

VOLUME 11

**AERATION:
Principles
and Practice**

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2 Principles

2.1 MASS TRANSFER PRINCIPLES

2.1.1 PHYSICAL MECHANISMS INVOLVED IN TRANSFER

Mass transfer refers to the movement of molecules or mass from one location to another due to a driving force. This movement can occur within one fluid phase or among a number of fluid phases. Of particular concern to mass transfer in aeration is the transfer between two phases. This chapter specifically addresses the transfer between a gas and a liquid, which can be considered to occur in three stages. Oxygen molecules are initially transferred from a gas phase to the surface of a liquid. Equilibrium is quickly established at the gas–liquid interface. The oxygen molecules then move from the interface into the main body of the liquid.

The diffusion process in the liquid phase is initially considered with emphasis on the speed of diffusive transport and the factors influencing it. Interphase transport between the gas and the liquid is then addressed to establish the relationship between the oxygen saturation concentration in the liquid and the oxygen concentration in the gas phase. The basic equation describing the transfer of oxygen from the gas to the liquid phase is developed with the factors affecting the important parameters. Finally, the basic equations used for design are presented along with the relationship between process water conditions and the clean water conditions used in manufacturers' specifications for their equipment.

2.1.2 FICK'S LAW—QUIESCENT CONDITIONS

The principles defining the movement of oxygen molecules are similar to those defined in Newton's law, which governs the transfer of momentum in fluid flow, and Fourier's law, which defines the transfer of heat when a temperature gradient is present (Bird et al., 1960). The following equation, Fick's law, describes the transfer process when a concentration gradient is present in the fluid and no convection occurs. In this process, Brownian motion of the molecules in the fluid provides the transport.

$$J = -D \frac{dC}{dy} \quad (2.1)$$

The left-hand side of the equation provides the rate of mass transfer per unit interfacial area or mass flux. The negative sign indicates that transfer occurs in the direction of a decreasing gradient from a higher concentration to a lower value, similar to sliding down hill. The proportionality factor in the equation, D , represents the diffusion coefficient or diffusivity and is used to define the linear dependency of the flux on the associated gradient.

Figure 2.1 shows a schematic of the diffusive transport of oxygen molecules into a quiescent tank. The upper liquid layer is kept saturated by input of oxygen from the outside. The lower liquid layer initially is devoid of oxygen. Brownian motion causes both water and oxygen molecules to be transported across the interface between the two layers. Due to this random motion of molecules, oxygen begins to penetrate to the lower layers of the liquid in the "y" direction. Figure 2.2 shows the lower liquid layer when one-half of the total volume has attained saturation. It should be noted that penetration is not to the same depth in all locations due to the random nature of the diffusive process. Finally, at an infinite time, as shown in Figure 2.3, the total volume of the lower layer is saturated.

By conducting a mass balance on an elemental slice within the liquid layer, the differential equation describing the change in concentration with time is given by Fick's second law of diffusion (Bird et al., 1960) as:

$$\frac{\partial C}{\partial t} = D \frac{\partial^2 C}{\partial y^2}$$

The equation describing the time-space distribution of the oxygen penetration into the above tank is given by (Sherwood et al., 1975).

or

$$C(t, y) = C_0 + (C_s - C_0) \operatorname{erfc} \left[\frac{y}{2\sqrt{Dt}} \right] \quad (2.2)$$

$$C(t, y) = C_0 + 2(C_s - C_0) \phi \left[-\frac{y}{2\sqrt{Dt}} \right]$$

The complementary error function, erfc , and the cumulative Gaussian error function, ϕ , are available on spreadsheet programs and tabulated in statistics and engineering texts (Blank, 1982; Carslaw and Jaeger, 1959).

An example of the rate of molecular diffusion into the upper 5 mm of the tank in Figure 2.1 is given below using the following parameters at 20°C after one hour:

oxygen saturation concentration, $C_s = 9.09 \text{ mg/L}$, $D = 1.83 \cdot 10^{-9} \text{ m}^2/\text{s}$, $C_0 = 0 \text{ mg/L}$, initial oxygen free water.

$$C(t, y) = 0 + 2(9.09 - 0) \phi \left[-\frac{5 \cdot 10^{-3}}{\sqrt{2 \cdot 1.83 \cdot 10^{-9} \cdot 3600}} \right]$$

$$= 2(9.09) \phi[-1.377] = 2(9.09) \cdot 0.0844$$

$$= 1.53 \text{ mg/L or } 16.8\% \text{ of saturation.}$$

This process is slow as demonstrated further for a 0.5 m tank using Equation (2.2). Figure 2.4 illustrates that oxygen penetrates only to a depth of 10 mm after one hour, increasing to about 50 mm after one day. After 100 days, significant oxygen penetration occurs to mid-depth, taking almost one year to reach the bottom of the tank and over 10 years to come close to saturation.

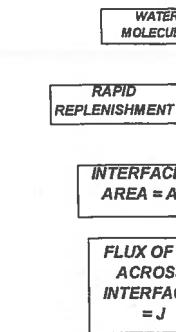


FIGURE 2.1 Oxy-

FIGURE 2.2 Oxy-

FIGURE 2.3 O-

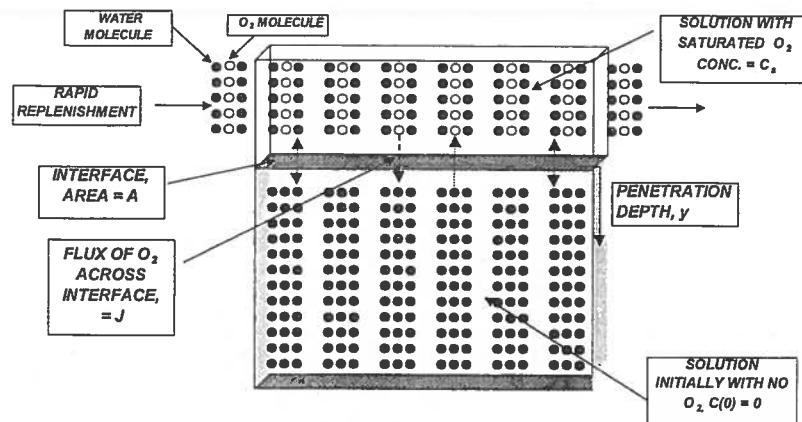


FIGURE 2.1 Oxygen diffusion schematic for quiescent solutions, $t = 0$.

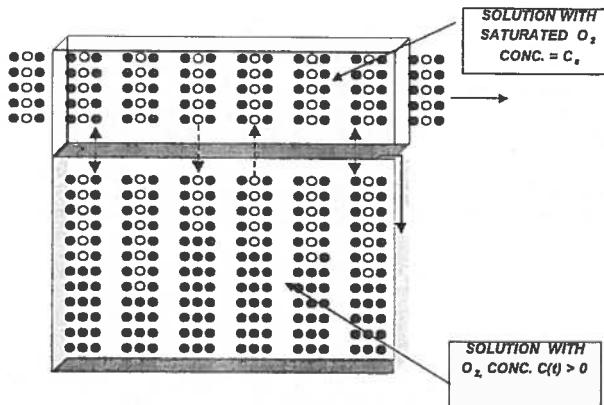


FIGURE 2.2 Oxygen diffusion schematic for quiescent solutions, $t = 1/2 t \text{ infinity}$.

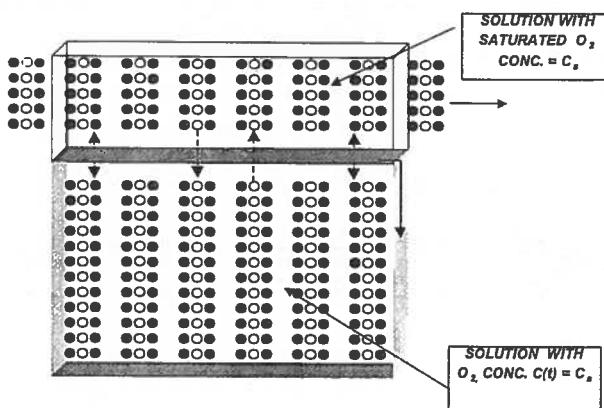


FIGURE 2.3 Oxygen diffusion schematic for quiescent solutions, $t = t \text{ infinity}$.

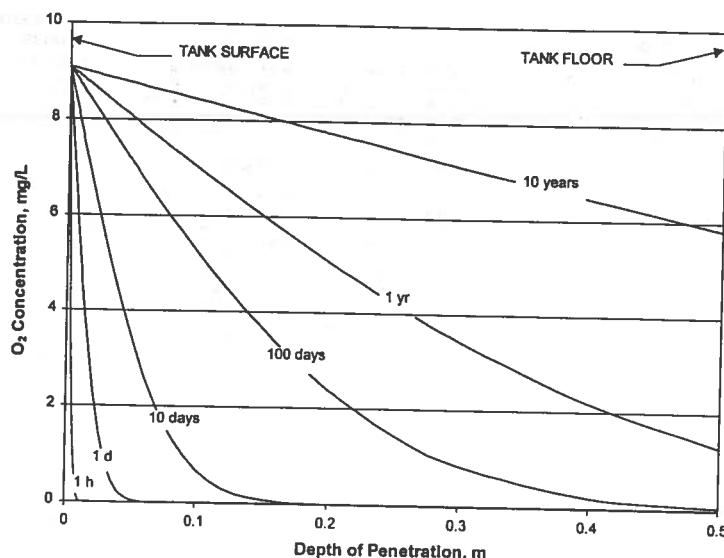


FIGURE 2.4 O_2 Profiles for molecular diffusion into a 0.5-m-deep tank.

Both the saturation and diffusivity values in Equation (2.2) are affected by temperature. Saturation decreases with increasing temperature (as discussed later), while diffusivity increases with temperature. The Wilke-Chang relationship (Reid et al., 1987) is an empirical correlation commonly used to describe the diffusivity, D_{AB} , of a dilute solution of A in solvent B as a function of molecular weight, M_B , and viscosity, μ_B , of the solvent, total volume, V_A , of the solute and absolute temperature, T .

$$D_{AB} = \frac{7.4 \times 10^{-12} T \sqrt{\phi M_B}}{\mu_B V_A^{0.6}} [=] \frac{m^2}{s} \quad (2.3)$$

When the solvent is water and the solute is dissolved oxygen, the Wilke-Chang expression is as follows.

$$D = \frac{6.85 \times 10^{-12} T}{\mu} [=] \frac{m^2}{s} \quad (2.4)$$

T is the absolute temperature in K, and μ is the viscosity of water in centipoises ($g/m \cdot s$). The viscosity of water decreases as temperature increases, and fluid exerts less resistance on the Brownian motion of the water molecules. Figure 2.5 illustrates the increase in diffusivity with increasing temperature according to the Wilke-Chang equation using $20^\circ C$ as the base. Note that the major impact of the temperature change on the diffusivity is due to the reduction in viscosity.

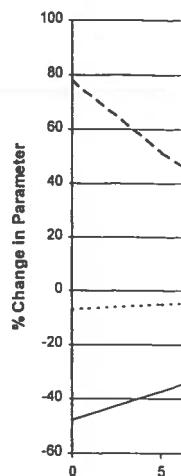


FIGURE 2.5 Relative change in diffusivity using Wilke-Chang equation.

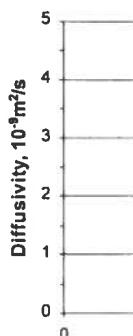


FIGURE 2.6 Effect of temperature on diffusivity.

An overall expression can be expressed as

Figure 2.6 shows using the typical heat of $2.1 \times 10^{-9} m^2/s$. The general profile.

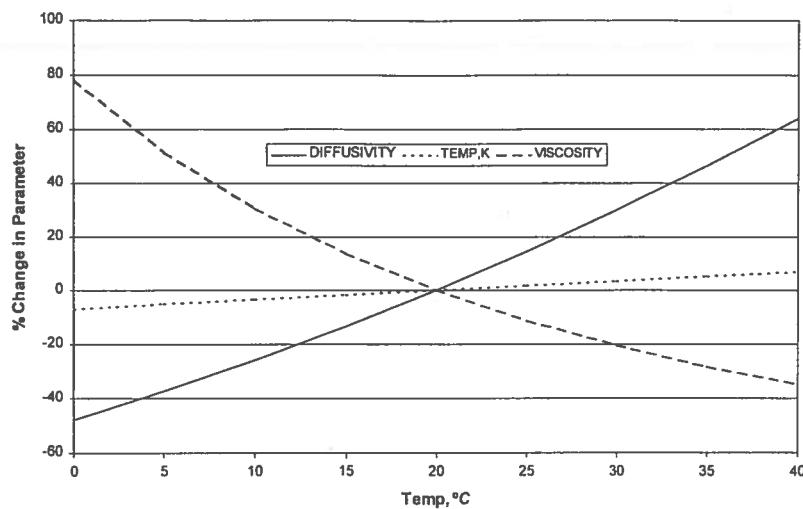


FIGURE 2.5 Relative effects of changes in temperature and viscosity on oxygen diffusivity using Wilke-Chang equation.

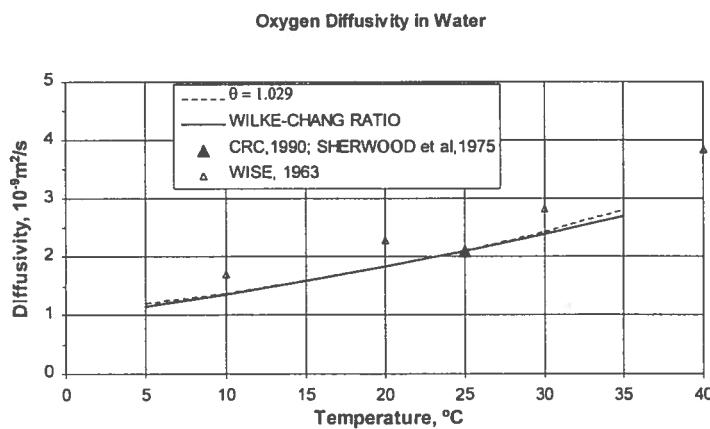


FIGURE 2.6 Effect of temperature on oxygen diffusivity.

An overall expression to relate the effect of temperature on the diffusivity value can be expressed as follows:

$$D_{t,^{\circ}C} = D_{20^{\circ}C} \theta^{t-20} \quad (2.5)$$

Figure 2.6 shows that a θ value of 1.029 fits the Wilke-Chang expression using the typical handbook value (Weast, 1989) for oxygen diffusivity at $25^{\circ}C$ of $2.1 \times 10^{-9} \text{ m}^2/\text{s}$. The data provided by Wise (1963) is somewhat higher but fits the general profile.

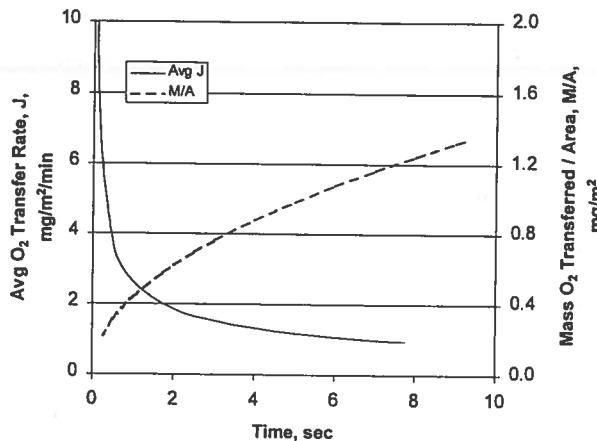


FIGURE 2.7 Initial rate and mass of oxygen transferred to water by Fick's diffusion at 20°C.

The total mass of oxygen transferred by diffusion, M , per unit interfacial area, A , into an infinitely deep tank (Sherwood et al., 1975), similar to the situation in Figure 2.1, is given as:

$$\frac{M}{A} = 2(C_s - C_0) \sqrt{\frac{Dt}{\pi}} \quad (2.6)$$

The average concentration, \bar{C} , attained over the depth of the tank, represented by d , can be obtained as follows:

$$\bar{C} = \frac{M}{Ad} = \frac{M}{V} = \frac{2(C_s - C_0)}{d} \sqrt{\frac{Dt}{\pi}} \quad (2.7)$$

The average flux of oxygen during the above time is obtained by dividing Equation (2.6) by the time of transfer to attain:

$$\bar{J} = \frac{M}{At} = 2(C_s - C_0) \sqrt{\frac{D}{\pi t}} \quad (2.8)$$

Figure 2.7 provides the average transfer rate, \bar{J} and total mass per unit area, M/A , during the first seconds of transfer. The initially high rates of transfer are quickly reduced as oxygen begins to build up in the layers adjacent to the interface. This outcome highlights the desirability of removing these upper layers by mixing them into the bulk solution (convective transport) to allow transfer to proceed more rapidly.

2.1.3 COMPARISON OF DIFFUSIVE TO CONVECTIVE TRANSPORT

Mixing and turbulence in the bulk solution destroy any concentration gradients in the major portion of the liquid with molecular diffusion occurring only in a thin

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2.1.4 GAS-

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layer at the interface. The mass flux is then defined in terms of the measured concentration difference and an empirically determined transfer coefficient, k_L , which represents the liquid film coefficient. This definition is expressed as follows.

$$J = k_L(C_s - C) \quad (2.9)$$

The mass flux can be expressed in terms of the change in the bulk liquid concentration by multiplying by the interfacial area per unit liquid volume, $a = \frac{A}{V}$.

$$J \frac{A}{V} = \frac{dC}{dt} = k_L a (C_s - C) \quad (2.10)$$

Integrating between the initial conditions and those at time, t , yields the following:

$$\int_{C_0}^C \frac{dC}{C_s - C} = k_L a \int_0^t dt$$

$$\frac{C_s - C}{C_s - C_0} = e^{-k_L a t} \quad (2.11)$$

When the initial concentration is zero, then the fraction saturation attained with time is given as follows.

$$\frac{C}{C_s} = 1 - e^{-k_L a t}; \quad C_0 = 0 \quad (2.12)$$

The fraction saturation obtained by molecular diffusion as a function of tank depth can be obtained by expressing Equation (2.7) as follows:

$$\frac{\bar{C}}{C_s} = \frac{2}{d} \sqrt{\frac{Dt}{\pi}}; \quad C_0 = 0 \quad (2.13)$$

Figure 2.8 shows the above two equations for a range of $k_L a$ values, from the high rates encountered in aeration tanks to the lower rates in natural water systems. To approximate the results from the field, it is obvious that molecular diffusion must occur in the thin, centimeters to microns surface layers of these systems. Turbulent or convective transport occurs over the bulk of the depth.

2.1.4 GAS-LIQUID TRANSFER

The mass transfer principles discussed above have not yet addressed the relationship between the gas and liquid phases. Figure 2.9 is a schematic of the two phases

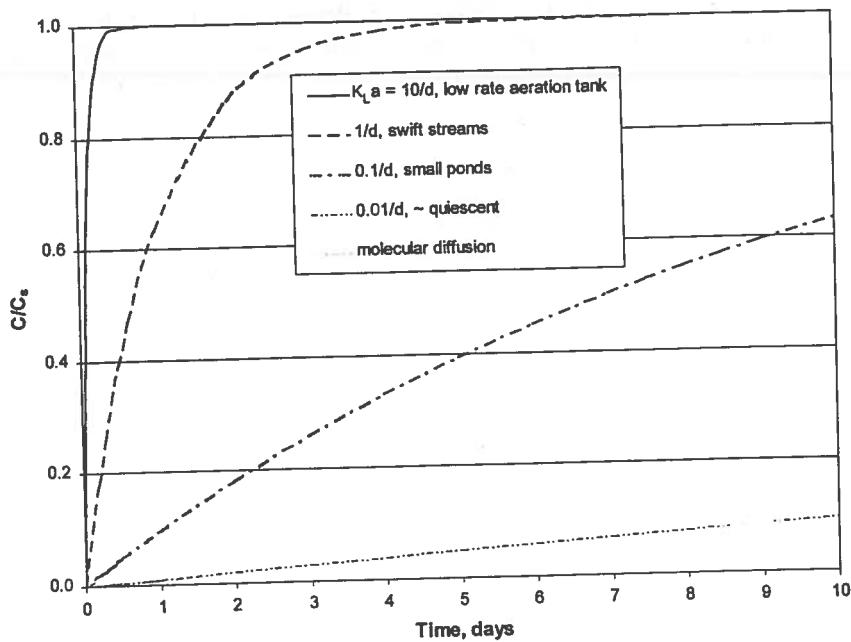


FIGURE 2.8 O₂ Transfer rates for field conditions compared to molecular diffusion at 20°C and 0.5 m depth.

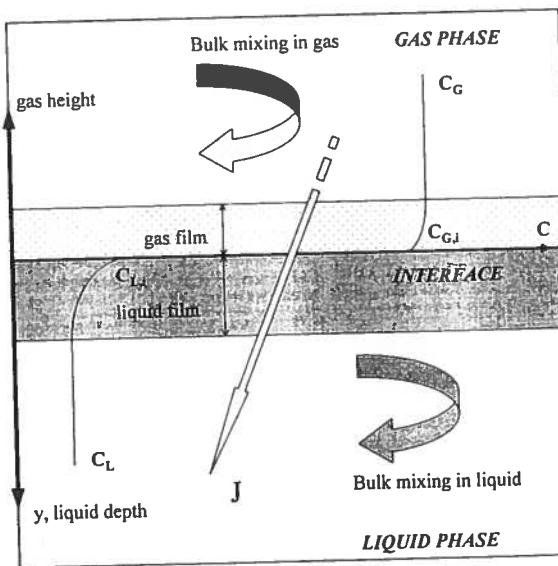


FIGURE 2.9 Two phase O₂ transfer schematic.

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showing two resistances to transfer, one in the gas phase and one in the liquid phase. The schematic also reveals a discontinuity occurring between the two phases.

2.1.4.1 Gas and Liquid Films

The oxygen flux is expressed using both liquid, k_L , and gas, k_G , film coefficients, similar to Equation (2.9), but with the concentration difference expressed in each phase from the bulk values, C_G and C_L , to the interface values, $C_{G,i}$ and $C_{L,i}$:

$$J = k_G(C_G - C_{G,i}) \quad \text{gas layer} \quad (2.14)$$

$$J = k_L(C_{L,i} - C_L) \quad \text{liquid layer} \quad (2.15)$$

Note that the oxygen flux through each layer is equal with no buildup of oxygen at the interface.

2.1.4.2 Henry's Law

The relationship between the concentrations at the interface is expressed by Henry's law as follows.

$$C_{G,i} = HC_{L,i} \quad (2.16)$$

This equation is an equilibrium relationship where the concentrations at the interface have the same activity or chemical potential (fugacity). Both concentrations are expressed in similar units, so H , the Henry's constant, is considered to be dimensionless, although actual units are $(\text{mg/L})_{\text{gas}} / (\text{mg/L})_{\text{liquid}}$. One must be careful when using handbook values for Henry's constant since it is also expressed as the inverse of the above and called a solubility or absorption coefficient.

2.1.4.3 Overall Driving Force

Combining the above three equations yields the following.

$$J = \left[\frac{1}{k_L} + \frac{1}{Hk_G} \right]^{-1} \left(\frac{C_G}{H} - C_L \right) \quad (2.17)$$

The first term in the above equation contains the resistances to transfer in both liquid, R_L , and gas, R_G , layers, while the driving force or concentration difference is expressed in terms of measurable concentrations in bulk gas and bulk liquid phases. The first term in brackets is the inverse of the total resistance to transfer (R_T) and can be expressed as follows.

$$\left. \begin{aligned} R_T &= R_L + R_G \quad \text{or} \\ \frac{1}{K_L} &= \frac{1}{k_L} + \frac{1}{Hk_G} \end{aligned} \right\} \quad (2.18)$$

K_L is the overall liquid film coefficient taking into account both gas and liquid phase resistances. The relative importance of both resistances can be evaluated using the following expression for the resistance due to the liquid film.

$$\frac{\%R_L}{100} = \frac{R_L}{R_T} = \frac{K_L}{k_L} = \frac{1}{1 + \frac{1}{H \frac{k_G}{k_L}}} \quad (2.19)$$

Using typical values of the gas to liquid film coefficient ratio, $\frac{k_G}{k_L}$, of 20 to 100, with a Henry's constant for oxygen of 29 at 20°C, shows that the liquid film resistance comprises more than 99.8 percent of the total resistance. The gas phase resistance is insignificant, typical of low solubility compounds such as oxygen and nitrogen. For oxygen transfer, $K_L \approx k_L$ and the gas side resistance can be ignored. Thus, turbulence and mixing has to be applied only to the liquid. The only impact of gas phase turbulence would be shear stress at the interface causing liquid phase turbulence.

2.1.4.4 Liquid Film Coefficient

There are a number of theories to describe the liquid film coefficient. Summaries of the earlier work, given in Sherwood et al. (1975), Aiba et al. (1965), and Eckenfelder and O'Connor (1961) are briefly reviewed here.

First proposed by Nernst in 1904, an equation for the two-film theory using stagnant gas and liquid films was derived by Lewis and Whitman in the 1920's to allow both gas and liquid resistances to be added in series. Through a gross simplification, linear concentration profiles were used in each of the films with sharp discontinuities between film and bulk phase concentration gradients. The liquid film coefficient was given as a function of a characteristic liquid film thickness, δ_L .

$$k_L = \frac{D}{\delta_L} \quad (2.20)$$

Although no predictive estimates of δ_L are available, it has been useful in predicting mass transfer rates with simultaneous chemical reaction based on data without reaction, as well as the impact of high mass transfer rates on heat transfer. Typical liquid films over which the concentration gradient occurs vary from 10 to 200 microns thick, depending on the level of turbulence in the bulk liquid (Hanratty, 1991).

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The penetration theory by Higbie in 1935 assumes a small fluid element at concentration, C_0 , is brought into contact with the interface for a short time, t , where diffusion into the element occurs as a transient process, decreasing with time. Equation (2.8) describes this process resulting in a value of the film coefficient as follows.

$$k_L = 2 \sqrt{\frac{D}{\pi t}} \quad (2.21)$$

(2.19)

The time of contact for bubble aeration is defined as the time for a single bubble to travel through liquid at a distance equal to its diameter, d_B , using the bubble velocity, v_B .

$$t = \frac{d_B}{v_B}$$

Mackay et al. (1991), summarizing results of Asher and Pankow from 1986 to 1990, illustrates the Higbie model gave a good description of CO₂ transfer through a clean air-water interface. The characteristic diffusional distance, given as

$$\delta_L = \sqrt{Dt}$$

was 42 μm at a contact time of 1 s. This thickness was much larger than the monomolecular interface thickness of 0.3 nm or 0.0003 μm.

Danckwerts (1951) expanded on the penetration theory by employing a wide spectrum of times instead of a single contact time, wherein an element of fluid would be exposed to the saturation concentration at the interface.

$$k_L = \sqrt{Dr} \quad (2.22)$$

The parameter, r , is the fractional rate of surface renewal.

In the three above models for the liquid film coefficient, values are not generally available except in the case of bubble aeration for the penetration model. Therefore, experimental measurement of the film coefficient is required.

O'Connor and Dobbins (1958) defined the surface renewal rate as a function of fluid turbulence parameters, a characteristic mixing length, l , and vertical velocity fluctuation, \bar{v} , as:

$$r = \frac{\bar{v}}{l}$$

This definition led to two expressions for the reaeration coefficient of streams based on the stream characteristics. One was for shallow streams where there is a

significant velocity gradient and shearing stress (nonisotropic turbulence), and the other was for deep streams where a significant velocity gradient and shearing stress do not exist (isotropic turbulence). In the case of deep streams, this expression led to the widely used equation for determining the stream reaeration coefficient based on stream velocity and depth.

$$\left. \begin{aligned} k_L &= \sqrt{\frac{DU}{H}} \\ k_L a &= \frac{k_L}{H} = \frac{DU^{1/2}}{H^{3/2}} \end{aligned} \right\} \text{deep streams} \quad (2.23)$$

O'Connor (1983) went further to describe the overall resistance to oxygen transfer as two resistances in series, similar to the two-film theory but both in the liquid film. A viscous laminar sublayer is adjacent to the interface and the other a turbulent mixed zone between the laminar sublayer and the bulk fluid.

$$\frac{1}{k_L} = \frac{1}{k_\delta} + \frac{1}{k_r}$$

Brumley and Jirka (1988), pg 316, indicate that the above conceptual models are on the right track. They attempt "to describe a process where dissolved gas enters a boundary layer by molecular diffusion and is subsequently transported into the bulk by turbulent mixing in such a way that the boundary layer remains thin". Recent evaluations of the liquid film coefficient consider the hydrodynamics near the interface with the velocity fluctuations normal to the interface (Hanratty, 1991). Hydrodynamic models describing eddy motion are being developed for relatively smooth surfaces and are not capable of addressing the complex situations in aeration tanks where the interfacial area is not known.

Clearly, there is no simple theoretical expression for the liquid film coefficient that would be suitable for all types of aeration systems. It will be a function of the energy input to the system, the interfacial area developed, and the hydrodynamics and velocity profile at the interface. Thus, the interfacial area is generally combined with the overall liquid film coefficient and data from empirical correlations are used to design systems.

2.2 APPLICATION TO OXYGEN TRANSFER

2.2.1 BASIC EQUATION

The oxygen saturation concentration, C^* , is defined as the value in equilibrium (at infinite time) with the concentration in the bulk gas phase, which is also the concentration at the interface since the gas side gradient is negligible.

$$C^* = \frac{C_G}{H} \quad (2.24)$$

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Substituting Equations (2.18) and (2.24) into (2.17) yields the oxygen flux.

$$J = K_L(C^* - C_L) \quad (2.25)$$

Multiplying by the interfacial area per unit volume, the change in oxygen concentration with time, similar to Equation (2.10) results.

(2.23)

$$\frac{dC_L}{dt} = K_L a(C^* - C_L) \quad (2.26)$$

Equation (2.26) is the basic equation used to describe oxygen transfer in actual aeration systems. The maximum rate of transfer occurs when the dissolved oxygen concentration in solution is zero. No transfer occurs when the dissolved oxygen concentration has attained equilibrium with the gas phase.

The oxygen transfer coefficient, $K_L a$, is the product of the liquid film coefficient, K_L and the interfacial area exposed to transfer in a given liquid volume, a . In all but the simplest systems, the individual values, K_L and a , are impossible to individually measure. Incorporating them into one coefficient, $K_L a$, provides the ability to obtain a measurable value in complex field aeration systems.

The saturation value, C^* , is also a measured value in aeration systems. Although oxygen saturation values in equilibrium with bulk atmospheric gas concentrations at the liquid surface have been tabulated, these conditions do not necessarily exist in aeration tanks. The actual values are impacted, especially for diffused aeration systems, by increased pressure from the release of gas below the water and by decreased bulk gas concentrations resulting from the transfer process of gas rising through the liquid.

2.2.2 FACTORS AFFECTING OXYGEN TRANSFER

From the basic equation defining oxygen transfer, Equation (2.26), the factors affecting each of the major parameters are discussed below.

2.2.2.1 Oxygen Saturation, C^*

Using the Henry's law definition for the saturation value, Equation (2.24), the oxygen saturation value is a function of both the oxygen gas phase concentration and the Henry's constant. From the ideal gas law

$$C_G = \frac{nM}{V} = \frac{pM}{RT} \quad (2.27)$$

For dry air, oxygen is 20.95 percent by volume, thus the oxygen partial pressure, p , is related to the total pressure, p_t , by:

(2.24)

$$p = 0.2095(p_t - p_v) \quad (2.28)$$

For open systems, both surface and diffused, the vapor pressure, p_v , is assumed saturated at the liquid temperature, with gas phase temperature having no effect on the vapor pressure or C_G . Only in well mixed closed systems, where there are significant differences in gas and liquid phase temperatures, would vapor pressures at the gas phase temperature be utilized (Mueller, 1979).

The total pressure is related to both the barometric pressure, P_b , and increased pressure from aerator submergence.

$$P_t = P_b + p_{d_e} \quad (2.29)$$

An effective pressure, p_{d_e} , is determined from shop or field data for specific equipment. Previous theoretical relationships for this term have proven faulty due to the complexity of mixing patterns in aeration systems.

2.2.2.1.1 Temperature

The Henry's law constant, H , increases with increasing temperature and dissolved solid concentrations, which causes a reduction in the oxygen saturation value. The Henry's constants for oxygen in Table 2.1 are back calculated from the observed oxygen saturation values from Benson and Krause (1984) and *Standard Methods* (APHA et al., 1995) at one atmosphere total pressure and no dissolved solids (0 chlorinity), C_s^* .

In specifying aerator performance, 20°C is used as a standard condition with the saturation value at one atmosphere total pressure. The temperature correction factor for the saturation value, τ , is then given by the following equation and illustrated in Figure 2.10.

$$\left. \begin{aligned} \tau &= \frac{C_s^*}{C_{s20}^*} \\ C_{s20}^* &= 9.09 \frac{mg}{L} \end{aligned} \right\} \quad (2.30)$$

2.2.2.1.2 Wastewater

To account for the effect of wastewater constituents on oxygen saturation, a β factor is introduced as the ratio of saturation in wastewater to tap water.

$$\beta = \frac{C_{s\text{wastewater}}^*}{C_s^*} \quad (2.31)$$

The major impact on wastewater saturation value is the inorganic dissolved solids. The chlorinity data in *Standard Methods*, (APHA et al., 1995), was scaled up to total dissolved solids using NaCl ($1.65 \times$ chlorinity) from 0 to 20,000 mg/L TDS. As indicated in *Standard Methods*, this scale-up, shown in Figure 2.11, assumes that the wastewater inorganic composition is similar to that in seawater. It is the

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TABLE 2.1
Henry's Constants for Oxygen as
a Function of Temperature

| Temperature, °C | C_s^* , mg/L | $H, \frac{(mg/L)_{air}}{(mg/L)_{water}}$ (*) |
|--------------------|-------------------|--|
| 0 | 14.62 | 20.3 |
| 10 | 11.29 | 25.1 |
| 20 | 9.09 | 29.8 |
| 30 | 7.56 | 34.0 |
| 40 | 6.41 | 37.6 |

$$(*)H = \frac{5530(14.7 - p_v(psia))}{C_s^* T(K)}$$

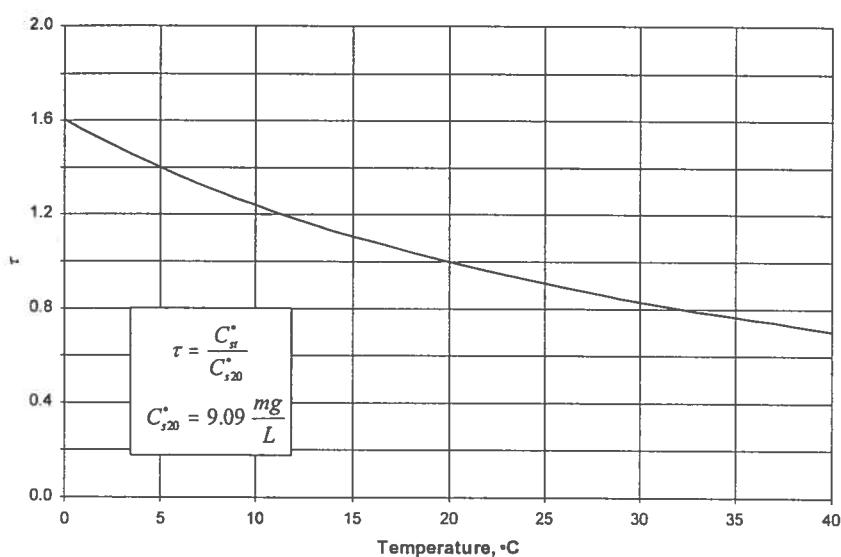


FIGURE 2.10 Effect of temperature on oxygen saturation.

consensus of the ASCE Committee on Oxygen Transfer Standards that this scale-up factor is sufficiently accurate for practical use (ASCE, 2001).

$$\beta = 1 - 5.7 \times 10^{-6} \times TDS \quad (2.32)$$

For municipal wastewater at TDS < 1500 mg/L, β is commonly taken as 0.99. For industrial wastewater such as pharmaceutical waste at a TDS of 10,000 mg/L, β will be as low as 0.94.

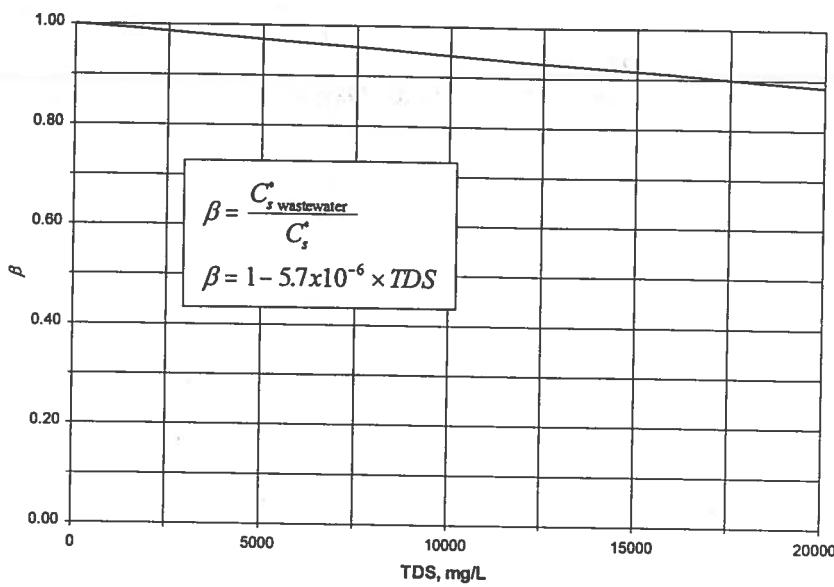


FIGURE 2.11 Effect of total dissolved solids on oxygen saturation.

In an evaluation of saturation values in upper Mississippi waters in Minnesota, Parkhill and Gulliver (1997) recommend taking additional DO measurements on distilled water samples to remove any bias associated with Winkler measurement and DO probe calibration errors. This correction factor is laudable to obtain continual calibration update of the DO probes. However, they also have taken Winkler measurements on the river samples and found β values lower than predicted by a TDS correction in May and June, 1994. Therefore the adequacy of the above TDS correction approach was questioned. It is the authors' opinion that as the river temperature warmed in the spring, algae growth may have occurred and caused an organic interference with the Winkler and not a true β value. Until further demonstration of the ability to run accurate titrimetric tests on water with differing organic concentrations, the above correction factor is recommended.

2.2.2.1.3 Submergence

At standard conditions of temperature (20°C) and pressure (1 atm), the effect of diffuser submergence on oxygen saturation is given by δ .

$$\delta = \frac{C_{s20}^*}{C_s^*} = \frac{P_s + P_{d_e} - p_v}{P_s - p_v} \quad (2.33)$$

Since δ is the measured value, the effective pressure can be defined.

$$P_{d_e} = (\delta - 1)(P_s - p_v) = \gamma_w d_e \quad (2.34)$$

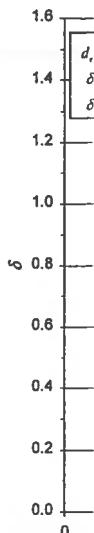


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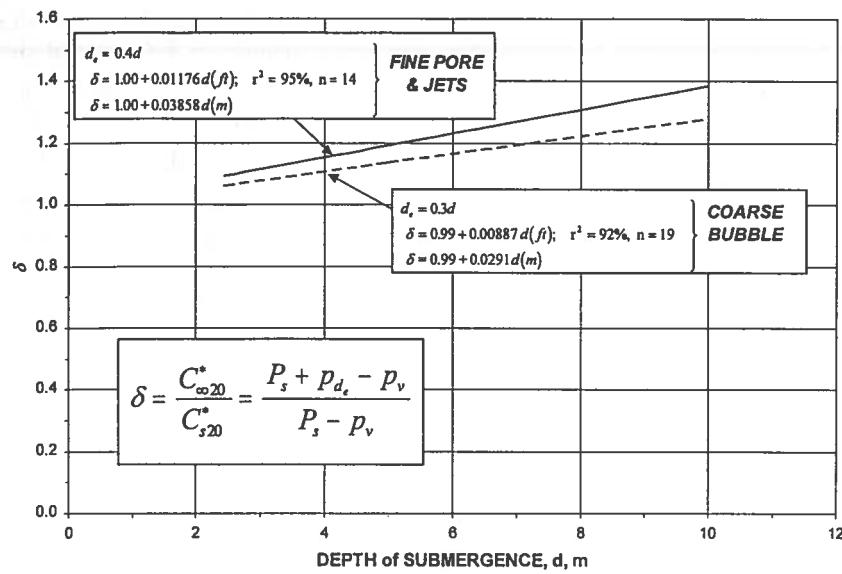


FIGURE 2.12 Effect of diffuser submergence on oxygen saturation.

The term d_e , representing the effective depth, is typically given as a fraction of the total depth of submergence, d .

To determine δ in the field, seven types of diffusers were used in clean water studies by Yunt et al. (1980), Mueller et al. (1982b), Mueller and Sauer (1986), and Mueller and Sauer (1987). Coarse bubble units provided significantly lower saturation values than fine pore and jet diffusers, as shown in Figure 2.12 and given below.

$$\left. \begin{array}{l} d_e = 0.4d \\ \delta = 1.00 + 0.01176d(\text{ft}) \quad r^2 = 95\%, n = 14 \\ \delta = 1.00 + 0.03858d(\text{m}) \end{array} \right\} \text{Fine Pore and Jets} \quad (2.35)$$

$$\left. \begin{array}{l} d_e = 0.3d \\ \delta = 0.99 + 0.00887d(\text{ft}) \quad r^2 = 92\%, n = 19 \\ \delta = 0.99 + 0.0291d(\text{m}) \end{array} \right\} \text{Coarse Bubble} \quad (2.36)$$

2.2.2.1.4 Barometric Pressure

The impact of barometric pressure on saturation is given by Ω , shown in Figure 2.13 and given as follows:

(2.33)

(2.34)

$$\Omega = \frac{C_{\infty P_b}^*}{C_{\infty P_s}^*} = \frac{P_b + p_{d_e} - p_v}{P_s + p_{d_e} - p_v} \approx \frac{P_b}{P_s} \quad (2.37)$$

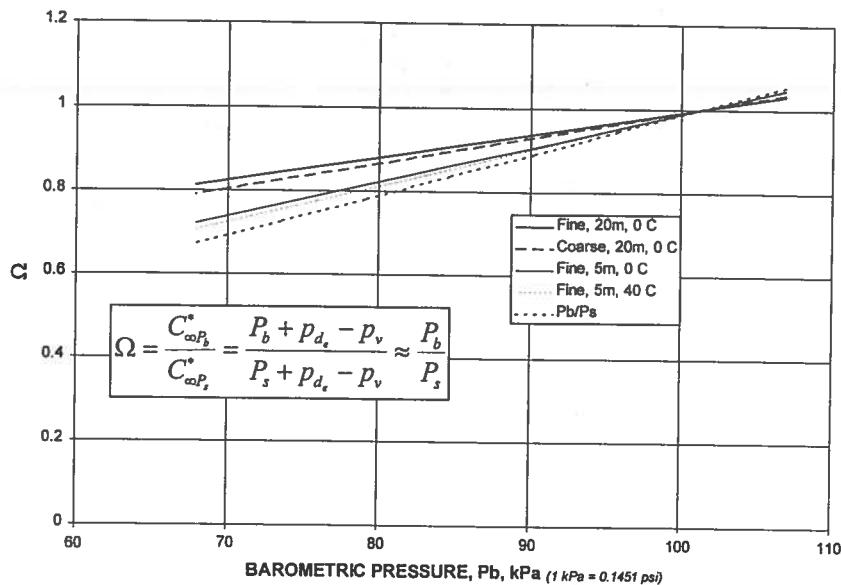


FIGURE 2.13 Effect of barometric pressure on oxygen saturation.

The approximation, which always gives lower values than actual, is satisfactory for tanks under six meters (20 feet) (ASCE, 1991) at barometric pressures and within five percent of standard atmospheric pressure. At greater aeration tank depths, the impact of reduced barometric pressure is less due to the large effect of hydrostatic head. Coarse bubble diffusers give slightly lower Ω values than fine pore, and vapor pressure has a minimal effect at normal temperatures. Since barometric pressure decreases with altitude, the following factor can be used for high altitude locations (Metcalf and Eddy, 1972).

$$P_b = P_{b0} \left(1 - \frac{alt(ft)}{30,000} \right) = P_{b0} \left(1 - \frac{alt(m)}{9100} \right)$$

For example, Denver, Colorado, at an altitude of 1500 meters (5000 feet), will have a surface saturation value of 7.6 mg/L at 20°C compared with 9.09 mg/L at sea level.

Using the above correction factors, the actual saturation value in an aeration tank under process conditions is given as $C_{\infty f}^*$.

$$\left. \begin{aligned} C_{\infty f}^* &= \tau \beta \Omega C_{\infty 20}^* \\ C_{\infty 20}^* &= \delta C_{s20}^* \end{aligned} \right\} \quad (2.38)$$

2.2.2.2 Oxygen Transfer Coefficient, $K_L a$

Both the liquid film coefficient and the interfacial area through which transfer occurs are affected by the type of aeration equipment employed and the turbulence level

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in the system. Data available from manufacturers on their specific equipment operating in tap water are typically given as a function of gas flow or power input at a temperature of 20°C. Therefore, the data must be adjusted to account for the temperature in the aeration tank and for the wastewater constituents.

2.2.2.2.1 Temperature

Increasing temperature increases $K_L a$ similar to the effect on diffusivity and liquid film coefficient using the following relationship.

$$K_L a_t = K_L a_{20} \theta^{t-20} \quad (2.39)$$

In the above equation, θ is dimensionally not homogeneous requiring a temperature in °C. An alternative would be to express the temperature impact in the exponential form.

$$K_L a_t = K_L a_{20} e^{\kappa(t-20)}$$

$$\kappa = \ln \theta$$

At present, the above has not been used in the aeration field, but would be a logical direction for the future. The value of θ is commonly taken as 1.024 (ASCE, 1993; Jensen, 1991), equivalent to κ of 0.0237/°C.

As indicated previously, the liquid film coefficient, K_L , is a function of diffusivity raised to the power of 0.5 to 1.0. Using the Wilke-Chang correlation, this equation would result in θ values of 1.028 and 1.029 respectively. For diffused saran tube and sparger aeration units, Bewtra et al. (1970) have measured a value of 1.02 while Landberg et al. (1969) have found a lower θ of 1.012 for surface aeration units. Figure 2.14 and 2.15 show θ for static mixers and dome diffusers (Mueller et al., 1982a; Mueller et al., 1983a) to vary from 1.028 at low gas flows (low turbulence levels) to 1.017 at high gas flows. Metzger and Dobbins (1967) have determined the average θ values for the liquid film coefficient to be 1.032 for low intensity mixing and 1.006 for high intensity mixing. Jensen (1991) correlating K_L data over three orders of magnitude has shown θ to decrease from a value of 1.047 at low turbulence to 1.006 at higher turbulence levels.

Since temperature also affects viscosity and surface tension, changes in the interfacial area as well as K_L may also result. Lacking information on this relatively complex impact of temperature, data on specific aeration systems is required from manufacturers if accurate temperature corrections are to be obtained.

2.2.2.2.2 Wastewater

The presence of dissolved organics in wastewater can have a significant effect on $K_L a$, typically much greater than all the other factors combined. An experimentally measured parameter, α , is defined to account for the wastewater effects.

$$\alpha = \frac{K_L a_{wastewater}}{K_L a_{tap\ water}} = \frac{K_L a_f}{K_L a} \quad (2.40)$$

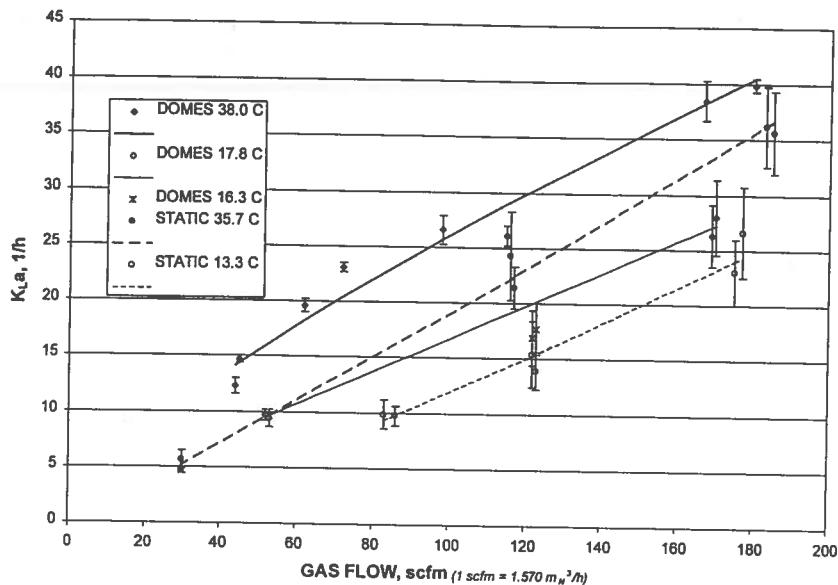


FIGURE 2.14 Effect of temperature on clean water K_La for dome and static aerators in a 9.1 m deep pilot plant.

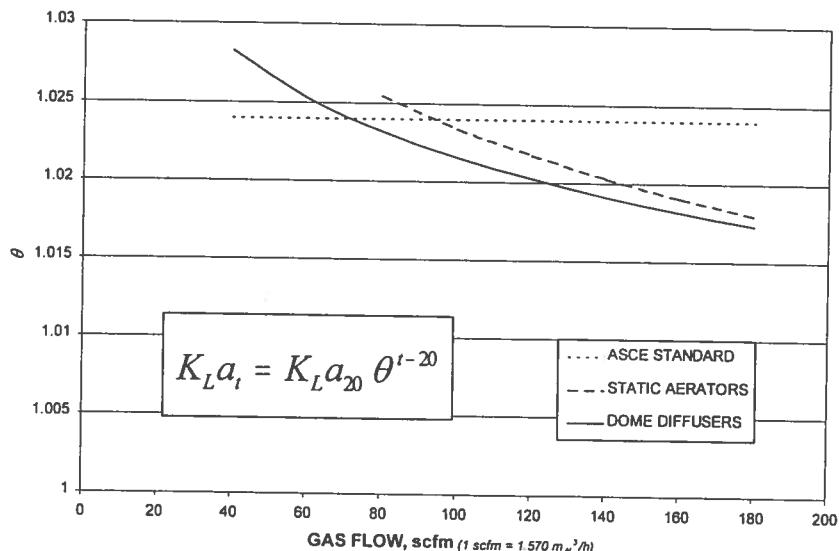


FIGURE 2.15 Effect of gas flow on θ for dome and static aerators in a 9.1 m deep pilot plant.

Surface active agents affect K_La due to a reduction in the liquid film coefficient but with an increase in surface area due to lowered surface tension (Wagner and Poepel, 1995). According to Mancy and Okun (1965), the resistance to oxygen transfer is caused mainly by a viscous hydration layer at the water surface and to a lesser extent by the interfacial film of adsorbed surfactant molecules.

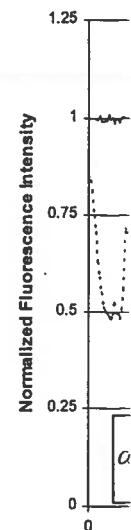


FIGURE 2.16 using fluoresce

Asher and co-workers (1995) conducted experiments using vertically oscillating air bubbles in surface reactors to study the effect of surface active agents on oxygen transfer. The figure shows the fluorescence intensity over time, which remains constant for a period before dropping sharply, indicating a change in the surface condition or oxygen transfer process.

When a surface active agent is added to a liquid, it can reduce the surface tension and increase the surface area, which can affect the mass transfer coefficient. The effect depends on the concentration and type of surfactant. At low concentrations, the surfactant molecules adsorb onto the liquid-gas interface, forming a monolayer that reduces the surface tension. As the concentration increases, the surfactant molecules begin to form aggregates, such as micelles, which can further reduce the surface tension and increase the surface area available for mass transfer.

Eckenfelder (1995) has shown that the effect of surfactants on bulk oxygen transfer is dependent on the surfactant concentration. At low concentrations, the surfactant can reduce the surface tension and increase the surface area, leading to an increase in the oxygen transfer rate. However, at high concentrations, the surfactant can form aggregates that can reduce the surface area and increase the surface tension, leading to a decrease in the oxygen transfer rate.

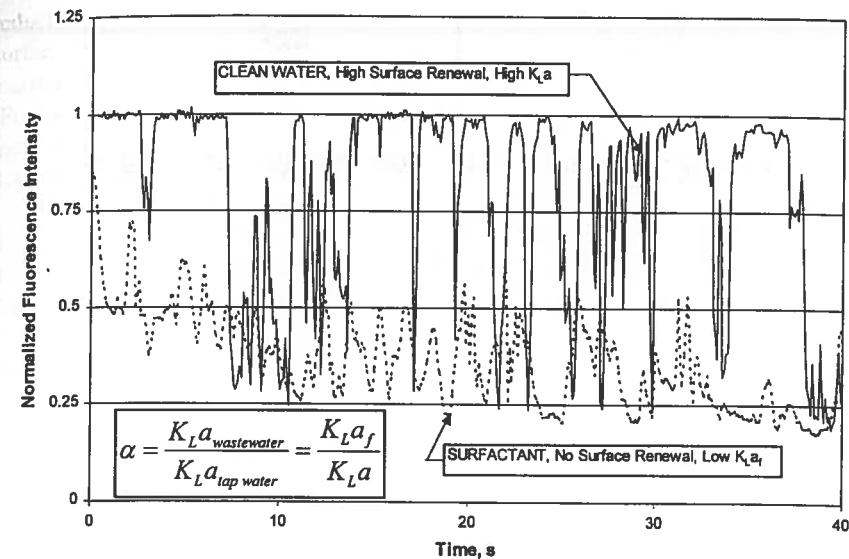
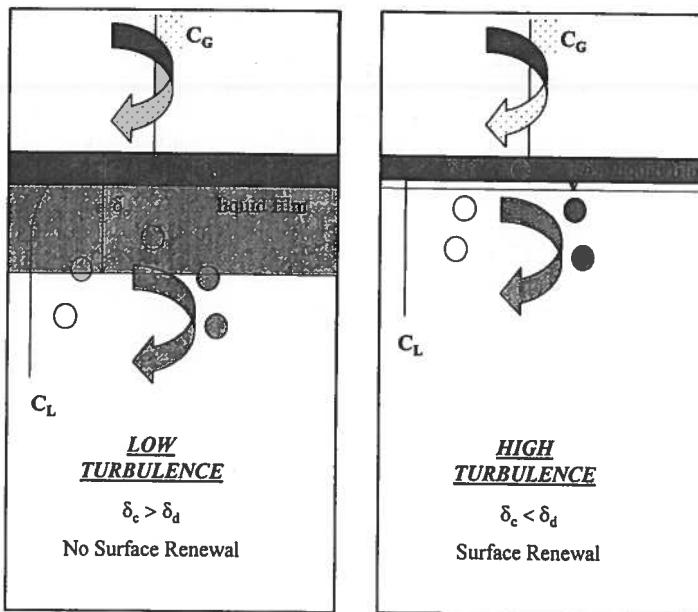
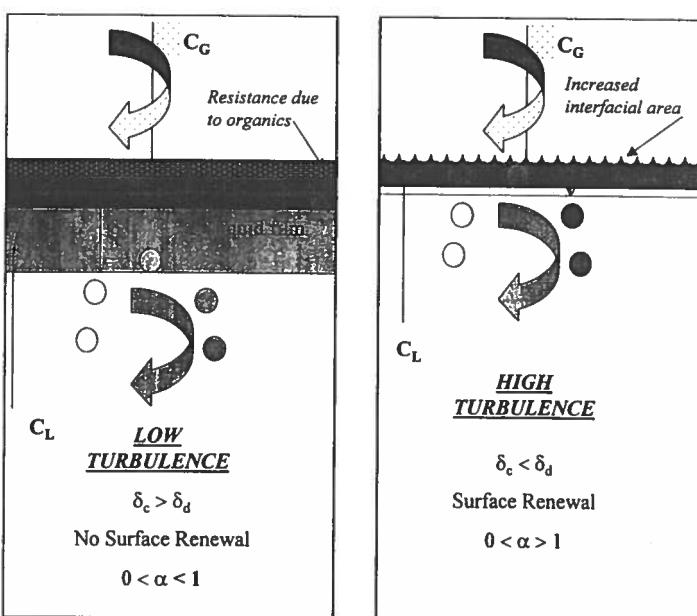


FIGURE 2.16 Impact of wastewater on $K_L a$ shown by velocity fluctuations at a water surface using fluorescence measurement (Asher and Pankow, 1991b; Asher, 1998).

Asher and Pankow (1991a,b) using surface fluorescence fluctuations with a vertically oscillating grid at different turbulence levels, showed a marked difference in surface renewal rates at the interface with and without a surfactant. Figure 2.16 shows the greater frequency and magnitude of the surface renewal rates in the clean water compared with that of the surfactant (Asher, 1998). As the peaks approach a value of 1.0, a high degree of surface renewal is occurring, typical of the clean water data. Figure 2.17 and 2.18 show the impact of turbulence and surfactants on the transfer process. For clean interfaces at low turbulence levels, the eddy caused by bulk mixing does not reach the surface. The concentration boundary layer, δ_c , is greater than the diffuse sublayer thickness, δ_d (~40 microns), so diffusion does not have enough time to saturate the eddy before returning to the bulk solution. For high turbulence, the eddy reaches the surface where it becomes saturated in the exposure time (~1 s) and then mixes into the bulk fluid. At these turbulence levels, the concentration boundary layer, δ_c , is less than the diffuse sublayer thickness, δ_d , where diffusion has enough time to saturate the eddy before leaving the surface.

When a surfactant is present at low turbulence levels, the concentration profile attains a greater depth, and δ_c increases. The additional resistance due to organics reduces transfer rate when compared with clean water. Due to the surfactant damping the turbulent motion through an increased shear stress at the interface, concentration fluctuations were never observed in the diffusive sublayer of Asher and Pankow, regardless of turbulence intensity of the grid system.

Eckenfelder (1970) indicates that for quiescent or laminar flow conditions, the bulk resistance to oxygen transfer is high and masks the surface resistance caused by the surfactant. In an intermediate range, low turbulence conditions, the bulk resistance to transfer is reduced and the surfactant interfacial resistance causes a significant

FIGURE 2.17 Turbulence impact on clean water O₂ transfer.FIGURE 2.18 Turbulence impact on wastewater O₂ transfer.

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reduction in transfer rate. At high turbulence levels, oxygen transfer depends on surface renewal and again, is not significantly affected by diffusion through interfacial resistances. Under these conditions, α may be >1.0 due to increased interfacial area (Figure 2.18). Both Lister and Boon (1973) and Otoski et al. (1979) contend that the increase in surface area does not offset the decrease in K_L with α always being less than one, which is most likely the case in full-scale systems.

For bubble systems, nonionic surfactants reduce oxygen transfer more strongly than anionic surfactants (Wagner and Poepel, 1995). They also show that surface tension measurements alone cannot be used to predict α values. Masutani and Stenstrom (1991) show that a measurement of dynamic surface tension was a potentially useful tool to determine the impact of surfactants on α . They also indicate that use of antifoam agents significantly decrease α .

During the course of biological oxidation of wastewater, the substances causing variations in $K_L\alpha$ are being removed. Thus, in a plug flow aeration tank, α will normally increase as flow progresses down the tank. Completely mixed, step feed, and selector processes (Mueller et al., 1996, 2000) will tend to minimize this large variation in α and operate closer to the effluent value.

After an aeration system has been operational for a time, field-measured $K_L\alpha_f$ values include not only the effect of the dissolved organics in the wastewater but also any deterioration in aerator characteristics. This effect is frequently found in fine pore diffusers when clogging or embrittlement occurs. An additional factor, F , is used to account for this diffuser aging process.

$$F = \frac{K_L\alpha_{f\text{service}}}{K_L\alpha_{f\text{new}}} \quad (2.41)$$

2.2.2.3 Dissolved Oxygen Concentration in Bulk Liquid, C_L

In setting a C_L value, two factors must be considered: the minimum dissolved oxygen concentration required by the activated sludge to maintain the maximum oxygen utilization rate, and the varying oxygen demands due to flow and organic load variations.

Activated sludge consists of microorganisms, the majority of which exist in biological floc particles. Data by Borkowski and Johnson (1967) indicate that a low oxygen concentration of 0.0004 mg/L is sufficient to maintain full activity of dispersed cells oxidizing carbonaceous organics. For oxygen to reach the active sites at the bacterial cell membranes, it must penetrate the liquid film surrounding the floc particle and diffuse through the floc matrix to the individual bacteria. Assuming a uniform oxygen uptake rate in the floc, the drop in dissolved oxygen concentration from the floc surface to the center of a spherical floc is given as follows (Wuhrmann, 1963).

$$C_L = C_m + \frac{A\gamma_f d_f^2}{24D_f}$$

Larger size floc particles and higher oxygen uptake rates require higher dissolved oxygen values as shown in Figure 2.19 (Mueller 1979). The greater floc sizes had

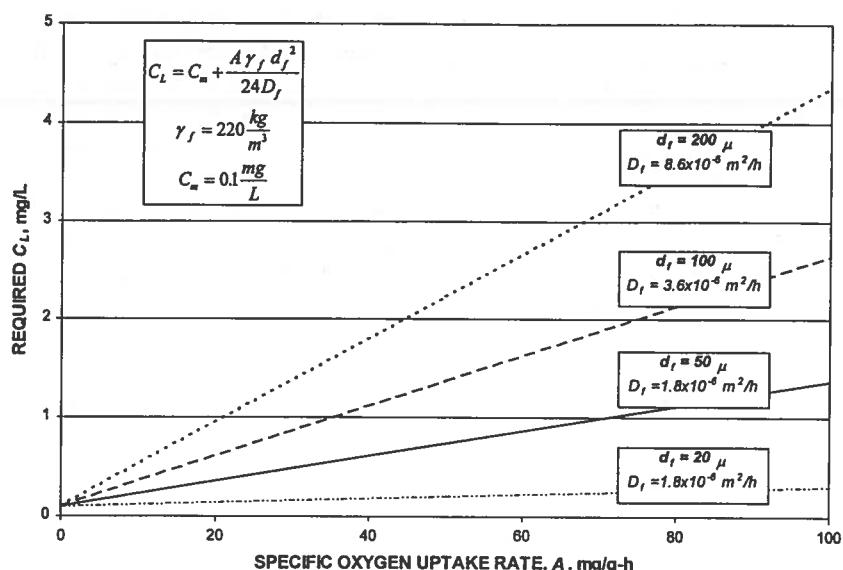


FIGURE 2.19 Impact of activated sludge mass transfer resistance on required O₂ concentration.

larger effective diffusivities. Argaman et al. (1995) shows that the effective diffusivity increases with increasing sludge volume index and specific surface area probably due to an increase in floc porosity. Activated sludge from an aeration tank at the Nancy (France) Metropolitan wastewater treatment plant had a mean diameter of 125 μ (Snidaro et al., 1997). Analysis after sonification revealed that the large floc were made up of more tightly bound 13 μ size microcolonies, which were in turn composed of 2.5 μ bacteria. A gel-like matrix of exopolymers provides the cohesion for these units. The loosely bound large floc should have greater porosity than the smaller more tightly bound floc, resulting in higher diffusivities.

For the typical size of activated sludge floc, 20 to 115 μ (Mueller et al., 1966), a dissolved oxygen concentration between 0.2 and 1.5 mg/L, typically 0.5–0.7 mg/L, is desirable. This parameter will insure the oxygen uptake rates of bacteria oxidizing carbonaceous organics are not oxygen limited. For nitrification to proceed at optimum rates, dissolved oxygen values > 2.0 mg/L are required (EPA, 1975). Stenstrom and Song (1991) show that the DO concentration for nitrification ranges from 0.5 to 2.5 mg/L depending on operational parameters and mass transport resistance. This level can go as high as 4.0 mg/L during an organic shock load.

To allow for variation in oxygen demand due to changing loads, a design C_L value of 2.0 mg/L is often used based on average load. Maximum load conditions should be evaluated to insure that C_L is above 0.5 mg/L to avoid septic conditions.

2.3 DESIGN EQUATIONS

In designing aeration systems, the basic equation used for the analysis is Equation (2.26), which is modified to account for the conditions at which manufacturers

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TABLE 2.2
Standard Conditions for Specification of
Aeration Equipment Performance

| Parameter | Condition U.S. Practice | Condition European Practice |
|---------------------|--|---|
| Type water | Tap water | Tap water |
| Water temperature | 20°C | 20°C |
| C_L | 0 mg/L | 0 mg/L |
| Barometric pressure | 1 atm | 1 atm |
| Air flow | 20°C and 36% relative humidity, $\gamma = 0.075 \text{ lb air/ft}^3$ $= 0.01736 \text{ lb O}_2/\text{ft}^3$ | 0°C and 0% relative humidity, $\rho = 1.293 \text{ kg air/m}^3$ $\approx 300 \text{ g O}_2/\text{m}^3$ |

specify the capabilities of their equipment. Specifications for aeration equipment are given based on clean water data under the conditions in Table 2.2 (ASCE, 1991; ATV, 1996).

2.3.1 STANDARD OXYGEN TRANSFER RATE, SOTR

The SOTR is the mass of oxygen transferred per unit time into a given volume of water and reported at standard conditions. The European literature also refers to this term as the oxygenation capacity (OC). The nomenclature used in the ASCE Standard is utilized throughout this text and the alternate value indicated as done here. Equation (2.26) is multiplied by the aeration tank volume and standard conditions employed.

$$SOTR = V \left(\frac{dC_L}{dt} \right)_{STD} = K_L a_{20} C_{\infty 20}^* V \quad (2.42)$$

Note that at standard conditions, the dissolved oxygen concentration is taken as zero thus providing the maximum driving force for transfer. As these equations are developed, an example calculation is performed in both the English and SI systems so that the units' conversion factors are clear (Table 2.3).

TABLE 2.3
SOTR Example Calculation

| SI | U.S. |
|--|---|
| $SOTR = 10.5 \frac{\text{mg}}{\text{L}} \times 1000 \text{ m}^3 \times \frac{8}{\text{h}} \times 10^{-3} \frac{\text{kg} \cdot \text{L}}{\text{mg} \cdot \text{m}^3}$ $= 84.0 \frac{\text{kg}}{\text{h}}$ | $SOTR = 10.5 \frac{\text{mg}}{\text{L}} \times 0.264 \text{ MG} \times \frac{8}{\text{h}} \times 8.34 \frac{\text{lb} \cdot \text{L}}{\text{mg} \cdot \text{MG}}$ $= 185 \frac{\text{lb}}{\text{h}}$ |

TABLE 2.4
Example Calculation for Specific Oxygenation Capacity, *oc*

SI and U.S.

$$oc = 10.5 \frac{mg}{L} \times \frac{8}{h} = 84 \frac{mg}{L \cdot h} = 84 \frac{g}{m^3 \cdot h}$$

The conditions for this computation will be an aeration tank of 1000 m³ (0.264 MG) at a water depth of 4.57 m (15 ft) with fine pore diffusers located at 4.27 m (14 ft) below the water surface. The saturation value calculated from Equation (2.35) is 10.59 mg/L, a measured value of 10.5 mg/L used in the computation. The clean water oxygen transfer coefficient of 8.0/h will be utilized within the range of actual values.

2.3.2 SPECIFIC OXYGENATION CAPACITY, *oc*

This parameter is often used in the European literature to designate the rate of change in oxygen concentration in an aeration tank. Simply put, it is Equation (2.26) at standard conditions.

$$oc = \left(\frac{dC_L}{dt} \right)_{STD} = K_L a_{20} C_{\infty-20}^* = \frac{SOTR}{V} = SOTR_v \quad (2.43)$$

In both systems, the calculation is the same as shown in Table 2.4.

This parameter has the same units as the oxygen uptake rate (OUR) of the system and gives a feel for reaction rate in the system. Note that both $K_L a$ and C_{∞}^* are a function of temperature, the former increasing and the latter decreasing. When defining the ratio of specific oxygenation capacity at any temperature to that at 20°C, Figure 2.20 shows that the impact of temperature on this product is much less than on the oxygen transfer rate or the oxygen saturation value.

$$\frac{oc_t}{oc} = \frac{K_L a_t}{K_L a_{20}} \frac{C_{\infty-t}^*}{C_{\infty-20}^*} = \theta^{t-20} \tau = \frac{OTR_v}{SOTR_v} \quad (2.44)$$

2.3.3 STANDARD AERATION EFFICIENCY, SAE

The SAE is the rate of oxygen transfer per unit power input, which may be based on either delivered (DP) or wire power (WP).

$$\left. \begin{aligned} SAE &= \frac{SOTR}{DP} \\ SAE &= \frac{SOTR}{WP} \end{aligned} \right\} \quad (2.45)$$

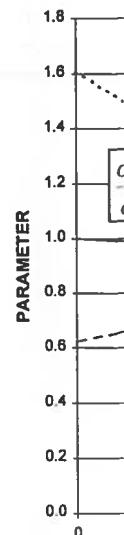


FIGURE 2.20

The overall individual efficiency (individual components), motor used to relate diffused aera

For diffused adiabatic conditions are given under gas flow conditions

The values both pressure Modern Ger a K of 0.285 the adiabatic nearly adiabatic

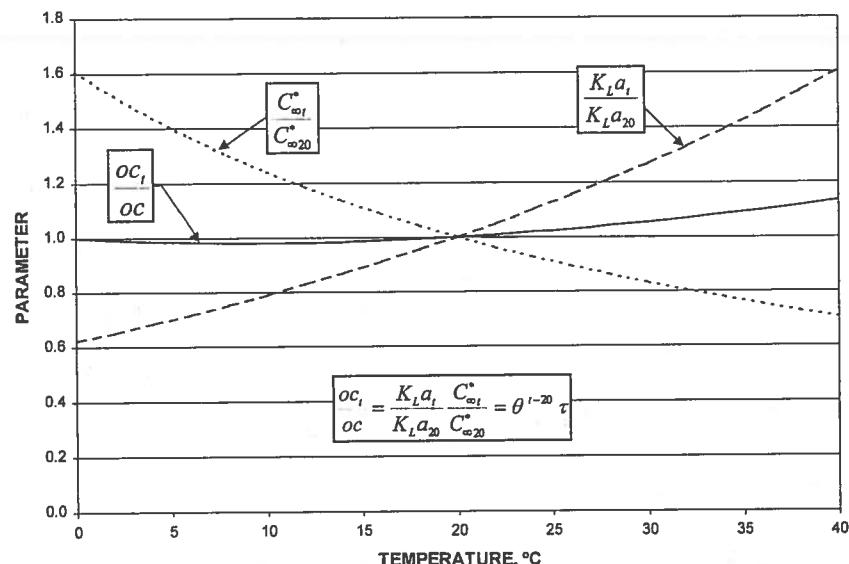


FIGURE 2.20 Impact of temperature on O₂ transfer at zero dissolved O₂ concentration.

The overall efficiency, e , of the aeration equipment is the product of the individual efficiencies of mechanical equipment. Typical efficiencies (EPA, 1983) of the individual components are: blowers (50 percent for older to 80 percent for newer units), motors (95 percent), coupling (95 percent) and gear box (95 percent). It is used to relate the consumed wire power to that which is delivered to the air for diffused aeration or to the liquid for mechanical aeration.

$$WP = \frac{DP}{e} \quad (2.46)$$

For diffused aeration, the delivered power of blowers is typically based on the adiabatic compression equation, AP , (Yunt, 1979). The equations below for power are given under both SI and English units due to the difference in units and standard gas flow conditions.

$$DP = AP = \frac{wRT_a}{K} \left[\left(\frac{P_d}{P_a} \right)^K - 1 \right] \quad (2.47)$$

The value of K is 0.283 for air in the U.S. (36 percent relative humidity) and both pressures are in absolute units (gage + standard atmospheric) as is temperature. Modern German literature on turbo compressors applies adiabatic compression with a K of 0.2857 for dry air. A note of caution must be expressed with respect to using the adiabatic compression equation for all blowers. Although many blowers are nearly adiabatic, some may be closer to polytropic in operation (Yunt, 1979).

The mass flow rate of air, w , is related to the air density and the volumetric flow rate of the influent air, which will be specified at standard conditions as given in Table 2.2.

$$\begin{aligned} w &= \rho_s G_s && \text{(SI)} \\ w &= \gamma_s G_s && \text{(US)} \end{aligned} \quad (2.48)$$

Using the gas constant, R , as follows with the standard conditions in Table 2.2 provides the power level for both SI and English units.

$$\begin{aligned} R &= 287 \frac{J}{kg \cdot ^\circ K} \\ &= 53.346 \frac{ft \cdot lb}{lb_m \cdot ^\circ R} \end{aligned}$$

$$\begin{aligned} AP(kW) &= 0.100G_s(m_N^3/h) \left[\left(\frac{P_d}{P_a} \right)^K - 1 \right] \\ AP(hp) &= 0.227G_s(scfm) \left[\left(\frac{P_d}{P_a} \right)^K - 1 \right] \end{aligned} \quad (2.49)$$

Note that the gas flows are given in terms of their standard conditions as (Normal) m_N^3/h and (standard) scfm. The pressures are expressed as follows. The discharge pressure includes the depth of water at the diffuser submergence as well as all the losses in the air piping and diffuser system. The inlet pressure at the blower is somewhat less than atmospheric due to losses in the air filtering system and inlet piping.

$$P_d = P_s + \gamma_w d + \Delta p_d$$

$$P_a = P_s - \Delta p_a$$

To illustrate use of these concepts, an example in the form of a tabular summary is given in Table 2.5.

Observing the 7.5 percent difference in power requirements using the U.S. and SI designations for standard gas flow conditions shows that the actual inlet air conditions are required to get an accurate estimate of power consumption.

For all aeration devices, wire power can be measured accurately using a recording polyphase wattmeter. An ammeter measuring current can also be used if both the voltage and power factor are known. For squirrel cage induction motors, a power factor of 0.9 is typical (Perry et al., 1984).

2.3.4 STANDARD

The SOTE is the transferred or dis parameter for difl

The mass frac

$$\frac{w_o}{w} = 0.20$$

In the English humidity provides

electric flow
given in

Table 2.2

TABLE 2.5
SAE Example Calculation

| | Parameter | SI | U.S. |
|--------|-----------------|-------------------------------------|----------------|
| (2.48) | P_s | 101.325 kPa | 14.7 psi |
| | Δp_d | 6.89 kPa | 1.0 psi |
| | d | 4.27 m | 14.0 ft |
| | γ_w | 9.81 kN/m ³ | 62.4 lb/cf |
| | $\gamma_w d$ | 41.85 kPa | 6.07 psi |
| | P_d | 150.1 kPa | 21.8 psia |
| | Δp_a | 0.69 kPa | 0.10 psi |
| | P_a | 100.6 kPa | 14.6 psia |
| | G_i | 1000 m _N ³ /h | 637 scfm* |
| | AP = DP | 12.0 kW | 17.37 hp* |
| | $K = 0.283$ | | |
| | e | 0.6 | 0.6 |
| | WP | 20.0 kW | 28.95 hp* |
| | SAE (delivered) | 7.0 kg/kWh | 10.65 lb/hp-h* |
| | SAE (wire) | 4.2 kg/kWh | 6.39 lb/hp-h* |

(2.49) * Not a direct scale-up (approximately 7.5 percent higher) from SI value due to the U.S. standard requiring compression at a temperature of 20°C and 36 percent relative humidity compared with 0°C for the SI with bone dry air. Gas flow based on similar SOTE values. Note that scfm × 1.570 = m_N³/h.

$$kW = \frac{\sqrt{3}EIpf}{1000}$$

2.3.4 STANDARD OXYGEN TRANSFER EFFICIENCY, SOTE

The SOTE is the fraction of oxygen supplied to the aeration tank, which is actually transferred or dissolved into the liquid at standard conditions. It is a major design parameter for diffused aeration systems.

$$SOTE = \frac{SOTR}{w_o} \quad (2.50)$$

The mass fraction of oxygen in dry air is as follows.

$$\frac{w_o}{w} = 0.2095 \frac{\text{mole O}_2}{\text{mole air}} \times 32 \frac{\text{g O}_2}{\text{mole O}_2} \times \frac{\text{mole air}}{28.964 \text{ g air}} = 0.2315 \frac{\text{g O}_2}{\text{g air}}$$

In the English system, taking into account the water vapor at 36 percent relative humidity provides a slightly lower value, 0.23 (ASCE, 1991).

TABLE 2.6
SOTE Example Calculation

| SI | ENGLISH |
|---|---|
| $SOTE = \frac{84.0 \frac{kg}{h}}{0.30 \times 1000 \frac{m^3}{h}} = 0.280$ | $SOTE = \frac{185 \frac{lb}{h}}{1.04 \times 637 scfm} = 0.2793$ |
| | $\%SOTE = 27.9\%$ |
| | $\%SOTE = 28.0\%$ |

Using Equation (2.48) provides the oxygen supply rate.

$$w_o(kg/h) = 0.2315 \times 1.293 G_s = 0.30 G_s (m^3/h)$$

$$w_o(lb/h) = 0.23 \times 0.075 G_s \times 60 \frac{\min}{h} = 1.04 G_s (scfm)$$

Inserting the above into Equation (2.50) provides the SOTE as a function of gas flow.

$$\left. \begin{aligned} SOTE &= \frac{SOTR(kg/h)}{0.30 G_s (m^3/h)} \\ SOTE &= \frac{SOTR(lb/h)}{1.04 G_s (scfm)} \end{aligned} \right\} \quad (2.51)$$

Using the results of the prior example calculations, the SOTE is expressed in Table 2.6. The slight difference in SOTE values is due to the roundoff in Equation 2.51.

2.3.5 APPLICATION TO PROCESS CONDITIONS

Under process conditions, the oxygen transfer rate must meet the demand of the biomass in the aeration tank. The dissolved oxygen level in the tank will always move toward a concentration that balances the transfer rate with the demand. At a steady state condition, these two rates will be equal and will serve as the basis for design.

The actual oxygen transfer rate under process conditions is defined similar to Equation (2.42).

$$OTR_f = V \left(\frac{dC_L}{dt} \right)_{PROCESS} = K_L a_f (C_{\infty f}^* - C_L) V \quad (2.52)$$

Dividing Equation (2.52) by (2.42) provides the ratio of the actual to the standard oxygen transfer rate.

$$\frac{OTR_f}{SOTR} = \frac{K_L a_f (C_{\infty f}^* - C_L)}{K_L a_{20} C_{\infty 20}^*}$$

TA
OT

OI
SC
Employing coefficient and design equation

Assuming 12,000 mg/L t provides the re

The remain calculations.

2.4 NOMEN

| | |
|--------|----------|
| a | m^{-1} |
| A | m^2 |
| A | mg/g |
| AE_f | kg/kW |
| AP | kW, h |
| C | mg/L |
| C_0 | mg/L |

TABLE 2.7
OTR_f and OTE_f Example Calculations

$$\tau = \frac{7.56 \frac{mg}{L}}{9.09 \frac{mg}{L}} = 0.83$$

$$\beta = 1 - 5.7 \times 10^{-6} \times 12000 \frac{mg}{L} = 0.93$$

$$P_b = 101.325 kPa \left[1 - \frac{1000m}{9100m} \right] = 90.19 kPa$$

$$P_d = 0.4 \times 41.85 kPa = 16.76 kPa; \quad p_v = 4.24 kPa$$

$$\Omega = \frac{90.19 + 16.76 - 4.24}{101.325 + 16.76 - 4.24} = 0.90 \quad \xrightarrow{\text{if } 10.59}$$

$$\frac{OTR_f}{SOTR} = 0.45 \times 1.024^{30-20} \left[\frac{0.83 \times 0.93 \times 0.90 \times 10.5 \frac{mg}{L} - 1.5 \frac{mg}{L}}{10.5 \frac{mg}{L}} \right] = 0.31 \quad \xrightarrow{\text{or } 315}$$

$$OTR_f = 0.31 \times 84 \frac{kg}{h} = 26.1 \frac{kg}{h} = 57.5 \frac{lb}{h} \quad \rightarrow 26.5$$

$$\%OTE_f = 0.31 \times 28.0\% = 8.7\%$$

(2.51)

Employing the previously defined correction factors for the oxygen transfer coefficient and saturation value yields the following ratio for the commonly used design equations.

$$\frac{OTR_f}{SOTR} = \frac{oc_f}{oc} = \frac{AE_f}{SAE} = \frac{OTE_f}{SOTE} = \frac{\alpha \theta^{f-20} (\tau \beta \Omega C_{\infty 20}^* - C_L)}{C_{\infty 20}^*} \quad (2.53)$$

Assuming an industrial wastewater with an α of 0.45, a TDS concentration of 12,000 mg/L being treated at 30°C, C_L of 1.5 mg/L and an altitude of 1000 m provides the results in Table 2.7.

The remaining process values use the same ratio as the OTR_f and % SOTE calculations.

2.4 NOMENCLATURE

| | | |
|-----------------|-----------------|--|
| a | m^{-1} | interfacial area/unit liquid volume |
| A | m^2 | interfacial area |
| A | mg/g-h | specific oxygen uptake rate |
| AE _f | kg/kWh, lb/hp-h | aeration efficiency under process conditions |
| AP | kW, hp | adiabatic delivered power |
| C | mg/L | oxygen concentration |
| C ₀ | mg/L | DO concentration at time zero |

(2.52)

the standard

| | | | | |
|-------------------|---|---|------------------|-----------------|
| C_G | mg/L | bulk gas phase oxygen concentration | ℓ | m |
| $C_{G,i}$ | mg/L | gas phase oxygen concentration at interface | M | g |
| C_L | mg/L | bulk liquid phase oxygen concentration | M | g/mole |
| $C_{L,i}$ | mg/L | liquid phase oxygen concentration at interface | M_B | g/mole |
| C_m | mg/L | oxygen concentration at center of floc | n | moles |
| C_s | mg/L | DO saturation concentration | oc | mg/L-h |
| C_s^* | mg/L | surface saturation concentration | oc_t | mg/L-h |
| $C_{s,20}^*$ | mg/L | surface saturation concentration at 20 °C, 9.09 mg/L | OTE _f | -, % |
| C_∞^* | mg/l | oxygen saturation concentration | OTR _f | kg/h, lb |
| $C_{\infty,20}^*$ | mg/l | clean water oxygen saturation concentration at diffuser depth and 20 °C | p | |
| $C_{\infty,f}^*$ | mg/l | oxygen saturation concentration under process (field) conditions | P _a | kPa, ps |
| D | m ² /s | coefficient of molecular diffusion of oxygen in (waste)water | P _b | kPa, ps |
| d | m | tank depth | P _{b0} | kPa, psi |
| D_{AB} | m ² /s | coefficient of molecular diffusion of solute A into solvent B | P _d | kPa, psi |
| d_B | m | bubble diameter | P_{d_t} | kPa, psi |
| D_f | m ² /h | diffusivity in floc | P _s | kPa, psi |
| d_f | m | floc diameter | P _t | kPa, psi |
| DP | kW, hp | delivered power | p _v | kPa, psi |
| e | -, % | overall efficiency of blower or compressor | r | s ⁻¹ |
| E | volts | measured voltage | R | J/(kg·K) |
| F | | diffuser aging factor on oxygen transfer coefficient | R _G | s/m |
| G_s | m _N ³ /h, scfm | airflow rate at standard conditions | R _L | s/m |
| H | (mg/L) _{gas} /(mg/L) _{liquid} | Henry's constant | R _T | s/m |
| H | m | stream depth | SAE | kg/kWh |
| I | amps | measured current | SOTE | -, % |
| K | | coefficient in adiabatic compression equation | SOTR | kg/h, lb |
| J | g/m ² -s | mass flux of oxygen | T | °K |
| k_G | m/s | gas film coefficient | t | °C |
| k_L | m/s | liquid film coefficient | t | s |
| K_L | m/s | overall liquid film coefficient | T _a | °K, °R |
| $K_L a$ | h ⁻¹ | oxygen transfer coefficient | TDS | mg/L |
| $K_L a_{20}$ | h ⁻¹ | clean water oxygen transfer coefficient at 20°C | U | m/s |
| $K_L a_t$ | h ⁻¹ | clean water oxygen transfer coefficient at temperature t | V | m ³ |
| kW | kW | measured wire power | V_A | m ³ |
| k_δ | m/s | liquid film coefficient in viscous laminar sublayer | \bar{v} | m/s |
| k_t | m/s | liquid film coefficient in turbulent sublayer | w | kg/h, lb |
| | | | w_o | kg/h, lb |
| | | | WP | kW, hp |
| | | | y | m |
| | | | Δp_a | kPa, psi |
| | | | Δp_d | kPa, psi |

| | | | |
|------------------|----------------------------|-----------------|--|
| surface | ℓ | m | characteristic mixing length |
| 1 | M | g | mass of oxygen transferred |
| nterface | M | g/mole | molecular weight |
| | M_B | g/mole | molecular weight of solvent B |
| | n | moles | number of moles in ideal gas law |
| | oc | mg/L·h | specific oxygenation capacity in clean water = SOTR _v |
| °C, 9.09 mg/L | oc _t | mg/L·h | specific oxygenation capacity in clean water at temperature, t, = OTR _v |
| tration at dif- | OTE _f | –, % | oxygen transfer efficiency under process conditions |
| process (field) | OTR _f | kg/h, lb/h | oxygen transfer rate under process conditions |
| oxygen in | p | | partial pressure of oxygen |
| solute A into | P _a | kPa, psia | absolute pressure upstream of blower |
| | P _b | kPa, psia | barometric pressure |
| | P _{b0} | kPa, psia | barometric pressure at zero altitude |
| | P _d | kPa, psia | absolute pressure downstream of blower |
| | P _{d_e} | kPa, psi | effective pressure |
| | P _s | kPa, psia | standard barometric pressure, 101.325 kPa, 14.696 psia |
| | P _t | kPa, psia | total pressure |
| | P _v | kPa, psi | vapor pressure |
| | r | s ⁻¹ | surface renewal rate |
| | R | J/(kg·K) | universal gas constant (286.88 J/kg·K) |
| ressor | R _G | s/m | resistance to oxygen transfer in gas phase |
| sfer coefficient | R _L | s/m | resistance to oxygen transfer in liquid phase |
| | R _T | s/m | total resistance to oxygen transfer |
| equation | SAE | kg/kWh, lb/hp-h | standard aeration efficiency |
| | SOTE | –, % | standard oxygen transfer efficiency |
| | SOTR | kg/h, lb/h | standard oxygen transfer rate |
| | T | °K | absolute temperature |
| | t | °C | temperature |
| | t | s | time |
| | T _a | °K, °R | absolute temperature of influent gas to blower |
| | TDS | mg/L | total dissolved solids concentration |
| | U | m/s | stream velocity |
| | V | m ³ | tank volume |
| | V _A | m ³ | total volume of solute A |
| | \bar{v} | m/s | vertical velocity fluctuation |
| ient at 20°C | w | kg/h, lb/h | mass flow rate of air |
| ient at tempera- | w _o | kg/h, lb/h | mass flow rate of oxygen |
| uminar sublayer | WP | kW, hp | wire power |
| sublayer | y | m | depth of penetration |
| | Δp_a | kPa, psi | pressure drop in inlet filters and piping to blower |
| | Δp_d | kPa, psi | pressure drop in piping and diffuser downstream of blower |

| | | |
|------------|---------------------------------------|---|
| α | | wastewater correction factor for oxygen transfer coefficient |
| β | | wastewater correction factor for oxygen saturation |
| δ | | depth correction factor for oxygen saturation |
| δ_c | m | concentration boundary layer thickness |
| δ_d | m | diffuse sublayer thickness |
| δ_L | m | liquid film thickness |
| ϕ | | association parameter of solvent B, for water $\phi = 2.6$ |
| γ_f | kg/m ³ | specific weight of dry floc |
| γ_s | lb/ft ³ | specific weight of standard gas, 0.075 lb/ft ³ |
| γ_w | N/m ³ , lb/ft ³ | specific weight of water |
| κ | | temperature correction factor for oxygen transfer coefficient expressed in exponential form |
| μ | g/m-s | absolute viscosity |
| μ_B | g/m-s | absolute viscosity of solvent B |
| θ | | temperature correction factor for oxygen transfer coefficient |
| ρ_s | kg/m ³ | density of standard gas |
| τ | | temperature correction factor for oxygen saturation |
| Ω | | pressure correction factor for oxygen saturation |

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3

3.1 INTROD

Diffused aeration pressure below a this definition. H mechanical pump include jet aerato and aspirating in

Although the 1927), major adva sludge pr aeration technolo aware of the imp flow rate on oxy material framew diffuser was gra (Martin, 1927). I tubes with fibrou methods (Federal gators sought m England, experin and glass and pu losses. A secret p placed in cast iro used for many ye

Meanwhile, i newly constructe using grids of pe jets. The Filtrc sand and had per (9 to 13 scfm) at North-Side plant CA; and Gastoni of the developmen Ferro Corporatio bonded with a va of choice. Perme (120 scfm). In ad and more recently

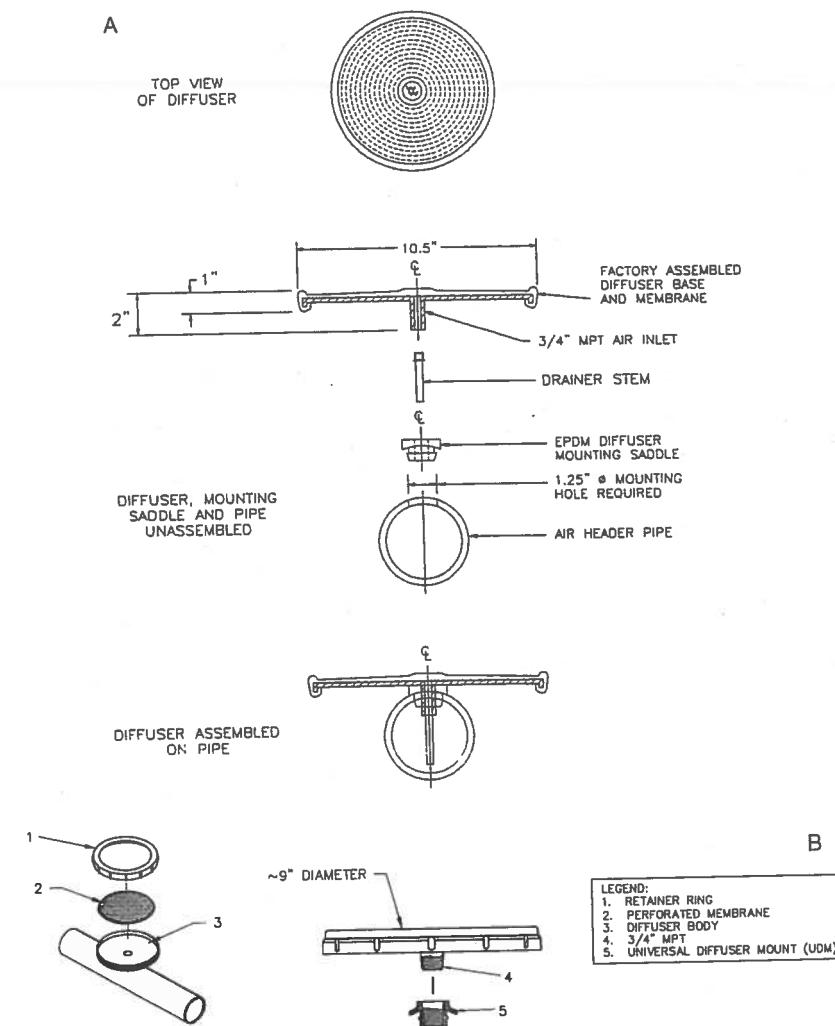


FIGURE 3.9 Several membrane disc configurations [(A) courtesy of Wilfey Weber, Inc., Denver, CO; (B) courtesy of EDI, Columbia, MO].

15.7 m³/h (1 to 10 scfm) per diffuser for the discs up to 30 cm (12 in) in diameter and 4.7 to 31.4 m³/h (3-20 scfm) per diffuser for the larger discs.

3.2.2 NONPOROUS DIFFUSER SYSTEMS

Nonporous diffusers differ from porous diffusers in that they use larger orifices or holes to discharge air. Introduced as early as 1893 these diffusers are available in a variety of shapes and materials. This section will describe these diffusers under the categories of fixed orifice, valved orifice, static tubes, perforated tubes, and other units.

3.2.2.1 Fixed Orifice

Fixed orifice openings in a (0.16 in) were openings close hole sizes that relatively coarse oxygen trans channel aeration have a proper diffusers are stainless steel on both side tube are designed to produce a thin membrane disc.

Many of these are equipped with legs to purge rates range from tubes normally employed to

3.2.2.2 Valved Orifice

Valved orifice off. Some discharge off. Several designs that opens a Delrin ball containing a A 7.6 cm (3 in) that opens devices operating at 12 scfm). They require high check valves for oxygen transfer service in a

3.2.2.3 Static Tubes

Static tube that deliver airlift pump

3.2.2.1 Fixed Orifice Diffusers

Fixed orifice diffusers vary from simple openings in pipes to specially configured openings in a number of housing shapes. Historically, orifices much below 4 mm (0.16 in) were susceptible to rapid clogging in wastewater, although even the coarser openings clogged under some wastewater conditions. These devices typically employ hole sizes that range from 4.76 to 9.5 mm (0.1875 to 0.375 in) in diameter producing relatively coarse bubbles (6 to 10 mm). As a result, these diffusers are not efficient oxygen transfer devices but find use in grit separation processes, influent and effluent channel aeration, aerobic sludge digestion and aeration of certain wastewaters that have a propensity to precipitate or easily foul porous diffusers. Today, fixed orifice diffusers are usually molded plastic devices containing a number of holes or slotted stainless steel tubes containing rows of holes along the top or sides and an open slot on both sides of the tube below the holes (Figure 3.10 A and B). The slots in the tube are designed to carry air as airflow increases or as holes plug. One manufacturer produces a slotted tube constructed of plastic that may be converted to a porous membrane diffuser with the placement of a synthetic fiber sheath over the tube.

Many of the fixed orifice diffusers are saddle mounted on the air header. Most are equipped with airflow control orifices to balance airflow. Some contain blowoff legs to purge liquid or relieve back pressure in the event of fouling. Typical gasflow rates range from 9.4 to 47.1 m³/h (6 to 30 scfm) depending on the unit. Perforated tubes normally are screwed into air headers in wideband configurations. Orifices are employed to control airflow distribution in the system.

3.2.2.2 Valved Orifice Diffusers

Valved orifice diffusers use a check valve to prevent backflow when the air is shut off. Some are designed to provide adjustment of the number or size of the air discharge openings. Orifice sizes are similar to those used in fixed orifice devices. Several designs incorporate a membrane (EPDM or other elastomer) as a diaphragm that opens and closes over orifices when air is on or off (Figure 3.11). Another uses a Delrin ball check valve that rides up and down a sleeve mounted inside a cylinder containing drilled holes. A third design employs a cast body with inner air chamber. A 7.6 cm (3 in) diameter plastic disc is retained in position by a steel spring wire that opens and closes over the air chamber depending upon airflow. All of these devices operate over a variety of airflows ranging from 9.4 to 18.8 m³/h (6 to 12 scfm). The units are typically mounted on the crown of the air header thereby requiring header blowoff provisions to purge the system of water in the event of a check valve failure. As with fixed orifice diffusers, these devices exhibit lower oxygen transfer efficiencies than the finer bubble porous diffusers and typically find service in grit separation, inlet/outlet channel aeration, and aerobic digestion.

3.2.2.3 Static Tubes

Static tube diffusers consist of a stationary vertical tube placed over an air header that delivers bubbles of air through drilled holes. The static tube is similar to an airlift pump. As air rises through the vertical tube, interference devices within the

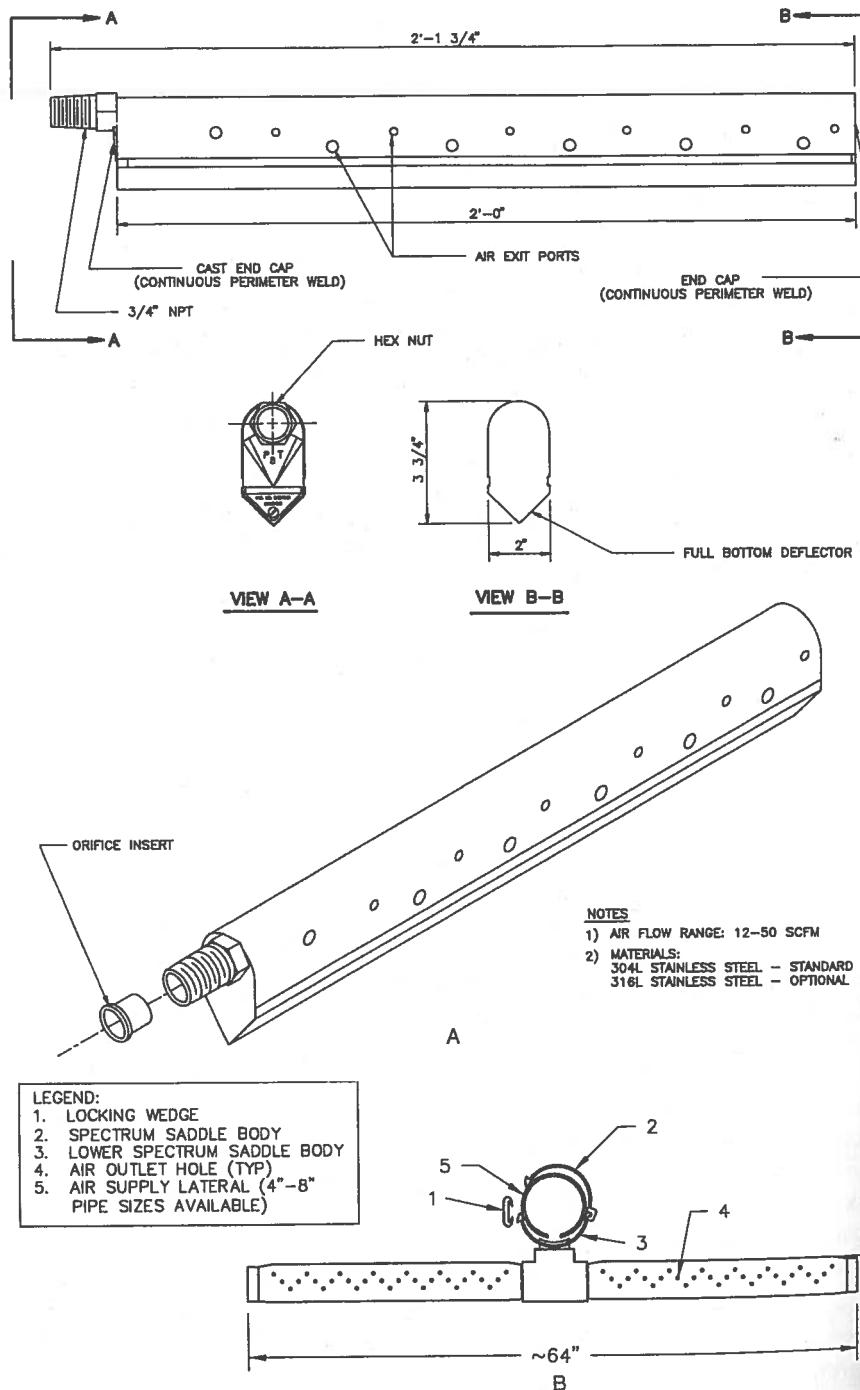
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are available
these diffusers
forated tubes,



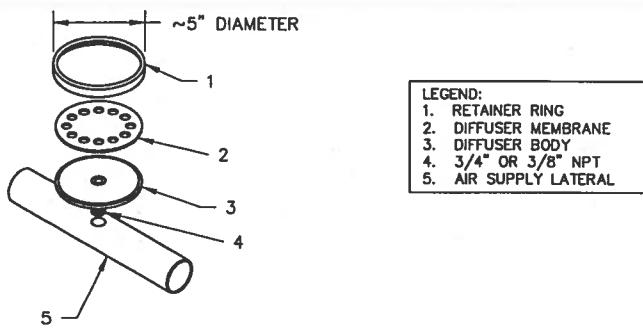


FIGURE 3.11 Selected coarse bubble diffusers (courtesy of EDI, Columbia, MO).

tube are designed to shear bubbles and mix the air and liquid, thereby promoting gas transfer. The vertical tubes are normally 0.3 to 0.45 m (12 to 18 in) in diameter and constructed of polypropylene or polyethylene. They are fixed to the tank bottom by stainless steel support stands. High-density polyethylene air piping is supported below the vertical tube. Holes drilled in the air header are normally of a size similar to fixed orifice diffusers. Airflow rates per tube vary with tube diameter but are typically in the range of 15.7 to 70.7 $\text{m}^3/\text{N}/\text{h}$ (10 to 45 scfm). Static tubes are most often applied to aerated lagoon systems, although some may be used in activated sludge processes.

3.2.2.4 Other Devices

3.2.2.4.1 Jets

Jet aeration combines liquid pumping with gas pumping to result in a plume of liquid and entrained air bubbles. A pumping system recirculates the wastewater from the aeration basin and ejects it through a nozzle assembly. The nozzle configurations may include a venturi or mixing chamber whereby gas and liquid are mixed in the motive field. At least one manufacturer produces a jet aerator containing an inner and outer jet configuration with mixing chamber. Gas is pumped through a separate header and is introduced into the recycled wastewater at the venturi or within the mixing chamber (Figure 3.12 and 3.13). The resultant gas-liquid plume is then directed back into the aeration tank through the jet. Jet aerators may be configured as directional devices or as clustered or radial devices. The piping and jets are normally constructed of polypropylene, fiberglass, or stainless steel.

Typically the wastewater recirculation pump is a constant-rate device, and the power turndown for the aerator is accomplished by varying the airflow rate. Air is delivered under pressure by a low head blower. As such, power is consumed both in the recirculation of the liquid and the delivery of the air. The gas-liquid plume normally contains very fine bubbles of gas, thereby classifying jets as fine bubble devices. Depending upon basin geometry and jet exit velocity, the horizontal plume rises rapidly within the basin intermixing with the basin contents. It is significant to note that the air-head loss through the jet is very low or negative due to the ejecting action of the motive fluid. Although it has been used in rectangular basins,

2-50 SCFM
EEL - STANDARD
EEL - OPTIONAL

4"
WI; (B) courtesy

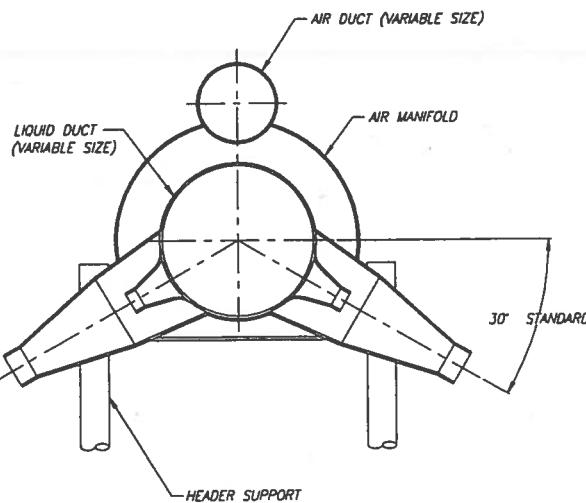


FIGURE 3.12 Unidirectional jet (courtesy of US Filter, Jet Tech Products, Edwardsville, KS).

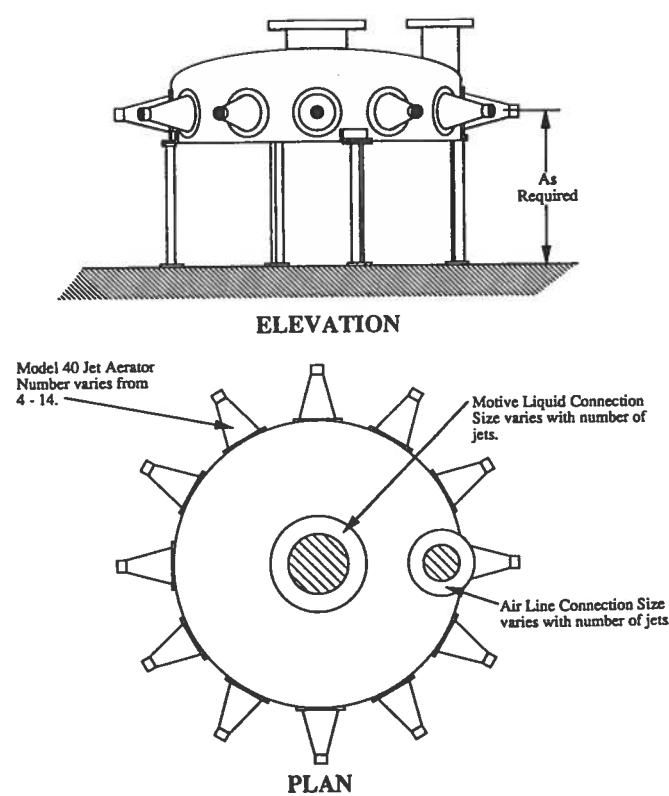


FIGURE 3.13 Radial jet (courtesy of US Filter, Jet Tech Products, Edwardsville, KS).

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3.2.2.4.3

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FIGURE 3.14

the directional feature of the device favors its application in oxidation ditches and circular basins.

3.2.2.4.2 Perforated Hose

Perforated hose typically consists of polyethylene tubing held on the floor of the basin by lead ballast. At least one manufacturer suspends the tubing from floats. The tubing contains slits or holes at the top of the tube to release air. Manifolds running along the basin length supply the air. Typically the tubing is mounted across the basin width. Applications of perforated tubing are limited to lagoon systems.

3.2.2.4.3 U-Tube Aeration

A U-tube system consists of a 9 to 150 m (30 to 500 ft) deep shaft that is divided into an inner and outer zone. As air is directed to the wastewater in the downcomer zone, the mixture travels to the bottom of the tube and then returns back to the surface for further treatment (Figure 3.14). The great depth to which the air-water mixture is subjected provides high dissolution due to the high oxygen partial pressures.

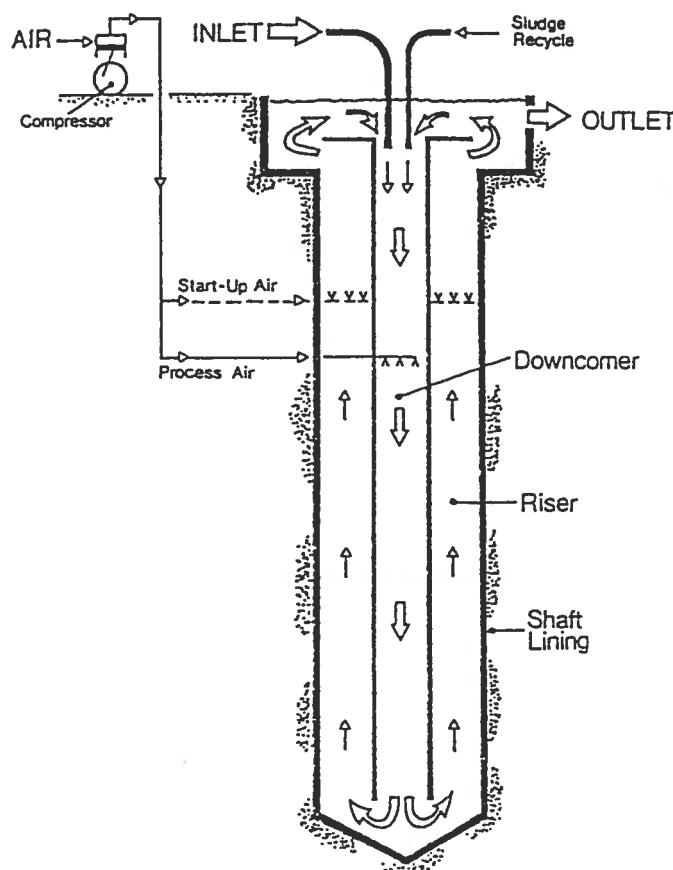


FIGURE 3.14 U-tube aerator.

CLEAN WATER TEST RESULTS

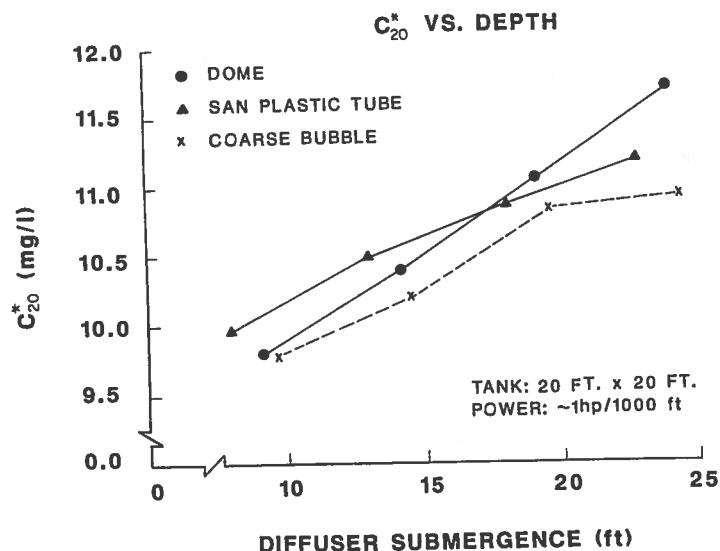


FIGURE 3.22 Diffuser submergence vs. DO saturation.

TABLE 3.1
Typical Delta Values for Diffused Aeration Devices

| Diffuser Type | Range of Delta | Range of Depth (m) |
|------------------|----------------|--------------------|
| Nonporous | | |
| Static tube | 1.08–1.16 | 4.2–5.2 |
| Perf tube | 1.05–1.15 | 2.7–7.3 |
| Porous | | |
| Plates | 1.25–1.28 | 5.6 |
| PM tubes | 1.07–1.21 | 2.1–4.6 |
| PM* disc | 1.05–1.30 | 2.8–7.4 |
| Cer disc | 1.09–1.18 | 4.3–5.4 |
| Cer dome | 1.13–1.14 | 2.9 |

* PM- Perforated membrane.

3.4.2.2 Oxygen Transfer Data

Typical values of SOTE (and SAE for some nonporous diffusers) for various diffuser types are presented in Tables 3.2 through 3.5. With the continuous changes occurring in the development of diffuser materials and shapes, it is difficult to make many generalizations about the performance of any given diffuser. However, as discussed previously, there are some factors that influence performance of an aeration system.

TABLE 3.2
Clean Water Oxygen Transfer Efficiency — Nonporous Diffusers

| Type and Placement | | Airflow Rate (m ³ /h/unit) | Submergence (m) | SOTE (%) | SAE (kg/kWh) | Reference |
|-------------------------------|----|--|--------------------|-------------|-----------------|----------------------|
| Fixed orifice perforated tube | S | 9.3–32.8 ✓ | 7.3 | 21–25 | — | Johnson, 1992 |
| | S | 8.6–40.0 ✓ | 5.2–5.6 | 11–18 | — | Johnson, 1992 |
| | S | 9.3–64.3 ✓ | 4.1–4.8 | 5–17 | — | Johnson, 1992 |
| | S | 16.0–39.6 ✓ | 3.0–3.8 | 6–14 | — | Johnson, 1992 |
| | S | 8.9–31.5 ✓ | 2.7 | 6–7 | — | Johnson, 1992 |
| | G | 7.5–23.2 ✓ | 3.0 | 7–8 | 1.3–1.5 | Yunt & Hancuff, 1988 |
| | G | 8.3–24.4 ✓ | 6.1 | 17–20 | 2.0–2.2 | Yunt & Hancuff, 1988 |
| | MW | 6.6–18.8 ✓ | 4.6 | 11–13 | 1.5–1.6 | Yunt & Hancuff, 1988 |
| | S | 12.9–51.3 ✓ | 4.1–4.8 | 9–13 | — | Johnson, 1992 |
| | MW | 18.7–57.0 ✓ | 3.0 | 6–7 | 1.3–1.5 | Yunt & Hancuff, 1988 |
| Sparger | MW | 19.8–60.2 ✓ | 4.6 | 10–11 | 1.5–1.6 | Yunt & Hancuff, 1988 |
| | MW | 19.8–59.2 ✓ | 6.1 | 15–17 | 1.8–1.9 | Yunt & Hancuff, 1988 |
| | G | 15.7–60.2 ✓ | 3.0 | 6–8 | 1.1–1.5 | Yunt & Hancuff, 1988 |
| | G | 15.7–65.6 ✓ | 4.2 | 11–15 | 1.5–1.8 | Semblex, 1987 |
| Static tube | G | 15.7–66.4 ✓ | 6.1 | 13–20 | 1.7–1.9 | Semblex, 1987 |
| | G | 24.4–51.0 ✓ | 4.2–4.6 | 8–12 | — | Johnson, 1992 |
| | ? | 37.0–68.3 ✓ | 5.2 | 12–15 | — | Johnson, 1992 |

1 m = 3.28 ft; 1.0 m³/h = 0.64 scfm; 1.0 kg/kWh = 1.644 lb/hp-h

G = Grid; S = Spiral roll; MW = Mid-width

TABLE 3.4
Clean Wat

Type

Porous plastic

Perforated membrane

1 m = 3.28 ft;

G = Grid; DS =

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TABLE 3.3
Clean Water Oxygen Transfer Efficiency — Aspirators and Jets

| Type and Placement | | Airflow Rate (m ³ /h/unit) | Submergence (m) | SOTE (%) | SAE (kg/kWh) | Reference |
|--------------------|--------|--|--------------------|-------------|-----------------|----------------------|
| Jets | Dir | 21.1–119 | 4.6 | 15–24 | 1.7–2.0 | Yunt & Hancuff, 1988 |
| | Clu | 7.1–86.3 | 3.0 | 8–14 | 1.1–1.6 | Yunt & Hancuff, 1988 |
| | Clu | 7.7–50.5 | 6.1 | 21–33 | 1.6–2.2 | Yunt & Hancuff, 1988 |
| Aspirator tube | 5.5 kw | — | 2.0 | — | 0.5–0.9 | Kayser, 1992 |
| | 15 kw | — | 2.5 | — | 0.4–0.8 | Kayser, 1992 |

1 m = 3.28 ft; 1.0 m³/h = 0.64 scfm; 1.0 kg/kWh = 1.644 lb/hp-h

Dir = Directional; Clu = Cluster

Some of these factors are discussed in further detail in the following sections. Since the power consumed in transferring oxygen to the liquid is most important in assessing system performance, estimates of SAE are presented in this section for a variety of devices. For diffused air devices, this figure typically requires a calculation

TABLE 3.4
Clean Water Oxygen Transfer Efficiency — Porous Tubes

| Type | Placement | Airflow Rate (m ³ /h/unit) | SOTE (%) at Depth | | | | Reference |
|---------------------|-----------|--|-------------------|-------|-------|-------|-------------------------------|
| | | | 2.1 m | 3.0m | 4.6m | 6.1m | |
| Porous plastic | G | 3.8–6.3 | — | — | 28–32 | — | EPA, 1989 |
| | DS | 4.7–11.0 | — | 10–16 | 16–24 | 22–32 | EPA, 1989 |
| | DS | 14.1–17.3 | — | 10–14 | 15–17 | 21–26 | EPA, 1989 |
| | S | 3.1–11.0 | — | 12–15 | 15–20 | 22–25 | EPA, 1989 |
| | S | 12.6–18.8 | — | 10–15 | 10–17 | 22 | EPA, 1989 |
| Perforated membrane | G | 4.7 | — | — | — | 45 | GSEE, Inc., 1998 |
| | G | 3.0–10.0 | — | 27–28 | — | — | Pöpel, 1991 |
| | S | 0.8–10.0 | 13–19 | 17–21 | 26–35 | — | Johnson, 1993; Pöpel, 1991 |
| | DS | 0.8–18.8 | 10–20 | 15–21 | 21–36 | 27–36 | EPA, 1989; Johnson, 1993 |

$$1 \text{ m} = 3.28 \text{ ft}; 1.0 \text{ m}^3/\text{h} = 0.64 \text{ scfm}$$

G = Grid; DS = Dual spiral roll; S = Spiral roll

of power required by a given blower under a given set of environmental conditions. In this case, the blower wire power consumption is related to the discharge pressure and the mass rate of air by the adiabatic compression of air. A discussion of this calculation is found in Chapter 4. The assumed values of system head loss, blower inlet and discharge temperatures, and combined blower/motor efficiency are presented as required for these calculations.

3.4.2.2.1 Diffuser Type

In diffused aeration, air bubbles, which are typically formed at an orifice (exceptions are jet and aspirator systems) near the bottom of the aeration basin, break off and rise through the liquid finally bursting at the surface. As the bubble begins to emerge from the orifice, the air-water interface is continuously being replenished causing a high surface renewal rate and thus, a high transfer rate. Once it breaks away from the orifice and theoretically reaches a terminal rise velocity, the effective liquid film thickness or surface renewal rate becomes constant. In an aeration tank, eddy currents normally will affect rise velocities, which are the sum of the terminal or "slip" velocity, v_s , of the bubble and the fluid velocity for the rising gas-liquid stream, v_w . As the bubble bursts at the surface, it sheds an oxygen-saturated film into the surface layers. Some surface aeration also occurs due to surface turbulence.

The size of the bubble released by a diffuser is related to the orifice diameter, surface tension, and liquid density when gas flows are low (typically less than 100 bubbles per minute). At the higher airflow rates used in wastewater aeration practice, bubble diameter is a function of gas flow rate, G_s , while frequency of formation remains constant yielding the following empirical expression.

TABLE 3.5
Clean Water Oxygen Transfer Efficiency — Porous Disc/Domes in Grid

| Type | Diffuser Density (%) | Airflow Rate ($m_N^3/h/unit$) | SOTE (%) at Depth | | | Reference |
|---------------------|----------------------|---------------------------------|-------------------|-------|-------|--------------------------|
| | | | 3.0 m | 4.6 m | 6.1 m | |
| Plastic plates | 10 | 35.6–84.7 | — | — | 30–40 | Johnson, 1993 |
| Ceramic disc, 24-cm | 7.5 | 1.4–4.7 | 20–22 | 27–33 | 34–37 | EPA, 1989 |
| | 11.7 | 1.3–4.6 | 21–24 | 30–34 | 35–41 | EPA, 1989 |
| | 15.1 | 1.1–4.1 | 22–25 | 31–34 | 38–41 | EPA, 1989 |
| Ceramic disc, 22-cm | 6.0–6.3 | 2.3–5.0 | — | 25–29 | 32–38 | EPA, 1989 |
| | 6.9–7.7 | 0.9–3.9 | — | 25–30 | 33–40 | EPA, 1989 |
| | 8.9–10.2 | 0.9–5.3 | — | 27–34 | 31–40 | EPA, 1989; Johnson, 1993 |
| Ceramic disc, 23-cm | 12.0–12.8 | 0.6–4.4 | — | 25–36 | 34–39 | EPA, 1989 |
| | 16.4–21.6 | 1.1–4.9 | — | 27–38 | 31–38 | EPA, 1989; Johnson, 1993 |
| Ceramic dome, 18-cm | 12.0 | 1.9 | — | 32–33 | — | Johnson, 1993 |
| Plastic disc, 18-cm | 4.8 | 0.8–3.9 | — | 23–31 | 28–40 | EPA, 1989 |
| | 6.1–6.3 | 0.8–3.9 | 16–23 | 25–32 | 30–41 | EPA, 1989 |
| | 8.1–8.4 | 0.8–3.9 | 20–24 | 27–37 | 31–44 | EPA, 1989 |
| | 10.7–12.1 | 0.8–3.9 | 17–23 | 27–35 | 33–47 | EPA, 1989 |
| | 17.3 | 0.8–3.9 | 18–26 | 27–34 | — | EPA, 1989; Johnson, 1993 |
| Plastic disc, 18-cm | 3.9 | 0.9–5.5 | 15–18 | 22–27 | — | EPA, 1989 |
| | 5.8 | 0.9–5.5 | 16–21 | 24–28 | — | EPA, 1989 |
| | 6.8 | 0.8–3.6 | — | 25–31 | — | EPA, 1989 |
| | 9.2 | 0.6–2.3 | 19–22 | 26–32 | — | EPA, 1989 |

1 m = 3.28 ft; 1.0 m_N^3/h = 0.64 scfm

$$d_b = C_1 G_s^n \quad (3.1)$$

C_1 and n are empirical constants. For porous diffusers (fine pore) where pore size is typically 0.1 to 0.3 mm, n is usually less than 1.0, and bubble diameters range from 1.5 to 3.0 mm. For nonporous diffusers where orifice sizes typically range from 5 to 25 mm or larger, n may be greater than 1.0 and, bubble diameters range from 20 to 40 mm. For these coarse bubble diffusers, it is believed that as gas flow increases, the turbulence tends to redivide the larger bubbles into smaller ones (Eckenfelder, 1959). An intermediate group includes diffusers that have pore sizes that may range from 2 to 5 mm, and bubbles exhibit diameters typically intermediate between the fine pore diffuser and the nonporous diffuser.

Bubble size and shape affect oxygen mass transfer in several ways. Barnhart (1966, 1969) has shown that about 25 percent of the total oxygen transferred in a 3.65 m (12 ft) deep tank occurred at bubble formation for a fine pore diffuser system. Using coarse bubble diffusers, considerably less transfer occurred during bubble formation. Barnhart has shown that the liquid film coefficient, k_L , increases as bubble

K_L (cm/hr)

FIGURE 3.2
Barnhart, 1969

size increases, decreases with about the location found that / The individual bubble size Finally, the size. The total At $R_c < 300$ et al., 1973 rectilinear, basin total

TABLE 3.6
Typical Clean Water Standard Aeration Efficiencies — Porous Diffusers
(Submergence 4.6 m)

| Type/Configuration | Airflow Rate (m ³ /h/diffuser) | SAE ^a (kg/kWh) |
|---------------------------------------|--|------------------------------|
| Plastic tube | | |
| Grid | 3.8–6.3 | 4.5–5.2 |
| Spiral | 3.1–11.0 | 2.4–3.2 |
| Spiral | 17.6–18.8 | 1.6–2.7 |
| Dual | 4.7–11.0 | 2.6–3.9 |
| Perforated membrane | | |
| Spiral | 0.8–10 | 4.2–5.7 |
| Dual | 0.8–18.8 | 3.4–5.8 |
| Ceramic disc | | |
| 18-cm grid | 0.6–5.5 | 3.6–5.2 |
| 22-cm grid | 0.6–5.0 | 4.1–6.1 |
| 24-cm grid | 1.1–4.7 | 4.4–5.5 |
| Ceramic dome | | |
| 18-cm grid | 0.8–3.9 | 3.4–6.0 |
| Perforated membrane disc ^b | | |
| 51-cm | 24–172 | 2.7–4.6 |
| 30-cm | 13–237 | 2.7–6.1 |
| 23-cm | 13–280 | 2.4–7.1 |
| Panel ^b | 4–74 | 3.1–6.9 |

1.0 m³/h = 0.64 scfm; 1.0 kg/kWh = 1.644 lb/hp-h

^a Wire power calculated from adiabatic compression relationship for T = 20°C, P = 1 atm, blower/motor efficiency = 70%, discharge pressure varies with diffuser type

^b Airflow rate — m³/h-m²

and the bubble residence time distribution, the total gas surface area in the basin decreases as the bulk bubble velocity increases.

Oxygen transfer efficiencies can therefore be related to diffuser type by means of the system parameters of bubble size and shape along with gas flow rate for a given basin geometry. Typically, for bubbles larger than about 1 to 2 mm, efficiency will decrease with increased bubble size down to some asymptotic value. Tables 3.2 through 3.5 illustrate that porous diffusers, which generally produce fine bubbles, will produce significantly higher efficiencies than nonporous large orifice diffusers. It should be noted that jet diffusers also generate fine bubbles due to cavitation and/or turbulence occurring in the region where gas is introduced into the recirculated liquid stream. Aspirating devices generally produce an intermediate bubble size that is less efficient than the porous diffuser or the jet.

An examination of Tables 3.4 through 3.8 indicate that among the porous diffuser systems, all appear to be similar in oxygen transfer efficiencies with the possible exception of certain membrane panel and high-density membrane disc configurations.

TABLE 3.7
SOTE vs.

| | |
|--------------|---------|
| Ceramic don | Diffuse |
| Ceramic disc | Diffuse |
| Ceramic disc | Diffuse |
| Rigid porous | Diffuse |
| Rigid porous | Diffuse |
| Nonrigid por | Diffuse |
| Perforated n | Diffuse |
| 23-cm perfo | Diffuse |
| membrane | Diffuse |
| EPDM perf | Diffuse |
| membrane | Diffuse |

^a Equation 3

^b One 23-cm

^c One 61-cm

1 m = 3.28

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TABLE 3.7
SOTE vs. Airflow for Selected Fine-Pore Diffusers in Clean Water (EPA, 1989)

| | Diffuser Type | Layout | Diffuser Submergence (m) | Diffuser Density (No. units/m ²) | SOTE (%) | Exponent "m" ^a |
|--------------------------------|--------------------|--------|--------------------------|--|----------|---------------------------|
| Ceramic dome | Grid | | 4.3 | 3.4 | 29.6 | -0.150 |
| Ceramic disc | Grid | | 3.7 | 2.8 | 31.7 | -0.133 |
| Ceramic disc | Grid | | 3.7 | 1.6 | 26.0 | -0.126 |
| Rigid porous plastic disc | Grid | | 4.0 | 3.7 | 27.9 | -0.097 |
| Rigid porous plastic tube | Double spiral roll | | 4.0 | 1.1 | 26.7 | -0.240 |
| Nonrigid porous plastic tube | Spiral roll | | 4.6 | 0.9 | 27.1 | -0.276 |
| Perforated membrane disc | Grid | | 4.3 | 0.9 | 29.2 | -0.195 |
| 23-cm perforated membrane disc | Grid | | 3.0 | 2.1 ^b | 18.9 | -0.110 |
| EPDM perforated membrane tube | Grid | | 3.0 | 2.1 ^c | 21.0 | -0.150 |

^a Equation 3.2

^b One 23-cm-diameter disc in a 76-cm-diameter column

^c One 61-cm-long tube in a 76-cm-diameter column

1 m = 3.28 ft

Reasons for these higher levels of performance are elaborated further in this section. A comparison of diffuser performance based on SAE is provided in Tables 3.2, 3.3 and 3.6. It can be seen that most of the devices generating the finer bubbles will also require significantly less power for a given transfer rate than the coarser bubble devices. What is also clear from this tabulation is that those devices requiring power for both the delivery of air and liquid will suffer lower values of SAE even though SOTE values may be high.

3.4.2.2.2 Diffuser Airflow Rate

As seen from Equation (3.1), bubble size depends on airflow rate. The airflow rate also affects bubble shape, bubble rise velocity, and system turbulence. As described above, airflow influences overall bubble surface area and therefore, oxygen transfer rate. It also will influence surface renewal rates and bubble size distributions. For porous diffusers, an increase in G_s will produce larger bubbles and higher bubble velocities, thereby decreasing total bubble surface area and oxygen transfer rate. Over the normal range of operation for a given basin geometry, aeration system, and diffuser type, the relationship between SOTE and diffuser airflow rate can be described by the following empirical relationship.

$$SOTE_a / SOTE_b = [G_{sa} / G_{sb}]^m \quad (3.2)$$

In this equation $SOTE_a$ and $SOTE_b$ equals SOTE values at gas flow rates G_{sa} and G_{sb} respectively, and "m" is a constant for a given diffuser and system configuration.

TABLE 3.8
Clean Water Oxygen Transfer Efficiency — Perforated Membrane
Panels/Discs in Grids

| Type | Diffuser Density (%) | Airflow Rate ^b (m ³ /h-m ²) | Specific SOTE ^c (%/m) | Reference |
|-----------------------|----------------------|---|----------------------------------|--------------------------------------|
| Panels | 5.0 | 37.2–74.4 | 4.6 | Pöpel & Wagner, 1991 |
| | 8.0 | 45.6–92.9 | 5.9–6.2 | Pöpel & Wagner, 1991 |
| | 31.0 | 4.7–16.9 | 7.5–10.1 | Pöpel & Wagner, 1991 |
| | 44.6 | 4.1–12.5 | 7.9–9.5 | Parkson, 1991 |
| | 98+ | 0.8–12.3 | 10.8–17.0 ^a | GSEE, 1986 |
| Disc — 51 cm | 6 | 27.0–172 | 3.6–5.6 | Huibregtse, 1987 |
| | 17.7 | 23.7–162 | 3.9–6.2 | Huibregtse, 1987 |
| Disc — 30 cm | 1.5–3.0 | 13.5–27.0 | 5.3–8.0 | Johnson, 1993 |
| | 4.1 | 49.0–312 | 3.6–6.2 | Eimco, 1986 |
| | 6.9–7.6 | 27.0–346 | 4.5–6.0 | Johnson, 1993 |
| | 6.8 | 59.1–237 | 4.5–7.8 | Johnson, 1993 |
| | 13.6 | 59.1–237 | 4.1–8.2 | Johnson, 1993 |
| Disc — 25 cm (fine) | 4.7 | 15.5–217 | 4.4–7.2 | Wilfey, 1998 |
| | 12.6 | 15.5–217 | 5.6–8.2 | Wilfey, 1998 |
| Disc — 25 cm (coarse) | 4.7–12.6 | 15.5–217 | 5.1–5.9 | Wilfey, 1998 |
| Disc — 23 cm | 1.6 | 20.2–292 | 3.2–7.9 | Wilfey, 1987 |
| | 3.2 | 20.2–255 | 4.6–6.9 | Wilfey, 1987 |
| | 4.4 | 66.0–140 | 5.4–6.7 | Johnson, 1993 |
| | 5.8–7.6 | 20.3–280 | 4.9–6.2 | Wilfey, 1987 |
| | 12.4–12.8 | 13.5–140 | 6.1–8.5 | Pöpel et al., 1993; Johnson, 1993 |
| Disc — 18 cm | 24.9 | 13.5–69 | 8.8–9.5 | Pöpel et al., 1993 |
| | 22.0 | 14.5 | 8.1 | Stenstrom, 1997 |

^a for diffuser submergence of 1.75 m

^b airflow rate per diffuser surface area

^c SOTE/H_s where H_s is diffuser submergence

1 m = 3.28 ft; 1.0 m³/h-m² = 0.059 scfm/ft²

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Gas flow rates are often reported on a per diffuser element basis for discs, domes, tubes, and nonporous diffusers. For plate and panel diffusers, airflow per effective projected surface area is used. In some cases, tubes are rated on a per tube length basis. When comparisons are made between diffusers of different shape or size, it is most useful to express airflow on an effective area basis. This expression is not difficult to apply for ceramic and plastic discs and plates, but requires an understanding of the contributing surface area for perforated membrane diffusers. For tube diffusers, the contributing area is often difficult to assess since airflow distribution is not only dependent upon the perforated (or porous) area but also on the means for distributing air to the media and the airflow rate.

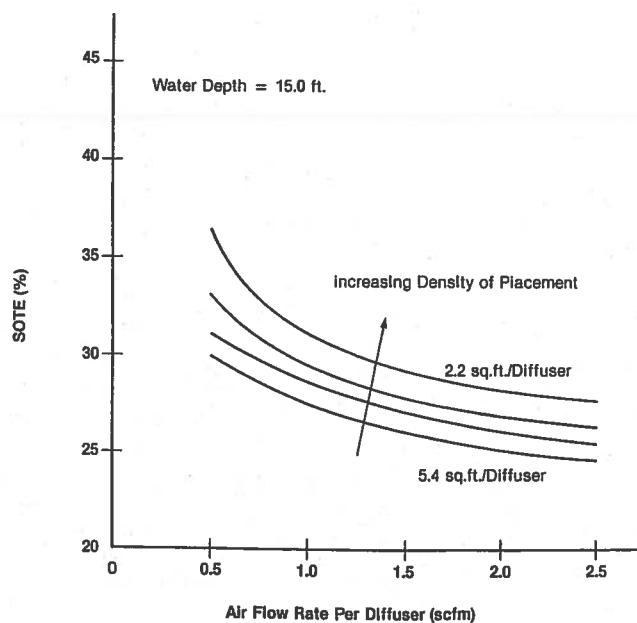
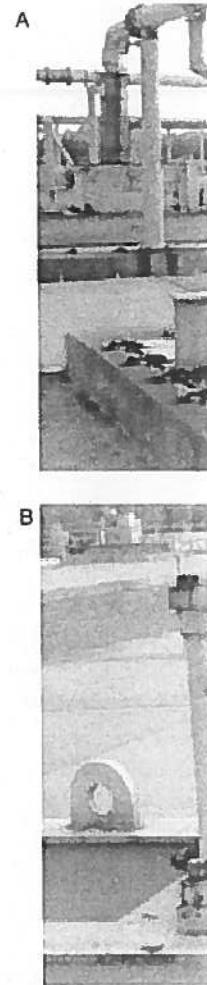


FIGURE 3.25 Impact of diffuser density on efficiency.

are illustrated in Tables 3.5 and 3.8. Generally, an increase in diffuser density results in an increase in SOTE for the same gas flow rate per diffuser. In 1976, Paulson tested dome diffusers in a 4.6 m (15 ft) deep tank and found a linear relationship between diffuser density and SOTE in the range of densities of 6.9 to 18.3 percent (Figure 3.25). Two airflow rates were evaluated in this work. Since that time, numerous other investigations have shown similar results (EPA, 1989). Huibregts et al. (1983) evaluated the effects of density of disc and dome placements in a 6.1 × 6.1 m (20 ft × 20 ft) test tank. Grid placements of 23.8 cm (9.375 in) ceramic disc diffusers were studied at densities of 7.6, 11.6 and 15 percent. Header spacing was held constant at 0.76 m (2.5 ft). At all three test submergences they found that SOTE increased with diffuser density, but the increase was not linear in all cases. A comparison between dome diffusers (17.8 cm [7 in] in diameter) and the same disc diffusers indicated that, at the same density of diffuser number, the discs were more efficient. This result can be attributed to the higher projected surface area provided by the disc, which was about 70 percent greater than the dome. Yunt and Hancuff (1979) reported similar findings for dome and disc performance. There appears to be an upper limit to diffuser density where little improvement in SOTE will be found. This limit will depend on the diffuser size, airflow rate, and spacing. For example, a 23 cm (9 in) disc diffuser, at a submergence of 4.3 m (14.2 ft) and gas flow of 1.6 m³/h (1 scfm) per diffuser, exhibited little increase in SOTE at densities > 14 percent (Sanitaire, 1976–1986). On the other hand, tests with a 51 cm (20 in) membrane disc indicated that SOTE increased to a density of 26 percent, but the increase was small. A 40 percent increase in the number of diffusers required to increase the density from 18 to 26 percent resulted in only a five percent increase



COLOR FIGURE
(A) aeration tank turb

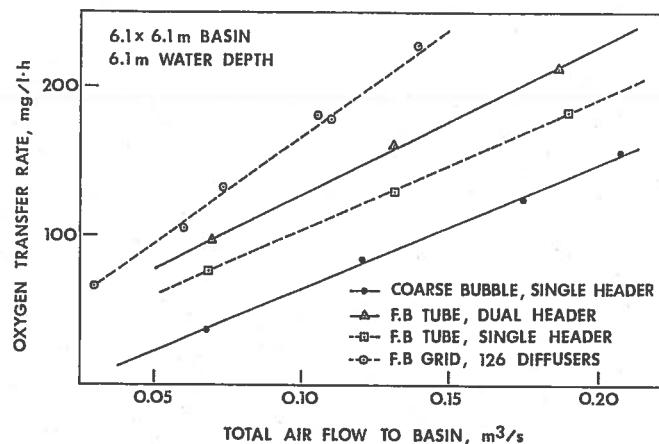


FIGURE 3.26 Efficiency vs. airflow for selected configurations. (From Huibregtse, G.L. et al. (1982). "Factors Affecting Fine Bubble Diffused Aeration," unpublished.)

TABLE 3.9
Clean Water Oxygen Transfer Efficiency Comparison for Selected Diffusers (EPA, 1989)

| Diffuser Type and Placement | Airflow Rate (m_N^3/h /diffuser) ^a | SOTE (%) at 4.6-m Submergence |
|----------------------------------|---|----------------------------------|
| Ceramic plates — grid | 35–85 $m_N^3/h \cdot m^2$ | 26–33 |
| Ceramic discs — grid | 0.6–5.3 | 25–40 |
| Ceramic domes — grid | 0.8–3.9 | 27–39 |
| Porous plastic discs — grid | 0.9–5.5 | 24–35 |
| Perforated membrane discs — grid | 0.8–3.9 | 16–38 |
| Rigid porous plastic tubes | | |
| Grid | 3.8–6.2 | 28–32 |
| Dual-spiral roll | 4.7–17.3 | 17–28 |
| Single-spiral roll | 3.1–18.8 | 13–25 |
| Perforated membrane tubes | | |
| Grid | 1.6–6.2 | 22–29 |
| Mid-width | 3.1–9.4 | 16–19 |
| Mid-width | 3.1–18.8 | 21–31 |
| Single-spiral roll | 3.1–9.4 | 15–19 |
| Coarse bubble diffusers | | |
| Dual-spiral roll | 5.2–15.5 | 12–13 |
| Mid-width | 6.6–7.1 | 10–13 |
| Single-spiral roll | 15.7–55.0 | 9–12 |

^a Except for plates

1 m = 3.28 ft; 1 m_N^3/h = 0.64 scfm; 1 $m_N^3/h \cdot m^2$ = 0.059 scfm/ft²

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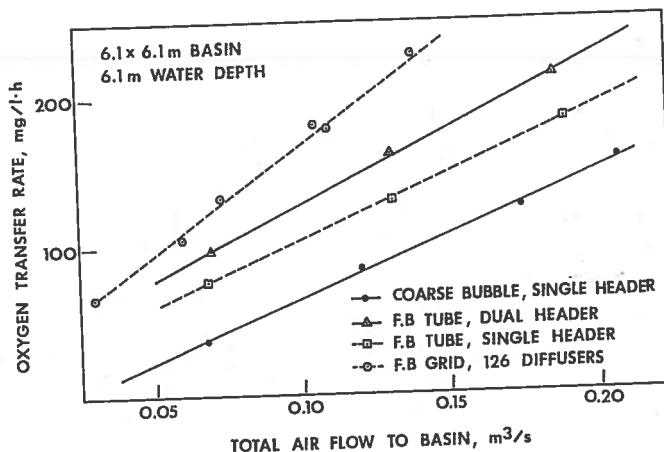


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| Ceramic discs — grid | 0.6–5.3 | 25–40 |
| Ceramic domes — grid | 0.8–3.9 | 27–39 |
| Porous plastic discs — grid | 0.9–5.5 | 24–35 |
| Perforated membrane discs — grid | 0.8–3.9 | 16–38 |
| Rigid porous plastic tubes | | |
| Grid | 3.8–6.2 | 28–32 |
| Dual-spiral roll | 4.7–17.3 | 17–28 |
| Single-spiral roll | 3.1–18.8 | 13–25 |
| Perforated membrane tubes | | |
| Grid | 1.6–6.2 | 22–29 |
| Mid-width | 3.1–9.4 | 16–19 |
| Mid-width | 3.1–18.8 | 21–31 |
| Single-spiral roll | 3.1–9.4 | 15–19 |
| Coarse bubble diffusers | | |
| Dual-spiral roll | 5.2–15.5 | 12–13 |
| Mid-width | 6.6–7.1 | 10–13 |
| Single-spiral roll | 15.7–55.0 | 9–12 |

^a Except for plates
1 m = 3.28 ft; 1 m³/h = 0.64 scfm; 1 m³/h·m² = 0.059 scfm/ft²

FIGURE 3.27

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3.4.2.2.5 Longitudinal The influence mean partial pressure the longer re demonstrated

Tank: 20 ft x 20 ft
 Power: ~ 1 hp delivered/1,000 cu ft for rigid porous plastic tubes
 Power: ~ 5 hp delivered/1,000 cu ft for ceramic domes

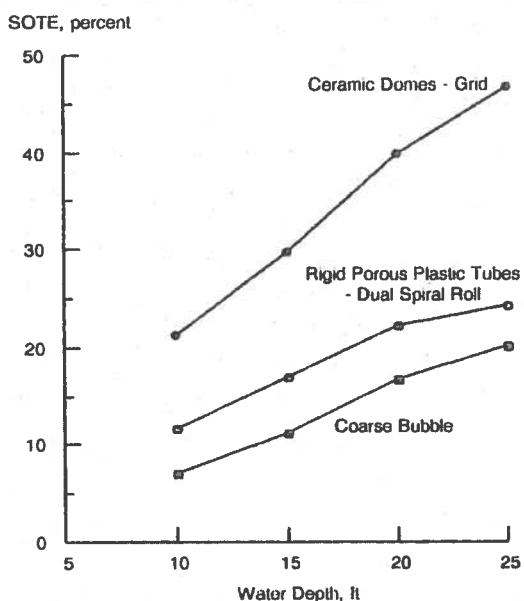


FIGURE 3.27 SOTE vs. submergence for selected diffusers.

configurations the value of v_w is only one to two times greater than v_s producing longer bubble residence times and concomitant higher efficiencies. Schmit et al. (1978) showed that mid-width configurations were more efficient than spiral roll when the SOTR (and airflow rate) increased or when the submergence increased for the same basin width. Bewtra and Nicholas (1964) and, later, Rooney and Huibregtse (1980) observed the same phenomenon.

Clearly, there is no simple relationship that can be used to express the relationship between placement and performance. Diffuser type, gas flow rate, and basin geometry all play an important role in the efficiency of the aeration system. Figure 3.26 taken from Huibregtse et al. (1983) summarizes the importance of diffuser pattern for several diffuser placements. Entering this curve for a given SOTR_v will indicate the relative amount of gas flow required to achieve that value for given configurations in the same basin geometry. In this work, which confirms much of the earlier research, the grid configuration is most efficient, followed by dual and single roll configurations. Table 3.9 also provides typical results of clean water tests for a variety of diffuser system placements.

3.4.2.2.5 Diffuser Submergence

The influence of diffuser submergence on SOTE is primarily the result of the higher mean partial pressure of oxygen in the basin (and thus a greater driving force) and the longer residence time of the bubble in contact with the water. This influence is demonstrated in Tables 3.2 through 3.9 and illustrated in Figure 3.27 for three

by foulants or media deterioration and is calculated as the ratio of the mass transfer coefficient, $K_L a$, of a fouled diffuser to that of a new diffuser, both measured in the same process wastewater. The value of F was theorized to decrease from 1.0 with time in service, but the actual model of the dynamics of this decrease could not be identified. A linear model was assumed for simplicity, and the fouling rate, f_F , was estimated for a number of sites. This controlled study using portable headers equipped with ceramic disc diffusers demonstrated that values of F appeared to correlate with foulant accumulation and the changes in uniformity of operating pores. These values ranged from 0.99 to 0.56 over the 12-month study. The lower values of F were from plants that received a significant industrial waste contribution.

It is noted that there was significant temporal variation in foulant accumulations at these plants. Further, the effect of foulant (or deterioration) may depend on position within the aeration tank. Foulant accumulations have been found to be highest at the influent end of plug flow tanks in some instances and randomly distributed in others (EPA, 1989). No definitive studies have been performed, however, to quantify the independent effects of fouling/deterioration temporally or spatially on OTE. Clearly the dynamics of fouling are not understood well enough to effectively apply the fouling factor correction to the oxygen transfer relationship for aeration system design.

3.5 DIFFUSED AIR SYSTEM DESIGN

A typical diffused air system is illustrated in Figure 3.43. The air supply system consists of blowers, air filters, air piping, and airflow control equipment, including flow meters and flow control valves. The diffusion system consists of a series of headers and lateral piping in the aeration tank and the associated diffusers. The system may be arranged in a series of grids (as depicted in the figure) so as to allow for proper airflow distribution or in laterals running longitudinally along one or both sides of the basin with diffusers placed on one or both sides of the lateral in a tapered or regular spacing format. Other diffuser arrangements are also used on occasion as described earlier in Section 3.3. The basin may be rectangular, square, circular, or oval with a number of different $l/w/d$ (or radius/depth) ratios. Aeration tanks may be laid out in series using common wall construction, folded arrangements, or individual, independent basins. This section presents the procedures and considerations required for the design and installation of a diffused air system. A number of steps are involved in the process. A brief outline of the process is first presented followed by a more detailed description of the design elements.

3.5.1 STEPS IN DESIGN

One suggested format used in the selection and design of a diffused air system is given below. It should be emphasized that there are any number of approaches that may be followed. The procedure given below has proven to be an effective approach for the design of most systems.

- determine flows and loads
- select a process flowsheet that meets the objectives of the system design

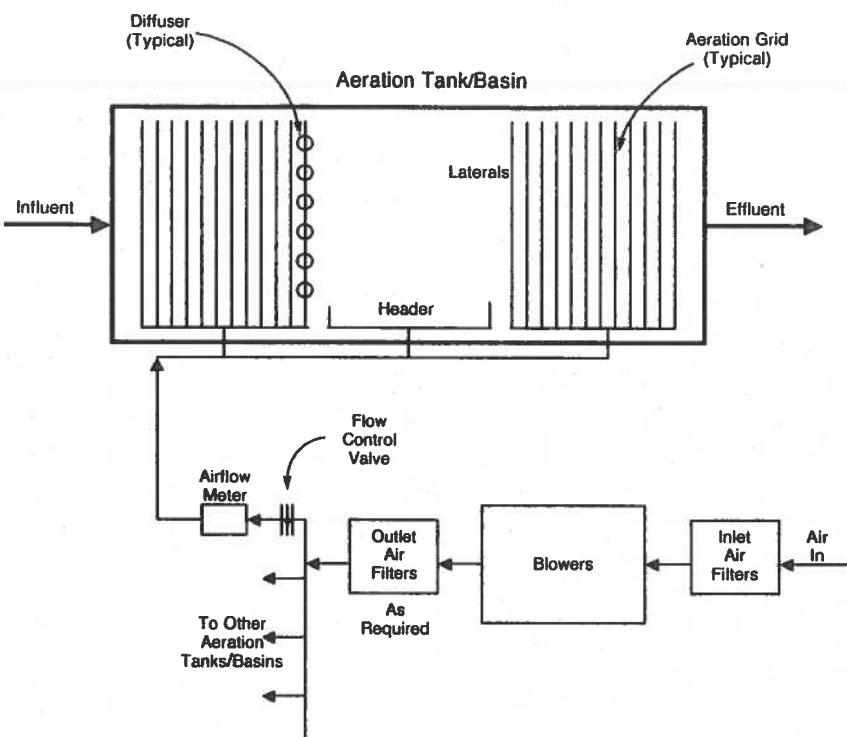


FIGURE 3.43 Schematic of diffused air system.

- establish design criteria for the process selected
- size the basins
- configure the basins
- determine the temporal and spatial oxygen demand for the process
- select the diffusers
- determine the appropriate airflow rates and their distribution
- check for mixing
- configure the diffuser system
- design the blower system
- review system flexibility
- design air piping
- select and design appropriate control system
- retrofit considerations

3.5.1.1 Determine Flows and Loads

Design wastewater flow and loads should be established for the entire range of operating conditions anticipated. From these, system oxygen requirements can be calculated. Load parameters of interest include carbonaceous oxygen demand, nitrogenous oxygen demand and any inorganic oxygen demand that might occur. Waste streams should include all return side-streams including sludge handling

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and internal recycle flows. Important load and flow conditions that must be determined are:

- minimum month to establish blower and diffuser turndown requirements
- average conditions (nitrifying and nonnitrifying), to establish normal operating conditions for blowers and other system components
- maximum month, to determine the maximum condition under which process oxygen requirements must be met to meet permit requirements
- peak day/ 4 hour peak (considering diurnal fluctuations), to establish the maximum operating point for all system components, including diffusers, air supply piping and blowers

3.5.1.2 Select Process Flowsheet

The selection of a process flowsheet depends on a number of factors. Among the more important considerations are:

- achievement of target pollutant removal (carbonaceous oxygen demand, nitrification, nitrogen removal, phosphorous removal, etc.)
- achievement of process stability (solid/liquid separation, qualitative or quantitative shock loads, etc.)
- site-related issues (footprint, near residential, etc.),
- low yield of biosolids
- low oxygen requirements
- the efficient removal of pollutants (e.g., plug flow vs. completely mixed)

The selection of the appropriate flowsheet will impact directly on the selection and design of an aeration system. Examples are cited below.

Conventional activated sludge processes designed for BOD and solids removal often use plug flow configurations or basins-in-series to achieve efficient removal of contaminants. The oxygen demand in these systems is highest near the influent end thereby requiring the highest transfer rate. If aeration is tapered by means of diffuser placement, the highest diffuser density, which is normally the most efficient, is used at the influent end. Counteracting this, however, is that the value of alpha is normally the lowest in this zone, and the requirement for airflow rates is the highest. Furthermore, there is a greater likelihood that diffuser fouling will take place where the load is highest. As a result, there may be a limit on the sizing and configuration of the basin due to the characteristics of the diffused air system that is selected and the wastewater that is being treated.

The requirements for ammonia oxidation will dictate longer MCRTs and greater oxygen requirements than the conventional carbonaceous systems. In fact, for many years, operators of conventional BOD removal facilities tried to avoid nitrification in the warmer seasons by maintaining low MCRTs to hold their oxygen requirements (and power consumption) down. It is now very evident that although nitrification will increase oxygen demand, the value of alpha in porous diffuser systems will significantly increase, resulting in oxygen transfer rates in nitrifying systems

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that are not much different (or even higher) than those for carbonaceous systems. Thus, the operation of nitrifying systems may not have any significant effect on blower sizing and power consumption. It is likely, however, that the distribution of oxygen demand through the system (plug flow) will differ significantly from the carbonaceous system.

Nitrogen removal may be accomplished in single-sludge systems by the incorporation of anoxic zones within the reactor system. This zone may be located at the influent or effluent end of the process and serves as the zone where nitrates are converted to nitrogen oxides and nitrogen. The flowsheet may have significant impacts on the aeration system where oxygen demand and alpha are concerned. If nitrate is reduced by organic matter in the influent stream, then some oxygen demand is satisfied reducing the requirements for oxygen (the nitrate serves as the electron acceptor in place of DO). Furthermore, the value of alpha for porous diffusers following the anoxic zone may be elevated by virtue of the removal of some organic matter. Whether to take advantage of these "credits" is a matter of engineering judgment. Often, they are ignored and presumed to add a degree of conservatism in the design. One important factor to consider in the aeration system design for this flowsheet is the type of diffuser. In some designs, a variable anoxic zone is used to provide greater flexibility in seasonal operation. Since these zones may be aerated or anoxic, diffusers may be idle for significant periods of time. Perforated membranes are often used for this application.

Phosphorus removal by biological methods will normally call for anaerobic zones located within the reactor system. Anoxic zones may also be incorporated into biological phosphorus removal plants where nitrification is required. The impact of these zones on alpha has been shown to be positive resulting in higher alpha values than observed for the carbonaceous removing facilities without these zones. It appears that alpha will approach the values found in high MCRT facilities that nitrify.

Process stability is often an important consideration in process design. The use of selectors has become popular in many new and retrofit designs to insure improved settleability of sludge. Inherent in biological nutrient removal schemes, aerobic, anoxic, or anaerobic selectors may be included in carbonaceous systems as well. These selectors will typically result in higher observed alpha values for porous diffuser systems as compared with systems without selectors. The magnitude of this improvement is not well documented, but it will be wastewater and selector design specific, likely approaching values found in high MCRT processes. The step aeration process may also be used to achieve process stabilization by attenuating the effects of load and flow on the system, approaching a completely mixed flow regime. As seen earlier, the step aeration process will even out spatial oxygen demand and alpha values but may result in somewhat lower mean-weighted alpha values. Often provided in plug flow systems to add operational flexibility, the engineer must evaluate the impact of this flow regime on oxygen transfer distribution. The ultimate in attenuating qualitative and quantitative shock loads to the aeration system is the completely mixed flow regime. This scheme is the easiest for designing and controlling the aeration system since there is little or no spatial variation in oxygen demand. The completely mixed flow regime generally results

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Site constraints may dictate flowsheet selection. Small footprints available for the facility may dictate the use of deep aeration tanks, the use of high purity oxygen systems, or deep shaft reactors. Each has unique characteristics that will affect aeration system design. All three systems will result in higher partial pressures of oxygen and therefore, higher transfer rates. The details of these systems are found elsewhere in this book.

Smaller communities may elect to use processes that are highly stable and require minimum operational requirements. Extended aeration systems, designed for high MCRT operation will have high total oxygen demands (mass of oxygen required per unit oxygen demand satisfied) where a significant portion of the oxygen is required for endogenous respiration. These systems may be designed in a number of configurations including oxidation ditches, aerobic or facultative lagoons, completely mixed processes, or conventional plug flow systems. Aeration system design for these processes will generally follow the same guidelines as that used for the flow regimes described above with the exception of the use of higher overall oxygen requirements. At the other extreme are the highly loaded, high-rate activated sludge systems sometimes used as a pretreatment step in industrial waste flowsheets. High-rate processes are characterized as systems with lower overall total oxygen requirements at the cost of higher biomass yields, as compared with conventional designs. Generally, they will exhibit lower porous diffuser alpha values than carbonaceous removal systems and will potentially produce a greater opportunity for diffuser fouling. Nonporous diffusers are excellent candidates for this process.

3.5.1.3 Establish Process Design Criteria — Oxygen Transfer Considerations

Several design criteria are important to the estimation of system oxygen requirements both temporally and spatially. They include:

- maximum wastewater temperature and the corresponding MCRT which are used to estimate maximum carbonaceous (and nitrogenous) oxygen requirements
- minimum wastewater temperature and the corresponding MCRT which are used to estimate minimum carbonaceous (and nitrogenous) oxygen requirements

- expected extent of denitrification, if system is designed to denitrify, to estimate oxygen "credits" in oxygen requirements calculations
- basin configuration, which will be used to estimate spatial distribution of oxygen demand
- wastewater flow distribution (step and recycle points and flows which will be used to estimate oxygen demand distribution)
- design life and process growth patterns

3.5.1.4 Size the Basins

The required sizing of the aeration basins, the anoxic, aerobic and anaerobic zones, and selectors is determined by the biological process design methodology selected by the engineer and is outside the scope of this discussion.

3.5.1.5 Configure the Basins

Once total reactor volumes are calculated, the number, size and shape of the basins must be determined. Basin dimensions are important considerations in aeration system design. Depth of submergence influences both the OTE, the value of the steady-state DO saturation concentration, and the static pressure that the blowers must overcome. The basin length to width ratio will affect spatial oxygen demand and the physical layout of the diffused air system. Points of wastewater inflow, recycle flows, and return sludge will affect the magnitude and distribution of oxygen demand. The selection of a single basin severely constrains the selection of diffusers and diffuser layout in that porous diffusers require routine servicing and must be readily accessible. To avoid basin shutdown, diffusers need to be placed on retrievable lifts and should be capable of long-term operation without maintenance.

3.5.1.6 Determine Temporal and Spatial Oxygen Demand

Oxygen demand is dictated by the quality and quantity of wastewater treated and will vary over the life of the facility, normally being lower in initial years of operation and increasing to the design life of the facility. Hourly, daily, and seasonal variations will also occur and must be estimated to ensure that process oxygen requirements are properly met in accordance with the process design objectives. An evaluation of the potential impacts of periodic low mixed liquor DO on process performance and operating characteristics should be performed to determine the range of conditions that should be considered in estimating oxygen requirements. The loading conditions normally considered are outlined in Section 3.7.1.1 above. *Design of Municipal Wastewater Treatment Plants, Vol. 1, Manual of Practice 8* (WEF, 1991) provides an excellent discussion of wastewater flow and loading considerations for design and should be consulted.

Typically, oxygen demand calculations will be made for a variety of process loading conditions as appropriate for the particular system. For example, ammonia oxidation may be required from spring through fall but not the remainder of the year. The calculation of nitrogenous oxygen demand would only be necessary during this period and may or may not control aeration system design depending upon loads

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important consideration when examining system flexibility under a variety of operating conditions. As described earlier, the influent end of plug flow basins produces high oxygen demands, low alpha values and greater opportunities for diffuser element fouling and deterioration. Since many of the nonporous diffuser systems are less susceptible to fouling and exhibit higher alpha values in wastewater, the use of hybrid aeration systems, which incorporate nonporous diffusers at the influent end and porous diffusers through the remainder of the aeration basin, is sometimes practiced. It should be noted, however, that most nonporous diffusers produce lower back pressures than porous diffusers and therefore require careful selection of airflow orifice controls to ensure appropriate airflow distribution throughout the system.

Designers attempt to provide sufficient process operational flexibility in their facility. This provision is often accomplished by providing several alternative flow regimes to handle a number of different process objectives and to improve system stability. Step feed or sludge reaeration may be used to supplement a conventional plug flow system to accommodate fluctuations in flow or load that would impact system performance. Process loading may be changed to accommodate different seasonal discharge permit requirements. The facility is normally designed in anticipation of future growth and, therefore, is typically underloaded early in the design period. All of these factors will affect the design of the aeration system and require that sufficient flexibility be provided to meet the variable oxygen demands that will occur. The components of the aeration system that must be designed to meet these changes include the blowers, air piping and appurtenances, and the diffusers. Air piping and blowers are addressed in later sections.

All diffusers have an allowable range of airflow rates that can be applied per unit. The range depends on size, shape, orifice diameter, and other characteristics of the device. The lower limits of this range are dictated by uniform airflow distribution from the system, and upper limits are those that cause diminishing improvements in oxygen transfer rate. To illustrate the constraints on airflow, consider the example of a typical ceramic disc diffuser. For this device, the allowable ratio of maximum to minimum airflow is about 5:1. Based on the change in OTE with airflow, the resulting ratio of maximum to minimum oxygen transfer rates would be approximately 4:1. It should be emphasized that diffuser density will play a significant role in this calculated turndown capacity. If turndown flexibility is dictated by growth over the design life of the facility, it is possible to provide only enough diffusers to meet initial diurnal and seasonal demands and to make provisions to add additional units over time to meet the ultimate demands of the system. In performing these calculations, it is important to consider mixing requirements as well as oxygen transfer rates. In systems operating under initial load conditions and in tapered aeration systems near the effluent end, mixing often controls airflow rate and may be an overriding consideration in diffuser layout and selection.

In the example above, the relationship between airflow rate and OTE was used to estimate oxygen transfer rate turndown. It is important to emphasize that this relationship may be different for different diffusers (see Table 3.7) and may change over time in process wastewater. When selecting a diffuser element, an examination of this relationship may be important. An example of this process is provided in the following. Figure 3.44 (Marx, 1998) provides data on the airflow rates and SOTE

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FIGURE 3.4

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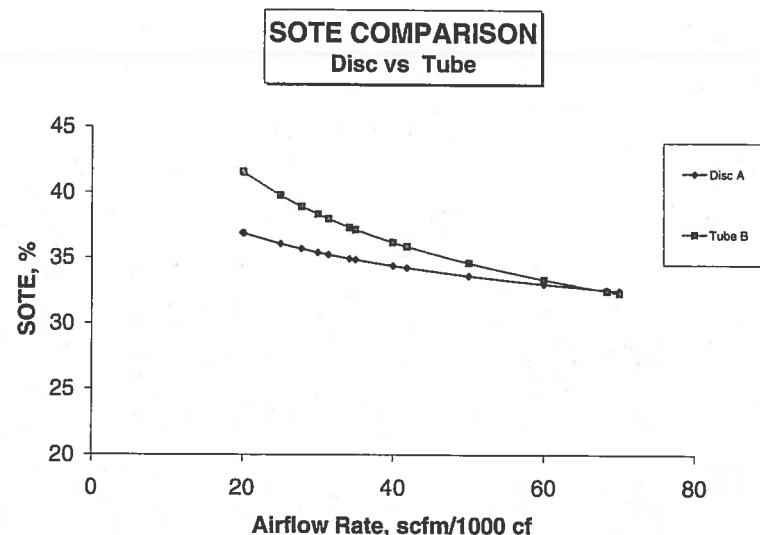


FIGURE 3.44 Comparison of performance of a disc and a tube vs. airflow (Marx, 1998).

values for two competing diffuser systems for two parallel grids in an aeration system. Disc A exhibits a lower sensitivity to airflow rate. Since the blower capacity is set, the maximum oxygen transfer rate is at the point where the two systems must provide the same SOTE. In this example, tube B will provide substantial benefit to the owner over diffuser B because the SOTE is much higher at average conditions where the system will normally operate. Note also that the turndown flexibility of disc A is significantly higher.

The operation of diffusers at their lowest allowable airflow rate has been shown to be the most efficient operating point for porous diffusers. It is tempting to operate a system at this low value but this practice can lead to operational problems. At low airflow, uniform air distribution across the diffuser may be difficult to obtain. Also at this low airflow, the head loss across the control orifice could also be low, requiring a change in orifice size to balance airflow throughout the entire system. If maldistribution occurs either along an individual diffuser header or within the entire grid of diffusers in the system, foulant deposition can begin, which may lead to premature fouling and poor performance of the entire system.

The reliability of a given diffuser system depends upon the mechanical integrity of the system and the maintenance required to ensure a high level of performance. Critical components to be considered in evaluating system integrity include the diffuser material, diffuser supports, diffuser connections, piping supports, and submerged air piping. Considerations for the diffuser material include physical and chemical resistance to the wastewater. Designers should incorporate mounting details that minimize build-up of stringy materials on diffuser piping. The supports and connections should be able to withstand stresses that will occur both during construction and operation. For example, tube-type diffusers will be subject to bending and relatively high stresses at the point of connection to the air header. Supports and air piping must be able to resist the dead weight of the equipment during installation as well as the buoyant

forces of the system under normal operation. Gasket materials must be flexible and resistant to chemical or biological attack.

Required maintenance of diffused air systems has been described above. All systems need some preventative maintenance, but porous diffusers are typically more susceptible to wastewater components that may lead to fouling or deterioration. Routine maintenance is site specific depending on wastewater characteristics, process loads, flow regime, and system operation. Maintenance is performed to control fouling and to replace diffuser components when they deteriorate. To maximize OTE and minimize costs, fouling must be controlled. As fouling progresses, head loss across the diffuser increases thereby increasing blower energy costs. This gradual increase in pressure must be considered in the design of porous diffused air systems. Typical designs allow for head loss to increase by about 3.4 to 10.3 kPa (0.5 to 1.5 psi) before cleaning. Management of fouling at a given installation includes the provision of effective wastewater pretreatment to remove most of the fibrous material and heavy suspended solids. Air bumping is sometimes recommended to remove some deposits from the diffuser. The incorporation of in situ acid gas cleaning may serve to slow down fouling rate in some wastewaters. For systems that do not provide portable removal of diffuser headers for inspection, basins should be designed to allow isolation and rapid dewatering of the basin for appropriate cleaning and inspection of diffuser systems. Access to plant water that can deliver a high flow at about 415 kPa (60 psig) should be provided for diffuser cleaning.

All diffusers may be subject to gradual deterioration although those constructed from ceramic and stainless steel have demonstrated very long service lives. Deterioration may be due to buildup of inorganic materials within the diffuser that cannot be removed by ordinary cleaning methods or through breakdown of the diffuser material itself. The rate of deterioration depends on wastewater characteristics and diffuser type. The useful service life of a diffuser is generally considered to have been reached when the cost of replacement offsets the increased operating cost of the deteriorated element.

An important element in the design of the aeration system is the appropriate selection of the diffuser. Special testing of candidate diffusers using test headers or pilot plants is often justifiable when wastewater characteristics are suspected to have a significant influence on diffuser performance and/or service life. Present worth cost analyses are appropriate for both selecting diffusers and evaluating cost effectiveness of diffuser replacement.

3.5.1.8 Determine Aeration Rates

There are a number of different approaches to the design of diffused air systems. The procedure described below represents an iterative process where total airflow is calculated from the required transfer rate, OTR_f , and the estimated transfer efficiency, SOTE, for the diffuser system that was selected. The number of diffusers is ultimately determined based on the calculated total airflow rate. To start the process, the designer must determine the diffuser pattern (e.g., full floor grid, spiral roll) and whether tapering of airflow to meet demand will be implemented by varying diffuser density (if tapering is, in fact, selected as a design factor). If the flow regime is plug

flow or basin or grids (typical influent to effluent whereas in deep some degree of distribution o

Once the have been selected oxygen demand may use Equations for the select submergence, equipment may then perform designer selected a total airflow airflow rate per and compared be readjusted density is used whether in a f in some other any diffuser type

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flow or basins-in-series, the aeration system may be laid out as a series of sectors or grids (typically three or four), each with a diffuser density that decreases from influent to effluent sectors. For completely mixed regimes, tapering is not practiced, whereas in dedicated step systems, the designer may or may not elect to provide some degree of diffuser tapering or may rely on adjustments in airflow rate for the distribution of oxygen along the tank length.

Once the oxygen requirements (AOR) have been calculated and the diffusers have been selected, it is possible to estimate the required airflow rate to meet the oxygen demand. Since the AOR will equal the OTR_f at steady state conditions, one may use Equation (2.53) to determine the standard oxygen transfer rate (SOTR) for a given grid within the tank. The designer can then determine the appropriate SOTE for the selected diffuser system. This value depends on the diffuser airflow rate, submergence, placement pattern, and diffuser density. It is often available from the equipment manufacturer. The calculation of total airflow rate for the given sector is then performed using Equation (2.51). An iterative process occurs whereby the designer selects an airflow rate per diffuser and estimates a diffuser density. Once a total airflow is calculated, the required number of diffusers for the preselected airflow rate per diffuser is determined. The diffuser density is subsequently calculated and compared with the estimated value. Either diffuser airflow rate or density can be readjusted until appropriate closure is achieved. It should be noted that diffuser density is used in its broadest definition to identify numbers of diffusers per sector whether in a full floor grid, located along one or two longitudinal walls, or placed in some other pattern. The design procedure described above should be effective for any diffuser type or configuration.

In these calculations, it is necessary for the designer to have information on field conditions (process water temperature, atmospheric pressure), beta, alpha and its spatial distribution, the target process water DO, and the steady-state DO saturation concentration at 20°C and 101.4 kPa (1atm). One issue that the designer often faces is identifying the source for information on clean water performance data for the diffusers and on the appropriate values of alpha to use. This source should be the manufacturer of the equipment that was selected, although the information is sometimes unavailable or has been collected using nonstandard methodology. Today, most reputable manufacturers test their equipment in clean water using approved standard methods, but the information may be limited to a range of airflow, submergence, diffuser density, and pattern outside the actual system that is being designed. In those cases, the designer needs to estimate values of SOTE, preferably with the guidance of the manufacturer who knows the equipment.

The selection of alpha is often more difficult. If the manufacturer is unable to provide documented evidence of typical values for the facility being designed, it will be necessary to estimate values from the literature. Typical values of alpha for municipal wastewater have been presented in this text, but values for industrial or combined industrial/municipal wastewater are more difficult to obtain. Often the designer must ask for pilot studies with the wastewater and the selected diffusers to determine realistic alpha values. Since alpha varies with time of treatment (distance along a plug flow basin), the designer must also estimate appropriate values of alpha for each design sector if a plug flow regime has been selected. It is a good design

strategy to be conservative in the estimate of alpha, especially for porous diffusers, and to provide sufficient flexibility in the aeration system because of the uncertainty of this value.

3.5.1.9 Check for Proper Mixing

Once airflow rates have been calculated, it is important to determine whether the diffused air system will provide sufficient mixing in each design sector. Details on mixing requirements are described in Section 3.4.4. As described in that section, mixing requirements are based on experience, and the designer must rely on the experience of the manufacturer (if any) and reported data in the literature.

3.5.1.10 Configure Diffuser System

After the number of diffusers has been selected, the diffuser system may be configured. Several iterations may be required to ensure that the entire range of oxygen demands can be met without exceeding the recommended airflow rate per diffuser. Important design considerations include basin inlet conditions, wastewater and airflow patterns within the basin, ability to isolate and dewater individual basins, access to diffusers within the basin and availability of plant water.

The distribution of influent wastewater and return sludge flows to the inlet end of the basin (or along the basin where step feed alternatives are selected) should be carefully considered. Depending upon basin size and configuration, it may be advisable to distribute these flows across the entire width of the basin. This distribution may minimize localized high velocity gradients and poor initial mixing in the inlet zone.

Provisions should be made for partially filling the basin without allowing the incoming flow to cascade directly onto the diffusers and in-basin piping. A drain system that permits each basin to be dewatered in a reasonable period of time (normally 8 to 24 hours) should be provided if diffusers are floor mounted and inaccessible for servicing at tank-side. The basin floor should be sloped to allow complete drainage to occur without ponding and to facilitate easy removal of residual solids. One arrangement that has been effectively used is the construction of a drain trough along the longitudinal wall of the basin, with the basin floor sloped to the trough and the trough sloped to drain to a collection sump or dewatering manhole.

Diffusers should be arranged in the tank to allow space for walking and access. Access is necessary both for installation and maintenance. Spacing between diffusers on adjacent laterals, between grids, and between each basin wall and adjacent diffusers should be examined. A minimum clear walkway space of about 50 cm (20 in) is usually adequate. Basin and diffuser cleaning require water at moderate pressures (approximately 400 to 700 kPa [60 to 100 psi]) at the nozzles. Hydrants with appropriate hose connections should be placed at frequent intervals (typically about 60 m [200 ft]).

3.5.1.11 Blower System Design

The description and design of the blower system are found in Chapter 4. Temporal variations in oxygen demand should be considered in selecting the appropriate

number of diffusers to meet annual oxygen demand.

3.5.1.12 Sector Configuration

Sufficient flow is required to effectively clean each sector. The system must be designed to meet oxygen demands.

Providing oxygen in ways. Where the designer/owner constructs all of the plant in the basins, both of the mechanical The decision and owner play a role in selecting the system.

In any economical operation installed than operate only in reserve. Similar the design year provided in the Not all hold the same.

Flexibility in oxygen demand or cleaning sectors or basins.

3.5.1.13 Air Supply

The air supply system diffusion system filters along the cleaning system blowers are designed to minimize line losses and used to remove debris from the pipe.

The air piping system. Piping must be designed to withstand pressure fluctuations and prevent air entrainment.

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Chapter 4. Temporal
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number of blowers. Typically, the blowers are sized to allow one blower to meet minimum oxygen requirements, one or more blowers operating at full capacity to meet annual average requirements, and two or more blowers operating at full capacity to meet peak hour requirements.

3.5.1.12 System Flexibility

Sufficient flexibility should be provided to enable the system to be operated cost effectively over the entire life of the facility. The review should consider how the system will be operated at start-up and at the design loading. Over that period, the system must have sufficient flexibility to handle temporal variations in loading and oxygen demand, including hour-to-hour, day-to-day, and year-to-year variations.

Providing flexibility for year-to-year variations can be accomplished in several ways. Where the design period is relatively long and steady growth is expected, the designer/owner could choose to build a facility in phases. Another option is to construct all facilities in the first phase, with provisions for operating only a portion of the plant in the early design period. An additional alternative is to construct all of the basins, buildings, and major yard piping in the first phase and stage construction of the mechanical equipment (blowers, in-basin piping, and diffusers), as necessary. The decision on these alternatives depends on funding, projected growth patterns, and owner preference. A cost-effectiveness analysis of the alternatives is helpful in selecting the appropriate plan.

In any event, the final design must provide sufficient flexibility to allow economical operation over the design life. For example, if more basins and blowers are installed than are required to handle initial loads, capability should be provided to operate only as many basins and blowers as needed while holding the others in reserve. Similarly, if the number of diffusers required in a given basin or sector for the design year is significantly greater than required during start-up, space may be provided in the laterals to accommodate the maximum number of diffusers required. Not all holders need be filled with diffusers early in the design life.

Flexibility for handling seasonal, hour-to-hour and day-to-day variations in demand or changes in flow regime must also be provided in the system design. This is most often accomplished by providing the capability to adjust airflow to various sectors or basins in response to spatial and temporal changes in demand.

3.5.1.13 Air Piping Design

The air supply system delivers atmospheric air or high purity oxygen to the air diffusion system. It consists of three basic components: air piping, blowers, and air filters along with other conditioning equipment including gas injection diffuser cleaning systems. The air piping delivers air from the blowers to the diffusers. The blowers are designed to develop sufficient pressure to overcome the static head and line losses and deliver the required airflow to the diffusion system. Air filters are used to remove particulates from the inlet air stream to the blowers and may also be used to protect porous diffusers from air-side foulants.

The air piping should be designed to permit cost-effective installation and operation. Piping materials should be selected to provide the degree of durability (including

resistance to mechanical damage, corrosion, and sunlight degradation) appropriate for the facility. Commonly used piping materials include carbon steel, stainless steel, ductile iron, fiberglass reinforced plastic (FRP), high-density polyethylene (HDPE), and polyvinyl chloride (PVC). Carbon steel, ductile iron, and FRP are the materials most often used for delivering air from the blowers to the basins because of their strength. Within the basin, stainless steel, HDPE, and PVC are often used because of their resistance to corrosion. The choice is typically made at the droplegs into the basin. The choice between stainless steel and PVC for the air headers depends on the structural requirements of the diffuser connection. Stainless steel is often used for tube diffusers because of the cantilevered load applied to the lateral piping. However, PVC has been successfully used in tube installations where the connection between tube and lateral pipe has been designed for this force.

Both permanent flow meters and flow points for portable meter installation need to be properly located to allow accurate airflow determinations. An adequate number of flow points should be provided as required by the control requirements of the facility.

Piping should be sized to provide acceptable head loss at maximum airflow, including a head loss between the last positive flow split and the farthest diffuser of less than 10 percent of the loss through the diffuser. Losses through the blower inlet filter, control valves, and fittings all need to be considered in establishing total blower discharge pressure requirements. Basic principals of fluid mechanics can be used to determine head loss in air piping systems. At the rates of flow and velocities found in these systems, air can be treated as an incompressible fluid within the pipe and the Darcy-Weisbach equation can be used to determine head loss. An excellent source for the details of air piping design can be found in the *Design Manual, Fine Pore Aeration Systems* (EPA, 1989).

3.5.1.14 Control System Design

The control system is selected to meet the objectives of the wastewater treatment facility. A description of aeration control systems is found in Chapter 9. The design of this system is beyond the scope of this text but can be found elsewhere (EPA, 1989).

3.5.1.15 Retrofit Considerations

The retrofit of an aeration system is site specific. Many of the same considerations that apply to new systems apply to retrofit installations. These considerations include process oxygen requirements, diffuser selection, and configuration of the aeration system. There are some factors, however, that the designer cannot control such as basin configuration and flow regime.

In most instances where diffused air systems are being retrofitted, the existing air piping sizes are adequate for upgrading the system. Because the total airflow rate may decrease due to the higher efficiency diffusers, the size of the existing blower discharge headers and air mains that deliver air to the basins will usually be sufficient. The drop pipes into the basin may also be large enough. Replacement and recalibration of air metering devices must be considered at this time. The designer must also carefully check to determine if air piping is properly located to provide the air

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on) appropriate stainless steel, polyethylene (HDPE), etc. The materials used because of their low cost and ease of installation depend on what is often the lateral piping. The connection

Installation needs adequate number of blowers of the facility. Maximum airflow, the best diffuser of the blower inlet meeting total blower capacity can be used to velocities found in the pipe and system. An excellent guide is the *ASCE Standard Practice Manual, Fine*

ewater treatment plant. The design criteria (EPA, 1989).

ne considerations include location of the aeration control such as

itted, the existing total airflow rate of the existing blower usually be sufficient. Design and recalibration designer must also provide the air

distribution and flow control capabilities required. Existing air distribution piping should be inspected for leaks, corrosion, and other conditions that may lead to premature failure.

Air filters will protect blowers from particulate intrusions but will not protect diffusers from air contaminants already in the downstream piping such as dirt, rust, or scale that were produced due to internal pipe corrosion, leaks or physical damage. Thorough cleaning of the air piping system may be required in some situations. Some designers prefer to provide air filters downstream of the blower discharge or in the drop pipes to protect new piping placed within the basin from debris accumulated in the older air distribution mains.

3.5.2 DESIGN EXAMPLE

The following example has been developed to illustrate one method for the design of a municipal wastewater activated sludge aeration system using diffused air aeration. The system will be a new design for 20 years into the future. The projected flow for this municipality is $0.232 \text{ m}^3/\text{s}$ (5.3 MGD). The current average flow is $0.114 \text{ m}^3/\text{s}$ (2.6 MGD). The loading and process conditions are presented below.

**Process Loading Conditions for Municipality — 20 Year Design
(lb/d = 2.205 × kg/d)**

| Variable | Min Month | Average Nonnitrifying Month | Average Nitrifying Month | Maximum Month Nitrifying | Peak Day Nonnitrifying |
|-------------------------|-----------|-----------------------------|--------------------------|--------------------------|------------------------|
| AOR, kg/d | 1621 | 2454 | 4392 | 5255 | 5515 |
| BOD ₅ , kg/d | 2494 | 2993 | 2993 | 3492 | 5805 |
| Temp, °C | 10 | 15 | 20 | 25 | 25 |
| Nitrifying | No | No | Yes | Yes | No |
| NOD, kg/d | — | — | 1924 | 1924 | — |
| Design DO, mg/l | 2.0 | 1.0 | 2.0 | 2.0 | 0.5 |
| Flow condition | Sustained | Sustained | Sustained | Sustained | Short term |

Secondary treatment is to be provided to meet discharge requirements. Nitrification is required in summer months. The design requires an average hydraulic residence time of six hours with an average MCRT during the winter of four days and six days during the summer, when nitrification is required. The selected flow regime for this municipality is a plug flow activated sludge process consisting of four parallel aeration basins, each 7.0 m (23 ft) wide by 40 m (132 ft) long with a sidewater depth of 4.6 m (15 ft) (Figure 3.45). Diffuser submergence is 4.3 m (14 ft). Four basins may appear to be a large number for this small plant but were selected because of the wide variation in the process loading from start-up conditions to the 20-year design value (a doubling in flow and load over the 20 years). This variation is an economic issue. Initial construction costs will be higher but additional basins are needed for maintenance of the diffusers. Furthermore, operating costs may be

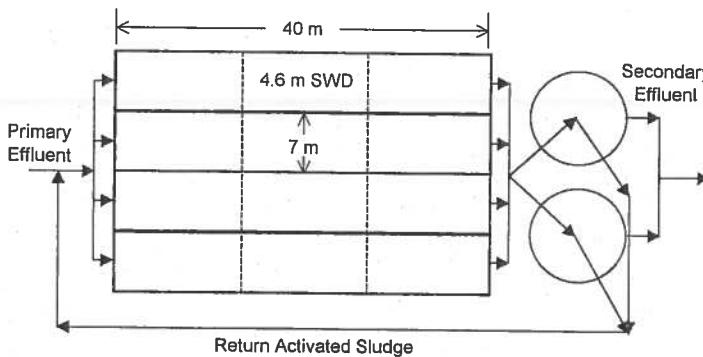


FIGURE 3.45 Design problem — aeration tank layout.

reduced since only the number of basins needed to satisfy maximum process oxygen requirements must be in service at any point in the life of the facility.

The next step in the aeration system design process is the estimation of spatial variations in process oxygen requirements along the plug flow basins. For the dimensions selected for these four parallel basins, it can be calculated that the hydraulic flow pattern for each basin would be approximated by three equal-sized basins in series. Therefore, it was decided that the air diffusion system would be segmented into three equal sized aeration zones. It was also determined that oxygen distribution would be achieved by tapering the diffusers in proportion to the oxygen demand in each of the three zones. The estimation of spatial oxygen demand was briefly described above and can be evaluated by appropriate biotreatment modeling or by the use of distribution factors obtained from practice (EPA, 1989). The actual oxygen requirements of each zone for one of the four parallel basins were calculated by oxygen demand distribution factors and appear below.

Actual Oxygen Requirements for One Basin — 20 Year Design (kg/d) (lb/d = 2.205 × kg/d)

| Zone | Minimum Month | Average Nonnitrifying Month | Average Nitrifying Month | Maximum Month Nitrifying | Peak Day Nonnitrifying |
|-------|---------------|-----------------------------|--------------------------|--------------------------|------------------------|
| 1 | 239 | 329 | 523 | 616 | 702 |
| 2 | 135 | 205 | 398 | 470 | 459 |
| 3 | 31 | 80 | 177 | 228 | 218 |
| Total | 405 | 614 | 1098 | 1314 | 1379 |

Following the estimation of AORs for each condition, the standard oxygen transfer rates (SOTRs) for each of the zones are calculated. The actual oxygen requirements (AOR) are equated to the field transfer rates (OTR_f) since the OTR_f s must satisfy the corresponding AORs. Equation (2.53) may then be used to estimate the individual SOTR values for each zone and flow condition. For this calculation,

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it is necessary to identify all of the parameters in the equation. These values are identified as follows:

- alpha values for each zone and flow condition were determined as follows

Alpha Values for Zone and Flow Condition

| Zone | Minimum Month | Average Nonnitrifying Month | Average Nitrifying Month | Maximum Month Nitrifying | Peak Day Nonnitrifying |
|------|---------------|-----------------------------|--------------------------|--------------------------|------------------------|
| 1 | 0.30 | 0.20 | 0.25 | 0.25 | 0.20 |
| 2 | 0.50 | 0.30 | 0.40 | 0.40 | 0.30 |
| 3 | 0.80 | 0.60 | 0.70 | 0.70 | 0.60 |

- Theta is 1.024; the values of wastewater temperature for each flow condition are given above.
- Omega, the pressure correction, is estimated as P_b/P_s ; the elevation of the plant is 305 m (1007 ft); the value of P_b at 305 m is 98.6 kPa (14.3 psi). Omega = 0.97.
- Tau, the temperature correction, is estimated from DO surface saturation values at the given wastewater temperature and is given as $Tau = C_{s,i}^*/9.09$.
- Beta is estimated to be 0.98.
- The value of $C_{s,20}^* = 10.5 \text{ mg/L}$ from clean water testing of the selected aeration device at a submergence of 4.3 m (14 ft).
- The value of C_L for each zone is given above.

Using Equation (2.53), the following values of SOTR were calculated for each zone and flow condition.

Standard Oxygen Transfer Rates for Each Basin — 20 Year Design (kg/d) (lb/d = 2.205 × kg/d)

| Zone | Minimum Month | Average Nonnitrifying Month | Average Nitrifying Month | Maximum Month Nitrifying | Peak Day Nonnitrifying |
|-------|---------------|-----------------------------|--------------------------|--------------------------|------------------------|
| 1 | 1039 | 1937 | 2752 | 3241 | 3898 |
| 2 | 347 | 787 | 1284 | 1568 | 1702 |
| 3 | 50 | 156 | 327 | 430 | 396 |
| Total | 1436 | 2880 | 4364 | 5239 | 5996 |

At this point, the designer must determine the performance characteristics for the diffused air device that was selected for this facility. If the design is preliminary, this information may be obtained from estimates in the literature such as the values

provided in this text, the *Design Manual, Fine Pore Aeration Systems* (EPA, 1989), or the open literature. Final designs dictate that this information should be obtained from the manufacturer(s) of the device under consideration. For this example, a hypothetical set of performance data is used for a 23 cm (9 in) perforated membrane disc in a full floor grid configuration as given below.

**Clean Water Test Performance Data-Perforated Membrane Disc (23 cm)
Submergence — 4.3 m (14 ft)**

| AIRFLOW (m ³ /h) | AIRFLOW (scfm) | SOTE @ Density-7.4% | SOTE@ Density-9.9% | SOTE@ Density-12.4% | SOTE@ Density-18.5% |
|--------------------------------|-------------------|------------------------|-----------------------|------------------------|------------------------|
| 0.78 | 0.5 | 30 | 33 | 36 | 38 |
| 1.57 | 1.0 | 28 | 30 | 32 | 34 |
| 2.35 | 1.5 | 27 | 29 | 31 | 32 |
| 3.14 | 2.0 | 26.5 | 28 | 28.5 | 31 |
| 3.93 | 2.5 | 26.3 | 27 | 28 | 30.5 |

The following design steps will use (Equation 2.51) in conjunction with the data in the table above. It is an iterative process whereby a value of SOTE is selected based on an estimate of diffuser density and diffuser airflow rate. A total airflow rate, G_s , is then calculated from Equation (2.51) and, for the selected airflow rate per diffuser, a total number of diffusers are calculated. The actual diffuser density is calculated and compared with the estimated value. A series of iterations follows until airflow per diffuser, diffuser density and SOTE are appropriate. Then, a calculation is performed to determine the SOTR at minimum allowable diffuser airflow rate, and this value is compared with the minimum oxygen requirement to determine whether more oxygen is provided than is required at this lower level of airflow (resulting in wasted energy at minimum turndown). At this point, adjustments may be made in diffuser density and airflow rate per diffuser to provide a more efficient design. Finally, a check must be made to determine whether sufficient mixing will be provided at minimum airflow rate per diffuser.

Zone 1

The first zone will need to satisfy the highest oxygen demands. It will, therefore, require the highest diffuser densities and airflow rates per diffuser. This zone is one-third of the basin length, 13.2 m (43.3 ft) and is 7.0 m (23 ft) wide. For this area, an 18.5 percent diffuser density was selected with airflow per diffuser of 3.93 m³/h (2.5 scfm), providing an SOTE of 30.5 percent. Peak day will control the design.

$$G_s = \frac{3898(\text{kg/d})}{0.3 \times 24 \times 0.305 \frac{\text{kg/d}}{\text{m}^3/\text{h}}} = 0.139 \times 3898 / 0.305 = 1776 \text{ m}^3/\text{h} (1127 \text{ scfm})$$

Using Equation (2.51)

$$\text{Number of diffusers} = 1776 \text{ m}^3/\text{h} / 3.93 \text{ m}^3/\text{h-diffuser} = 452$$

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$$\text{Check density: Density} = \frac{452 \text{ diffusers} \times 0.038 \text{ m}^2/\text{diffuser}}{(7 \text{ m} \times 13.2 \text{ m})} = 0.187 \text{ or } 18.7$$

percent (vs. 18.5 percent selected). This figure is acceptable and conservative.

Check SOTR at minimum acceptable airflow/diffuser: [Minimum airflow = 0.78 m³/h/diffuser (0.5 scfm); SOTE = 38 percent]: SOTR = 452 diffusers × 0.78 m³/h/diffuser × 0.38/0.139 = 964 kg/d (2142 lb/d). This figure compares with 1039 kg/d (2291 lb/d) at minimum flow; thus, demand controls airflow rate, not minimum allowable airflow, and excessive energy will not be consumed at minimum turndown.

Check mixing: Select $G = 60 \text{ sec}^{-1}$, and minimum airflow rate is calculated at 1.52 m³/h/m² (0.09 scfm/ft²) by Equation (3.6). Minimum mixing airflow required will be 1.52 m³/h/m² × 7 m × 13.2 m = 140 m³/h (90 scfm). At minimum allowable airflow rate per diffuser, minimum airflow will be 0.78 m³/h × 452 diffusers = 353 m³/h (226 scfm). This rate exceeds minimum mixing requirement; therefore, mixing requirement does not control airflow rate, and sufficient mixing will occur at minimum turndown.

Zone 2

In Zone 2, the peak day SOTR requirements control the design. Several alternative diffuser density/airflow rate combinations are possible. Select a diffuser density of 12.4 percent and airflow rate of 3.14 m³/h (2.0 scfm), which would yield an SOTE of 28.5 percent. Using the same calculation procedure illustrated above, the following design information is obtained.

- (1) $G_s = 830 \text{ m}^3/\text{h}$ (488 scfm)
- (2) Number of diffusers = 264.
- (3) Calculated density = 10.9 percent; this figure is significantly lower than estimated (12.4 percent).

Try 9.9 percent at an airflow of 3.14 m³/h producing an SOTE of 28 percent.

- (4) New $G_s = 845 \text{ m}^3/\text{h}$ (536 scfm).
- (5) New number of diffusers = 269.
- (6) New density = 11 percent; this is a little better and conservative. Additional iterations will not be necessary.
- (7) Check SOTR at minimum acceptable airflow/diffuser: At allowable minimum airflow of 0.78 m³/h/diffuser, SOTE = 33 percent; SOTR = 498 kg/d (1105 lb/d) which compares with an oxygen demand (SOTR) of 347 kg/d (764 lb/d) at minimum flow. Since the allowable minimum airflow controls airflow to Zone 2 during minimum wastewater flow, the target DO will be exceeded during this period, and some energy will be wasted.
- (8) Check minimum mixing requirements. The required airflow for adequate mixing of Zone 2 would be 140 m³/h (90 scfm), the same as Zone 1 (step 5). At allowable minimum airflow per diffuser, the total airflow would be 210 m³/h (130 scfm). Therefore, mixing requirement does not control airflow rate in this zone.

Zone 3

The maximum month SOTR controls oxygen requirements in Zone 3. Estimating a diffuser density of 7.4 percent and airflow rate per diffuser of $1.57 \text{ m}^3/\text{N/h}$ (1.0 scfm), the SOTE would be 28 percent. The calculations follow.

- (1) $G_s = 213 \text{ m}^3/\text{N/h}$ (136 scfm)
- (2) Number of diffusers = 136.
- (3) Calculated density is 5.6 percent. This calculation compares with estimated value of 7.4 percent. By linear extrapolation, estimate a value of SOTE = 25.5 percent for a density of 5.6 percent and airflow of $1.57 \text{ m}^3/\text{N/h}$ per diffuser.
- (4) New $G_s = 226 \text{ m}^3/\text{N/h}$ (143 scfm).
- (5) New number of diffusers = 144.
- (6) New calculated density = 5.9 percent; this estimate is acceptable.
- (7) Check SOTR at minimum allowable airflow rate per diffuser. At allowable airflow of $0.78 \text{ m}^3/\text{N/h}/\text{diffuser}$ (0.5 scfm/diff), the estimated SOTE will be 28 percent by linear extrapolation; $\text{SOTR} = 226 \text{ kg/d}$ (504 lb/d) compared with an SOTR required at minimum flow of 50 kg/d (111 lb/d). As in Zone 2, the minimum allowable airflow rate per diffuser controls airflow in this zone during minimum wastewater flow conditions resulting in higher DO values and wasted energy.
- (8) Check minimum mixing requirements. The required airflow is again $140 \text{ m}^3/\text{N/h}$ (90 scfm) for adequate mixing of Zone 3, the same as Zones 1 and 2. At minimum allowable airflow rate per diffuser, the total airflow rate in this zone = $112 \text{ m}^3/\text{N/h}$ (72 scfm), which indicates that mixing will control airflow in Zone 3. The minimum airflow rate allowable due to mixing considerations would be $0.97 \text{ m}^3/\text{N/h}/\text{diffuser}$ (0.6 scfm/diffuser). Note that this exacerbates the already excessive oxygen transfer in this zone as calculated in (7) above.

Summary

Aeration rates were calculated for each flow condition and zone for the diffuser densities selected above. They are tabulated below.

Summary of Airflow Rates for Flow Condition and Zone — 20 Year Design Airflow — $\text{m}^3/\text{N/h}$ (scfm = $0.637 \times \text{m}^3/\text{N/h}$)

| Zone | Number of Diffusers | Minimum Month | Average Nonnitrifying Month | Average Nitrifying Month | Maximum Month Nitrifying | Peak Day Nonnitrifying |
|-------------|---------------------|---------------|-----------------------------|--------------------------|--------------------------|------------------------|
| 1 | 452 | 380 | 816 | 1195 | 1453 | 1776 |
| 2 | 269 | 210* | 353 | 615 | 765 | 845 |
| 3 | 144 | 140** | 140** | 165 | 226 | 204 |
| Basin Total | 865 | 730 | 1309 | 1975 | 2444 | 2825 |
| Syst. Total | 3460 | 2920 | 5236 | 7900 | 9776 | 11300 |

* Controlled by minimum allowable airflow rate/diffuser; ** mixing controlled.

Once the design and the system through the 20-year life with respect to the design year most other this results in to avoid mixing.

Zone 1 and operation calculations

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Several options place fewer diffusers. This leads to zero DO in Zones 2 and 3. Avoid mixing also helps to bring DO up. It is that operating at lower DO levels also help to bring DO up. As an alternative, operate at lower DO levels. stream as designed life will be longer. In the design, we will avoid mixing in the operation.

A second option would perhaps, Zone 3, have proper values advantage of lower loads to the plant.

Once the system is configured, it will furnish air at the desired DO level.

3. Estimating a
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| 5 | 845 |
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| 5 | 11300 |

Once these calculations are performed, the designer should review the system design and identify any drawbacks that may affect the construction or operation of the system. A calculation of the system capacity at start-up and one-half way through the design life is instructive assuming a linear increase in load over the 20-year life. At start-up, it appears that Zone 1 will not be significantly inefficient with respect to excess aeration capacity except during minimum month flow conditions (i.e., the airflow rate per diffuser will be greater than the minimum allowable for all flow conditions except minimum month). In Zone 2, the aeration system will need to be operated at minimum allowable airflow per diffuser during average, nonnitrifying periods and minimum month periods during the start-up years. Observation of the data in the table above indicates that Zone 3 is mixing limited in the design year for low flow and average winter months. It is also mixing limited for most other flow conditions early in the design period. As previously mentioned, this results in higher operating costs than would occur if all zones were operated to avoid mixing limitations.

Zone 1 has been designed for a diffuser density that may create construction and operational difficulties. These characteristics are described more fully in the calculations that follow.

Finally, it is normally desirable that the airflow rate per diffuser in each zone be about the same to minimize head loss and difficulties with airflow control that may lead to poor airflow distribution and premature fouling. For average flow-nitrification conditions, the airflow is 2.64, 2.29 and 1.15 m^3_N/h /diffuser (1.68, 1.46 and 0.73 scfm/diffuser) for Zones 1, 2, and 3 respectively.

Several options are available to address these concerns. One design option is to place fewer diffusers in Zone 1 without changing the allowable airflow rate per diffuser. This would allow greater spacing between diffusers but would result in low to zero DO in that zone, thereby passing system oxygen demand downstream to Zones 2 and 3. The design could be modified so that Zone 3 could be operated to avoid mixing limited conditions some, or all of the time. This modification would also help to balance unit airflow rates in the three zones. A drawback to this strategy is that operation at low DO in Zone 1 may cause sludge bulking some of the time. As an alternative to removing diffusers from Zone 1, this zone could be deliberately operated at low airflow rates, and therefore, low DO forcing a greater load downstream as described above. This strategy is tempting during the earlier years of design life when there is excess capacity in the system. During the later periods in the design, when oxygen demands increase and nitrification becomes more critical, the operation can revert to the original design airflows.

A second design option would be to operate the basins in a step-feed mode. This option would allow part of the influent load to be introduced into Zone 2 and, perhaps, Zone 3. If this option is selected, it will be necessary to reevaluate the proper values of alpha and AOR distribution in the zones. Step-feed offers an advantage of superior sludge management during qualitative or quantitative shock loads to the plant but may produce lower treatment efficiency during some periods.

Once the diffuser number and airflow rates are determined, the designer may configure the diffuser system. A full floor grid was selected. Assume that one drop-leg will furnish air to each of the three zones. Each zone has a floor area of $7 m \times 13.2 m$,

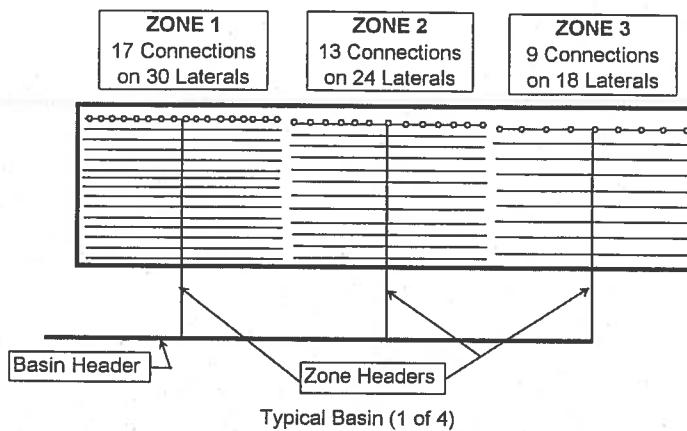


FIGURE 3.46 Design problem — diffuser system layout.

or 92.4 m^2 ($23 \text{ ft} \times 43.6 \text{ ft} = 996 \text{ ft}^2$). Designers often provide extra baseplates in each zone for contingency. The calculations for each zone follow.

Zone 1

There are 452 diffusers in Zone 1, or $0.20 \text{ m}^2/\text{diffuser}$ ($2.20 \text{ ft}^2/\text{diffuser}$). This would require a 0.45 m (1.5 ft) spacing, center-to-center. Dividing the tank width by this spacing results in $7 \text{ m}/0.45 = 15.5$, say, 15 laterals placed on each side of the drop-leg main header. Note that the equal spacing between laterals will be about 46 cm (18 in), which is the minimum desirable spacing between laterals containing 23 cm (9 in) disc diffusers. Typically, the designer will leave approximately a 60 cm (24 in) clearance between the end of the headers and the wall, approximately 30 cm (12 in) spacing at the end of the zone, and will allow about 60 cm clearance at the central main header. This would leave $13.2 \text{ m} - 0.6 \text{ m} - 0.3 \text{ m} - 0.6 \text{ m} = 11.7 \text{ m}$ for diffuser baseplates (about 38.3 ft). At a minimum spacing between discs of 33 cm (13 in) center-to-center, each lateral could accommodate $11.7 \text{ m}/0.33 \text{ m} = 35.5$, say 34 diffusers for a total of $15 \times 34 = 510$ diffusers, or a 13 percent contingency. Leave four baseplates empty per lateral, uniformly distributed along the longitudinal axis of the zone. See Figure 3.46 for the layout of this system.

Zone 2

There are 269 diffusers in Zone 2, or $0.34 \text{ m}^2/\text{diffuser}$ ($3.7 \text{ ft}^2/\text{diffuser}$) with a spacing of 0.58 m (1.9 ft) center-to-center. Use $7 \text{ m}/0.58 = 12$ laterals in this zone on each side of the main header. Each lateral should accommodate a minimum of $269/12 = 22$ diffusers. Adding a 20 percent contingency will place 26 baseplates on each lateral spaced at 45 cm (17.5 in) centers. Leave 4 baseplates empty per lateral.

Zone 3

By the same type of calculations, there will be nine laterals in Zone 3. Each lateral will contain 18 diffuser pods, of which, two will be blank, providing a contingency of about 12 percent in this zone.

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The next step of the design will be the selection and sizing of the blowers, followed by the final piping design, filter selection and control layout. An example of blower calculations is found in Chapter 4. Details of piping design and layout along with control system selection and design may be found in the EPA *Design Manual, Fine Pore Aeration Systems* (1989).

3.6 NOMENCLATURE

| | | |
|-------------------|------------------------------------|--|
| A_d | m^2 | total projected area of diffuser media |
| A_t | m^2 | total surface area of aeration basin |
| AE_f | kg/kWh, lb/hp-h | aeration efficiency under process conditions |
| C | mg/L | surfactant concentration |
| C_1 | | empirical coefficient |
| C_L | mg/L | bulk liquid phase oxygen concentration |
| C_{20}^* | mg/l | clean water oxygen saturation concentration at diffuser depth and 20°C |
| $C_{\infty 20}^*$ | mg/l | clean water oxygen saturation concentration at diffuser depth and 20°C |
| DWP | cm of water | dynamic wet pressure |
| d_B | cm | bubble diameter |
| E | W | power transferred to the fluid |
| F/M | $\text{lb BOD}_5/\text{d-lb MLSS}$ | food to microorganism ratio |
| G | s^{-1} | root mean square velocity gradient |
| G_s | $\text{m}_N^3/\text{h, scfm}$ | airflow rate at standard conditions |
| G_{sd} | $\text{m}_N^3/\text{h-diff}$ | airflow rate per diffuser at standard conditions |
| H | m | sidewater depth |
| H_s | m | diffuser submergence |
| K_L | cm/h | overall liquid film coefficient |
| $K_L \alpha$ | h^{-1} | oxygen transfer coefficient |
| $K_L \alpha_{20}$ | h^{-1} | clean water oxygen transfer coefficient at 20°C |
| m | | alpha factor for surfactant data |
| m | | empirical constant |
| n | | empirical coefficient |
| OTE | | oxygen transfer efficiency |
| OTE_f | -, % | oxygen transfer efficiency under process conditions |
| OTR_f | kg/h, lb/h | oxygen transfer rate under process conditions |
| P_1 | kPa, psia | absolute pressure at the surface |
| P_2 | kPa, psia | absolute pressure at the depth of injection |
| SAE | kg/kWh, lb/hp-h | standard aeration efficiency |
| SOTE | -, % | standard oxygen transfer efficiency |
| $SOTE_a$ | -, % | standard oxygen transfer efficiency at gas flow G_{sa} |
| $SOTE_b$ | -, % | standard oxygen transfer efficiency at gas flow G_{sb} |
| SOTR | kg/h, lb/h | standard oxygen transfer rate |

| | | |
|----------|--------------------|---|
| SRT | d | solids retention time |
| t | °C | temperature |
| V | m ³ | tank volume |
| W | W/m ³ | power dissipation |
| α | | wastewater correction factor for oxygen transfer coefficient |
| β | | wastewater correction factor for oxygen saturation |
| δ | | depth correction factor for oxygen saturation |
| μ | N-s/m ² | absolute viscosity |
| θ | | temperature correction factor for oxygen transfer coefficient |
| τ | | temperature correction factor for oxygen saturation |
| Ω | | pressure correction factor for oxygen saturation |

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Two safety systems are provided. Combination vacuum/pressure relief valves in the headspace of the first and last stages open automatically if excessively high or low pressures occur. A second system continuously monitors hydrocarbon concentrations in the first and last stages so an air purge can be initiated if concentrations become unacceptable.

6.2.1 GAS TRANSFER KINETICS

To properly design the aerators in the closed tank systems, the oxygen supply must be properly balanced with the oxygen demand in each reactor, similar to an air aeration system. The major difference between the two systems is that the oxygen partial pressure in the headspace of the HPO is not known as it is with an air system. It requires a mathematical model to predict this concentration. In its early development work, Union Carbide utilized such a model. Independent of this work, in 1973 a model was developed at Hydroscience, Inc. (presently Hydroqual) to evaluate the system for an industrial client (Mueller et al., 1973; 1978). This model utilized non-steady state equations that were rapidly solved numerically to obtain steady-state solutions. Subsequently, Clifft (1988; 1992) solved the non-steady state equations as true dynamic models and began to evaluate control strategies. Yuan et al. (1993) and Stenstrom et al. (1989) developed similar models to evaluate calibration requirements and to use in oxygen transfer compliance testing. More recently, Yin and Stenstrom (1996) have evaluated both feed forward and feed back control strategies. This section will present the basic principles involved in the models with the steady-state results.

Gas transfer occurs for at least four constituents when pure oxygen is introduced into an aeration tank as shown in Figure 6.7. Oxygen is transferred from the gas to the liquid phase. Nitrogen and inert gases such as argon, originally present in the liquid phase or produced in a prior denitrification reaction, are transferred to the gas phase. Carbon dioxide, produced by the biological reaction, is transferred to the gas phase. Since dry gas is introduced into the gas phase from the oxygen generation equipment, water vapor is transferred to the gas phase until it reaches the saturated vapor pressure.

Using the CSTR schematic in Figure 6.8, two mass balance equations are required for each parameter of concern, one for the liquid phase and one for the gas phase.

Liquid Phase:

$$V_{L,n} \frac{dC_{L,i,n}}{dt} = Q(C_{L,i,n-1} - C_{L,i,n}) + K_L a_{f,i,n} V_{L,n} (C_{\text{off},i,n}^* - C_{L,i,n}) + r_{v,i,n} V_{L,n} \quad (6.1)$$

Gas Phase:

$$V_{G,n} \frac{dC_{G,i,n}}{dt} = G_{n-1} C_{G,i,n-1} - G_n C_{G,i,n} - K_L a_{f,i,n} V_{L,n} (C_{\text{off},i,n}^* - C_{L,i,n}) \quad (6.2)$$

An additional equation defining the gas phase concentration as a function of partial pressure in the gas phase is as follows from Chapter 2.

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xxygen is introduced from the gas to nally present in the transferred to the gas transferred to the gas > oxygen generation reaches the saturated

equations are required for the gas phase.

$$\dots + r_{v,i,n} V_{L,n} \quad (6.1)$$

$$\dots - C_{L,i,n} \quad (6.2)$$

tion as a function of

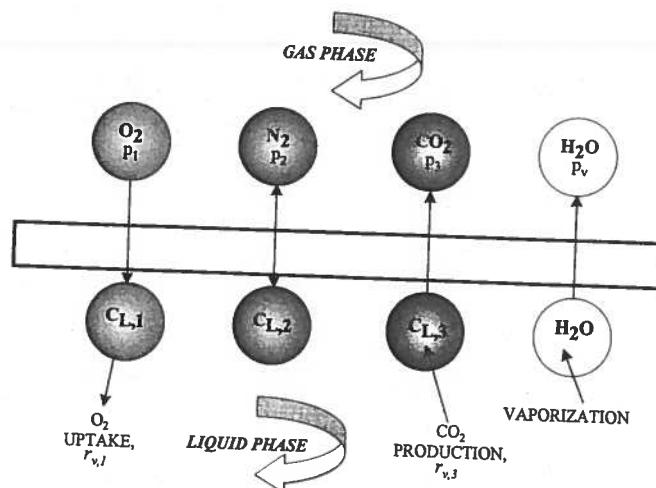


FIGURE 6.7 Gas transfer constituents in HPO system.

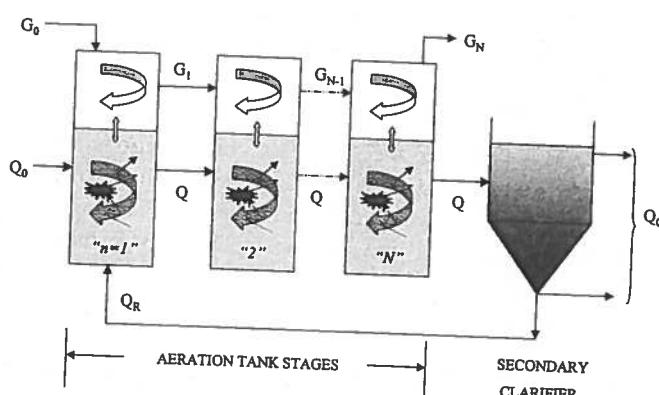


FIGURE 6.8 HPO completely mixed series tank reactors (CSTR) schematic.

$$C_{G,i,n} = \frac{M_i}{RT} p_{i,n} \quad (6.3)$$

The linkage between the two phases is provided by using the Henry's law relationship, corrected for field conditions as given in Chapter 2. Note that the pressure correction factor (Ω) is not included since the actual partial pressure as defined in Equation 6.3 is used to define the gas phase concentration.

Both Phases:

$$C_{\text{eff},i,n}^* = \frac{\tau \beta \delta C_{G,i,n}}{H_{20,i}} \quad (6.4)$$

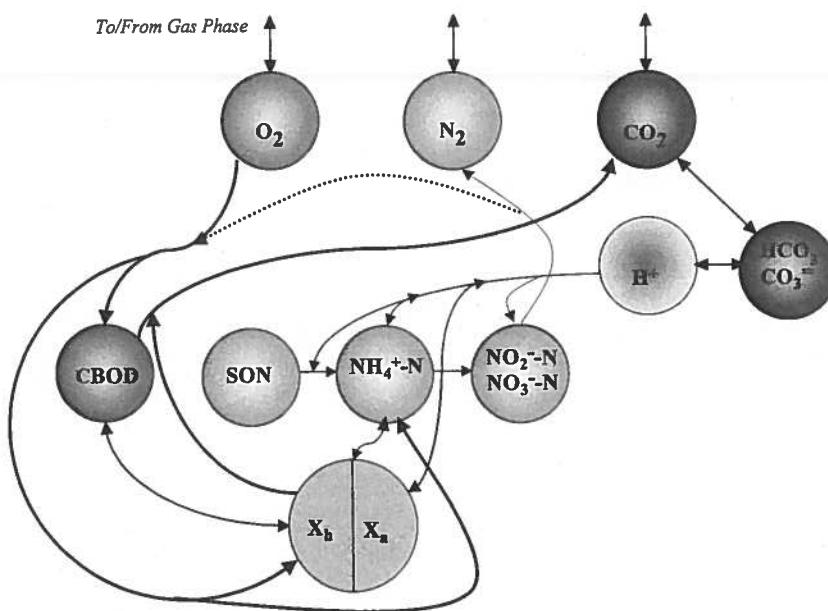


FIGURE 6.9 Biological and chemical reaction schematic in liquid phase.

Equations 6.3 and 6.4 provide the gas phase concentrations and the saturation values as a function of the headspace partial pressures in each stage. For each parameter of concern, the liquid phase concentration and gas phase partial pressure are unknown with Equations 6.1 and 6.2 available to solve them. However, an additional unknown always exists in the gas phase, i.e., the gas flow. A final equation, defining the total pressure in the gas phase, allows simultaneous solutions of the equation set for each stage.

$$\sum_{i=1}^l p_{i,n} = p_{T,n} - p_v \quad (6.5)$$

The equations using first order BOD removal kinetics and bacterial respiration for the reaction rate term, r_v , were originally solved numerically to a steady-state solution using CSMP. Later, when more complex nitrification and sulfite oxidation kinetics were utilized with bacterial growth, Famularo (1975) developed a solution technique using the steady state equations by stepping up the recycle stream in small increments.

A simplified schematic of the biological reactions occurring during carbon oxidation, CBOD removal, is shown in Figure 6.9. The buildup of CO_2 in the gas phase causes a significant reduction in the tank pH. The amount of CO_2 production is related to the oxygen consumption by the respiratory quotient, RQ , with the pH calculated from the first equilibrium constant for the CO_2 system.

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$$pH = pK_1 + \log \frac{[HCO_3^-]}{[CO_2]} \quad (6.6)$$

If the reactor is designed for nitrification, significant reductions in bicarbonate alkalinity will occur while denitrification will produce alkalinity. A further discussion of the alkalinity effects and a more complex evaluation of the carbonate equilibrium system are given in Mueller et al. (1978; 1980).

In Equations 6.1 and 6.2, the mass transfer coefficient for oxygen can be calculated from measured field calibrations. It can also be determined from aeration equipment specifications as a function of the power level as given in Chapter 2.

$$\left. \begin{aligned} K_L a_{20} &= \frac{SAE}{C_{\infty 20}^*} \left(\frac{WP}{V} \right) \\ C_{\infty 20}^* &= \delta C_{s20}^* = 89.09 \text{ mg/L} \end{aligned} \right\} \quad (6.7)$$

Note that the above SAE value is based on the manufacturer's specifications using air and not high purity oxygen. The mass transfer coefficients for the other gases can be corrected for diffusivity. This has some impact when large volatile organics are being stripped from solution. For the smaller inorganic gases, O₂, CO₂, and N₂, the diffusivity difference is small and has often been ignored. In laboratory experiments, Speece and Humenick (1973) have shown that CO₂ has the same K_L value as O₂ and that N₂ is 89 percent that of O₂. The field transfer coefficients are then determined, similar to the air aeration systems.

$$K_L a_{f,i,n} = \alpha_n \theta^{l-20} K_L a_{20,i,n} \quad (6.8)$$

(6.5)

6.2.2 APPLICATIONS OF STEADY-STATE KINETICS

Figure 6.10 shows the gas phase parameters for the Batavia Phase III data (Albertsson et al., 1970). The measured data for oxygen and gas flow are given with the solid lines representing calculated values from a model employing the above mechanisms (Mueller et al., 1973). Gas flow significantly decreases from the influent to the vent from the last stage in this three-stage reactor system. Oxygen partial pressure decreases successively from stage to stage as CO₂ and N₂ increase. Figure 6.11 provides the liquid phase concentrations along with the pH. CO₂ increases in successive stages due to its high solubility, yielding effluent concentrations >250 mg/L with a resulting pH of 6.3. In the parallel air system at Batavia, the pH remained near the raw wastewater pH of 7.1 due to continual CO₂ stripping to the atmosphere. Lower RQ values result at higher organic loading rates, probably due to incomplete oxidation of the organics.

return sludge. However, design of the aeration tank should provide an overflowing weir outlet so any *Nocardia* growth will not accumulate on the surface of the system.

6.3.2 I-SO™ DESIGN EXAMPLE

A proprietary computer spreadsheet is used by the manufacturer for design of the I-SO™ systems. The manufacturer was requested to provide a design for the conditions shown in Table 6.1. These conditions are similar to those used earlier for the fine pore system except the hydraulic detention time was reduced to 2 h from the prior 6 h, reflecting the greater transfer capabilities of the HPO system. This would require about 3200 mg/L MLSS, triple that required in the air system. Note that an α value of 0.5 is used in the design for comparison to the fine pore system. Manufacturer's tests in municipal wastewater have shown α to be above 0.8 (Storms, 2001) thus making this design conservative.

From preliminary designs using four tanks, each with three zones similar to the air system design, it became obvious that mixing controlled the design with much greater power utilization than required for oxygen transfer. Therefore, three aeration tanks operating in parallel, each completely mixed were chosen for the final design. The tanks were also circular to eliminate dead spaces where sludge settling might occur.

The results of this design yielded a 40 hp (29.8 kW) unit in each of the three tanks as shown in Table 6.2. At the design power level, the diameter of influence or the mixing diameter is significantly greater than the tank diameter, which should provide complete suspension of the solids. The design capacity of the 29.8 kW Oxygenator unit is 20 percent higher than that needed for the peak load and about 50 percent higher than that needed for the average load. Thus, the generating unit would be operated at significant turndown from full capacity. Peak hourly loads would require minimum liquid oxygen due to this available capacity.

The aeration efficiencies, $3.5 \text{ kgO}_2/\text{kWh}$ (wire) at peak to $2.8 \text{ kgO}_2/\text{kWh}$ (wire) at average monthly conditions, are somewhat lower than those reported from field units, 4.3 to $5.5 \text{ kgO}_2/\text{kWh}$ (wire). This may be due to the low α value of 0.5 used in the design example as mentioned previously. When the generation power in Table 6.3 is taken into account for the average load, the field aeration efficiency decreases to $1.24 \text{ kgO}_2/\text{kWh}$ (wire).

Figure 6.21 shows the fraction of the area covered by the floating hood varied from 8.6 to 23 percent of the total tank surface area as a function of aeration tank depth. A clarifier design was also conducted to get a sense of the relative size of the two units. Using a range of realistic overflow rates in Figure 6.22, the clarifier surface area is significantly greater than the aeration tanks, typical of HPO systems. Figure 6.23 gives a schematic layout of the plant using 9.1 m (30 ft) deep aeration tanks with two clarifiers at $24.4 \text{ m}^3/\text{m}^2/\text{d}$ (600 gpd/sf). Table 6.3 summarizes the monthly power requirements and total costs of the I-SO™ system including the generation costs using a single-bed vacuum pressure swing adsorption (VPSA) system and liquid oxygen (LO_x) costs to handle load variability. All equipment would be leased, the lease costs estimated as 73 percent of the total monthly costs. These are not bid values and may be lower under competitive bidding. The unit costs per volume treated for the above cost estimates are $\$0.039/\text{m}^3$ ($\$0.148/1000 \text{ gal}$).

TABLE 6.1
HPO Design Conditions

$Q = 5.3 \text{ MGD} = 0.232 \text{ m}^3/\text{s}$

| Tank Type | peak day | max mo | avg mo | avg no nit | min mo |
|-------------------------|--------------|-----------------|---------------------------------|-------------|-------------|
| Influent | 12800 | 7700 | BOD ₅ , lb/d 6600 | 6600 | 5500 |
| Influent | 5805 | 3492 | BOD ₅ , kg/d 2993 | 2993 | 2494 |
| CSTR | 12160 | 11588 | OTRF, lb/d 9685 | 5412 | 3575 |
| PLUG FLOW SYSTEM | | | | | |
| zone 1 | 6187 | 5430 | 4613 | 2904 | 2109 |
| 2 | 4053 | 4147 | 3513 | 1804 | 1191 |
| 3 | 1920 | 2010 | 1559 | 704 | 275 |
| CSTR | 5515 | 5255 | OTRF, kg/d 4392 | 2454 | 1621 |
| PLUG FLOW SYSTEM | | | | | |
| zone 1 | 2806 | 2463 | 2092 | 1317 | 956 |
| 2 | 1838 | 1881 | 1593 | 818 | 540 |
| 3 | 871 | 912 | 707 | 319 | 125 |
| All Tanks | 6252 | 3761 | MLSS, X, mg/L* 3223 | 3223 | 2686 |
| # tanks | 4 | maximum | | | |
| | 2 | minimum | | | |
| HRT, hr | 2 | | | | |
| Vol, m ³ | 1671 | | | | |
| SWD, m | 4.57 | minimum | | | |
| | ≥ 9.14 | maximum | | | |
| ELEV = | 1000 | ft = 305 m | | | |
| Pb = | 14.21 | psi = 97.95 kPa | | | |
| OMEGA | 0.97 | | | | |
| ALPHA | 0.5 | all zones | | | |
| BETA | 0.99 | | | | |
| SRT, day | 4 | | | | |
| DO in tanks | 4 | mg/L | | | |

* Assuming net sludge wastage (ΔM) = 0.45 * BOD₅ load.
For SRT = V*X/ΔM; X = SRT*ΔM/V.

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TABLE 6.2
I-SO™ Design

| | | | | | |
|-------|-------------------------------------|---|-----------------------------|----------------------------|--|
| in mo | No. Aeration Tanks | 3 | | | |
| 5500 | Vol./tank | 557.1 m ³ | 19672 ft ³ | 0.147 MG | |
| | Depth | 9.1 m | 30 ft | | |
| | Diameter | 8.810 m | 28.9 ft | | |
| 2494 | No. I-SO™ Units | 3 | 1/tank | | |
| | Impeller Size | 0.610 m | 24 in | | |
| | Motor Power | 29.83 kW | 40 hp | | |
| | Actual Power | 21.85 kW | 29.3 hp | | |
| 3575 | Hood Dia | 3.657 m | 12 ft | | |
| | Hood As | 10.51 m ² | 113.1 ft ² | | |
| | Power Level | 39.22 W/m ³ | 199.1 hp/MG | | |
| 2109 | Mixing Dia. | 19.45 m | 63.8 ft | | |
| 1191 | Oxygen Transfer Capabilities | | | | |
| 275 | I-SO™ Capacity | 92.1 kgO ₂ /h | 203 lbO ₂ /h | | |
| | Peak Req'd. | 76.6 kgO ₂ /h | 169 lbO ₂ /h | | |
| | Avg. Req'd. | 61.0 kgO ₂ /h | 135 lbO ₂ /h | | |
| 1621 | Peak AEf | 3.51 kgO ₂ /kWh | 5.76 lbO ₂ /hp-h | mixer power only | |
| | Avg AEf | 2.79 kgO ₂ /kWh | 4.59 lbO ₂ /hp-h | mixer power only | |
| 956 | Avg AEf | 1.24 kgO ₂ /kWh | 2.13 lbO ₂ /hp-h | mixer and generation power | |
| 5 | No. Secondary Clarifiers | 2 | | | |
| 125 | Overflow Rate | 24.44 m ³ /m ² /d | 600 gpd/sf | | |
| | Diameter | 22.9 m | 75 ft | | |

TABLE 6.3
I-SO™ Design Power Requirements and Costs (April 1998)

| Item | Number | Type | Unit Cost | Monthly Cost |
|---|--------|--------------|-----------|--------------|
| I-SO™ Units, 40 hp, 24" | 3 | Lease/mo* | \$1,500 | \$4,500 |
| VPSA, single-bed (2% downtime) | 1 | Lease/mo* | \$13,000 | \$13,000 |
| LOx, 1000 cf (Supplemental + Backup) | 70 | Purchase/mo | \$5 | \$350 |
| Power for Generation, 1000 kWh (Avg O ₂ Demand incl. turn down) | 60 | Purchase/mo | \$50 | \$3,000 |
| Power for Aeration, 1000 kWh | 47.8 | Purchase/mo | \$50 | \$2,391 |
| Total Power and Lease Costs | | | | \$23,241 |
| Site Preparation, i = 8%, n = 20 yr | | Construction | \$75,000 | \$627 |
| Total Monthly Cost | | | | \$23,868 |

* Conservative estimate, not actual bid value.

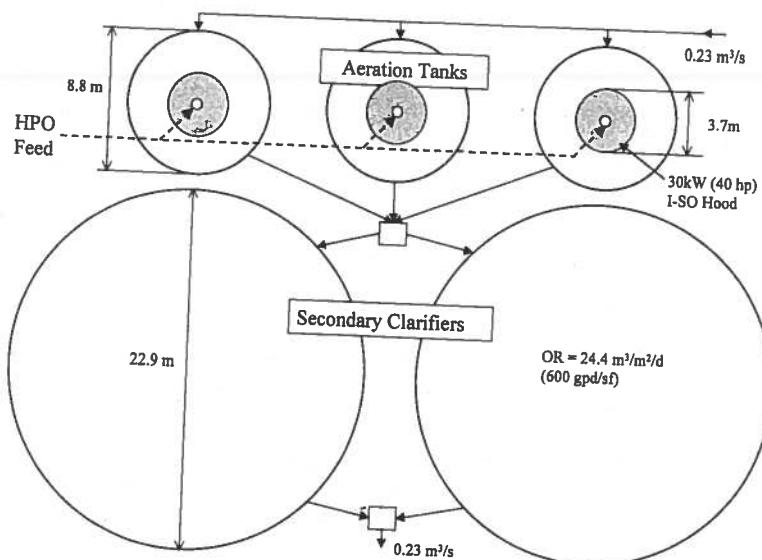


FIGURE 6.23 Schematic of in-situ oxygenator layout for 9.1 m deep aeration tanks.

6.4 NOMENCLATURE

| | | |
|-------------------|---|--|
| AE_f | kg/kWh, lb/hp-h | aeration efficiency under process conditions |
| C_G | mg/L | bulk gas phase concentration |
| C_L | mg/L | bulk liquid phase concentration |
| C_{s20}^* | mg/L | surface saturation concentration at 20°C, 9.09 mg/L |
| $C_{\infty 20}^*$ | mg/l | clean water oxygen saturation concentration at diffuser depth and 20°C |
| C_{wf}^* | mg/l | process water oxygen saturation concentration |
| F/M | lb BOD ₅ /d-lb MLSS | food to microorganism ratio |
| G | m ³ /h | gas flow rate |
| G_{90} | m ³ /h | gas flow rate that obtains 90% oxygen utilization |
| H_{20} | (mg/L) _{gas} /(mg/L) _{liquid} | Henry's constant at 20°C, 29.8 from Table 2.1 |
| HRT | h | hydraulic detention time |
| $K_L a_f$ | h ⁻¹ | oxygen transfer coefficient under process conditions |
| $K_L a_{20}$ | h ⁻¹ | clean water oxygen transfer coefficient at 20°C |
| LO _x | | liquid oxygen |
| M | g/mole | molecular weight |
| N_o | lb/hp-h | standard aeration efficiency = SAE |
| OR | m ³ /m ² -d, gpd/sf | clarifier overflow rate |
| p | atm | partial pressure of constituent in gas phase |
| P_b | kPa, psia | barometric pressure |

ample using three
a 3.66 m (12 ft)

30

or I-SO™ design

| | | | |
|-------------------|---|---|----------------|
| p _t | atm | total pressure | Budd, W |
| p _v | atm | vapor pressure | S |
| pK ₁ | | first equilibrium constant for CO ₂ system, 6.35 at 25°C | Cheng, A |
| Q | m ³ /h | liquid flow rate | V |
| R | m ³ -atm/gmole-K | universal gas constant (8.205*10 ⁻⁵ m ³ -atm/gmole-K) | I |
| r _v | mg/L-h | reaction rate | Cliffit, R. |
| RQ | mole CO ₂ /mole O ₂ | respiratory quotient, CO ₂ production/O ₂ utilization | E |
| SAE | kg/kWh, lb/hp-h | standard aeration efficiency | Cliffit, R. |
| SWD | m | sidewater depth | E |
| T | °K | absolute temperature | Famularo, |
| t | °C | temperature in aeration basin | In |
| t | h | time | Fullerton, |
| V | m ³ | aeration tank volume | O |
| V _G | m ³ | gas phase volume | W |
| V _L | m ³ | liquid phase volume | Geselbrac |
| VPSA | | vacuum pressure swing adsorption system | an |
| WP | kW, hp | wire power | Er |
| X | mg/L | mixed liquor suspended solids concentration, MLSS | Gilligan, T |
| α | | wastewater correction factor for oxygen transfer coefficient | Gilligan, T |
| β | | wastewater correction factor for oxygen saturation | of |
| δ | | depth correction factor for oxygen saturation | Krugerwor |
| ΔM | kg/d, lb/d | net sludge production rate | Mueller, J. |
| θ | | temperature correction factor for oxygen transfer coefficient | Eq |
| τ | | temperature correction factor for oxygen saturation | wa |
| Ω | | pressure correction factor for oxygen saturation | Mueller, J. |
| <i>subscripts</i> | | | |
| i | | constituent | Coi |
| n | | reactor number | Mueller, J. |
| | | | Oxy |
| | | | 99% |
| | | | Okun, D. A. |
| | | | Wo |
| | | | Okun, D. A. |
| | | | Sew |
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| | | | Detro |
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| | | | 600/S |
| | | | Robbins, M. I |
| | | | Waste |
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