

CHAPTER 10

AERATION AND STRIPPING

GAS TRANSFER

The process of aeration is used for transferring oxygen to waste water. Stripping is for removing volatile gases such as hydrogen sulphide, hydrogen cyanide and ammonia from waste water. Both aeration and stripping involve a gas-liquid mass-transfer process in which the driving force is created by a departure from equilibrium. In other words, the driving force in the gas phase is a partial pressure gradient and, in the liquid phase, is a concentration gradient.

The transfer of a gas can be treated as a four-step process. The first step involves passage of the gas from the vapour phase to the gas-liquid interface. The gas must then pass through a gas film on the vapour side of the interface. The third step is the passage of the gas through a liquid film on the liquid side of the interface. In the final step the gas must be dispersed throughout the bulk of the liquid. General conditions are such that one of the steps is rate-limiting and the overall gas-transfer rate can be calculated on the basis of this step. The remaining steps are insignificant in the overall process.

In stagnant conditions diffusion of the gas through the bulk solution is generally the slowest step and an expression for molecular diffusion can be used to predict the transfer rate.

The diffusion can be calculated by means of Fick's Law:

$$N = -D \cdot A \cdot \frac{dc}{dy} \quad (10.1)$$

where

N = mass transfer per unit time

A = the cross-sectional area across which diffusion occurs

$\frac{dc}{dy}$ = the concentration gradient perpendicular to the cross-sectional area, A

D = diffusion coefficient

If, however, the solution is sufficiently agitated either by natural turbulence or by mechanical mixing, the rate of transfer through the gas-liquid interface becomes the controlling factor. For sparingly soluble gases such as oxygen and carbon dioxide, the resistance of the liquid film controls the rate of gas transfer, while for highly soluble gases such as ammonia, the transfer rate is controlled by the resistance of the gas phase.

Gas solubility

The equilibrium concentration of a gas in contact with a liquid can be calculated by Henry's Law:

$$C_{eq} = \frac{1}{H} \cdot p \quad (10.2)$$

where

C_{eq} = the equilibrium concentration of the gas in solution

H = Henry's Constant

p = the partial pressure of the gas in the gas phase

Table 10.1 gives the equilibrium concentration of oxygen in contact with distilled water in the temperature range of 0-30°C.

Table 10.2 gives Henry's constant as a function of temperature for several important gases. Henry's Constant is roughly proportional to the temperature; i.e. with increased temperature the solubility of a gas decreases. At 10°C the amount of oxygen dissolved in distilled water is approximately 11.3 mg/l, whereas at 30°C the maximum solubility is about 7.44 mg/l. As can be seen the solubility changes significantly with the temperature.

Henry's Constant is also influenced by the presence of dissolved solids. The combined effects of dissolved solids and temperature on the solubility of oxygen in water is expressed by the following equation (Gameson and Robertson, 1955):

$$C_{eq} = \frac{475 - 2.65 \cdot C_{ds}}{33.5 + (T - 273)} \quad (10.3)$$

where

C_{ds} = the concentration of total dissolved solids expressed in g/l

T = the absolute temperature expressed on °K

TABLE 10.1

Dissolved oxygen saturation values in distilled water

Temperature (°C)	Dissolved oxygen (mg/l)									
	0.0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9
0	14.65	14.61	14.57	14.53	14.49	14.45	14.41	14.37	14.33	14.29
1	14.25	14.21	14.17	14.13	14.09	14.05	14.02	13.98	13.94	13.90
2	13.86	13.82	13.79	13.75	13.71	13.68	13.64	13.60	13.58	13.53
3	13.49	13.46	13.42	13.38	13.35	13.31	13.28	13.24	13.20	13.17
4	13.13	13.10	13.06	13.03	13.00	12.96	12.93	12.89	12.86	12.82
5	12.79	12.76	12.72	12.69	12.66	12.62	12.59	12.56	12.53	12.49
6	12.46	12.43	12.40	12.36	12.33	12.30	12.27	12.24	12.21	12.18
7	12.14	12.11	12.08	12.05	12.02	11.99	11.96	11.93	11.90	11.87
8	11.84	11.81	11.78	11.75	11.72	11.70	11.67	11.64	11.61	11.58
9	11.55	11.52	11.49	11.47	11.44	11.41	11.38	11.36	11.33	11.30
10	11.27	11.24	11.22	11.19	11.16	11.14	11.11	11.08	11.06	11.03
11	11.00	10.98	10.95	10.93	10.90	10.87	10.85	10.82	10.80	10.77
12	10.75	10.72	10.70	10.67	10.65	10.62	10.60	10.57	10.55	10.52
13	10.50	10.48	10.45	10.43	10.40	10.38	10.36	10.33	10.31	10.33
14	10.26	10.24	10.22	10.19	10.17	10.15	10.12	10.10	10.08	10.06
15	10.03	10.01	9.99	9.97	9.96	9.92	9.90	9.88	9.86	9.84
16	9.82	9.79	9.77	9.75	9.73	9.71	9.69	9.67	9.65	9.63
17	9.61	9.58	9.56	9.54	9.52	9.50	9.48	9.46	9.44	9.42
18	9.40	9.38	9.36	9.34	9.32	9.30	9.29	9.27	9.25	9.23
19	9.21	9.19	9.17	9.15	9.13	9.12	9.10	9.08	9.06	9.04
20	9.02	9.00	8.98	8.97	8.95	8.93	8.91	8.90	8.88	8.86
21	8.84	8.82	8.81	8.79	8.77	8.75	8.74	8.72	8.70	8.68
22	8.67	8.65	8.63	8.62	8.60	8.58	8.56	8.55	8.53	8.52
23	8.50	8.48	8.46	8.45	8.43	8.42	8.40	8.38	8.37	8.36
24	8.33	8.32	8.30	8.29	8.27	8.25	8.24	8.22	8.21	8.19
25	8.18	8.16	8.14	8.13	8.11	8.10	8.08	8.07	8.06	8.04
26	8.02	8.01	7.99	7.98	7.96	7.95	7.93	7.92	7.90	7.89
27	7.87	7.86	7.84	7.83	7.81	7.80	7.78	7.77	7.75	7.74
28	7.72	7.71	7.69	7.68	7.66	7.65	7.64	7.62	7.61	7.59
29	7.58	7.56	7.55	7.54	7.52	7.51	7.49	7.48	7.47	7.45
30	7.44	7.42	7.41	7.40	7.38	7.37	7.35	7.34	7.32	7.31

It must be emphasized that this equation is developed under the conditions that the pressure is 760 mm of mercury and that clean water is in contact with wet air. The solubility of oxygen in waste water is less than in pure water. Roughly, the solubility in municipal waste water is about 95% that in pure water; the solubility of oxygen in various types of industrial waste water may be even less.

TABLE 10.2

Henry's Constant ($\times 10^{-3}$) (Atm $^{-1}$)

Gas	Temperature ($^{\circ}$ C)						
	0	5	10	15	20	25	30
Acetylene	0.72	0.84	0.96	1.08	1.21	1.33	1.46
Air (atm)	0.43	0.49	0.55	0.61	0.66	0.72	0.77
Carbon dioxide	0.73	0.88	1.04	1.22	1.42	1.64	1.86
Carbon monoxide	0.35	0.40	0.44	0.49	0.54	0.58	0.62
Hydrogen	0.58	0.61	0.64	0.66	0.68	0.70	0.73
Ethane	0.13	0.16	0.19	0.23	0.26	0.30	0.34
Hydrogen sulphide	26.80	31.50	36.70	42.30	48.30	54.50	60.90
Methane	0.22	0.26	0.30	0.34	0.38	0.41	0.45
Nitrous oxide	0.17	0.19	0.22	0.24	0.26	0.29	0.30
Nitrogen	0.53	0.60	0.67	0.74	0.80	0.87	0.92
Nitric oxide	-	1.17	1.41	1.66	1.98	2.25	2.59
Oxygen	0.25	0.29	0.33	0.36	0.40	0.44	0.48

In aeration tanks the gas is generally introduced into the water at a depth of 3-5 m below the surface. In this case the gas solubility in the tank is influenced by the fact that the partial pressure of the gas in the bubbles decreases as they rise to the surface of the solution. It is therefore necessary to use a mean saturation value (C_{sm}) for the oxygen introduced at the bottom of an aeration tank (Oldshue, 1956):

$$C_{sm} = C_{eq} \left(\frac{P}{2} + \frac{O_{ta}}{42} \right) \quad (10.4)$$

where

P = absolute pressure in atm

O_{ta} = percent concentration of oxygen in the air leaving the aeration tank

In this context it must be underlined that Henry's Law is an ideal law and gives only approximate values. It is preferable to use solubility data if these are available.

Mass transfer

Lewis and Whitman (1924) developed equations for the transfer rate controlled by the gas-film resistance as well as for the transfer rate controlled by the liquid-film resistance:

$$N = K_L \cdot A(C_{eq} - C) = K_G \cdot A(p - p_{eq}) \quad (10.5)$$

where

N = mass transfer per unit time

A = area of cross-section

C_{eq} = concentration at equilibrium (saturation)

p = partial pressure in the gas phase

p_{eq} = partial pressure at the interface

K_L = liquid-film defined as D_L/Y_L

K_G = gas-film defined as D_G/Y_G

D_L = diffusion coefficient in the liquid

D_G = diffusion coefficient in the gas

See also Fig. 10.1.

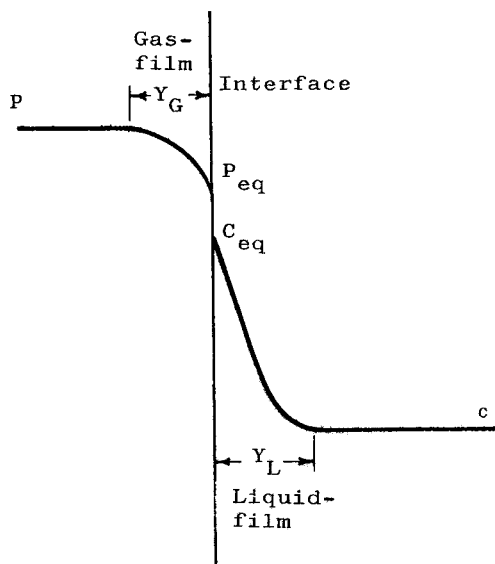


Fig. 10.1. Schematic representation of interfacial mass transfer.

The liquid-film-controlled process can be expressed in concentration units by dividing by the volume, V :

$$\frac{1}{V}N = \frac{dc}{dt} = K_L \cdot \frac{A}{V}(C_{eq} - c) \quad (10.6)$$

$K_{L,a} = K_L \cdot \frac{A}{V}$ is termed the overall film coefficient.

The transfer coefficient, K_L , is affected by a number of variables. In general, the liquid-film coefficient increases with increasing temperature according to

$$K_L(t) = K_{L,20^\circ} \cdot 1.028^{t-20} \quad (10.7)$$

t = temperature ($^\circ\text{C}$)

For $K_{L,a}$ in a bubble aeration system, the equation becomes

$$K_{L,a}(t) = K_{L,a,20^\circ\text{C}} \cdot 1.02^{t-20} \quad (10.8)$$

The presence of surface-active agents in the waste water has a significant effect on K_L and $\frac{A}{V}$. A decrease in surface tension will decrease the size of the bubbles generated, which will increase $\frac{A}{V}$. In some instances the increase in $\frac{A}{V}$ will exceed the decrease in K_L , with the overall effect that the transfer rate increases. Generally, $K_{L,a}$ decreases with increasing concentration of impurities in water. Fig. 10.2 shows a characteristic change in the coefficient α , as a function of BOD_5 of water.

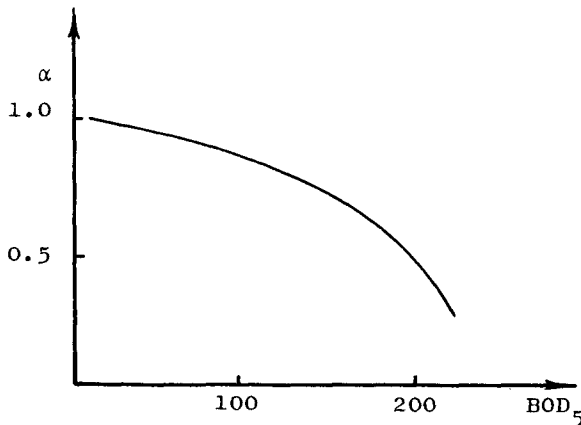


Fig. 10.2. A typical BOD_5/α relationship.

The coefficient, α , is defined as the ratio of $K_{L,a}$ for waste water to that for distilled water.

Generally $K_{L,a}$ varies with the depth (H) of the bubble diffusion system according to the equation:

$$\frac{(K_{L,a})_1}{(K_{L,a})_2} = \left(\frac{H_1}{H_2} \right)^n \quad (10.9)$$

The exponential, n , has a value close to 0.7 for most systems.

The overall transfer coefficient will increase with the degree of turbulent mixing.

In natural streams $\frac{A}{V}$ is equal to $\frac{1}{H}$, where H is the depth of the stream. O'Connor (1958) has found the following relationship:

$$K_L = \left(\frac{D_L \cdot U}{H} \right)^{1/2} \quad (10.10)$$

where U = the average velocity of the stream.

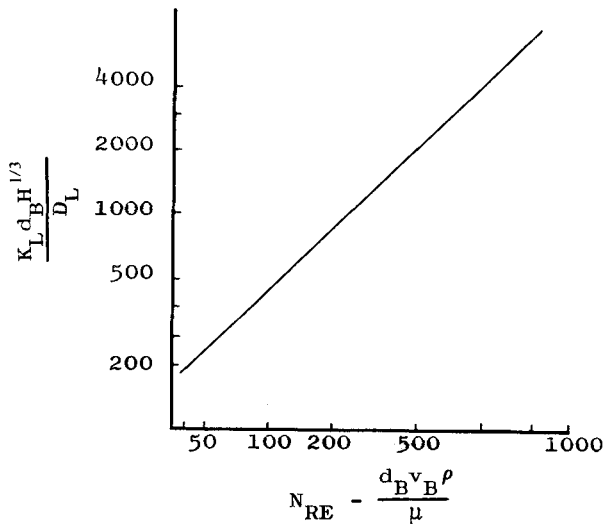


Fig. 10.3. Correlation of bubble-aeration data.

Fig. 10.3 shows the correlation of bubble aeration. v_B is the velocity of the bubbles. From this correlation it is possible to find K_L (Eckenfelder, 1959).

The diameter, d_B , of the bubbles can be expressed in terms of G^n , where G is the gas flow. This will lead to the following equation, when a mathematical expression for Fig. 10.3 is used:

$$K_{L,a} = \frac{K^1 \cdot H^{2/3} \cdot G^{(1-n)}}{V} \quad (10.11)$$

where K^1 is a constant.

For a wide range of commercial diffusion devices the value $(1 - n)$ has been found to vary between 0.8 and 1.0. It is less than unity for small, orificed diffusion units. As shown, the value of the depth exponential is 0.67. However, it has been found to be 0.71 - 0.77 for plate diffusers (King, 1955), 0.78 for a four-nozzle sparger placed in the centre line of a cylindrical tank (Eckenfelder, 1959), 0.45 for diffusion tubes along the side wall of a wide ring in a regular tank and 0.65 for impingement-type units (Dreier, 1956).

The aeration equipment commonly used is air diffusion units, turbine aeration equipment or surface aeration units.

The diffused-aeration equipment can either be units producing small bubbles from a porous medium (tubes or plates constructed of porcelain or carborundum tightly wrapped in, for example, nylon or polyether sponge), or units using a large orifice to produce the air bubbles. The performance of all air diffusion units can be calculated using the following equation:

$$K_{L,a} = \frac{K^1 \cdot G^n \cdot H^m (C_{eq} - C) \cdot 1.02^{(t-20)} \cdot \alpha \cdot \beta}{W^p \cdot V} \quad (10.12)$$

where $K_{L,a}$, K^1 , V , G , H , t , C_{eq} , C and α have been defined previously, W = the width of the aeration tank, n , m and p = exponentials characteristic of the aeration device, and β = a correction coefficient with a value close to 1.0.

The turbine aeration system disperses compressed air by the pumping action of a rotating propeller (see Fig. 10.4). The unit can be defined by the relationship:

$$\frac{K_{L,a}}{V} = K^1 \cdot I^x \cdot G^n \cdot D^y \cdot (C_{eq} - C) \cdot 1.02^{(t-20)} \quad (10.13)$$

where

I = the propeller speed (m/s)

D = the propeller diameter

x and y = characteristic exponentials

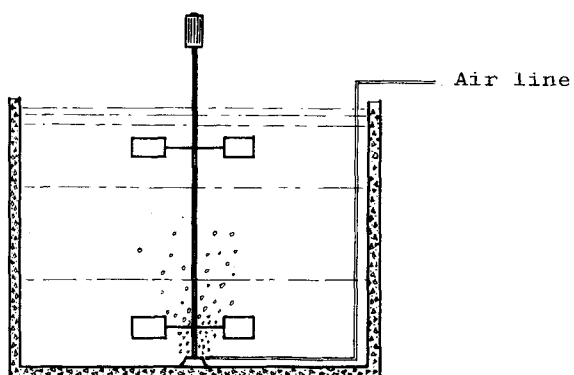


Fig. 10.4. Turbine aerator

Surface aeration units are of two types. One type employs a draft tube, the other a surface propeller, but in both types the gas transfers occurs by the surface exposure of a large volume of liquid (see Fig. 10.5).

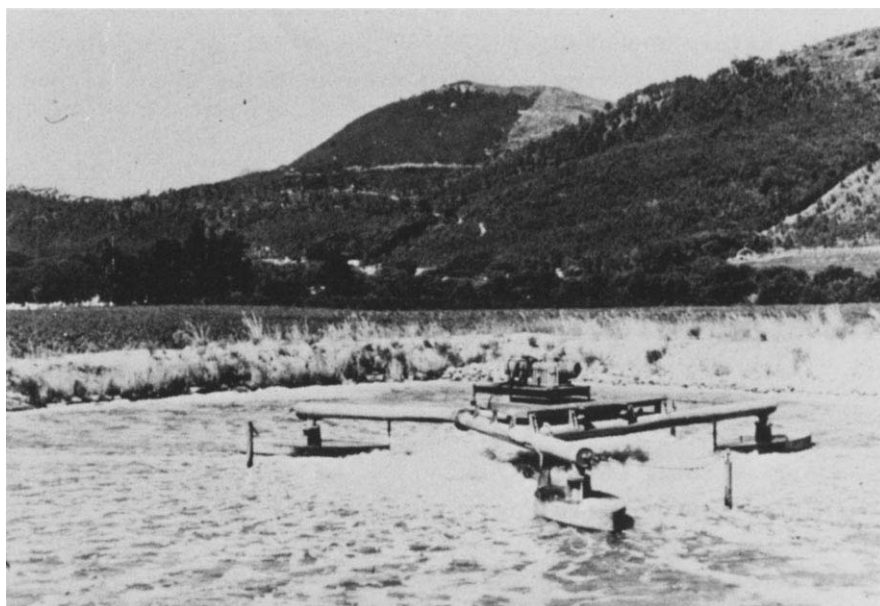


Fig. 10.5. Surface impeller.

Often it is possible to use the following equation for this equipment:

$$N_t = N_0 \frac{C_{eq} - C}{C_{eq,20}} \cdot 1.028^{t-20} \cdot \alpha \quad (10.14)$$

where

N_t = kg O_2 /hp-h at $t^\circ C$ and N_0 = kg O_2 /hp-h at $20^\circ C$ zero dissolved oxygen. The other symbols are defined above. N_0 is characteristic of the aeration unit and the speed (rpm). N_0 is in the order of 2 kg/hp-h for most applied units and decreases as the volume of liquid increases.

Although several equations have been developed to estimate the gas-transfer efficiency of aeration devices, it seems necessary in most cases actually to measure the oxygen-transfer efficiency. This measurement involves the chemical removal of dissolved oxygen by the addition of sodium sulphite, using cobalt as a catalyst. The oxygen concentration is then measured during aeration under specified conditions and the overall transfer coefficient can be calculated on the basis of equation (10.6).

Design of an absorption or stripping tower

Fig. 10.6 shows the principle of a countercurrent tower. The tower may be either a packed or a spray tower filled with bubble-cap trays, or of any internal construction to bring about a good gas-liquid contact.

The following relationships are valid:

$$Y = \frac{(y)}{1 - y} = \frac{p}{P_t - p} \quad (10.15)$$

and

$$G_s = G(1 - y) = \frac{G}{1 + Y} \quad (10.16)$$

where

G = gas stream total moles/ h/m^2

y = mole fraction of diffusing solute

p = partial pressure

Y = mole ratio of diffusing solute

G_s = moles/ h/m^2 of nondiffusing, essentially insoluble gas

P_t = total pressure

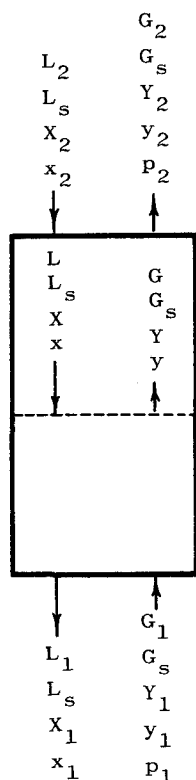


Fig. 10.6. Principles of countercurrent tower.

Similarly, the following equation is valid for the liquid stream:

$$X = \frac{x}{1 - x} \quad (10.17)$$

$$L_s = L(1 - x) = \frac{L}{1 + X} \quad (10.18)$$

where

L = liquid stream moles/h/m²

x = mole fraction of soluble gas

X = mole ratio of soluble gas

L_s = moles/h/m² of non-volatile solvent

Since the solvent gas and solvent liquid are essentially unchanged in quantity as they pass through the tower, it is convenient to express the material balance in terms of these.

The balance in the lower part of the tower (see Fig. 10.6) can be expressed by

$$G_s(Y_1 - Y) = L_s(X_1 - X) \quad (10.19)$$

This is the equation of a straight line, the so-called operating line, which has a slope of $\frac{L_s}{G_s}$ and passes through (X_1, Y_1) . The operating line also passes through the point (X_2, Y_2) .

In Fig. 10.7 the operating line is plotted together with the equilibrium solubility curve, plotted in terms of the mole ratio.

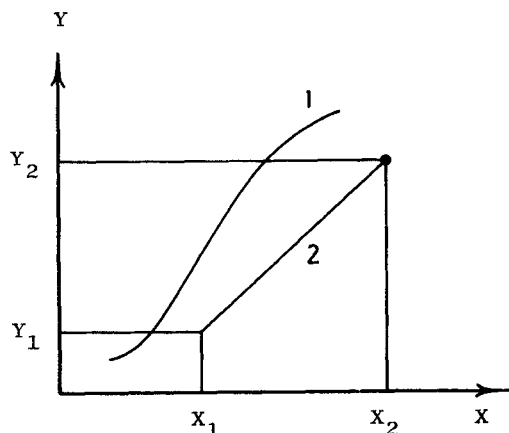


Fig. 10.7. Equilibrium curve (1) and operating line (2) for a stripping process.

For an absorption tower the operating line always lies above the equilibrium solubility curve while, for a stripping tower, the operating line is always below the equilibrium solubility curve (see Fig. 10.7).

If we consider a packed or spray tower of unit area cross-section, it is convenient to describe the interfacial surface between the gas and liquid as a function of the dispersion of the liquid in the thin film over the packing. The following equation is valid:

$$dS = a dZ \quad (10.20)$$

where

S = area of the interfacial expressed as m^2/m^2 tower cross-section

a = m^2 interfacial surface/ m^3 packed volume

Z = the height, (m), of the tower

The amount of solid in the gas passing the differential section of the tower is $G \cdot y$ mole/h/m², and the rate of mass transfer to the liquid, $d(G \cdot y)$. This can be related to the mass transfer coefficient as follows (compare with equation 10.5):

$$d(G \cdot y) = K_y \cdot a(y - y_{eq})dZ \quad (10.21)$$

where K_y = the overall transfer coefficient (see also equation 10.31).

Both G and Y vary from one part of the tower to another, but G_s does not. Therefore it is more convenient to use G_s in these expressions:

$$d(G \cdot y) = G_s \cdot d\left(\frac{Y}{1 - y}\right) = G_s \frac{dy}{(1 - y)^2} = \frac{Gdy}{1 - y} \quad (10.22)$$

The mass-transfer coefficient for diffusion of one component through a second (the solvent) includes a term involving the average concentration, Y_m , of the non-diffusing gas along the path of the diffusion. If the concentration of solute varies considerably from one end of the tower to another, the quantity $K_G \cdot a(1 - y)_m$ will be much more constant than $K_G \cdot a$ alone. Therefore equation (10.21) will be transformed to

$$d(G \cdot y) = \frac{Gdy}{1 - y} = \frac{(K_G a (1 - y)_m) (y - y_{eq})dZ}{(1 - y)_m} \quad (10.23)$$

or

$$\frac{(1 - y)_m}{(1 - y)} \cdot \frac{dy}{y - y_{eq}} = \frac{K_G \cdot a (1 - y)_m \cdot dZ}{G} \quad (10.24)$$

Equation (10.24) may be integrated to obtain, Z , in terms of $K_G \cdot a$, but for many situations the first term on the left-hand side is very close to unity. Since the number of transferred units N_{tog} is defined as

$$N_{tog} = \int_{y_2}^{y_1} \frac{(1 - y)_m \cdot dy}{(1 - y) (y - y_{eq})} \quad (10.25)$$

then

$$Z = N_{tog} \cdot H_{tog} \quad (10.26)$$

N_{tog} can as shown be related to the height of the packing and the height per transfer unit, termed H_{tog} . The height per transfer unit is an experimental quantity, but it is more convenient to use it than $K_G \cdot a$ and other mass-transfer coefficients in the design of towers. H_{tog} has the dimension of length. The subscript, og, is used, as seen in the terms N_{tog} and H_{tog} to indicate that these terms are based on an overall driving force $y - y_{\text{eq}}$ within the gas phase. These terms therefore represent the vertical distance between the operating line and equilibrium curve at any liquid concentration on a graph plotted in mole fractions.

The quantity $(1-y)_m$ is the average concentration of non-diffusing gas at either end of the diffusion path. $(1-y)$ is the concentration of the main body of the gas and $(1-y_{\text{eq}})$ that at the liquid gas interface.

$$(1-y)_m = \frac{(1-y_{\text{eq}}) - (1-y)}{\ln \frac{(1-y_{\text{eq}})}{(1-y)}} \cong \frac{(1-y_{\text{eq}}) + (1-y)}{2} \quad (10.27)$$

For all ordinary purposes the arithmetic mean is entirely satisfactory, and equation (10.25) is changed to:

$$N_{\text{tog}} = \int_{y_2}^{y_1} \frac{dy}{y - y_{\text{eq}}} + \frac{1}{2} \ln \frac{1-y_2}{1-y_1} \quad (10.28)$$

The calculation of the number of transfer units for dilute mixtures can be simplified. When the gas mixture is dilute, the second term of equation (10.27) becomes negligible and the equation may be simplified as follows:

$$N_{\text{tog}} = \int_{y_2}^{y_1} \frac{dy}{y - y_{\text{eq}}} \quad (10.29)$$

If the equilibrium curve and the operating line in terms of mole fraction are considered as straight lines, it is possible to re-write equation (10.29) as:

$$N_{\text{tog}} = \frac{y_1 - y_2}{(y - y_{\text{eq}})_m} \quad (10.30)$$

Equation (10.30) demonstrates that one overall gas-transfer unit is obtained when the change in gas composition equals the average of the overall driving force causing the change. Let us consider the diagram shown in Fig. 10.8. The line (3) is vertically halfway between the operating line (2) and the equilibrium curve (1). The step CFD, which corresponds to one transfer unit, has been constructed by drawing the horizontal line CEF, so that CE is equal to EF, and continuing vertically to D.

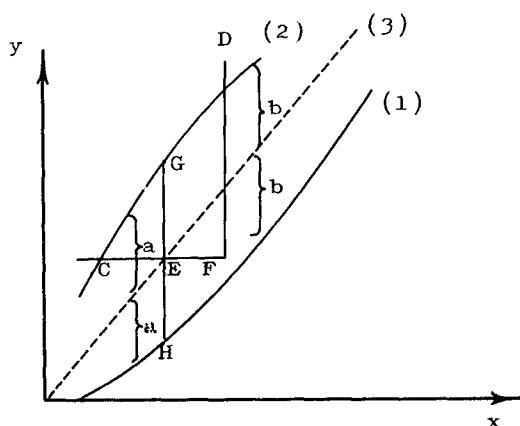


Fig. 10.8. Graphical determination of transfer units (absorption).

$y_G - y_H$ may be considered as the average driving force for the exchange in gas composition $y_O - y_F$ corresponding to this step. As GE is equal to EH and if the operating line is straight $DF = 2 \cdot GE = GH$, and the step CFD corresponds to one transfer unit. In a similar way the other transfer units were stepped off.

The resistance to mass transfer in absorption and stripping processes in the case both the gas film and liquid film are controlling factors can be calculated on the basis of the following equation:

$$\frac{1}{K_y \cdot a} = \frac{1}{K_G \cdot a} + \frac{m}{K_L \cdot a} \quad (10.31)$$

where m = the slope of the equilibrium solubility curve (mole fraction in the gas/mole fraction in the liquid).

By comparison of equation (10.23) with (10.26) H_{tog} can be expressed by the contribution of individual phase resistances, H_{tG} and H_{tL} :

$$H_{tog} = H_{tG} + \frac{mG}{L} H_{tL} \frac{(1-x)_m}{(1-y)_m} \quad (10.32)$$

For diluted solutions, the ratio of concentrations of nondiffusing substances will be nearly unity, and:

$$H_{tog} = H_{tG} + \frac{mG}{L} H_{tL} \quad (10.33)$$

As mentioned previously, absorption or stripping of very insoluble gases such as oxygen, hydrogen or carbon dioxide, is controlled by resistance to mass transfer in the liquid, for which H_{tL} is a direct measure. H_{tL} can be found for common packing from the empirical expression:

$$H_{tL} = \emptyset \left(\frac{L}{0.31 \cdot \mu_L} \right)^\eta Sc_L^{0.5} \quad (10.34)$$

where $\emptyset$ and η can be found from Table 10.3 for different packings.

L = the flowrate kg/h/m^2

Sc_L = the dimensionless Schmidt number = $\mu_L / \rho_L \cdot D_L$

μ_L = the viscosity (kg/m/h)

ρ_L = specific gravity

D_L = diffusion coefficient

In some instances $H_{tog} \approx H_{tG}$. This almost obtains for the stripping of ammonia from water into air, but in this case the liquid-foam resistance is still not completely negligible despite the fact that ammonia is very soluble in water.

It is possible to calculate H_{tG} from empirical data:

$$H_{tG} = \frac{\alpha \cdot G^\beta}{L^\gamma} \cdot Sc_G^{0.5} \quad (10.35)$$

where α , β and γ = empirical constants, Sc_G = the dimensionless Schmidt number, $Sc_G = \mu_G / \rho_G \cdot D_G$, and G and L = the gas and liquid flowrates measured in kg/h/m^2 .

TABLE 10.3

Liquid-film height of transfer unit

$$H_{tL} \cdot \phi \left(\frac{L}{\mu_L} \right)^\eta Sc_L^{0.5}.$$

 $H_{tL} = m$, $L = \text{kg/h/m}^2$, $\mu_L = \text{kg/m/h}$, $Sc_L = \text{dimensionless (Schmidt number)}$

Packing	ϕ	η	Range of L
Rashig rings:			
3/8 in.	$3.15 \cdot 10^{-4}$	0.46	} 1,800-68,000
1/2 in.	$7.05 \cdot 10^{-4}$	0.35	
1 in.	$2.30 \cdot 10^{-3}$	0.22	
1.5 in.	$2.56 \cdot 10^{-3}$	0.22	
2 in.	$2.88 \cdot 10^{-3}$	0.22	
Berl saddles:			
1/2 in.	$1.43 \cdot 10^{-3}$	0.28	
1 in.	$1.26 \cdot 10^{-3}$	0.28	
1.5 in.	$1.34 \cdot 10^{-3}$	0.28	
3-in. partition rings (stacked staggered)	0.0168	0.09	13,000-63,000
Spiral rings (stacked staggered):			
3-in. single spiral	$1.95 \cdot 10^{-3}$	0.28	1,800-68,000
3-in. triple spiral	$2.49 \cdot 10^{-3}$	0.28	13,000-63,000
Drip-point grids (continuous flue):			
No. 6146	$3.51 \cdot 10^{-3}$	0.23	15,000-135,000
No. 6295	$1.50 \cdot 10^{-3}$	0.31	11,000-100,000

From the data of Sherwood et al., and Molstad et al.

Values of the empirical constants are listed in Table 10.4.

The diameter of the tower is calculated on the basis of the minimum liquid rate for wetting and on the so-called flooding point. The minimum liquid rate for wetting l_w , can be calculated from the following equation:

$$l_w = \frac{L}{\rho_L \cdot a} \quad (10.36)$$

where

 ρ_L = the density of the liquid kg/m^3 a = surface area of the packing m^2/m^3

TABLE 10.4

Gas-film height of transfer unit

$$H_{tG} = \frac{\alpha G^\beta}{L^\nu} Sc_G^{0.5}$$

 $H_{tG}=m$, $G=\text{kg/h/m}^2$, $L=\text{kg/h/m}^2$, $Sc_G=\text{dimensionless (Schmidt number)}$

Packing	α	β	ν	Range of	
				G	L
Rashig rings:					
3/8 in.	2.39	0.45	0.47	900-2,300	2,300-6,800
1 in.	9.31	0.39	0.58	900-3,600	1,800-2,300
	8.53	0.32	0.51	900-2,700	2,300-20,000
1.5 in.	26.4	0.38	0.66	900-3,200	2,300-6,800
	2.66	0.38	0.40	900-3,200	6,800-20,000
2 in.	4.06	0.41	0.45	900-3,600	2,300-20,000
Berl saddles:					
1/2 in.	62.8	0.30	0.74	900-3,200	2,300-6,800
	0.741	0.30	0.24	900-3,200	6,800-20,000
1 in.	2.09	0.36	0.40	900-3,600	1,800-20,000
1.5 in.	6.14	0.32	0.45	900-4,500	1,800-20,000
3-in.partition rings (stacked staggered)	1338	0.58	1.06	700-4,100	13,000-20,000
Spiral rings (stacked staggered):					
3-in.single spiral	2.17	0.35	0.29	600-3,200	13,000-45,000
3-in.triple spiral	21.7	0.38	0.60	900-4,500	2,300-13,000
Drip-point (continuous flue):					
No. 6146	4.02	0.37	0.39	600-4,500	13,000-30,000
No. 6295	5.40	0.17	0.27	450-4,500	9,000-52,000

From the data of Fellienger and Pigford (1952) and Molstad et al. (1943)

The flooding point has been defined as the gas velocity at which a liquid layer forms on top of the packing. Based on experimental data, the following equation can be used for the determination of l_w at the flooding point:

$$\frac{l_w}{d_h^{2/3}} (1000 \mu_L)^{0.1} = f \left(\frac{L}{G} \sqrt{\frac{\rho_G}{\rho_L}} \right) \quad (10.37)$$

where d_h = the hydraulic diameter of the packing and μ_L = the viscosity, but in kg/m.s. The function is shown in Fig. 10.9.

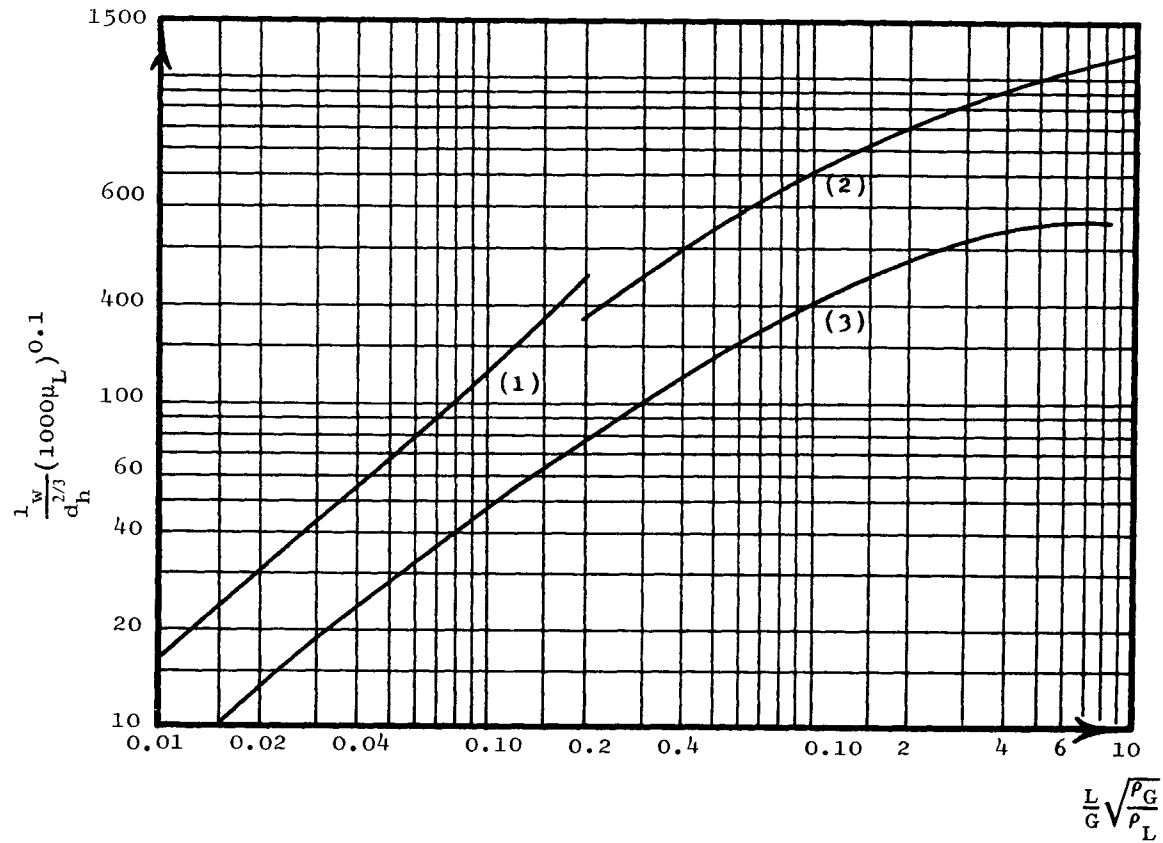


Fig. 10.9. Plot for determination of flooding point.
 (1) Grids. (2) Stacked rings. (3) Random packings of rings.

TABLE 10.5

Characteristic packing data

Packing	Dimensions (inch)			Number per m ³	Surface (m ² /m ³)	Poro- sity (%)	d _h =hydrau- lic diam. (m)	$\frac{3}{2} 10^3 d_h$	Gas flow entry tower (m/s)
	Diam.	Height	Thick- ness						
Coke	-	3	-	-	49	0.50	0.041	8.3	0.54-0.96
"	-	1-2	-	-	115	0.40	0.014	1.7	0.26
"	-	1	-	-	131	0.45	0.014	1.7	0.15-0.26
Broken stone	-	2	-	-	62.5	0.46	0.029	4.9	0.51-0.60
"	-	1/2-1 1/4	-	-	144	0.40	0.011	1.15	0.13
<u>Grids:</u>									
Wood	1	1	1/4	-	98.5	0.75	0.019	2.65	1.5 -2.4
"	1	2	1/4	-	88.5	0.75	0.019	2.65	1.7 -2.5
<u>Jagged grids:</u>									
Wood	4	4	1/2	-	19.5	0.89	0.089	26.5	2.4 -3.6
"	2	2	3/8	-	42.5	0.83	0.041	8.4	2.1 -3.3
"	1 1/2	1 1/2	3/16	-	54.0	0.89	0.033	6.1	2.1 -3.0
<u>Stacked</u>									
<u>Rashig rings:</u>									
Stoneware	4	4	3/8	950	62.5	0.73	0.047	10.2	1.6 -2.4
"	3	3	3/8	2300	82	0.66	0.032	5.7	1.1 -1.5
"	3	3	1/4	2300	82	0.76	0.037	7.1	1.7
"	2	2	1/4	7400	118	0.67	0.023	3.5	0.86
"	2	2	3/16	7400	118	0.72	0.024	3.7	0.89
Metal	2	2	1/16	6180	98.5	0.92	0.037	7.1	0.72-0.90
"	1	1	1/16	47600	194	0.86	0.018	2.4	0.57-0.69
"	1/2	1/2	1/32	370000	377	0.87	0.009	0.85	0.3
<u>Random</u>									
<u>packings of</u>									
<u>Rashig rings:</u>									
Stoneware	3	3	3/8	1840	65.5	0.72	0.044	9.2	0.67-1.2
"	2	2	1/4	5820	92	0.74	0.032	5.7	0.54-0.86
"	2	2	3/16	6000	95	0.79	0.033	6.0	0.63-0.93
"	1 1/2	1 1/2	3/16	14100	125	0.73	0.023	3.5	0.51-0.81
"	1	1	3/32	46000	184	0.80	0.017	2.2	0.42-0.60
"	3/4	3/4	3/32	106000	236	0.74	0.013	1.5	-
"	1/2	1/2	1/16	370000	377	0.73	0.008	0.72	0.19
<u>Bell-saddles:</u>									
Stoneware	1/2	-	-	528000	460	0.65	0.0057	0.43	} not indicated
"	1	-	-	81000	258	0.69	0.0107	1.10	
"	1 1/2	-	-	22900	165	0.72	0.017	2.21	
"	2	-	-	8000	120	0.72	0.024	3.72	

Partly after G.A. Morris and J. Jackson: Absorption Towers, 1953.

The flooding point represents the upper limit for the operation of the tower. Operating conditions of the tower can be improved by increasing the gas flow. Usually a gas flow of 50-60% of the flow corresponding to the flooding point is used. The diameter of the tower is found by the following procedure:

1. Based on L, G the specific gravity of the liquid and the gas, μ_L and Fig. 10.9 is found $\frac{l_w}{d_h^{2/3}} \cdot 10^3$. d_h is shown in Table 10.5 for different packing materials.
2. l_w and d_h must be chosen so, that l_w is greater than $0.08 \text{ m}^3/\text{m/h}$ for common packing including rashig rings less than 7.5 cm and greater than $0.12 \text{ m}^3/\text{h}$ for rashig rings larger than 3 in.
3. Generally, $0.4 \text{ m}^3/\text{m/h}$ can be considered as the upper limit for all types of packing.
4. Based on equation (10.36) and the total flows (kg/m^2) it is possible to find the area of cross-section of the absorption stripping tower.

Application of aeration

Air is introduced into waste water for a variety of reasons. The possibility of removing dissolved inorganic substances such as divalent iron or manganese by oxidation has already been mentioned. Preaeration is sometimes used in cases where raw waste water lacks dissolved oxygen. Furthermore, diffused compressed air is used in aerated flotation chambers and, finally, air is used in biological treatment of waste water to meet the metabolic oxygen requirements of the treatment organisms and to provide mixing within the reaction vessel. Oxygen is a very important factor in all aerobic biological oxidation processes and therefore deserves primarily consideration in process design operations. In the presence of high concentrations of organic compounds aerobic biological oxidation may even be limited by the rate at which oxygen is transferred to the system by natural or artificial aeration.

Application of stripping

The stripping process is used to remove volatile gases such as hydrogen sulphide, hydrogen cyanide and ammonia. The removal of ammonia by stripping is used in the treatment of municipal waste water, but it has also been suggested for the treatment of industrial waste water or for the regeneration of the liquid used for elution of ion exchangers (Jørgensen, 1976). The rate at which carbon dioxide, hydrogen sulphide, ammonia and hydrogen cyanide can be removed by air stripping is highly dependent on pH, since all four of these volatile gases are acids or bases. This is illustrated for the removal of hydrogen sulphide from water at various pH values in Fig. 10.10.

Also some organic solvents such as acetone and methylethylketone can be removed from waste water by stripping.

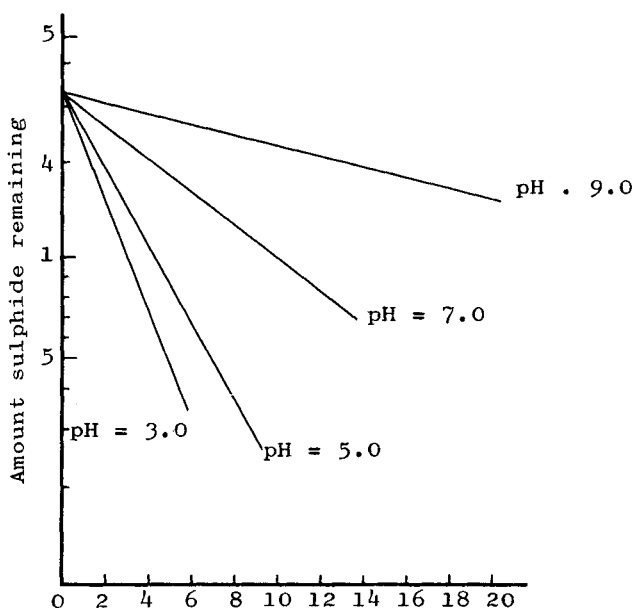


Fig. 10.10. Sulphide remaining versus time at pH 3.0, 7.0 and 9.0.

Example 10.1

The design data for an aeration plant are required.

The plant must treat 0.5 m^3 of water per minute and provide an oxygen concentration of between 8.5 and 9.0 mg/l, i.e. between 84.9 and 89.9% saturation at ambient temperature (16°C).

An aerator supplying $15 \text{ m}^3/\text{h}$ is available and the area of the contact is $1.2 \cdot 1.2 \text{ m}^2$.

Solution

Based on Henry's Law:

$$p = H \cdot X$$

we find that $X = 5.66 \cdot 10^{-6}$ since $p = 0.21 \text{ atm}$.

This gives an oxygen concentration of

$$5.66 \cdot 10^{-6} \cdot 32 \cdot 55.5 \cdot 10^3 = 10.1 \text{ mg/l.}$$

Based on equation (10.34) H_{tL} can be found:

$$\mu_L = \frac{0.01 \text{ g}}{\text{cm} \cdot \text{sec}} = \frac{0.01 \cdot 10^{-3} \text{ kg} \cdot 3600 \cdot 100}{\text{m} \cdot \text{h}} = 3.6 \text{ kg/m/h}$$

$$Sc = \frac{\mu_L}{\rho_L \cdot D_L} = 558 \text{ for water}$$

If we choose a packing of 0.024 m size, $\phi = 2.3 \cdot 10^{-3}$ and $\eta = 0.22$, we find $H_{tL} = 0.37$ m.

Based on a diagram giving the equilibrium data (use Henry's Law or Table) and the operating line, it is possible to see that one transfer unit is sufficient. This means, that the height of the aeration vessel should be a minimum of 0.37 m.

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