7. AIR STRIPPING

7.1 Physico-chemical Principles of Air Stripping

The stripping process is used to remove volatile gases, such as hydrogen sulfide, hydrogen cyanide and ammonia by blowing air through the waste water. The process is therefore to be considered as a transfer from a liquid phase to a gas phase. The basic principle of this process of nitrogen removal is illustrated in Fig. 7.1.

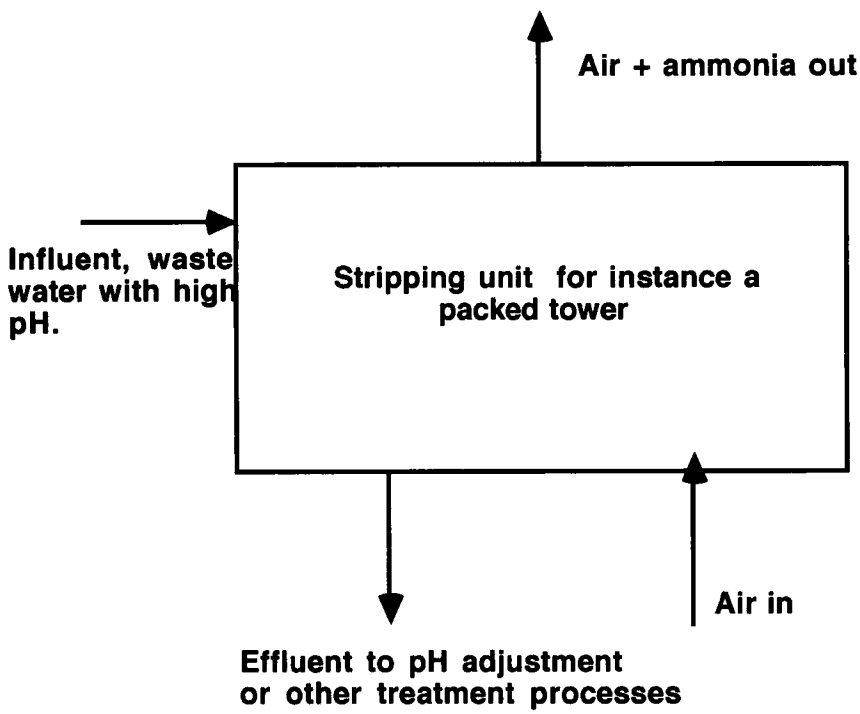

Figure 7.1. The principle of air stripping.

The rate at which ammonia can be removed by air stripping is highly dependent on pH, because the exchange between the two forms, ammonium, which is the ionic form, and ammonia, which is a highly water-soluble gas, is an acid-base reaction. The ammonia stripping is based on the following reaction:

$$NH_4^+ <=> NH_3 + H^+ \qquad (7.1)$$

The equilibrium constant for this process is $10^{-9.25}$ at 18°C, which means that:

$$\frac{[NH_3]\,[H^+]}{[NH_4^+]} = 10^{-9.25} \qquad (7.2)$$

By separating H^+ in this equation and converting to a logarithmic form, we get:

$$pH = 9.25 + \log \frac{[NH_3]}{[NH_4^+]} \qquad (7.3)$$

Knowing the ammonium concentration in an aquatic ecosystem, this relationship can be used to estimate the toxicity level of the water, see Section 1.4. From equation (7.3) we can see that at pH = 9.25, 50% of the total ammonia-nitrogen is in the form of ammonia and 50% in the form of ammonium. Correspondingly the ratio between ammonia and ammonium is 10 at pH 10.25 and 100 at pH 11.25. A graph showing the ratio ammonia to ammonium is given in Fig. 7.2. Consequently it is necessary to adjust the pH to 10 or more before the stripping process is used. The pK_a value, which is the negative logarithm to the equilibrium constant, is dependent on the presence of other ions, or expressed in another way, of the ionic strength of the influent. The ionic strength is defined by the following expression:

$$I = \sum 1/2\ C Z^2 \qquad (7.4)$$

where C = the molar concentration of the considered ions and Z = the charge.

On the basis of the ionic strength, it is possible to find the activity coefficient, f, from:

$$-\log f = \frac{0.5 \cdot Z^2 \cdot \sqrt{I}}{\sqrt{I} + 1} \qquad (7.5)$$

where I = ionic strength, Z = charge and f = activity coefficient. The activity coefficient, f, is defined as the activity a, divided by the concentration c. The activity is used in the mass equations to replace the concentrations, if the ionic strength is sufficient high to play a significant role, see also below.

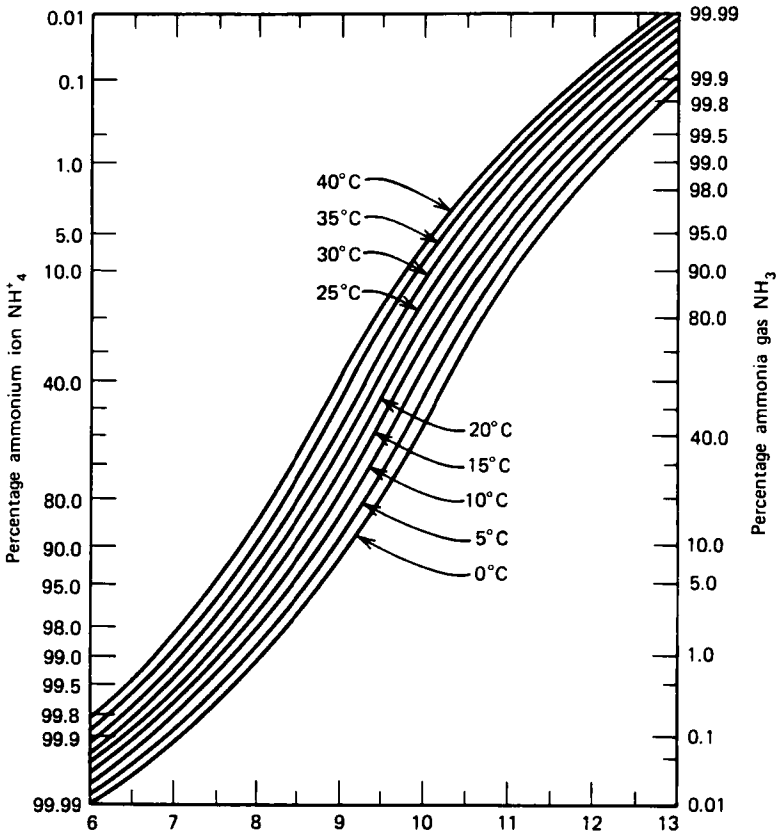

Fig. 7.2. Distribution of ammonia and ammonium as function of pH and temperature.

If the ionic strength plays a role, the concentrations in equation (7.2) are replaced by activities. As pH is defined from the activity of hydrogen ions, (7.2) will be changed to the following expression in this case:

$$\frac{[NH_3] * a_{H^+}}{[NH_4^+]} = 10^{-9.25} * f \qquad (7.6)$$

Equation (7.3) will be changed correspondingly:

$$pH = 9.25 - \log f + \log \frac{[NH_3]}{[NH_4^+]} \qquad (7.7)$$

or

$$pH = 9.25 + \frac{0.5 * Z^2 * \sqrt{I}}{\sqrt{I} + 1} + \log ([NH_3] / [NH_4^+]) \qquad (7.8)$$

As seen from equation (7.8) the ratio ammonia / ammonium is disfavored by increased ionic strength, implying that a higher pH is need to obtain the same stripping effect at higher ionic strength.

Table 7.1 gives the activity coefficients for different ionic charges, calculated from the equation (7.5).

TABLE 7.1
Activity coefficient f at different ionic strengths

I	$\frac{\sqrt{I}}{1+\sqrt{I}}$	f for Z = 1	f for Z = 2	f for Z = 3
0	0	1.00	1.00	1.00
0.001	0.03	0.95	0.82	0.64
0.005	0.07	0.93	0.74	0.51
0.01	0.09	0.90	0.66	0.40
0.02	0.12	0.87	0.57	0.28
0.05	0.18	0.81	0.43	0.15
0.1	0.24	0.76	0.33	0.10
0.2	0.31	0.70	-	-
0.5	0.41	0.62	-	-

I = ionic strength, Z = charge, f = activity coefficient

Since calcium hydroxide is the cheapest source of hydroxide ions, it is most

often used for adjustment of pH before the stripping process. The addition of calcium hydroxide leads to an increased ionic strength. However, the ionic strength of most waste waters, after addition of sufficient calcium hydroxide to obtain a pH of 10 or above, is only in the order of 0.05-0.1, which implies that the increase of pH needed to obtain the same stripping effect as for distilled water is approximately only 0.1.

7.2 Process Variables

As much as 13 g ammonia gas is soluble at room temperature in 100 ml water. Due to this very high solubility of ammonia in water a large quantity of air is required to transfer ammonia effectively from the water to the air. In principle there are three different configurations of stripping units, as shown in Fig. 7.3; see Montgomery (1985).

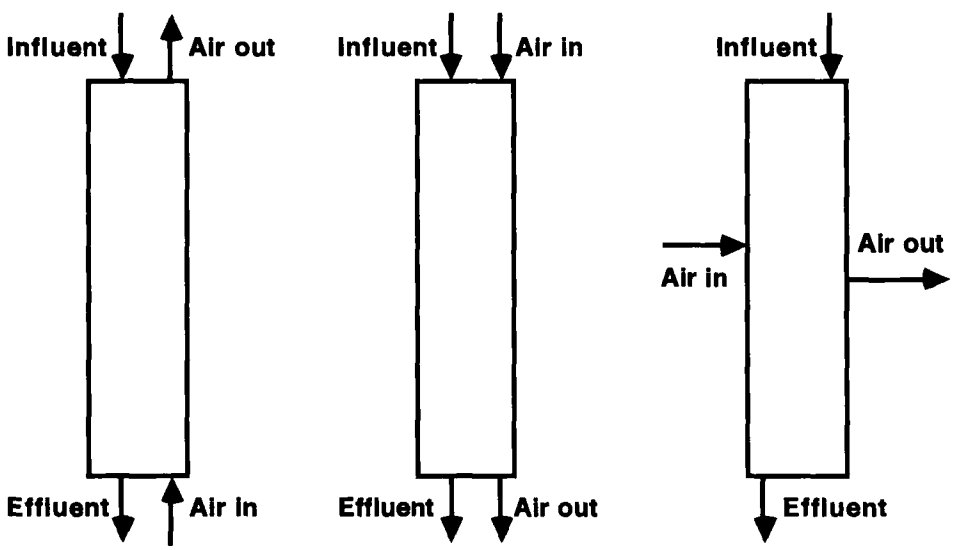

Figure 7.3. Configuration of air stripping units. From left to right: countercurrent, cocurrent and cross flow.

The efficiency of the process depends on:

1. **pH**, according to the considerations mentioned above. Equations (7.2) and

(7.3) may be applied and in case where the ionic strength is significant, equations (7.7) and (7.8) are used.

2. **The temperature**. The solubility of ammonia decreases with increasing temperature. The efficiency at three temperatures - 0°C, 20°C and 40°C - is plotted versus the pH in Fig. 7.4 and versus the tower height in Fig. 7.5

3. **The quantity of air per m^3 of water treated**. At least 3000 m^3 of air per m^3 of water are required (see Fig.7.6).

4. **The height of the stripping tower**. The relationship between the efficiency and the quantity of air is plotted for three heights - Figs. 7.5 and 7.6.

5. **The specific surface of the packing** (m^2/m^3). Greater specific surface results in greater efficiency.

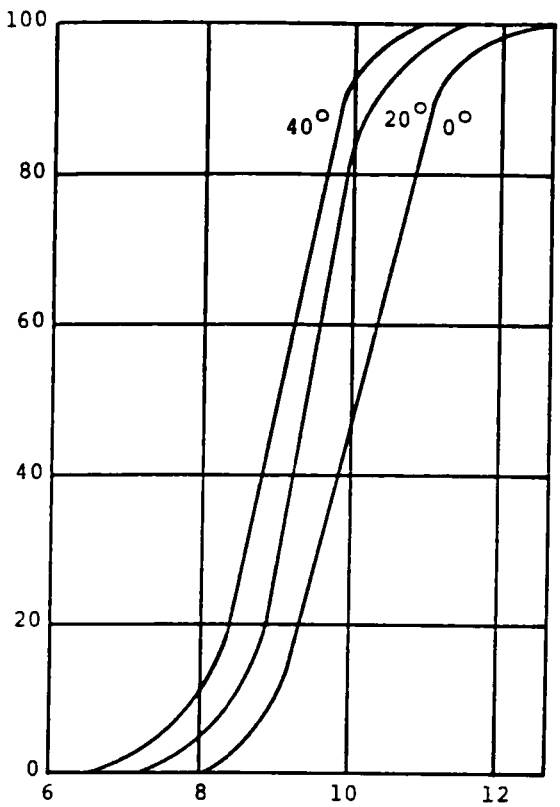

Fig. 7.4. Stripping efficiency as function of pH at three different temperatures.

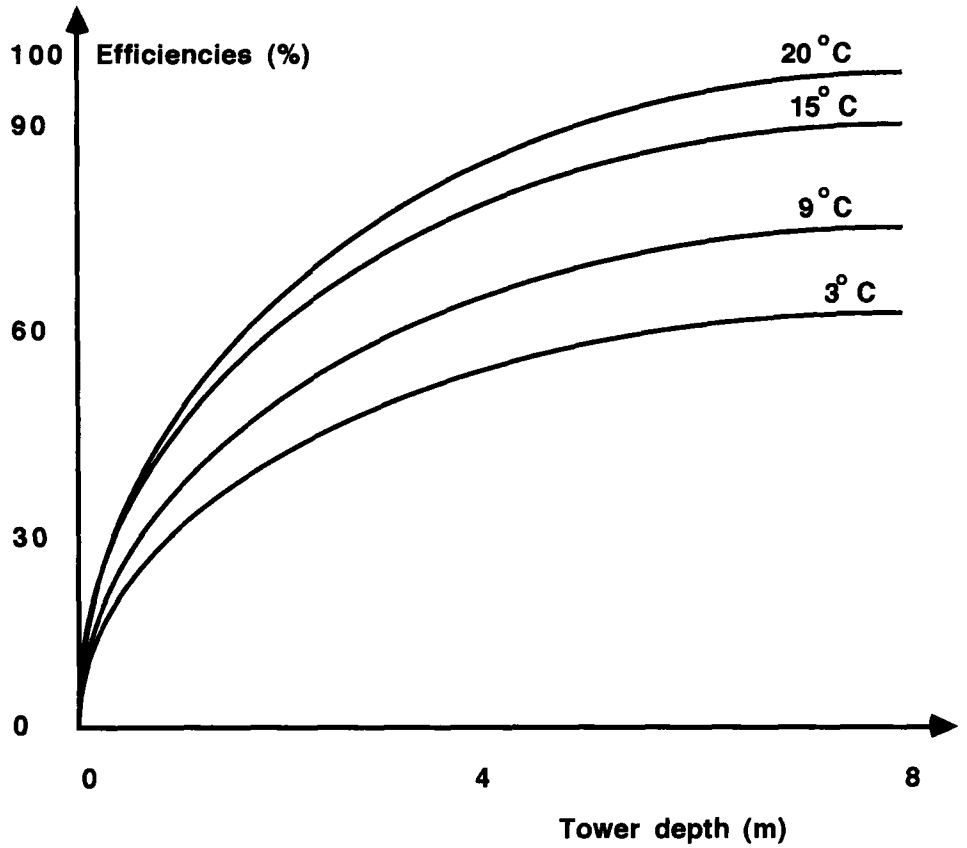

Figure 7.5. Effect of water temperature on ammonia stripping. 4 m^3 air is used per liter of waste water. The efficiencies are plotted versus the tower height for various temperatures.

Figure 7.7 demonstrates the principle of a stripping tower. The waste water treatment plant at Lake Tahoe, California, includes a stripping process. 10,000 m^3 of waste water is treated per 24h at a cost of approximately 8 US cents (1992) per m^3. The capital cost is in the order of 20 US cents per m^3 (based on 16% depreciation and interest per year of the investment).

The cost of stripping is therefore relatively moderate, but the process has two crucial limitations:

1. It is practically impossible to work at temperatures below 5-7°C. The large quantity of air will cause considerable evaporation, which results in the

water in the tower freezing.

2. Deposition of calcium carbonate can reduce the efficiency or even block the tower.

Due to limitation 1) it will be necessary to use warm air for the stripping during winter in temperate climates, or to install the tower indoors. This makes the process too costly for plants in areas with more than 10,000 inhabitants and limits the application for treatment of bigger volumes to tropical or possibly subtropical latitudes.

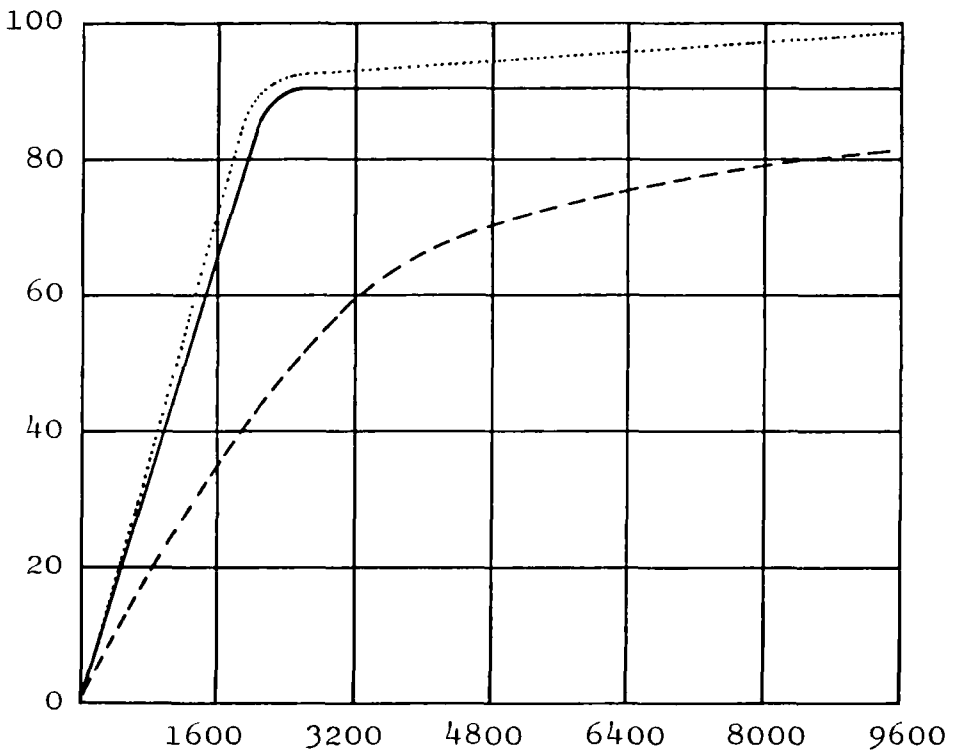

Figure 7.6. Efficiency as function of m^3 of air per m^3 of water for three different tower heights ..... line = 8 m, _____ line = 6.7 m, --- line = 4 m.

A very important shortcoming of some technological solutions is, that *they do not consider a total environmental solution, as they solve one problem but create a new one.* The stripping process is a characteristic example, since *the*

ammonia is removed from the waste water *but transferred to the atmosphere, unless recovery of* ammonia is carried out. In each specific case it is necessary to assess whether the air pollution problem created is greater that the water pollution problem solved. If a significant amount of municipal waste water were be treated by air stripping, the ammonia removed by air would make a crucial contribution to the air pollution problem of nitrogenous compounds on a regional basis.

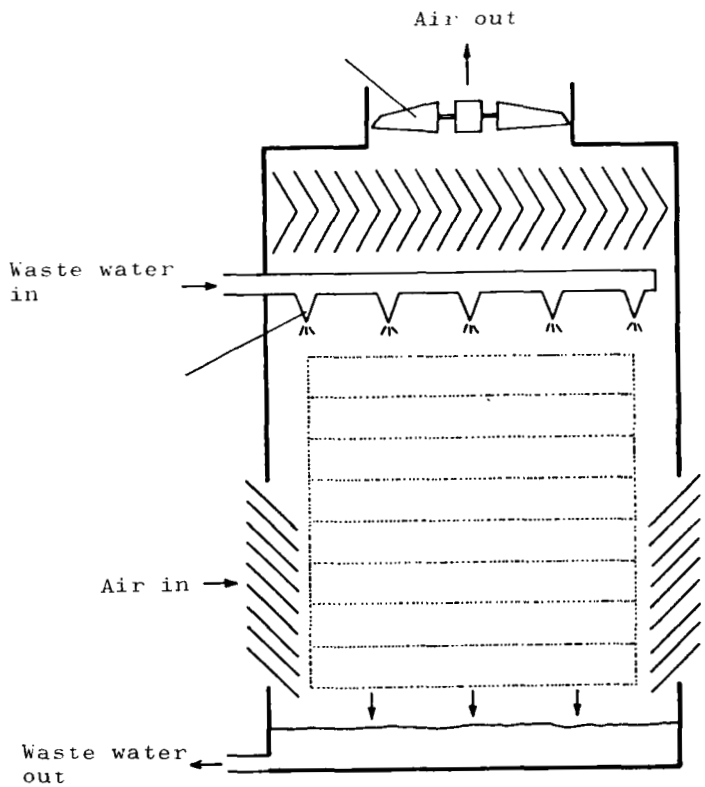

Fig. 7.7. The principle of a stripping tower.

7.3. Gas Transfer

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 is a concentration gradient in the liquid phase.

The transfer of a gas can be treated as a four-step process. The first step of a stripping process involves passage of the dissolved gas from the liquid phase to the gas-liquid interface. The second step is the passage of the gas through a liquid film on the liquid side of the interface The gas must then pass through a gas film on the vapor side of the interface. The gas must in the final step be dispersed throughout the bulk of the gas. 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 most often 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:

$$N = -D * A * \frac{dc}{dy} \qquad (7.9)$$

where

N = mass transfer per unit time

A = the cross-sectional area across which diffusion occurs

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:

$$C_{eq} = p / H \qquad (7.10)$$

where

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

H = Henry's Constant

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

Henry's Constant is roughly proportional to the temperature; i.e., with increased temperature the solubility of a gas decreases. Figure 7.8 gives the relation between solubility of ammonia and the temperature. As can be seen, the solubility changes significantly with the temperature; see also Figs. 7.2, 7.4 and 7.5.

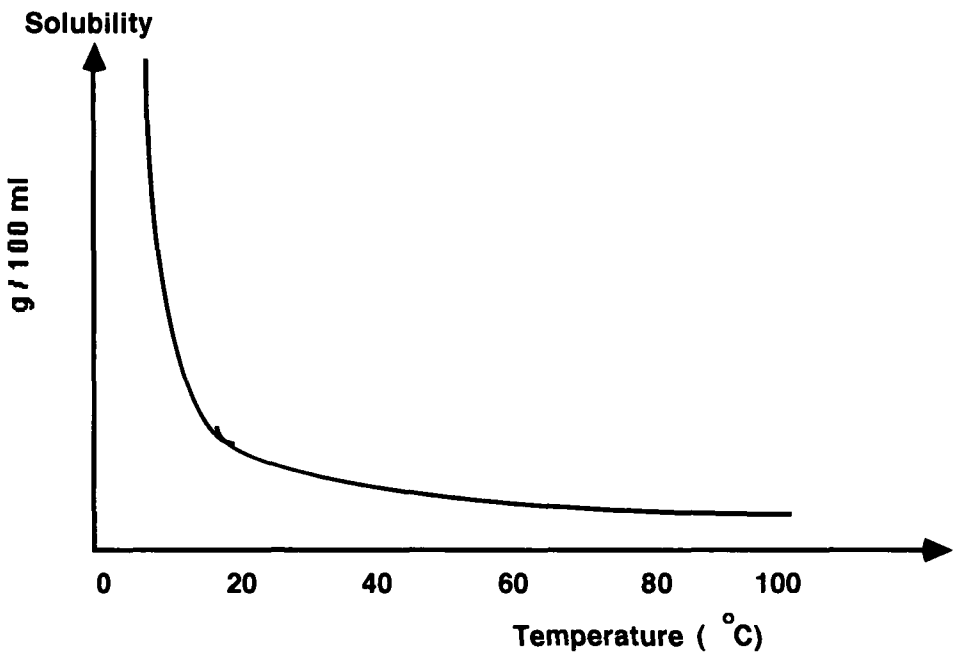

Figure 7.8. The solubility of ammonia plotted versus the temperature.

The temperature dependence of Henry's constant may be found by

use of one of the following two equations: (Srinath and Loehr 1974 and Montgomery 1985)

$$H = 0.268 \cdot \exp(0.0525 \cdot t\,^{o}C) \quad \text{bar} \tag{7.11}$$

$$H = 3754 / (1.987 \cdot (273 + t\,^{o}C) + 6.135 \quad \text{bar} \tag{7.12}$$

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 are expressed by the following equation (Gameson and Robertson, 1955):

$$C_{eq} = \frac{475 - 2.65 \cdot C_{ds}}{33.5 + (T - 273)} \tag{7.13}$$

where

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

T = the absolute temperature expressed in K

It must be emphasized that this equation is developed under the conditions that the pressure is 760 mm Hg and that clean water is in contact with wet air.

In this context it must be stressed 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:

$$N = K_L \cdot A(C_{eq} - C) = K_G \cdot A(p - p_{eq}) \tag{7.14}$$

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 coefficient defined as D_L/Y_L
K_G = gas-film coefficient defined as D_G/Y_G
D_L = diffusion coefficient in the liquid
D_G = diffusion coefficient in the gas

Figure 7.9 shows a schematic representation of the liquid-gas mass transfer.

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

$$\frac{1}{V}N = \frac{dc}{dt} = K_L * \frac{A}{V}(C_{eq} - C) \qquad (7.15)$$

$K_{L,a} = K_L * (A/V)$ 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:

$$K_L(t) = K_{L,20°} \cdot 1.028^{(t-20)} \qquad (7.16)$$

t = temperature (°C).

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

$$K_{L,a}(t) = K_{L,a,20°} \cdot 1.02^{(t-20)} \qquad (7.17)$$

The presence of surface-active agents in the waste water has a significant effect on K_L and A/V (area to volume ratio). A decrease in surface tension will decrease the size of the bubbles generated, which will increase A/V. In some instances the increase in 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. A coefficient, b, defined as the as the ratio of $K_{L,a}$ for waste water to that for distilled water, is used to account for the influence of

the impurities in the waste water on $K_{L,a}$. Figure 7.10 shows a characteristic change in the coefficient b, as a function of BOD_5 of water.

The liquid film resistance is usually not of importance for ammonia stripping. It is therefore possible to relate the transfer process directly to the gas film resistance, which in practice is performed by empirical relations between the resistance coefficients and the tower packing.

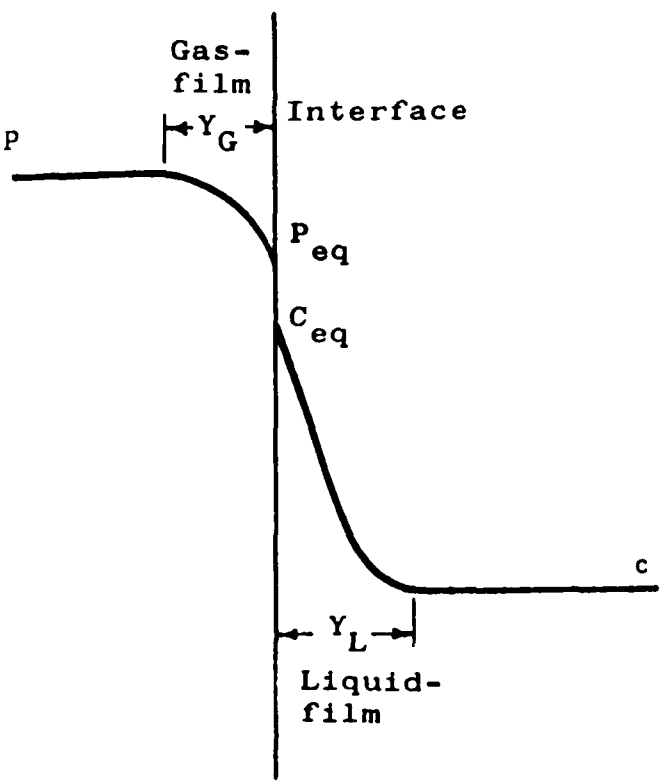

Figure 7.9. Schematic representation of interfacial mass transfer.

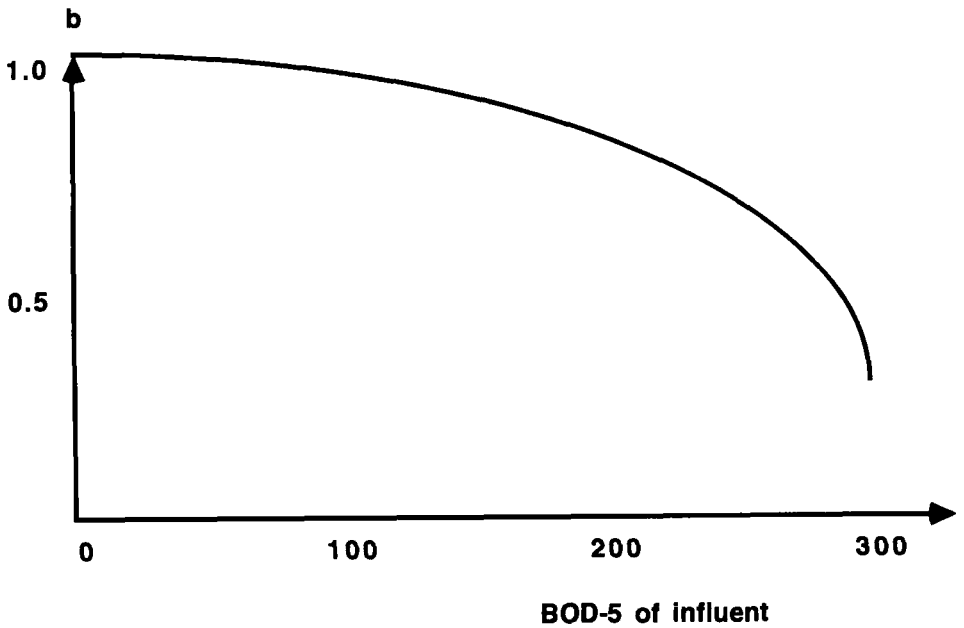

Figure 7.10. A typical BOD5 / b relationship.

7.4. Design of Stripping Tower

Figure 7.11 shows the application of the mass conservation principle on 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 ($y \ll 1$):

$$Y = \frac{y}{1-y} = \frac{p}{P_t - p} \tag{7.18}$$

and

$$G_s = G(1-y) = \frac{G}{1+Y} \tag{7.19}$$

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 non-diffusing, essentially insoluble gas

P_t = total pressure

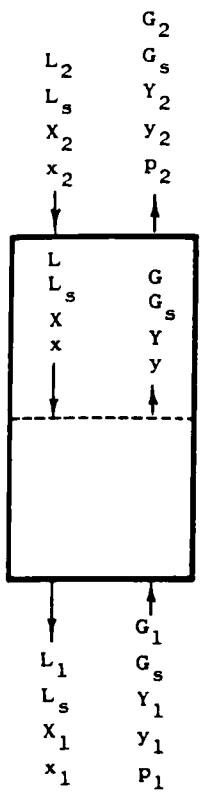

Figure 7.11. Principles of mass conservation applied to countercurrent tower.

Similarly, the following equation is valid for the liquid stream ($x << 1$):

$$X = \frac{x}{1 - x} \tag{7.20}$$

$$L_s = L(1 - x) = \frac{L}{1 + X} \qquad (7.21)$$

where

L = liquid stream moles/h/m^2

x = mole fraction of soluble gas

X = mole ratio of soluble gas

L_s = moles /h / m^2 of non-volatile solvent

Since the solvent gas (air) and solvent liquid (water) 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. 7.11) can be expressed by

$$G_s (Y_1 - Y) = L_s (X_1 - X) \qquad (7.22)$$

This is the equation of a straight line, the so-called operating line, which has a slope of 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. 7.12 the operating line is plotted together with the equilibrium solubility curve, which may be found from Henry's law and plotted in terms of the mole ratio.

For a stripping tower, the operating line is always below the equilibrium solubility curve (see Fig. 7.12).

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:

$$dS = a^*dZ \qquad (7.23)$$

where

S = area of the interface 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.

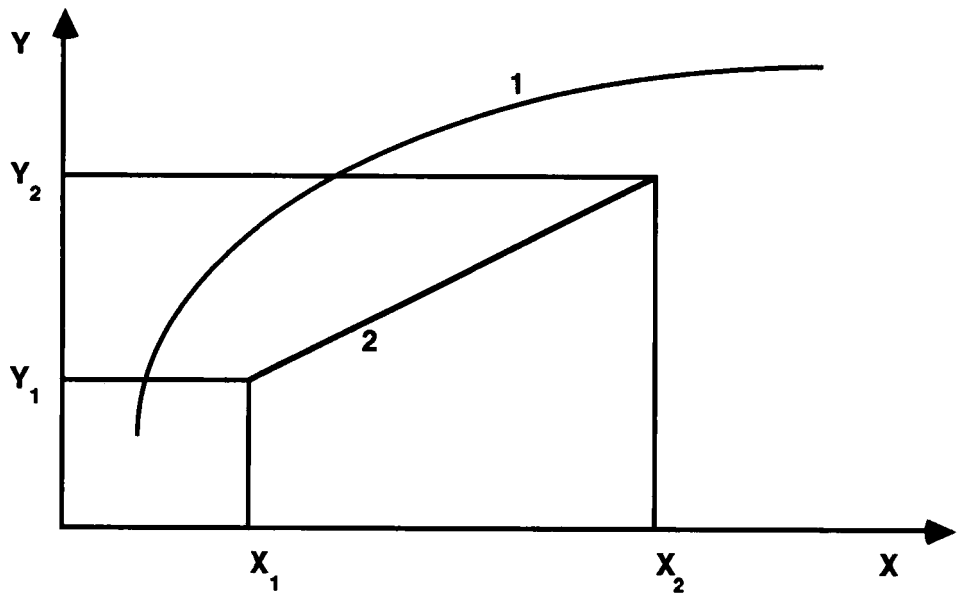

Figure 7.12. Equilibrium curve (1) and operating line (2) for a stripping process.

The amount of solid in the gas passing the differential section of the tower is $G * y$ mole/h/m^2, and the rate of mass transfer to the liquid, $d(G * y)$. This can be related to the mass transfer coefficient as follows:

$$d(G * y) = K_y * a(y - y_{eq})dZ \qquad (7.24)$$

where K_y = the overall transfer coefficient.

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:

$$d(G * y) = G_s * d\left(\frac{y}{1 - y}\right) = G_s \frac{dy}{(1 - y)^2} = \frac{Gdy}{1 - y} \qquad (7.25)$$

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 $K_G{}^*a(1-y)_m$ will be much more constant than $K_G{}^*a$ alone. Therefore, equation (7.25) will be transformed to

$$d(G * y) = \frac{Gdy}{1 - y} = \frac{(K_G{}^*a\,(1 - y)_m)\,(y - y_{eq})\,dZ}{(1 - y)_m} \qquad (7.26)$$

or

$$\frac{(1 - y)_m}{(1 - y)} * \frac{dy}{y - y_{eq}} = \frac{K_G * a\,(1 - y)_m * dZ}{G} \qquad (7.27)$$

Equation (7.27) may be integrated to obtain, Z, in terms of $K_G{}^*a$, 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

$$N_{tog} = \int_{y2}^{y1} \frac{(1 - y)_m * dy}{(1 - y)\,(y - y_{eq})} \qquad (7.28)$$

then

$$Z = N_{tog} * H_{tog} \qquad (7.29)$$

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 $K_G{}^*a$ and other mass-transfer coefficients in the design of towers. H_{tog} has the dimension of length. The subscript, tog, is used, as seen in the terms N_{tog} and H_{tog} to indicate that these terms are based on an overall driving force $y - y_{eq}$ 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 $(1-y_{eq})$ that at the liquid gas interface.

$$(1 - y)_m = \frac{(1 - y_{eq}) - (1 - y)}{\ln \frac{(1 - y_{eq})}{(1 - y)}} \approx \frac{(1 - y_{eq}) + (1 - y)}{2} \qquad (7.30)$$

For all ordinary purposes the arithmetic mean is entirely satisfactory, and equation (7.28) is changed to

$$N_{tog} = \int_{y_2}^{y_1} \frac{dy}{y - y_{eq}} + 1/2 \ln \frac{1 - y_2}{1 - y_1} \qquad (7.31)$$

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 (7.29) becomes negligible and the equation may be simplified as follows:

$$N_{tog} = \int_{y_2}^{y_1} \frac{dy}{y - y_{eq}} \qquad (7.32)$$

If the equilibrium curve and the operating line in terms of mole fraction are considered as straight lines, it is possible to rewrite equation (7.32) as:

$$N_{tog} = \frac{y_1 - y_2}{(y - y_{eq})_m} \qquad (7.33)$$

Equation (7.33) demonstrates that one overall gas-transfer unit is obtained when the change in gas composition equals the average of the overall driving forces causing the change. Let us consider the diagram shown in Fig. 7.13. The line (3) is vertically half-way 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.

TABLE 7.2

Liquid-film height of transfer unit

$$HtL = \varphi \left(\frac{L}{\mu L}\right)^{n} \cdot ScL^{0.5}$$

HtL = m, L=kg/h/m2, μL = kg/m/h, ScL = dimensionless (Schmidt number)

Packing	φ	n	Range of L
Raschig rings:			
3/8 in.	$3.15 \cdot 10^{-4}$	0.46	
1/2 in.	$7.05 \cdot 10^{-4}$	0.35	
1 in.	$2.30 \cdot 10^{-3}$	0.22	1,800-68,000
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. (1940), and Molstad et al. (1943)

$Y_G - Y_H$ may be considered as the average driving force for the exchange in

gas composition y_0 - y_F corresponding to this step. As GE is equal to EH and if the operating line is straight DF = 2 * GE = GH, and the step CFD corresponds to one transfer unit. In a similar way the other transfer units are stepped off.

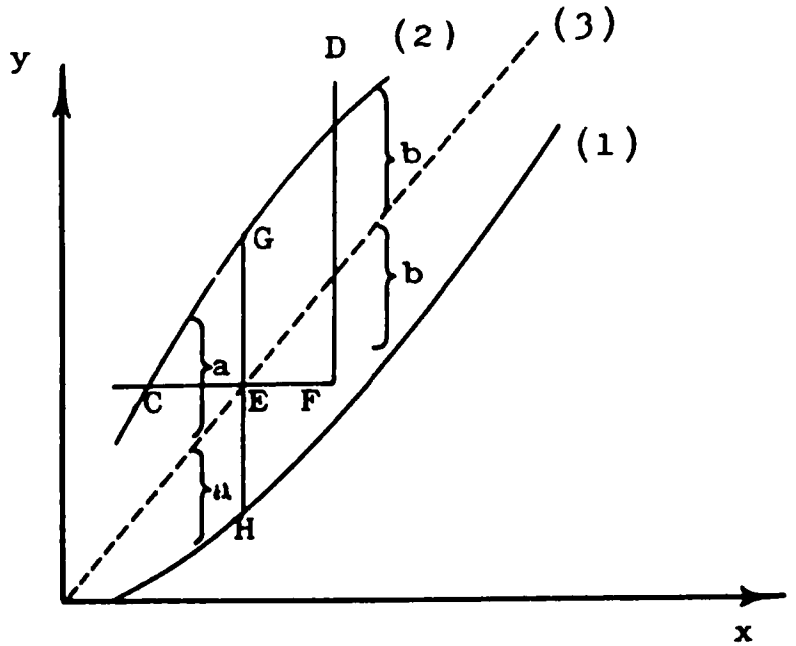

Figure 7.13. Graphical determination of transfer units (absorption).

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:

$$\frac{1}{K_y * a} = \frac{1}{K_G * a} + \frac{m}{K_L * a} \qquad (7.34)$$

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

By comparing equation (7.26) with (7.29), H_{tog} can be expressed by the contribution of individual phase resistances, H_{tG} and H_{tL}:

$$H_{tog} = H_{tG} + \frac{mG}{L} H_{tL} \frac{(1 - x)m}{(1 - y)m} \qquad (7.35)$$

For diluted solutions, the ratio of concentrations of non-diffusing substances will be nearly unity, and:

$$H_{tog} = H_{tG} + \frac{mG}{L} H_{tL}, \qquad (7.36)$$

where L is the flowrate in kg / h / m^2.

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 material from the empirical expression

$$H_{tL} = \varphi \left\{ \frac{L}{0.31 * \mu_L} \right\}^n * Sc_L^{0.5} \qquad (7.37)$$

where φ and n can be found from Table 7.2 for different packings.

L = the flow rate kg/h/m^2

Sc_L = the dimensionless Schmidt number = $\mu_L / p_L * D_L$

μ_L = the viscosity (kg/m/h)

p_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 although ammonia is very soluble in water.

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

$$H_{tG} = \frac{\alpha * G^{\beta}}{L^{\gamma}} * Sc_G^{0.5} \qquad (7.38)$$

where α, ß and γ are empirical constants, Sc_G = the dimensionless Schmidt number, $Sc_G = \mu_G / p_G * D_G$, G and L = the gas and liquid flow rates respectively measured in kg /h / m^2. p_G is the specific gravity of the gas.
The diameter of the tower is calculated on the basis of the minimum liquid rate for wetting and on the so-called flooding point.

Values of the empirical constants are listed in Table 7.3.

The minimum liquid rate for wetting l_w, can be calculated from the following equation:

$$l_w = \frac{L}{d_L * a} \qquad (7.39)$$

where

d_L = the density of the liquid kg/m^3

a = surface area of the packing m^2 / m^3

L = See Table 7.2

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:

$$\frac{l_w * (1000\,\mu_L)^{0.1}}{d_h^{2/3}} = f\left(\frac{L}{G}\sqrt{\frac{p_G}{p_L}}\right) = f(Q) \qquad (7.40)$$

where d_h = the hydraulic diameter of the packing and μ_L = the viscosity in kg/m*s.

Table 7.3 is based on data of Fellinger and Pigford (1952) and Molstad et al. (1943).

The function is shown in Fig. 7.14, where

$$Z = \frac{l_w * (1000\,\mu_L)^{0.1}}{d_h^{2/3}}$$ is expressed as a function of Q.

TABLE 7.3

Gas-film height of transfer unit

$$HtG = \frac{\alpha G^{\beta}}{L^{\gamma}} ScG^{0.5}$$

HtG=m, G=kg/h/m2, L =kg/h/m2, ScG=dimensionless (Schmidt number)

Packing	a	ß	y	Range of G	Range of L
Raschig rings:					
3/8 in.	.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
4 in	1.80	0.40	0.40	5,000-10,000	2,500-20,000
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

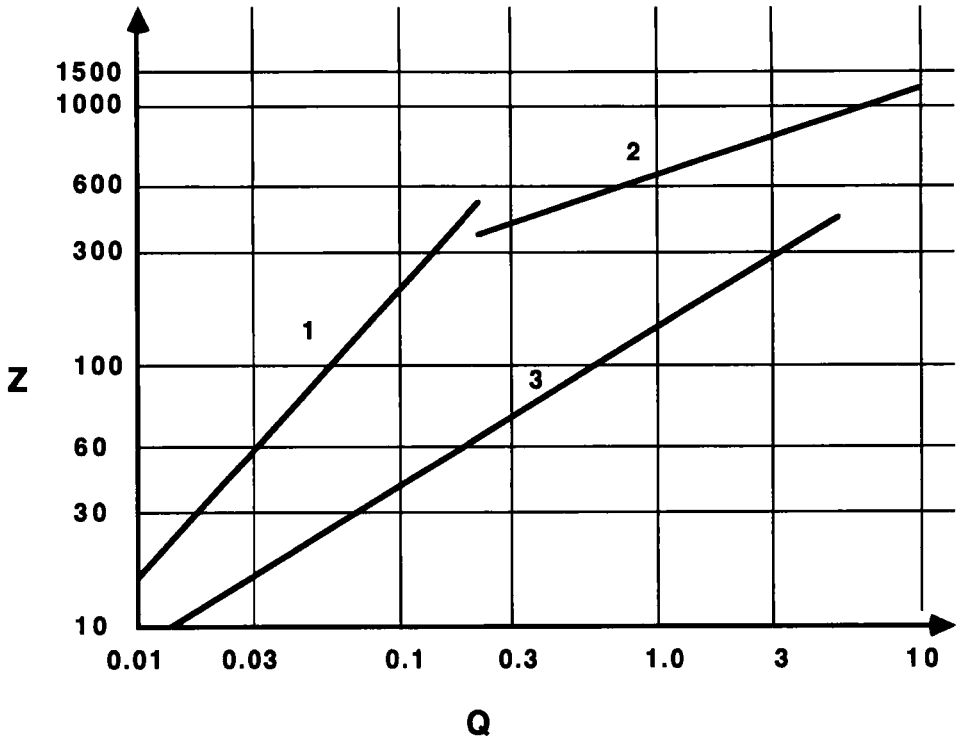

Figure 7.14. Plot for determination of flooding point. (1) Grids. (2) Stacked rings. (3) Random packing of rings.

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. 7.14, is found ($l_w/d_h^{2/3} * 10^3$). $d_h^{2/3}$ is shown in Table 7.4 for different packing materials.
2. l_w and d_h must be chosen so, that l_w is greater than 0.08 $m^3/m/h$ for common packing including raschig rings less than 7.5 cm, and greater than 0.12 m^3/h for raschig rings larger than 7.5 cm.
3. Generally, 0.4 $m^3/m/h$ can be considered as the upper limit for all types of packing.

4.Based on equation (7.39) and the total flows (kg/m^2) it is possible to find the area of cross-section of the absorption stripping tower.

Table 7.4
Characteristic packing data

Packing	Dimensions (inch) Diam.	Height	Thick-ness	Number per m^3	Surface m^2/m^3	Poro-sity (-)	dh=hydrau-lic. diam. (m)	$10^3 dh^{3/2}$	Gas flow entry tower (m/s)
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
"	-	1/2-1/4	-	-	144	0.40	0.011	1.15	0.13
Grids:									
"	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
Jagged grids:									
"	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
Stacked Raschig rings:									
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
Random packings of Raschig rings:									
Stoneware	3	3	3/8	1810	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

TABLE 7.4 (continued)

Packing	Dimensions (inch)			Number	Surface	Poro-	dh=hydrau-		Gas flow
			Thick-			sity	lic. diam.		entry tower
	Diam.	Height	ness	per m^3	m^2/m^3	(-)	(m)	$10^3 dh^{3/2}$	(m/s)
"	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
Berl-saddles:									
Stoneware	1/2	-	-	528000	460	0.65	0.0057	0.43	
"	1	-	-	81000	258	0.69	0.0107	1.10	not
"	1 1/2	-	-	22900	165	0.72	0.017	2.21	indicated
"	2	-	-	8000	120	0.72	0.024	3.72	

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

7. 5. Practical Experience

The best results in practice are achieved by use of countercurrent packed towers; see Ødegaard (1988). The water is distributed on the top of the packing with distribution trays or spray nozzles. For a high air to water ratio, a mist eliminator is necessary at the air outlet. Random packing of Raschig rings or saddles or grids, made of metal, ceramic, plastic or even impregnated wood, can be used.

Stripping ponds, see Fig. 7.15, might be used to remove 30-50% ammonia, but higher efficiencies can hardly be expected, even by introduction of agitation of the pond surface. It might, however, be practical to install stripping ponds as supplement to stripping tower to account for peak loadings.

Figures 7.16 and 7.17, taken from Fetting (1989), are constructed to facilitate the design in practice.

1. The operating temperature is selected for determination of Henry's constant; see equations 7.11 and 7.12.

2. The minimum ratio air to water, A/W can be derived from a simple mass balance

$$A/W = 1244*p*(1 - ef) / H \quad (7.41)$$

where p is the total pressure, ef is the required efficiency, i.e., the ratio between the concentration of ammonia in the effluent and in the influent. It can be recommended to multiply the minimum value of A/W by 1.2 -2.3 in practice.

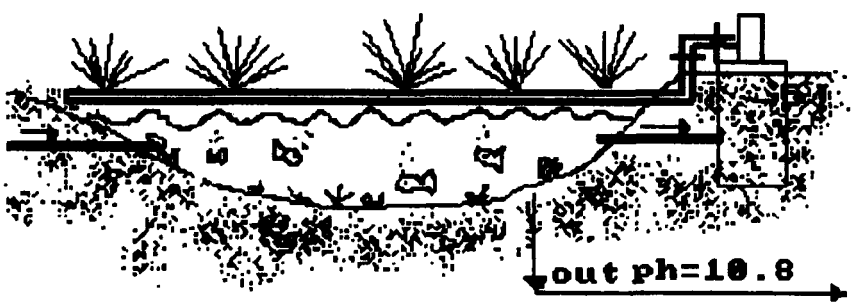

Figure 7.15. Ammonia stripping pond system. (Drawn by Morten V. Jørgensen).

3. The stripping factor R is found, based upon the selected A/W ratio, s:

$$R = h * s / 1244 \quad (7.42)$$

4. Figure 7.16 gives the number of transfer units, when R and the fraction removed are known. Figure 7.16. is valid for countercurrent operation, while Fig. 7.17 is constructed for single-stage cross-flow operation. Note that this latter figure uses the inverse stripping factor and the fraction remaining.

Due to the growing concern over air pollution problems, including the dry and wet deposition of nitrogen components as an increasing source of nutrients to

fresh and marine waters, it is necessary in most cases to combine the stripping unit with an absorption unit. The removed ammonia is absorbed in sulfuric acid for production of ammonium sulfate, which can be used as fertilizer. Figure 7.18 shows a flow chart of the combination of stripping and absorption.

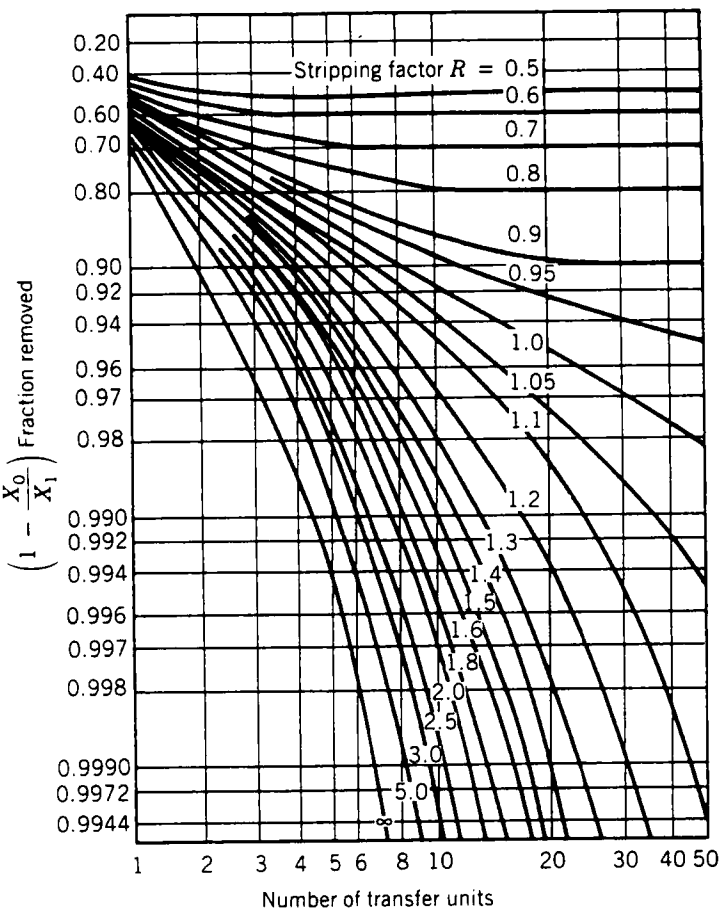

Figure 7.16. Number of transfer units for counter current operation as a function of removal efficiency and stripping factor, R. Reproduced from Fetting (1989).

High efficiency in ammonia removal requires adjustment of pH to about 11.0 before the stripping process. It implies that the pH after the stripping must be readjusted. The pH might drop about 0.2 by the stripping process due to removal of

ammonia, but a pH of 6-8 is required for the effluent.

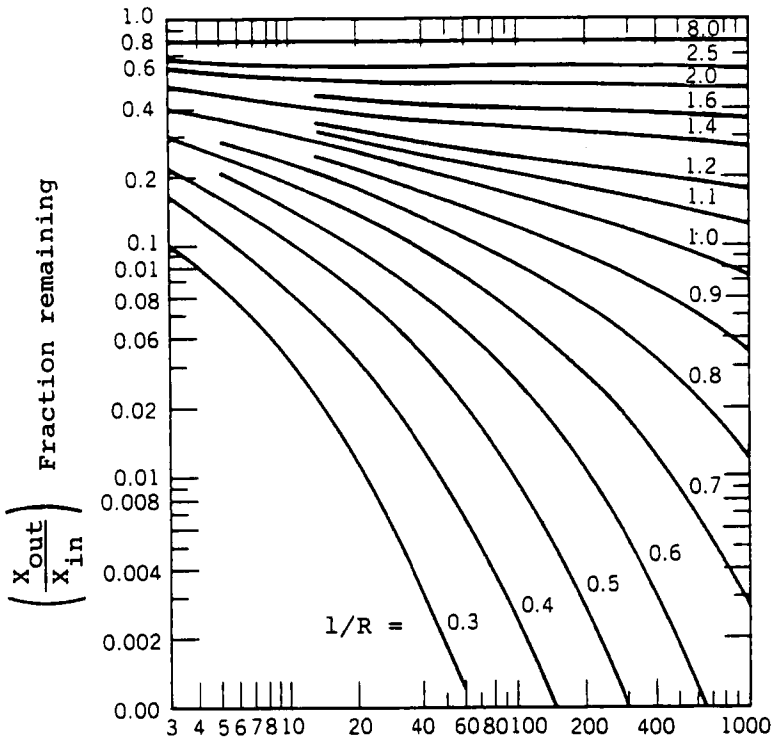

Figure 7.17. Number of transfer units for a single-stage cross-flow operation as a function of the concentration of ammonia remaining in water and of stripping factor, R. Reproduced from Fetting (1989).

The readjustment of pH can be carried out by recarbonization. Carbon dioxide is easily obtained from incineration of bio-gas, sludge or solid waste. Sulfuric acid might also be applied, but it is a less cost-effective alternative, which can only be recommended if there is no easy access to carbon dioxide.

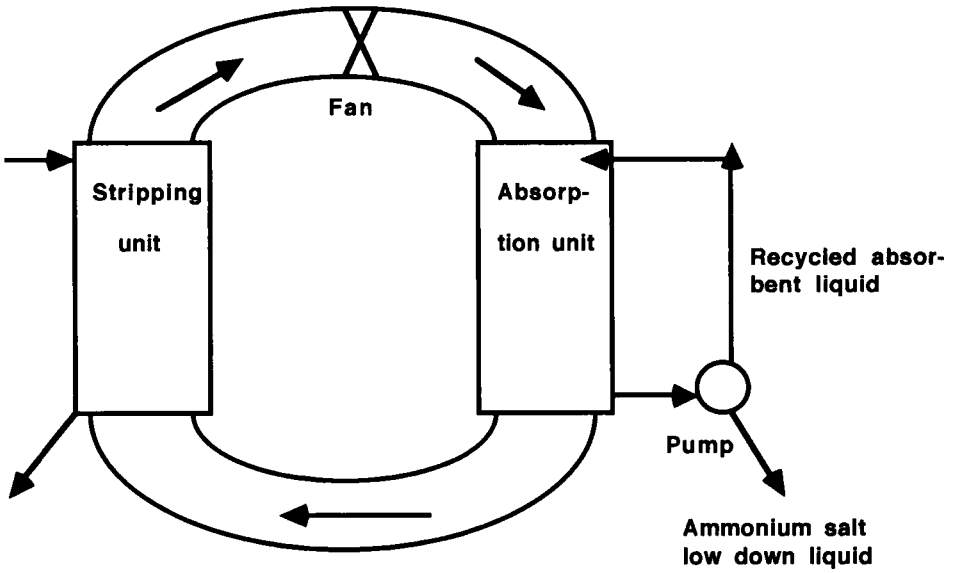

Figure 7.18. Process for stripping and recovery of ammonia.

7.6. Application of stripping

The stripping process is used to remove volatile gases such as hydrogen sulfide, hydrogen cyanide as well as ammonia. The removal of ammonia by stripping is used in the treatment of municipal waste water, where it has found very little application due to the problems mentioned in Section 7.2. Generally it can be concluded that the method is not economic in a temperate climate for large flows of waste water with relatively small concentrations of ammonia, as is found in municipal waste water. An additional problem is the air pollution caused by the removed ammonia, see Section 7.1. A recovery of ammonia by absorption in acid is possible, but the value of the recovered ammonia as ammonium sulfate is less than the costs of the recovery process.

The process has, however, found application at two well-known waste water treatment plants: at Lake Tahoe and in Pretoria. The flow chart of the latter plant is shown in Fig. 7.19. The main problem behind this solution is, however, not a pollution problem, but the scarcity of water.

If the concentration of ammonia is higher and the volume of waste water to

be treated smaller, the process becomes more favorable. This is for instance the case for the reject water, produced by dewatering of municipal sludge. The concentration here is 2-5 times higher than in municipal waste water and the process has therefore found some application for the treatment of this water particularly where the treatment plant is too small to handle the reject water in addition to the waste water.

Stripping has also been suggested for the treatment of industrial waste water and for the regeneration of the liquid used for eluting ion exchangers (Jørgensen, 1975). In these cases ammonia is removed from relatively small volumes and is present in high concentrations. As the amount of air needed is roughly independent of the ammonia concentration, see equation (7.41), the cost per kg of ammonia removed is much lower at high ammonia concentrations. The method therefore becomes much more attractive for industrial waste water with high ammonium concentrations or for recovery of elution liquids, used for regeneration of ion exchangers. Up to now stripping has not been used widely for treatment of industrial waste water, but with the growing demand for nitrogen removal, it is anticipated that the application of the method will increase in the coming decade.

Typical concentrations in waste water originating from production of ammonia, meat-bone-meal or fish meal are in the order of 500-1000 mg/l or 10-25 times higher than for municipal waste water. Elution liquids after regeneration of ion exchange columns may contain even higher ammonia concentrations and have already a high pH (see also Chapter 8).

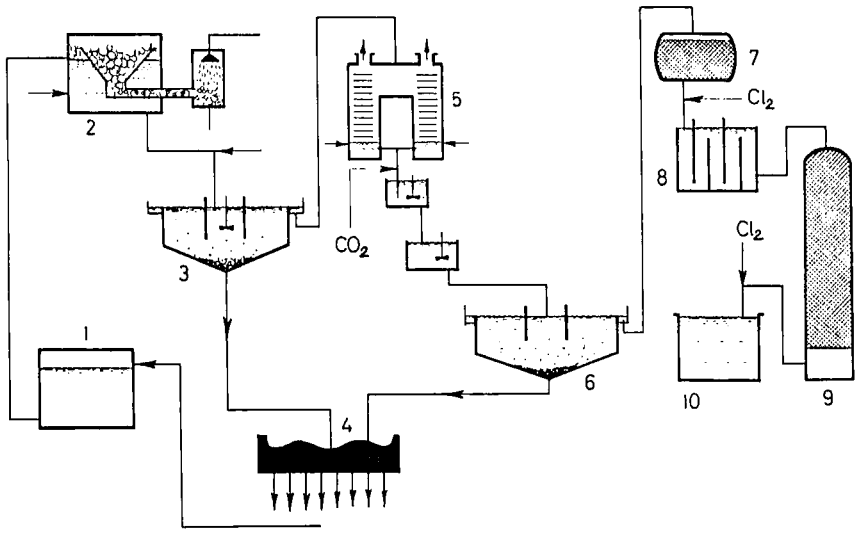

Figure 7.19. Waste water treatment plant, Pretoria. After mechanical-biological treatment (not shown) there follows 1) an algae pond, 2) aeration 3) lime precipitation 4) sludge drying 5) air stripping of ammonia 6) recarbonization 7) sand filtration 8) chlorination 9) adsorption on activated carbon 10) a second chlorination.