model on experimental data from pilot or field facilities. It represents a variety of solids' loss processes including at least: (a) viral lysis of microbial and metazoan cells; (b) hydrolysis of solids by exocellular bacterial and fungal enzymes; (c) hydrolysis of solids by intracellular ("intestinal") protozoan, rotiferan, and nematodal enzymes; (d) simple dissolution;

(e) abiotic hydrolysis; and (f) the respiration of all the organisms present. The decay rate is a constant regardless of solids' retention time.

**Dead Biomass,  $X_{vd}$**

accumulation in aeration = inflow – outflow + "decay"

$$\frac{d(VX_{vd})}{dt} = Q_r X_{vdr} - (Q + Q_r) X_{vd} + f_d k_d V X_{va} \quad (11.4)$$

$$X_{vd} = f_d k_d \Theta_X X_{va}$$

where  $f_d$  = the fraction of active biomass converted to dead (inert) suspended solids by the various decay processes (dimensionless)

$X_{vd}$  = the dead (inert) biomass in the aeration tank (kg VSS/m<sup>3</sup>)

In McKinney's original model, the missing part  $(1 - f_d)$  of the decayed active biomass is the substrate oxidized during endogenous respiration. In some more recent models, the missing active biomass is assumed to be oxidized and converted to biodegradable and unbiodegradable soluble matter.

**Particulate Substrate,  $X_s$**

accumulation in aeration = inflow – outflow – hydrolysis

$$\frac{d(VX_{vs})}{dt} = QX_{so} + Q_r X_{sr} - (Q + Q_r) X_s - k_h V X_s \quad (11.5)$$

$$\frac{X_s}{X_{so}} = \frac{\Theta_X}{(1 + k_h \Theta_X) \tau}$$

where  $k_h$  = the first-order particulate substrate hydrolysis rate (per s)

$X_s$  = the particulate substrate concentration in the aeration tank (kg/m<sup>3</sup>)

$X_{so}$  = the particulate substrate concentration in the raw or settled wastewater (kg/m<sup>3</sup>)

$\tau$  = the hydraulic retention time (s)

=  $V/Q$

The particulate substrate comprises the bulk of the biodegradable organic matter in municipal wastewater, even after primary settling. The analytical method used to measure  $X_s$  will depend on whether it is regarded as part of the sludge solids (in which case, the units are VSS) or as part of the substrate (in which case, the units are BOD<sub>5</sub>, ultimate carbonaceous BOD, or biodegradable COD). In Eq. (11.5), the focus is on substrate, and the analytical method is unspecified.

**Inert Influent Particulate Organic Matter,  $X_{vi}$**

accumulation in aeration = inflow – outflow

$$\frac{d(VX_{vi})}{dt} = QX_{vio} + Q_r X_{vir} - (Q + Q_r) X_{vi} \quad (11.6)$$

$$\frac{X_{vi}}{X_{vio}} = \frac{\Theta_X}{\tau}$$

where  $X_{vi}$  = the concentration of inert organic suspended solids in the aeration tank that originated in the raw or settled wastewater (kg VSS/m<sup>3</sup>)

$X_{vio}$  = the concentration of inert organic suspended solids in the raw or settled wastewater (kg VSS/m<sup>3</sup>)

### **Inert Particulate Mineral Matter, $X_m$**

accumulation in aeration = inflow – outflow

$$\frac{d(VX_m)}{dt} = QX_{mo} + Q_r X_{mr} - (Q + Q_r)X_m \quad (11.7)$$

$$\frac{X_m}{X_{mo}} = \frac{\Theta_X}{\tau}$$

where  $X_m$  = the concentration of inert suspended mineral matter in the aeration tank (kg/m<sup>3</sup>)

$X_{mo}$  = the concentration of inert suspended mineral matter in the raw or settled wastewater (kg/m<sup>3</sup>)

The influent suspended mineral matter is mostly colloidal clay. However, in many wastewaters, additional suspended inorganic solids are produced by abiotic oxidative processes (e.g., ferric hydroxide) and by carbon dioxide stripping (e.g., calcium carbonate).

The mixed liquor suspended solids (MLSS) concentration includes all the particulate organic and mineral matter,

$$X = X_{va} + X_{vd} + X_{vi} + X_{vs} + X_m \quad (11.8)$$

whereas, the mixed liquor volatile suspended solids (MLVSS) includes only the particulate organic matter,

$$X_v = X_{va} + X_{vd} + X_{vi} + X_{vs} \quad (11.9)$$

where  $X$  = the total suspended solids concentration in the aeration tank (kg/m<sup>3</sup>)

$X_v$  = the volatile suspended solids concentration in the aeration tank (kg VSS/m<sup>3</sup>)

In Eq. (11.9), particulate substrate ( $X_{vs}$ ) is measured as VSS for consistency with the other particulate organic fractions. The quantity of total suspended solids determines the capacities of solids handling and dewatering facilities, and the quantity of volatile suspended solids determines the capacity of sludge stabilization facilities.

### **Soluble Substrate, $S_s$**

accumulation in aeration = inflow – outflow – uptake for reproduction

– uptake for maintenance + hydrolysis

$$\frac{d(VS_s)}{dt} = QS_{so} + Q_r S_s - (Q + Q_r)S_s - \frac{\mu VX_{va}}{Y_a} \quad (11.10)$$

$$-k_m VX_{va} + k_h VX_s$$

$$X_{va} = \frac{Y_a}{1 + (k_d + k_m Y) \Theta_X} \cdot \frac{\Theta_X}{\tau} \cdot \left[ S_{so} - S_s + \left( \frac{k_h \Theta_X}{1 + k_h \Theta_X} \right) X_{so} \right]$$

where  $k_m$  = the specific maintenance energy demand rate of the active biomass (kg substrate per kg active biomass per s)

$S_s$  = the concentration of soluble readily biodegradable substrate in the aeration tank, analytical method unspecified (kg/m<sup>3</sup>)

$S_{so}$  = the concentration of soluble readily biodegradable substrate in the raw or settled wastewater, analytical method unspecified (kg/m<sup>3</sup>)

$Y_a$  = the true growth yield of the active biomass from the soluble substrate (kg active biomass VSS per kg soluble substrate)

The soluble substrate is the only form of organic matter that can be taken up by bacteria, fungi, and algae. Its concentration and the concentration of particulate substrate should be measured as the ultimate carbonaceous biochemical oxygen demand (CBOD<sub>u</sub>). If it is measured as COD, then the biodegradable fraction of the COD must be determined.

Weichers et al. (1984) describe a kinetic technique for determining the soluble readily biodegradable COD ( $S_{\text{CODrb}}$ ) based on oxygen uptake rate (OUR). First, the OUR of the mixed liquor is measured under steady loading and operating conditions. This should be done at several different times to establish that the load is steady. Then the influent load is shut off, and the OUR is measured at several different times again. There should be an immediate drop in OUR within a few minutes followed by a slow decline. The immediate drop represents the readily biodegradable COD. The calculation is as follows:

$$S_{\text{CODrb}} = \frac{(R_{\text{O}_2sl} - R_{\text{O}_2nl})V}{f_{\text{ox}}Q} \quad (11.11)$$

where  $f_{\text{ox}}$  = the fraction of the consumed COD that is oxidized by rapidly growing bacteria (dimensionless)

$$= 1 - \beta_x Y_h$$

$$\approx 0.334 \text{ (Weichers et al., 1984)}$$

$R_{\text{O}_2nl}$  = the oxygen uptake rate immediately after the load is removed (kg/s)

$R_{\text{O}_2sl}$  = the oxygen uptake rate during the steady load (kg/s)

$Q$  = the wastewater flow rate during loading (m<sup>3</sup>/s)

$S_{\text{CODrb}}$  = the soluble readily biodegradable COD (kg/m<sup>3</sup>)

$V$  = the reactor volume (m<sup>3</sup>)

The maintenance energy demand is comprised of energy consumption for protein turnover, motility, maintenance of concentration gradients across the cell membrane, and production of chemical signals and products. In the absence of external substrates, the maintenance energy demand of single cells is met by endogenous respiration.

Eq. (11.10) can also be written:

$$1 = \frac{Y_a}{1 + (k_d + k_m Y)\Theta_x} \cdot \frac{Q \left[ S_{so} - S_s + \left( \frac{k_h \Theta_x}{1 + k_h \Theta_x} \right) X_{so} \right]}{VX_{va}} \cdot \Theta_x \quad (11.12)$$

The first factor on the right-hand side of Eq. (11.12) can be thought of as an observed active biomass yield,  $Y_{ao}$ :

$$Y_{ao} = \frac{Y_a}{1 + (k_d + k_m Y)\Theta_x} \quad (11.13)$$

which is the true growth yield corrected for the effects of decay and maintenance. We can also define a specific substrate uptake rate by active biomass,

$$q_a = \frac{Q \left[ S_{so} - S_s + \left( \frac{k_h \Theta_x}{1 + k_h \Theta_x} \right) X_{so} \right]}{VX_{va}} \quad (11.14)$$

where  $q_a$  = the specific uptake rate of substrate by the active biomass (kg substrate per kg VSS per s).

With these definitions, Eq. (11.12) becomes,

$$Y_{ao} q_a \Theta_x = 1 \text{ (dimensionless)} \quad (11.15)$$

Combining Eqs. (11.10) and (11.14) also produces,

$$q_a = \frac{\mu}{Y_a} + k_m \quad (11.16)$$

and

$$\frac{1}{\Theta_x} = Yq_a - (k_d + k_m Y_a) \quad (11.17)$$

which should be compared with Eq. (11.3).

### ***Inert Soluble Organic Matter, $S_i$***

accumulation in aeration = inflow – outflow

$$\frac{d(VS_i)}{dt} = QS_{io} + Q_r S_i - (Q + Q_r)S_i \quad (11.18)$$

$$S_i = S_{io}$$

where  $S_i$  = the inert soluble organic matter concentration in the aeration tank, analytical method unspecified (kg/m<sup>3</sup>)

$S_{io}$  = the inert soluble organic matter concentration in the raw or settled wastewater, analytical method unspecified (kg/m<sup>3</sup>).

The results of model calibrations suggest that roughly a fifth to a fourth of the particulate and soluble organic matter in municipal wastewater is unbiodegradable. This is an overestimate, because the organisms of the sludge produce a certain amount of unbiodegradable organic matter during the decay process.

### **Input-Output Variables**

Pilot plant and field data and rules of thumb are often summarized in terms of traditional input-output variables. There are two sets of such variables, and each set includes the SRT as defined by Eq. (11.2) above. Refer again to Fig. 11.1.

The first set is based on the organic matter removed from the wastewater and consists of an observed volatile suspended solids yield and a specific uptake rate of particulate and soluble substrate:

### ***Observed Volatile Suspended Solids Yield, $Y_{vo}$***

$$Y_{vo} = \frac{Q_w X_{vw} + (Q - Q_{vw})X_{ve}}{Q(C_{so} - S_{se})} \quad (11.19)$$

where  $Y_{vo}$  = the observed volatile suspended solids yield based on the net reduction in organic matter (kg VSS/kg substrate)

$C_{so}$  = the total (suspended plus soluble) biodegradable organic matter (substrate) concentration in the settled sewage, analytical method not specified (kg substrate/m<sup>3</sup>)  
 $= X_{s0} + S_{s0}$

$S_{se}$  = the soluble biodegradable organic matter concentration in the final effluent (kg COD/m<sup>3</sup> or lb COD/ft<sup>3</sup>).

Because the ultimate problem is sludge handling, stabilization, and disposal, the observed yield includes all the particulate organic matter in the sludge, active biomass, endogenous biomass, inert organic solids, and residual particulate substrate. Sometimes particulate mineral matter is included, too.

### Specific Uptake (Utilization) Rate, $q_v$

$$q_v = \frac{Q(C_{so} - S_{se})}{VX_v} \quad (11.20)$$

where  $q_v$  = the specific uptake (utilization) rate of biodegradable organic matter by the volatile suspended solids (kg BOD<sub>5</sub>/kg VSS·s).

The specific uptake rate is sometimes reported in units of reciprocal time (e.g., “per day”), which is incorrect unless all organic matter is reported in the same units (e.g., COD). The correct traditional units are kg BOD<sub>5</sub> per kg VSS per day.

As a consequence of these definitions,

$$Y_{vo} q_v \Theta_X \equiv 1 \text{ (dimensionless)} \quad (11.21)$$

This is purely a semantic relationship. There is no assumption regarding steady states, time averages, or mass conservation involved. If any two of the variables are known, the third can be calculated. Simple rearrangement leads to useful design formulas, e.g.,

$$\frac{VX_v}{\Theta_X} = Y_{vo} Q(C_{so} - S_{se}) \quad (11.22)$$

$$X_v = \frac{Y_{vo} \Theta_X (C_{so} - S_{se})}{\tau} \quad (11.23)$$

Note also that these variables can be related to the biokinetic model variables as follows:

$$Y_{vo} q_v = Y_{ao} q_a \quad (11.24)$$

The second set of variables defines the suspended solids yield in terms of the organic matter supplied. The specific uptake rate is replaced by a “food-to-microorganism” ratio. Refer to [Fig. 11.1](#).

### Observed Volatile Suspended Solids Yield, $Y'_{vo}$

$$Y'_{vo} = \frac{Q_w X_{vw} + (Q - Q_w) X_{ve}}{QC_{so}} \quad (11.25)$$

where  $Y'_{vo}$  = the observed yield based on the total (both suspended and soluble) biodegradable organic matter in the settled sewage (kg VSS/kg BOD<sub>5</sub>).

### Food-to-Microorganism ratio (F/M or F:M)

$$F_v = \frac{QC_{so}}{VX_v} \quad (11.26)$$

where  $F_v$  = the food-to-microorganism ratio (kg COD/kg VSS·s).

Synonyms for F/M are *loading*, *BOD loading*, *BOD loading factor*, *biological loading*, *organic loading*, *plant load*, and *sludge loading*. McKinney’s (1962) original definition of F/M as the ratio of the BOD<sub>5</sub> and VSS concentrations (not mass flows) is still encountered. Synonyms for McKinney’s original meaning are *loading factor* and *floc loading*.

The SRT is defined previously in Eq. (11.2), so,

$$Y'_{vo}F_v\Theta_X \equiv 1 \text{ (dimensionless)} \quad (11.27)$$

Again, rearrangement leads to useful formulas:

$$\frac{VX_v}{\Theta_X} = Y'_{vo}QC_{so} \quad (11.28)$$

$$X_v = \frac{Y'_{vo}\Theta_X QC_{so}}{\tau} \quad (11.29)$$

These two sets of variables are related through the removal efficiency:

$$E = \frac{C_{so} - S_{se}}{C_{so}} = \frac{q_v}{F_v} = \frac{Y'_{vo}}{Y_{vo}} \quad (11.30)$$

where  $E$  = the removal efficiency (dimensionless).

Pilot-plant data provide other useful empirical formulas:

$$\frac{1}{\Theta_X} = aq_v - b \quad (11.31)$$

$$\frac{1}{\Theta_X} = a'F_v - b' \quad (11.32)$$

$$q_v = a''F_v + b'' \quad (11.33)$$

where  $a, a', a'', b, b', b''$  = empirical constants (units vary).

The input-output variables are conceptually different from the model variables, even though there are some analogies.  $F_v$ ,  $q_v$ ,  $Y_{vo}$ , and  $Y'_{vo}$  are defined in terms of the total soluble and particulate substrate supplied, and all the volatile suspended solids, even though these included inert and dead organic matter and particulate substrate. The model variables  $q_a$  and  $Y_{ao}$  include only soluble substrate, hydrolyzed (therefore, soluble) particulate substrate, and active biomass.

However, the input-output variables are generally easier and more economical to implement, because they are measured using routine procedures, whereas determination of the model parameters and variables requires specialized laboratory studies. The input-output variables are also those used to develop the traditional rules of thumb, so an extensive public data is available that may be used for comparative and design purposes.

### Substrate Uptake and Growth Kinetics

If a pure culture of microbes is grown in a medium consisting of a single soluble kinetically limiting organic substrate and minimal salts, the specific uptake rate can be correlated with the substrate concentration using the Monod (1949) equation:

$$q = \frac{q_{\max}S_s}{K_s + S_s} \quad (11.34)$$

where  $q$  = the specific uptake rate of the soluble substrate by the microbial species (kg substrate per kg biomass per s)

$q_{\max}$  = the maximum specific uptake rate (kg substrate per kg microbial species per s)

$K_s$  = the Monod affinity constant (kg substrate/m<sup>3</sup>)

**TABLE 11.1** Typical Parameter Values for the Conventional, Nonnitrifying Activated Sludge Process for Municipal Wastewater at Approximately 15 to 25°C

Parameter	Symbol	Units	Typical	Range (%)
True growth yield	$Y_a$	kg VSS/kg COD	0.4	±10
		kg VSS/Kg CBOD <sub>5</sub>	0.7	±10
Decay rate	$k_d$	Per day	0.05	±100
Maintenance energy demand	$k_m$	kg COD/kg VSS d	0.2	±50
		kg CBOD <sub>5</sub> /kg VSS d	0.07	±50
Maximum specific uptake rate	$q_{max}$	kg COD/kg VSS d	10	±50
		kg CBOD <sub>5</sub> /kg VSS d	6	±50
Maximum specific growth rate	$\mu_{max}$	Per day	4	±50
Affinity constant	$K_s$	mg COD/L	500 (Total COD)	±50
		mg COD/L	50 (Biodegradable COD)	±50
		mg CBOD <sub>5</sub> /L	100	±50
First-order rate constant	$k$	L/mg VSS d	0.02 (Total COD)	±100
			0.2 (Biodegradable COD)	±100
			0.06 (BOD)	±100

Source: Joint Task Force of the Water Environment Federation and the American Society of Civil Engineers. 1992. *Design of Municipal Wastewater Treatment Plants: Volume I. Chapters 1–12*, WEF Manual of Practice No. 8, ASCE Manual and Report on Engineering Practice No. 76. Water Environment Federation, Alexandria, VA; American Society of Civil Engineers, New York. 1992.

Goodman, B.L. and Englande, A.J., Jr. 1974. "A Unified Model of the Activated Sludge Process," *Journal of the Water Pollution Control Federation*, 46(2): 312.

Lawrence, A.W. and McCarty, P.L. 1970. "Unified Basis for Biological Treatment Design and Operation," *Journal of the Sanitary Engineering Division, Proceedings of the American Society of Civil Engineers*, 96(3): 757.

Jordan, W.J., Pohland, F.G., and Kornegay, B.H. (no date). "Evaluating Treatability of Selected Industrial Wastes," p. 514 in *Proceedings of the 26<sup>th</sup> Industrial Waste Conference, May 4, 5, and 6, 1971*, Engineering Extension Series No. 140, J.M. Bell, ed. Purdue University, Lafayette, IN.

Peil, K.M. and Gaudy, A.J., Jr. 1971. "Kinetic Constants for Aerobic Growth of Microbial Populations Selected with Various Single Compounds and with Municipal Wastes as Substrates," *Applied Microbiology*, 21:253.

Wuhrmann, K. 1954. "High-Rate Activated Sludge Treatment and Its Relation to Stream Sanitation: I. Pilot Plant Studies," *Sewage and Industrial Wastes*, 26(1): 1.

Button (1985) has collected much of the published experimental data for  $K_s$ ,  $q_{max}$ , and  $Y$ .

A number of difficulties arise when the Monod formula is applied to activated-sludge data. First, when pure cultures are grown on minimal media,  $K_s$  takes on a value of a few mg/L for organic substrates and a few tenths of a mg/L, or less, for inorganic nutrients. However, in the mixed culture, mixed substrate environment of the activated sludge, when lumped variables like VSS, CBOD<sub>5</sub>, and COD are used,  $K_s$  typically takes on values of tens to hundreds of mg/L of CBOD<sub>5</sub> or COD. (See Table 11.1.) This is partly due to the number of different substrates present, because when single organic substances are measured as themselves,  $K_s$  values more typical of pure cultures are found (Sykes, 1999).

The apparent variation in  $K_s$  is also due to the neglect of product formation and the inclusion of unbiodegradable or slowly biodegradable microbial metabolic end-products in the COD test. If end-products and kinetically limiting substrates are measured together as a lumped variable, the apparent  $K_s$  value will be proportional to the influent substrate concentration (Contois, 1959; Adams and Eckenfelder, 1975; Grady and Williams, 1975):

$$K_s \propto C_{s0} \quad (11.35)$$

This effect is especially important in pilot studies of highly variable waste streams. It is necessary to distinguish the substrate COD from the total soluble COD. The unbiodegradable (nonsubstrate) COD is usually defined to be the intercept of the  $q_v$  vs.  $S_{se}$  plot on the COD axis. The intercept is generally on the order of 10 to 30 mg/L and comprises most of the soluble effluent COD in efficient plants.

Mixing and flocculation also affect the apparent  $K_s$  value. This was first demonstrated theoretically by Powell (1967) and then empirically by Baillod and Boyle (1970).

Wastewater treatment plants are operated to produce low soluble substrate concentrations, and the correlation between soluble substrate and  $q_a$  can be approximated as a straight line (called “first-order” kinetics):

$$q_a = kS_s \quad (11.36)$$

where  $k$  = the first-order rate constant ( $\text{m}^3/\text{kg VSS}\cdot\text{s}$ ).

In pure cultures grown on minimal media, Eq. (11.36) is observed whenever  $S_s$  is much smaller than  $K_s$ , and  $k$  is approximately  $q_{\text{max}}/K_s$ . However, in activated sludge plants treating complex wastes, individual pure substances are removed at zero-order kinetics ( $S_s \gg K_s$ ;  $q \approx q_{\text{max}}$ ) down to very low concentrations. Individual substances are removed at different constant rates, and in batch cultures tend to disappear sequentially (Wuhrmann, 1956; Tischler and Eckenfelder, 1969; Gaudy, Komolrit, and Bhatia, 1963). If a lumped variable like COD is used to measure soluble organic matter, the overall pattern fits Eq. (11.36).

### Biokinetic Caveat

From the discussion on the *Effect of Tank Configuration on Removal Efficiency* in Chapter 9, Section 9.2, it might be assumed that organic matter removal in plug flow aeration tanks and sequencing batch reactors would be greater than that in mixed-cells-in-series, which, in turn, would be more efficient than completely mixed reactors. Unfortunately, this is not true. In the case of mixed microbial populations consuming synthetic or natural wastewater, the effluent *soluble* organic matter concentration is *not* affected by reactor configuration (Badger, Robinson, and Kiff, 1975; Chudoba, Strakova, and Kondo, 1991; Haseltine, 1961; Kroiss and Ruider, 1977; Toerber, Paulson, and Smith, 1974). This is true regardless of how the organic matter is measured: biochemical oxygen demand, chemical oxygen demand, or total organic carbon. The appropriate design procedure is to assume that all reactors are completely mixed, regardless of any internal baffling.

Biological processes do not violate the laws of reaction kinetics. Instead, it is the simplistic application of these laws that is at fault. The soluble organic matter concentration in the effluents of biological reactors is not residual substrate (i.e., a reactant). Rather, it is a microbial product (Baskir and Spearing, 1980; Erickson, 1980; Grady, Harlow and Riesing, 1972; Hao and Lau, 1988; Rickert and Hunter, 1971).

It should not be assumed that reactor configuration has no effect. Sequencing batch reactors and mixed-cells-in-series reactors with short compartmental detention times (less than 10 min) suppress activated sludge filamentous bulking and are the preferred configuration for that reason.

Furthermore, in the case of *particulate* or *emulsified* substrates, ideal plug flow configurations, like SBRs, produce lower effluent substrate concentrations than do CSTRs (Cassidy, Efundiev, and White, 2000).

### Return Sludge Flow Rate

The return sludge flow rate can be calculated using the Benefield–Randall (1977) formula:

$$\frac{Q_r}{Q} = \frac{1 - \frac{\tau}{\Theta_x}}{\frac{X_r}{X} - 1} \approx \frac{1}{\frac{X_r}{X} - 1} \quad (11.37)$$

where  $Q$  = the settled sewage flow rate ( $\text{m}^3/\text{s}$ )

$Q_r$  = the return activated sludge (RAS) flow rate ( $\text{m}^3/\text{s}$ )

$V$  = the aeration tank volume ( $\text{m}^3$ )

$X$  = the concentration of volatile suspended solids in the aeration tank mixed liquor, MLVSS ( $\text{kg VSS}/\text{m}^3$ )

$X_r$  = the concentration of volatile suspended solids in the return (or recycle) activated sludge ( $\text{kg VSS}/\text{m}^3$ )

$\Theta_X$  = the solids' retention time (s)  
 $\tau$  = the hydraulic retention time,  $V/Q$  (s)

This is another rearrangement of the clarifier solids balance, Eq. (11.1).

### Steady State Oxygen Consumption

The steady state total oxygen demand balance on the whole system applicable to any activated sludge process is as follows:

$$R_{O_2} = Q(C_{so} - S_{se}) + 4.57Q(C_{TKNo} - S_{TKNe}) - 2.86R_{N_2} - 1.98 \frac{VX_v}{\Theta_X} \quad (11.38)$$

where  $C_{so}$  = the influent soluble plus particulate substrate concentration (kg COD/m<sup>3</sup>)  
 $C_{TKNo}$  = the influent soluble plus particulate total kjeldahl nitrogen (TKN) concentration (kg N/m<sup>3</sup>)  
 $Q$  = the settled or raw sewage flow rate (m<sup>3</sup>/s)  
 $R_{N_2}$  = the nitrogen gas production rate (kg N<sub>2</sub>/s)  
 $R_{O_2}$  = the oxygen utilization rate (kg O<sub>2</sub>/s)  
 $S_{se}$  = the final effluent soluble organic matter concentration (kg COD/m<sup>3</sup>)  
 $S_{TKNe}$  = the final effluent soluble TKN concentration (kg N/m<sup>3</sup>)  
 $V$  = the aeration tank volume (m<sup>3</sup>)  
 $X_v$  = the volatile suspended solids concentration in the aeration tank (kg VSS/m<sup>3</sup>)  
 $\Theta_X$  = the solids' retention time (s)  
 4.57 = the oxygen demand of the TKN for the conversion of TKN to HNO<sub>3</sub> (kg O<sub>2</sub>/kg TKN)  
 2.86 = the oxygen demand of nitrogen gas for the conversion of nitrogen gas to HNO<sub>3</sub> (kg O<sub>2</sub>/kg N<sub>2</sub>)  
 1.98 = the oxygen demand of the VSS, assuming complete oxidation to CO<sub>2</sub>, H<sub>2</sub>O, and HNO<sub>3</sub> (kg O<sub>2</sub>/kg VSS)

Some authors mistakenly use 1.42 as the oxygen demand of the biomass, but this ignores the oxygen demand of the reduced nitrogen in the biomass solids.

If there is no denitrification, the term for nitrogen gas production,  $R_{N_2}$ , is zero. In municipal wastewaters, there is no denitrification unless there is first nitrification, because all influent nitrogen is in the reduced forms of ammonia or organically bound nitrogen. However, some industrial wastewaters (principally explosives and some agricultural chemicals) contain significant amounts of nitrates.

If there is no nitrification, the influent TKN is merely redistributed between the soluble TKN output,  $QS_{TKNe}$ , and the nitrogen incorporated into the waste solids,  $VX/\Theta_X$ . The oxygen demand of the nitrogen in the waste solids exactly cancels the oxygen demand of the TKN removed. In the case of the input-output variables, the oxygen utilization rate becomes,

$$R_{O_2} = Q(C_{so} - S_{se}) - 1.42 \frac{VX_v}{\Theta_X} \quad (11.39)$$

$$R_{O_2} = (1 - 1.42Y_{vo})Q(C_{so} - S_{se}) \quad (11.40)$$

where 1.42 = the oxygen demand of the VSS assuming oxidation to CO<sub>2</sub>, H<sub>2</sub>O, and NH<sub>3</sub> (kg O<sub>2</sub>/kg VSS).

In the case of the Gram–Pearson–McKinney–Pirt model, the oxygen uptake rate in the aeration tank would be as follows:

oxygen uptake = substrate oxidized + biomass oxidized + maintenance

$$R_{O_2} = \left( \frac{1}{Y_{ca}} - 1 \right) \mu VX_{ca} + (1 - f_d)k_d VX_{ca} + k_{mc} VX_{ca} \quad (11.41)$$

where  $k_{mc}$  = the maintenance energy demand as COD of substrate per COD of biomass (kg COD/kg COD s)  
 $X_{ca}$  = the active biomass concentration in the aeration tank as COD rather than VSS (kg COD/m<sup>3</sup>)  
 $Y_{ca}$  = the true growth yield of the active biomass as COD on the substrate COD (kg COD/kg COD)

Substitution from Eqs. (11.3), (11.4), (11.5), and (11.10) produces,

$$R_{O_2} = Q(X_{cso} + S_{cso} - S_{cs}) - \frac{V(X_{ca} + X_{cd} + X_{cs})}{\Theta_X} \quad (11.42)$$

And, if the inert influent organic solids are added [Eq. (11.6)], one gets again Eq. (11.36) (with all the particulate organics reported as COD):

$$\begin{aligned} R_{O_2} &= Q(X_{cio} + X_{cso} + S_{cso} - S_{cs}) - \frac{V(X_{ca} + X_{cd} + X_{cs} + X_{ci})}{\Theta_X} \\ &= Q(C_{co} - S_{cs}) - \frac{VX_c}{\Theta_X} \end{aligned} \quad (11.43)$$

### Minimum Oxygen Concentration

The rate of carbonaceous BOD<sub>5</sub> removal is reduced at oxygen concentrations below about 0.5 mg/L (Orford, Heukelekian, and Isenberg, 1963). However, most authorities require higher aeration tank DOs. For nonnitrifying systems, the Joint Task Force (1988) recommends a DO of 2 mg/L under average conditions and 0.5 mg/L during peak loads. The Wastewater Committee (1997) and the Technical Advisory Board (1980) require a minimum DO of 2 mg/L at all times.

### Temperature

Field and laboratory data indicate that the BOD<sub>5</sub> removal efficiency increases from about 80–85% to 90–95% as the temperature increases from about 5 to 30°C (Benedict and Carlson, 1973; Hunter, Genetelli, and Gilwood, 1966; Keefer, 1962; Ludzack, Schaffer, and Ettinger, 1961; Sawyer, 1942; Sayigh and Malina, 1978). Increases above 30°C do not improve BOD<sub>5</sub> removal efficiency, and increases above 45°C degrade BOD removal efficiency (Hunter, Genetelli, and Gilwood, 1966).

The true growth yield coefficient,  $Y$ , does not vary with temperature.

The values of  $k_d$  and  $k_m$  are so uncertain that temperature adjustments may not be warranted; however, most engineers make temperature corrections to  $k_d$ . This is justified by the reduction in predation and consequent increase in solids' production that occurs at low temperatures (Ludzack, Schaffer, and Ettinger, 1961). Low-temperature operation also results in poorer flocculation and a greater amount of dispersed fine solids. This is also due to reduced predation. The temperature correction to the decay rate would be as follows (Grady, Daigger, and Lim, 1999; Joint Task Force, 1992):

$$\frac{k_d(T_1)}{k_d(T_o)} = 1.04^{T_1 - T_o} \quad (11.44)$$

The parameters of the Monod function vary with temperature approximately as follows (Novak, 1974; Giona et al., 1979):

$$\frac{\mu_{\max}(T_1)}{\mu_{\max}(T_o)} = \frac{q_{\max}(T_1)}{q_{\max}(T_o)} \cong 1.10^{T_1 - T_o} \quad (11.45)$$

$$\frac{K_s(T_1)}{K_s(T_o)} \cong 1.075^{T_1 - T_o} \quad (11.46)$$

Lin and Heinke (1977) analyzed 26 years of data from each of 13 municipal plants and concluded that the temperature dependence of the first-order rate coefficient was,

$$\frac{k(T_1)}{k(T_0)} \cong 1.125^{T_1 - T_0} \quad (11.47)$$

## pH

The optimum pH for the activated sludge process lies between 7 and 7.5, but pH does not substantially affect BOD<sub>5</sub> removal between about 6 and 9 (Keefer and Meisel, 1951). BOD<sub>5</sub> removal falls sharply outside that range and is reduced by about 50% at pHs of 5 or 10.

## Nutrients

Most municipal and many industrial wastewaters have a proper balance of nutrients for biological waste treatment. However, some industrial wastes may be deficient in one or more required elements. The traditional rule of thumb is that the BOD<sub>5</sub>:N and BOD<sub>5</sub>:P mass ratios should be less than 20:1 and 100:1, respectively (Helmers et al., 1951). Sludge yields and treatment efficiencies fall when the BOD<sub>5</sub>:P ratio exceeds about 220 (Greenberg, Klein, and Kaufman, 1955; Verstraete and Vissers, 1980).

Some industrial wastes are deficient in metals, especially potassium. Table 11.2 is the approximate composition of bacterial cells, and it may be used as a guide to nutrient requirements.

Waste-activated sludge is typically about 70% volatile solids, and the volatile solids contain about 7% N and 3% P. (See Table 11.3.)

## Poisons

Carbonaceous BOD<sub>5</sub> removal is not affected by salinity up to that of seawater (Stewart, Ludwig, and Kearns, 1962). Approximate concentrations at which some poisons become inhibitory are indicated in Tables 11.4 and 11.5.

## Traditional Rules of Thumb

Nowadays, most regulatory authorities have approved certain rules of thumb. An example is shown in Table 11.6 (Wastewater Committee, 1997). The rules of thumb should be used in the absence of pilot-plant data or when the data are suspect.

## Carbonaceous BOD Removal

### Solids' Retention Time

The solids' retention time should be short enough to suppress nitrification but long enough to achieve essentially complete soluble CBOD removal. At 25°C, a solids' retention time of 1 to 2 days suffices for nearly complete soluble BOD<sub>5</sub> removal, but 5 days SRT will be needed at 15°C. Satisfactory flocculation may require 3 days SRT, and the hydrolysis of particulate BOD<sub>5</sub> may require 4 days SRT (Grady, Daigger, and Lim, 1999).

If conditions are otherwise favorable, nitrification can occur at SRTs less than 3 days at warmer temperatures. Under these circumstances, aeration tank dissolved oxygen concentrations less than 2 mg/L may partially inhibit nitrification (Mohlman, 1938), but the aeration tank DO should not be so small as to limit soluble BOD<sub>5</sub> uptake and metabolism. Regulators often specify an absolute minimum DO of 1 mg/L.

It should be noted that ammonia is toxic to fish at concentrations around 1 mg/L (Table 8.7), and in many cases, the regulatory authority will require nitrification to prevent fish kills. Nonnitrifying processes are acceptable only where the receiving water provides dilution sufficient to avoid toxicity.

### System Biomass and Waste Solids Production

Once the SRT is determined, the system biomass and waste solids production rate may be calculated.

If pilot-plant data are available, the SRT determines the specific uptake rate and food-to-microorganism ratio via Eqs. (11.31) and (11.32). The definitions of  $q_v$  and  $F_v$  directly produce the required MLVSS,

**TABLE 11.2** Approximate Composition of Growing Bacteria and Nutrient Requirements for Biological Treatment

Component	Weight Percentage <sup>1,2,3</sup>			Eckenfelder's Guidelines <sup>4</sup> (mg substance/kg BOD)
	Total Weight	Dry Weight	Mole Ratio	
Water	80	—	—	—
Solids	20	100	—	—
Ash	—	7	—	—
Volatile Solids	—	93	—	—
C	—	54	6.5	—
O	—	23	2.1	—
N	—	9.6	1.0	<sup>a</sup>
H	—	7.4	10.7	—
P	—	3	0.14	<sup>b</sup>
K	—	1	0.04	4500
Mg	—	0.7	0.04	3000
Na	—	0.5	0.03	50
S	—	0.5	0.02	—
Ca	—	0.5	0.02	6200
Fe	—	0.025	—	12,000
Cu	—	0.004	—	146
Mn	—	0.004	—	100
Co	—	—	—	130
CO <sub>3</sub>	—	—	—	2700
Mo	—	—	—	430
Se	—	—	—	0.0014
Zn	—	—	—	160
Proteins	—	50	—	—
RNA	—	20	—	—
Carbohydrates	—	10	—	—
Lipids	—	7	—	—
DNA	—	3	—	—
Inorganic ions	—	3	—	—
Small molecules	—	3	—	—

<sup>a</sup> N(kg/d) = [0.123·f + 0.07·(0.77 - f)]·ΔMLVSS(kg/d)/0.77, where f = the biodegradable fraction of the MLVSS ≈ 0.6 to 0.4 at SRTs of 1 to 4 days, respectively.

<sup>b</sup> P(kg/d) = [0.026·f + 0.01·(0.77 - f)]·ΔMLVSS(kg/d)/0.77, where f = the biodegradable fraction of the MLVSS ≈ 0.6 to 0.4 at SRTs of 1 to 4 days, respectively.

Sources:

<sup>1</sup> Bowen, H.J.M. 1966. *Trace Elements in Biochemistry*. Academic Press, New York.

<sup>2</sup> Porter, J.R. 1946. *Bacterial Chemistry and Physiology*. John Wiley & Sons, Inc., New York.

<sup>3</sup> Watson, J.D. 1965. *Molecular Biology of the Gene*. W.A. Benjamin, Inc., New York.

<sup>4</sup>Eckenfelder, W.W., Jr. 1980. "Principles of Biological Treatment," p. 49 in *Theory and Practice of Biological Wastewater Treatment*, K. Curi and W.W. Eckenfelder, Jr., eds. Sijthoff & Noordhoff International Publishers BV., Germantown, MD.

$VX_v$ . Furthermore, the assumption of perfect clarification ( $X_e = 0$ ) produces an upper limit on the waste sludge production rate,  $Q_w X_w$ , via the definition of the SRT, Eq. (11.2).

If a calibrated model is available, the MLVSS can be calculated from Eq. (11.9) with substitutions from Eqs. (11.4), (11.5), (11.6), and (11.10):

$$VX_v = Q\Theta_X \left[ \frac{Y_a(1 + f_d k_d \Theta_X)}{1 + (k_d + k_m Y_a)\Theta_X} \left( S_{so} - S_s + \frac{k_h \Theta_X X_{so}}{1 + k_h \Theta_X} \right) + \frac{X_{vso}}{1 + k_h \Theta_X} + X_{vio} \right] \quad (11.48)$$

Except for particulate substrate, the solids terms on the right-hand side must have units of VSS. In the case of particulate substrate, in its first appearance (in parentheses), it must have the units appropriate

**TABLE 11.3** Typical Activated Sludge Compositions

Parameter	Typical	Range	References
Volatile solids (% of TSS), municipal	70	65 to 75	2, 6, 4
Volatile solids (% of TSS), industrial	—	Up to 92	3
Nitrogen (% of VSS)	7	—	1, 6, 7
Phosphorus (% of VSS), conventional plants	2.6	1.1 to 3.8	1, 5, 6, 7, 8, 9
Phosphorus (% of VSS), enhanced P-removal plants	5.5	4.5 to 6.8	10

*Sources:*

<sup>1</sup> Ardern, E. and Lockett, W.T. 1914. "Experiments on the Oxidation of Sewage Without the Aid of Filters," *Journal of the Society of Chemical Industry*, 33(10): 523.

<sup>2</sup> Babbitt, H.E. and Baumann, E.R. 1958. *Sewerage and Sewage Treatment*, 8th ed. John Wiley & Sons, Inc., New York.

<sup>3</sup> Eckenfelder, W.W., Jr., and O'Connor, D.J. 1961. *Biological Waste Treatment*. Pergamon Press, Inc., New York.

<sup>4</sup> Joint Committee of the Water Pollution Control Federation and the American Society of Civil Engineers. 1977. *Wastewater Treatment Plant Design*, Manual of Practice No. 8. Water Pollution Control Federation, Washington, DC; American Society of Civil Engineers, New York.

<sup>5</sup> Levin, G.V., Topol, G.J., Tarnay, A.G., and Samworth, R.B. 1972. "Pilot-Plant Tests of a Phosphate Removal Process," *Journal of the Water Pollution Control Federation*, 44(10): 1940.

<sup>6</sup> Levin, G.V., Topol, G.J., and Tarnay, A.G. 1975. "Operation of Full-Scale Biological Phosphorus Removal Plant," *Journal of the Water Pollution Control Federation*, 47(3): 577.

<sup>7</sup> Martin, A.J. 1927. *The Activated Sludge Process*. Macdonald and Evans, London.

<sup>8</sup> Metcalf, L. and Eddy, H.P. 1916. *American Sewerage Practice: Vol. III Disposal of Sewage*, 2nd ed. McGraw-Hill Book Co., Inc., New York.

<sup>9</sup> Mulbarger, M.C. 1971. "Nitrification and Denitrification in Activated Sludge Systems," *Journal of the Water Pollution Control Federation*, 43(10): 2059.

<sup>10</sup> Scaff, M.R., Pfeffer, F.M., Lively, L.D., Witherow, J.L., and Priesing, C.P. 1969. "Phosphate Removal at Baltimore, Maryland," *Journal of the Sanitary Engineering Division, Proceedings of the American Society of Civil Engineers*, 95(SA5): 817.

to the true growth yield constant,  $Y_o$ , which are usually kg VSS per kg COD (or BOD<sub>5</sub>). In its second appearance, in the numerator of the third fraction, it must have units of VSS. The calculation of the waste solids' production rate proceeds, as above, from the definition of the SRT.

### Aeration Tank Volume and Geometry

The aeration tank volume should be adjusted so that it can carry between 160 and 240% of the suspended solids required to treat the annual average flow and load (Table 8.13).

Aeration tanks are normally rectangular in plan and cross-sectional and much longer than they are wide or deep. Width and depth are controlled by the aeration system employed, and the length determines the hydraulic retention time. Diffusers typically have submergence depths of 12 to 20 ft, with 15 ft being common. HRTs of a few to several hours are generally required. (See Table 11.6.) Substrate removal in batch systems is generally complete in ½ hr, and the longer HRTs are required to promote flocculation and the hydrolysis of particulate substrate. Smaller HRTs produce larger values of  $X$ , so the lower limit on HRT is set by the mass transfer limits of the aeration system and by the allowable mass flux on the secondary clarifier. Very large HRTs are uneconomical.

Many aeration tanks incorporate a plug-flow selector at the inlet end to control filamentous bulking. The usual design choices are mixed-cells-in-series and sequencing batch reactors. The objective of using a plug-flow selector is to create a zone of relatively high substrate concentration near the inlet, which favors the growth of zoogloal species. Because of the speed of soluble substrate uptake, a selector consisting of mixed-cells-in-series must have very short HRTs in each cell, about 10 min maximum. Figure 11.2 shows a typical plug-flow tank with selector. POTWs are typically designed for a peak load some 20 years distant, so it is necessary to check the as-built selector HRTs to make sure they are short enough under the initial low hydraulic loads.

**TABLE 11.4** Approximate Threshold Concentrations for Inhibition of Activated Sludges by Inorganic Substances

Substance	Threshold Concentration for Inhibition (mg/L)		
	Nonnitrifying	Nitrifying	Denitrifying
Ammonia (NH <sub>3</sub> )	480	—	—
Arsenic (As)	0.1	—	—
Arsenate (AsO <sub>2</sub> )	—	—	1.0
Barium (Ba)	—	—	0.1
Borate (BO <sub>4</sub> )	10	—	—
Cadmium (Cd)	1	5	1.0
Calcium (Ca)	2500	—	—
Chromium(Cr VI)	1	0.25	0.05
Chromium(Cr III)	50	—	0.01
Copper (Cu)	0.1	0.05	20
Cyanide (CN)	0.5	0.3	0.1
Iron (Fe)	1000	—	—
Lead (Pb)	0.1	—	0.05
Magnesium (Mg)	—	50	—
Manganese (Mn)	10	—	—
Mercury (Hg)	0.1	—	0.006
Nickel (Ni)	1	0.5	5.0
Silver (Ag)	5	—	0.01
Sulfate (SO <sub>4</sub> )	—	500	—
Sulfide (S <sup>=</sup> )	25	—	—
Zinc (Zn)	0.1	0.1	0.1

Source: EPA. 1977. *Federal Guidelines: State and Local Pretreatment Programs. Volume I.* EPA—430/9–76–017a and *Volume II. Appendices 1–7.* EPA—430/9–76–017b. Environmental Protection Agency, Office of Water Programs Operations, Municipal Construction Division, Washington, DC.

Knoetze, C., Davies, T.R., and Wiechers, S.G. 1980. “Chemical Inhibition of Biological Nutrient Removal Processes,” *Water SA*, 6(4): 171.

The Joint Task Force (1992) summarizes a number of design recommendations and notes that there is no consensus on the design details for selectors. Anoxic denitrification zones in semiaerobic (nitrification/denitrification) plants are effective selectors, because they deny the filamentous microbes access to oxygen.

### Air Supply and Distribution

Air supply is generally based on the maximum 1-hr BOD<sub>5</sub> load on the aeration tank, which is about 280% of the annual average BOD<sub>5</sub> load (Table 8.13).

In mixed-cells-in-series, the oxygen uptake rate is highest in the inlet compartment and lowest in the outlet compartment. Consequently, the rate of oxygen supply must be “tapered.” A commonly recommended air distribution is given in Table 11.7. This should be checked against the mixing requirements, which are about 10 to 15 scfm per 1000 cu ft of aeration volume for diffused air grid systems and 15 to 25 scfm per 1000 cu ft of aeration volume for spiral flow systems (Joint Task Force, 1988).

### Secondary Clarifiers

Secondary clarifiers (not the bioprocess) control final effluent quality, and engineers must exercise special care in their design.

In order to avoid floc breakup, the Aerobic Fixed-Growth Reactors Task Force (2000) recommends that the outlets of aeration tanks should not have waterfalls higher than 0.2 m, and that all piping, channels, and structures between the aeration tank and the clarifier should have peak velocities less than 0.6 m/s. Transfer channels should not be aerated, and 5 min of hydraulic flocculation should be provided

**TABLE 11.5** Approximate Threshold Concentrations for Inhibition of Activated Sludges by Organic Substances

Substance	Threshold Concentration for Inhibition (mg/L)		
	Nonnitrifying	Nitrifying	Denitrifying
Acetone	—	840	—
Allyl alcohol	—	19.5	—
Allyl chloride	—	180	—
Allyl isothiocyanate	—	1.9	—
Analine	—	0.65	—
Benzidine	500	—	—
Benzyl thiuronium chloride	—	49	—
CARBAMATE	0.5	0.5	—
CARBARYL	—	—	10
Carbon disulfide	—	35	—
CEEPRYN™	100	—	—
CHLORDANE™	—	0.1	10
2-chloro-6-trichloro-methyl-pyridine	—	100	—
Creosol	—	4	—
Diallyl ether	—	100	—
Dichlorophen	—	50	—
Dichlorophenol	0.5	5.0	—
Dimethyl ammonium dimethyl dithiocarbamate	—	19.3	—
Dimethyl paranitroso aniline	—	7.7	—
DITHANE	0.1	0.1	10
Dithiooxamide	—	1.1	—
EDTA	25	—	—
Ethyl urethane	—	250	—
Guanadine carbonate	—	19	—
Hydrazine	—	58	—
8-Hydroxyquinoline	—	73	—
Mercaptothion	10	10	10
Methylene blue	—	100	—
Methylisothiocyanate	—	0.8	—
Methyl thiuronium sulfate	—	6.5	—
NACCONOL™	200	—	—
Phenol	—	4	0.1
Piperidinium cyclopentamethylene dithiocarbamate	—	57	—
Potassiumthiocyanate	—	300	—
Pyridine	—	100	—
Skatole	—	16.5	—
Sodium cyclopentamethylene dithiocarbamate	—	23	—
Sodium dimethyl dithiocarbamate	—	13.6	—
Streptomycin	—	400	—
Strychnine hydrochloride	—	175	—
Tetramethyl thiuram disulfide	—	30	—
Tetramethyl thiuram monosulfide	—	50	—
Thioacetamid	—	0.14	—
Thiosemicarbazide	—	0.18	—
Thiourea	—	0.075	—
Trinitrotoluene	20	—	—

Source: EPA. 1977. *Federal Guidelines: State and Local Pretreatment Programs. Volume I.* EPA—430/9–76–017a and *Volume II. Appendices 1–7.* EPA—430/9–76–017b. Environmental Protection Agency, Office of Water Programs Operations, Municipal Construction Division, Washington, DC.

Knoetze, C., Davies, T.R., and Wiechers, S.G. 1980. "Chemical Inhibition of Biological Nutrient Removal Processes," *Water SA*, 6(4): 171.

**TABLE 11.6** Summary of Recommended Standards for Wastewater Facilities

Treatment Scheme	Food-to-Microorganism Ratio, $F$ (kg BOD <sub>5</sub> /kgVSSday)	Mixed Liquor Total Suspended Solids $X$ (mg TSS/L)	Aeration Tank Load (kg BOD <sub>5</sub> /m <sup>3</sup> ·day)	Air Supply (m <sup>3</sup> /kg BOD <sub>5</sub> )	Return Sludge Flow $Q_r$		Secondary Clarifier Overflow Rate (dm <sup>3</sup> /m <sup>2</sup> ·s)	Secondary Clarifier Solids' Flux (kg TSS/m <sup>2</sup> ·d)
					% Design Ave Flow			
					min	max		
Conventional, nonnitrifying, plug flow	0.2–0.5	1000–3000	0.64	94	15	100	0.56	245
Conventional, nonnitrifying, complete mix	0.2–0.5	1000–3000	0.64	94	15	100	0.56	245
Step aeration	0.2–0.5	1000–3000	0.64	94	15	100	0.56	245
Contact—stabilization	0.2–0.6	1000–3000	0.8	94	50	150	0.56	245
Extended aeration	0.05–0.1	3000–5000	0.24	128	50	150	0.47	171
Single—stage nitrification	0.05–0.1	3000–5000	0.24	—	—	—	0.47	171
Two—stage nitrification, carbonaceous stage	—	—	—	—	15	100	0.56	245
Two—stage nitrification, nitrification stage	—	—	—	—	50	200	0.38	171

Note: The design waste sludge flow will range from 0.5 to 25% of the design average flow, but not less than 10 gpm.

Source: Wastewater Committee, Great Lakes—Upper Mississippi River Board of State Public Health and Environmental Managers. 1997. *Recommended Standards for Wastewater Facilities*, 1997 Edition of the Health Education Services, Inc., Albany, NY.

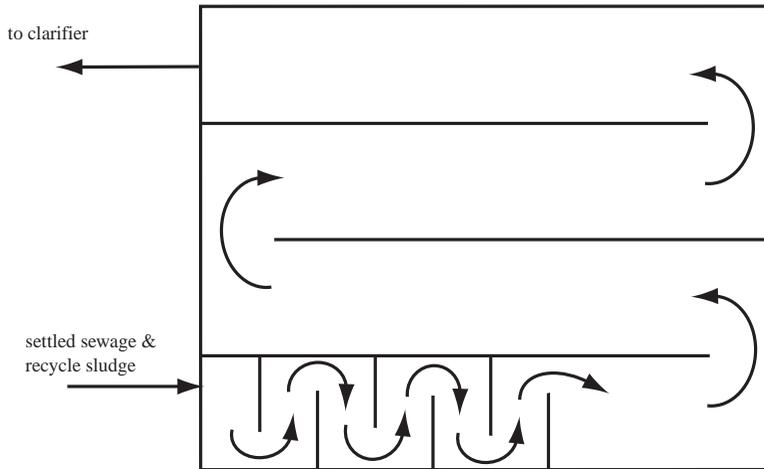


FIGURE 11.2 Conventional four-pass aeration tank with selector.

TABLE 11.7 Distribution of Oxygen Consumption Along Plug Flow and Mixed-Cells-In-Series Aeration Tanks

Aeration Tank Volume	Carbonaceous Demand (%)	Carbonaceous Plus Nitrogenous Demand (%)
First fifth	60	46
Second fifth	15	17
Third fifth	10	14
Fourth fifth	10	13
Last fifth	5	10

Source: Boon, A.G. and Chambers, B. 1985. "Design Protocol for Aeration Systems — U.K. Perspective," in *Proceedings — Seminar Workshop on Aeration System Design, Testing, Operation, and Control*, EPA 600/9-85-005, W.C. Boyle, ed. Environmental Protection Agency, Risk Reduction Engineering Laboratory, Cincinnati, OH.

as the mixed liquor enters the clarifier or just before it enters. In rectangular clarifiers, the flocculation chamber is separate and has a low headloss diffusion inlet. In circular clarifiers, the flocculator is part of or an extension of the feed well.

The Joint Task Force (1992) indicates that shape has little influence on annual average effluent SS concentration between surface overflow rates of 400 and 800 gallon per square foot per day.

The side water depth of circular clarifiers should be at least 11 ft if their diameter is about 40 ft and at least 15 ft if the diameter is about 140 ft. Increasing improvements in effluent SS quality accrue as side water depths increase to 18 ft. Rectangular tanks may be somewhat shallower.

Secondary activated sludge clarifiers accumulate solids during peak flows if their allowable solids flux is exceeded. The clarifier depth should include an allowance for the increased sludge blanket depth. The Aerobic Fixed-Growth Reactors Task Force (2000) recommends that the allowance be less than 0.6 to 0.9 m above the average top of the sludge blanket.

The side water depth is also related to the overflow rate. For primary clarifiers, intermediate clarifiers, and trickling filter clarifiers with floor slopes greater than 1:12, the relationships are as follows (Aerobic Fixed-Growth Reactors Task Force, 2000):

$$v_{\max} \leq 0.182H_{sw}^2; 1.8 \text{ m} \leq H_{sw} \leq 3.0 \text{ m} \quad (11.49)$$

$$v_{ave} \leq 0.092H_{sw}^2; 1.8 \text{ m} \leq H_{sw} \leq 3.0 \text{ m} \quad (11.50)$$

$$v_{max} \leq 0.556H_{sw}; 3.0 \text{ m} \leq H_{sw} \leq 4.6 \text{ m} \quad (11.51)$$

$$v_{ave} \leq 0.278H_{sw}; 3.0 \text{ m} \leq H_{sw} \leq 4.6 \text{ m} \quad (11.52)$$

where  $H_{sw}$  = the side water depth (m)  
 $v_{ave}$  = the average overflow rate (m/h)  
 $v_{max}$  = the maximum overflow rate (m/n = h)

For activated sludge, trickling filter-activated sludge, and trickling filter-solids contacting with bottom slopes greater than 1:12, the relationships are as follows (Aerobic Fixed-Growth Reactors Task Force, 2000):

$$v_{max} \leq 0.182H_{cw}^2; 1.8 \text{ m} \leq H_{cw} \leq 3.0 \text{ m} \quad (11.53)$$

$$v_{ave} \leq 0.092H_{cw}^2; 1.8 \text{ m} \leq H_{cw} \leq 3.0 \text{ m} \quad (11.54)$$

$$v_{max} \leq 0.556H_{cw}; 3.0 \text{ m} \leq H_{cw} \leq 4.6 \text{ m} \quad (11.55)$$

$$v_{ave} \leq 0.278H_{cw}; 3.0 \text{ m} \leq H_{cw} \leq 4.6 \text{ m} \quad (11.56)$$

where  $H_{cw}$  = the clear water depth, the depth of supernatant water above the maximum elevation of the sludge blanket (m).

Weir loading is unimportant, but placement is important. In rectangular tanks, the effluent launders should be placed in the last one-fourth to one-third of the tank length. Circular tanks should have double launders placed seven-tenths of the radius from centerline. Wall-mounted launders should have baffles placed beneath them to deflect upward flows along the wall. Launders cantilevered inboard can also deflect the upward flows. Baffles should project horizontally at least 18 in. in a 30 ft diameter clarifier, and the projection should increase above 18 in. by about 0.2 in./ft of diameter up to a maximum projection of 48 in. from the clarifier wall. The effluent weirs should be protected from scum by vertical baffles.

### Operational and Design Problems

The chief operational problem of the conventional activated-sludge process is filamentous bulking. The filamentous bacteria responsible for bulking are strict aerobes (some are microaerophilic) and are limited to the catabolism of small organic molecules (simple sugars, volatile fatty acids, and short-chain alcohols). They grow faster than the zoogloal bacteria at low concentrations of oxygen, nutrients, and substrates, and come to dominate the activated sludge community under those conditions. When this happens, the flocs settle slowly, and the secondary clarifier may fail to achieve adequate solids/liquid separation.

Completely mixed aeration tanks are especially prone to bulking, because the substrate concentration is low everywhere in such a tank. However, high-rate completely mixed processes, which produce relatively high effluent BODs, do not usually bulk. Mixed-cells-in-series selectors can produce bulked sludges if the HRT of the first cell is longer than about 10 min. Ideal plug flow tanks (see "Sequencing Batch Reactors" below) may produce bulked sludges if the aeration system cannot maintain at least 1 or more mg/L of oxygen at the inlet or if the influent waste is weak and largely soluble. However, most filamentous microbes are strictly aerobic, even the microaerophilic species. For that reason, semiaerobic designs in which the initial biomass-sewage contacting chamber is anoxic have become almost standard practice.

Some facilities have inadequate return sludge and waste sludge capacity or lack of control over flow rates and monitoring of flow rates, or both. Excessive return flow may hydraulically overload the secondary clarifier. Inadequate sludge return or wasting may allow solids to reside too long in the clarifier, producing gases and rising sludge. This is a special problem in nitrification facilities. Pumps and pipes are susceptible to plugging from debris.

Porous diffusers occasionally clog on the air supply side or the mixed liquor side (Joint Task Force, 1988). Airside clogging is due to suspended solids in the flow. This may be derived from dust in the local atmosphere, corrosion of the air piping, dislodgement of air supply pipe liner, leftover construction debris, or leaks that admit mixed liquor during out-of-service periods. Clogs develop in the mixed liquor side because of high soluble BOD concentrations, high soluble iron concentrations, low mixed liquor DO, high C:N or C:P ratios in the feed, and low unit airflow (especially due to uneven air distribution).

Disc, brush, and surface aerators are liable to accumulate ice in cold weather and require protective enclosures. They also tend to accumulate debris that passes through the preliminary treatment process. In aeration tanks deeper than about 15 ft, draft tubes are necessary to ensure the whole depth is mixed (Joint Task Force, 1992). Otherwise, the MLSS may settle out, forming odor-generating sludge deposits and minimizing sewage-biomass contact.

All aeration systems (except bubbleless membranes) produce aerosols and strip volatile organic compounds. Buffer strips around the facility sometimes provide adequate dilution of the contaminants before breezes reach the surrounding community. In other cases, capture and treatment of the aeration tank off-gases may be required.

Aeration tanks generally emit a nonoffensive musty odor. Other odors may indicate inadequate aeration.

## Sequencing Batch Reactors

In recent years, sequencing batch reactors (SBR) have become quite common. They combine high turbulence (which promotes high mass transfer rates from the sewage to the activated sludge flocs) with batch reaction conditions (which tends to suppress filamentous microbes). They operate as follows:

- Starting out empty, the tank is first filled; any needed chemicals are added during the filling.
- The full tank is then stirred and aerated (as needed), and the reactions proceed.
- After mixing and reacting, the tank is allowed to stand quiescently to settle out the MLSS.
- The tank supernatant is drained off.
- The tank may sit idle between the draining and filling operations, while valves and pumps are switched.

The SBR is the fill-and-draw operating mode used by Arden and Lockett (1914) and many others studying the activated sludge process. It is also the mode of operation of what used to be called “contact beds,” a form of sewage treatment employed around the turn of the century and a predecessor of the trickling filter (Dunbar, 1908; Metcalf and Eddy, 1916).

### Phase Scheduling

The design problem is scheduling the filling, stirring, draining, and idle phases of the cycle so as to meet the plant design flow rate.

If water production is to be continuous, at least one tank must be filling and one draining at each moment. Consider the schedule shown in Fig. 11.3. The total cycle time for a single tank is the sum of the times for filling, reacting, settling, draining, and idling. The plant flow first fills Tank No. 1. Then the flow is diverted to Tank No. 2, and so on. Tank No. 1 is again available after its cycle is completed. Raw water will be available at that moment, if another tank has just completed filling. This means that the cycle time for a single tank must be equal to or less than the product of the filling time and the number of available tanks:

$$nt_f \geq t_f + t_r + t_s + t_d + t_i \quad (11.57)$$

where  $n$  = the number of tanks (dimensionless)  
 $t_d$  = the (clear) supernatant decanting time (s)  
 $t_f$  = the filling time (s)  
 $t_i$  = the idle time (s)  
 $t_r$  = the reaction time (s)  
 $t_s$  = the settling time (s)

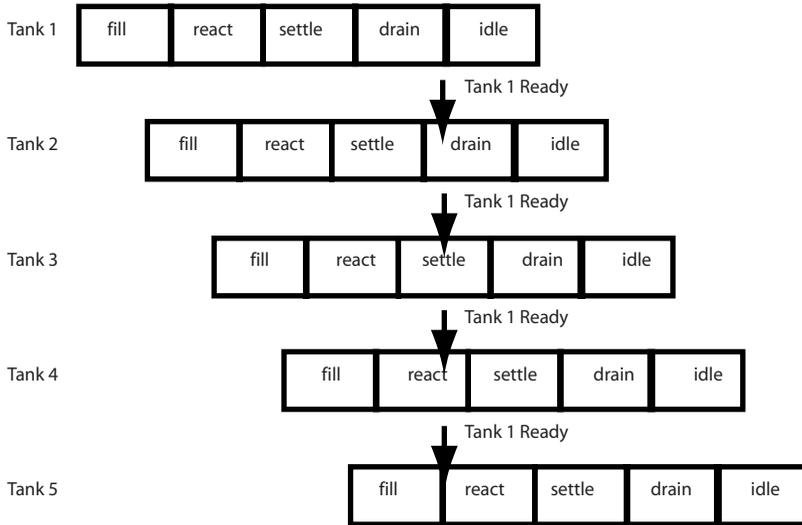


FIGURE 11.3 Schedule for sequencing batch reactor operation.

The idle time is freely adjustable, but the other times are set by the hydraulic capacity of the inlet and outlet devices and the time required for reaction and settling. A similar analysis of the draining operation leads to the conclusion that the product of the number of tanks and the draining time must equal the cycle time:

$$nt_d \geq t_f + t_r + t_s + t_d + t_i \quad (11.58)$$

The number of tanks in both equations must be the same, so the filling time must equal the draining time. They are otherwise freely adjustable within the limits of the hydraulic system.

An SBR may not be a viable option for very short phase time, say less than 30 min. With very short phase times, electric motors do not achieve a significant duration of steady state operation. Instead, they are always in the start-up mode or just recently exited from it, and they have not had sufficient time to cool from the heavy currents drawn during start-up. Under these conditions, the motors are prone to overheating and burnout. Also, a sequence of short-duration phases requires rapid valve and pump switching to occur within a few seconds. This is intrinsically difficult, because of the weights of the equipment parts, and it imposes high mechanical stresses on them. Finally, even if quick switching can be achieved, it may cause water hammer in the conduits.

The SBR is suitable where the duration of each phase is long, say several hours, so that only a few cycles are needed each day. This is the case in many activated sludge plants.

Headloss is a problem, too. In continuous flow tanks, the headloss amounts to several centimeters, at most. However, in SBRs, the headloss is the difference between the high water level and low water level, and this may amount to a few meters. In flat country, these headlosses become a significant problem, because they must be met by pumping.

Ideal plug flow tanks constructed as many-mixed-cells-in-series are more efficient than SBR systems. The total tankage in an SBR system is the product of the number of tanks, the filling time, and the flow rate. The tankage required for an ideal plug flow reactor is the product of the flow and processing time. If the same conversion efficiency is required of both systems, the processing times will be equal, and the ratio of the system volumes will be as follows:

$$\frac{V_{SBR}}{V_{PF}} = \frac{t_f + t_r + t_s + t_d + t_i}{t_r + t_s} > 1 \quad (11.59)$$

where  $V_{SBR}$  = the volume of the sequencing batch reactor ( $m^3$ )

$V_{PF}$  = the volume of the equivalent plug flow reactor and associated clarifier ( $m^3$ ).

Plug flow reactors do not incorporate settling; a separate clarifier is required for that. Consequently, the settling time of the clarifier associated with the plug flow reactor must be included in the denominator.

### Volume, Plan Area, and Depth

The SBR volume must be sufficient to contain the required mass of MLSS for  $BOD_5$  removal and to satisfy the maximum MLSS concentration limits of the aeration system and the subsequent settling/thickening phase. The desired SRT is chosen, and the required specific uptake rate, food-to-microorganism ratio or MLVSS mass is determined as described above for conventional  $BOD_5$  removal. Aeration and settling/thickening determine the maximum MLVSS concentration, and the aeration volume follows directly.

The SBR tank must also function as a secondary clarifier/thickener, and all the design criteria applying to activated sludge settling and thickening apply to the SBR tank, too. This means that there will be a minimum plan area set by the design overflow rate and solids' flux and a minimum side water depth. The tank must also be large enough to store the sludge solids retained for the next filling phase and whatever the minimum clear supernatant depth is needed over the sludge. The settled/thickened sludge volume can be estimated from the sludge volume index (SVI). This is true only for batch settling processes; the SVI is not relevant to continuous flow clarifiers.

### Solids' Wasting

In principal, solids can be wasted (to set the required SRT) at any point in the SBR cycle. However, it must be remembered that the secondary clarifier is a selector, too—one that selects for those microbes that can form activated sludge flocs. Therefore, solids should be wasted after the settling/thickening phase and before the filling phase.

## Membrane Activated Sludge

### Membrane Types and Uses

There are five basic types of membrane processes that are useful in wastewater treatment (Stephenson et al., 2000):

**Hyperfiltration (reverse osmosis)** — Selective separation of small solutes (relative molecular weights less than a few hundred; diameter less than 1 nm) by pressure differential; depends on differing solubilities of water and solutes in dense, polymeric membrane material

**Electrodialysis** — Selective separation of small ions (relative molecular weights less than a few hundred; diameters less than 1 nm) by voltage differential; depends on magnitude, density, and sign of electrical charge on ion and ion exchange properties of dense, polymeric membrane material

**Nanofiltration (leaky reverse osmosis)** — Separation of molecules and polymers (relative molecular weights a few hundred to 20,000; diameters 1 to 10 nm) by pressure differential; depends on solubility and diffusion of solutes in membrane material and sieving; dense or porous polymeric or porous inorganic membrane materials are available

**Ultrafiltration** — Separation of large polymers, colloids, and viruses (relative molecular weights 10,000 to 500,000; diameters 0.01 to 0.1  $\mu m$ ) by pressure differential; porous polymeric or inorganic membrane material separates suspended solids by size by sieving

**Microfiltration** — Separation of bacteria, protozoan cysts, eukaryotic cells, metazoa, and activated sludge flocs (relative molecular weights above 500,000; diameters above 0.1  $\mu m$ ) by pressure differential; porous polymeric or inorganic membrane material separates suspended solids by size by sieving

As the particle sizes increase, the pressure differential required for liquid/solid separation declines sharply, and operational and capital costs become more attractive.

The applications of membranes in biological treatment include the following:

- Solids/liquid separation, including protozoan cysts, bacteria, and viruses, (eliminating clarifiers, tertiary filters, and disinfection, and stabilizing operation via positive SRT control under widely varying loading conditions)
- Bubbleless oxygen transfer (yielding 100% gas transfer efficiency and no stripping of other gases or volatile organic carbon)
- Selective substrate removal from waste and/or isolation of biomass from poisons (eliminating some wastewater pretreatments)

At present, the best-established membrane application is the separation of activated sludge flocs from the mixed liquor.

### System Configuration

Proprietary membrane activated sludge units are marketed by a number of companies (Stephenson et al., 2000), including AquaTech (BIOSUF), Bioscan A/S (BIOREK), Degremont (BRM), Enviroquip (Kubota MBR), Kubota (KUBOTA MBR), Membratex (ADUF and MEMTUF), Rhodia Group/Orelis (Pleiade and Ubis), Mutsui Chemicals, Inc. (AMSEX and Ubis), Vivendi Group/USF Gütling (Kopajet), Vivendi Group/USF Memcor (Membio), Vivendi Group/OTV (Biosep), Wehrle-Werk AG (Biomembrat), Weir Envig (ADUF), and Zenon Environmental (ZENOGEM and ZEEWEED). To date, most existing installations are relatively small and provide on-site treatment of gray water, night soil, landfill leachate, or high-strength industrial wastes, but municipal facilities are becoming common.

The usual membrane activated sludge process consists of an aeration tank and a microfiltration or ultrafiltration membrane system for solids/liquid separation; these processes are often called extractive membrane bioreactors (EMBR). The membrane replaces the secondary clarifier and any tertiary granular filtration units. Most installations can operate at MLSS concentrations up to 15,000 to 20,000 mg/L, which substantially reduces the aeration tank volume.

The membrane system may be external to the aeration tank (side stream) or submerged within it. The membranes may be hollow fiber, plate-and-sheet, tubular, or woven cloth. Except for hollow fiber membranes, operation is usually side stream. The majority of current industrial installations use tubular, side stream modules with pore sizes of 1 to 100 nm (Stephenson et al., 2000). Membranes operate with cross-flow velocities of 1.6 to 4.5 m/s to reduce fouling (Stephenson et al., 2000). Fouling, however, is inevitable due to biomass accumulation and accumulation of mineral solids, like calcium carbonate and ferric hydroxide, all of which are formed in the reactor. Mineral scale formation can sometimes be minimized by pretreatments such as pH reduction. Membrane systems usually include a cleaning mechanism that may include air scouring, back flushing, chlorination, and removal and cleaning and/or replacement. Membrane removal for cleaning and/or replacement is determined by the allowable maximum pressure differential and required minimal fluxes.

In industrial applications, specific fluxes range anywhere from 5 to 200 cu dm per sq m per bar per hour, and pressure differentials across the membrane range from 0.2 to 4 bar, depending on application (Stephenson et al., 2000). The usual industrial system employs pressures of 1.5 to 3 bar and achieves specific fluxes less than 100 cu dm per sq m per bar per hour.

The adoption of membrane activated sludge systems for municipal wastewater treatment involves several considerations (Günder, 2001). Operating fluxes are generally limited to 10 to 20 cu dm per sq m per hour at pressure differentials of 0.15 to 0.60 bar. The electric power requirements of municipal EMBR plants are generally twice that of conventional plants at an MLSS concentration of 15,000 mg/L and may be four times the conventional plant's usage if the MLSS concentration reaches 25,000 mg/L.

Mixed liquors become increasingly non-Newtonian as the concentration of SS exceeds a couple thousand mg/L. In the first instance, this shows up as rapidly increasing viscosity (35% greater than water at 3000 mg/L and double that of water at 7000 mg/L), and this, in turn, impairs all mass transfer processes dependent on viscosity, such as membrane flux rate, gas and heat transfer rates, settling/thickening, pumping, and transport via channels or pipes. The mixed liquor is gel-like and exhibits ductile flow behavior due to the extracellular polysaccharides excreted by the microbes and to the increased numbers

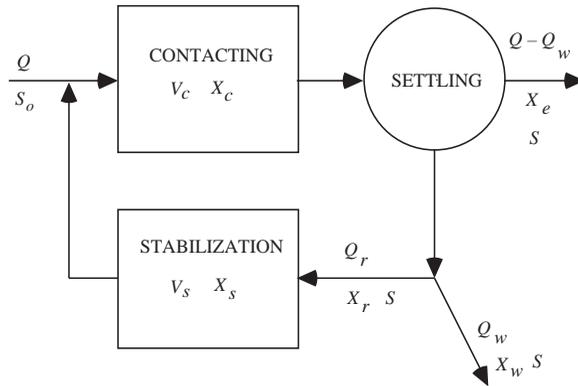


FIGURE 11.4 The contact-stabilization process.

of filaments in the sludge. Slowly rotating mixers produce a rotating core near the mixer axis and are surrounded by an unmixed stagnant zone. Gas bubbles tend to coalesce into very large bubbles, and small bubbles may remain trapped for long time periods in the mixed liquor.

Oxygen consumption, waste sludge production, fraction VSS in the MLSS, and soluble effluent COD appear to be similar to those in conventional plants, but the effluent BOD<sub>5</sub>, coliform count, and plaque-forming units in the EMBR's effluent are much less than in a conventional plant's effluent.

Bubbleless aeration membrane systems called membrane aeration bioreactors (MABR) also are in use. These are more properly classified as fixed film reactors, because the biomass adheres to the wastewater side of the membrane.

Anaerobic membrane bioreactors are also available.

## The Contact-Stabilization Process

Contact-stabilization (Zablatsky, Cornish, and Adams, 1959) is an important modification of the conventional, nonnitrifying activated sludge process. Synonyms are *sludge reaeration* (Arden and Lockett, 1914a, 1914b), *bioflocculation* (Martin, 1927), *biosorption* (Ullrich and Smith, 1951) and *step-aeration* (Torpey, 1948). Variants are the *Hatfield Process* and the *Kraus Process* (Haseltine, 1961).

The distinguishing feature of contact-stabilization is the inclusion of a tank for the separate aeration of the return activated sludge (Fig. 11.4). The practical effect is that a much smaller total aeration tank volume is needed to hold the required system biomass, because the biomass is held at a higher average concentration.

The basic design principle proposed by Haseltine (1961) is that the distribution of activated sludge solids between the contacting and stabilization tanks does not affect the required system biomass. Haseltine's data, collected from 36 operating facilities, may be expressed mathematically as follows:

$$V_c X_{vc} + V_s X_{vs} = \frac{Q(C_{bo} - S_{be})}{q_v} = \frac{QC_{bo}}{F} \quad (11.60)$$

- where
- $C_{bo}$  = the settled sewage total BOD<sub>5</sub> (kg BOD<sub>5</sub>/m<sup>3</sup>)
  - $F$  = the food-to-microorganism ratio (kg BOD<sub>5</sub>/kg VSS·s)
  - $Q$  = the settled sewage flow rate (m<sup>3</sup>/s)
  - $q_v$  = the specific uptake rate (kg BOD<sub>5</sub>/kg VSS·s)
  - $S_{be}$  = the soluble effluent CBOD<sub>5</sub> (kg BOD<sub>5</sub>/m<sup>3</sup>)
  - $V_c$  = the volume of the contacting tank (m<sup>3</sup>)
  - $V_s$  = the volume of the stabilization tank (m<sup>3</sup>)
  - $X_{vc}$  = the concentration of VSS in the contacting tank (kg/m<sup>3</sup>)

$X_{vs}$  = the concentration of VSS in the stabilization tank (kg/m<sup>3</sup>)

The usual relationship between SRT, specific uptake rate, and decay rate applies, except the decay rate should be corrected for the distribution of solids between the contacting and stabilization tanks. Jenkins and Orhon (1973) recommend:

$$k_d = \beta \cdot \left[ \frac{0.74q}{(1-\beta)3.2 + q_v} \right] \quad (11.61)$$

where  $k_d$  = the specific decay rate (per day)  
 $q_v$  = the specific uptake rate (kg COD/kg VSS·d)  
 $\beta$  = the weight fraction of the total activated sludge inventory that is in the stabilization basin (dimensionless)

Most engineers make the contacting and stabilization tanks of equal volume; some regulatory bodies require that at least one-third of the total aeration tankage be in the contacting tank (Wastewater Committee, 1997). Another rule of thumb is that the contacting period, counting recycle flows, should be less than 1 hr. Excessively long contacting periods negate the benefits of the process.

The contacting tank is usually constructed as mixed-cells-in-series. No particular design is used for the stabilization tank, because substrate removal does not occur in this tank.

The return sludge flow rate can be approximated using the Benefield–Randall formula [Eq. (11.37)].

Although there are some solids' losses in the stabilization tank due to microbial respiration (mostly from predation upon bacteria and fungi), the suspended solids' concentrations entering and leaving the stabilization tank are nearly equal.

The total oxygen consumption rate of a contact/stabilization system is the same as that of a conventional system operated at the same F/M ratio (Haseltine, 1961). The rate of oxygen consumption in the stabilization tank is probably about the same as that in the final 60 to 80% of the conventional, mixed-cells-in-series aeration tank. The recommendations of Boon and Chambers (1985), given in Table 11.7, would suggest that the stabilization tank could account for 25 to 40% of the total oxygen consumption. However, the uncertainties in this process require considerable flexibility in the capacity of the air distribution system. It is probably desirable to size the air piping and diffuser systems so that either the contacting tank or the stabilization tank could receive 100% of the estimated total air requirement.

## The Extended Aeration Process

F. S. Barckhoff, the plant operator in East Palestine, OH, discovered the extended aeration process in about 1947 (Knox, 1958, 1959–60). Originally, it was called “aerobic digestion,” because its purpose is to stabilize waste activated sludge in the aeration tank rather than in separate sludge digestion facilities. In older designs, there was no intentional wasting of activated sludge. The only solids to leave the system were those in the final effluent. The result was SRTs on the order of months to a year or more. The chief advantage of the extended aeration process was the elimination of waste sludge processing facilities. The chief disadvantages were the increase in oxygen consumption (needed to decompose the activated sludge solids) and the lack of kinetic control.

Full-scale plants fed dairy wastes have operated for periods up to a year without sludge wasting (Forney and Kountz, 1959; Kountz and Forney, 1959). Laboratory units fed glucose and soluble nutrients and managed so that no solids left the systems (other than those removed in sampling) have operated for periods of three years (SRT of 500 days) without net solids accumulation (Gaudy et al., 1970; Gaudy, Yang, and Obayashi, 1971). However, the MLSS in the laboratory units fluctuated widely during this time.

Nowadays, extended aeration plants normally incorporate solids wasting facilities, and the name really means plants with long aeration periods (usually 24 hr) and long SRTs (about 30 days).

In terms of the Gram–Pearson theory, the ideal extended aeration plant is operated so that sludge growth just equals sludge decay. Consequently, one has,

**TABLE 11.8** Typical Parameter Values for the Extended Aeration Activated Sludge Process at Approximately 20°C

Parameter	Symbol	Units	Value
True yield	$Y_v$	kg VSS/kg COD	0.4
		kg VSS/kg BOD <sub>5</sub>	0.7
Decay rate	$k_d$	Per day	0.012
Maximum specific growth rate	$\mu_{\max}$	Per day	0.04
Maximum uptake rate	$q_{\max}$	kg COD/kg VSS d	0.05
Affinity constant	$K_s$	mg CBOD <sub>5</sub> /L	10

Source: Goodman, B.L. and Englande, A.J., Jr. 1974. "A Unified Model of the Activated Sludge Process," *Journal of the Water Pollution Control Federation*, 46(2): 312.

Middlebrooks, E.J. and Garland, C.F. 1968. "Kinetics of Model and Field Extended-Aeration Wastewater Treatment Units," *Journal of the Water Pollution Control Federation*, 40(4): 586.

Middlebrooks, E.J., Jenkins, D., Neal, R.C., and Phillips, J.L. 1969. "Kinetics and Effluent Quality in Extended Aeration," *Water Research*, 3(1): 39.

Morris, G.L., Van Den Berg, L., Culp, G.L., Geckler, J.R., and Porges, R. 1963. *Extended Aeration Plants and Intermittent Watercourses*, Public Health Service Pub. No. 999-WP-8. Department of Health, Education and Welfare, Public Health Service, Division of Water Supply and Pollution Control, Cincinnati, OH.

$$\mu = k_d \quad (11.62)$$

$$S_s = \frac{k_d K_s}{Yq_{\max} - k_d} \quad (11.63)$$

If there were no inert organic and inorganic solids inflow or production, all the biomass eventually would be oxidized, and the SRT would become infinite. However, no actual extended aeration facility can operate at infinite SRT. Most wastewaters contain significant amounts of clay and other inert inorganic solids, and various inert organic solids and inorganic precipitates are formed in the aeration tank. All of these gradually accumulate in the system, and in the absence of a clarifier, they must be discharged in the settled effluent.

Extended aeration plants accumulate suspended solids until the secondary clarifier fails, producing periods of very high effluent suspended solids concentrations (Morris et al., 1963). These sludge discharges usually are associated with high flows or bulking. Bulking is always a hazard, because most extended aeration plants incorporate completely mixed aeration tanks.

The SRTs employed in extended aeration are sufficient to permit year-round nitrification in most locales; however, traditional designs did not provide sufficient aeration capacity to support nitrification.

Extended aeration plants are frequently built as oxidation ditches or as aerated ponds. Both of these designs may incorporate anoxic zones. In the case of oxidation ditches that employ surface aerators for both aeration and mixing, the bottom may accumulate a deposit of activated sludge. In the case of ponds, the placement of surface aerators may result in unaerated zones (Nicholls, 1975; Price et al., 1973). These anoxic zones can be the sites of denitrification, and extended aeration plants may accomplish substantial nitrogen removal even in the absence of sludge wasting.

Some values of the Gram–Pearson kinetic model parameters are given in Table 11.8. It should be noted that the decay rate appears to decline at long SRTs. For SRTs up to about 40 days, Goodman and Englande (1974) recommend the following:

$$k_d = 0.48 \times 0.75^{\ln \Theta_x / \ln 2} \times 1.075^{T-20} \quad (11.64)$$

where  $k_d$  = the decay rate (per day)

$T$  = the temperature (°C)  
 $\Theta_x$  = the solids' retention time (days)

For SRTs on the order of hundreds of days, the decay rate predicted by Eq. (11.64) may be high by a factor of about two.

Extended aeration plants have usually been recommended for small or isolated installations where regular, trained operating staffs are difficult to obtain. This practice results in serious operation and maintenance problems, most typically (Guo, Thirumurthi, and Jank, 1981):

- Clogging of comminutors
- Clogging of sludge return lines
- Clogging of air diffusers
- Failure of skimmers
- Icing of clarifier outlet weirs
- Insufficient biomass
- Insufficient aeration
- Offensive odors

Extended aeration plants are also subject to bulking and scum formation, but this is a result of the combination of completely mixed aeration tanks and long SRTs.

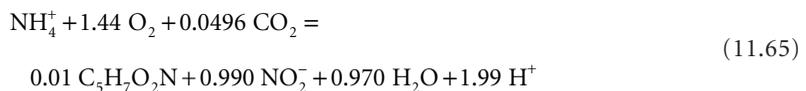
## Nitrification

Nitrification is required to prevent fish kills due to ammonia toxicity and to prepare for nitrogen removal by denitrification. In single-sludge systems, the SRT required for nitrification also applies to the heterotrophs. This produces a large MLVSS inventory and large aeration tanks. Two-sludge processes separate CBOD removal from nitrification, and the resulting total system MLVSS and aeration tankage are substantially reduced. However, an additional, intermediate clarifier is needed.

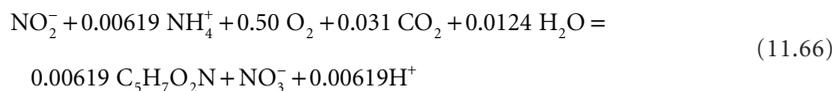
## Microbiology

Nitrification is the biological conversion of ammonia to nitrate. Aerobic, chemoautotrophic bacteria do it in two steps (Scheible and Heidman, 1993).

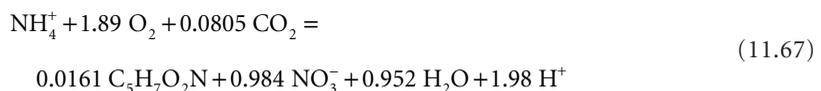
### *Nitrosomonas, Nitrospira, Nitrosococcus, and Nitrosolobus:*



### *Nitrobacter, Nitrospina, and Nitrococcus:*



### **Overall:**



The first step is slow and controls the overall rate of conversion. Consequently, in most systems, only small amounts of nitrite are observed, and the process can be represented as a one-step conversion of

ammonia to nitrate. Less than 2% of the ammonia-nitrogen is incorporated into new cells; the rest is oxidized.

### Biokinetics

At low ammonia concentrations, the rate of growth of the rate-controlling *Nitroso*- genera can be correlated with the ammonia concentration using Monod kinetics:

$$\mu_n = \frac{\mu_{\max n} \cdot S_{na}}{K_{na} + S_{na}} \quad (11.68)$$

where  $K_{na}$  = the ammonia affinity constant (kg NH<sub>3</sub>-N/m<sup>3</sup>)  
 $S_{na}$  = the total ammonia concentration (kg NH<sub>3</sub>-N/m<sup>3</sup>)  
 $\mu_n$  = the specific growth rate of the *Nitroso*- genera (per sec)  
 $\mu_{\max n}$  = the maximum specific growth rate of the *Nitroso*- genera (per sec)

The values of the kinetic parameters are usually estimated as follows (Knowles, Downing, and Barrett, 1965; Parker et al., 1975; Scheible and Heidman, 1993):

**pH < 7.2**

$$\mu_{\max n} = 0.47 \cdot \exp\{0.098(T - 15)\} \cdot [1 - 0.833(7.2 - \text{pH})] \cdot \left( \frac{S_{O_2}}{S_{O_2} + 1.3} \right) \quad (11.69)$$

**7.2 < pH < 9**

$$\mu_{\max n} = 0.47 \cdot \exp\{0.098(T - 15)\} \cdot \left( \frac{S_{O_2}}{S_{O_2} + 1.3} \right) \quad (11.70)$$

**Any pH**

$$K_{na} = 10^{0.051T - 1.158} \quad (11.71)$$

where  $K_{na}$  = the affinity constant (mg NH<sub>3</sub>-N/L)  
pH = the aeration tank pH (standard units)  
 $S_{O_2}$  = the aeration tank dissolved oxygen concentration (mg/L)  
 $T$  = the aeration tank temperature (°C)  
 $\mu_{\max N}$  = the maximum growth rate of the *Nitroso*- genera (per day)

Scheible and Heidman (1993) recommend a constant value of the affinity constant of 1 mg NH<sub>3</sub>-N/L because of the high degree of variability in the reported values.

The maximum growth rate of the nitrifiers is much smaller than that of the heterotrophs, so nitrification is sensitive to SRT at cold temperatures.

At 20°C, the affinity coefficient is predicted to be about 0.73 mg N/L. This means that the rate of nitrification is independent of ammonia concentration (zero order) down to concentrations on the order of 1 mg NH<sub>3</sub>-N/L. The practical consequence of this is that reactor configuration has little effect on overall ammonia removal. Aeration tanks are sized assuming complete mixing, which is conservative, because the effective ammonia specific uptake rate in mixed-cells-in-series is  $\mu_{\max n}/Y_n$ .

The maximum specific growth rates of the *Nitroso*- genera fall off sharply above pH 9, and the estimators do not apply above that pH.

The nitrification process produces nitric acid [Eq. (11.67)], and in poorly buffered waters, this may cause a significant decline in pH: 1 g NH<sub>3</sub>-N reduces the alkalinity by 7.14 g (as CaCO<sub>3</sub>).

The affinity constant for the DO correction, given here as 1.3 mg/L, may be as high as 2 mg/L in some systems. Scheible and Heidman (1993) recommend a value of 1 mg/L.

### Ammonia Inhibition

Un-ionized ammonia, i.e.,  $\text{NH}_3$ , is inhibitory to the *Nitroso*- and the *Nitro*- genera. The inhibition threshold concentrations of un-ionized ammonia for the *Nitroso*- group is about 10 to 150 mg  $\text{NH}_3/\text{L}$ ; for the *Nitro*- group, it is about 0.1 to 1 mg  $\text{NH}_3/\text{L}$ . Inhibition of the *Nitro*- group results in the accumulation of nitrite.

The usual way to handle these effects is to adopt the Haldane kinetic model for the specific growth rate (Haldane, 1930):

$$\mu_n = \frac{\mu_{\max n} \cdot S_{na}}{K_{na} + S_{na} + \frac{S_{na}^2}{K_i}} \quad (11.72)$$

where  $K_i$  = the Haldane inhibition constant (kg  $\text{NH}_3\text{-N}/\text{m}^3$ ).

In heterotroph-free cultures, the inhibition constant has been reported to be about 20 mg N/L at 19°C and pH 7 (Rozich and Castens, 1986). The basis here is the total ammonia concentration, both  $\text{NH}_3\text{-N}$  and  $\text{NH}_4^+\text{-N}$ . Under the given conditions, the ratio of total ammonia concentration to un-ionized ammonia would be about 250:1. At a full-scale nitrification facility treating landfill leachate, the observed inhibition constant was 36 mg/L of total ammonia nitrogen (Keenan, Steiner, and Fungaroli, 1979). The wastewater temperature varied from 0 to 29°C, and the pH varied from 7.3 to 8.6.

### Other Inhibitors

A list of other inhibitors and the approximate threshold concentration for nitrification inhibition is given in [Tables 11.4](#) and [11.5](#). The reduction in nitrification rate in some industrial wastewaters due to inhibitors can be severe. Adams and Eckenfelder (1977) give some laboratory data for nitrification rates for pulp and paper, refinery, and phenolic wastes that are only about 0.1% of the rates in municipal wastewater. The reported rates are low by an order of magnitude even if ammonia inhibition is accounted for.

### Carbon:Nitrogen Ratio of Feed

The usual C:N ratio in municipal wastewater is about 10 to 15. However, in many industrial wastewaters, it may be higher or lower. In general, increasing the C:N ratio increases the heterotrophic biomass and the “endogenous” solids in the mixed liquor.

Because the heterotrophs can metabolize at much lower oxygen concentrations than can the nitrifiers, and at higher C:N ratios, the heterotrophs can reduce the aeration tank DO below the levels needed by the nitrifiers. Consequently, at least in plug flow tanks, the zone of active nitrification moves toward the outlet end of the aeration tank, and a longer aeration period may be needed.

The elevated MLSS concentrations also require larger aeration tanks, if for no other reason than the solids’ flux on the secondary clarifier must be limited.

### Design Solids’ Retention Time

The minimum solids’ retention required for nitrification ranges from about 3 days or less at 25°C to over 18 days at 12 to 15°C (Grady, Daigger, and Lim, 1999). The usual design procedure is to do the following (Metcalf & Eddy, Inc., 2002):

- Calculate the specific growth rate [Eqs. (11.68 through 11.71)] and the decay rate [Eq. (11.41) and [Table 11.1](#)] for the expected operating conditions and for the design load and permit conditions. Note that the design load includes a peaking factor ([Table 8.13](#)). If nitrification is required year-round, these calculations must be done for each distinct season.
- Calculate the required SRT [Eq. (11.3)].

- Multiply the calculated SRT by a safety factor of about 1.5 to obtain the design SRT. Note that when the peaking factor is accounted for, the design SRT will accommodate a loading that is at least three times the annual average load.

### True Growth Yield

The yield of nitrifiers normally includes the *Nitroso*- and *Nitro*- genera and may be approximated by Eq. (11.67) (Scheible and Heidman, 1993). The estimated true growth yield coefficient for the nitrifiers as a whole is 0.13 g VSS/g NH<sub>3</sub>-N or 0.028 g VSS/g NBOD. At least 98% of the ammonia-nitrogen is oxidized to nitrate.

The nitrifier “decay” rate is usually assumed to be the same as that of the heterotrophs. This is reasonable, because the so-called decay rate is really a predation/lysis effect.

### Nitrifier Biomass

The nitrifier population is normally only a small portion of the MLSS. If it is assumed that the heterotrophs are carbon-limited and the nitrifiers are NH<sub>3</sub>-N-limited, then the nitrifier biomass can be estimated from a simple nitrogen balance:

$$\underbrace{\frac{VX_{vn}}{Y_{on} \cdot \Theta_X}}_{\text{N consumed by nitrifiers}} = \underbrace{Q(C_{TKNo} - S_{TKN})}_{\text{total N removed}} - \underbrace{f_{nh} \cdot \frac{VX_{vh}}{\Theta_X}}_{\text{N removed by heterotrophs}} \quad (11.73)$$

where  $f_{nh}$  = the fraction of nitrogen in the heterotrophs (kg N/kg VSS)  
 $\approx 0.070$  kg N/kg VSS (Table 11.3)

$S_{TKN}$  = the soluble TKN (not ammonia) of the final effluent (kg TKN/m<sup>3</sup>)

$C_{TKNo}$  = the TKN (soluble plus particulate) of the settled wastewater (kg TKN/m<sup>3</sup>)

$Q$  = the settled sewage flow rate (m<sup>3</sup>/s or ft<sup>3</sup>/sec)

$V$  = the volume of the aeration tank (m<sup>3</sup> or ft<sup>3</sup>)

$X_{vh}$  = the concentration of heterotrophs in the aeration tank (kg VSS/m<sup>3</sup>)

$X_{vn}$  = the concentration of nitrifiers in the aeration tank (kg VSS/m<sup>3</sup>)

$Y_{on}$  = the observed yield for nitrifier growth on ammonia (kg VSS/kg NH<sub>3</sub>-N)

$\Theta_X$  = the solids’ retention time (sec)

### Organic Nitrogen Production

NPDES permits are written in terms of ammonia-nitrogen, because of its toxicity. However, a substantial portion of the effluent TKN in nitrifying facilities is soluble organic nitrogen. As a rough guide, the ratio of TKN to NH<sub>3</sub>-N in settled effluents is about 2:1 to 3:1 (Barth, Brenner, and Lewis, 1968; Beckman et al., 1972; Clarkson, Lau, and Krichen, 1980; Lawrence and Brown, 1976; Mulbarger, 1971; Prakasam et al., 1979; Stankewich, no date).

### Two-Stage Nitrification

In the two-stage nitrification process, the nitrification step is preceded by a roughing step, either activated sludge or trickling filter, which is designed to remove about one-half to three-quarters of the settled sewage BOD<sub>5</sub>. The benefits of this scheme are that the total biomass carried in the plant and the oxygen consumption are reduced. The costs are an additional clarifier and increased waste sludge production. The aeration tankage requirements of two-stage nitrification are comparable to those of nonnitrifying facilities, which suggests that most conventional plants can be readily upgraded to nitrification without major expense. Mulbarger (1971) estimates the increase in capital cost above that of a nonnitrifying facility to be about 10%.

The CBOD and TKN loads to the second stage are the expected effluent quality of the first stage. The first stage effluent BOD<sub>5</sub> is a semi-free design choice; it determines the specific uptake rate for the first

stage. The first stage effluent TKN is the TKN not incorporated into the first stage's waste activated sludge. It may be estimated by,

$$\underbrace{QS_{\text{TKN}}}_{\text{1st stage soluble effluent TKN}} = \underbrace{QC_{\text{TKNo}}}_{\text{settled sewage total TKN}} - \underbrace{f_{nh} \cdot \frac{VX_{vh}}{\Theta_x}}_{\text{TKN in 1st stage waste activated sludge}} \quad (11.74)$$

The EPA model described above can represent the nitrification kinetics of the second stage. The kinetics of CBOD removal in the second stage is not well known. The second-stage influent CBOD is a microbial product formed in the first stage, not residual settled sewage CBOD, and its removal kinetics may not be well represented by the data in [Tables 11.1](#) and [11.8](#).

## Denitrification

A variety of proprietary denitrification processes are being marketed, including A/O™ (U.S. Patent No. 4,056,465), Bardenpho™ (U.S. Patent No. 3,964,998), and BIO-DENITRO™ (U.S. Patent No. 3,977,965). The Joint Task Force (1992) lists other patents.

## Microbiology

Many aerobic heterotrophic bacteria, especially pseudomonads, can utilize nitrate and nitrite as a terminal electron acceptor. The half-cell reactions are as follows:



The oxygen reduction half-cell is as shown:



Consequently, the electronic equivalent weights of nitrite and nitrate are 1.713 and 2.857 g O<sub>2</sub>/g NO<sub>3</sub>-N, respectively. The nitrate equivalent weight is confirmed by Wuhrmann's (1968) data.

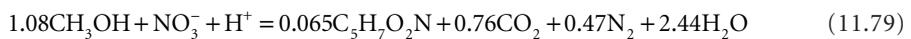
## Growth Stoichiometry

The energy available to the denitrifiers is sharply reduced, from about 1 ATP per electron pair for oxygen-based oxidations to about 0.4 ATP per electron pair (Sykes, 1975). This substantially reduces waste activated sludge production.

Smakel (1977) gives the theoretical growth stoichiometry for the anoxic growth of heterotrophic bacteria growing on methanol and nitrate plus nutrient salts as the following:



This closely approximates the empirical formula given by McCarty, Beck, and St. Amant (no date):



The calculated true growth yields for organic matter and nitrate nitrogen are 0.172 g VSS per g COD and 0.684 g VSS per g NO<sub>3</sub>-N.

## Growth Kinetics

The denitrifier growth rate can be limited by the nitrate concentration or the COD concentration, and values of the Gram model kinetic parameters have been reported for both cases. Laboratory results for

**TABLE 11.9** Typical Parameter Values for Laboratory-Scale Denitrifying Activated Sludge Processes Using Methanol at Approximately 20°C

Parameter	Symbol	Units	Typical
True growth yield	$Y_{\text{COD}}$	kg VSS/kg COD	0.17
	$Y_{\text{NO}_3}$	kg VSS/kg NO <sub>3</sub> -N	0.68
Decay rate	$k_d$	Per day	0.04
Maximum specific growth rate	$\mu_{\text{max}}$	Per day	0.3
Maximum uptake rate	$q_{\text{max}}$	kg COD/kg VSS·d	2
		kg NO <sub>3</sub> -N/kg VSS·d	0.5
Affinity constant	$K_s$	mg BOD <sub>5</sub> /L	150
		mg COD <sub>total</sub> /L	75
		mg COD <sub>biodeg</sub> /L	10
		mg NO <sub>3</sub> -N/L	0.08

Sources: Johnson, W.K. 1972. "Process Kinetics for Denitrification," *Journal of the Sanitary Engineering Division, Proc. ASCE*, 98(SA4): 623.

McClintock, S.A., Sherrard, J.H., Novak, J.T., and Randall, C.W. 1988. "Nitrate Versus Oxygen Respiration in the Activated Sludge Process," *Journal of the Water Pollution Control Federation*, 60(3): 342.

Moore, S.F. and Schroeder, E.D., 1970. "An Investigation of the Effects of Residence Time on Anaerobic Bacterial Denitrification," *Water Research*, 4(10): 685.

Moore, S.F. and Schroeder, E.D. 1971. "The Effect of Nitrate Feed Rate on Denitrification," *Water Research*, 5(7): 445.

Stensel, H.D., Loehr, R.C., and Lawrence, A.W. 1973. "Biological Kinetics of Suspended-Growth Denitrification," *Journal of the Water Pollution Control Federation*, 45(2): 249.

some of these parameters are summarized in Table 11.9. Scheible and Heidman (1993) give additional data. While the values for the true growth yields, the microbial decay, and perhaps the affinity coefficient for nitrate are reliable, the maximum specific growth rate and uptake rate are questionable. Large-scale pilot studies have produced maximum specific uptake rates that range from about 0.1 to 0.4 g NO<sub>3</sub>-N per g MLVSS d at 20°C (Ekama and Marais, 1984; Parker et al., 1975).

A commonly used kinetic model for the rate of denitrification is as follows (Metcalf & Eddy, Inc., 2002):

$$r_{\text{NO}_3} = f_{dn} X_{va} \left( \frac{1 - 1.42 Y_{hox}}{2.86} \right) \left( \frac{q_{\text{max}} S_s}{K_s + S_s} \right) \left( \frac{S_{\text{NO}_3}}{K_{\text{NO}_3} + S_{\text{NO}_3}} \right) \left( \frac{K_{\text{O}_2}}{K_{\text{O}_2} + S_{\text{O}_2}} \right) \dots + \frac{1.42 f_{nd} k_d X_{va}}{2.86} \quad (11.80)$$

where  $f_{dn}$  = the fraction of the active heterotrophic biomass that can denitrify (dimensionless)

$K_{\text{NO}_3}$  = the affinity constant for nitrate-limited denitrification (kg NO<sub>3</sub>-N/m<sup>3</sup>)

$K_{\text{O}_2}$  = the oxygen inhibition constant for denitrification (kg O<sub>2</sub>/m<sup>3</sup>)

$K_s$  = the affinity constant for substrate-limited denitrification (kg COD/m<sup>3</sup>)

$k_d$  = the decay rate of the active biomass (per s)

$r_{\text{NO}_3}$  = the volumetric nitrate-nitrogen consumption rate (kg NO<sub>3</sub>-N/m<sup>3</sup> s)

$S_{\text{NO}_3}$  = the nitrate-nitrogen concentration (kg N/m<sup>3</sup>)

$S_{\text{O}_2}$  = the aeration tank oxygen concentration (kg O<sub>2</sub>/m<sup>3</sup>)

$S_s$  = the soluble rapidly biodegradable COD concentration in the aeration tank (kg COD/m<sup>3</sup>)

$Y_{hox}$  = the aerobic heterotrophic true growth yield (kg VSS/kg COD)

$X_{va}$  = the active heterotrophic biomass (kg VSS/m<sup>3</sup>)

1.42 = the oxygen equivalent of the biomass under nonnitrifying conditions (kg O<sub>2</sub>/kg VSS)

2.86 = the oxygen equivalent of nitrate-nitrogen (kg O<sub>2</sub>/kg N)

This formulation assumes that the MLVSS are partitioned as suggested by McKinney or the IWA's Activated Sludge Model. It includes the active biomass' growth and decay. The formulation of the rate as a product on Monod-like terms also assumes simultaneous kinetic limitation by the electron acceptors and the carbon substrate. This latter assumption is unproven empirically, but it may be qualitatively correct.

### Oxygen

The threshold for oxygen inhibition of denitrification is about 0.1 mg/L, and the denitrification rate is reduced by 50% at oxygen concentrations around 0.2 to 0.3 mg/L (Focht and Chang, 1975). At 2.0 mg/L of oxygen, the denitrification rate is reduced by 90%.

Even conventional activated sludge processes denitrify if nitrate or nitrite is present, losing as much as 40% of the applied TKN (Johnson, 1959; Van Huyssteen, Barnard, and Hendrikz, 1990). This is generally believed to be due to microscale anoxic zones in the mixed liquor or inside the sludge flocs.

### Inhibitors

The reduction of nitrite is inhibited by nitrite. In unacclimated cultures, the nitrite reduction rate falls by about 80% when the nitrite-nitrogen concentration exceeds about 8 mg/L (Beccari et al., 1983). This may be due to the accumulation of free nitrous acid, because there is an unusually sharp fall in the rate of nitrite reduction as the pH falls below 7.5.

Other reported inhibitors are as follows (Painter, 1970):

- Metal chelating agents (e.g., sodium diethyldithiocarbamate, orthophenanthroline, potassium cyanide, and 4-methyl-1:2-dimercaptobenzene)
- Cytochrome inhibitors (e.g., 2-*n*-heptyl-4-hydroxyquinoline-N-oxide)
- *P*-chloromercuribenzoate
- Hydrazine
- Chlorate
- Copper

Tables 11.4 and 11.5 summarize some quantitative data on inhibition.

### Temperature

Focht and Chang (1975) reviewed  $Q_{10}$  data for a variety of nitrification processes (including mixed and pure cultures and suspended and film systems), and Lewandowski (1982) reviewed theta values for wastewater. The average theta for all these processes is about 1.095, and the type of reductant does not appear to affect the value.

### pH

The optimum pH for the reduction of nitrate is about 7 to 7.5 (Beccari et al., 1983; Focht and Chang, 1975; Parker et al., 1975). The rate of nitrate reduction falls sharply outside that range to about half its maximum value at pHs of 6.0 and 8.0.

The optimum range of pH for nitrite reduction appears to be narrowly centered at 7.5. At pH 7.0, it is only 20% of its maximum value, and at pH 8.0, it is about 70% of its maximum (Beccari et al., 1983).

## Semi-Aerobic Denitrification

The simplest denitrification facility is the "semi-aerobic" process developed by Ludzack and Ettinger (1962). This process is suitable for moderate removals of nitrate.

A schematic of the process train is shown in Fig. 11.5. In this process, mixed liquor from the effluent end of a nitrifying aeration tank is recirculated to an anoxic tank ahead of the aeration tank, where it is mixed with settled sewage. The anoxic tank is mixed but not aerated. The heterotrophs of the activated sludge utilize the nitrates in the mixed liquor to oxidize the CBOD of the settled sewage, and the nitrates are reduced to nitrogen gas.

The nitrate removal efficiency may be approximated by the following:

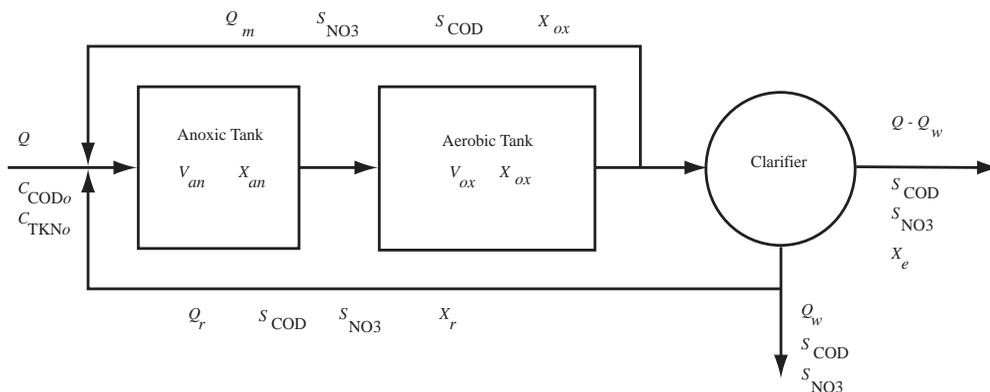


FIGURE 11.5 Ludzack-Ettinger (1962) semi-aerobic process.

$$E_{\text{NO}_3} = \frac{Q_m + Q_r}{Q + Q_m + Q_r} \quad (11.81)$$

where  $E_{\text{NO}_3}$  = the nitrate removal efficiency (dimensionless)

$Q$  = the settled sewage flow ( $\text{m}^3/\text{s}$  or  $\text{ft}^3/\text{sec}$ )

$Q_m$  = the mixed liquor return flow ( $\text{m}^3/\text{s}$  or  $\text{ft}^3/\text{sec}$ )

$Q_r$  = the return sludge flow ( $\text{m}^3/\text{s}$  or  $\text{ft}^3/\text{sec}$ )

Because of economic limitations on the amount of return flows, the practical removal efficiency is limited to maybe 80%. If high removal efficiencies are needed, a separate stage denitrification facility must be built.

It is assumed that the CBOD of the settled sewage is sufficient to reduce the nitrate produced in the aerobic tank, and this is usually the case. However, it may be necessary to add additional reductant in some cases.

It is important to distinguish the anoxic SRT from the aerobic SRT, because nitrification occurs only under aerobic conditions. These may be estimated as follows:

$$\Theta_{X_s} = \Theta_{X_{an}} + \Theta_{X_{ox}} = \frac{V_{an}X_{an} + V_{ox}X_{ox}}{Q_w X_w + (Q - Q_w)X_e} \quad (11.82)$$

$$\Theta_{X_{an}} = \frac{V_{an}X_{an}}{Q_w X_w + (Q - Q_w)X_e} \quad (11.83)$$

$$\Theta_{X_{ox}} = \frac{V_{ox}X_{ox}}{Q_w X_w + (Q - Q_w)X_e} \quad (11.84)$$

where  $Q$  = the settled sewage flow ( $\text{m}^3/\text{s}$ )

$Q_m$  = the mixed liquor return flow ( $\text{m}^3/\text{s}$ )

$Q_r$  = the return sludge flow ( $\text{m}^3/\text{s}$ )

$X_{an}$  = the volatile suspended solids' concentration in the anoxic tank ( $\text{kg}/\text{m}^3$ )

$X_e$  = the volatile suspended solids' concentration in the settled effluent ( $\text{kg}/\text{m}^3$ )

$X_{ox}$  = the volatile suspended solids' concentration in the aerobic (oxic) tank ( $\text{kg}/\text{m}^3$ )

$X_w$  = the volatile suspended solids' concentration in the waste activated sludge ( $\text{kg}/\text{m}^3$ )

$V_{an}$  = the volume of the anoxic compartment ( $\text{m}^3$ )

$V_{ox}$  = the volume of the aerobic (oxic) compartment ( $\text{m}^3$ )

$\Theta_{X_s}$  = the system solids' retention time, SSRT (sec)

$\Theta_{X,an}$  = the anoxic SRT (sec)  
 $\Theta_{X,ox}$  = the aerobic (oxic) SRT (sec)

The MLVSS concentrations in the anoxic and aerobic compartments are nearly equal.

The presence of an anoxic zone ahead of the nitrification zone has no effect on the rate of nitrification (Jones and Sabra, 1980; Sutton, Jank, and Vachon, 1980). Consequently, the aerobic SRT may be chosen using the procedures for single-stage nitrification discussed above.

The anoxic HRT is normally 1 to 4 hr, and the anoxic SRT is generally about 30% of the system SRT. The maximum specific nitrate uptake rates observed in field units are about one-fourth to one-half of the rate quoted in Table 11.9. Also,  $q_{\max \text{ NO}_3}$  declines as the anoxic and system SRT increases (Jones and Sabra, 1980).

Sutton et al. (1978) indicate that removal efficiency for filterable (soluble) TKN in a semiaerobic process can be estimated by the following:

**At 24 to 26°C:**

$$E_{\text{TKN}_{\text{filt}}} = \frac{0.98\Theta_{X,ox}}{0.26 + \Theta_{X,ox}} \quad (11.85)$$

**At 14 to 16°C:**

$$E_{\text{TKN}_{\text{filt}}} = \frac{1.05\Theta_{X,ox}}{1.00 + \Theta_{X,ox}} \quad (11.86)$$

**At 7 to 8°C:**

$$E_{\text{TKN}_{\text{filt}}} = \frac{1.11\Theta_{X,ox}}{5.04 + \Theta_{X,ox}} \quad (11.87)$$

where  $E_{\text{TKN}_{\text{filt}}}$  = the removal efficiency for filterable total kjeldahl nitrogen (dimensionless).

These results are supported by the data of Sutton, Jank, and Vachon (1980) and by the data of Jones and Sabra (1980).

Burdick, Refling, and Stensel (1982) provide kinetic data for the removal of nitrate from municipal wastewater in the anoxic zone at 20°C:

$$q_{\text{NO}_3\text{-N}} = 0.03F_{an} + 0.029 \quad (11.88)$$

$$q_{\text{NO}_3\text{-N}} = 0.12 \left( \frac{1}{\Theta_{Xs}} \right)^{0.706} \quad (11.89)$$

where  $q_{\text{NO}_3\text{-N}}$  = the anoxic zone specific nitrate consumption rate (kg NO<sub>3</sub>-N/kg MLTSS·d)  
 $F_{an}$  = the anoxic zone food-to-microorganism ratio (kg BOD<sub>5</sub>/kg MLTSS·d).

Typical design values for the A/O™ and BARDENPHO™ processes are given in Table 11.10. Recent BARDENPHO™ designs are tending toward the lower limits tabulated.

A nitrogen/COD balance on the system and anoxic tank produces:

$$\begin{aligned} Q(C_{\text{TKNo}} - S_{\text{TKN}}) &= (Q + Q_m + Q_r)(S_{\text{NO}_3} - S_{\text{NO}_3an}) + \dots \\ &\dots + f_{\text{Nh}} \cdot \{ Y_{\text{oCODox}} \cdot (C_{\text{CODo}} - S_s) + \dots \\ &\dots + Y_{\text{oNO}_3an} \cdot \left( 1 - \frac{Y_{\text{oCODox}}}{Y_{\text{oCODan}}} \right) \cdot \left[ (Q_m + Q_r)S_{\text{NO}_3} - (Q + Q_m + Q_r)S_{\text{NO}_3an} \right] \} \end{aligned} \quad (11.90)$$

**TABLE 11.10** Typical Design Values for Commercial Semiaerobic Processes

Design Parameter	A/O™ Value	BARDENPHO™ Value
Anaerobic HRT (h)	0.5–1.0	0.6–2.0
First anoxic HRT (h)	0.5–1.0	2.2–5.2
Oxic HRT (h)	3.5–6.0	6.6–19.0
Second anoxic HRT (h)	—	2.2–5.7
Reaeration HRT (h)	—	0.5–2.0
Oxic SRT (d)	—	>10
F/M (kg BOD/kg MLVSS·d)	0.15–0.25	—
MLVSS (mg/L)	3000–5000	3000
Return sludge flow (% settled sewage flow)	20–50	—
Mixed liquor recycle flow (% settled sewage flow)	100–300	400

Sources: Barnard, J.L. 1974a. "Cut P and N Without Chemicals," *Water & Wastes Engineering*, 11(7): 33.

Barnard, J.L. 1974b. "Cut P and N Without Chemicals," *Water & Wastes Engineering*, 11(8): 41.

Roy F. Weston, Inc. 1983. *Emerging Technology Assessment of Biological Phosphorus Removal: 1. PHOSTRIP PROCESS; 2. A/O PROCESS; 3. BARDENPHO PROCESS*. Environmental Protection Agency, Wastewater Research Division, Municipal Environmental Research Laboratory, Cincinnati, OH.

Weichers, H.N.S. et al., eds. 1984. *Theory, Design and Operation of Nutrient Removal Activated Sludge Processes*, Water Research Commission, Pretoria, South Africa.

where  $C_{\text{COD}_0}$  = the soluble plus particulate COD of the settled sewage (kg COD/m<sup>3</sup>)

$C_{\text{TKN}_0}$  = the soluble plus particulate TKN in the settled wastewater (kg N/m<sup>3</sup>)

$f_{\text{Nh}}$  = the weight fraction of nitrogen in the heterotrophic biomass (kg N/kg VSS)

$S_{\text{NO}_3}$  = the final effluent nitrate-nitrogen concentration (kg N/m<sup>3</sup>)

$S_{\text{NO}_3\text{an}}$  = the nitrate-nitrogen concentration in the effluent of the anoxic tank (kg N/m<sup>3</sup>)

$S_s$  = the soluble effluent COD (kg COD/m<sup>3</sup>)

$S_{\text{TKN}}$  = the soluble TKN in the final effluent (kg N/m<sup>3</sup>)

$Q$  = the settled sewage flow rate (m<sup>3</sup>/s)

$Q_m$  = the mixed liquor recirculation rate (m<sup>3</sup>/s)

$Q_r$  = the return sludge flow rate (m<sup>3</sup>/s)

$Y_{\text{oCODan}}$  = the observed heterotrophic yield from COD consumption under anoxic (denitrifying) conditions (kg VSS/kg COD)

$Y_{\text{oCODox}}$  = the observed heterotrophic yield from COD consumption under aerobic conditions (kg VSS/kg COD)

$Y_{\text{oNO}_3\text{an}}$  = the observed heterotrophic yield from nitrate-nitrogen consumption under anoxic (denitrifying) conditions (kg VSS/kg NO<sub>3</sub>-N)

This ignores the nitrifying biomass on the grounds that it is small.

The observed yields can be estimated by,

$$Y_{\text{oCODan}} = \frac{Y_{\text{CODan}}}{1 + k_d \Theta_{xan}} \approx \frac{0.17}{1 + k_d \Theta_{xan}} \quad (11.91)$$

$$Y_{\text{oCODox}} = \frac{Y_{\text{CODox}}}{1 + k_d \Theta_{xox}} \approx \frac{0.40}{1 + k_d \Theta_{xox}} \quad (11.92)$$

$$Y_{oNO_3an} = \frac{Y_{NO_3an}}{1 + k_d \Theta_{Xan}} \approx \frac{0.68}{1 + k_d \Theta_{Xan}} \quad (11.93)$$

In Eq. (11.90), the settled sewage flow rate, the influent TKN, and the target effluent nitrate-nitrogen concentrations are known. The effluent soluble TKN can be estimated as in the designs for single-stage nitrification; it is two or three times the effluent ammonia-nitrogen concentration. The effluent ammonia-nitrogen concentration and the effluent soluble BOD<sub>5</sub> and COD are fixed by the aerobic SRT. The nitrate-nitrogen concentration in the effluent of the anoxic stage is often small, and in that case, it may be neglected. The return sludge flow rate may be estimated using the Benefield–Randall formula [Eq. (11.37)]. The observed anoxic and aerobic heterotrophic yields must be calculated using the anoxic and aerobic SRT, respectively.

The unknown in Eq. (11.90) is the required mixed liquor recirculation rate. Generally, recirculation rates of 100 to 400% of the settled sewage flow are employed. Higher rates produce more complete nitrate removal, but very high rates are uneconomical. If nearly complete nitrogen removal is required, then either a two- or three-sludge system fed external reductant like methanol should be constructed. These are discussed below.

Once the flow rates are known, simple mass balances around the system can be used to calculate the various mass conversions:

**Anoxic consumption of nitrate and production of nitrogen gas:**

$$R_{NO_3an} = R_{N_2an} = (Q_m + Q_r)S_{NO_3} - (Q + Q_m + Q_r)S_{NO_3an} \quad (11.94)$$

**Anoxic consumption of COD:**

$$R_{CODan} = QC_{CODo} + (Q_m + Q_r)S_{COD} - (Q + Q_m + Q_r)S_{CODan} \quad (11.95)$$

$$R_{CODan} = \frac{Y_{oNO_3an}}{Y_{oCODan}} \cdot R_{NO_3an} = 3.98 \cdot R_{NO_3an} \quad (11.96)$$

**Aerobic consumption of COD:**

$$R_{CODox} = Q(C_{CODo} - S_{COD}) - R_{CODan} \quad (11.97)$$

where  $C_{CODo}$  = the total COD in the settled wastewater (kg/m<sup>3</sup>)  
 $R_{CODan}$  = the rate of COD consumption in the anoxic tank (kg/s)  
 $R_{CODox}$  = the rate of COD consumption in the aerobic tank (kg/s)  
 $R_{N_2an}$  = the rate of nitrogen gas production in the anoxic tank (kg/s)  
 $R_{NO_3an}$  = the rate of nitrate-nitrogen consumption in the anoxic tank (kg/s)  
 $S_{COD}$  = the soluble COD in the final effluent  
 $S_{CODan}$  = the soluble COD in the effluent of the anoxic tank (kg/m<sup>3</sup>)  
 $S_{NO_3}$  = the concentration of nitrate-nitrogen in the settled effluent (kg/m<sup>3</sup>)  
 $S_{NO_3an}$  = the concentration of nitrate-nitrogen in the effluent of the anoxic tank (kg/m<sup>3</sup>)  
 $Y_{oCODan}$  = the observed yield of heterotrophs from COD in the anoxic tank (kg VSS/kg COD)  
 $Y_{oNO_3an}$  = the observed yield of heterotrophs from nitrate-nitrogen in the anoxic tank (kg VSS/kg N)

Note that the effluent COD is fixed by the aerobic SRT. The COD in the effluent of the anoxic tank can be calculated using Eq. (11.95). If it is negative, supplemental reductant, usually methanol, is required.

The waste sludge production rate is given by,

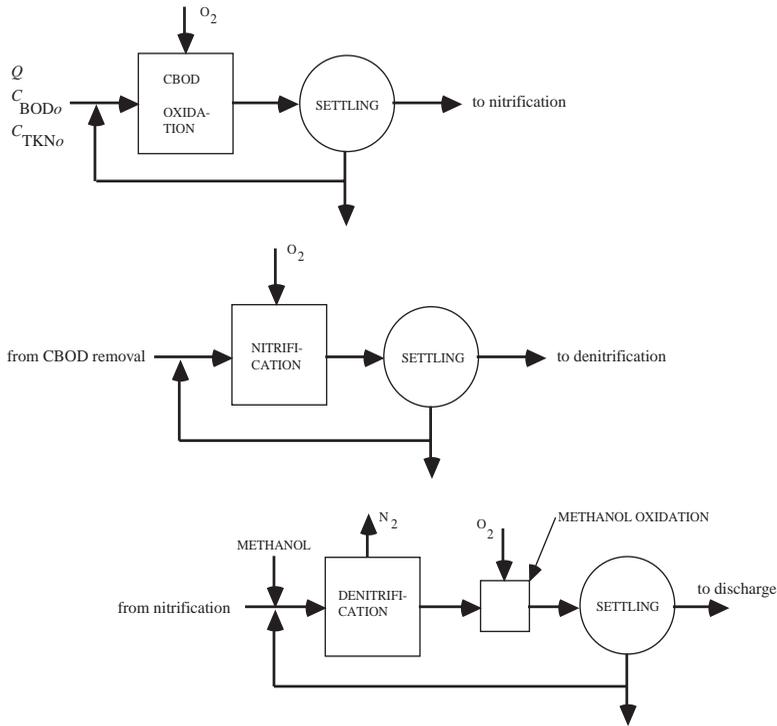


FIGURE 11.6 The three-sludge nitrogen removal scheme (Mulbarger, 1971).

$$Q_w X_w \leq \frac{V_{an} X_{an}}{\Theta_{Xan}} + \frac{V_{ox} X_{ox}}{\Theta_{Xox}} = Y_{oCODox} \cdot R_{CODox} + Y_{oCODan} \cdot R_{CODan} \quad (11.98)$$

The aerobic SRT is fixed by the specified effluent ammonia-nitrogen concentration, and the anoxic SRT is set to minimize the nitrate-nitrogen concentration in the anoxic effluent. The MLSS is a semifree choice, so the only unknown is the total tankage.

## Two- and Three-Stage Denitrification

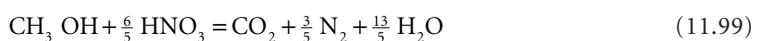
If low effluent nitrate concentrations are needed, a separate stage denitrification process fed supplementary reductant is required. A schematic process train for separate stage denitrification is shown in Fig. 11.6, which is Mulbarger's (1971) three-sludge process. This scheme was modified and built by the Central Contra Costa Sanitary District and Brown and Caldwell, Engineers (Horstkotte et al., 1974). The principal changes are (1) the use of lime in the primary clarifier to remove metals and (2) the substitution of a single-stage nitrification process for the two-stage process used by Mulbarger.

The nitrification stage can be designed using the procedures described above.

The denitrification stage can be designed using the methanol and nitrate kinetic data in Table 11.11.

Any nonfermentable organic substance can be used as the reductant for nitrate. The usual choice is methanol, because it is the cheapest. Fermentable substances like sugar or molasses are not used, because a substantial portion of the reductant can be dissipated as gaseous end-products like hydrogen.

Disregarding microbial growth, the stoichiometry of methanol oxidation is as follows:



**TABLE 11.11** Separate Stage Denitrification Design Values <sup>1-3</sup>

Design Variable	Reference 1	Reference 2	Reference 3
Anoxic HRT (h)	3	1.6	0.82
Oxic HRT (h)	0.4	0.3	0.79
MLVSS (mg/L)	1390	2460	2600
F/M (kg COD/kg VSS.d)	0.73		0.79
Anoxic SRT (d)	38	7.6	29
Methanol/nitrate ratio (kg methanol/kg NO <sub>3</sub> -N)	5.83	3.07	3.54
Effluent NO <sub>3</sub> -N (mg/L)	0.9	0.7	0.8
System nitrogen removal (%)	91	>69	—
Temperature (°C)	12–22	10	16–19

*Sources:*

<sup>1</sup> Barth, E.F., Brenner, R.C., and Lewis, R.F. 1968. "Chemical-Biological Control of Nitrogen and Phosphorus in Wastewater Effluent," *Journal of the Water Pollution Control Federation*, 40(12): 2040.

<sup>2</sup> Mulbarger, M.C. 1971. "Nitrification and Denitrification in Activated Sludge Systems," *Journal of the Water Pollution Control Federation*, 43(10): 2059.

<sup>3</sup> Horstkotte, G.A., Niles, D.G., Parker, D.S., and Caldwell, D.H. 1974. "Full-Scale Testing of a Water Reclamation System," *Journal of the Water Pollution Control Federation*, 46(1): 181.

This suggests a methanol dosage of at least 1.9 g methanol per g nitrate-nitrogen. In order to account for growth and the presence of oxygen and nitrite in the feed, McCarty, Beck, and St. Amant (no date) recommend a dosage of the following:

$$S_{meth} = 2.47S_{NO_3o} + 1.53S_{NO_2o} + 0.87S_{O_2o} \quad (11.100)$$

where  $S_{meth}$  = the required methanol dosage (mg CH<sub>3</sub>OH/L)

$S_{NO_3o}$  = the initial nitrate-nitrogen concentration (mg NO<sub>3</sub>-N/L)

$S_{NO_2o}$  = the initial nitrite-nitrogen concentration (mg NO<sub>2</sub>-N/L)

$S_{O_2o}$  = the initial dissolved oxygen concentration (mg O<sub>2</sub>/L)

In general, excess methanol is supplied to ensure that nearly complete nitrate removal is achieved. This means that the nitrate-limited kinetic parameters control. The excess methanol must be removed prior to discharge. This is accomplished by a short-detention aerobic stage inserted between the anoxic stage and the clarifier. The aerobic stage also serves as a nitrogen gas-stripping unit.

The sludge yield from the denitrification plant can be estimated via the formula suggested by McCarty, Beck, and St. Amant (no date):

$$X_v = 0.53S_{NO_3o} + 0.32S_{NO_2o} + 0.19S_{O_2o} \quad (11.101)$$

where  $X_v$  = the active heterotrophic biomass produced (mg VSS/L).

## Biological Phosphorus Removal

The biochemical phosphorus removal process is called *polyphosphat überKompensation* or "polyphosphate overplus" (Harold, 1966). This is not to be confused with *luxury uptake*, which occurs when growth is inhibited by lack of an essential nutrient. Except for in the PHOSTRIP™ process, phosphorus is removed from wastewater by incorporation into the biomass and subsequent biomass wastage from the system. The steady state system phosphorus balance is as follows:

$$C_{p0} = S_p - f_{ph} \frac{VX_v}{Q\Theta_x} \quad (11.102)$$

where  $C_{p_0}$  = the total influent phosphorus concentration (kg/m<sup>3</sup>)  
 $f_{ph}$  = the concentration of phosphorus in the volatile suspended solids (kg P/kg VSS)  
 $Q$  = the settled sewage flow rate (m<sup>3</sup>/s)  
 $S_p$  = the soluble effluent phosphorus concentration (kg/m<sup>3</sup>)  
 $V$  = the aeration tank volume (m<sup>3</sup>)  
 $X_v$  = the mixed liquor volatile suspended solids' concentration (kg VSS/m<sup>3</sup>)  
 $\Theta_x$  = the solids' retention time (s)

Growing bacteria typically contain about 3% by wt (dry) phosphorus, of which 65% is incorporated into deoxyribonucleic acid (DNA) and ribonucleic acid (RNA), 15% into phospholipids, and the remainder as acid-soluble phosphate esters (Roberts et al., 1955). Nearly all algae, bacteria, and fungi can accumulate more than that (Harold, 1966), and the bacterium *Acinetobacter* has been reported to contain as much as 16% phosphorus (Eagle et al., 1989). The increased phosphorus content consists of volutin granules. Volutin is a long-chain polymer of orthophosphate ions. The polymer is highly charged and is associated with K<sup>+</sup>, Ca<sup>2+</sup>, and Mg<sup>2+</sup> ions, RNA, polybetahydroxybutyric acid (PHB), proteins, and phospholipids (Eagle et al., 1989).

Conventional activated sludge plants typically remove about 40% of the influent phosphorus (EPA, 1977). The phosphorus content of activated sludges with volutin-containing bacteria generally consists of 4 to 7% by wt of the total suspended solids, and the plants generally achieve better than 90% phosphorus removal (Scalf et al., 1969; Vacker, Connell and Wells, 1967).

A number of proprietary phosphorus removal processes are being marketed that induce volutin accumulation in bacteria, including A<sup>2</sup>/O<sup>TM</sup> (U.S. Patent No. 4,056,465), BARDENPHO<sup>TM</sup> (U.S. Patent No. 3,964,998), PHOSTRIP<sup>TM</sup> and PHOSTRIP II<sup>TM</sup> (U.S. Patent No. 4,042,493; 4,141,822; 4,183,808; and 4,956,094) and VIP<sup>TM</sup> (U.S. Patent 4,867,883). The Joint Task Force (1992) lists several other patents and processes.

## Microbiology

The underlying mechanism of volutin accumulation is the depression of the synthesis of three enzymes (Harold, 1966): alkaline phosphatase (which hydrolyzes extracellular phosphate esters), polyphosphate kinase (which synthesizes polyphosphate chains by transferring orthophosphate from adenosine triphosphate), and polyphosphatase (which hydrolyzes polyphosphate chains). Bacteria that are subjected to repeated anaerobic/aerobic cycles might have three to five times the normal amount of these enzymes..

Volutin formation occurs in activated sludge when it is exposed to an anaerobic/aerobic cycle (Wells, 1969). The currently accepted biochemical model for polyphosphate overplus is (Bowker and Stensel, 1987; Buchan, 1983; Fuhs and Chen, 1975; Marais, Loewenthal and Siebritz, 1983; Yall and Sinclair, 1971):

- During the anaerobic phase, volutin is hydrolyzed to orthophosphate, and the hydrolysis energy is used to absorb short-chain, volatile fatty acids (VFA, principally acetic acid) that are polymerized into polybetahydroxybutyric acid (PHB) and related substances like polybetahydroxyvaleric acid (PHV) (Bowker and Stensel, 1987). Orthophosphate is released to the surrounding medium, and cellular phosphorus levels fall to less than one-half normal (Smith, Wilkinson, and Duguid, 1954).
- Under subsequent aerobic conditions, the PHB (and PHV) are oxidized for energy and used for cell synthesis. The volutin is resynthesized by using some of the oxidation energy to reabsorb and polymerize the released orthophosphate.

Consequently, there is an alternation between volutin-rich aerobic and PHB-rich anaerobic states. Peak volutin levels are reached about 1 hr after re-exposure to oxygen or nitrate. Under prolonged aeration, the volutin is broken down and incorporated into DNA, RNA, and phospholipids.

The anaerobic phase shuts down the metabolism of most aerobic bacteria, except *Acinetobacter* spp. This permits *Acinetobacter* species to absorb volatile fatty acids without competition from other aerobic bacteria. The result is an activated sludge enriched in volutin-accumulating *Acinetobacter* bacteria. A prolonged anaerobic phase may encourage the growth of anaerobic and facultatively anaerobic bacteria that can ferment sewage organic matter to VFA and other small molecules for use by the *Acinetobacter*. However, if the raw wastewater lacks the required VFAs, acetate should be added from another source.

## Kinetics

The kinetics of phosphorus uptake and release are poorly established in the public literature. In the early laboratory studies, it was reported that the release and subsequent uptake of orthophosphate each require about 3 hr for completion (Fuhs and Chen, 1975; Shapiro, 1967). The anaerobic phase in A/O™ systems generally has a hydraulic retention time of 30 to 90 min, BARDENPHO™ systems generally incorporate an anaerobic HRT of 1 to 2 hr, and PHOSTRIP™ plants operate with an anaerobic HRT of 5 to 20 hr (Bowker and Stensel, 1987). In PHOSTRIP™, the anaerobic phase is applied to settled sludge solids in a sidestream operation. It is nowadays recommended that the anaerobic phase HRT not exceed 3 hr (to avoid phosphorus releases not associated with VFA uptake), and that the anaerobic SRT be about 1 day (Grady, Daigger, and Lim, 1999).

BPR removal plants utilize system SRTs of 2 to 4 days (Grady, Daigger, and Lim, 1999). BPR plants remove phosphorus by incorporating it into active biomass, and longer SRTs result in lower active biomass concentrations and cells depleted in reserves of all kinds, which reduce overall phosphorus removal.

Because anaerobic conditions induce phosphorus release, it is important to waste sludge solids as soon after aerobic growth as possible and before the sludge enters any anaerobic environment.

## Soluble Organic Matter Requirement

If effluent phosphorus concentrations less than 1 mg/L are desired, the settled wastewater should have a soluble BOD<sub>5</sub> to soluble P ratio of at least 15 or a total BOD<sub>5</sub> to total P ratio of at least 35 (Bowker and Stensel, 1987). Phosphorus-accumulating organisms preferentially consume acetate, and if the wastewater lacks acetate, consistent P removal will not occur.

## Effects of Oxygen and Nitrate

Oxygen and nitrate (in nitrifying plants) are carried into the anaerobic zone by sludge and mixed liquor recycles. Airlift pumps, which are sometimes used for sludge recycle, will saturate water with oxygen. This permits some growth of aerobic bacteria in the anaerobic zone and reduces the competitive advantage of *Acinetobacter* spp. The net effect of oxygen and nitrate is to reduce the effective BOD<sub>5</sub>:P ratio below that required for optimum phosphorus uptake. The reduction can be estimated from the initial oxygen or nitrate concentrations. The fraction of the removed COD that is actually oxidized is  $1 - 1.42Y_o$ . For rapidly growing bacteria, the observed yield is nearly equal to the true growth yield of 0.4 g VSS per g COD, so the fraction oxidized is about 43%. In the case of denitrification, the theoretical ratio is 3.53 kg COD removed for every kg of nitrate-nitrogen that is reduced (McCarty, Beck, and St. Amant, no date); reported values of the ratio range from 2.2 to 10.2 (Bowker and Stensel, 1987).

## Chemical Requirements

Other than VFA in the influent, additional chemicals are not required for BPR. However, if the discharge permit has stringent limits on effluent phosphorus or if the VFA of the influent is variable, some provision for chemical precipitation of phosphate should be made. The preferred precipitant is alum, because, unlike ferric salts, the aluminum-phosphate precipitates (sterrettite, taranakite, and variscite) are insoluble under reducing conditions (Goldshmid and Rubin, 1978).

Lack of VFAs can be offset by the addition of acetate. Some facilities obtain acetate internally by an acid-phase-only anaerobic fermentation of sludge solids (Metcalf & Eddy, Inc., 2002). Unheated digesters with SRTs of 3 to 5 days (to inhibit methanogenesis) are fed primary solids only (to minimize phosphate in the recycle). The expected VFA yield is about 0.1 to 0.2 g VFA per g VSS fed to the digesters.

## Phosphorus-Laden Recycle Streams

Sludge thickening and stabilization operations usually result in solubilization of phosphates. These streams should be pretreated with alum to remove soluble phosphate prior to admixture with the raw or settled sewage flow, and the alum-phosphate sludge should be dewatered and sent to final disposal.

## Effluent Solids

Effluent solids are also enriched in phosphorus (beyond normal activated sludge levels), and tertiary solids removal may be needed to meet permit conditions.

## Empirical Design

Weichers et al. (1984) recommend the following semiempirical formula for estimating phosphorus uptake:

$$\Delta C_{\text{PoX}} = C_{\text{CODo}} \left\{ \left[ \frac{(1 - f_{\text{CODsu}} - f_{\text{CODpu}}) Y_a}{1 + k_d \Theta_X} \right] \left( \gamma + f_{\text{Pu}} f_{X_{au}} k_d \Theta_X \right) + f_{\text{Pu}} \left( \frac{f_{\text{CODpu}}}{\beta_X} \right) \right\} \quad (11.103)$$

where  $C_{\text{CODo}}$  = the total influent COD (mg/L)

$\Delta C_{\text{PoX}}$  = the phosphorus removal from the wastewater by incorporation into the sludge (mg/L)

$f_{\text{CODpu}}$  = the unbiodegradable fraction of the particulate influent COD (dimensionless)

$\approx 0.13$  for municipal wastewater (Weichers et al., 1984)

$f_{\text{CODsu}}$  = the unbiodegradable fraction of the soluble influent COD (dimensionless)

$\approx 0.05$  for municipal wastewater (Weichers et al., 1984)

$f_{X_{an}}$  = the anaerobic mass fraction of the MLSS (dimensionless)

$f_{X_{au}}$  = the unbiodegradable fraction of the active biomass (dimensionless)

$\approx 0.20$  for municipal wastewater (Weichers et al., 1984)

$f_{\text{Pu}}$  = the phosphorus concentration in the unbiodegradable volatile solids (mg P/mg VSS)

$\approx 0.015$  mg P/mg VSS for municipal wastewater (Weichers et al., 1984)

$k_d$  = the decay coefficient of the heterotrophic bacteria, per day

$\approx 0.24/\text{day}$  (Weichers et al., 1984)

$S_{\text{CODanrb}}$  = the readily biodegradable COD in the anaerobic reactor (mg/L)

$Y_a$  = the true growth yield of the heterotrophic bacteria (mg VSS/mg COD)

$\approx 0.45$  mg VSS/mg COD (Weichers et al., 1984)

$\beta_X$  = the COD of the VSS (mg COD/mg VSS)

$\approx 1.48$  mg COD/mg VSS (Weichers et al., 1984)

$\phi$  = the excess phosphorus removal propensity factor (dimensionless)

$\approx (S_{\text{CODanrb}} - 25) f_{X_{an}}$

$\gamma$  = the fraction of phosphorus in the active biomass (mg P/mg VSS)

$= 0.35 - 0.29 \exp(-0.242\phi)$

$\Theta_X$  = the solids' retention time (days)

Eq. (11.103) distinguishes between the active biomass, which absorbs and releases phosphorus, and inert (endogenous) biomass. The phosphorus content of the active biomass is supposed to vary between about 3 and 35% by wt of the VSS. The phosphorus content of the inert biomass is estimated to be about 1.5% of VSS.

The anaerobic mass fraction is the fraction of the system biomass held in the anaerobic zone. Because MLSS concentrations do not vary substantially from one reactor to the next in a series of tanks, this is also approximately the fraction of the total system reactor volume in the anaerobic zone.

## Flow Schematics and Performance

Flow schematics of the principal phosphorus removal processes are shown in [Fig. 11.7](#).

The PHOSTRIP™ process removes phosphorus by subjecting a portion of the recycled activated sludge to an anaerobic holding period in a thickener. The phosphorus-rich supernatant is then treated with lime, and the phosphorus leaves the system as a hydroxylapatite  $[\text{Ca}_5(\text{PO}_4)_3\text{OH}]$  sludge. The phosphorus-

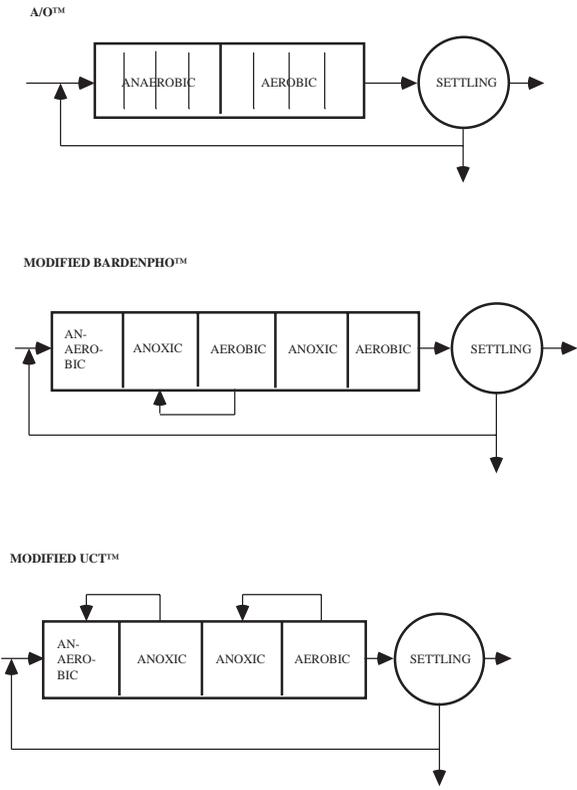


FIGURE 11.7(a) Biological phosphorus removal schemes.

depleted activated sludge solids in the thickener are returned to aeration. The PHOSTRIP™ process consistently produces soluble effluent phosphorus concentrations under 1 mg/L and usually produces total effluent phosphorus concentrations of about 1 mg/L, although excursions of effluent suspended solids sometimes degrade performance (Roy F. Weston, Inc., 1983). The process cannot be used to remove phosphorus from a nitrifying sludge, but it can be applied to the first stage of a two-stage nitrification plant. The PHOSTRIP™ process is best suited to low hardness waters so that the lime treatment step does not produce unwanted calcium carbonate sludges.

The A/O™ process nitrifies, partially denitrifies, and partially removes phosphorus. Total effluent phosphorus concentrations are generally 1.5 to 3.0 mg/L (Roy F. Weston, Inc., 1983). Phosphorus is removed from the system in the waste activated sludge.

The BARDENPHO™ process also nitrifies, partially denitrifies, and partially removes phosphorus. Supplemental methanol for denitrification and alum for phosphate precipitation may be required if consistently high removals of nitrogen and phosphorus are needed (Roy F. Weston, Inc., 1983). Phosphorus is removed from the system in the waste activated sludge.

Except for the PHOSTRIP™ process, all the schemes for biological phosphorus removal produce phosphorus-rich waste activated sludges. If these sludges are subjected to digestion, phosphorus-rich supernatants are produced, and the supernatants require a phosphorus removal treatment prior to recycle.

All the processes show degradation of total phosphorus removal when secondary clarifiers are subjected to hydraulic overloads. If consistently low total effluent phosphorus concentrations are required, effluent filters should be considered.

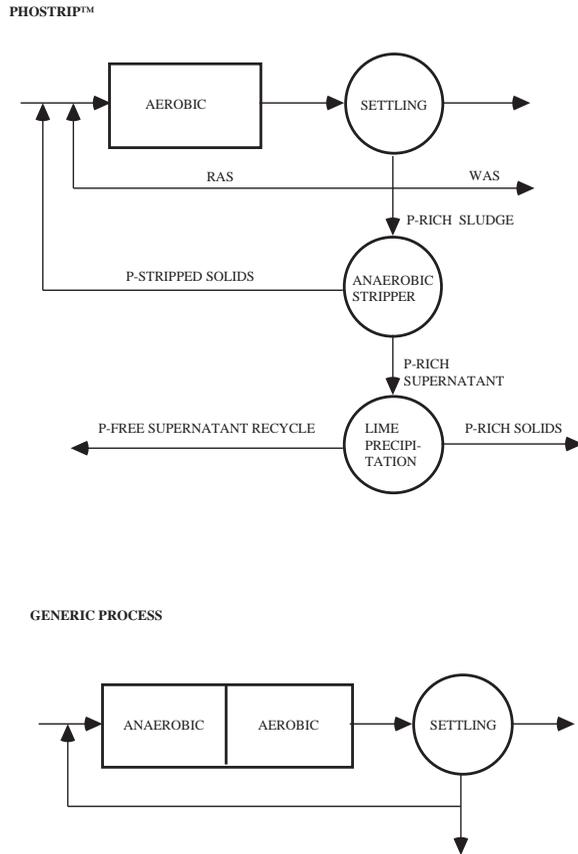


FIGURE 11.7(b) Biological phosphorus removal schemes.

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### 11.3 Aerobic Fixed-Film Processes

The fundamental problems of any percolation system are the hydraulic and pneumatic transmissibilities of the media. The initial solution to the relatively low transmissibility of natural soils was the intermittent sand filter, which was developed by Frankland in Great Britain and the Lawrence Experiment Station in

Massachusetts (Bruce and Hawkes, 1983). These filters are constructed of relatively coarse sands with controlled size grading. However, it was discovered at the Lawrence Experiment Station that high transmissibilities and good organic removals were obtainable by utilizing gravels. Latter studies increased the size of the media to crushed rock of a few inches diameter.

## Trickling Filters

Trickling filters (*bacteria beds*, *biological filters*, *percolating filters*) consist of a thin film of wastewater flowing over a packing (media) that holds an aerobic surface biofilm. Aerobic conditions are maintained by the flow of air through the packing voids. Fig. 11.8 depicts a typical arrangement. The packing is supported by filter blocks that are slotted on top to admit the treated wastewater. The slots connect to a horizontal passageway in the lower part of the block. The passageways connect to a collection channel that gathers all wastewater leaving the filter. The drainage channel is also connected to the peripheral chimneys that admit air to the packing.

The first trickling filter plant was constructed in Salford, England, in 1892; the first American trickling filter was built in Atlanta, GA, in 1903 (Peters and Alleman, 1982). The traditional trickling filter classification by hydraulic and  $BOD_5$  loading is given in Table 11.12. The Joint Task Force (1992) regards this classification scheme to be obsolete.

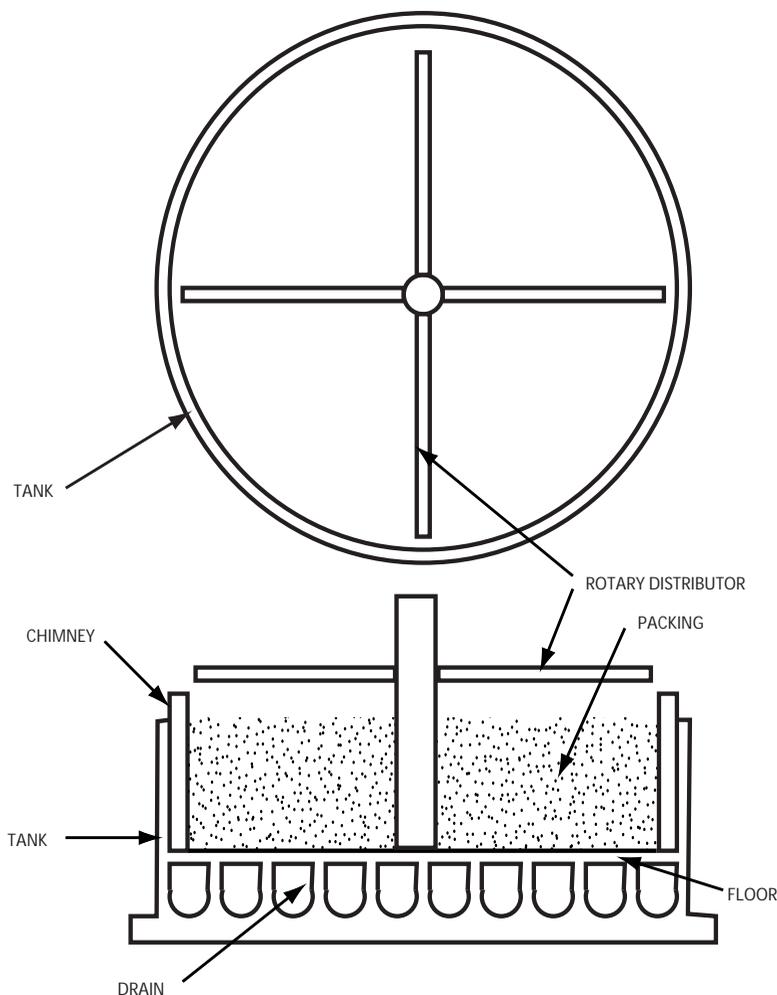


FIGURE 11.8 Trickling filter schematic.

**TABLE 11.12** Trickling Filter Classification

Class	Media	Hydraulic Loading (mgad)	BOD <sub>5</sub> Loading (lb/d/1000 ft <sup>3</sup> )	Depth (ft)	BOD <sub>5</sub> Removal <sup>a</sup> (%)	Recirculation/Nitrification
Low	Stone	1–4	5–20	5–10	75–85	No/fully
Intermediate	Stone	4–10	15–30	6–8	50–70	Yes/partial
High	Stone	10–40	30–60	3–8	40–80	Yes/none
Super	Plastic	15–90	30–150	<40	65–85	Yes/some
Roughing	Stone/plastic	60–180	<300	3–20	40–85	No/none

<sup>a</sup> After settling

Sources: Joint Task Force of the Water Environment Federation and the American Society of Civil Engineers. 1992. *Design of Municipal Wastewater Treatment Plants: Volume I. Chapters 1–12*, WEF Manual of Practice No. 8, ASCE Manual and Report on Engineering Practice No. 76. Water Environment Federation, Alexandria, VA; American Society of Civil Engineers, New York.

Schwinn, D.E. and Gasset, R.B., eds. 1974. *Process Design Manual for Upgrading Existing Wastewater Treatment Plants*. Environmental Protection Agency, Technology Transfer, Washington, DC.

Low-rate filters sometimes employ recirculation during low flow periods to keep the media wet. Low-rate and intermediate-rate filters produce nuisance flies. They also slough in the early spring when the filter fly larvae become active and disrupt the biofilm. During the sloughing, the filter effluent will contain more suspended solids than its influent. High-rate filters do not usually harbor filter flies, and they discharge solids at low rates more or less continuously. The low rate of solids discharge is not sloughing, because it is not larva-induced, seasonal, or flashy.

### Packings (Media)

Trickling filter packings (media) can be classified as *random*, *stacked*, or *modular*:

- Random media consist of relatively small individual pieces of crushed stone or ceramic or plastic shapes that are gently poured into the filter box so that the orientations of the pieces are random.
- Stacked media consist of individual pieces that are placed by hand in a specified orientation.
- Modular media consist of relatively large integrated arrays of plates or channels. These are placed by hand in fixed orientations, and individual modules may be shaped in the field to fit.

Older filters built before 1960 consist almost entirely of crushed stone placed by hand-shovel and wheelbarrow and randomly oriented. Hand placement is required to avoid damage by dumping to the under drain system and to the media. The stone is usually sieved and washed on site to remove fines and long or flat shapes. The usual size range is 3 to 4 in. (76 to 114 mm). Stone must be durable against freeze-thaw cycles and chemically and biologically inert to the wastewater and microbes. Some furnace slags and burnt clays leach iron, which discolors the effluent. The usual test of durability is a loss of less than 10% by wt after 20 cycles of soaking for 16 to 18 hr in a saturated sodium sulfate solution at 70°F followed by drying at 230°F (Anonymous, 1936; ASTM Committee C-9, 1993).

Other random packings include ceramic, carbon, metal, or plastic rings (Raschig, Lessing, cross-partition, and spiral) and saddles (Berl and Intalox). Rings are short, hollow cylinders and different kinds have different designs for partition walls within the cylinder. For example, Raschig rings have no inner partitions, Lessing rings have a single cross-wall running down the axis of the cylinder, and cross-partitioned rings have two intersecting walls running down the axis. Saddles are strips of material that have been bent into a semicircular curve; the surface of the strip may be grooved or folded.

Stacked media are almost never used, because the careful hand labor required makes them expensive to install, because they are prone to channeling.

Modular media consist of arrays of parallel sheets of plastic joined together. Early designs used corrugated sheets with the corrugation channels aligned vertically. Two kinds are in use: vertical-flow, fully corrugated bundles (VFC) and vertical-flow, semicorrugated bundles (VFS). The latter alternate

**TABLE 11.13** Physical Properties of Trickling Filter Media

Media	Unit Size (in.)	Specific Weight (lb/ft <sup>3</sup> )	Specific Surface Area (ft <sup>2</sup> /ft <sup>3</sup> )	Void Space (%)
Modular plastic	24 × 24 × 48	2–6	25–35	94–97
Random plastic	Varies	2–5	25–50	>94
Redwood	47½ × 47½ × 35¾	10	14	76
Blast furnace slag	2–3	68	29	49
Rock (granite)	1–3	90	19	46
Rock (granite)	4	70	13	60

Source: Schwinn, D.E. and Gassett, R.B., eds. 1974. *Process Design Manual for Upgrading Existing Wastewater Treatment Plants*. Environmental Protection Agency, Technology Transfer, Washington, DC.

flat sheets with corrugated sheets. In recent years, the corrugations have been arranged so that the flow is diagonal through a bundle, with alternating layers directing the flow to opposite sides. This arrangement is called cross flow (XF). VFC is preferred for strong wastewaters and filters with high organic loadings. XF is preferred for weak wastewaters and low organic loadings (Joint Task Force, 1992). Plastic media must be resistant to ultraviolet radiation, disintegration, erosion, aging, all common acids and alkalis, organic compounds, and microbial attack (Wastewater Committee, 1997). The media must not be toxic to the biofilm or leach toxic materials. The media should be able to support foot traffic for distributor maintenance. If it cannot, walkways must be provided.

An important property of any media is the specific surface area,  $a$ , which is defined to be the total area available for biofilm attachment divided by the bulk volume of the media. The surface area actually covered by biofilm and wastewater depends on the hydraulic loading, and media-specific minimum flows are required to achieve the maximum treatment capacity. Some typical values of  $a$  are given in Table 11.13.

In order to promote free airflow, the Wastewater Committee (1997) specifies maximum specific surface areas of 30 sq ft per cu ft (100 m<sup>2</sup>/m<sup>3</sup>) for carbonaceous BOD<sub>5</sub> removal and 45 sq ft per cu ft (150m<sup>2</sup>/m<sup>3</sup>) for nitrification.

Rock media filters that depend on natural draft for aeration are usually limited to a maximum media depth of 10 ft (Wastewater Committee, 1990); the usual practice is to build them no deeper than 6 ft. Greater depths can be used if forced aeration is provided. Natural draft modular plastic media filters are typically 16 to 26 ft deep (Wastewater Committee, 1997).

## Hydraulics and Pneumatics

### Hydraulic Retention Time

The current model for the hydraulic regime in trickling filters is as follows (Suschka, 1987):

- There is a biofilm layer attached to the media surface.
- On the liquid side of the biofilm, there is a heterogeneous layer of porous biofilm and pore water; the biofilm is similar in morphology to a forest of microscopic mushrooms.
- Because of intermittent dosing, wastewater waves flow freely over the mixed biofilm-pore water layer.
- The remainder of the media voids are filled with air.

Most of the liquid on the media surface is in the heterogeneous layer, rather than in the freely flowing waves. The intermittent wastewater waves periodically disrupt and mix with the pore water in the biofilm, and this gives the flow a significant complete-mix and/or dispersed-flow character. Even in steady flow over biofilm-free filters, there is some dispersion, and it increases as the hydraulic load decreases, approaching complete mixing and streaming (Muslu, 1986).

For clean modular media, without a biofilm, the thickness of the freely flowing water film can be estimated by the following (Suschka, 1987):

$$h_{fl} = \sqrt[3]{\frac{3q}{\gamma \sin \alpha}} \quad (11.104)$$

where  $h_{fl}$  = the thickness of the free liquid (m)  
 $q$  = the wastewater flow rate per unit width of media surface ( $\text{m}^3/\text{m}\cdot\text{s}$ )  
 $\alpha$  = the inclination of the media surface measured from the horizontal (rad)  
 $\gamma$  = the unit weight of the liquid ( $\text{N}/\text{m}^3$ )

However, once the biofilm and its associated heterogeneous layer develop, the effective water layer thickness is determined by experiment to be approximately (Suschka, 1987),

$$h_{el} = 3.4 \times \sqrt[3]{\frac{3q}{\gamma \sin \alpha}} \quad (11.105)$$

where  $h_{el}$  = the effective water film thickness (m).

Suschka (1987) gives an order of magnitude estimate of the HRT of trickling filters with modular media:

$$\tau = \frac{2H}{(Q/V)^{0.75}} \quad (11.106)$$

where  $H$  = the depth of media (m)  
 $Q$  = the total wastewater flow rate including recirculated flow ( $\text{m}^3/\text{h}$ )  
 $V$  = the media volume ( $\text{m}^3$ )  
 $\tau$  = the HRT (min)

Equation (11.106) appears to be most representative of media with specific surface areas of about 100 to 120  $\text{m}^2/\text{m}^3$ . The HRT of media with specific surface areas of about 40  $\text{m}^2/\text{m}^3$  is only about one-third of the HRT predicted by Eq. (11.106), and the HRT of media with specific surface areas around 200  $\text{m}^2/\text{m}^3$  is about three times the equation value.

### Distributor Systems

The preferred distributor system consists of rotary sprinkler arms driven by variable speed electric motors (Joint Task Force, 1992). The wastewater should be distributed as uniformly as possible over the filter cross section; the maximum areal application rate should not vary by more than 10% from one point to another on the cross section (Wastewater Committee, 1990).

Impulse-driven arms are not desirable, because they fail to achieve adequate wetting. Fixed nozzles require high maintenance, and uniform distribution of the wastewater is nearly impossible (Joint Task Force, 1992).

Traditional designs attempted to provide a steady flow to each area of the filter cross section, but recent studies suggest that this results in poor wetting of the media surface, with perhaps as little as one-third of the media being used (Joint Task Force, 1992). Recent recommendations focus on the instantaneous dosing or flushing intensity (Ger. *spülkraft*) on a subarea due to the passage of a distributor arm. This is normally expressed as the total depth deposited on the area per passage of an arm and is calculated as follows (Joint Task Force, 1992):

$$I = \frac{Q + Q_r}{nfA} \quad (11.107)$$

where  $A$  = the plan area of the filter ( $\text{m}^2$ )  
 $f$  = the rotational speed (frequency) of the distributor (Hz, rev/s)  
 $I$  = the instantaneous dosing or flushing intensity (m/pass)  
 $n$  = the number of distributor arms or passes per revolution (dimensionless)  
 $Q$  = the settled sewage flow rate ( $\text{m}^3/\text{s}$ )  
 $Q_r$  = the flow rate of the recirculated filter effluent ( $\text{m}^3/\text{s}$  or  $\text{ft}^3/\text{sec}$ )

Older practice results in flushing intensities on the order of 2 to 10 mm per pass, but field experience suggests it should be somewhere at least a few hundred mm per pass (Joint Task Force, 1992). The flushing intensity may be site-, media-, application-specific, or all, and the selection of the distributor motor and controller should allow maximum flexibility in the selection of rotational speeds.

The recommendation for high flushing intensities is supported by chemical engineering experience with packed beds for liquid/gas mass transfer operations (Leva, 1953). The efficiency of such units appears to increase with hydraulic loading nearly up to the point of flooding because of improved wetting. Packed towers are often operated in what is called the *loading region*, which occurs when the liquid rate is high enough that the gas pressure drop across the packing varies as a power of the gas flow rate that is greater than 2. In this region, Eqs. (11.109) and (11.112) do not apply.

## Drainage

The usual requirements are (Wastewater Committee, 1997):

- The under-drain filter blocks should collect wastewater from the entire floor of the filter.
- The inlet slots in the top of the filter blocks must cover at least 15% of the filter plan area.
- The filter block passageways and collection drains may be only half full at instantaneous peak flows (to permit air flow), and they should be designed to provide a minimum velocity of 2 ft/sec (0.61 m/s) at an invert slope of 1%.

## Aeration

The trickling filter is an aerobic process, and the oxygen needed is supplied by air flowing through the packed bed. This airflow is developed by natural convection due to density differences between the air inside and outside the bed or by blowers.

As long as the pressure drop is small, say 10% or less, air may be treated as an incompressible fluid, and the usual headloss equations apply. Piping losses may be estimated using the Darcy–Weisbach equation, and minor losses may be estimated as multiples of the velocity head. Conversion of headlosses to pressure drops requires the specific weight of air:

$$\Delta p = \gamma h_L \quad (11.108)$$

where  $g$  = the acceleration due to gravity ( $\text{m/s}^2$ )  
 $h_L$  = the headloss experienced by the flowing air (m)  
 $M_r$  = the relative molecular weight of air (28.96 g/mol)  
 $\Delta p$  = the pressure drop (Pa)  
 $p_{atm}$  = local atmospheric pressure (Pa)  
 $R$  = the gas constant (8.314 510 J/mol·K)  
 $T$  = the absolute temperature (K)  
 $\gamma$  = the unit weight of air ( $\text{N/m}^3$ )  
 $= p_{atm} M_r g / RT$

The pressure drop in the filter bed may be estimated using the MacDonald et al. (1979) modification of the Ergun (1952) equation [Eq. (9.234)]. For random and modular plastic media, this can be rewritten as follows:

$$\frac{\Delta p}{\gamma} = h_L = \frac{1}{6} f_m (Ha) \left( \frac{1}{\epsilon^3} \right) \frac{v_s^2}{g} \quad (11.109)$$

where  $A_p$  = the surface area of a single media “particle”  
 $a$  = the specific surface area of the filter media, i.e., the total media surface area divided by the total volume of the filter bed ( $\text{m}^2/\text{m}^3$ )  
 $= 6(1 - \epsilon) / d_{eq}$  for granular media;  
 $d_{eq}$  = the equivalent particle diameter (m)  
 $= 6 (V_p / A_p)$

$f_m$  = MacDonald–El-Sayed–Mow–Dullien (1979) friction factor (dimensionless)  
 $\geq 180 (1 - \varepsilon/\text{Re}) + 1.8$  (for smooth media)  
 $\leq 180 (1 - \varepsilon/\text{Re}) + 4$  (for rough media)  
 $H$  = the height of the filter (m)  
 $\text{Re}$  = the bed Reynold's number (dimensionless)  
 $= \rho v_s d_{eq} / \mu$  or  $\rho v_s / \mu \cdot 6(1 - \varepsilon) / a$   
 $V_p$  = the volume of a single media particle ( $\text{m}^3$ )  
 $v_s$  = the superficial (approach) velocity of the air (m/s)  
 $\varepsilon$  = the bed porosity (dimensionless)  
 $\mu$  = the dynamic viscosity ( $\text{N}\cdot\text{s}/\text{m}^2$ )

In these relations,

$$Ra = \varepsilon \quad (11.110)$$

$$Aa = P \quad (11.111)$$

where  $A$  = the total cross-sectional plan area of the bed, including voids and media ( $\text{m}^2$ )  
 $P$  = the wetted perimeter of the media in any planar cross section (m)  
 $R$  = the bed hydraulic radius, i.e., the void volume divided by the surface area of the media (m)

The void volume in the bed must be reduced by the liquid and biomass volumes in the filter, and, in the case of crushed stone, the grain diameter should be increased by twice the film thickness.

If no blower is installed, the airflow is developed by the natural draft. The natural draft is the gas phase pressure difference developed between the air in the bottom of the trickling and the surrounding air. It may be estimated from the following (Schroeder and Tchobanoglous, 1976):

$$\Delta p = \frac{p_{atm} M_r g H}{R} \left( \frac{1}{T_1} - \frac{1}{T_2} \right) \quad (11.112)$$

where  $\Delta p$  = the natural draft developed by the temperature difference (Pa)  
 $T_1$  = the cooler absolute temperature (K)  
 $T_2$  = the warmer absolute temperature (K)

The airflow through natural draft filters may stagnate for temperature differences of less than  $3^\circ\text{C}$  (Joint Task Force, 1992). If temperature differences this small are expected to be common, forced air circulation is needed.

In natural draft systems, the air passages consist of the pore spaces in the media, the under drains, and “chimneys” placed around the periphery of the bed and connecting the under drains with the atmosphere. The under drains and effluent piping must have at least 50% of their cross-sectional area open to air movement at the peak instantaneous hydraulic flow rate (including the recirculated flow) (Wastewater Committee, 1997).

Plastic media filters for municipal wastewater should have at least  $0.1 \text{ ft}^2$  of ventilation area for each 3 to 4.6 m of filter circumference and 1 to 2  $\text{m}^2$  of ventilation area per 1000  $\text{m}^3$  of media (Joint Task Force, 1992). The total cross-sectional area of the chimneys in stone filters must be at least 15% of the plan filter area (Wastewater Committee, 1997).

The minimum airflows recommended by the Joint Task Force (1992) are as follows:

***Roughing filters loaded at 120 to 320 kg BOD<sub>5</sub>/m<sup>3</sup>/day***

$$Q_{air} = 0.9 \cdot PF \cdot K \cdot Q \cdot (C_{\text{BOD}i} - S_{\text{BOD}}) \quad (11.113)$$

***Standard rate filters loaded at 40 to 80 kg BOD<sub>5</sub>/m<sup>3</sup> per day:***

$$Q_{air} = PF \cdot K \cdot Q \cdot (C_{\text{BOD}i} - S_{\text{BOD}}) \quad (11.114)$$

where  $C_{\text{BOD}_i}$  = the total BOD<sub>5</sub> in the flow applied to the filter (kg/m<sup>3</sup>)  
 $K$  = the volumetric air requirement per unit mass of BOD<sub>5</sub> removed [75 m<sup>3</sup> air (1 atm, 0°C)/kg BOD<sub>5</sub>]  
 $PF$  = the load peaking factor (dimensionless)  
 $Q$  = the total hydraulic flow rate applied to the filter, including recirculated flow (m<sup>3</sup>/s)  
 $Q_{\text{air}}$  = the airflow rate [m<sup>3</sup> air (1 atm, 0°C)/s or ft<sup>3</sup> air (SC)/sec]  
 $S_{\text{BOD}}$  = the soluble BOD<sub>5</sub> in the filter effluent (kg/m<sup>3</sup>)

The suggested volumetric air requirement amounts to an oxygen-to-BOD<sub>5</sub> mass ratio of about 21:1 and an implied oxygen transfer efficiency of 10%.

### Empirical CBOD Removal Formulas

There are many empirical and semitheoretical formulas for the design of trickling filters (Roberts, 1973). The more commonly encountered ones are listed below in chronological order. Designers should bear in mind that the validity of these formulas is under active review, and it appears that all of them fail to account for (1) the fraction of the media that is wetted and (2) the liquid dosing pattern. It is recommended that trickling filter designs be based on large-scale pilot studies. In this regard, the older literature for stone and plastic media contains mainly data from filters that may have been hydraulically underloaded, and these reports may seriously underestimate filter capacities.

It should also be borne in mind that all the models assume that the effluent BOD<sub>5</sub> is merely residual influent BOD<sub>5</sub> that the microbes did not remove and metabolize. However, it has been shown that in laboratory-scale filters fed known, soluble substrates, about 85% of the soluble organic matter in the effluent is a high molecular weight ( $M_r > 1000$ ) microbial product (Namkung and Rittman, 1986). This means that all trickling filter formulas, including those derived theoretically, have no mechanistic meaning, and the model parameter values can be expected to vary with wastewater composition and media configuration. All models should be treated as regression models.

#### National Research Council Formula

The NRC (Mohlman et al., 1946; Fair et al., 1948) formula was derived from correlations using data from stone media trickling filters at U.S. Army bases during World War II:

$$E = \frac{100}{1 + 0.0085 \sqrt{\frac{W}{VF}}} \quad (11.115)$$

where  $E$  = the BOD<sub>5</sub> removal efficiency, settled influent sewage to settled effluent, not counting recycle (%)  
 $F$  = the Mountfort (1924) recirculation factor (dimensionless)  
 $= (1 + R) / [1 + (1 - f_{\text{av}})R]^2$   
 $f_{\text{av}}$  = the “available” fraction of the BOD<sub>5</sub>, usually assumed to be about 0.9 (decimal fraction)  
 $R$  = the recirculation ratio, i.e., the ratio of the settled sewage flow to the recycled treated flow (dimensionless)  
 $V$  = the volume of the filter bed (ac·ft)  
 $W$  = the BOD<sub>5</sub> load in the settled sewage not counting the recirculated flow (lb/day)

Note the traditional units. The settled effluent will contain some particulate BOD<sub>5</sub> depending on the efficiency of the final clarifier. The scatter about this formula is very large, and it is only a rough guide to trickling filter performance. The filter load,  $W$ , does not include adjustment for the recirculated flow. The Mountfort recirculation factor,  $F$ , accounts for any recirculation effect.

The NRC recommends the same formula for the second stage of two-stage filters with an adjustment for the reduced BOD of the effluent from the first filter. The second-stage efficiency is as follows:

$$E_2 = \frac{100}{1 + 0.0085 \sqrt{\frac{W_1}{VF(1-e_1)^2}}} \quad (11.116)$$

where  $E_2$  = the BOD<sub>5</sub> removal efficiency of the second stage (%)  
 $e_1$  = the fractional BOD<sub>5</sub> removal of the first-stage filter (decimal fraction)  
 $W_1$  = the BOD<sub>5</sub> load in the effluent of the first-stage clarifier (lb/day)

The data scatter about Eq. (11.116) is even worse for the second-stage plants.

Two-stage filtration is a solution to the structural and aeration problems associated with deep piles of stone. These include collapse of the pile, crushing of the deeper media layers, and low airflow rates. The second stage functions as the deeper layers of tall stone filter, not as a separate treatment process.

### **The Germain Formula**

The Germain (1966) formula is the most commonly used design equation for plastic media. It is based on the theoretical and empirical work of Velz (1948), Howland (1958), and Schulze (1960):

$$\frac{C_{\text{BOD}_{5e}}}{C_{\text{BOD}_o}} = \exp\left(-\frac{K_G H}{(Q/A)^{0.5}}\right) \quad (11.117)$$

where  $A$  = the plan area of the trickling filter (ft<sup>2</sup>)  
 $C_{\text{BOD}_o}$  = the BOD<sub>5</sub> in the settled sewage not including the recirculated flow (mg/L)  
 $C_{\text{BOD}_{5e}}$  = the BOD<sub>5</sub> in the settled effluent (mg/L)  
 $K_G$  = the Germain treatability factor [ $s^{-0.5} \cdot m^{-0.5}$  or  $(\text{gpm})^{0.5}/\text{ft}^2$ ]  
 $\approx 0.088 (\text{gpm})^{0.5}/\text{ft}^2$  for settled domestic sewage  
 $Q$  = the flow rate of the settled sewage not including any recirculated flow (gpm)

Equation (11.117) applies to trickling filters with and without recirculation. In both cases,  $Q$  is the settled sewage flow rate without any adjustment for the volume of the recirculated flow, and  $C_{\text{BOD}_o}$  is the BOD<sub>5</sub> of the settled sewage from the primary clarifier, again without any adjustment for the BOD<sub>5</sub> in the recirculated flow. In other words, Germain's experiments show no effect of recirculation upon filter performance.

### **Eckenfelder's Retardant Model**

The Eckenfelder (1961) retardant formula was derived from the assumption that the rate of BOD<sub>5</sub> removal decreases as the contacting time increases. Eckenfelder applied his model to data from stone filters treating domestic sewage and obtained:

$$\frac{C_{\text{BOD}_{5e}}}{C_{\text{BOD}_o}} = \frac{1}{1 + \frac{2.5H^{0.67}}{(Q/A)^{0.50}}} \quad (11.118)$$

where  $C_{\text{BOD}_o}$  = the settled sewage BOD<sub>5</sub> (mg/L)  
 $C_{\text{BOD}_{5e}}$  = the settled trickling filter effluent BOD<sub>5</sub> (mg/L)  
 $A$  = the plan area of the filter (ac)  
 $H$  = the depth of media (ft)  
 $Q$  = the settled sewage flow rate (mgd)

### **Galler-Gotaas Correlation**

Galler and Gotaas (1964) performed multiple linear (logarithmically transformed) regression on a large sample of published operating data and obtained the following formula, which is one of several versions they derived:

$$C_{\text{BOD}_{5e}} = \frac{0.464 C_{\text{BOD}_i}^{1.19} \left( \frac{q + q_r}{q} \right)^{0.28} (q + q_r)^{0.13}}{(1 + H)^{0.67} T^{0.15}} \quad (11.119)$$

where  $C_{\text{BOD}_i}$  = the  $\text{BOD}_5$  in the mixture of settled sewage and recirculated flow applied to the filter (mg/L)  
 $C_{\text{BOD}_{5e}}$  = the  $\text{BOD}_5$  in the settled, final effluent (mg/L)  
 $H$  = the depth of media (ft)  
 $q$  = the settled sewage flow rate (mgad)  
 $q_r$  = the recirculation rate (mgad)  
 $T$  = the sewage temperature ( $^{\circ}\text{C}$ )

The multiple correlation coefficient for the logarithmic form of Eq. (11.119) is 0.974.

Equation (11.119) was modified by Blain and McDonnell (1965) in order to account for strong correlations among the data set:

$$C_{\text{BOD}_{5e}} = \frac{0.860 C_{\text{BOD}_o}^{1.31} q^{0.11}}{\left( \frac{q + q_r}{q} \right)^{0.35} H^{0.68} T^{0.57}} \quad (11.120)$$

where  $C_{\text{BOD}_o}$  = the  $\text{BOD}_5$  of the settled sewage applied to the filter, not including the recirculated flow or  $\text{BOD}_5$  (mg/L).

The logarithmic form of Eq. (11.120) has a multiple linear correlation coefficient of 0.869.

### ***Bruce–Merkens Correlation***

An empirical formula developed by Bruce and Merkens (1973) fits a variety of plastic and stone media:

$$\frac{C_{\text{BOD}_{5e}}}{C_{\text{BOD}_i}} = \exp\left(-\frac{K_{15} \theta^{T-15} a}{Q/V}\right) \quad (11.121)$$

where  $a$  = the specific surface area of the media in  $\text{m}^2/\text{m}^3$   
 $C_{\text{BOD}_i}$  = the  $\text{BOD}_5$  in the mixture of settled sewage and recirculated flow applied to the filter (mg/L)  
 $C_{\text{BOD}_{5e}}$  = the  $\text{BOD}_5$  in the settled, final effluent (mg/L)  
 $K_{15}$  = the reaction rate coefficient at  $15^{\circ}\text{C}$   
 $\approx 0.037$  m/d  
 $Q$  = the settled sewage flow rate in  $\text{m}^2/\text{d}$   
 $V$  = the media volume in  $\text{m}^3$   
 $\theta$  = the Streeter–Phelps temperature correction coefficient (dimensionless)  
 $\approx 1.08$

Note that depth is not a factor in this model.

### ***Institution of Water and Environmental Managers***

The formula recommended by the Institution of Water and Environmental Management is as follows (Joint Task Force, 1992):

$$\frac{C_{\text{BOD}_{5e}}}{C_{\text{BOD}_i}} = \frac{1}{1 + \frac{K \theta^{T-15} a^m}{(Q/V)^n}} \quad (11.122)$$

where  $a$  = the specific surface area of the filter media ( $\text{m}^2/\text{m}^3$ )  
 $C_{\text{BOD}_i}$  = the  $\text{BOD}_5$  in the mixture of settled sewage and recirculated flow applied to the filter (mg/L)  
 $C_{\text{BOD}_{5e}}$  = the  $\text{BOD}_5$  in the settled, final effluent (mg/L)

$K = 0.0204 \text{ m}^{m+n}/\text{d}^n$  for stone and random media and  $0.400 \text{ m}^{m+n}/\text{d}^n$  for modular plastic media  
 $m = 1.407$  for stone and random media and  $0.7324$  for modular plastic media  
 $n = 1.249$  for stone and random media and  $1.396$  for modular plastic media  
 $Q$  = the settled sewage flow rate ( $\text{m}^3/\text{d}$ )  
 $V$  = the filter volume ( $\text{m}^3$ )  
 $\theta$  = the Streeter–Phelps temperature coefficient,  $1.111$  for stone and random media and  $1.089$  for modular plastic media

Note that media depth is not a factor in this model.

### Relative Reliability of Formulae

Benjes (1978) compared the predictions of the NRC formula, Eckenfelder's equation, and the Galler–Gotaas correlation to data from 20 treatment plants. Data from these treatment plants were not used in the original studies, so Benjes' work is a test of the predictive capabilities of the models. In general, the coefficient of determination (the correlation coefficient squared) was 0.50 for the NRC formula and the Galler–Gotaas correlation; it was 0.56 for the Eckenfelder equation. All of the formulas tend to predict better effluents than are actually achieved, especially when the effluent  $\text{BOD}_5$  is large; the Eckenfelder formula yields the smallest predicted effluent  $\text{BOD}_5$ . The divergences from observed data are greatest for effluent  $\text{BOD}_5$  above 30 mg/L. Below 30 mg/L, the three formulas are about equally accurate, but errors of plus or minus 50% in the predicted effluent  $\text{BOD}_5$  should be expected.

### Effect of Bed Geometry, Recirculation, and Hydraulic Load

Trickling filter models that incorporate Howland's HRT formula [like Eq. (11.117)] predict that if the media volume is held constant and the media depth is increased, the  $\text{BOD}_5$  removal efficiency will improve. This occurs because the liquid film thickness increases as  $Q/A$  does, and the HRT increases because the total volume of liquid in the bed is larger. These models also predict that increasing recirculation will improve removal efficiency, for the same reason.

The NRC formula also predicts that recirculation will improve removal efficiency. Again, this is because of increased contact between the wastewater and the media.

In the case of empirical equations like the Galler–Gotaas correlation, the prediction falls out of the regression, and no mechanism is implied.

It is now believed that increased hydraulic loading improves removal efficiency, but the improvement is due to more complete wetting of the media surface and a resulting increase in the effective media surface area. The high-intensity, low-frequency dosing recommended by the Joint Task Force (1992) is intended to provide maximum wetting of the surface area.

The effects of wetting can be reported as changes in the Germain treatability constant,  $K_G$ , which increases with hydraulic loading up to some maximum value, at which it is supposed that the media is completely wet (Joint Task Force, 1992). In a study by Dow Chemical, Inc., the reaction rate constant increased with hydraulic load up to about  $0.75 \text{ gpm}/\text{ft}^2$ , above which the rate coefficient was constant (Joint Task Force, 1992). Albertson uses a Germain treatability constant of  $0.203 \text{ L}^{0.5}/\text{s}^{0.5} \cdot \text{m}^2$  for fully wetted media at the reference conditions of  $20^\circ\text{C}$ , 6.1 m depth, and 150 mg/L influent  $\text{BOD}_5$ .

The Dow study also indicated that  $\text{BOD}_5$  removal per unit media volume was independent of media depth for any given hydraulic load that achieved complete wetting of the media. This result was supported by Bruce and Merkens (1973) and by others (Joint Task Force, 1992), and the relationship between the Germain treatability coefficient and the media depth can be represented as follows:

$$\frac{K_{G1}}{K_{G2}} = \sqrt{\frac{H_2}{H_1}} \quad (11.123)$$

so that  $K_G\sqrt{H}$  is a constant. The consequence of this is that the removal efficiency should depend on the volumetric rather than the areal hydraulic loading rate.

Albertson (Joint Task Force, 1992) recommends that the Germain constant also be corrected for the strength of the applied sewage:

$$K_G(H, C_{\text{BOD}_0}, T) = 0.203 \left( \frac{6.1}{H} \right)^{0.5} \left( \frac{150}{S_o} \right)^{0.5} 1.035^{T-20} \quad (11.124)$$

where  $C_{\text{BOD}_0}$  = the BOD<sub>5</sub> of the settled sewage not including the recirculated flow (mg/L)

$H$  = the media depth (m)

$K_G(H, C_{\text{BOD}_0}, T)$  = the Germain treatability constant at 20°C for the given filter depth and influent BOD<sub>5</sub>, in L<sup>0.5</sup>/(s<sup>0.5</sup>·m<sup>2</sup>)

$T$  = the wastewater temperature (°C)

Equation (11.124) only applies to vertical-flow, modular plastic media. The Germain treatability constant for shallow cross-flow media is significantly smaller.

### Effect of Temperature on Carbonaceous BOD Removal

The BOD<sub>5</sub> removal efficiency of trickling filters depends on the temperature of the slime layer. In the case of continuously dosed, high-rate filters, this is probably the temperature of the applied wastewater. However, the slime layer in intermittently dosed, low-rate filters spends most of its time in contact with the air circulating through the filter, and the slime temperature will be somewhere between the air temperature and the water temperature. If recirculation is practiced, the applied wastewater temperature tends to approach the ambient air temperature.

Schroepfer et al. (1952) derived the following formulas for the effects of temperature on the BOD<sub>5</sub> removal efficiency of rock filters:

#### *Low-rate, intermittently dosed filters:*

$$E_2 - E_1 = 0.62(T_2 - T_1); \quad r \cong 0.7 \quad (11.125)$$

#### *High-rate, continuously dosed filters:*

$$E_2 - E_1 = 0.34(T_2 - T_1); \quad r \cong 0.6 \quad (11.126)$$

where  $E$  = the percentage BOD<sub>5</sub> removal efficiency of the combined filter and secondary clarifier based on the total BOD<sub>5</sub> load (%)

$T$  = the wastewater temperature (°F).

Recirculation greatly increases the seasonal variation in BOD<sub>5</sub> removal efficiency. In a study of 17 stone filters in Michigan, Benzie, Larkin, and Moore (1963) found that the summer efficiency was 21 percentage points higher than the winter efficiency when recirculation was practiced but only 6 points higher without recirculation.

### Tertiary Nitrification

A trickling filter may be classified as tertiary or nitrifying only as long as the soluble BOD<sub>5</sub> in the applied flow is less than about 12 mg/L and the BOD<sub>5</sub>:TKN ratio is less than about 1 (Joint Task Force, 1992).

Nitrification in trickling filters is regarded to be mass transport limited as long as the ammonia concentration is greater than a few mg/L. The limiting flux may be that of oxygen or ammonia. According to the Williamson–McCarty (1976a, 1976b) analysis, the oxygen flux is rate limiting whenever

$$\frac{S_{\text{O}_2}}{S_{\text{NH}_3}} = \frac{S_{\text{O}_2,i}}{S_{\text{NH}_3,i}} < \frac{D_{f\text{NH}_3} \cdot v_{\text{O}_2} \cdot M_{r\text{O}_2}}{D_{f\text{O}_2} \cdot v_{\text{NH}_3} \cdot M_{r\text{NH}_3}} \quad (11.127)$$

where  $D_{f\text{NH}_3}$  = the diffusivity of ammonia nitrogen in the biofilm ( $\text{m}^2/\text{s}$ )  
 $D_{f\text{O}_2}$  = the diffusivity of oxygen in the biofilm ( $\text{m}^2/\text{s}$ )  
 $M_{r\text{NH}_3}$  = the relative molecular weight of ammonia nitrogen (17 g/mol)  
 $M_{r\text{O}_2}$  = the relative molecular weight of oxygen (32 g/mol)  
 $S_{\text{NH}_3}$  = the ammonia nitrogen concentration in the bulk liquid film (mg  $\text{NH}_3\text{-N/L}$ )  
 $S_{\text{NH}_3i}$  = the ammonia nitrogen concentration at the interface of the liquid and biofilm (mg/L)  
 $S_{\text{O}_2}$  = the oxygen concentration in the bulk liquid film (mg/L)  
 $S_{\text{O}_2i}$  = the oxygen concentration at the interface of the liquid and biofilm (mg/L)  
 $\nu_{\text{NH}_3}$  = the stoichiometric coefficient of ammonia nitrogen for nitrifier growth (dimensionless)  
 $\nu_{\text{O}_2}$  = the stoichiometric coefficient of oxygen for nitrifier growth (dimensionless)

Williamson and McCarty estimate that nitrification will be oxygen limited whenever the oxygen:ammonia ratio in the liquid film is less than 2.7 kg  $\text{O}_2$  per kg  $\text{NH}_3\text{-N}$ . As a practical matter, this means that tertiary nitrifying filters are oxygen-limited whenever the ammonia nitrogen concentration is larger than about 2 to 4 mg/L. In this range, the nitrification rate is also zero order with respect to ammonia.

### Okey–Albertson Design Procedure

Okey and Albertson (Joint Task Force, 1992) developed an empirical procedure for tertiary nitrifying filters. They first define a transition ammonia concentration for the boundary between oxygen limitation and ammonia limitation of nitrification. This varies with oxygen saturation of the liquid film approximately as shown in Table 11.14 (Okey and Albertson, 1989). They then assume the filter depth can be divided into an upper oxygen-limited region and a lower ammonia-limited region. The volumes of the two regions are calculated separately and then added to get the total media volume. The filter area is calculated by assuming a minimum hydraulic loading rate, which includes recirculated flow. This procedure is restricted to applied wastewaters containing less than 12 mg/L of soluble  $\text{BOD}_5$ , a  $\text{BOD}_5\text{:TKN}$  ratio of less than 1 and a combined  $\text{BOD}_5$  plus TSS concentration of 30 mg/L. It is assumed that modular media having a specific surface area of  $138 \text{ m}^2/\text{m}^3$  is employed.

### Oxygen-Limited Media Volume

Calculate the oxygen-limited media volume by assuming a nitrification rate of  $1.2 \text{ g NH}_3\text{-N}/\text{m}^2/\text{d}$  and a media-specific surface area of  $138 \text{ m}^2/\text{m}^3$ . The rate is constant between 10 and  $30^\circ\text{C}$ , and it is reduced below  $10^\circ\text{C}$  by using a Streeter–Phelps theta value of 1.035:

$$K_o(10^\circ\text{C} \leq T \leq 30^\circ\text{C}) = 1.2 \frac{\text{g NH}_3\text{-N}}{\text{m}^2 \cdot \text{d}} \quad (11.128)$$

$$K_o(T < 10^\circ\text{C}) = 1.2 \frac{\text{g NH}_3\text{-N}}{\text{m}^2 \cdot \text{d}} \times 1.035^{T-10}$$

**TABLE 11.14** Ammonia Concentrations for the Transition from Oxygen to Ammonia: Limitation of Nitrification in Trickling Filters

Oxygen Saturation (%)	Transition Ammonia Concentration (mg $\text{NH}_3\text{-N/L}$ )		
	$10^\circ\text{C}$	$20^\circ\text{C}$	$30^\circ\text{C}$
25	1.0	0.9	0.8
50	2.2	1.8	1.5
75	3.3	2.6	2.1
100	4.3	3.5	2.9

Source: Okey, R.W. and Albertson, O.E. 1989. "Diffusion's Role in Regulating Rate and Masking Temperature Effects in Fixed Film Nitrification," *Journal of the Water Pollution Control Federation*, 61(4): 500.

The oxygen-limited media volume will be

$$V_o = \frac{(Q+Q_r)(C_{TKNo} - S_{NH_3t})}{K_o a} \quad (11.129)$$

where  $a$  = the specific surface area of the media ( $138 \text{ m}^2/\text{m}^3$ )

$C_{TKNo}$  = the total kjeldahl nitrogen concentration in the applied flow, including the TKN of the recirculated flow (mg TKN/L)

$K_o$  = oxygen-limited nitrogen oxidation rate ( $\text{g NH}_3\text{-N}/\text{m}^2 \cdot \text{d}$ )

$Q + Q_r$  = the applied flow ( $\text{m}^3/\text{d}$ )

$S_{NH_3t}$  = the transition ammonia nitrogen concentration obtained from [Table 11.14](#) (mg  $\text{NH}_3\text{-N}/\text{L}$ )

$V_o$  = the oxygen-limited media volume in ( $\text{m}^3$ )

### Ammonia-Limited Media Volume

Use the empirical ammonia oxidation rate:

$$K_n(7^\circ\text{C} \leq T \leq 30^\circ\text{C}) = 1.2 \left( \frac{S_{NH_3e}}{S_{NH_3t}} \right)^{0.75} \frac{\text{g NH}_3\text{-N}}{\text{m}^2 \cdot \text{d}} \quad (11.130)$$

and calculate the required ammonia-limited volume by,

$$V_n = \frac{(Q+Q_r)(S_{NH_3t} - S_{NH_3e})}{K_n a} \quad (11.131)$$

where  $K_n$  = the ammonia-limited nitrogen oxidation rate ( $\text{g NH}_3\text{-N}/\text{m}^2 \cdot \text{d}$ )

$S_{NH_3e}$  = the required effluent ammonia nitrogen concentration (mg  $\text{NH}_3\text{-N}/\text{L}$ )

$V_n$  = the ammonia-limited media volume ( $\text{m}^3$ )

### Filter Depth

The filter depth is the total media volume divided by the required minimum hydraulic load,  $47 \text{ m}^3/\text{m}^2/\text{d}$ :

$$H = \frac{V_n + V_o}{(Q+Q_r)/q} \quad (11.132)$$

where  $q$  = the required hydraulic loading rate in  $\text{m}^3/\text{m}^2 \cdot \text{d} \geq 0.54 \text{ dm}^3/\text{m}^2 \cdot \text{s} = 47 \text{ m}^3/\text{m}^2 \cdot \text{d}$ .

### Aeration and Dosing

Okey and Albertson recommend that forced ventilation be employed with a minimum oxygen supply of  $50 \text{ kg O}_2$  per  $\text{kg O}_2$  consumed, i.e., an oxygen transfer efficiency of 2%.

They also recommend instantaneous dosing intensity of 25 to 250 mm per pass and occasional flushing at greater than 300 mm per pass.

### Combined BOD and TKN Removal

Heterotrophic bacteria grow more rapidly than nitrifiers and are less sensitive to low oxygen concentrations, so when the influent  $\text{BOD}_5$  is high, the heterotrophs tend to displace the nitrifiers that become limited to the deep, low  $\text{BOD}_5$  region of the filter. TKN removal rates decrease as the  $\text{BOD}_5$ :TKN ratio increases and as the temperature increases. At  $15^\circ\text{C}$ , the TKN oxidation rate is approximately

$$K_{TKN}(15^\circ\text{C}) \cong 1.086 \left( \frac{C_{\text{BOD}_5^i}}{C_{\text{TKN}^i}} \right)^{-0.44} \quad (11.133)$$

where  $K_{\text{TKN}}(15^\circ\text{C}) =$  the total kjeldahl nitrogen removal rate ( $\text{g TKN}/\text{m}^2 \cdot \text{d}$ )  
 $C_{\text{BOD}_5i} =$  the total  $\text{BOD}_5$  concentration in the applied flow ( $\text{mg TKN}/\text{L}$ )  
 $C_{\text{TKNi}} =$  the total TKN concentration in the applied flow ( $\text{mg TKN}/\text{L}$ )

At  $25^\circ\text{C}$ , the TKN oxidation rate is likely to be less than half that indicated by Eq. (11.133); at  $10^\circ\text{C}$ , the rate may be 50% higher. The scatter about Eq. (11.133) is large; the standard deviation of the estimate is  $0.17 \text{ g TKN}/\text{m}^2/\text{d}$ .

The Albertson–Okey (Joint Task Force, 1992) procedure for combined  $\text{BOD}_5$  and ammonia removal is discussed below.

### Media Volume

Assuming nitrification controls, determine the influent  $\text{BOD}_5$  and TKN, and estimate the TKN oxidation rate for the indicated  $\text{BOD}_5$ :TKN ratio. Reduce this estimate by 1 standard deviation, i.e.,  $0.17 \text{ g TKN}/\text{m}^2/\text{d}$ . Use the corrected oxidation rate to calculate the required media volume:

$$K_{\text{TKNdes}} \cong 1.086 \left( \frac{C_{\text{BOD}_5i}}{C_{\text{TKNi}}} \right)^{-0.44} - 0.17 \quad (11.134)$$

$$V_{\text{TKN}} = \frac{(Q + Q_r)(C_{\text{TKNi}} - S_{\text{TKNse}})}{K_{\text{TKNdes}} a} \quad (11.135)$$

where  $a =$  the specific surface area of the media ( $\text{m}^2/\text{m}^3$ )  
 $= 98 \text{ m}^2/\text{m}^3$ , for design purposes

$K_{\text{TKNdes}} =$  the design TKN oxidation rate for  $15^\circ\text{C}$  ( $\text{g TKN}/\text{m}^2 \cdot \text{d}$ )

$S_{\text{TKNse}} =$  the required settled effluent TKN ( $\text{mg TKN}/\text{L}$ )

$V_{\text{TKN}} =$  the total media volume required for combined  $\text{BOD}_5$  and TKN removal ( $\text{m}^3$ )

### Aeration and Distribution

Powered ventilation of the towers is required to prevent oxygen limitation. The oxygen requirement (at 2% transfer efficiency) is  $50 \text{ kg O}_2$  per kg oxygen demand removed:

$$R_{\text{O}_2\text{des}} = 50(Q + Q_r) \left[ 1.2(C_{\text{BOD}_5i} - C_{\text{BOD}_5e}) + 4.57(C_{\text{TKNi}} - C_{\text{TKNe}}) \right] \quad (11.136)$$

where  $C_{\text{BOD}_5e} =$  the total  $\text{BOD}_5$  in the filter underflow ( $\text{g}/\text{m}^3$ )

$C_{\text{BOD}_5i} =$  the total  $\text{BOD}_5$  in the applied flow ( $\text{g}/\text{m}^3$ )

$C_{\text{TKNe}} =$  the total TKN in the filter underflow ( $\text{g}/\text{m}^3$ )

$C_{\text{TKNi}} =$  the total TKN in the applied flow ( $\text{g}/\text{m}^3$ )

$Q + Q_r =$  the applied flow ( $\text{m}^3/\text{s}$ )

$R_{\text{O}_2\text{des}} =$  the design oxygen requirement in ( $\text{g}/\text{s}$ )

The minimum hydraulic application rate to ensure adequate wetting is  $47 \text{ m}^3/\text{m}^2/\text{d}$ . The wastewater application should be intermittent with an instantaneous dosing rate of 15 to 300 mm per pass.

### Predator Control

Trickling filters mimic the physical ecology of brooks. Consequently, they become populated with significant numbers of insect larvae, worms, and snails that are adapted to eating microbial films. No steady state is possible in such an ecosystem, and the biofilm and its predators exhibit oscillations that are out of phase. All trickling filters exhibit such oscillations, but the oscillations in nitrifying filters are more pronounced, because the nitrifying bacteria grow more slowly than the heterotrophic bacteria. In fact, predator infestations can denude significant portions of the media surface, thereby preventing nitrification. Such denudation is followed by predator starvation and population decline, which permit reestablishment of the nitrifying film. There is, of course, no nitrification until the film has regrown.

Parker et al. (1989) recommend that tertiary nitrifying filters be flooded and then backwashed about once per week to kill and/or remove insect larvae, worms, and snails that consume the nitrifying biomass.

### **Effect of Temperature on Nitrification**

The Joint Task Force (1992) concluded that temperature had little effect on the performance of tertiary nitrifying filters. The effects that are commonly reported were attributed instead to deficiencies in wetting, biofilm consumption by predators, overgrowth by heterotrophs, ammonia concentrations, and oxygen transfer limitations. In fact, if filters are liquid film mass transfer limited, as assumed in the Logan–Hermanowicz model, then the controlling rate process is diffusion, and the Streeter–Phelps (1925) temperature correction factor for the process is about 1.024, which leads to relatively small temperature effects.

The data for combined BOD<sub>5</sub> removal and nitrification is limited. It appears that the uncorrected TKN oxidation rate given by Eq. (11.134) should be decreased by about 50% if the temperature is 25°C or more, and that it may be increased by about 50% if the temperature is 10°C.

### **pH**

The pH requirements of activated sludge apply to BOD<sub>5</sub> removal in trickling filters as well. Tertiary nitrifying filters may develop low pHs as a result of nitric acid formation. Limited data presented by Huang and Hopson (Parker et al., 1975) suggest that the nitrification rate actually increases up to about pH 8.4; at pH 7.1, the rate is only 50% of the maximum, and at pH 7.8, it is 90% of the maximum. This is somewhat contrary to the pH effect reported by Downing and Knowles for activated sludge, in which they indicated no increase in nitrification rate above pH 7.2.

### **Inhibitors**

The list of inhibitors given for activated sludge in [Tables 11.4 and 11.5](#) probably applies to trickling filter nitrification as well.

### **Secondary Clarifiers**

See “Secondary Clarifiers” in Section 11.2.

## **Intermittent Sand Filters**

The first treatment plant to incorporate the process was built in 1871 in Merthyr-Tydvil, Wales. The first American intermittent sand filter was installed in Medfield, MA, in 1887. Intermittent sand filters are still commonly used in rural or isolated areas where land costs are low and operating staff scarce. They also have been used to upgrade stabilization pond effluents.

### **Filtration Media**

The earliest intermittent sand filters were constructed by removing the topsoil and vegetation from natural deposits of coarse sand and leveling the sand surface. Nowadays, the filtration media are engineered to specification.

Intermittent sand filtration fell out of favor after World War II, and current design regulations seldom include specific mention of them. However, the older standards of the Wastewater Committee (1968) and the Technical Advisory Board (1962) do. A combination of their more stringent limits and the recommendations of the Committee of the Sanitary Engineering Division on Filtering Materials (1936) produces media with the following characteristics:

- Depth: at least 36 in. (but not more than 48 in.)
- Effective size ( $d_{10}$ ): if the wastewater is applied by flooding, 0.3 mm to 0.6 mm; if it is applied by rotating distributor, 0.4 mm to 1.0 mm
- Uniformity coefficient ( $d_{60}/d_{10}$ ): less than 3.5
- Homogeneity of placement: media should be free of strata or veins of varying size
- Media cleanliness: practically free of clay, loam, soft limestone, or other material subject to disintegration; less than 1% organic matter; less than 3% acid-soluble matter

- Sand grain composition: siliceous sand without calcareous or argillaceous matter
- Shape: rounded, not angular; do not use crushed stone

The combinations of depth, effective size, and dosing method are intended to avoid short-circuiting of the flow through the filter. Especially in the case of flooding, if coarse sands are used, most of the flow will enter the bed near the inlet, and large parts of the bed will receive little or no flow.

A gravel layer supports and retains the filter media. A minimum of three gravel layers is required:

- Technical Advisory Board (1962)
  - From 2 in. below to 2 in. above the drains — grain sizes of 2 in. to 1.5 in.
  - From 2 to 4 in. above the drains — grain sizes of 1 in. to ½ in.
  - From 4 to 6 in. above the drains — grain sizes of ½ in. to ¼ in.
- Wastewater Committee (1968)
  - From below the drains to 6 in. above them, three layers declining in size upwards of — grain sizes of 1.5 in. to ¾ in.; grain sizes of ¾ in. to ¼ in.; and grain sizes of ¼ in. to ⅛ in.

### Filter Size and Number

A minimum of two filters is needed to treat continuous raw sewage flow. Each filter should be less than ¾ to 1 acre in plan area (Metcalf and Eddy, 1930). The beds are normally rectangular and grouped around a central distribution box. Nowadays, they are constructed in filter boxes to avoid groundwater contamination. Unless the filter is covered, or is surrounded by a high curb, its surface needs to be protected from dirt and dust by a buffer strip of pavement around its edge, and the grounds farther away should be grassed and mowed regularly with collection of the clippings (Keefer, 1940). Vegetation will grow on the filter surface and must be periodically removed.

### Dosing

Although early intermittent sand filters accepted raw sewage, nowadays, at least primary settling or microscreening is required. The BOD<sub>5</sub> and SS loading should not exceed 2.5 lb per 1000 ft<sup>2</sup>/d (Technical Advisory Committee, 1962). The hydraulic loads should not exceed the following (Wastewater Committee, 1968):

- Normal settled sewage: 3 gal/ft<sup>2</sup>/day
- Strong settled sewage: 1 gal/ft<sup>2</sup>/day
- Settled trickling filter effluent: 10 gal/ft<sup>2</sup>/day

Removal efficiencies are insensitive to small deviations from these recommendations (Furman, Calaway, and Grantham, 1955).

Wastewater is applied to the filter at least two to not more than four times per day in depths of at least 3 in each time (Keefer, 1940). An application rate of once per day impairs removal efficiency (Furman, Calaway, and Grantham, 1955). The higher dosing frequency is more appropriate in the south and to weaker sewages (Imhoff, Müller, and Thistlewaite, 1971). In the case of strong wastes, several days may be required between doses to avoid clogging. Babbitt and Baumann (1958) and Keefer (1940) cite Barbour's recommendation of an instantaneous dosing rate of 1 ft<sup>3</sup>/sec for every 5000 ft<sup>2</sup> of surface for about 20 min. Imhoff, Müller, and Thistlewaite (1971) recommend that dosing by flooding be completed in less than 15 min.

The distribution system may be as follows:

- Impulse-driven rotary, perforated pipes like those used in trickling filters
- A manifold of different pipes discharging onto splash plates placed not more than 30 to 60 ft apart over the filter surface
- A ridge and furrow surface

Intermittency of flow may be achieved via dosing siphons (traditional) or pumps (modern and preferred).

## Cleaning

Filter surfaces gradually accumulate a fibrous humus surface mat that must be removed by periodic raking. Cleaning may be indicated if the applied dose takes more than 20 to 30 min to infiltrate the surface. The rakings may be recycled to agricultural land like other sewage sludges if they pass the EPA's Rule 503 requirements (see below). The rakings may amount to 7 to 9 yd<sup>3</sup>/mil gal (Keefer, 1940).

The bulk of the biological action occurs in the top 5 to 8 in. of the bed, and sometimes this part of the bed clogs. Clogging may be relieved by harrowing the surface or by removing and replacing the top ¾ to 2 in. of sand (Keefer, 1940; Babbitt and Baumann, 1958). If several filters are available, it may be feasible to relieve clogging by taking one unit at a time out of service.

In northern climates, ice accumulation may clog the filter surface. This may be alleviated by plowing the surface into ridges and furrows (Babbitt and Baumann, 1958), and by increasing the depth of dosage to 1 ft (Imhoff, Müller, and Thistlewaite, 1971). In extreme climates, the filters should be covered.

## Drainage

The underdrains should be at least 4 in. in diameter, placed not more than 10 ft on centers and sloped to the outlet. They may be open-joint (1/4 to 3/8 in. spacing with tar paper cover over the top half of the diameter) or perforated, vitrified clay or concrete.

## Aeration

Intermittent sand filters are aerated by the wave of sewage moving through the filter. This wave expels the deoxygenated air in the media pores and draws fresh air in behind it. Separate aeration systems are not needed.

## Performance

If the average dosing rate for normal sewage is less than about 50,000 gal/ac/day, the effluent will be clear, odorless, colorless, and nitrified (Babbitt and Baumann, 1958). An effluent BOD<sub>5</sub> of a few mg/L and virtually no SS can be expected (Keefer, 1940).

## Rotating Biological Contactors

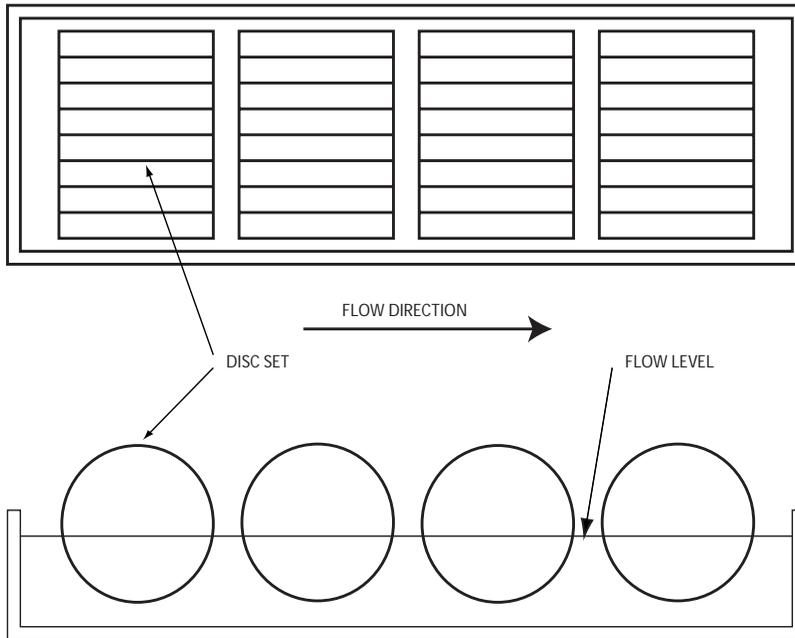
The rotating biological contactor (RBC) is also known as *rotating biological disc*, *rotating biological surface* (RBS), *bio-disc*, *rotating filter*, and *rotating biological filter*, and is marketed under Aero-Surf™, Bio-Surf™, BioSpiral™, and Surfact™. It consists of a number of partially submerged discs mounted on a rotating shaft. The discs provide a surface for the attachment of microbes and their predators. The rotation mixes the tank contents, aerates, and promotes mass transfer to the attached biomass. Oxygen is also transferred to the biomass when it is lifted out of the tank by the rotation and exposed to the air. Nowadays, many RBC installations have supplemental diffused air aeration.

Weigand in Germany patented the earliest RBC in 1900 (Peters and Alleman, 1982). It consisted of a rotating, hollow cylinder made of wooden slats. In 1925, Doman used rotating galvanized steel discs for the biomass support. Lack of suitable materials delayed the further development of the RBC until the 1950s, when Popel and Hartmann made improved discs out of expanded polystyrene. The J. Conrad Stengelin Co. in Germany manufactured large discs for use in a municipal treatment plant in Stuttgart, which went operational in 1960. The first RBC application in the U.S. was at the Eiler Cheese Co. in DePere, WI, in 1970. The first full-scale municipal RBC plant was built in Pewaukee, WI (Joint Task Force, 1992).

## Configuration

The usual process train consists of preliminary treatment, primary settling, staged RBC treatment, final clarification, and disinfection. Solids are not usually recycled from the final clarifier underflow to the RBC.

The combination biomass support-aeration system consists of shaft-mounted modules of corrugated, high-density polyethylene discs (Fig. 11.9). The shaft is usually square in cross section. The typical motor-driven module is about 12 ft in diameter, and 25 ft long and contains 2750 ft<sup>3</sup> of media (Aerobic



**FIGURE 11.9** Rotating biological contactor.

Fixed-Growth Reactors Task Force, 2000). The surface area available for biofilm support on a single shaft will be between 100,000 ft<sup>2</sup> (low density, 36 ft<sup>2</sup>/ft<sup>3</sup>) to 150,000 ft<sup>2</sup> (high density, 55 ft<sup>2</sup>/ft<sup>3</sup>) depending on the intended use. Low-density media are used for the initial CBOD-removal-only stages in order to avoid clogging inside the discs; high-density media is used in later nitrification stages where thin biofilms are desired. Motor-driven systems sometimes include diffused air for supplemental air supply.

Diffused air drives are also available. Diffused air-driven modules have cups on their peripheries to capture the rising air bubbles and generate torque. They also have larger diameters, typically 16 ft, to increase the available torque. Rising air bubbles may improve mass transfer to the discs and scour excessive biomass from them.

The side water depth is typically 5 ft, and approximately 40% of the disc area is submerged. The disc rotational speed is about 1 to 1.6 rpm. The modules are usually rotated by constant-speed, variable-torque electric motors connected to the shafts by gearboxes, chain and sprocket, or V-belt and pulley transmissions.

The modules are installed in tanks configured as mixed-cells-in-series in order to prevent short-circuiting. Each cell may contain more than one module. If baffles are installed between modules in a single cell, they should be removable for maintenance. The inlet piping should allow for step-feed among the cells.

Modules may be installed with the shafts parallel to the flow (small tanks) or across the flow (large tanks).

Tanks are covered to avoid media deterioration by ultraviolet light, to avoid the accumulation of algal mats, and to protect the biofilm from excessive high or low temperatures.

The total area of the enclosed media determines tank liquid volumes. A common requirement is 0.12 gal/ft<sup>2</sup> of media (Aerobic Fixed-Growth Reactors Task Force, 2000). This results in a HRT of about 1.4 hr.

### **CBOD Removal Kinetics**

The Benjes empirical formula is commonly used for RBC design. This model is conservative in that it predicts larger module volumes than do some other models. It is an adaptation of the Germain formula used in trickling filter design (Aerobic Fixed Growth Reactors Task Force, 2000):

$$\frac{C_{\text{BOD}_{5e}}}{C_{\text{BOD}_o}} = \exp\left[-K_B(V/Q)^{0.5}\right] \quad (11.137)$$

where  $C_{\text{BOD}_o}$  = the BOD<sub>5</sub> in the settled primary sewage (mg/L)  
 $C_{\text{BOD}_{5e}}$  = the BOD<sub>5</sub> in the final settled effluent (mg/L)  
 $K_B$  = the Benjes treatability constant (gpm<sup>0.5</sup>/ft<sup>1.5</sup>)  
 $\approx 0.3$  gpm<sup>0.5</sup>/ft<sup>1.5</sup> at 13°C  
 $\approx 0.2$  gpm<sup>0.5</sup>/ft<sup>1.5</sup> at 7°C  
 $\approx 0.15$  gpm<sup>0.5</sup>/ft<sup>1.5</sup> at 5°C  
 $Q$  = the settled sewage flow rate (gpm)  
 $V$  = the total module volume (cu ft)

The value of 0.3 gpm<sup>0.5</sup>/ft<sup>1.5</sup> for  $K_B$  represents the median of the scattered data. The envelope values are 0.2 and 0.4 gpm<sup>0.5</sup>/ft<sup>1.5</sup>.

The Streeter–Phelps theta value for temperature corrections to  $K_B$  is poorly known. Designers have used values between 1.01 and 1.05. Most manufacturers estimate that the removal rate per unit area falls by about one-third as the temperature declines from 18 to 6°C.

The settled effluent BOD<sub>5</sub> in mg/L also can be estimated as 15 times the areal organic loading on the discs in lb BOD<sub>5</sub>/1000 ft<sup>2</sup>/day.

### Dissolved Oxygen

A dissolved oxygen level of at least 2 mg/L should be maintained in each cell.

### Biofilm Control

The active biofilm thickness is limited to about 0.1 mm. Biofilm thickness on the media should be monitored by load cells or strain gages. Excessive biofilm thickness may be controlled by diffused air scouring, increasing rotational speed, reversing rotational direction, taking the module out of service (resting), removing baffles between modules to decrease BOD concentrations around a module, and chemical stripping.

### Nitrification

The TKN available for nitrification is estimated by (Aerobic Fixed-Growth Reactors Task Force, 2000):

$$\text{TKN}_a = \text{TKN}_t - 0.055\text{BOD}_{5r} - 1 \quad (11.138)$$

where  $\text{BOD}_{5r}$  = the BOD<sub>5</sub> removed in the CBOD removal stages (mg/L)  
 $\text{TKN}_a$  = the available TKN (mg/L)  
 $\text{TKN}_t$  = the TKN in the influent of the RBC's first stage (mg/L)

The TKN of any recycle streams returned to the primary clarifier must be included.

The areal nitrification rate is normally represented by a Monod-like formula (Aerobic Fixed-Growth Reactors Task Force, 2000):

$$q_n = \frac{q_{n\max} S_{na}}{K_{na} + S_{na}} \quad (11.139)$$

where  $K_{na}$  = the Monod affinity constant for nitrification by RBC (mg NH<sub>3</sub>-N/L)  
 $\approx 0.4$  mg NH<sub>3</sub>-N/L  
 $q_n$  = the areal ammonia nitrogen uptake rate (g NH<sub>3</sub>-N/m<sup>2</sup> d)  
 $q_{n\max}$  = the maximum areal ammonia nitrogen uptake rate (g NH<sub>3</sub>-N/m<sup>2</sup> d)  
 $\approx 1.5$  g NH<sub>3</sub>-N/m<sup>2</sup> d (0.3 lb NH<sub>3</sub>-N/1000 ft<sup>2</sup>/day) at 13°C  
 $S_{na}$  = the ammonia nitrogen concentration in the stage (mg NH<sub>3</sub>-N/L)

The temperature dependence of nitrification in RBCs is poorly known. The Wastewater Committee (1992) recommends that nitrification at 5°C requires 2.5 times the media area needed at 13°C. Temperatures higher than 13°C do not increase the nitrification rate.

### Staging

Because Eq. (11.137) is a plug flow model, the RBC tankage must be compartmentalized as mixed-cells-in-series and the shafts distributed among the cells. A minimum of four stages should be constructed. Generally, several (at least two) parallel, multistage trains will be constructed. The design areal BOD load and the design peak influent flow and strength are used to determine the required number of first-stage shafts. Low-density media is assumed. The total influent is divided equally among the parallel trains.

Equation (11.137) is used to estimate the total media volume required for CBOD removal, and this is converted into the required number of shafts (at 2750 ft<sup>3</sup>/shaft). This estimate is multiplied by a safety factor of 1.25.

The areal BOD load on the first stage should be less than 2.5 lb soluble BOD<sub>5</sub> per 1000 ft<sup>2</sup>/day and 6.4 lb total BOD<sub>5</sub>/1000 ft<sup>2</sup>/day (Joint Task Force, 1992). A design first-stage loading of 6 lb total BOD<sub>5</sub>/1000 ft<sup>2</sup>/day is commonly used. Loadings above this level tend to produce excessive biofilm thickness and possible structural overloading, varying motor loads and rotational speed, oxygen limitation and odors, reduced BOD removal efficiency, and nuisance growths of *Beggiatoa* spp. If recycles are returned to the primary clarifier, their BOD load must be included. The first stage will remove the greatest amount of BOD, and it may require supplementary diffused air.

The remaining shafts calculated from Eq. (11.137) are distributed uniformly among the parallel trains. Staging is achieved by placing a removable baffle between each shaft in a train.

The nitrification staging proceeds similarly. The settled effluent of the CBOD removal process is the influent of the nitrification process. The media area is determined by the NPDES permit conditions, the expected available TKN determined by Eq. (11.138), and the areal nitrification rate determined by Eq. (11.139). This area is usually multiplied by a safety factor of 1.25. The use of Eq. (11.130) implies a completely mixed tank, but each shaft may be baffled, and this will give an additional safety factor.

### Waste Solids Production

The Aerobic Fixed-Growth Reactors Task Force (2000) estimates waste solids production to be between 0.5 and 0.8 kg per kg BOD<sub>5</sub> removed, with the higher yields occurring in cold weather and at high organic loading rates.

The underflow from the secondary clarifier will contain about 2.5 to 3% solids and may be gravity thickened to about 5% solids.

### Operation and Design Problems

Some of the early RBC installations exhibited an excessive rate of mechanical and structural failures, but newer installations do not.

Present-day operation and design problems include the following (Aerobic Fixed-Growth Reactors Task Force, 2000):

- Excessive first-stage loadings: remove baffles between first and second stages to reduce BOD concentrations in the first stage; increase aeration by increasing rotational speed or adding supplementary diffused air; bypass some influent to second or later stages
- Excessive or uneven biomass accumulation on the discs: reduce loadings on the discs as above; clean discs; take discs out of service to starve biomass
- Inadequate solids control and loss of fine SS in final effluent: reduce high hydraulic loadings by equalization; recirculate some secondary clarifier underflow sludge to the RBC first stage to promote flocculation of fines; add coagulant prior to secondary clarifier inlet
- Corrosion of metal supports

## Anaerobic Rotating Biological Contactors

Tait and Friedman developed the anaerobic rotating biological contactor (AnRBC) at bench-scale in 1980 and studied it as a sludge digestion process.

Dimensional analysis indicates that the relevant process variables are as follows (Phoon, 1982):

- The dimensionless methane yield: mass of methane produced per mass of COD fed,  $R_{\text{CH}_4}/R_{\text{COD}}$ , kg/kg
- The disc Reynolds number:  $fd/D_l$
- The product of the disc rotational speed and the hydraulic retention time:  $\tau f$

In Phoon's study, the gas production and COD loading data for AnRBCs fed primary sewage sludge could be correlated by,

$$\ln\left(\frac{R_{\text{CH}_4}}{R_{\text{COD}}}\right) = a \ln(\tau f) + b \quad (11.140)$$

where  $a, b$  = empirical constants, possibly depending on the disc Reynold's number (dimensionless)

$d$  = the disc diameter (m)

$f$  = the disc rotational speed (rev/sec, Hz)

$R_{\text{CH}_4}$  = the rate of production of methane (kg/s)

$R_{\text{COD}}$  = the rate of feed of COD (kg/s)

$\tau$  = the HRT (s)

## Combination Fixed-Growth Suspended-Growth Processes

A number of combined fixed-growth suspended-growth processes are available, most of which are patented.

### CAPTOR™ and LINPOR™

The CAPTOR™ processes use polyether foam pads about  $30 \times 25 \times 12.5$  mm in size, and the LINPOR™ process uses  $10 \times 13$  mm cubes. The pads and cubes are suspended in the aeration tank and retained by effluent screens. They retain a substantial amount of the biomass and allow high biomass concentration up to 9000 mg/L without overloading the secondary clarifier. The CAPTOR™ system uses about 25 pads/1000 m<sup>3</sup> of mixed liquor.

### Moving Bed Biofilm Reactor (MBBR)

The moving bed biofilm reactor marketed by the Kaldnes Miljøteknologi Corporation consists of polyethylene cylinders 10 mm in diameter and 7 mm long suspended in the aeration tank. The packed aeration tank has a specific surface area on the cylinder of 500 m<sup>2</sup>/m<sup>3</sup>. Again, the packing allows high biomass concentrations. Perforated plates retain the cylinders. The secondary clarifier captures any lost biofilm. All the clarifier underflow is wasted; no sludge recirculation is attempted.

### PACT™

DuPont markets the powdered activated carbon treatment (PACT™) process in which powdered activated carbon is added to the aeration basin. The usual application is to industrial wastewaters containing toxic or recalcitrant compounds that are sorbable to activated carbon. The PAC forms a matrix with the biomass, and the biomass may degrade the sorbed matter as well as suspended and soluble organic matter. The chief advantages are improvement in the following (Joint Task Force, 1992):

- Settling rates
- Dewaterability
- Biorefractory compound removal
- Odor removal
- Color removal

The PACT™ process can be sized to nitrify, if desired. Many industrial facilities practice single-pass use of the carbon with disposal, but carbon regeneration is feasible in larger plants.

## Contact Aeration

In contact aeration units, narrowly spaced, vertically oriented, flat-plate media fill the upper portion of the aeration tank and are completely and continuously submerged by the mixed liquor. Aeration is provided by diffused air on the tank bottom, and the bubbles rise between the plates, providing oxygen, enhancing mass transfer to the plates, and scouring excess biomass from them. Some of the early designs used wooden laths, granular media, wire screens, brushwood, and even corncobs. Synonyms are *nidus racks*, *tank filters*, *Imhoff filters*, and *Emscher filters*.

A variety of early and recent designs are described by Black and Phelps (1911), Clark and Gage (1913), Bach (1937a, 1937b), Buswell (Anonymous, 1923), Imhoff (1926a, 1926b), Buswell and Pearson (1929), Kato and Sekikawa (1968), and Rusten (1984). The first German plant was used to treat phenolic coke oven wastes at the Helene coalmines near Essen in the mid 1920s. The first American installation was built at Jacksonville, TX, in 1927.

## Configuration

Contact aeration systems include preliminary treatment to remove grit and debris, primary settling and scum removal, two-stage contact aeration with intermediate and final settling, disinfection, and sludge stabilization.

The final development in the U.S. was the Hays process ca. 1930 (Griffith, 1943). A Hays plant generally incorporates two aeration basins separated by an intermediate clarifier as well as primary and final clarifiers. Sludges are collected from each clarifier and aeration tank and are wasted without any recycle to the process. The aeration tanks were filled with vertical asbestos plates and have hopper bottoms to collect scoured biomass. The first municipality to adopt the Hays process was Elgin, TX, in about 1938.

## Design Criteria

The design load is taken to be the maximum 3-hr load and flow, and is estimated to be 150% of the annual average (Griffith, 1943). The contact plates are flat or corrugated and 8 ft high  $\times$  4 ft wide  $\times$  ¼ in. thick and spaced 2 in. center-to-center normal to the flow direction. The plates are evenly distributed between the two equal-size aeration tanks, and each aeration tank is divided into halves by baffles. There is normally 6 in. of clearance above the air diffusers on the bottom and 4 in. clearance below the water surface.

The design criteria for U.S. Army plants built in the 1940s were as listed below (Mohlman et al., 1946; Fair et al., 1948):

- The first-stage BOD loading was set at an average rate of 6.4 lb BOD/1000 ft<sup>2</sup>/day.
- The second-stage BOD loading should be the same, 6.4 lb BOD/1000 ft<sup>2</sup>/day.
- The aeration tanks have equal HRTs of about 1.5 to 2.5 hr each.
- The primary settling HRT is 2.5 hr.
- The intermediate clarifier HRT is 1.5 hr.
- The final clarifier HRT is 2.5 hr (Fair et al., 1948).
- The required air supply is at least 1.5 ft<sup>3</sup> air/gal sewage split 60/40 between the first and second aeration tanks.

The total system volume is not much less than that of an activated sludge plant.

The BOD removal efficiency can be estimated by (Mohlman et al., 1946; Fair et al., 1948),

$$E = \frac{100}{1 + 0.248w_a^{0.746}} \quad (11.141)$$

where  $E$  = the annual average BOD<sub>5</sub> removal efficiency, settled influent sewage to settled effluent (%)  
 $w_a$  = the annual average mass loading rate of BOD<sub>5</sub> (lb/1000 ft<sup>2</sup>/hr). Equation (11.141) applies to each stage separately and to both stages combined. The scatter is significant, and only five plants were used to develop the correlation.

The principal problems of the contact-aeration process are as follows (Mohlman et al., 1946):

- Inadequate oxygen transfer with the resulting development of anaerobic conditions in the biofilm and odor production, especially hydrogen sulfide
- Excessive biofilm buildup requiring frequent plate cleaning

### Modern Submerged Fixed-Growth Processes

The Joint Task Force (1992) reviewed a large number of fixed submerged packing systems, consisting of high-density media that rest on the tank floor (BIOCARBON™, BIOFOR™, DENITE™, DYNASAND™, FITRAZUR™, FLOPAC™, NITRAZUR™, and OXYAZUR™); low-density media that float and require some sort of retaining system (BIOFILTER™, BIOSTYR™, and DENIPOR™); media suspended in the aeration tank (RINGLACE™ and BIO-2™); and fully submerged rotating biological contactors (SURFACT™).

Some of these processes operate downflow (BIOCARBONE™), and others operate upflow (BIOFOR™). Some packings retain the biomass so that settling of the effluent is unnecessary. However, all packings must be periodically backwashed to remove excess growth. The empty bed HRTs are generally about 1 to 1.5 hr, so the units are small compared to equivalent activated sludge processes. Control systems and instrumentation are relatively complex (Metcalf & Eddy, Inc., 2002).

### Trickling Filter-Solids Contacting and Trickling Filter-Activated Sludge

Trickling filters and suspended growth processes have been combined in two ways: the trickling filter-activated sludge (TFAS) process and the trickling filter-solids contact (TFSC) process. In either case, the trickling filter usually discharges directly (without clarification) to an activated sludge plant.

In the TFAS, the activated sludge process may be nitrifying or nonnitrifying, but nonnitrifying is more common. Soluble BOD removal occurs in the trickling filter, and influent particulate BOD and lost trickling filter biomass are decomposed (at least partially) in the activated sludge unit. For a nonnitrification design, the roughing trickling filter is loaded at about 1.2 to 4.8 kg BOD/m<sup>3</sup>/day, and the activated sludge process has a HRT of 10 to 60 min and an SRT of 2 to 7 days. The overflow rate on the secondary clarifier is usually 2.0 to 3.5 m/h at peak flow (Metcalf & Eddy, Inc., 2002).

If the influent wastewater contains a significant amount of toxic substances, and if nitrification is to be guaranteed, a secondary clarifier may be installed between the trickling filter and the aeration tank. In this arrangement, the underflow of the intermediate clarifier is wasted and not recycled in order to remove any sorbed poisons. The process loadings are the same as for the TFAS scheme (Metcalf & Eddy, Inc., 2002).

In the TFSC, more of the BOD removal is transferred to the trickling filter, and the solids contacting process serves more as a flocculation step. Nitrification may occur in the trickling filter at the lower loadings. The trickling filter is loaded at about 0.3 to 1.2 kg BOD m<sup>3</sup>/day, and the solids contacting tank has an HRT of 10 to 60 min and an SRT of 0.3 to 2 days. The peak clarifier overflow rate is between 2 and 3 m/h (Parker and Bratby, 2001). TFSC plants also incorporate a flocculator/clarifier for the secondary settling step and may include sludge reaeration. Both of these processes are intended to improve solids flocculation and capture. The return sludge from the flocculator/clarifier or reaeration tank is pumped to the inlet end of the solids contact tank.

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## 11.4 Ponds

### General Considerations

#### Classification

Ponds (lagoons) are artificial bodies of surface water designed and constructed to provide wastewater treatment and storage. The most common designs are as follows:

- **Mechanically aerated** — Oxygen supplied by surface aerators, diffused air, etc.
- **Facultative** — Surface layers aerobic by atmospheric reaeration and photosynthesis, bottom layers anaerobic
- **Maturation (polishing, tertiary)** — A long detention time pond that receives the effluent of a secondary treatment system
- **Anaerobic** — Fully anaerobic, treating concentrated wastes and producing methane

Depending on the NPDES permit, the discharge from a pond may be as follows:

- **Continuous** — Continuous, unregulated discharge to a receiving water
- **Intermittent, controlled, or hydrograph controlled** — Discharge only during specified periods and in specified amounts
- **Total containment, nonoverflow, or evaporation/infiltration** — Discharge only via evaporation and/or infiltration (if permitted)

## Configuration

A pond system will consist of the following:

- Preliminary screening and grit removal to protect pumps and surface aerators against clogging by debris and to minimize grit accumulation
- The ponds
- Final clarifiers and/or filters to remove suspended biomass
- Disinfection
- Ancillary equipment and storage sheds
- All-weather access roads and parking
- Groundwater monitoring wells
- Fencing and buffer strips

Pond sites should be graded to divert storm water away from the dikes to prevent toe erosion (Wastewater Committee, 1997). In order to prevent liner rupture by buoyant uplift or differential settling, the pond bottom should be at least 4 ft above the top of the groundwater table and 10 ft above any bedrock.

The pond should be subdivided into at least three cells-in-series (Wastewater Committee, 1997). To prevent short-circuiting, no cell may be larger than 40 ac in plan area. Piping should permit isolation and bypassing of each cell for maintenance. The cells are generally rectangular with length-to-width ratios of 3:1 to 4:1. The minimum depth for any pond is 2 ft. Specific types of ponds may have other minimum depths.

Gloyna (1971) provides many recommendations on the design of pond appurtenances.

## Dikes and Liners

Cells may share common-wall dikes (Wastewater Committee, 1997). The dikes should be at least 8 ft wide on top for vehicular traffic and have both inner and outer side slopes no steeper than 1 vertical on 3 horizontal. Inner side slopes should be no flatter than 1 vertical on 4 horizontal. The minimum freeboard is 3 ft. The dikes shall be constructed of impervious soil and compacted to 90% standard proctor density. The dikes may not contain or rest on vegetation, peat, roots, or other unstable material. Prior to filling, dikes shall be covered with 4 in. topsoil and seeded with perennial, low-growing, spreading grasses from the outside toe to 2 ft above the pond bottom on the interior slope. Additional riprap may be required on the inner slope to prevent wave erosion and on the outer slope to prevent flood damage from nearby streams or lakes. Riprap on the inner slope should extend from 1 ft above the high water level to 2 ft below the low water level.

Dikes should be inspected regularly for erosion damage by wind, waves, and runoff and for damage by burrowing animals and long-rooted plants (Gloyna, 1971).

Pond bottoms should be flat, impervious, and compacted to at least 90% standard proctor density. Unless the pond is permitted for groundwater recharge, most jurisdictions will require an impermeable pond liner. Typically, the water permeation through the bottom should be less than 500 gal/ac/day.

## Piping

Inlet, outlet, and transfer piping should be noncorroding, provided with adequate access for cleaning and maintenance, and valved to permit flow control. Ductile iron is preferred by some jurisdictions, but the usual materials of sanitary sewerage construction (concrete and fired clay) are acceptable. Antiseepage collars should be provided wherever the piping pierces a liner. All piping must be protected from settlement, erosion, and ice.

Influent piping should discharge below the lowest operating level of the pond to avoid nuisance. If the inlet pipe is horizontal, it may lie on the top of the pond bottom, but it should be supported, especially at the discharge end, and provided with a discharge apron at least 2 ft square to protect the liner. A vertical upwards discharge is preferred. This may be a simple bend or discharge bell, but it needs a supporting structure to avoid damage to the liner. Although traditional designs favor single inlet pipes, multiple discharge ports or manifolds give better distribution of the flow across the pond and are preferred. This is less of an issue in mixed, mechanically aerated ponds, but it is a serious problem in anaerobic ponds, which normally treat sludges that accumulate around the inlet pipe's exit. The point of discharge should be located about one-third of the way from the upstream end of the point to the outlet.

Transfer piping should be located so as to minimize short-circuiting. This is usually accomplished by placing a given cell's inlets and outlets catcorner. The discharge should be horizontally or vertically upward, and discharge aprons should be provided to avoid bottom erosion. The pipe should be supported.

Outlet piping is usually submerged and located at least 10 ft horizontally from the toe of the dike and at least 2 ft above the pond bottom. Ponds that stratify may be equipped with multilevel takeoffs. For constant discharge and/or shallow warm ponds, floating surface overflow takeoffs may be installed. All outlet piping should be equipped with scum baffles and should be valved to prevent the receiving water from entering the pond at flood.

Ponds should be equipped with emergency spillways to prevent dike failure by overtopping during extreme inflow events, diversion (bypass) valves and pipes, drainage valves and pipes, and provisions for water level control.

### Pond Loadings and Effluent Quality

The areal BOD loading rate largely determines the performance of all waste stabilization ponds naturally aerated by photosynthesis and oxygen uptake from the atmosphere. The effluent BOD<sub>5</sub> data from a large number of ponds fall below the following envelope (Aguirre and Gloyna, 1970):

$$C_{\text{BOD}_e} = 55 + 0.21w_a - 45e^{-0.038w_a} \quad (11.142)$$

where  $C_{\text{BOD}_e}$  = the total effluent BOD<sub>5</sub> (mg/L)

$w_a$  = the areal BOD<sub>5</sub> loading rate (lb BOD<sub>5</sub>/ac/day).

The data were collected from tertiary (maturation), facultative, and anaerobic ponds. Performance data on specific processes are given below.

### Kinetic Models

Ponds are sometimes modeled assuming first-order decay of BOD and coliform. Common flow models are (a) for *each cell* in mixed-cells-in-series,

$$\frac{C_e}{C_o} = \left( \frac{1}{1 + k\tau_1} \right) \quad (11.143)$$

and (b) for *each cell* in plug-flow-with-dispersion,

$$\frac{C_e}{C_o} = \frac{4a \cdot \exp\left[\frac{1}{2}(1+a)Pe\right]}{(1+a)^2 \cdot \exp[aPe] - (1-a)^2} \quad (11.144)$$

where  $a = \sqrt{1 + (4kD_a/u^2)}$ , (dimensionless)

$C_e$  = the effluent concentration (mg/L or cells per 100 mL)

$C_o$  = the influent concentration (mg/L or cells per 100 mL)

$D_a$  = the axial dispersion coefficient (m<sup>2</sup>/s)

$k$  = the first-order decay rate (per s)  
 $L$  = the length of the cell (m)  
 $Pe$  = the turbulent Peclet number (dimensionless)  
 $= uL/D_a$   
 $Q$  = the total flow through the cell ( $m^3/s$ )  
 $u$  = the horizontal velocity (m/s)  
 $V_1$  = the volume of one cell ( $m^3$ )  
 $\tau_1$  = the detention time of one cell (s)  
 $= V_1/Q$

Axial dispersion is strongly influenced by wind and should be determined *in situ* by tracer studies.

### BOD Decay Rate

The Wastewater Committee (1997) recommends a BOD<sub>5</sub> decay rate of 0.28 per day at 20°C and 0.14 per day at 1°C. Other authorities make similar recommendations (Reed et al., 1995). The Streeter–Phelps theta value for BOD decay is about 1.036.

### Coliform Decay Rate

Pathogen decay is believed to be the result of predation, starvation, settling, and ultraviolet light irradiation during the daytime. The decay rates of bacteria in ponds is not well established, but it seems clear that highly loaded ponds are less effective than lightly loaded ponds (Gloyna, 1971). Oakley et al. (2000) recommend the following coliform decay rates:

#### *Heavily loaded facultative ponds:*

$$k = 0.48 \times 1.18^{T-20} \quad (11.145)$$

#### *Facultative ponds:*

$$k = 0.90 \times 1.04^{T-20} \quad (11.146)$$

#### *Maturation ponds:*

$$k = 0.81 \times 1.09^{T-20} \quad (11.147)$$

where  $k$  = the coliform decay rate (per day)  
 $T$  = the pond temperature (°C).

### Effluent Solids Removal

The effluent BOD of any pond consists almost entirely of suspended heterotrophic bacteria and algae. If required, effluent suspended solids removal can be achieved by intermittent sand filtration or land application.

### Mechanically Aerated Ponds

Mechanically aerated ponds are extended aeration plants. There is no sludge recycle, although secondary clarification to capture the net solids production is usual. Detention times are on the order of 1 month, and oxygen is supplied by surface aeration or by submerged air diffusers. Aerated ponds are widely used in the treatment of wastewaters at isolated industrial facilities.

### Configuration

Mechanically aerated ponds systems generally consist of the following:

- Preliminary treatment to remove grit, debris and rags to prevent nuisance and clogging of aerators, pumps, and weirs
- The aerated pond
- Secondary clarification without solids recycle
- Disinfection

The pond must consist of at least two aerated cells-in-series followed by a final polishing cell with volume equal to at least 30% of the combined volume of the aerated cells. The depth should be 10 to 15 ft.

### Aerators

Aerators must provide oxygen and mixing. The minimum mixing energy for low-speed mechanical aerators may be estimated by Rich's equation (Metcalf & Eddy, Inc., 2002):

$$P_v = 0.004X + 5; \quad X \leq 2,000 \text{ mg/L} \quad (11.148)$$

where  $P_v$  = the required mixing power (kW per 1000 m<sup>3</sup>)  
 $X$  = the MLSS (mg/L).

Equation (11.148) includes a safety factor of 1.25 to 1.5. The threshold mixing power for solids suspension is about 1.5 to 1.45 kW per 1000 m<sup>3</sup>.

Closely spaced aerators interact, reducing oxygen transfer and bottom scour. Widely spaced aerators produce unmixed, unaerated dead zones. The maximum spacing should be 0.8 m per kW of installed power but not more than 75 m (Metcalf & Eddy, Inc., 2002). For 75 kW aerators, a minimum spacing of 70 ft and a maximum water depth of 15 ft are recommended (Price, Conway, and Cheely, 1973). Surface aerators in deep basins should be equipped with bottom impellers or draft shrouds and downward-mixing surface impellers should be installed between the aerators, or both.

### Kinetics

Equation (11.143) should be applied separately to each of the aerated cells (Wastewater Committee, 1997).

### Facultative Ponds

Because of their simplicity, facultative ponds are the most common municipal pond systems. Facultative ponds depend upon natural surface aeration and photosynthesis for oxygen. The sediments and the water layer adjacent to them are usually anaerobic.

Facultative ponds are classified as *controlled-discharge*, in which the rate of discharge is a fixed, constant rate, or *flow-through*, in which there is no control of the discharge rate.

### Configuration

In all cases, the system configuration includes at least the following:

- Preliminary treatment to remove grit, debris and rags to prevent nuisance and clogging of piping, pumps, and weirs
- The pond
- Disinfection

Both controlled-discharge and flow-through facultative ponds should consist of two equal-volume *primary* cells-in-series followed by a *secondary* cell of equal or greater (preferred) volume (Wastewater Committee, 1997). The maximum depth in the primary cells should be 6 ft, and the minimum operating depth should be 2 ft. The secondary cells should be at least 8 ft deep to promote suspended solids settling.

Gloyna (1976) recommends a depth of 1 m for tropical and subtropical ponds and 1.5 m for temperate ponds with significant seasonal variations in temperature. In either case, the design surface area is for the operating depth of 1 m.

## Areal BOD Loading Rate

In controlled-discharge systems and flow-through systems, the annual average areal BOD<sub>5</sub> loading rate on the primary cells should be limited to 15 to 35 lb BOD<sub>5</sub>/ac/day at the mean operating depth (Wastewater Committee, 1997). Detention time in controlled-discharge systems should be 180 days for the volume between the minimum and maximum operating depths. In flow-through systems, the detention time must be at least 90 to 120 days. Longer detention times are needed in cold climates.

In flow-through systems, the primary cells should be designed to maximize BOD removal at peak flow.

## Empirical Models

An empirical adjustment of the Gloyna–Espino (1969) formula for facultative ponds yields the following (Reed et al., 1995):

$$V = 0.035QC_{\text{BOD}_0} \cdot 1.099^{(35-T)/250} \quad (11.149)$$

where  $C_{\text{BOD}_0}$  = the influent BOD<sub>5</sub> (mg/L)

$I$  = the mean solar radiation (Langley's)

$Q$  = the influent sewage flow rate (m<sup>3</sup>/d)

$T$  = the average pond temperature for the coldest month (°C)

$V$  = pond volume (m<sup>3</sup>)

About 85 to 95% BOD removal can be expected for facultative ponds having this volume. In the case of domestic wastewaters, temperature is the most important factor in pond design, and the design temperature should be for the coldest month of the year.

## Sulfides

Anaerobic bacteria in the pond sediments reduce sulfate to sulfide, which is toxic to algae at concentrations above about 6 mg/L. This reduces the oxygen supply to the heterotrophic population, which in turn, reduces BOD removal efficiency and produces nuisance odors. This usually becomes a problem when the influent sulfate concentration exceeds 500 mg/L. The sulfide concentration can be estimated by the following (Gloyna and Espino, 1969):

$$S_{\text{S}^{2-}} = (0.000118w_a + 0.00166\tau + 0.0553)S_{\text{SO}_4^{2-}} \quad (11.150)$$

where  $S_{\text{S}^{2-}}$  = the pond sulfide concentration (mg/L)

$S_{\text{SO}_4^{2-}}$  = the influent sulfate concentration (mg/L)

$w_a$  = the ultimate BOD areal load (lb BOD<sub>u</sub>/ac/day)

$\tau$  = the pond detention time (days)

The sulfide concentration can be reduced by reducing the areal BOD<sub>u</sub> loading rate, which has the effect of increasing the pond area.

## Bottom Sludge

Sludge and grit accumulate on the bottom of facultative ponds. The amount of sludge at any time can be estimated from the following:

$$m_s = \frac{R_{\text{SS}_0}}{k_s} (1 - e^{-k_s t}) \quad (11.151)$$

where  $k_s$  = the first-order organic sludge decay rate (per day)

$\approx 0.002 \times 1.35^{T-20}$  (Gloyna, 1971)

$m_s$  = the mass of organic sludge in the primary cells (kg)

$R_{\text{SS}_0}$  = the settleable solids input rate in the influent wastewater (kg/day)

$t$  = the age of the sludge deposit since the last cleaning (days)

Municipal sewage transports about 95 g VSS/cap/day (Table 8.14). Perhaps two-thirds is settleable. A significant portion of the inorganic fraction of the sewage sludges is intracellular matter and is solubilized during decay and exits in the pond discharge. The grit content of municipal sewage is highly variable between cities, ranging from 0.4 to 10.5 ft<sup>3</sup>/mil gal. (Joint Task Force, 1992). Grit does not decay but merely accumulates.

### Insects

Mosquitoes are a potential seasonal problem and disease vector in facultative ponds (Gloyna, 1971). They may be controlled by the removal of any emergent vegetation from the pond, as such vegetation provides breeding sites.

### Maturation Ponds

The principle purpose of maturation (tertiary, polishing) ponds is the elimination of pathogens and parasites. They are placed last in the process train, and they treat the stable effluents of activated sludge plants, trickling filters, and other ponds. The maturation pond effluent may be disinfected.

Gloyna (1971) recommends maturation pond detention times of 7 to 10 days and depths of 1 m. Leffel et al. (1977) recommend that maturation ponds be constructed as three equal-volume cells-in-series, each cell having a detention time of 5 days. They also observe that the effluent will contain substantial algal concentrations unless the depths are 3 to 8 ft and the detention times are reduced to 48 hr/cell.

### Coliform Decay Rate

An apparent first-order decay rate for fecal coliform that incorporates ultraviolet sunlight irradiation may be approximated by the following (Mayo, 1989):

$$k = k_d + \frac{k_s I_o (1 - e^{-k_e H})}{k_e H} \quad (11.152)$$

$$\approx 0.108 + 0.579 \times 10^{-3} \times \frac{I_o}{H}; \text{ at } 26 \text{ to } 34 \text{ }^\circ\text{C}; r = 0.68$$

where  $H$  = the mean depth of the pond (m)  
 $I_o$  = the average daily solar irradiation at the pond surface (cal/cm<sup>2</sup>)  
 $k$  = the overall bacterial die-off coefficient (per day)  
 $k_d$  = the dark bacterial die-off coefficient (per day)  
 $k_e$  = the light extinction coefficient in the pond (per m)  
 $k_s$  = the specific solar die-off coefficient (cm<sup>2</sup>/day/cal)

Some additional data on coliform decay in ponds are given above in the “Coliform Decay Rate” section.

### Insects

Mosquitoes are a potential seasonal problem and disease vector in maturation ponds (Gloyna, 1971). They may be controlled by the removal of any emergent vegetation from the pond, as such vegetation provides breeding sites. Maturation ponds also may be stocked with top feeding minnows to prey on the larvae. Midges may also develop in lightly loaded ponds.

### Anaerobic Ponds

Anaerobic ponds are designed to treat high-strength wastes and function as anaerobic digesters. Methane production is the key objective, as it minimizes odors and yields a useful by-product. At larger installations, it may be practicable to collect the methane produced. The usual expectation is about 12 ft<sup>3</sup> of total gas (1 atm, 32°F) per day per population equivalent, of which 65% by vol will be methane. The liquid effluent from anaerobic ponds generally contains about 100 to 250 mg/L of BOD<sub>5</sub>, and the overall BOD<sub>5</sub> removal efficiency is about 50 to 60% (Aguirre and Gloyna, 1970; Dinges, 1982).

## Configuration

Anaerobic ponds are almost always constructed as the first stage of a two-stage anaerobic/facultative (or aerated) pond system.

Most anaerobic ponds are uncovered, and areal loading rates need to be high enough to minimize oxygen penetration. A loading of about 90 lb BOD<sub>5</sub>/ac/day will limit oxygen penetration to the top 2 ft of the pond, and a loading of 130 lb BOD<sub>5</sub>/ac/day will limit it to the top foot (Oswald, 1968). The usual loading is 150 to 1000 lb BOD<sub>5</sub>/ac/day (Aguirre and Gloyna, 1970).

The liquid detention time in the supernatant should be kept relatively short, say 2 to 4 days, to minimize the top surface area. Liquid detention times longer than 4 days result in lower BOD removal (Dinges, 1982).

Pond depths should be 4.5 to 6.0 m (Dinges, 1982). An allowance should be made for sludge solids storage. Domestic wastewaters produce about 0.04 ft<sup>3</sup> of raw wet primary sewage sludge per cap per day and 0.07 ft<sup>3</sup> of raw wet waste activated sludge per cap per day (Fair and Geyer, 1954).

The active sludge digestion volume in an unmixed tank can be approximated as follows (Fair and Geyer, 1954):

$$V = \left[ Q_f - \frac{2}{3}(Q_f - Q_d) \right] t_d \quad (11.153)$$

where  $Q_d$  = the volumetric flow rate of digested sludge (m<sup>3</sup>/s)  
 $Q_f$  = the volumetric flow rate of fresh sludge (m<sup>3</sup>/s)  
 $t_d$  = the required digestion time (s)  
 $V$  = the active sludge digestion volume in the digester (m<sup>3</sup>)

Additional volume must be allowed for digested sludge storage and for supernatant formation. In general, the volume of digested wet primary sewage sludge is about one-fourth the volume of fresh primary sewage sludge, and the volume of a mixture of digested primary and waste activated sludge is about 40% of that of the raw mixture (Fair and Geyer, 1954). The solids' retention time required for substantially complete digestion, assuming adequate seeding, is about 75 days at 50°F, 56 days at 60°F, and 42 days at 70°F.

Some meatpacking wastes are relatively warm because of hot water use in cleanup operations, and the decay of high-strength wastes releases significant metabolic energy. Pond designers attempt to conserve this heat by minimizing total surface area, top, sides, and bottom. Small top areas also minimize oxygen transfer into the pond, which permits the accumulation of methanogenic organisms.

Below 15°C, and for a pond loading of 425 lb BOD<sub>5</sub>/ac/day, the gas production rate in anaerobic ponds is about 500 ft<sup>3</sup> (1 atm, 32°F) per ac per day (Oswald, 1968). Above 15°C, the gas production rate increases nearly linearly up to a rate of 6600 ft<sup>3</sup> (1 atm, 32°F) at 23°C at the same loading.

## Kinetics

The required detention time of anaerobic ponds may be correlated with the input and output BOD concentrations by the following (Gloyna, 1971):

$$\tau = \frac{\frac{C_{\text{BOD}_o} - 1}{C_{\text{BOD}_e}}}{k \left( \frac{C_{\text{BOD}_e}}{C_{\text{BOD}_o}} \right)^n} \quad (11.154)$$

where  $k$  = an empirical rate constant (per s)  
 $n$  = an empirical exponent (dimensionless).

In tropical countries, a rate constant of about 6 per day and an exponent of about 4.8 have been reported.

## Insects and Odors

Floating scum and manure may support houseflies and stable flies, which are both nuisances and disease vectors. Scum and manure should be broken up and submerged by hosing, the pond should be covered to permit gas collection, or both (Gloyna, 1971).

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## 11.5 Land Application

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A wide variety of land application schemes and objectives have evolved, including the following:

- Irrigation of food and fiber crops, sod farms, tree farms, pasture, and golf courses for tertiary wastewater treatment, commercial crop and animal production and recreation
- Wetlands for tertiary wastewater treatment and wildlife conservation
- Rapid infiltration of tertiary effluents for groundwater recharge and storage
- Bioretention facilities for storm water runoff interception and treatment
- Overland flow for secondary wastewater treatment

Crop irrigation, wetlands, and overland flow are described below.

All land treatment schemes require at least the following:

- Preliminary treatment to remove debris, rags and grit to prevent nuisance and clogging of piping and pumps
- Primary settling and skimming to remove settleable solids and scum

- A storage/treatment pond
- A wastewater distribution system
- The wastewater application area
- A drainage collection system
- Disinfection
- Ancillary equipment and storage sheds, influent/effluent monitoring facilities, all-weather roads, fencing, and buffer strips

Most land application systems also incorporate secondary pretreatment for organic matter removal and stabilization. A commonly used secondary treatment process is the facultative pond, which also serves as a flow-storage device when wastewater cannot be applied to land.

Most jurisdictions also enforce one or more of the following restrictions:

- A flat prohibition on site runoff, which requires water storage during freezing or wet weather and the determination of allowable infiltration and percolation rates
- Limitations on the scheduling of wastewater applications and the kinds of crops and animals raised to protect consumers, workers, and animals from infection
- Limitations on the kinds of crops irrigated and pretreatment schemes tailored to meet the needs of specific crops, especially regarding salts
- Limits on the nitrogen fluxes into the underlying groundwater, which may control crop types and usually requires drainage systems for the interception of the percolating, treated wastewater
- Buffer strips and restrictions on application methods to minimize odor nuisance and pathogen and parasite spread by aerosols

The *stabilized, disinfected* sludges produced by modern wastewater treatment plants also are often (if not usually) disposed of on land. The sludges are first treated to stabilize them and destroy pathogens and parasites, and they must be monitored for a variety of metals and other contaminants. The 503 Rule for sludge disposal on land is described below.

## **Crop Irrigation (Slow-Rate Infiltration)**

Crop irrigation is most widespread in arid areas, where the water is of major economic value. In these applications, the water requirement is usually seasonal, during summer drought, and the nutrient benefit of the wastewater is negligible. The health of consumers, plant workers, and animals, which have direct contact with the crop and/or irrigated area, is a major concern. Designers also must consider crop management as well as the quality of the drainage water produced.

There are two general types of crop irrigation (Pettygrove et al., 1984; Reed, Crites, and Middlebrooks, 1995). In Type I Slow-Rate Infiltration, the focus is on crop production, and crop yield (and land utilization) is maximized. Salinity control is a major concern. In Type II Slow-Rate Infiltration, cropping is necessary but secondary. Land use is limited to the minimum required for the allowable water flux or nitrate flux.

### **Treatment Mechanisms**

Soils filter and flocculate particulate organics, which the soil flora and faunae then consume or decompose. The flora and faunae include a wide variety of bacteria, fungi, insects, insect larvae, “bugs,” and “worms.” Bacteria and fungi directly absorb and metabolize soluble organics. These decomposition processes produce the slowly degrading humus that increases the soil’s water retention, air and water permeability, resistance to compaction, nutrient availability, and anion and cation exchange capacity (Taylor and Ashcroft, 1972). Some of the particulate organics are pathogens and parasites, and soil microbes prey upon and destroy them.

Nitrifying bacteria oxidize ammonia nearly completely to nitrate. Some of the nitrate enters the groundwater, some of it is denitrified and lost as nitrogen gas, and the crop consumes some of it.

Phosphates and many metal ions are adsorbed or ion-exchanged by soil clay minerals and humus.

## Drainage Water Quality

Crop irrigation systems are capable of high levels of wastewater treatment, and the typical drainage water composition is as follows (Sopper and Kardos, 1973):

- $BOD_5 = 1 \text{ mg/L}$
- $COD = 10 \text{ mg/L}$
- $NO_3^- = \text{TKN} - \text{crop uptake}$
- $PO_4^{3-} = 0.5 \text{ mg/L}$
- Coliform count = very low, but not potable

## Inorganic Contaminants

Wastewaters containing substantial concentrations of metals and metalloids are not suitable for land irrigation treatment (Table 8.5). Cadmium, copper, molybdenum, selenium, and zinc are accumulated by many crops, making them toxic to consumers (Wright, 1973). Boron, copper, nickel, and zinc are also toxic to many crops and may cause process failure. Metals entering conventional wastewater treatment plants tend to concentrate in the waste sludges, and EPA regulations prohibit the application of metal-enriched sludges to irrigation plots.

## Pathogens and Parasites

There is a significant health hazard to using sewage on crops and pasture, and all states require that sewage be disinfected prior to use. A common requirement is a fecal coliform count less than 1000/100 ml (Table 8.5).

In general, wastewaters should not be used to irrigate crops eaten raw, like lettuce, celery, and tomatoes or on root crops, like carrots and potatoes, even if they are cooked. Treated wastewater may be used to irrigate grasses like alfalfa, corn, oats, rye, and wheat. In these permissible cases, the edible portion of the plant is exposed to sunlight, which is germicidal, and it is usually cooked.

## Nutrients

If the crop needs additional nitrogen and phosphorus beyond that in the soil and wastewater, commercial fertilizers and stabilized or composted sewage sludge solids should be used. Dried primary solids typically are 1 to 1.5% by wt phosphorus as  $P_2O_5$  and 1 to 3.5% by wt nitrogen as  $NH_4NO_3$ ; secondary solids are typically 3 to 5% by wt  $P_2O_5$  and 4 to 7% by wt  $NH_4NO_3$ ; and digested solids are 0.5 to 0.7% by wt  $P_2O_5$  and 1 to 4% by wt  $NH_4NO_3$  (Babbitt and Baumann, 1958). Sludges also release nutrients more slowly, which makes nutrient capture by the crop more likely.

## Salinity

Deflocculation often occurs when the sodium/calcium ratio becomes large, and clays disperse. The usual design criterion is the sodium adsorption ratio (SAR):

$$SAR = \frac{\{Na^+\}}{\sqrt{\frac{1}{2}(\{Ca^{2+}\} + \{Mg^{2+}\})}} \quad (11.155)$$

where SAR = the sodium adsorption ratio (dimensionless)

$\{Ca^{2+}\}$  = the calcium ion concentration (meq/L)

$\{Mg^{2+}\}$  = the magnesium ion concentration (meq/L)

$\{Na^+\}$  = the sodium ion concentration (meq/L)

An SAR less than 3 indicates a low risk of permeability loss, and an SAR of 8 or more indicates a high risk (Table 8.5; Taylor and Aschcroft, 1972).

High total salt concentrations also inhibit many crops. An electrical conductivity less than 0.75 mmhos per cm (25°C) minimizes the deleterious effects on most crops (Table 8.6; Taylor and Aschcroft, 1972).

Hard, alkaline wastewaters also exhibit large increases in pH upon evapotranspiration, which also injures some crops. This is usually explained by a carbonate/bicarbonate equilibrium shift (Stumm and Morgan, 1996):

$$\frac{m_{\text{CO}_3^{2-}}}{\left(m_{\text{HCO}_3^-}\right)^2} = \frac{K_1}{K_2 K_H p_{\text{CO}_2} V} \quad (11.156)$$

where  $K_1$  = the first acid dissociation constant of carbonic acid (mol/L)  
 $K_2$  = the second acid dissociation constant of carbonic acid (mol/L)  
 $K_H$  = the Henry's law constant for carbon dioxide and water (mol/L Pa)  
 $m_{\text{CO}_3^{2-}}$  = the mass of carbonate in solution (mol)  
 $m_{\text{HCO}_3^-}$  = the mass of bicarbonate in solution (mol)  
 $p_{\text{CO}_2}$  = the partial pressure of carbon dioxide in contact with the wastewater (Pa)  
 $V$  = the volume of wastewater (m<sup>3</sup>)

As the solution volume evaporates, the dissolved carbonate-to-bicarbonate ratio rises, and so does pH.

Besides the pH effect, evaporation of hard, alkaline waters precipitates calcium carbonate, which may reduce soil permeability and cement it, and increases the SAR.

Most of these effects can be offset by proper management of irrigation water, providing excess water so that the various concentration limits are not exceeded due to evapotranspiration and so that salts may be flushed from the soil.

### Storage Requirements

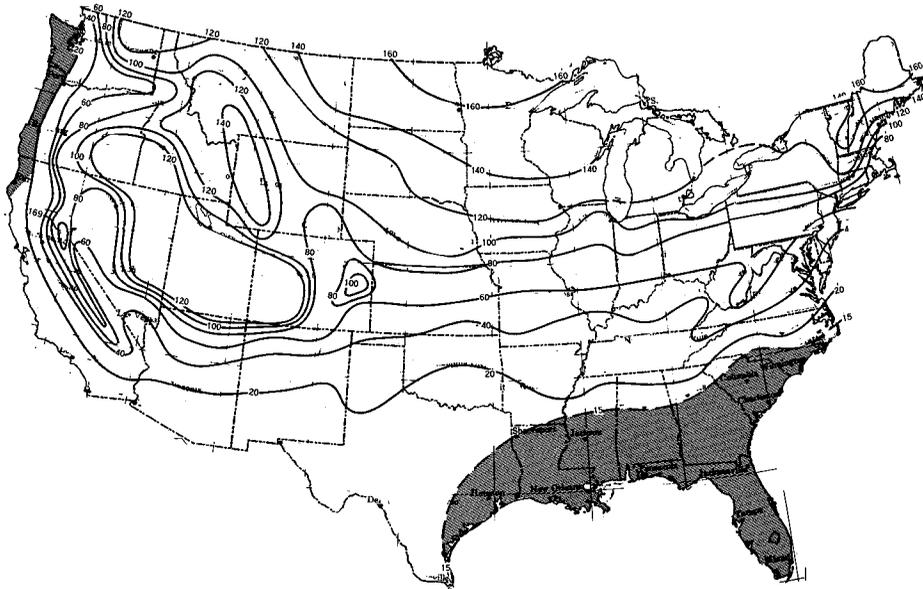
Public health, crop management, and weather determine the wastewater application schedule and, consequently, the required storage volume. In general, irrigation should cease about 1 month before harvest or grazing and when workers are in field so that pathogens and parasites might die away. Usually, there is no irrigation before cultivating, planting, and harvest, because the ground must be dry and firm for equipment. Irrigation also ceases when the ground is frozen or saturated by rainfall in order to prevent runoff.

The computer programs EPA-1, Cold Climates, and EPA-3, Moderate Climates, estimate the storage days required by cold weather (Whiting, 1976). Cold climates have an average January air temperature less than 40°F. This includes all states along the Canadian border and Alaska except for Ohio, Pennsylvania, and southern New York. Given climate data, EPA-1 classifies all the days in each annual cold period from November in 1 year through April of the next as favorable for irrigation or unfavorable. An unfavorable day has a mean daily air temperature of less than 32°F plus a snow depth on the ground of 1 in. or more plus a precipitation of more than ½ in. Favorable days are otherwise. During an unfavorable day, all the wastewater flow must be stored. On a favorable day, all of that day's wastewater flow plus a volume equal to ½ day's flow in storage (if any) may be applied. EPA-1 applies the procedure to each cold season of record, and assigns the annual storage requirements a return period using the log Pearson Type III method. [Figure 11.10](#) is a map of the isochrones with a 10-year return period.

EPA-3 is a modification of EPA-1 that applies roughly between 35 and 42 °N latitude. In general, the sewage application rate and storage requirement are determined by the crop's consumption water use, which can be obtained from agricultural extension services.

The EPA-2, Wet Climate Model, estimates storage days due to wet weather (Whiting, 1976). In this case, storage is required when the soil is saturated down to plow depth. Any excess of precipitation or irrigation over the amount of evapotranspiration will produce runoff. This model typically applies to the coastal regions of Washington, Oregon, and northern California, southern North Carolina, South Carolina, Florida, the Gulf Coast states, and southeast Texas.

Water climate storage volumes are best estimated using a water budget for each month of a wet year with a return period of 10 to 25 years (Crites et al., 1977):



Note: Shaded areas indicate storage controlled by prolonged wet spells.

FIGURE 11.10 Storage days due to frozen ground and snow cover.

$$\text{precipitation} + \text{wastewater} = \text{evapotranspiration} + \text{percolation to ground water} + \text{runoff} \quad (11.157)$$

Any runoff would go to storage. The usual units are cm or in. of water over the application area.

### Irrigation

Irrigation imitates natural precipitation. Operators apply wastewater to a selected field for several hours to a day, and then the field is rested for a few days to a couple of weeks. Intermittent flooding and draining of the soil pushes out old air ahead of the water wave and draws in new air behind it. While one field is resting, the operators shift the wastewater to another. The duration of the resting period is determined by the field's infiltration capacity, the current weather, the crop's drought tolerance, and the rate of biodegradation of the applied organic and inorganic matter. All of these vary significantly throughout the growing season and even from one field to the next at the same site. Consequently, engineers must design storage volumes, distribution networks, and control structures that give operators the greatest possible flexibility in irrigation quantity and timing.

In the case of Type I Slow-Rate Infiltration systems (crop production), the volume of wastewater applied during a single irrigation is the difference between the available water capacity (AWC) and the management-allowed deficit (MAD) (Pettygrove et al., 1984). The available water capacity is the difference between the soil's field capacity and the crop's wilting point (usually about 15 bar tension). For a clay-loam or a silty-clay-loam, the AWC may be 2 to 2.5 in. of water per ft of soil depth, whereas for a sand or fine sand, it may be as little as 0.5 to 1.0 in./ft. The MAD is usually 30 to 50% of the AWC of the whole root zone. It varies by stage of crop growth. The Department of Agriculture's Soil Conservation Service and Cooperative Extension can provide details for local sites.

The depth of water needed to replenish the AWC is as follows (Pettygrove et al., 1984):

$$d_{irr} = \frac{f_{mad} d_{awc} (1 + f_{lch})}{E_u} \quad (11.158)$$

where  $d_{awc}$  = the depth of available water in the crop's root zone (cm)  
 $d_{irr}$  = the depth of wastewater applied during one irrigation (cm)  
 $E_u$  = the unit application efficiency (decimal)  
 $f_{lch}$  = the fractional leaching requirement (decimal)  
 $f_{mad}$  = the fractional management-allowed deficit (decimal)

The unit application efficiency accounts for the fact that the application is not perfectly uniform and that some water is lost to mist and evaporation. The efficiency depends on the kind of distribution system and ensures that all parts of the field get the desired minimum amount of water. The leaching fraction is the amount of water required to control salt buildup. The leaching volume becomes the drainage water that must be captured. The timing of the irrigation is by soil moisture measurements. Pettygrove et al. (1984) provide details.

In the case of Type II Slow-Rate Infiltration systems (wastewater treatment), the depth applied is the net of evapotranspiration plus percolation less precipitation (Pettygrove et al., 1984):

$$d_{irr} = (d_{et} - d_{ppt} + d_{prc}) \frac{t_{max}}{30} \quad (11.159)$$

where  $d_{et}$  = the month's total evapotranspiration rate (cm)  
 $d_{prc}$  = the month's total percolation out of the root zone (cm)  
 $d_{ppt}$  = the month's total precipitation (cm)  
 $t_{max}$  = the maximum time between irrigations (days)

The analysis is done month by month throughout the year. The net difference between evapotranspiration and precipitation should be the 90% exceedance value, i.e., a relatively wet month. The allowable monthly percolation is a soil property.

## Scheduling

For design purposes, the maximum time between irrigations can be estimated from the following (Pettygrove et al., 1984):

$$t_{max} \approx \frac{f_{mad} d_{awc}}{r_{et\ max}} \quad (11.160)$$

where  $r_{et\ max}$  = the 10-year return maximum evapotranspiration rate for the month (cm/d). The actual duration of any irrigation must be less than  $t_{max}$ , and Pettygrove et al. (1984) recommend that the irrigation rate be high enough for it to be completed in less than 75% of  $t_{max}$ .

## Irrigation Methods

Wastewater distribution types include sprinkler, drip pipe, surface flooding (border strip), and ridge-and-furrow distribution. Design recommendations and details are given by Booher (1974), Crites et al., 1977; Dillon, Hiler, and Vittetoe (1972), Fry and Gray (1971), Hart (1975), Jensen (1980), Pair (1975), Pettygrove et al. (1984), Schulbach and Meyer (1979), and the Soil Conservation Service (1974 to date).

Sprinkler systems (spray irrigation) are subdivided into fixed-in-place, hand-moveable, end-pull tractor-mobile, side-wheel self-mobile, center pivot self-mobile, and traveling gun self-mobile. The sprinklers usually have impact nozzles mounted on risers or horizontal, elevated pipes, but some orchard systems use perforated pipes laid on the ground. Fixed-in-place systems are not restricted as to ground slope or crop, and the minimum application rate needed to activate the impact-nozzles is only about 0.12 cm/h (Reed, Crites, and Middlebrooks, 1995). Hand-moveable systems can be placed on grades up to 20%

and require minimum application rates of 0.25 cm/hr to operate the nozzles. Side-wheel, center pivot, and traveling gun self-mobile systems are hydraulically self-mobile and require maximum ground slopes of 10 to 15% and minimum application rates of 0.25 to 0.75 cm/hr to operate. Side-wheel and center pivot self-mobile systems also have maximum crop height limitations. Sprinklers generate aerosols that contain wastewater contaminants. Most jurisdictions require and specify buffer zones and/or mist barriers to prevent contamination of surrounding properties. A common requirement is at least 400 ft of buffer or 200 ft of buffer with a mist barrier of trees (Powell, 1976).

Drip pipe systems generally have no slope restrictions and very low minimum application rates (0.05 cm/hr). The orifices are subject to clogging and require prescreening. They may not be suitable where salinity control is an issue.

Surface flooding (border strip) requires relatively flat fields with grades from 0.5 to 7% and low to moderate infiltration rates for even distribution. The crops must be able to withstand inundation. The minimum widths are set by farm equipment requirements. The maximum widths depend on the uniformity of distribution and are usually less than 100 ft. Clay soil strips are generally longer, up to 1200 ft, and sandy soil strips are generally shorter, up to 300 ft. Unit wastewater flows range from 0.01 to 0.04 ft<sup>3</sup> per ft width per sec for clay-loam soils to 0.03 to 0.11 ft<sup>3</sup> per ft width per sec for sandy-loam soils. Application depths range from 7 to 8 in. for clayey-loam and clay soils to 4 to 5 in. for sandy-loam and sandy soils.

Straight ridge-and-furrow systems are restricted to relatively flat fields (less than 0.25% grade), but furrows that follow contours can be laid across slopes up to 8%. Furrow length and spacing for uniform distribution depend on local soil conditions and may require annual adjustment. Sandy soils (with high infiltration rates) require closely spaced, short furrows, whereas clay soils (with low infiltration rates) need widely spaced, long furrows.

### Drainage and Tailwater Collection Systems

Subsoil drains are necessary to intercept percolating water to minimize contamination of the groundwater resource and, especially in arid areas, to prevent salinization of the root zone by the upward migration of the groundwater. The top of the groundwater table should be at least 1 to 1.5 m deep (Pettygrove et al., 1984). In the San Joaquin valley, the pipe spacing is estimated from the following:

$$L = \sqrt{\frac{4k(H^2 - d^2)}{q}} \quad (11.161)$$

- where
- $d$  = the diameter of the drainpipe (m)
  - $H$  = the height of the groundwater table above the drainpipe invert (m)
  - $k$  = Darcy's permeability coefficient (m/s)
  - $L$  = the distance between pipe center lines (m)
  - $q_{perc}$  = the rate of percolation water flow (m<sup>3</sup>/m<sup>2</sup> s)

The design of the subsurface drains depends on the hydraulic conductivity of the soil (down to at least 10 ft) and the rate of irrigation. The conductivity is measured *in situ* using prescribed borehole techniques. Drains are normally placed below the groundwater table.

Surface flooding, graded border, and ridge-and-furrow distribution systems also require tailwater capture and recycling systems. In arid areas, the tailwater may require treatment prior to reuse.

### Area Requirement for Water

In the long run, Eqs. (11.156) and (11.158) apply to all systems. Setting runoff to zero and rewriting to make the area explicit, one gets,

$$A_{ro} = \frac{V_{irr}}{d_{et} + d_{perc} - d_{ppt}} \quad (11.162)$$

where  $A_{ro}$  = the total irrigation area required to prevent runoff ( $m^2$ )  
 $V_{irr}$  = the total irrigation volume for some specified period ( $m^3$ ). The area required to prevent runoff should be compared to the area required for nitrogen control; the larger is the required field area.

For the whole year,  $V_{irr}$  is the total annual volume of wastewater less any storage/treatment pond evaporation. However, because of shutdowns for weather, public health protection, cultivation, harvest, etc., this volume must be applied in less than 1 year, perhaps in as little as a few months in severe northern climates. Once the schedule of nonapplication periods is determined, each application period should be analyzed using Eq. (11.162). During such a period, all the wastewater that arrives on-site during the period plus any wastewater taken out of storage should be considered. The analysis should be done for a relatively wet year, say a 10-year rainfall.

### Area Requirement for Nitrogen

Because many rural areas depend on untreated groundwater for domestic use, nitrate contamination of the groundwater under the irrigation area is an issue. The required land area for nitrogen removal depends upon the wastewater nitrogen inputs, crop uptake, denitrification, ammonia volatilization, and removal by percolation:

$$\frac{V_{irr} C_{No}}{A_N} = m_{cp} + m_{den} + m_{vol} + \frac{f_{prc} V_{irr} S_{NO_3 prc}}{A_N} \quad (11.163)$$

where  $A_N$  = the area required for nitrogen control ( $m^2$ )  
 $C_{No}$  = the total nitrogen content of the applied wastewater including TKN and nitrate/nitrite, if any (g N/ $m^3$ )  
 $f_{prc}$  = the fraction of the applied wastewater becoming drainage (decimal fraction)  
 $m_{cp}$  = the crop nitrogen uptake (g N/ $m^2$ )  
 $m_{den}$  = the amount of nitrate/nitrite lost to denitrification (g N/ $m^2$ )  
 $m_{vol}$  = the amount of ammonia lost to volatilization (g N/ $m^2$ )  
 $S_{NO_3}$  = the allowable nitrate concentration in the drainage water (g N/ $m^3$ )  
 = 10 mg N/L, for public health reasons  
 $V_{irr}$  = the volume of wastewater applied during the growing season ( $m^3$ )

The area required for nitrogen control should be compared to the area required to prevent runoff; the larger is the required field area. The fraction of the applied wastewater that becomes drainage depends on the irrigated area, so an iterative calculation is needed. The losses due to crop uptake can be determined from the agricultural literature. Some values are given in [Table 11.15](#). Denitrification and volatilization are usually taken to be some fraction (roughly 20%) of the applied nitrogen load.

### Area Requirement for Organic Matter

Organic loading is not usually a consideration in municipal wastewater treatment, because the applied wastewater is stable pond effluent and because the amounts are small compared to the soil humus (Loehr et al., 1979). Manure application rates are typically on the order of 10 to 20 tons/acre/year, and food processing, winery, brewery, and dairy wastewaters have been applied at rates up to 300 lb BOD<sub>5</sub>/ac/day without nuisance as long as the fields were allowed to dry between applications and the soil remained permeable (Loehr et al., 1979; Reed, Crites, and Middlebrooks, 1995).

Oils, grease, and fats need special consideration. Industrial oils are toxic to soil microorganisms at levels around 2 to 4% by wt and to sugar beet, rape, lawn grasses, yams, carrots, turnips, and radishes at soil levels of 0.5% by wt (Overcash and Pal, 1979). Fats, oils, and greases also greatly reduce soil permeability and may accumulate in the soil and form surface mats in severe northern climates.

The usual remedy for soil clogging is rest plus cultivation to break up the surface mats and soil clogs and to redistribute the oils. Sometimes application areas benefit by seeding with soil from established fields that have been receiving oily and fatty wastes.

**TABLE 11.15** Crop Nutrient Requirements per Growing Season<sup>a,b</sup>

Crop	lb N/ac	lb P/ac
Forage		
Alfalfa (legume)	155–220	—
Bermuda grass	480–600	—
Canary grass	226–359	40
Red clover (legume)	77–126	—
Sweet clover (legume)	158	—
Fescue	275	29
Hay	130	—
Field Crops		
Barley	63	—
Corn	155	31
Cotton	66–100	17
Oats	53	—
Soybeans	94–113	22
Wheat	50–76	20
Forest		
Deciduous <5 yr	100	—
Deciduous >5 yr	30–50	—
Evergreen <5 yr	60	—
Evergreen >5 yr	20–30	—

*Sources:*

<sup>a</sup> Crites, R., et al. 1977. *Process Design Manual for Land Treatment of Municipal Wastewater*, EPA 625/1-77-008, US EPA, US ACE, US DA, p. B-5.

<sup>b</sup> Powell, G.M. 1976. *Land Treatment of Municipal Wastewater Effluents: Design Factors II*, Environmental Protection Agency, Technology Transfer, Cincinnati, OH, Table IV-2, p. 42.

## Overland Flow

### Configuration

Overland flow systems may be regarded as trickling filters that use grass as the biomass support packing. They are alternatives to nonnitrifying activated sludge, trickling filters, and ponds on sloping ground, where excavation costs are excessive, or on relatively impermeable soils with infiltration rates less than 2 in./hr. An overland flow system consists of the following (Crites et al., 1977):

- Preliminary treatment to remove grit, rags and debris to prevent nuisance and clogging of pumps, piping, and weirs
- Primary settling and scum removal or fine screening
- A distribution box to control which slope is in use
- A top-of-slope side-overflow distribution channel or, more usually, a sprinkler system
- A grassy slope
- A toe-of-slope collection channel
- Disinfection
- Ancillary equipment and storage sheds, influent/effluent monitoring facilities, all-weather roads, fencing, and buffer strips

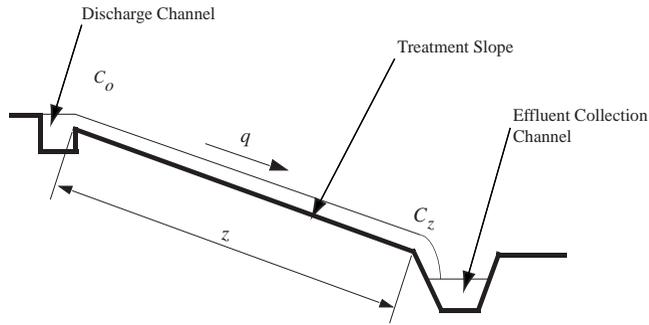


FIGURE 11.11 Overland flow definition sketch.

Overland flow systems sometimes are used to treat pond effluents, but they are poor at algae removal. The crop, almost always grass, must be harvested regularly, and the slopes must permit the safe use of the usual kinds of agricultural equipment.

Wastewater is usually applied using permanent (buried distribution piping) solid-set sprinklers, but perforated pipes and side-overflow weirs are acceptable if energy dissipaters are used to prevent erosion of the slope. If a side-overflow weir is used, wastewater should enter the distribution channel via some sort of manifold system. The momentum of the influent wastewater entering one end of a cross-flow, side-overflow channel makes uniform distribution nearly impossible unless the influent flow is controlled to be constant, which allows a single adjustment of the side-overflow weir. If the influent flow varies, low discharges concentrate near the inlet end of the distribution channel, and high flow discharges concentrate near the dead end of the channel.

In order to get good wastewater distribution, the slopes must be planar. The dip should be between 2 and 6% (to prevent slope erosion) and perpendicular to the top-of-slope distribution system. The strike should be parallel to the distribution system. Slopes are typically 45 to 60 m long if sprinklers are used, and 30 to 45 m long if perforated pipes or side-overflow weirs are used (Reed, Crites, and Middlebrooks, 1995). The minimum slope length is 20 m.

### Operation

Wastewater application is intermittent for 5 to 7 days per week and 6 to 12 hr (typically 8 hr) per day (Crites et al., 1977; Reed, Crites, and Middlebrooks, 1995). During the summer growing season, as much as 8 in. pond effluent per week may be applied, but the application rate should be reduced by half during the spring and fall, and applications should cease during the winter. In the case of screened sewage or primary effluent, and if “crop” is not used for food or pasture, a typical hydraulic loading is 2.5 to 8 in./week. The lower rate should be used if the slope is greater than 6% and in cold weather. Lagoon effluents may be loaded at 6 to 16 in./week. Algal removal may be inadequate.

### Kinetic Model

BOD<sub>5</sub> removal along the slope is approximately exponential (Fig. 11.11; Reed et al., 1984):

$$\frac{C_{\text{BOD}}(z) - C_{\text{BODmin}}}{C_{\text{BOD}_0}} = a \cdot \exp(-kz/q^n) \quad (11.164)$$

where  $a$  = an empirical coefficient dependent on  $q$  (dimensionless)

$C_{\text{BODmin}}$  = the minimum achievable BOD<sub>5</sub> (mg/L)  
 $\approx 5$  mg/L

$C_{\text{BOD}_0}$  = the BOD<sub>5</sub> of the applied wastewater (mg/L)

$C_{\text{BOD}}(z)$  = the BOD<sub>5</sub> of the wastewater a distance  $z$  from the distribution channel (mg/L)

$k$  = an empirical rate constant [(m<sup>3</sup>/m h)<sup>n</sup>/m]

- $q$  = the hydraulic load ( $\text{m}^3$  of flow per m of width per hr)
- $n$  = an empirical exponent dependent on  $q$  (dimensionless)
- $z$  = the distance from the distribution weir *along the slope* (not its horizontal projection) (m)

The values of  $a$ ,  $k$ , and  $n$  are not well known, and they are reported for experimental sites during the summer and fall with air temperatures ranging from 48 to 95°F. For screened sewage, the semilogarithmic intercept  $a$  declines from about 0.65 to 0.5, and the slope  $k/q^n$  increases from 0.018 to 0.033 as the hydraulic load declines from 0.45 to 0.10  $\text{m}^3/\text{m h}$ . In the case of primary effluent, the intercept  $a$  falls from about 0.7 to about 0.35, and the slope  $k/q^n$  varies between 0.02 and 0.025, as the hydraulic load declines from 0.37 to 0.1  $\text{m}^3/\text{m h}$ . There is no data for slope lengths beyond 40 m.

### Slope Area and Drainage

The slope area (true, not projected horizontally) is as follows:

$$A_s = \frac{V_{app} z}{qt_{app}} \quad (11.165)$$

- where  $A_s$  = the true, not projected, slope area ( $\text{m}^2$ )
- $q$  = the hydraulic load ( $\text{m}^3$  wastewater per m strip width per hr)
- $t_{app}$  = the duration of the application (hr)
- $V_{app}$  = the applied wastewater volume, including the daily flow and any wastewater taken from storage ( $\text{m}^3$ )
- $z$  = the true, not projected, length of the strip (m)

The collection channel should be designed for the maximum 24-hr storm with a 25-year return period plus the design maximum week sewage flow. If the channels are earthen or grassy, the channel velocities should be less than 1.5 ft/sec to avoid scour. The collection channel should have a freeboard at maximum flow of 8 in.

The treated effluents of high-strength industrial wastewaters may be recycled to dilute the contaminant concentrations in the applied flow.

### Performance

The probable best effluents in the southern U.S. at an application rate of 4 in./week are as follows:

- $\text{BOD}_5 \approx 8 \text{ mg/L}$
- $\text{SS} \approx 8$  to 16  $\text{mg/L}$
- P removal  $\approx 30$  to 40  $\text{lb/ac yr}$ , if cropped
- $\text{N} \approx 3 \text{ mg/L}$  total (no nitrification, cropped)
- Coliform  $\approx 90$  to 95% removal

Rainfall does not impair  $\text{BOD}_5$  removal efficiency, because the increased flow increases the water depth on the slope, which limits the decrease in hydraulic detention time; raindrops increase water film turbulence and mass transfer to the biofilm; and rain dilutes the effluent. Consequently, cessation of operation is not required during rainy periods.

Organic loads ought to be limited to less than 100  $\text{kg BOD}_5/\text{ha/day}$  to avoid anaerobic conditions in the wastewater film and soil (Reed, Crites, and Middlebrooks, 1995).

### Constructed Treatment Wetlands

A wetland is a community of hydrophytic plants growing in a hydric soil. Kadlec and Knight (1996) summarize the various kinds of wetlands. At least seasonally, hydric soils are saturated with water, often submerged and anaerobic. Anaerobiosis may proceed even when the soils dry out, if the organic matter content is high enough. Hydric soils may contain little or much organic matter, be acidic to neutral in

pH, have a wide variety of hydraulic conductivities and water-holding capacities, have variable nutrient content, and have variable cation exchange capacity. Typical hydrophytic plants include floating and rooted mosses (*Fontinalis*, *Sphagnum*), rooted ferns (*Belchnum*, *Osmunda*, *Salvinia*), conifers (*Pinus*), floating, submerged, and emergent herbs (*Lemna*, *Elodea*, *Sagittaria*), shrubs (*Lyonia*), and deciduous trees (*Acer*, *Quercus*, *Populus*) (Kadlec and Knight, 1996). An initial planting of reeds (*Phragmites*), bulrushes (*Scirpus*), cattails (*Typha*), or sedges is sometimes recommended (Crites et al., 1988; Reed, Crites, and Middlebrooks, 1995).

Although constructed treatment wetlands are intended primarily for wastewater treatment, they are also valued and often justified by their roles in wildlife conservation and flood mitigation. Design and operation must preserve these values, too, which limit the kinds of wastewaters that may be applied.

Leachate contamination of groundwater is a concern in some areas, and constructed wetlands are often lined.

Constructed treatment wetlands require a minimum pretreatment of screening, grit removal, settling, scum removal, and disinfection. It is especially important to remove settleable solids that may form noisome sludge deposits near the inlets. Many wetlands receive secondary effluents and function as effluent polishing processes. Other wetlands are used to treat acid mine drainage.

Wetlands may be classified as free water surface (FWS), in which water flows over the top of the soil, or subsurface flow (SSF), in which water flows through the soil. They are usually riparian and subject to flooding, so embankments must be protected from erosion. In either case, the principal agent of wastewater stabilization is the microbial biofilm attached to various surfaces. SSF systems generally have more surface area for biofilm attachment than do FWS systems and can accept higher organic and nitrogen loadings. In general, SSF wetlands are preferred for the treatment of domestic wastewaters because of the following:

- They are somewhat more efficient than FWS systems due to their larger biofilm attachments areas.
- They suppress insect vectors.
- They exclude human and animal contact with the wastewater.

### Free Water Surface (FWS) Wetlands

FWS wetlands consist of three to four squarish cells-in-series (Watson, 1989). Tracer studies show that the flow regime is complex and not adequately described as plug-flow-with-dispersion; the actual flow regime strongly affects the apparent reaction rate values (Kadlec et al., 1993). FWS bottoms are generally on impermeable soils, but a liner system may be required to prevent groundwater contamination.

Oxygen is available to the nitrifying bacteria from atmospheric exchange and algal photosynthesis. Many aquatic macrophytes also have a tubular airway system that directly connects their (aerobic) roots to the atmosphere. These airways may also transport a small amount of oxygen to the otherwise anaerobic soil.

The depth of flow is typically 6 to 24 in., and the areal requirement for the hydraulic load is typically 20 to 75 ac/mgd for secondary treatment and 10 to 50 ac/mgd for secondary effluent polishing. The hydraulic detention times range from a few to several days (Watson, 1989).

Organic loading is not normally considered a major factor in FWS wetland design. Observed organic loads range from 18 to 116 kg BOD<sub>5</sub>/ha/day, with a recommended maximum of 110 kg/ha/day (Watson, 1989).

The correlation between inlet and outlet BOD<sub>5</sub>s is as follows (Kadlec and Knight, 1996):

$$C_{\text{BOD}_e} = 0.173C_{\text{BOD}_o} + 4.7; \quad (r^2 = 62; n = 440) \quad (11.166)$$

where  $C_{\text{BOD}_e}$  = the total effluent BOD<sub>5</sub> (mg/L)  
 $C_{\text{BOD}_o}$  = the total inlet BOD<sub>5</sub> (mg/L).

The data are highly scattered, and the standard error of the outlet BOD is 13.6 mg/L. Inlet BODs ranged from 10 to 680 mg/L, and outlet BODs ranged from 0.5 to 227 mg/L. The effluent BODs appear to be independent of hydraulic loading between 0.27 and 25.4 cm/day.

Nitrogen removal in FWS wetlands proceeds by ammonia volatilization, biomass incorporation and burial, and nitrification/denitrification. The first-order mass transfer coefficients for the areal reaction rates at 20°C and their theta values are as follows (Kadlec and Knight, 1996):

**Mineralization of organic nitrogen:**

$$k_{\min} = 17 \times 1.05^{T-20} \quad (11.167)$$

**Nitrification of ammonia:**

$$k_{\text{nit}} = 18 \times 1.04^{T-20} \quad (11.168)$$

**Denitrification of nitrite/nitrate:**

$$k_{\text{dn}} = 35 \times 1.09^{T-20} \quad (11.169)$$

where  $k_{\text{dn}}$  = the areal denitrification rate coefficient (m/yr)  
 $k_{\min}$  = the areal mineralization rate coefficient (m/yr)  
 $k_{\text{nit}}$  = the areal nitrification rate coefficient (m/yr)

The minimum effluent organic nitrogen concentration for a FWS wetland is about 1.5 mg N/L. The effluent ammonia nitrogen concentration is generally less than 0.1 mg N/L. The Water Environment Federation recommends the following concentration estimators (Reed, Crites, and Middlebrooks, 1995):

$$\ln(S_{\text{NH}_e}) = 0.688 \ln(S_{\text{NH}_o}) + 0.655 \ln(q) - 1.107 \quad (11.170)$$

$$C_{\text{Ne}} = 0.193 C_{\text{No}} + 1.55q - 1.75 \quad (11.171)$$

where  $C_{\text{Ne}}$  = the effluent total nitrogen concentration, ammonia plus nitrate (mg N/L)  
 $C_{\text{No}}$  = the influent total nitrogen concentration, ammonia plus nitrate (mg N/L)  
 $q$  = the areal hydraulic load (cm/d)  
 $S_{\text{NH}_e}$  = the effluent ammonia nitrogen concentration (mg N/L)  
 $S_{\text{NH}_o}$  = the influent ammonia nitrogen concentration (mg N/L)

Initially, in a new wetland, phosphorus removal occurs by adsorption to minerals on the surface of the soil. Later, removal is by incorporation into biomass and settling. The settling material is liable to scour and discharge during storm events. In the long run, phosphorus removals average 30 to 60%, and the first-order mass transfer rate coefficient for the areal P removal rate is about 10 m/yr (Reed, Crites, and Middlebrooks, 1995).

Suspended solids removal declines as the hydraulic load increases. At loads between 0.4 and 75 cm/d, the effluent SS concentration of a FWS wetland may be approximated as follows (Reed, Crites, and Middlebrooks, 1995):

$$X_{\text{SSe}} = (0.1139 + 0.00213q) X_{\text{SSo}} \quad (11.172)$$

where  $q$  = the hydraulic load (cm/d)  
 $X_{\text{SSe}}$  = the effluent SS concentration (mg/L)  
 $X_{\text{SSo}}$  = the influent SS concentration (mg/L)

**Subsurface Flow (SSF) Wetlands**

SSF wetlands consist of a constructed permeable bottom, often crushed limestone or gravel, through which the wastewater flows. Under the assumption that the root zone is less than 30 cm deep, the usual

bed depth is 45 to 60 cm (Kadlec and Knight, 1996). The bed must be able to transmit the expected maximum wastewater flow and the design precipitation in the face of degradation of the bed's hydraulic conductivity without surface flooding. Hydraulic conductivities for the selected bed material must be measured to ensure adequate hydraulic capacity. A liner system is needed to keep the wastewater flow out of the groundwater.

The principal source of oxygen in SSF beds is the atmosphere. Under average design conditions, the beds are not completely saturated and contain a vadose zone open to the atmosphere. Aeration is improved by intermittent or varying hydraulic loading, which flushes air out of and into the vadose zone. The effluents of SSF wetlands contain very little oxygen, generally less than 1 mg/L. In some jurisdictions, the effluents must be reaerated before discharge.

The areal requirement for the hydraulic load is 5 to 20 ac/mgd for secondary and polishing wetlands (Watson, 1989).

For SSF wetlands, the correlation between inlet and outlet BOD<sub>5</sub>s is as follows (Kadlec and Knight, 1996):

***Soil beds with reeds (Phragmites):***

$$C_{\text{BOD}_e} = 0.11C_{\text{BOD}_o} + 1.87; \quad (r = 0.74; n = 73) \quad (11.173)$$

***Gravel beds:***

$$C_{\text{BOD}_e} = 0.33C_{\text{BOD}_o} + 1.4; \quad (r^2 = 0.48; n = 100) \quad (11.174)$$

where  $C_{\text{BOD}_e}$  = the total effluent BOD<sub>5</sub> (mg/L)  
 $C_{\text{BOD}_o}$  = the total influent BOD<sub>5</sub> (mg/L).

Again, the data are highly scattered. For the soil/reed systems, the inlet BODs ranged from 1 to 330 mg/L, and outlet BODs ranged from 1 to 50 mg/L. For the gravel systems, the influent BODs ranged from 1 to 57 mg/L, and the effluent BODs ranged from 1 to 36 mg/L. The effluent BODs appear to be independent of hydraulic loading between 0.27 and 25.4 cm/d.

Nitrogen removal in SSF wetlands proceeds by ammonia volatilization, biomass incorporation and burial, and nitrification/denitrification.

The first-order reaction coefficients for the mass transfer rate coefficients at 20°C and their theta values are as follows (Kadlec and Knight, 1996):

***Mineralization of organic nitrogen:***

$$k_{\text{min}} = 35 \times 1.05^{T-20} \quad (11.175)$$

***Nitrification of ammonia:***

$$k_{\text{nit}} = 34 \times 1.04^{T-20} \quad (11.176)$$

***Denitrification of nitrite/nitrate:***

$$k_{\text{dn}} = 50 \times 1.09^{T-20} \quad (11.177)$$

where  $k_{\text{dn}}$  = the areal denitrification rate coefficient (m/yr)  
 $k_{\text{min}}$  = the areal mineralization rate coefficient (m/yr)  
 $k_{\text{nit}}$  = the areal nitrification rate coefficient (m/yr)

These rates are about twice those of FWS systems because of the greater biomass in SSF.

Suspended solids removal declines as the hydraulic load increases. Between 0.4 and 75 cm/day, the effluent SS concentration of a SSF wetland may be approximated as follows (Reed, Crites, and Middlebrooks, 1995):

$$X_{SSe} = (0.1058 + 0.0011q)X_{SSi} \quad (11.178)$$

where  $q$  = the hydraulic load (cm/d)  
 $X_{SSe}$  = the effluent SS concentration (mg/L)  
 $X_{SSi}$  = the influent SS concentration (mg/L)

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## 11.6 Bioremediation and Composting

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Gas treatment processes and compost piles consist of permeable, moist solid packings that support aerobic biodegradation on their surfaces. In the case of gas treatment, the substrate is contained in the contaminated gas stream passing through the packing voids. Contaminated soils have a thin layer of adsorbed material on their surfaces. In the case of composting, the substrate is the pile itself.

### Treatment of Gases and Soils

A *biopile*, *biofilter*, *bioscrubber*, *biosparger*, *biovent*, *biologically activated foam*, or *soil filter* is an aerobic packed-bed bioreactor used for the biological treatment of contaminated gases and/or soils. They may be regarded as a subclass of trickling filter. Bioventing, biosparging, or soil venting typically refers to the *in situ* treatment of contaminated soils. Biopiles are piles of contaminated soils that are excavated and transported to a nearby treatment facility. The other terms refer to systems used to remove odors and biodegradable volatile organic carbon (VOC) compounds from process off-gases.

### Water

The ideal water content for soil systems is between 10 and 20% by wt or 70 to 95% of field capacity, and the extreme limits are 5 to 30% by wt (von Fahnestock et al., 1996). It is necessary that the packing be unsaturated to permit air permeation. In the case of biopile for contaminated soil remediation, the water is added as the soil is being prepared for placement on the pad or liner. In the case of biofilters and bioscrubbers, the raw gas is humidified.

In some applications, the gas flow may dry the packing, and provision for water addition is needed. This can be done by placing drip pipes on top of the packing. However, the packing should not be flooded or wetted to the point where gas flow is impeded. Off-gases from liquid and/or sludge processing may be saturated with water vapor and may deposit water in the packing. In these cases, predrying the gas stream may be necessary.

### Nutrients

It is generally necessary to seed the packing with a suite of acclimated microbes and to provide nutrients and trace minerals. Nutrients may be available from compost and soils used as packing in bioscrubbers. [Tables 11.2](#) and [11.3](#) can be used as guides to the quantities needed. The usual rules of C:N less than 20:1 and C:P less than 100:1 apply. The carbon content is that of the contaminated gas. In the case of biopile for contaminated soil remediation, the nutrients are added as the soil is being prepared for placement on the pad or liner.

### pH

The pH of the sorbed and capillary water should be somewhat above 7. The maximum range for biodegradation is 6 to 9.

## Poisons

Tables 11.4 and 11.5 may be used as guides.

## Bioventing and Biosparging

Bioventing is the aeration of vadose and capillary zone subsurface soils *in situ* (Leeson and Hinchee, 1995). In biosparging, the air is injected into the contaminated saturated zone, and biodegradation occurs in the overlying vadose and capillary zones (O'Brien and Gere, 1995). In either case, the purpose is to stimulate the biodegradation of soil contaminants by the indigenous microbes. Soil venting is usually intended merely to strip VOCs from the vadose or capillary zone with the expectation that gas treatment will be provided at the ground surface.

The system generally consists of a system of air injection and extraction wells and soil gas monitoring wells. An extracted gas treatment process will be required if the extracted gas contains high concentrations of VOC. The details of the installation depend critically on the three-dimensional position of the spilled contaminants, the local distribution of soil types, and the location of the vadose, capillary, and saturated zones, all of which must be determined by borehole studies. It may be necessary to model the pattern of gas flows through the soils in order to optimize well placement.

Groundwaters with high iron and manganese concentrations may plug the soil around the air injection diffusers.

If the organic contaminant is stripped into the flowing air rather than degraded *in situ*, it will be necessary to collect and treat the off-gas (Leeson and Hinchee, 1995). In general, substances with vapor pressures near atmospheric are stripped rather than degraded *in situ*. These substances are typically the lighter fractions of gasoline like butanes and propylene. If the organic contaminant does not volatilize, it may not become available to the microbes. In general, this includes alkanes larger than undecane, polynuclear aromatic hydrocarbons, phenols, and long-chain alcohols and organic acids. Typically, these are found in diesel fuel. The ideal subjects of bioventing and biosparging have vapor pressures on the order of  $10^{-3}$  to  $10^{-1}$  atm. These substances are commonly found in jet fuels.

Some soils have very low water contents, and bioventing may further dry them. Although biodegradation has been observed at moisture contents as low as 5%, it is very slow, and optimal decay rates occur at moisture contents of 15 to 25% (Leeson and Hinchee, 1995). Water can be added to shallow soils by surface irrigation, but deeper soils require trenches, infiltration galleries, or dry wells (O'Brien and Gere, 1995).

If needed, nutrients can be added to soils by the same techniques used to add water, however, most soils do not require nutrient addition.

Biodegradation half-lives range from days for metabolic intermediates, to months for polynuclear aromatic hydrocarbons, to years for chlorinated hydrocarbons (Leeson and Hinchee, 1995).

## Biopiles

In the case of bioremediation of VOC-contaminated soils, the soil becomes the biopile, and the biopile is a temporary construct. The complete system will include the following (von Fahnestock et al., 1996):

- The biopile treatment unit
- A contaminated soil storage pad
- A processing area to grind and screen the soil and to add microbial seed (soil from previous biopiles), bulking agents (if needed), nutrients, and water
- Access roads, equipment maneuvering areas, equipment and monitoring sheds, and fencing

Typical biopile dimensions for contaminated soil are 50 ft wide  $\times$  60 ft long  $\times$  4.5 ft high. They contain about 500 yd<sup>3</sup> of soil. Storage piles are somewhat larger. The fenced site area will be about 1 ac. Soil is moved by front-end loader and mixed. Typically, contaminated soil requires about 4 months of aeration for complete removal of the VOCs.

The biopile sits on a concrete pad or on compacted clayey soil plus an impermeable liner (40 to 60 mil HDPE). If necessary, pile leachate can be collected by sloping the pad/liner toward one end and installing a perforated PVC drain and sump. In any case, the surrounding soil should be graded and bermed to

collect any leachate and to prevent surface runoff from entering the pile. Leachate and runoff may require collection and treatment.

Directly on top of the pad or liner is a layer of clean graded sand about 1 ft deep. A manifold of perforated PVC pipe rests on the clean sand layer and is covered by a protective 8 in. of pea gravel. The manifold legs are 2 to 4 in. in diameter and about 30 ft long. They are valved to permit air distribution control. Each leg is capped at one end and connected to an unperforated header at the other. The manifold extracts soil air by vacuum or pumps atmospheric air into it.

In vacuum extraction systems, the off-gases are collected for further treatment or recycling. Also, because the soil gas is water-saturated at the pile temperature, vacuum extraction systems generally require water knockout systems and knockout water collection and treatment.

In dry climates, additional water can be supplied by a drip pipe on top of the soil.

At the top of the biopile, a 20 mil HDPE cover excludes precipitation. The upper liner requires a tie-down system. The biopile also contains a soil gas sampling system so the extent of treatment can be monitored.

### **Biofilters and Bioscrubbers**

Biofilters and bioscrubbers are used to treat the off-gases from other processes. The earliest proposal of biofilters for odor control was made by Bach in 1923, and applications of the concept began in the 1950s (Leson and Winer, 1991).

A biofilter or bioscrubber consists of the following (Leson and Winer, 1991):

- A raw gas collection system
- A raw gas heat exchanger to cool or heat the gas
- A raw gas humidifier
- A filter box that contains a packing supported over an air diffusion system and a supplementary water addition system
- An off-gas monitoring system

If the foul gas contains a mixture of hydrogen sulfide and VOCs, a two-stage system is recommended (Joyce and Sorensen, 1999). The first stage contains the autotrophic, sulfide-oxidizing *Thiobacillus* spp., and the second stage contains a mixed heterotrophic biofilm that oxidizes the VOCs. *Thiobacillus* produces sulfuric acid, and the stages are biologically separated by biofilm pH.

The removal rates for different classes of substances decline along the series alcohols > esters > ketones > aromatics > alkanes (Deshusses and Johnson, 2000). In general, substances with high dimensionless Henry's law coefficients or high octanol-water partition coefficients resist removal. Biofilters are not recommended if the following obtain (Nash and Siebert, 1996):

- There are economical, simple alternatives.
- Contaminant recovery and recycling are practicable.
- The contaminants are not fully known.
- The contaminants are insoluble in water.
- The contaminants have sterically hindered structures that resist biodegradation.
- Halogenated organics are present.

The best results to date have been obtained with sulfides. However, sulfuric acid is produced which lowers the biofilm pH unless it is flushed with water.

Biofilters should be sized using the data obtained from pilot studies. Packing depths are typically 1 to 2 m with gas detention times of 1 to 2 min and pressure drops of 15 to 30 cm or more of water (Nash and Siebert, 1996). Typical contaminant unit degradation rates range from 10 to 100 g/m<sup>3</sup>/hr, and gas application rates up to 300 m<sup>3</sup>/m<sup>2</sup>/hr are feasible (Leson and Winer, 1991).

The packing may be any permeable material, including plastic trickling filter media, granular activated carbon, diatomaceous earth, graded sands, soil, composting material, artificial foams (Hun, 1998), etc. A number of proprietary systems are available. Some packings, like compost, are able to hold sorbed and capillary water. Nonabsorbent media like those used in trickling filters require intermittent water application to keep the biofilm wet.

The pores of the wet packing must be large enough to permit gas flow. A  $d_{60}$  greater than 4 mm and an organic matter content of at least 55% are recommended (Leson and Winer, 1991). A porosity of at least 25% or higher is recommended, and 80% is preferred (Leson and Winer, 1991). Bulking agents may be required to maintain the porosity, especially where compost materials are employed. Soils should be free of clay and silt (von Fahnstock et al., 1996).

Raw gases should contain about 95% relative humidity in order to avoid drying of the packing and/or microbes. The raw gas may be humidified by spray humidifiers. The filter may produce leachate. This must be collected for treatment.

Waste gas temperatures should be between 20 and 40°C for optimal results. High temperatures will pasteurize the biofilter.

Some biodegradable contaminants exhibit substrate toxicity, and dilution of the waste gas with ambient air may be needed. VOC concentrations should be less than 3 to 5 g/m<sup>3</sup>.

## Composting

Composting is the aerobic conversion of waste solids to commercially viable humus for soil conditioning. The preferred starting material is yard waste, but properly prepared municipal solid wastes and POTW sludges are acceptable. In normal household and commercial use, individuals will come into close, unprotected contact with the compost. Consequently, it is mandatory that the compost be free of pathogens and parasites, hard sharp objects like broken glass and metals, and nuisances like odors and plastics.

Composting is normally done in open-air windrows on slabs, but enclosed in-tank processing is also practiced. In either case, the composting material may produce odors if aeration is inadequate and may combust spontaneously if water is inadequate. Storage piles of raw materials attract insects, rodents, and larger scavengers, which are nuisances and which may be disease vectors. The composting piles contain large numbers of fungi that are allergens to many people. The finished product should be free of such problems.

## Municipal Solid Waste

The processing of municipal solid waste (MSW) entails the following (Haug, 1993; Hickman, 1999; Skitt, 1972):

- Collection
- Bag breaking
- Hand sorting
- Trommel screening
- Magnetic separation
- Shredding
- Addition as needed of water, wood chips (bulking agent), energy amendment (sawdust), nutrients, pH adjustment, and seed
- Composting pile
- Product screening to recover bulking agent
- Curing pile
- Fine screening and packaging

The collection of MSW is normally done by others and delivered to the site during normal business hours. A storage bin will be needed to facilitate processing.

The MSW is first put through low power flails to break open plastic trash bags and permit further separation. The flailed material is then spread on a picking floor to a depth of less than 1 ft, and pickers pass through the MSW and collect salvage, hazardous material, and noncompostible or inappropriate material. Bag removal can be done at this stage.

The sorted material is then transported by conveyor belt to trommel screens for the removal of dirt, stones, bricks, cans, bottles, and other small dense debris. For MSW, the usual trommel opening is 4 in. Because of MSW variability, trommels must have adjustable speeds and inclines. Very long trommels with multiple cascades are required. About 50 to 60% removal of dirt, etc., can be expected.

Screening is followed by magnetic separation of ferrous metals. The metal is usually recycled to mini steel mills and must be nearly free of paper. Generally, two to three stages of magnetic separation are needed, perhaps with intermediate air classification.

The remaining MSW is then shredded by hammer mill. Generally, 1.5 to 3 in. pieces are preferred for windrows, and 0.5 to 1.5 in. sizes are preferred for in-tank processing. Shredding must be done after screening and magnetic separation, because shredders typically imbed broken glass, dirt, and metal into the paper, which makes their subsequent separation impossible and degrades the scrap value.

A number of additives are mixed into the shredded waste as needed. These may include fresh compost for seeding (1 to 5% by wt), water (up to 50 to 60% by wt), a bulking agent (for pile permeability), sawdust (for pile temperature maintenance to >50°C), nutrients (C:N ≈ 30 to 35 by wt; C:P ≈ 75 to 150 by wt) and pH adjustment (7 to 8). If shredded tires are used for bulking, the metal content of the rubber is a concern. The usual bulking agent is 1 to 2 in. wood chips.

The mixture is then placed in windrows on slabs or in tanks. Transfer is done by front-end loader or conveyor belt. Slabs and tanks must have leachate collection systems. Windrows are mixed and turned twice a week by machine, and tanks are mixed by augers. Windrows are aerated by vacuum piping placed on the slab, and tanks have air diffusers. The off-gas should be collected and treated for odor and fungus spore control. The off-gas from windrow vacuum systems also requires water knockouts.

The composting temperature is controlled by the aeration system. The usual air requirement is 10 to 30 scf per lb compost per day. The optimum temperature range for newsprint and other cellulosic wastes is 45 to 50°C; for freshly prepared MSW mixtures it is 55 to 60°C; and for stabilized MSW it is around 40°C (Haug, 1993).

Composting requires about 5 weeks and converts about 40 to 50% of the volatile solids in MSW to humus. The fresh compost is screened to remove bulking agent and transferred to an unaerated, unmixed curing pile, where it is held for about 1 month to cure. A portion of the fresh compost is used to seed the raw MSW. Finally, the cured compost is finely screened to remove all objectionable material and packaged for sale to vendors. The product must be free of pathogens, parasites, metal, glass, ceramics, plastics, and bulking agent.

### **Sewage Sludges and Garbage**

The system configuration for POTW sludges and garbage is nearly the same as for MSW composting (Hay and Kuchenrither, 1990). The initial steps of bag breaking, sorting, screening, magnetic separation, and shredding are not needed. Water, nutrients, and pH adjustments are also not normally needed. However, a bulking agent is always required, and sawdust is often needed to offset excessive moisture in the raw material. Holding times and conversions are the same as for MSW.

### **Material, Heat, and Air Balances**

The wet weight of the raw compost mixture, which is needed to size material handling equipment, slabs, and tanks, is as follows (Haug, 1993):

$$W_m = W_s + W_r + W_b + W_a + W_w \quad (11.179)$$

where  $W_a$  = the wet weight of the energy amendment, usually sawdust (kg)  
 $W_b$  = the wet weight of the bulking agent, usually wood chips (kg)

- $W_m$  = the wet weight of the raw compost mixture (kg)  
 $W_r$  = the wet weight of the recycled fresh compost seed (kg)  
 $W_s$  = the wet weight of the raw waste material (substrate) to be composted, e.g., sewage sludge, garbage, or MSW substrate (kg)  
 $W_w$  = the weight of any water added to the mixture (kg)

The weight of the waste material (substrate) is known, and the weight of the recycled seed compost is proportional to it. The other material weights are determined by considerations of permeability, average moisture content, and pile temperature. The sawdust amendment and the added water are mutually incompatible additions. Water is added only to dry wastes, and sawdust is added only to wet wastes.

In terms of dry weights (total solids, TS), this becomes as follows:

$$\begin{aligned}
 f_{sm}W_m &= f_{ss}W_s + f_{sr}W_r + f_{sb}W_b + f_{sa}W_a \\
 X_m &= X_s + X_r + X_b + X_a
 \end{aligned}
 \tag{11.180}$$

- where
- $f_{sa}$  = the weight fraction of dry solids in the energy amendment (dry wt/wet wt, dimensionless)
  - $f_{sb}$  = the weight fraction of dry solids in the bulking agent (dry wt/wet wt, dimensionless)
  - $f_{sm}$  = the weight fraction of dry solids in the raw compost mixture (dry wt/wet wt, dimensionless)
  - $f_{sr}$  = the weight fraction of dry solids in the recycled compost seed (dry wt/wet wt, dimensionless)
  - $f_{ss}$  = the weight fraction of dry solids in the waste material (substrate) to be composted (dry wt/wet wt, dimensionless)
  - $X_a$  = the dry weight of the energy amendment (kg TS)
  - $X_b$  = the dry weight of the bulking agent (kg TS)
  - $X_m$  = the dry weight of the raw compost mixture (kg TS)
  - $X_r$  = the dry weight of the recycled compost seed (kg TS)
  - $X_s$  = the dry weight of the substrate (kg TS)

The organic matter in the pile is usually reported as volatile solids:

$$\begin{aligned}
 f_{vm}f_{sm}W_m &= f_{vs}f_{ss}W_s + f_{vr}f_{sr}W_r + f_{vb}f_{sb}W_b + f_{va}f_{sa}W_a \\
 X_{vm} &= X_{vs} + X_{vr} + X_{vb} + X_{va}
 \end{aligned}
 \tag{11.181}$$

- where
- $f_{va}$  = the weight fraction of volatile solids in the dry energy amendment (VS/TS, dimensionless)
  - $f_{vb}$  = the weight fraction of volatile solids in the dry bulking agent (VS/TS, dimensionless)
  - $f_{vm}$  = the weight fraction of volatile solids in the dry raw compost mixture (VS/TS, dimensionless)
  - $f_{vr}$  = the weight fraction of volatile solids in the dry recycled compost seed (VS/TS, dimensionless)
  - $f_{vs}$  = the weight fraction of volatile solids in the dry raw waste (VS/TS, dimensionless)
  - $X_{va}$  = the weight of dry volatile solids in the energy amendment (kg VS)
  - $X_{vb}$  = the weight of dry volatile solids in the bulking agent (kg VS)
  - $X_{vm}$  = the weight of dry volatile solids in the raw compost mixture (kg VS)
  - $X_{vr}$  = the weight of dry volatile solids in the recycled compost seed (kg VS)
  - $X_{vs}$  = the weight of dry volatile solids in the raw waste (kg VS)

A portion of the volatile solids is biodegradable:

$$\begin{aligned}
 f_{bm}f_{vm}f_{sm}W_m &= f_{bs}f_{vs}f_{ss}W_s + f_{br}f_{vr}f_{sr}W_r + f_{bb}f_{vb}f_{sb}W_b + f_{ba}f_{va}f_{sa}W_a \\
 X_{bm} &= X_{bs} + X_{br} + X_{bb} + X_{ba}
 \end{aligned}
 \tag{11.182}$$

where  $f_{ba}$  = the weight fraction of biodegradable volatile solids in the energy amendment (kg biod VS/kg VS)  
 $f_{bb}$  = the weight fraction of biodegradable volatile solids in the dry bulking agent (kg biod VS/kg VS)  
 $f_{bm}$  = the weight fraction of biodegradable volatile solids in the dry raw compost mixture (kg biod VS/kg VS)  
 $f_{br}$  = the weight fraction of biodegradable volatile solids in the dry recycled compost seed (kg biod VS/kg VS)  
 $f_{bs}$  = the weight fraction of biodegradable volatile solids in the dry waste (kg biod VS/kg VS)  
 $X_{ba}$  = the weight of dry biodegradable volatile solids in the energy amendment (kg VS)  
 $X_{bb}$  = the weight of dry biodegradable volatile solids in the bulking agent (kg VS)  
 $X_{bm}$  = the weight of dry biodegradable volatile solids in the raw compost mixture (kg VS)  
 $X_{br}$  = the weight of dry biodegradable volatile solids in the recycled compost seed (kg VS)  
 $X_{bs}$  = the weight of dry biodegradable volatile solids in the raw waste (kg VS)

The decay of the biodegradable material consumes oxygen and produces heat and determines the airflow rate.

The biodegradable fraction on any component is closely related to its lignin content and may be approximated by the following (Haug, 1993):

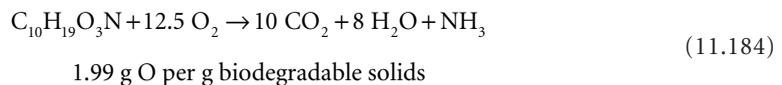
$$f_{bi} = 0.83 - 0.028f_{li} \quad (11.183)$$

where  $f_{bi}$  = the fraction of component  $i$  that is biodegradable (kg biod VS/kg VS)  
 $f_{li}$  = the fraction of component  $i$  that is lignin (kg lignin/kg VS).

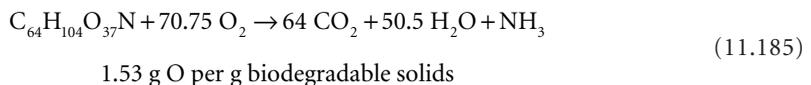
The fraction of the recycled seed that is biodegradable may be assumed to be zero, and any sawdust amendment may be assumed to be 100% biodegradable. Raw wastes are commonly 40 to 50% biodegradable. About 15 to 30% of wood chips are lost due to mechanical breakage in the screening process. The small wood fibers become part of the compost product. The bulking agent also slowly degrades and becomes part of the compost product. At least 5% of a wood-chip bulking agent degrades in each pass through the composting pile.

The air requirement depends on the composition of the waste and any biodegradable additions. Some of the stoichiometric oxidation reactions of typical compost components are as follows (Haug, 1993):

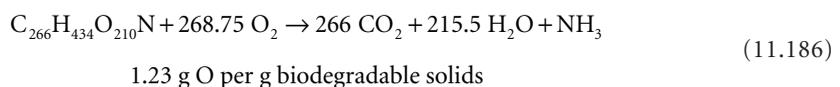
**Combined POTW sludges:**



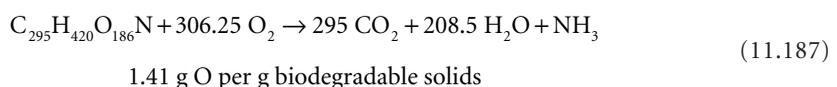
**MSW:**



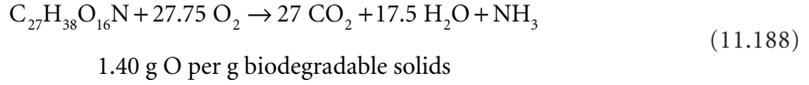
**Mixed paper:**



**Wood:**



**Yard waste:**



The airflow strips water from the compost pile. It may be assumed that exit air is saturated with water vapor at the pile temperature (Haug, 1993):

$$\log_{10} p_{vsat} = -\frac{2238}{T} + 8.896 \quad (11.189)$$

$$\omega_{sat} = \frac{0.6221 p_{vsat}}{p_{atm} - p_{vsat}} \quad (11.190)$$

where  $p_{atm}$  = the total atmospheric pressure, dry air plus water vapor (mm Hg)  
 $p_{vsat}$  = the vapor pressure of water in saturated air at the specified temperature (mm Hg)  
 $T$  = the absolute temperature (K)  
 $\omega_{sat}$  = the saturated specific humidity (kg H<sub>2</sub>O/kg air)

The required airflow for water removal is, therefore, as follows:

water in air = water in raw compost mixture – water in final compost product

$$(\omega_{sat} - \omega) \rho_a Q_a = (1 - f_{sm}) W_m - \frac{(1 - f_{sr})(X_m - X_{bm})}{f_{sr}} \quad (11.191)$$

where  $Q_a$  = the volumetric airflow rate (m<sup>3</sup>/s)  
 $\rho_a$  = the air density (kg/m<sup>3</sup>)  
 $\omega$  = the specific humidity of the ambient air (kg H<sub>2</sub>O/kg air)

At 77°F (25°C) and 70% relative humidity, which are typical summer conditions, air holds 0.014 lb water per lb dry air (23.8 mm Hg). At 131°F (55°C), which is a typical compost pile temperature, saturated air holds 0.115 lb water per lb dry air (118 mm Hg).

The airflow rate also affects the pile temperature. The heat balance for a compost pile is as follows:

$$h_{hiv} = [C_{pa}(T_p - T_a) + \omega_{sat} C_{pv}(T_p - T_a) + \lambda(\omega_{sat} - \omega)] \rho_a Q_a \quad (11.192)$$

where  $C_{pa}$  = the constant pressure-specific heat of dry air (kJ/kg K)  
 $\approx 1$  kJ/kg K (0.24 Btu/lbm °F)  
 $C_{pv}$  = the constant pressure specific heat of water vapor (kJ/kg K)  
 $\approx 1.93$  kJ/kg K (0.46 Btu/lbm °F)  
 $h_{hiv}$  = the lower heating value of the raw waste plus sawdust amendment (kJ/kg)  
 $T_a$  = the ambient air temperature (K)  
 $T_p$  = the compost pile temperature (K)  
 $\lambda$  = the latent heat of evaporation of water (kJ/kg)  
 $\approx 2371$  kJ/kg at 55°C (1019 Btu/lbm)

The lower heating value is required, because water exits the pile as vapor.

The higher heating value of the dry raw compost mixture can be estimated from the modified Dulong formula (Tchobanoglous, Theisen, and Vigil, 1993):

$$h_{hiv} = 145P_C + 610 \left( P_H - \frac{P_O}{8} \right) + 40P_S + 10P_N \quad (11.193)$$

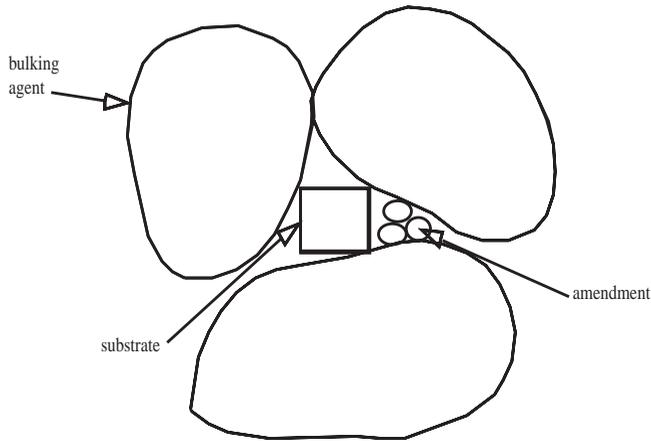


FIGURE 11.12 Free air space in solid waste mixtures.

where  $h_{hhv}$  = the higher heating value of the dry, ash-free biodegradable material in the mixture (Btu/lbm VS)

$P_C$  = the percent weight of carbon in the dry, ash-free mixture (%)

$P_H$  = the percent weight of hydrogen in the dry, ash-free mixture (%)

$P_N$  = the percent weight of nitrogen in the dry, ash-free mixture (%)

$P_O$  = the percent weight of oxygen in the dry, ash-free mixture (%)

$P_S$  = the percent weight of sulfur in the dry, ash-free mixture (%)

The lower heating value is simply the higher heating value less the latent heat of evaporation of the water formed:

$$h_{lhv} = h_{hhv} - 18\lambda f_H \quad (11.194)$$

where  $f_H$  = the weight fraction of hydrogen in the dry, ash-free biodegradable solids (lbm H/lbm VS).

Airflow cools the pile and dries it. Which factor controls depends on the starting condition of the mixture, the desired moisture content of the composted waste, and the maximum pile temperature.

### Free Airspace

Following agricultural practice, the void volume includes substrate water (which the bulking agent may absorb) and free airspace. The volume of the raw mixture is equal to the free airspace volume plus the volume of the individual component particles (Fig. 11.12):

$$\begin{aligned} \text{mix vol} = & \text{free air vol} + \text{waste particle vol} + \text{bulking agent particle vol} \\ & + \text{amendment particle vol} \end{aligned} \quad (11.195)$$

$$\frac{W_m}{\gamma_m} = \epsilon_{fa} \frac{W_m}{\gamma_m} + \frac{W_s}{\gamma_s} + (1 - \epsilon_b) \frac{W_b}{\gamma_b} + (1 - \epsilon_a) \frac{W_a}{\gamma_a}$$

where  $\epsilon_a$  = the void fraction of the amendment (dimensionless)

$\epsilon_b$  = the void fraction of the bulking agent (dimensionless)

$\epsilon_{fa}$  = the free airspace fraction of the raw mixture (dimensionless)

$\epsilon_s$  = the void fraction of the raw waste (substrate) (dimensionless)

= 0, by assumption for wet wastes like sludges

$\gamma_a$  = the wet, bulk density of the amendment, including voids (kg/m<sup>3</sup>)

$\gamma_b$  = the wet bulk density of the bulking agent, including voids (kg/m<sup>3</sup>)

$\gamma_m$  = the wet bulk density of the raw mixture, including voids (kg/m<sup>3</sup>)  
 $\gamma_s$  = the wet bulk density of the raw waste, including voids (kg/m<sup>3</sup>)

Raw wastes like POTW sludges are mostly water, even after dewatering. Therefore, the void volume of sludges is zero. Bulking agents are added to the wet solids to maintain voids for airflow. The free airspace fraction of the mixture,  $\epsilon_{fa}$ , should be at least 20%, and the optimum is 30 to 35%. Absorption of water from the wet raw waste by the bulking agent shrinks the volume of the raw waste in the mix and creates additional free airspace.

The amount of water adsorbed is limited by the absorptive capacity of the bulking agent or the amount of water in the substrate. Ignoring the amendment volume, which is small, this leads to two formulas for the volume of the raw mixture without free water films (Haug, 1993):

***Bulking agent absorptive capacity limited:***

absorbed water = final bulking agent water – initial bulking agent water

$$W_{wabs} = \left( \frac{f_{sb}}{f_{sb}^{\min}} - f_{sb} \right) W_b - (1 - f_{sb}) W_b \quad (11.196)$$

$$W_{wabs} = \left( \frac{f_{sb}}{f_{sb}^{\min}} - 1 \right) W_b$$

$$\frac{W_m}{\gamma_m} = \epsilon_m \frac{W_m}{\gamma_m} + \left[ \frac{W_s}{\gamma_s} - \left( \frac{f_{sb}}{f_{sb}^{\min}} - 1 \right) \frac{W_b}{\rho} \right] + (1 - \epsilon_b) \frac{W_b}{\gamma_b} \quad (11.197)$$

***Substrate surface water limited:***

water absorbed = initial substrate water – final substrate water

$$W_{wabs} = (1 - f_{ss}) W_s - \left( \frac{f_{ss}}{f_{ss}^{\max}} - f_{ss} \right) W_s \quad (11.198)$$

$$W_{wabs} = \left( 1 - \frac{f_{ss}}{f_{ss}^{\max}} \right) W_s$$

$$\frac{W_m}{\gamma_m} = \epsilon_m \frac{W_m}{\gamma_m} + \left[ \frac{W_s}{\gamma_s} - \left( 1 - \frac{f_{ss}}{f_{ss}^{\max}} \right) \frac{W_s}{\rho} \right] + (1 - \epsilon_b) \frac{W_b}{\gamma_b} \quad (11.199)$$

where  $f_{sb}^{\min}$  = the minimum fraction of dry solids in the bulking agent when it is saturated with water (kg dry solid/kg wet solid)

$f_{ss}^{\max}$  = the maximum fraction of dry solids in the waste (kg dry solid/kg wet solid)

$W_{wabs}$  = the water absorbed by the bulking agent (kg)

$\rho$  = the mass density of water (kg/m<sup>3</sup>)

Both formulas are more conveniently written in terms of bulk volume mixing ratios (Haug, 1993):

$$r_{bs} = \frac{\text{bulking agent volume}}{\text{substrate volume}} = \frac{W_b/\gamma_b}{W_s/\gamma_s} \quad (11.200)$$

$$r_{mb} = \frac{\text{mixture volume}}{\text{bulking agent volume}} = \frac{W_m/\gamma_m}{W_b/\gamma_b} \quad (11.201)$$

$$r_{ms} = \frac{\text{mixture volume}}{\text{waste volume}} = \frac{W_m/\gamma_m}{W_s/\gamma_s} \quad (11.202)$$

**Bulking agent absorptive capacity limited:**

$$1 = (1 - \epsilon_m)r_{ms} + \frac{\gamma_b}{\rho} \left( \frac{f_b}{f_b^{\min}} - 1 \right) r_{bs} - (1 - \epsilon_b)r_{bs} \quad (11.203)$$

**Substrate water limited:**

$$1 = (1 - \epsilon_m)r_{ms} - (1 - \epsilon_b)r_{bs} + \frac{\gamma_s}{\rho} \left( 1 - \frac{f_{ss}}{f_{ss}^{\max}} \right) \quad (11.204)$$

For wet substrates like POTW sludges, the mixture is mostly bulking agent; a typical mixture to bulking agent ratio,  $r_{mb}$ , is 1.1 (Haug, 1993).

The design unknown is the bulking agent to waste substrate ratio,  $r_{bs}$ . Both values produced by Eqs. (11.196) and (11.197) are used to calculate the weight of bulking agent required. Then, the amount of water absorbed by the bulking agent is computed for both cases. The smaller amount of absorbed water controls. The volume of the mixture is then estimated using  $r_{mb}$ .

## Kinetics

For garbage and raw POTW sludges, the peak oxygen consumption rates are on the order of 4 to 14 mg O<sub>2</sub>/g VS/hr (14 to 50 scm/ton/hr) (Haug, 1993). Newsprint and MSW absorb oxygen at much lower rates, on the order of 0.5 mg O<sub>2</sub>/g VS/hr. These rates control the capacity of the aeration system.

The decay of many waste materials (substrates) can be represented as a pseudo first-order reaction if the wastes are subdivided into fast and slow reacting portions (Haug, 1993):

$$\frac{dX_{bs}}{dt} = k_{df}X_{bsf} + k_{ds}X_{bss} \quad (11.205)$$

where  $k_{df}$  = the decay rate of the fast-reacting biodegradable waste material (substrate) (per day)  
 $k_{ds}$  = the decay rate of the slow-reacting biodegradable waste material (substrate) (per day)  
 $X_{bsf}$  = the dry weight of the fast-reacting biodegradable waste material (substrate) (kg VS)  
 $X_{bss}$  = the dry weight of the slow-reacting biodegradable waste material (substrate) (kg VS)

The fast-reacting fractions are poorly known. Reported values for POTW sludges range from about one-fifth to two-fifths (Haug, 1993). The fast rates are often five to ten times the slow rates.

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## 11.7 Sludge Stabilization

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The treatment goal is the production of a stabilized sludge that will not produce offensive odors or attract disease vectors. The stabilization of wastewater sludges can occur aerobically or anaerobically. Stabilization means that the bulk of the organic solids is converted to metabolic end-products that resist further degradation and that do not produce nuisance odors or attract vectors. In aerobic digestion, these products are carbon dioxide, water, various inorganic salts, and humic/fulvic materials. In anaerobic digestion, the products are methane, carbon dioxide, various inorganic salts, and humic/fulvic materials.

Vector attraction is considered to be reduced if the following occurs (EPA, 1993; Stein et al., 1995):

- The VS content of the sludge is reduced by at least 38% by aerobic or anaerobic digestion.
- An anaerobically digested sludge loses less than 17% of its VS content upon further anaerobic batch-digestion at 30 to 37°C for 40 additional days.
- An aerobically digested sludge containing less than 2% solids loses less than 15% of its VS content upon further aerobic batch-digestion at 20°C or 30 additional days; sewage sludges containing more than 2% solids should be diluted to 2% solids prior to testing.
- The specific oxygen uptake rate of an aerobically digested sewage sludge is reduced to 1.5 mg O<sub>2</sub>/g TS/hr at 20°C.
- A sludge is digested aerobically at an average temperature of 45°C (minimum 40°C) for at least 14 days.
- The sludge is lime-stabilized (see below).
- The moisture content of the stabilized sewage sludge is less than 25%.
- The moisture content of an unstabilized sewage sludge is less than 10%.
- The sewage sludge is injected below the ground surface; no sludge may be on the surface within 1 hr of injection; Class A sludges must be injected within 8 hr of its discharge from a pathogen reduction process.
- The sewage sludge is incorporated into the soil by plowing and disking within 8 hr of land application.

Aerobic digestion consumes energy because of the aeration requirement. Anaerobic digestion is a net energy producer because of the methane formed, but often, the only economic use of the methane is

digester and space heating. Aerobic digestion seldom produces offensive odors (the usual smell being mustiness), but failed anaerobic digesters can produce foul odors.

Sludge digestion substantially reduces the numbers of pathogens and parasites, but it does not qualify as a sludge disinfection process. If digested sludges are to be applied to land, disinfection by heat treatment and/or lime stabilization is required.

## **Anaerobic Digestion**

Anaerobic digestion stabilizes organic sludges by converting them to gas and humus. The principal interest is the methane content of the gas, which is usable as a fuel. It is, however, a dirty gas, containing carbon dioxide, greasy aerosols, hydrogen sulfide, water vapor, and nitrogen, and it requires cleanup before use. Also, its fuel value is low compared to natural gas (pure methane). The upshot is that many facilities that have access to cheap commercial fuels burn off the methane to control its fire and explosion hazards. Some of the digester gas may be used for space heating, as this requires only sulfide removal. Iron sponge scrubbers usually remove hydrogen sulfide, which is relatively cheap.

The humus is suitable as a soil conditioner as long as its heavy metal and pathogen/parasite content is low. These must be monitored regularly. Most of the municipally produced humus is spread on farmland, either as a wet sludge or a dewatered solid. Some of it is incinerated, although it makes better sense to incinerate the raw sludge and capture its fuel value for the burning process.

### **Facilities Arrangement**

According to the ASCE survey of POTWs, about three-fourths of the plants employing anaerobic digestion have single-stage, heated (95°F), mixed tanks (Leininger, Sailor, and Apple, 1983). About one-fourth of the plants have two-stage systems with heated, mixed primary tanks followed by unheated, unmixed secondary tanks. The secondary tanks are intended to capture and thicken digested solids. However, the suspended solids produced by the primary tanks settle poorly because of gas flotation, particle size reduction due to digestion, and particle size reduction due to mixing (Brown and Caldwell, Consulting Engineers, Inc., 1979). In full-scale field units, only about one-third of the suspended solids entering the secondary digester will settle out (Fan, 1983). The construction of secondary digesters does not appear to be warranted.

Plants that attempt to concentrate digested sludges by gravity thickening (in secondary digesters or otherwise) frequently recycle the supernatant liquor to the primary settlers. This practice results in digester feed sludges that contain substantial portions of previously digested, inert organic solids, perhaps 25 to 50% of the VS in the feed. Consequently, the observed volatile solids destructions are significantly reduced, frequently to as little as 30%. Such low destructions merely indicate the presence of recirculated inert volatile solids and do not imply that incomplete digestion is occurring. However, if substantial fractions of inert organics are being recirculated, then the digesters are oversized or their hydraulic retention time and treatment capacity are reduced.

Typical digestion tanks are circular in plan with conical floors for drainage. Typical tank diameters are 24 m, and typical sidewall depths are 8 m (Leininger, Sailor, and Apple, 1983). Almost half of the digesters are mixed by recirculated digester gas, nearly one-fourth by injection from the roof via gas lances. About two-fifths of the digesters are mixed by external pumps, which generally incorporate external heat exchangers. So-called egg-shaped digesters permit more efficient mixing and are gradually replacing the old-fashioned cylindrical digesters in new facilities.

Digesters are almost always heated by some kind of external heat exchanger fueled with digester gas. In cold climates, gas storage is usually practiced, usually in floating gasholder covers or flexible membrane covers.

### **Microbiology and Pattern of Digestion**

Many of the bacteria responsible for anaerobic digestion are common intestinal microbes (Kirsch and Sykes, 1971). They are fastidious anaerobes; molecular oxygen kills them. The preferred temperature is

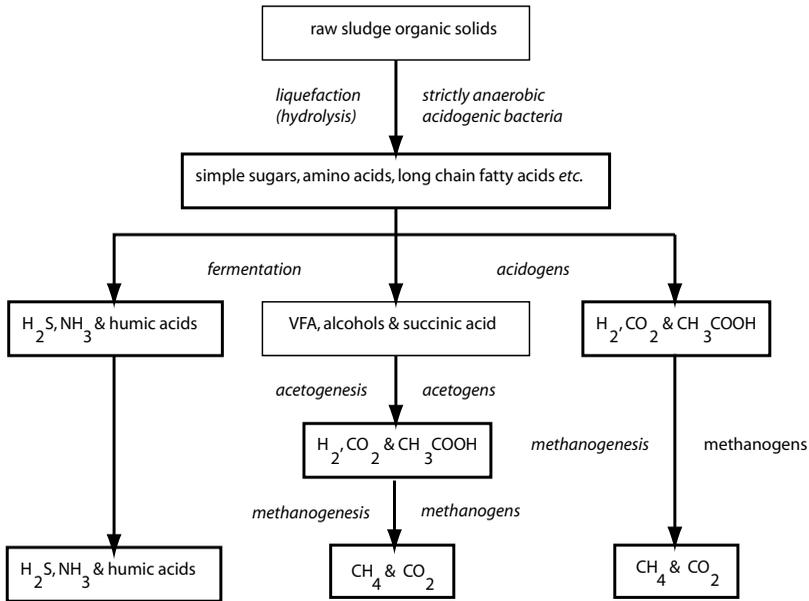
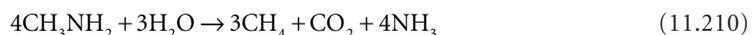


FIGURE 11.13 Pattern of organic solids decomposition in anaerobic digestion.

37°C (human body temperature), and the preferred pH range is about 6.5 to 7.5. Facultatively anaerobic bacteria like the coliforms comprise only a few percent of the population or less.

The pattern of anaerobic decomposition of wastewater sludges is indicated in Fig. 11.13. The primary ecological division among the bacteria in digesters is between acidogens and methanogens. The acidogen population (as a whole) hydrolyzes the cellulose and other complex carbohydrates, proteins, nucleotides, and lipids to simple organic molecules and ferments them to hydrogen gas, carbon dioxide, acetic acid, and other volatile fatty acids (VFA), other organic acids, alcohols, ammonia, hydrogen sulfide, and humic and fulvic acids. Three carbon and larger VFA, other organic acids, and alcohols are converted to acetic acid, possibly hydrogen gas and carbon dioxide, by a subgroup of the acidogenic population called the acetogens.

The methanogenic population (as a whole) converts acetic acid, carbon dioxide, and hydrogen, formic acid, methanol and tri-, di-, and monomethylamine to methane (Whitman, Bowen, and Boone, 1992).



No other substrates are known to support growth of the methanogens. Nearly all the known methanogens (except for *Methanotrix soehngenii*, which grows only on acetic acid) grow by reducing carbon dioxide with hydrogen, but the largest source of methane in digesters, about 70%, is derived from the lysis of acetic acid (McCarty, 1964). The methanogens are autotrophs and derive their cell carbon from carbon dioxide.

The composition of typical digested municipal sludge is given in Table 11.16, and the fate of various wastewater sludge components during digestion is indicated in Table 11.17 (Woods and Malina, 1965).

**TABLE 11.16** Approximate Composition of Digested Sludge

Item	Primary Digester Sludge		Secondary Digester Settled Sludge Solids with Associated Interstitial Water		Secondary Digester Decanted Supernatant Liquor	
	Median	Range	Median	Range	Median	Range
Total solids, TS (% by wt)	3.1	3.0–4.0	4.0	2.5–5.5	1.5	1.0–5.0
Volatile solids (% of TS)	58.0	49.0–65.0	51.0	44.0–60.0	50.0	1.0–71.0
Total suspended solids (mg/L)	—	—	—	—	(2205)	(143–7772)
Volatile suspended solids (mg/L)	—	—	—	—	(1660)	(118–3176)
pH	7.0	6.9–7.1	—	—	(7.2)	(7.0–8.0)
Alkalinity (mg/L as CaCO <sub>3</sub> )	2751	1975–3800	—	—	—	(1349–3780)
Volatile fatty acids (mg/L as HAc)	220	116–350	—	—	—	(250–322)
NH <sub>3</sub> -N (mg/L)	998	300–1100	—	—	—	(480–853)
BOD (mg/L)	—	—	—	—	2282	600–2650
COD	—	—	—	—	—	11–7000
Total P (mg/L)	—	—	—	—	—	(63–143)

Note: Numbers in parenthesis are taken from Brown and Caldwell (1979). The other numbers are taken from Leininger et al. (1983).

For the survey by Leininger et al. (1983), the range is the first and third quartile points of the distribution and represents the limits of the middle 50% of the reported data.

Sources: Brown and Caldwell, Consulting Engineers, Inc. 1979. *Process Design Manual for Sludge Treatment and Disposal*, EPA 625/1–79–011. Environmental Protection Agency, Municipal Environmental Research Laboratory, Office of Research and Development, Center for Environmental Research Information, Technology Transfer, Cincinnati, OH.

Leininger, K.V., Sailor, M.K., and Apple, D.K. 1983. *A Survey of Anaerobic Digester Operations, Final Draft Report*, ASCE Task Committee on Design and Operation of Anaerobic Digesters. American Society of Civil Engineers, New York.

**TABLE 11.17** Fate of Wastewater Sludge Constituents During Anaerobic Digestion

Item	Distribution of Feed Material in Products (% by wt)		
	Gas	Liquid	Solid
Carbon	54 <sup>a</sup>	26	10
Nitrogen (as N <sub>2</sub> )	11	70	9
Volatile solids	60	30	10
Carbohydrate	86	9	5
Fats/lipids	80	14	6
Protein	55	32	13

<sup>a</sup> Experimental error.

Source: Woods, C.E. and Malina, J.F., Jr. 1965. "Stage Digestion of Wastewater Sludge," *Journal of the Water Pollution Control Federation*, 37(11): 1495.

Overall, about 60% of the sludge organic matter is converted to gas, about 30% ends up as soluble organic matter, and 10% ends up in the residual humus solids. Over 80% of the influent carbohydrates and lipids are gasified, and most of the remainder ends up as soluble products. Only about half the proteins and other nitrogenous organics are gasified, and nearly one-third is converted to soluble products.

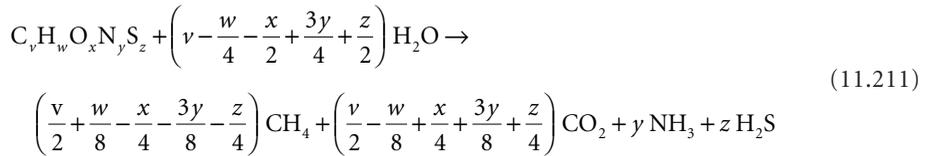
### Gas Stoichiometry

The principal benefit of anaerobic digestion is the methane gas produced, which can be used as a fuel. Typical municipal digesters produce a gas that is approximately 65% by vol. methane, 30% carbon dioxide, 2.6% nitrogen, 0.7% hydrogen, 0.4% carbon monoxide, 0.3% hydrogen sulfide, and about 0.2% other illuminants (Pohland, 1962). The fuel value of the raw gas is about 620 Btu/scf (1 atm, 32°F). Digester

gas is saturated with water vapor at the digestion temperature (42 mm Hg at 35°C) and contains an aerosol of small grease and sludge particles that is formed as gas bubbles burst at the liquid surface. The aerosols and hydrogen sulfide must be removed prior to transmission and burning.

The fuel value of the gas resides in the methane content. At 25°C (77°F), methane has a lower heating value (product water remains a vapor) of 21,500 Btu/lb (959 Btu/ft<sup>3</sup>) and a higher heating value (product water condenses) of 23,900 Btu/lb (1,064 Btu/ft<sup>3</sup>) (Van Wylene, 1963). Methane has an autoignition temperature of 650°C and lower and upper explosion limits in air of 5.3 and 15% by vol., respectively (Dean, 1992).

The quality and quantity of this gas is determined by the chemical composition of the volatile solids that are destroyed. This can be estimated using the following modification of Buswell's (1965) stoichiometry:



Organic nitrogen is released as ammonia, which reacts with water to form ammonium hydroxide and to trap some of the carbon dioxide produced:



The net result is that each mole of ammonia traps one mole of carbon dioxide. Accounting for this effect, the expected mole fractions of methane, carbon dioxide, and hydrogen sulfide are as follows:

$$f_{CH_4} = \frac{4v + w - 2x - 3y - 2z}{8(v - y + z)} \quad (11.213)$$

$$f_{CO_2} = \frac{4v - w + 2x - 5y + 2z}{8(v - y + z)} \quad (11.214)$$

$$f_{H_2S} = \frac{z}{v - y + z} \quad (11.215)$$

The estimate for hydrogen sulfide given in Eq. (11.215) is a maximum. The usual hydrogen sulfide concentration in digester gas is about 1% by vol., but it is variable (Joint Task Force, 1992). There are three general processes that reduce its gas phase concentration. First, hydrogen sulfide is a fairly soluble gas, about 100 times as soluble as oxygen, and much of it remains in solution. Second, hydrogen sulfide is also a weak acid, and the two-step ionization, which liberates bisulfide and sulfide, is pH dependent:



At 35°C and pH 7, most of the hydrogen sulfide exists as bisulfide, which further increases the amount of sulfide that remains in the sludge. Third, sulfide forms highly insoluble precipitates with many metals and can be trapped in the digested sludge as a metallic sulfide. The most common form is ferrous sulfide.

Carbohydrates and acetic acid produce gases that are 50/50 methane and carbon dioxide by vol. Proteins and long-chain fatty acids produce gases that are closer to 75/25 methane and carbon dioxide by vol.

Equation (11.211) indicates that anaerobic digestion is a pseudohydrolysis reaction, and that the weight of the gases produced may exceed the weight of the solids destroyed because of the incorporation of water. However, in the case of carbohydrates and acetic acid, there is no water incorporation, and the weight of the gases equals the weight of the carbohydrate/acetic acid destroyed. There is significant water incorporation in the destruction of long-chain fatty acids, and the weight of the gases formed may be 50% greater than the weight of the fatty acid destroyed. In the case of protein fermentation, there is significant water incorporation, but the trapping of carbon dioxide by ammonia yields a gas weight that is lower than the protein weight.

Primary sludges tend to contain more fats and protein and less carbohydrate than secondary sludges. Consequently, secondary sludges produce less gas but with a somewhat higher methane content. On the basis of volatile solids destroyed, the gas yields are as follows:

- Primary Sewage Solids (Buswell and Boruff, 1932):
  - 1.25 g total gas per g VS destroyed
  - 1.16 L total gas (SC) per g VS destroyed
  - CH<sub>4</sub>:CO<sub>2</sub>::67:33, by vol.
- Waste Activated Sludge and Trickling Filter Humus (Fair and Moore, 1932c):
  - 0.71 g total gas per g VS destroyed
  - 0.66 L total gas (SC) per g VS destroyed
  - CH<sub>4</sub>:CO<sub>2</sub>::71:29, by vol.

It is easier to estimate the methane production from a COD balance on a digester. Because it is an anaerobic process, all the COD removed from the sludge ends up in the methane produced. The COD:CH<sub>4</sub> ratio can be estimated from,



Consequently, the COD of 1 mole of methane is 64 g or 4 g COD/g CH<sub>4</sub>. Because 1 mole of any gas occupies 22.414 L at standard conditions (0°C, 1 atm), the ratio of gas volume to COD is 0.350 L CH<sub>4</sub> (SC) per g COD removed.

### Kinetics

Sludge digesters are usually designed to be completely mixed, single-pass reactors without recycle. This is due to the fact that digested sludges are only partially settleable, and solids capture by sedimentation is impractical. Consequently, the hydraulic retention time of the system is also the solids' retention time:

$$\Theta_x = \frac{VX}{QX} = \frac{V}{Q} = \tau \quad (11.219)$$

where  $Q$  = the raw sludge flow rate (m<sup>3</sup>/s)  
 $V$  = the digester's volume (m<sup>3</sup>)  
 $X$  = the suspended solids' concentration in the digester (kg/m<sup>3</sup>)  
 $\Theta_x$  = the solids' retention time (s)  
 $\tau$  = the hydraulic retention time (s)

Stewart (1958) first demonstrated and Agardy, Cole, and Pearson (1963) confirmed the applicability of Gram's (1956) model for the activated sludge process to anaerobic digestion. The important relationships are as follows:

$$\mu = Yq \quad (11.220)$$

$$q = \frac{q_{\max}S}{K_s + S} \quad (11.221)$$

$$\mu = \frac{\mu_{\max} S}{K_s + S} \quad (11.222)$$

$$\frac{1}{\Theta_x} = \mu - k_d \quad (11.223)$$

- where  $K_s$  = Monod's affinity constant (kg COD/m<sup>3</sup>)  
 $k_d$  = the "decay" rate (per s)  
 $q$  = the specific uptake (or utilization) rate (kg COD/kg VSS·s)  
 $q_{\max}$  = the maximum specific uptake (or utilization) rate (kg COD/kg VSS·s)  
 $S$  = the kinetically limiting substrate's concentration (kg COD/m<sup>3</sup>)  
 $\mu$  = the specific growth rate (per s)  
 $\mu_{\max}$  = the maximum specific growth rate (per s)  
 $\Theta_x$  = the solids' retention time (s)

The rate-limiting step in the conversion of organic solids to methane is the fermentation of saturated long-chain fatty acids to acetic acid (Fan, 1983; Novak and Carlson, 1970; O'Rourke, 1968). Kinetic constants for growth on selected volatile fatty acids, long-chain fatty acids, and hydrogen are given in Table 11.18. The minimum solids' retention time for satisfactory digestion is determined by using the kinetic parameters for long-chain fatty acid fermentation.

The kinetic parameters apply to the "biodegradable" fraction of the lipids in municipal wastewater sludges. This really means the fraction converted to gas; the remaining lipid is conserved in other soluble and particulate microbial products. O'Rourke (1968) estimates that 72% of the lipids in municipal sludges can be gasified at 35°C. The gasifiable fraction falls to 66% at 25°C and to 59% at 20°C. Lipids are not gasified below 15°C.

**TABLE 11.18** Gram Model Kinetic Parameters for Anaerobic Digestion at 35°C (Preferred Design Values Shown Boldface)

Gram Model Parameter	Substrate				
	H <sub>2</sub>	Acetic Acid	Propionic Acid	Butyric Acid	Long-Chain Fatty Acids
$\mu_{\max}$ (per day)	1.06	0.324	0.318	0.389	0.267 <b>(0.267)</b>
$q_{\max}$ (kg COD/kg VSS·d)	24.7	6.1	9.6	15.6	6.67 <b>(6.67)</b>
$K_s$ (mg COD/L)	569 (mm Hg)	164	71	16	2000 <b>(1800)</b>
$k_d$ (per day)	-0.009	0.019	0.01	0.027	0.038 <b>(0.030)</b>
$Y$ (kg VS/kg COD)	0.043	0.041	0.042	0.047	0.054 <b>(0.040)</b>

Sources: Lawrence, A.W. and McCarty, P.L. 1967. *Kinetics of Methane Fermentation in Anaerobic Waste Treatment*, Tech. Rept. No. 75. Stanford University, Department of Civil Engineering, Stanford, CA.

O'Rourke, J.T. 1968. *Kinetics of Anaerobic Waste Treatment at Reduced Temperatures*. Ph.D. Dissertation. Stanford University, Stanford, CA.

Shea, T.G., Pretorius, W.A., Cole, R.D., and Pearson, E.A. 1968. *Kinetics of Hydrogen Assimilation in Methane Fermentation*, SERL Rept. No. 68-7. University of California, Sanitary Engineering Research Laboratory, Berkeley, CA.

Speece, R.E. and McCarty, P.L. 1964. "Nutrient Requirements and Biological Solids Accumulation in Anaerobic Digestion: Advances in Water Pollution Research," in *Proceedings of the International Conference, London, September, 1962, Vol. II*, W.W. Eckenfelder, Jr., ed. Pergamon Press, New York.

The Wastewater Committee (1997) limits the solids loading to primary anaerobic digesters to a maximum of 80 lb VS per 1000 ft<sup>3</sup> per day (1.3 kg/m<sup>3</sup>·d), providing the units are completely mixed and heated to 85 to 95°F. The volume of secondary digesters used for solids capture and storage may not be included in the loading calculation. If the feed sludges have a VS content of 2%, the resulting hydraulic retention time of the primary digester is 16 days.

The median HRT employed for mixed, heated primary digesters treating primary sludge only is 20 to 25 days (Joint Task Force, 1992). The implied safety factor for 90% conversion of volatile solids to gas is about 2 to 2.5. Excluding HRTs less than 10 days, the median HRT for mixed, heated digesters fed a blend of primary and waste activated sludges is about 30 to 35 days. For either sludge, the modal HRT is 20 to 25 days.

## Temperature

Nearly all municipal digesters are heated to about 35°C (or 95°F), which is in the mesothermal range to which the intestinal microbes that dominate the process are adapted. Anaerobic ponds used to treat packinghouse wastes may not be heated directly, but much of the process wastewater is hot, and the metabolic heat contributes to maintaining a temperature above ambient.

Psychrophilic digesters operate below 30°C. Psychrophilic digestion is quite common in anaerobic ponds, septic tanks, and wetlands. Psychrophilic digestion is slower than mesophilic or thermophilic digestion. Also, below about 30°C, the fraction of the sludge solids converted to gas is reduced (Maly and Fadrus, 1971). The reduction is roughly linear, and at 10°C, the fraction of solids converted to gas is only about 60% of the conversion at 30°C. Long-chain fatty acids accumulate below about 18°C, which causes foaming (Fan, 1983).

Thermophilic digesters operate above 40°C, usually at 50°C or somewhat higher. The chief advantages of thermophilic digestion are as follows (Buhr and Andrews, 1977):

- More rapid completion of the digestion process, as indicated by the cumulative gas production
- Better dewatering characteristics
- Disinfection of pathogens and parasites, if operated above 50°C

There are, however, several reports of disadvantages (Pohland, 1962; Kirsch and Sykes, 1971):

- Accumulation of volatile fatty acids and reduction of pH
- Accumulation of long-chain fatty acids
- Reduced gas production per unit volatile solids fed
- Reduced methane content of the gas
- Offensive odors
- Reduced volatile solids destruction
- Impaired dewatering characteristics
- Increased heating loss rates (although the tanks are smaller, and the smaller surface area somewhat offsets the higher heat flux)

The systems reporting impaired digestion were not operating stably, and there is a general opinion that thermophilic digestion is difficult to establish and maintain. Part of the problem may lie in the reduced suite of microbes that are able to grow thermophilically. Naturally occurring thermophilic environments are rare, and there are few thermophiles among the acidogens and methanogens in primary sewage sludge, which is derived entirely from psychrophilic and mesophilic environments. In any particular plant, the conversion from mesophily to thermophily may require some time for proper seeding by the few thermophiles in the influent sludge. If the digester goes sour in the interim, the seeding may fail to take. It should be noted that many of the failed thermophilic digestion experiments were conducted using laboratory-scale units. The seeding problem here is compounded by the Poisson nature of the sludge sampling process; it is likely that none of the small samples needed for laboratory work will contain any

thermophiles. It should be noted that the full-scale thermophilic plant at Los Angeles was a stable process (Garber, 1954).

The effect of temperature on digestion can be represented in terms of Gram's model. For digestion temperatures in the range 20 to 35°C, Parkin and Owen (1986) recommend the following:

$$q_{\max} = 6.67 \times 1.035^{T-35} \quad (\text{kg COD/kg VSS} \cdot \text{d}) \quad (11.224)$$

$$K_s = 1.8 \times 0.8993^{T-35} \quad (\text{g COD/L}) \quad (11.225)$$

$$k_d = 0.030 \times 1.035^{T-35} \quad (\text{per day}) \quad (11.226)$$

$$Y = 0.040 \quad (\text{g VSS/g COD}) \quad (11.227)$$

The base values are referenced to 35°C and are derived from O'Rourke's (1968) work. True growth yields do not vary significantly with temperature. The microbial decay rate is so poorly known that temperature adjustments may not be warranted. Note that the effective  $\theta$  for the combined temperature effect on the maximum uptake rate and the affinity constant is about 1.15, which is much larger than the values reported for gasification rates.

## pH

Methane production only occurs between pH 5 and 9; the optimum pH is near 7 and falls rapidly as the pH increases or decreases. Price's (1963) data may be summarized as follows:

- Peak rate of methane formation at pH 7
- 90% of peak rate at pH 6.5 and 7.5
- 75% of peak rate at pH 6 and 8
- 50% of peak rate at pH 5.8 and 8.4
- 25% of peak rate at pH 5.4 and 8.8

## Inhibitors

Digestion inhibitors are listed in [Table 11.19](#). Some of the metals listed are also nutrients. The optimum concentration for  $\text{Na}^+$ ,  $\text{K}^+$ , or  $\text{NH}_4^+$  is 0.01 mol/L; the optimum concentration for  $\text{Ca}^{2+}$  or  $\text{Mg}^{2+}$  is 0.005 mol/L (Kugelman and McCarty, 1965).

Poisons must be in soluble form to be effective. Cobalt, copper, iron, lead, nickel, zinc, and other heavy metals form highly insoluble sulfides ranging in solubility from  $10^{-5}$  to  $10^{-11}$  mg/L, which eliminates the metal toxicity (Lawrence and McCarty, 1965). Heavy metals may comprise as much as 10% of the volatile solids without impairing digestion if they are precipitated as sulfides. The requisite sulfide may be fed as sodium sulfide or as various sulfate salts, which are reduced to sulfide.

It should be noted that methanogens reduce mercury to mono- and dimethylmercury, which are volatile and insoluble and may be present in the digester gas. Alkyl mercurials are toxic.

Halogenated methane analogs like chloroform, carbon tetrachloride, and Freon are toxic to methanogens at concentrations on the order of several mg/L (Kirsch and Sykes, 1971).

## Moisture Limitation

The stoichiometry of anaerobic digestion indicates that water is consumed and may be limiting in low moisture environments. Anaerobic digestion proceeds normally at total solids concentrations up to 20 to 25% by wt (Wujcik, 1980). Above about 30% TS, the rate of methane production is progressively reduced and ceases at 55% TS. In this range, the methanogens appear to be water-limited rather than salt-, ammonia-, or VFA-limited. Above 55% TS, acid production is inhibited.

**TABLE 11.19** Anaerobic Digestion Inhibitors

Substance	Effect	Concentration Units	Concentration
Inorganic			
Ammonia-nitrogen	Moderate	mg/L	1500–3000
	Strong	mg/L	3000
Calcium	Moderate	mg/L	2500–4500
	Strong	mg/L	8000
Chromium (III)	Strong	mg/L	180–420 (total)
Chromium (VI)	Strong	mg/L	3.0 (soluble)
	Strong	mg/L	200–260 (total)
Copper	Strong	mg/L	0.5 (soluble)
	Strong	mg/L	50–70 (total)
Magnesium	Moderate	mg/L	1000–1500
	Strong	mg/L	3000
Nickel	Strong	mg/L	1.0 (soluble)
	Strong	mg/L	30 (total)
Potassium	Moderate	mg/L	2500–4500
	Strong	mg/L	12,000
Sodium	Moderate	mg/L	3500–5500
	Strong	mg/L	8000
Zinc	Strong	mg/L	1.0 (soluble)
Organic			
Acetaldehyde	50% activity	mmol/L	10
Acrylic acid	50% activity	mmol/L	12
Acrylonitrile	50% activity	mmol/L	4
Acrolein	50% activity	mmol/L	0.2
Aniline	50% activity	mmol/L	26
Catechol	50% activity	mmol/L	24
Chloroform	“inhibitory”	mg/L	0.5
3-Chloro-1,3-propanediol	50% activity	mmol/L	6
1-Chloropropane	50% activity	mmol/L	1.9
1-Chloropropene	50% activity	mmol/L	0.1
2-Chloropropionic acid	50% activity	mmol/L	8.0
Crotonaldehyde	50% activity	mmol/L	6.5
Ethyl acetate	50% activity	mmol/L	11.0
Ethyl benzene	50% activity	mmol/L	3.2
Ethylene dichloride	“inhibitory”	mg/L	5
Formaldehyde	50% activity	mmol/L	2.4
Kerosene	“inhibitory”	mg/L	500
Lauric acid	50% activity	mmol/L	2.6
Linear alkylbenzene sulfonate	“inhibitory”	mg/L	1% of dry solids
Nitrobenzene	50% activity	mmol/L	0.1
Phenol	50% activity	mmol/L	26
Propanol	50% activity	mmol/L	90
Resorcinol	50% activity	mmol/L	29
Vinyl acetate	50% activity	mmol/L	8

Source: Parkin, G.F. and Owen, W.F. 1986. “Fundamentals of Anaerobic Digestion of Wastewater Sludge,” *Journal of Environmental Engineering*, 112(5): 867.

## Mixing

For high-rate digesters fed unthickened sludges, the required mixing power is about 0.2 to 0.3 hp per 1000 cu ft (Joint Task Force, 1992). Alternatively, the required rms velocity gradient is 50 to 80 per sec, and the turnover time is 30 to 45 min.

In conventional digesters, mixing becomes impaired at VS loading rates above 0.3 lb per cu ft per day (4 kg/m<sup>3</sup>·d) (Metcalf & Eddy, Inc., 1991). Ammonia toxicity limits VS loadings to about 0.2 lb per cu ft per day (3.2 kg/m<sup>3</sup>·d) (Joint Task Force, 1992).

Sludge pumping becomes a problem at about 8 to 12% TS (Brisbin, 1957). In this range, the Hazen–Williams *C* coefficient should be reduced by 60 to 75%.

## Heat Balance

The higher heating value of raw or digested sludge solids is as follows (Fair and Moore, 1932):

### *Primary sludge:*

$$\Delta H = 29P^{4/3} \quad (\text{Btu/lb TS}) \quad (11.228)$$

### *Waste activated sludge:*

$$\Delta H = 25P^{4/3} \quad (\text{Btu/lb TS}) \quad (11.229)$$

where *P* = the percentage of volatile solids in the total solids (%).

Note that the heating value does not vary linearly with the volatile solids content of the sludge solids. For primary sludges, the higher heating value extrapolated to 100% VS is 13,500 Btu per lb; for waste activated sludges, the higher heating value extrapolated to 100% VS is 11,600 Btu per lb.

The principal use of digester gas is digester heating. The steady state heat transfer due to conduction through material is directly proportional to the temperature difference and the area normal to the heat flow, and it is inversely proportional to the thickness of the material. A heat transfer coefficient, *k*, may be defined for any pure substance by,

$$Q_h = \frac{kA\Delta T}{L} \quad (11.230)$$

where *A* = the area normal to the heat flux (m<sup>2</sup>)  
*k* = the heat transfer coefficient (J/m·s·K)  
*L* = the thickness of the medium conducting the heat (m)  
*Q<sub>h</sub>* = the heat flow (J/s)  
*ΔT* = the temperature difference across the conducting medium (K)

Heat transfer coefficients for some materials are given in [Table 11.20](#).

If a wall or roof is made up of several layers of different substances, then an overall heat transfer coefficient, *K*, can be calculated by summing the temperature drops across each component and by noting that each component transmits the same heat flux and has the same area:

$$\frac{1}{K} = \frac{L_1}{k_1} + \frac{L_2}{k_2} + \dots + \frac{L_{n-2}}{k_{n-2}} + \frac{L_{n-1}}{k_{n-1}} \quad (11.231)$$

where *K* = the overall heat transfer coefficient (J/m<sup>2</sup>·s·K). As a matter of convenience, the thickness of the composite is adsorbed into the definition of the overall transfer coefficient.

**TABLE 11.20** Heat Transfer Coefficients for Various Materials

Material	Transfer Coefficient, $k$ (Btu·in./ft <sup>2</sup> ·hr·°F)
Air	0.17
Brick	3.0–6.0
Concrete	2.0–3.0
Earth, dry	10
Earth, wet	30
Mineral wool insulation	0.26–0.29
Steel	5.2–6.0

Source: Joint Task Force of the Water Environment Federation and the American Society of Civil Engineers. 1992. *Design of Municipal Wastewater Treatment Plants: Volume II. Chapters 13–20*, WEF Manual of Practice No. 8, ASCE Manual and Report on Engineering Practice No. 76. Water Environment Federation, Alexandria, VA; American Society of Civil Engineers, New York.

A complete heat balance on an anaerobic digester is as follows:

$$\begin{aligned}
 \underbrace{\Delta H_{req}}_{\text{heat required}} &= \underbrace{C_p \rho Q (T_{dig} - T_{slu})}_{\text{raw sludge heating}} + \underbrace{K_r A_r (T_{dig} - T_{air})}_{\text{heat through roof}} + \underbrace{K_w A_w (T_{dig} - T_{grd})}_{\text{heat through wall}} \\
 &+ \underbrace{K_f A_f (T_{dig} - T_{grd})}_{\text{heat through floor}} - \underbrace{H_{met} Q (X_{vo} - X_{ve})}_{\text{metabolic heat}}
 \end{aligned} \tag{11.232}$$

- where
- $A$  = area normal to heat flux (m<sup>2</sup>)
  - $C_p$  = constant pressure specific heat of water (J/kg)
  - $H_{met}$  = metabolic heat release (J/kg·VS)
  - $\Delta H_{req}$  = heat requirement (J/s)
  - $K$  = overall heat transfer coefficient (J/m<sup>2</sup>·s·K)
  - $Q$  = sludge flow rate (m<sup>3</sup>/s)
  - $T_{air}$  = air temperature (K)
  - $T_{dig}$  = digester temperature (K)
  - $T_{grd}$  = ground temperature (K)
  - $T_{slu}$  = sludge temperature (K)
  - $X_{ve}$  = effluent VSS (kg/m<sup>3</sup>)
  - $X_{vo}$  = influent VSS (kg/m<sup>3</sup>)
  - $\rho$  = mass density of water (kg/m<sup>3</sup>)

When VS are destroyed, approximately 80% of their fuel value is retained in the methane formed, and 20% is liberated to the digesting sludge as metabolic heat (Fan, 1983). A somewhat conservative estimate of the metabolic heat release is 2000 Btu/lb VS destroyed (1100 cal/g VS destroyed). This raises the possibility of autothermal anaerobic digestion. Ignoring the heat losses by conduction and setting the heat requirement to zero, one gets,

$$X_{vo} - X_{ve} = \frac{C_p \rho (T_{dig} - T_{slu})}{H_{met}} \tag{11.233}$$

Assuming 60% VS destruction and a sludge temperature increase of 40°F, the influent VS concentration for autothermal mesophilic digestion is about 5% by wt. This is equivalent to about 8% by wt TS, which is near the pumping limit.

## Aerobic Digestion

### Configuration

Aerobic digestion is usually restricted to smaller facilities where the cost of aeration is offset by the simplicity of the operation and facilities (Joint Task Force, 1992).

The digesters are constructed as open, unheated tanks. A variety of plan geometries have been built, including rectangular, circular, and annular tanks (Joint Task Force, 1992). Side water depths range from 10 to 25 ft. Aerobic digesters are liable to foam, and freeboard heights of 1.5 to 4 ft are required to retain the foam. Aeration and mixing are usually provided by diffused air systems, either coarse or fine bubble. Airflow rates of 20 to 40 scfm per 1000 cu ft are needed for mixing. Diffused air permits better control of dissolved oxygen and reduces heat losses, which is important in cold climates. Mechanical surface aerators have lower maintenance costs, but they produce greater heat losses, increase foam production, and are more liable to reduce oxygen transfer efficiency due to foam.

Typical process loadings are 24 to 140 lb VS per 1000 ft<sup>3</sup>/day, and reactor volume allowances are 3 to 4 ft<sup>3</sup> per cap (Schwinn and Gasset, 1974). Many small facilities store digested sludge in the digester for substantial time periods prior to disposal (e.g., because of seasonal land application), and allowances must be made for this additional storage.

Federal regulations require minimum solids' retention times of 40 days at 20°C and 60 days at 15°C and a minimum volatile solids destruction of 38% (Environmental Protection Agency, 1993).

Typical compositions of aerobic digester supernatants are summarized in [Table 11.21](#).

### Microbiology

The process of aerobic sludge digestion is a continuation of phenomena occurring in the activated sludge process. In some installations, soluble substrate levels are low, and heterotrophic growth of bacteria is small. Initially, there may also be some endogenous respiration by bacteria starved for substrate. The principal digestion process is the predation and scavenging by “worms,” rotifers, and protozoa of the bacteria and other sludge solids and cell lysis by viruses. Some bacteria may hydrolyze particulate matter, and the digestion processes of the predators and scavengers may release soluble substrates that support some heterotrophic growth and the growth of their phages. Additional soluble substrate may be released when the phages lyse the cells of their bacterial hosts.

**TABLE 11.21** Properties of Aerobic Digester Supernatants

Parameter (Units)	Mean Value	Range
BOD (mg/L)	500	9–1700
Soluble BOD (mg/L)	51	4–183
COD (mg/L)	2600	228–8140
Suspended solids (mg/L)	3400	46–11,500
Alkalinity (mg/L, as CaCO <sub>3</sub> )	—	473–514
pH	7	5.9–7.7
TKN (mg/L)	170	10–400
Total P (mg/L)	98	19–241
Soluble P (mg/L)	26	2.5–64

Sources: Schwinn, D.E. and Gasset, R.B., eds. 1974. *Process Design Manual for Upgrading Existing Wastewater Treatment Plants*. Environmental Protection Agency, Technology Transfer, Washington, DC.

## Kinetics

The usual assumptions are that the volatile solids may be divided into an inert fraction and a biodegradable fraction with destruction that obeys first-order kinetics (Adams, Eckenfelder, and Stein, 1974). For a completely mixed digester, the volatile solids destruction may be modeled as follows:

$$\frac{X_{vd} - X_{vi}}{X_{vo} - X_{vi}} = \frac{1}{1 + k_d \Theta_X} \quad (11.234)$$

where  $k_d$  = the decay rate (per sec)  
 $X_{vd}$  = the VSS in the digester ( $\text{kg}/\text{m}^3$ )  
 $X_{vi}$  = the inert or unbiodegradable VSS ( $\text{kg}/\text{m}^3$ )  
 $X_{vo}$  = influent VSS ( $\text{kg}/\text{m}^3$ )  
 $\Theta_X$  = the solid's retention time (sec)

A typical decay rate at 20°C is 0.08 to 0.12 per day (Brown and Caldwell, 1979). The decay may decline with increasing suspended solids concentrations. For bench-scale units, Reynolds (no date) reported a decline from 0.72 per d at a TSS of 8400 mg/L to 0.34 per d at a TSS of 22,700 mg/L. The sludge was digested at room temperature. Reynolds decay rates are substantially higher than other reported rates; his sludges were obtained from a contact-stabilization plant.

At very long HRTs, the digester VSS concentration,  $X_{vd}$ , approaches the inert or unbiodegradable VSS concentration,  $X_{vi}$ . Typically, about 50 to 60% of the volatile solids in waste activated sludge are biodegradable (Reynolds, no date).

It should be noted that the suspended ash (TSS minus VSS) is solubilized during digestion, and its concentration declines in parallel with the decline in VSS (Eckenfelder, 1956; Reynolds, no date). However, the solubilized solids remain in the liquid as part of the sludge and are not removed unless a dewatering process is applied to the sludge.

For temperatures above 15°C, SRTs range from 10 to 15 days for waste activated sludge and 15 to 20 days for primary sludge and for mixtures of waste activated and primary sludges (Schwinn and Gasset, 1974).

## Temperature

The variation of the decay rate with temperature is given approximately by the following:

$$k_d = 0.332 \left\{ 1 - \exp \left[ -0.0403(T - 8) \right] \right\}; \quad R^2 = 0.53 \quad (11.235)$$

where  $k_d$  = the decay rate (per day)  
 $T$  = the digestion temperature (°C)

Equation (11.235) was derived from the data summarized by Brown and Caldwell (1979) using the Thomas graphical method for fitting the BOD<sub>5</sub> curve. All the data were used, and the data span the temperature range 10 to 64°C. The derived curve lies somewhat above the hand-drawn curve presented in the report. The scatter about either line is very large, and digestion rates should be based upon pilot studies.

An examination of the plotted data suggests that the digestion rate reaches a maximum of 0.23 per day at a digestion temperature of 40°C. There is no clear thermophilic digestion range, which may reflect the limited number of thermophilic eukaryotes. There are no thermophilic rotifers or worms, which are the dominant predators affecting aerobic digestion.

## pH

The comments on the activated sludge process apply here as well. Organic solids destruction is not appreciably affected between pH 6 and 9. However, as long as the dissolved oxygen concentration is above 2 mg/L, aerobic digesters will nitrify, and the pH will fall in poorly buffered waters.

## Inhibitors

See [Tables 11.4](#) and [11.5](#). If nitrification is desired, the special requirements of the nitrifying bacteria will control.

## Oxygen Requirements

The general oxygen balance for activated sludge also applies to aerobic digestion. For nonnitrifying digesters,

$$R = 1.42Q(X_{vo} - X_{ve}) \quad (11.236)$$

and for nitrifying digesters,

$$R = 1.98Q(X_{vo} - X_{ve}) \quad (11.237)$$

See the comments about oxygen requirements in the activated sludge process, especially the requirements of the nitrifying bacteria.

## Mixing Requirements

The power required to mix thickened sludges may be estimated from the following (Zwietering, 1958; Reynolds, no date):

$$P = 0.00475\mu^{0.3} X_{TSS}^{0.298} V \quad (11.238)$$

where  $P_{\min}$  = minimum required mixing power (hp)

$V$  = digester volume (1000 gal)

$X_{TSS}$  = the TSS concentration (mg/L)

$\mu$  = the liquid viscosity (centipoise)

## Autothermal Thermophilic Digestion

The heat balance given above for anaerobic digesters also applies to aerobic digesters. However, in aerobic digestion, all the higher heating values of the destroyed volatile solids are released as metabolic heat, so the break-even point for autothermal digestion is a feed sludge containing between 1 and 2% VS. This is equivalent to about 2 to 3% TS, which is well within the limits for good mixing.

In European practice, waste activated sludges are first thickened to at least 2.5% by wt VSS (Joint Task Force, 1992). The digesters are cylindrical with a height-to-diameter ratio of 0.5 to 1.0. They are operated in the fill-and-draw mode with two temperature phases per cycle. The first temperature phase is 35 to 50°C and is intended to stabilize the sludge. The second phase is 50 to 65°C and is intended to reduce pathogens. The HRT for the digester is 5 to 6 days, with a minimum HRT of 20 hr in either temperature phase. Aeration is generally by diffused air. Substantial foaming occurs; foam cutters are needed to control foam accumulation. Nitrification does not occur at thermophilic temperatures, which reduces the oxygen requirement.

## Land Disposal of Sludges

The general requirements for pathogens in sewage sludges applied to land are given in [Table 11.22](#) (EPA, 1993, 1994). The Class A requirements apply to sewage sludges used on home gardens and lawns. Class A specifications also may be used for sludges applied to agricultural land, forest, public contract sites, or reclamation sites. Class B specification may be used for sludges applied to agricultural land, forests, public contract sites, and reclamation sites if certain site usage, cropping, and pasturing restrictions are met.

**TABLE 11.22** General Pathogen Restrictions for the Disposal of Sewage Sludges

Pathogen	Class A Requirements (Lawn and Garden)	Class B Requirements (Agricultural Land)
Fecal coliform	All alternatives: <1000 MPN per g total solids	$<2 \times 10^6$ per g total solids
Salmonella	All alternatives: <3 MPN per 4 g total solids	—
Enteric viruses	Alternatives 3 through 6: <1 PFU per 4 g total solids	—
Helminth ova	Alternatives 3 through 6: <1 ova per 4 g total solids	—

Sources: Environmental Protection Agency. 1993. "Standards for the Use or Disposal of Sewage Sludge," *Federal Register*, 58(32): 9248.

Environmental Protection Agency. 1994. "Standards for the Use or Disposal of Sewage Sludge," *Federal Register*, 59(38): 9095.

### Class A Sludges

Sewage sludges can be graded Class A if they are subjected to certain pathogen reduction treatments. There are six alternative treatments.

#### Alternative 1

This alternative requires heat inactivation of pathogens. Specific testing for enteric viruses and helminth ova is not required.

The minimum temperature for heat inactivation regardless of contacting time is 50°C (EPA, 1993). The contacting time required depends on the solids content of the sludge, the temperature, and the mode of heating. Required contacting times for disinfection temperatures of 50°C and above are given by equations that follow. Note that each method of heating and sludge concentration has an absolute minimum contacting time regardless of temperature.

- If the sludge solids concentration is 7% by weight or higher, heating of the bulk sludge is by heat exchanger, use Eq. (11.239) below. The minimum contacting time regardless of disinfection temperature is 20 min.

$$t_c = \frac{131\,700\,000}{10^{0.1440T}} \quad (11.239)$$

where  $t_c$  = the contacting time in days  
 $\geq 20$  min regardless of temperature  
 $T$  = the inactivation temperature in °C  
 $\geq 50^\circ\text{C}$

- If the sludge solids concentration is 7% by weight or higher, sludge heating is by diffused hot gas or hot immiscible liquid use Eq. (11.239) above, with the proviso that the minimum contacting time is 15 min, regardless of disinfection temperature.
- If the sludge concentration is less than 7% solids by weight, use Eq. 11.240 below to determine contacting times, with the proviso that the minimum contacting time regardless of inactivation temperature is 30 min.

$$t_c = \frac{50\,070\,000}{10^{0.1440T}} \quad (11.240)$$

where  $t_c$  = the contacting time in days  
 $\geq 30$  min regardless of temperature  
 $T$  = the inactivation temperature in °C  
 $\geq 50^\circ\text{C}$

### ***Alternative 2***

This method combines pH and heat inactivation:

- The pH of the sludge shall be kept above 12 for 72 hr.
- During the period of elevated pH, the sludge temperature shall be kept at 52°C or higher for at least 12 hr.
- After heat/pH treatment, the sludge shall be air-dried to a solids content of at least 50% by weight.

### ***Alternative 3***

No treatment method is specified; rather the sludge quality is further constrained. The sewage sludge shall be tested periodically for enteric viruses and viable helminth ova as well as fecal coliform and *Salmonella*, and it will be graded Class A as long as the conditions of [Table 11.22](#) are met. Failure to meet the tabulated requirements requires pathogen treatment and retesting of the treated sludge.

### ***Alternative 4***

No pathogen reduction treatment is required. However, the sludge must meet the conditions in [Table 11.22](#) at the time of sale or usage.

### ***Alternative 5***

The sludges must meet the fecal coliform and *Salmonella* standards in [Table 11.22](#) at the time of sale and usage and must be subjected to pathogen reduction treatments specified in Appendix B of the Rule.

### ***Alternative 6***

The sludges must meet the fecal coliform and *Salmonella* standards in [Table 11.22](#) at the time of sale and usage and must be subjected to pathogen reduction treatments approved by the permitting authority.

## **Class B Sludges**

### ***Alternative 1***

The sludge must meet the fecal coliform standard in [Table 11.22](#) at the time of usage.

### ***Alternative 2***

The sludge must be subjected to a pathogen reduction treatment specified in Appendix B of the rule.

### ***Alternative 3***

Sludges graded Class B must be subjected to pathogen reduction processes approved by the permitting authority. Also, there are site usage restrictions:

- Food crops that touch the sludge/soil mixture and are not entirely above ground may not be harvested for 14 months after sludge application.
- Food crops with harvested parts below the ground may not be harvested for 20 months after sludge application, if the sludge remains on the land surface for 4 months or longer prior to incorporation into the soil.
- Food crops with harvested parts below the ground may not be harvested for 38 months after sludge application when the sludge lies on the land surface for less than 4 months before incorporation into the soil.
- Food crops, feed crops, and fiber crops shall not be harvested for 30 days after sludge application.
- Animals may not be pastured on the land for 30 days after sludge application.
- Turf may not be harvested for 1 year after sludge application if the turf will be used as a lawn or if there is a high potential for public contact with the turf.
- Public access to land with a high potential to public exposure shall be restricted for 1 year after sludge application.
- Public access to land with a low potential for public exposure shall be restricted for 30 days after sludge application.

## Lime Stabilization

Lime stabilization requires that the sludge pH be raised above 12 and preferably above 12.5. The pH of 12 or more must be held for at least 2 hr without additional lime, and a pH of at least 11.5 must be held without additional lime for another 2 hr (Stein et al., 1995). At these pHs, there is substantial hydrolysis of organic molecules, and the lime dosage is controlled by the amount of solids to be treated. Following are some guidelines as to hydrated lime dosages (Joint Task Force, 1992):

**Primary sludge** — 0.1 to 0.15 kg Ca(OH)<sub>2</sub> per kg suspended solids

**Activated sludge** — 0.3 to 0.5 kg Ca(OH)<sub>2</sub> per kg suspended solids

**Septage** — 0.1 to 0.3 kg Ca(OH)<sub>2</sub> per kg suspended solids

At pH 12, a minimum contacting time of at least 72 hr is required for stabilization (Class A, Alternative 2). The hydrolysis reactions consume lime, so additional dosages of lime may be required during the contacting period.

Fecal streptococci are not inactivated unless the contacting time is at least 1 hr, and high pH does not affect nematodes, protozoa, or mites.

Lime stabilization releases organic nitrogen as ammonia, and some means of controlling the emitted ammonia gas is required.

## Concentration and Loading Limits for Sewage Sludge

Concentration and loading limits for various contaminants are given in [Table 11.23](#) (EPA, 1993, 1994). Bulk or packaged sewage sludges may not be land applied if any pollutant exceeds its *ceiling* concentration. Furthermore, packaged sewage sludge may not be land applied if any pollutant exceeds the *monthly average concentration* or *annual loading limit*. Sludges may not be applied to agricultural land, forest, public contact sites, or land reclamation sites if any pollutant exceeds either the *monthly average* or *cumulative limits*. If sludge is applied to a home garden or lawn, the pollutants may not exceed the *monthly average limits*.

Sewage sludges cannot be land applied if the ground is flooded, frozen, covered with snow, if it drains to a wetland or surface water, or is within 10 m of a surface water. Sewage sludge may not be land applied if it would adversely affect a threatened or endangered species.

Sewage sludge may be applied to agricultural land, forest, public contact sites, or land reclamation sites at or below the usual agronomic limits. For agricultural land, the agronomic rate is typically 10 tonne per ha per yr; for forest and public contact sites, it is 18 tonne per ha per yr; for reclamation sites, a single application of 112 tonne per ha is typical (Stein et al., 1995). Sewage sludges are applied annually to agricultural land between harvest and planting. It is applied to forest annually to once every few years.

**TABLE 11.23** Limits on Pollutants Applied to Land

Pollutant	Ceiling Concentration (mg/kg)	Monthly Average Concentration (mg/kg)	Annual Loading Rate (kg/ha/365 days)	Cumulative Loading Limit (kg/ha)
Arsenic	75	41	2.0	41
Cadmium	85	39	1.9	39
Chromium	3000	1200	150	3000
Copper	4300	1500	75	1500
Lead	840	300	15	300
Mercury	57	17	0.85	17
Nickel	420	420	21	420
Selenium	100	36	5.0	100
Zinc	7500	2800	140	2800

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