

Absorption of Gases

12.1. INTRODUCTION

The removal of one or more selected components from a mixture of gases by absorption into a suitable liquid is the second major operation of chemical engineering that is based on interphase mass transfer controlled largely by rates of diffusion. Thus, acetone can be recovered from an acetone–air mixture by passing the gas stream into water in which the acetone dissolves while the air passes out. Similarly, ammonia may be removed from an ammonia–air mixture by absorption in water. In each of these examples the process of absorption of the gas in the liquid may be treated as a physical process, the chemical reaction having no appreciable effect. When oxides of nitrogen are absorbed in water to give nitric acid, however, or when carbon dioxide is absorbed in a solution of sodium hydroxide, a chemical reaction occurs, the nature of which influences the actual rate of absorption. Absorption processes are therefore conveniently divided into two groups, those in which the process is solely physical and those where a chemical reaction is occurring. In considering the design of equipment to achieve gas absorption, the main requirement is that the gas should be brought into intimate contact with the liquid, and the effectiveness of the equipment will largely be determined by the success with which it promotes contact between the two phases. The general form of equipment is similar to that described for distillation in Chapter 11, and packed and plate towers are generally used for large installations. The method of operation, as will be seen later, is not the same. In absorption, the feed is a gas introduced at the bottom of the column, and the solvent is fed to the top, as a liquid; the absorbed gas and solvent leave at the bottom, and the unabsorbed components leave as gas from the top. The essential difference between distillation and absorption is that in the former the vapour has to be produced in each stage by partial vaporisation of the liquid which is therefore at its boiling point, whereas in absorption the liquid is well below its boiling point. In distillation there is a diffusion of molecules in both directions, so that for an ideal system equimolecular counterdiffusion takes place, though in absorption gas molecules are diffusing into the liquid, with negligible transfer in the reverse direction, as discussed in Volume 1, Chapter 10. In general, the ratio of the liquid to the gas flowrate is considerably greater in absorption than in distillation with the result that layout of the trays is different in the two cases. Furthermore, with the higher liquid rates in absorption, packed columns are much more commonly used.

12.2. CONDITIONS OF EQUILIBRIUM BETWEEN LIQUID AND GAS

When two phases are brought into contact they eventually reach equilibrium. Thus, water in contact with air evaporates until the air is saturated with water vapour, and the air is absorbed by the water until it becomes saturated with the individual gases. In any mixture of gases, the degree to which each gas is absorbed is determined by its partial pressure. At a given temperature and concentration, each dissolved gas exerts a definite partial pressure. Three types of gases may be considered from this aspect—a very soluble one, such as ammonia, a moderately soluble one, such as sulphur dioxide, and a slightly soluble one, such as oxygen. The values in Table 12.1 show the concentrations in kilograms per 1000 kg of water that are required to develop a partial pressure of 1.3, 6.7, 13.3, 26.7, and 66.7 kN/m² at 303 K. It may be seen that a slightly soluble gas requires a much higher partial pressure of the gas in contact with the liquid to give a solution of a given concentration. Conversely, with a very soluble gas a given concentration in the liquid phase is obtained with a lower partial pressure in the vapour phase. At 293 K a solution of 4 kg of sulphur dioxide per 1000 kg of water exerts a partial pressure of 2.7 kN/m². If a gas is in contact with this solution with a partial pressure SO₂ greater than 2.7 kN/m², sulphur dioxide will be absorbed. The most concentrated solution that can be obtained is that in which the partial pressure of the solute gas is equal to its partial pressure in the gas phase. These equilibrium conditions fix the limits of operation of an absorption unit. Thus, in an ammonia–air mixture containing 13.1 per cent of ammonia, the partial pressure of the ammonia is 13.3 kN/m² and the maximum concentration of the ammonia in the water at 303 K is 93 kg per 1000 kg of water.

Table 12.1. Partial pressures and concentrations of aqueous solutions of gases at 303 K

Partial pressure of solute in gas phase (kN/m ²)	Concentration of solute in water kg/1000 kg water		
	Ammonia	Sulphur dioxide	Oxygen
1.3	11	1.9	—
6.7	50	6.8	—
13.3	93	12	0.008
26.7	160	24.4	0.013
66.7	315	56	0.033

Whilst the solubility of a gas is not substantially affected by the total pressure in the system for pressures up to about 500 kN/m², it is important to note that the solubility falls with a rise of temperature. Thus, for a concentration of 25 per cent by mass of ammonia in water, the equilibrium partial pressure of the ammonia is 30.3 kN/m² at 293 K and 46.9 kN/m² at 303 K.

In many instances the absorption is accompanied by the evolution of heat, and it is therefore necessary to fit coolers to the equipment to keep the temperature sufficiently low for an adequate degree of absorption to be obtained.

For dilute concentrations of most gases, and over a wide range for some gases, the equilibrium relationship is given by Henry's law. This law, as used in Chapter 11, can be written as:

$$P_A = \mathcal{H}C_A \quad (12.1)$$

where: P_A is the partial pressure of the component A in the gas phase,
 C_A is the concentration of the component in the liquid, and
 \mathcal{H} is Henry's constant.

12.3. THE MECHANISM OF ABSORPTION

12.3.1. The two-film theory

The most useful concept of the process of absorption is given by the two-film theory due to WHITMAN⁽¹⁾, and this is explained fully in Volume 1, Chapter 10. According to this theory, material is transferred in the bulk of the phases by convection currents, and concentration differences are regarded as negligible except in the vicinity of the interface between the phases. On either side of this interface it is supposed that the currents die out and that there exists a thin film of fluid through which the transfer is effected solely by molecular diffusion. This film will be slightly thicker than the laminar sub-layer, because it offers a resistance equivalent to that of the whole boundary layer. According to Fick's law (Volume 1, equation 10.1) the rate of transfer by diffusion is proportional to the concentration gradient and to the area of interface over which the diffusion is occurring. Fick's law is limited to cases where the concentration of the absorbed component is low. At high concentrations, bulk flow occurs and the mass transfer rate, which is increased by a factor C_T/C_B , is governed by Stefan's law, equation 12.2. Under these circumstances, the concentration gradient is no longer constant throughout the film and the lines AB and DE are curved. This question has been discussed in Chapter 10 of Volume 1, but some of the important features will be given here.

The direction of transfer of material across the interface is not dependent solely on the concentration difference, but also on the equilibrium relationship. Thus, for a mixture of ammonia or hydrogen chloride and air which is in equilibrium with an aqueous solution, the concentration in the water is many times greater than that in the air. There is, therefore, a very large concentration gradient across the interface, although this is not the controlling factor in the mass transfer, as it is generally assumed that there is no resistance at the interface itself, where equilibrium conditions will exist. The controlling factor will be the rate of diffusion through the two films where all the resistance is considered to lie. The change in concentration of a component through the gas and liquid phases is illustrated in Figure 12.1. P_{AG} represents the partial pressure in the bulk of the gas phase and P_{Ai} the partial pressure at the interface. C_{AL} is the concentration in the bulk of the liquid phase and C_{Ai} the concentration at the interface. Thus, according to this theory, the concentrations at the interface are in equilibrium, and the resistance to transfer is centred in the thin films on either side. This type of problem is encountered in heat transfer across a tube, where the main resistance to transfer is shown to lie in the thin films on either side of the wall; here the transfer is by conduction.

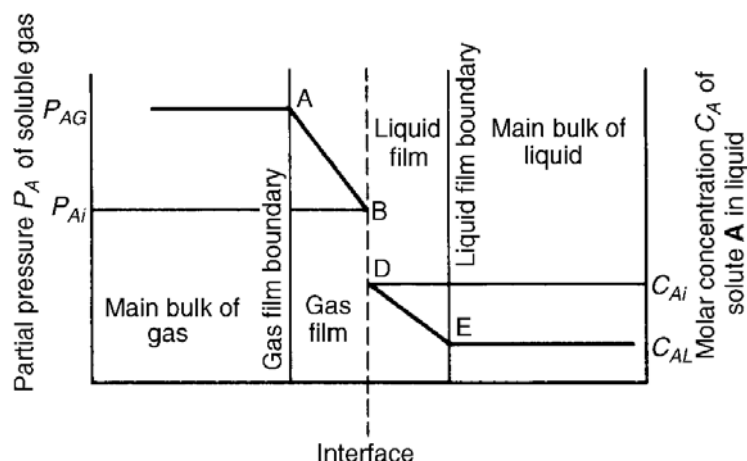


Figure 12.1. Concentration profile for absorbed component A

12.3.2. Application of mass transfer theories

The preceding analysis of the process of absorption is based on the two-film theory of WHITMAN⁽¹⁾. It is supposed that the two films have negligible capacity, but offer all the resistance to mass transfer. Any turbulence disappears at the interface or free surface, and the flow is thus considered to be laminar and parallel to the surface.

An alternative theory described in detail in Volume 1, Chapter 10, has been put forward by HIGBIE⁽²⁾, and later extended by DANCKWERTS⁽³⁾ and DANCKWERTS and KENNEDY⁽⁴⁾ in which the liquid surface is considered to be composed of a large number of small elements each of which is exposed to the gas phase for an interval of time, after which they are replaced by fresh elements arising from the bulk of the liquid.

All three of these proposals give the mass transfer rate N'_A directly proportional to the concentration difference ($C_{Ai} - C_{AL}$) so that they do not directly enable a decision to be made between the theories. However, in the Higbie–Danckwerts theory $N'_A \propto \sqrt{D_L}$ whereas $N'_A \propto D_L$ in the two-film theory. DANCKWERTS⁽³⁾ applied this theory to the problem of absorption coupled with chemical reaction but, although in this case the three proposals give somewhat different results, it has not been possible to distinguish between them.

The application of the penetration theory to the interpretation of experimental results obtained in wetted-wall columns has been studied by LYNN, STRAATMEIER, and KRAMERS⁽⁵⁾. They absorbed pure sulphur dioxide in water and various aqueous solutions of salts and found that, in the presence of a trace of Teepol which suppressed ripple formation, the rate of absorption was closely predicted by the theory. In very short columns, however, the rate was overestimated because of the formation of a region in which the surface was stagnant over the bottom one centimetre length of column. The studies were extended to columns containing spheres and again the penetration theory was found to hold, there being very little mixing of the surface layers with the bulk of the fluid as it flowed from one layer of spheres to the next.

Absorption experiments in columns packed with spheres, 37.8 mm diameter, were also carried out by DAVIDSON *et al.*⁽⁶⁾ who absorbed pure carbon dioxide into water. When a small amount of surface active agent was present in the water no appreciable mixing was

found between the layers of spheres. With pure water, however, the liquid was almost completely mixed in this region.

DAVIDSON⁽⁷⁾ built up theoretical models of the surfaces existing in a packed bed, and assumed that the liquid ran down each surface in laminar flow and was then fully mixed before it commenced to run down the next surface. The angles of inclination of the surfaces were taken as random. In the first theory it was assumed that all the surfaces were of equal length, and in the second that there was a random distribution of surface lengths up to a maximum. Thus the assumptions regarding age distribution of the liquid surfaces were similar to those of HIGBIE⁽²⁾ and DANCKWERTS⁽³⁾. Experimental results were in good agreement with the second theory. All random packings of a given size appeared to be equivalent to a series of sloping surfaces, and therefore the most effective packing would be that which gave the largest interfacial area.

In an attempt to test the surface renewal theory of gas absorption, DANCKWERTS and KENNEDY⁽⁸⁾ measured the transient rate of absorption of carbon dioxide into various solutions by means of a rotating drum which carried a film of liquid through the gas. Results so obtained were compared with those for absorption in a packed column and it was shown that exposure times of at least one second were required to give a strict comparison; this was longer than could be obtained with the rotating drum. ROBERTS and DANCKWERTS⁽⁹⁾ therefore used a wetted-wall column to extend the times of contact up to 1.3 s. The column was carefully designed to eliminate entry and exit effects and the formation of ripples. The experimental results and conclusions are reported by DANCKWERTS, KENNEDY, and ROBERTS⁽¹⁰⁾ who showed that they could be used, on the basis of the penetration theory model, to predict the performance of a packed column to within about 10 per cent.

There have been many recent studies of the mechanism of mass transfer in a gas absorption system. Many of these have been directed towards investigating whether there is a significant resistance to mass transfer at the interface itself. In order to obtain results which can readily be interpreted, it is essential to operate with a system of simple geometry. For that reason a laminar jet has been used by a number of workers.

CULLEN and DAVIDSON⁽¹¹⁾ studied the absorption of carbon dioxide into a laminar jet of water. When the water issued with a uniform velocity over the cross-section, the measured rate of absorption corresponded closely with the theoretical value. When the velocity profile in the water was parabolic, the measured rate was lower than the calculated value; this was attributed to a hydrodynamic entry effect.

The possible existence of an interface resistance in mass transfer has been examined by RAIMONDI and TOOR⁽¹²⁾ who absorbed carbon dioxide into a laminar jet of water with a flat velocity profile, using contact times down to 1 ms. They found that the rate of absorption was not more than 4 per cent less than that predicted on the assumption of instantaneous saturation of the surface layers of liquid. Thus, the effects of interfacial resistance could not have been significant. When the jet was formed at the outlet of a long capillary tube so that a parabolic velocity profile was established, absorption rates were lower than predicted because of the reduced surface velocity. The presence of surface-active agents appeared to cause an interfacial resistance, although this effect is probably attributable to a modification of the hydrodynamic pattern.

STERNLING and SCRIVEN⁽¹³⁾ have examined interfacial phenomena in gas absorption and have explained the interfacial turbulence which has been noted by a number of workers in

terms of the Marangoni effect which gives rise to movement at the interface due to local variations in interfacial tension. Some systems have been shown to give rise to stable interfaces when the solute is transferred in one direction, although instabilities develop during transfer in the reverse direction.

GOODRIDGE and ROBB⁽¹⁴⁾ used a laminar jet to study the rate of absorption of carbon dioxide into sodium carbonate solutions containing a number of additives including glycerol, sucrose, glucose, and arsenites. For the short times of exposure used, absorption rates into sodium carbonate solution or aqueous glycerol corresponded to those predicted on the basis of pure physical absorption. In the presence of the additives, however, the process was accelerated as the result of chemical reaction.

Absorption of gases and vapour by drops has been studied by GARNER and KENDRICK⁽¹⁵⁾ and GARNER and LANE⁽¹⁶⁾ who developed a vertical wind tunnel in which drops could be suspended for considerable periods of time in the rising gas stream. During the formation of each drop the rate of mass transfer was very high because of the high initial turbulence. After the initial turbulence had subsided, the mass transfer rate approached the rate for molecular diffusion provided that the circulation had stopped completely. In a drop with stable natural circulation the rate was found to approach 2.5 times the rate for molecular diffusion.

12.3.3. Diffusion through a stagnant gas

The process of absorption may be regarded as the diffusion of a soluble gas **A** into a liquid. The molecules of **A** have to diffuse through a stagnant gas film and then through a stagnant liquid film before entering the main bulk of liquid. The absorption of a gas consisting of a soluble component **A** and an insoluble component **B** is a problem of mass transfer through a stationary gas to which Stefan's law (Volume 1, Chapter 10) applies:

$$N'_A = -D_V \frac{C_T}{C_B} \frac{dC_A}{dz} \quad (12.2)$$

where N'_A is the overall rate of mass transfer (moles/unit area and unit time),
 D_V is the gas-phase diffusivity,
 z is distance in the direction of mass transfer, and
 C_A , C_B , and C_T are the molar concentrations of **A**, **B**, and total gas, respectively.

Integrating over the whole thickness z_G of the film, and representing concentrations at each side of the interface by suffixes 1 and 2:

$$N'_A = \frac{D_V C_T}{z_G} \ln \frac{C_{B2}}{C_{B1}} \quad (12.3)$$

Since $C_T = P/RT$, where **R** is the gas constant, T the absolute temperature, and P the total pressure. For an ideal gas, then:

$$N'_A = \frac{D_V P}{RT z_G} \ln \frac{P_{B2}}{P_{B1}} \quad (12.4)$$

Writing P_{Bm} as the log mean of the partial pressures P_{B1} and P_{B2} , then:

$$P_{Bm} = \frac{P_{B2} - P_{B1}}{\ln(P_{B2}/P_{B1})} \quad (12.5)$$

$$\begin{aligned} N'_A &= \frac{D_V P}{RT z_G} \frac{P_{B2} - P_{B1}}{P_{Bm}} \\ &= \frac{D_V P}{RT z_G} \left[\frac{P_{A1} - P_{A2}}{P_{Bm}} \right] \end{aligned} \quad (12.6)$$

Hence the rate of absorption of **A** per unit time per unit area is given by:

$$N'_A = k'_G P \left[\frac{P_{A1} - P_{A2}}{P_{Bm}} \right] \quad (12.7)$$

or:
$$N'_A = k_G (P_{A1} - P_{A2}) \quad (12.8)$$

where:
$$k'_G = \frac{D_V}{RT z_G}, \quad \text{and} \quad k_G = \frac{D_V P}{RT z_G P_{Bm}} = \frac{k'_G P}{P_{Bm}} \quad (12.9)$$

In the great majority of industrial processes the film thickness is not known, so that the rate equation of immediate use is equation 12.8 using k_G . k_G is known as the gas-film transfer coefficient for absorption and is a direct measure of the rate of absorption per unit area of interface with a driving force of unit partial pressure difference.

12.3.4. Diffusion in the liquid phase

The rate of diffusion in liquids is much slower than in gases, and mixtures of liquids may take a long time to reach equilibrium unless agitated. This is partly due to the much closer spacing of the molecules, as a result of which the molecular attractions are more important.

Whilst there is at present no theoretical basis for the rate of diffusion in liquids comparable with the kinetic theory for gases, the basic equation is taken as similar to that for gases, or for dilute concentrations:

$$N'_A = -D_L \frac{dC_A}{dz} \quad (12.10)$$

On integration:
$$N'_A = -D_L \left[\frac{C_{A2} - C_{A1}}{z_L} \right] \quad (12.11)$$

where: C_A, C_B are the molar concentrations of **A** and **B**,
 z_L is the thickness of liquid film through which diffusion occurs, and
 D_L is the diffusivity in the liquid phase.

Since the film thickness is rarely known, equation 12.11 is usually rewritten as:

$$N'_A = k_L (C_{A1} - C_{A2}) \quad (12.12)$$

which is similar to equation 12.8 for gases.

In equation 12.12, k_L is the liquid-film transfer coefficient, which is usually expressed in $\text{kmol/s m}^2(\text{kmol/m}^3) = \text{m/s}$. For dilute concentrations:

$$k_L = \frac{D_L}{z_L}$$

12.3.5. Rate of absorption

In a steady-state process of absorption, the rate of transfer of material through the gas film will be the same as that through the liquid film, and the general equation for mass transfer of a component **A** may be written as:

$$N'_A = k_G(P_{AG} - P_{Ai}) = k_L(C_{Ai} - C_{AL}) \tag{12.13}$$

where P_{AG} is the partial pressure in the bulk of the gas, C_{AL} is the concentration in the bulk of the liquid, and P_{Ai} and C_{Ai} are the values of concentration at the interface where equilibrium conditions are assumed to exist. Therefore:

$$\frac{k_G}{k_L} = \frac{C_{Ai} - C_{AL}}{P_{AG} - P_{Ai}} \tag{12.14}$$

These conditions may be illustrated graphically as in Figure 12.2, where ABF is the equilibrium curve for the soluble component **A**.

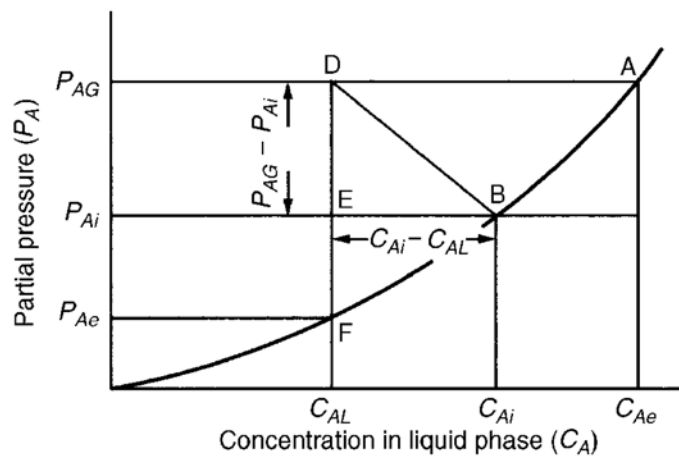


Figure 12.2. Driving forces in the gas and liquid phases

- Point D (C_{AL}, P_{AG}) represents conditions in the bulk of the gas and liquid.
 - P_{AG}) is the partial pressure of **A** in the main bulk of the gas stream, and
 - C_{AL}) is the average concentration of **A** in the main bulk of the liquid stream.
- Point A (C_{Ae}, P_{AG}) represents a concentration of C_{Ae} in the liquid in equilibrium with P_{AG} in the gas.

Point B (C_{Ai} , P_{Ai}) represents the concentration of C_{Ai} in the liquid in equilibrium with P_{Ai} in the gas, and gives conditions at the interface.

Point F (C_{AL} , P_{Ae}) represents a partial pressure P_{Ae} in the gas phase in equilibrium with C_{AL} in the liquid.

Then, the driving force causing transfer in the gas phase is:

$$(P_{AG} - P_{Ai}) \equiv DE$$

and the driving force causing transfer in the liquid phase is:

$$(C_{Ai} - C_{AL}) \equiv BE$$

Then:

$$\frac{P_{AG} - P_{Ai}}{C_{Ai} - C_{AL}} = \frac{k_L}{k_G}$$

and the concentrations at the interface (point B) are found by drawing a line through D of slope $-k_L/k_G$ to cut the equilibrium curve in B.

Overall coefficients

In order to obtain a direct measurement of the values of k_L and k_G the measurement of the concentration at the interface would be necessary. These values can only be obtained in very special circumstances, and it has been found of considerable value to use two overall coefficients K_G and K_L defined by:

$$N'_A = K_G(P_{AG} - P_{Ae}) = K_L(C_{Ae} - C_{AL}) \quad (12.15)$$

K_G and K_L are known as the overall gas and liquid phase coefficients, respectively.

Relation between film and overall coefficients

The rate of transfer of **A** may now be written as:

$$N'_A = k_G[P_{AG} - P_{Ai}] = k_L[C_{Ai} - C_{AL}] = K_G[P_{AG} - P_{Ae}] = K_L[C_{Ae} - C_{AL}]$$

Thus:

$$\begin{aligned} \frac{1}{K_G} &= \frac{1}{k_G} \left[\frac{P_{AG} - P_{Ae}}{P_{AG} - P_{Ai}} \right] \\ &= \frac{1}{k_G} \left[\frac{P_{AG} - P_{Ai}}{P_{AG} - P_{Ai}} \right] + \frac{1}{k_G} \left[\frac{P_{Ai} - P_{Ae}}{P_{AG} - P_{Ai}} \right] \end{aligned} \quad (12.16)$$

From the previous discussion:

$$\frac{1}{k_G} = \frac{1}{k_L} \left[\frac{P_{AG} - P_{Ai}}{C_{Ai} - C_{AL}} \right]$$

$$\begin{aligned} \text{Thus:} \quad \frac{1}{K_G} &= \frac{1}{k_G} + \frac{1}{k_L} \left[\frac{P_{AG} - P_{Ai}}{C_{Ai} - C_{AL}} \right] \left[\frac{P_{Ai} - P_{Ae}}{P_{AG} - P_{Ai}} \right] \\ &= \frac{1}{k_G} + \frac{1}{k_L} \left[\frac{P_{Ai} - P_{Ae}}{C_{Ai} - C_{AL}} \right] \end{aligned}$$

$(P_{Ai} - P_{Ae})/(C_{Ai} - C_{AL})$ is the average slope of the equilibrium curve and, when the solution obeys Henry's law, $\mathcal{H} = dP_A/dC_A \approx (P_{Ai} - P_{Ae})/(C_{Ai} - C_{AL})$.

$$\text{Therefore:} \quad \frac{1}{K_G} = \frac{1}{k_G} + \frac{\mathcal{H}}{k_L} \quad (12.17)$$

$$\text{Similarly:} \quad \frac{1}{K_L} = \frac{1}{k_L} + \frac{1}{\mathcal{H}k_G} \quad (12.18)$$

$$\text{and:} \quad \frac{1}{K_G} = \frac{\mathcal{H}}{K_L} \quad (12.19)$$

A more detailed discussion of the relationship between film and overall coefficients is given in Volume 1, Chapter 10.

The validity of using equations 12.17 and 12.18 in order to obtain an overall transfer coefficient has been examined in detail by KING⁽¹⁷⁾. He has pointed out that the equilibrium constant \mathcal{H} must be constant, there must be no significant interfacial resistance, and there must be no interdependence of the values of the two film-coefficients.

Rates of absorption in terms of mole fractions

The mass transfer equations can be written as:

$$N'_A = k''_G(y_A - y_{Ai}) = K''_G(y_A - y_{Ae}) \quad (12.20)$$

$$\text{and:} \quad N'_A = k''_L(x_{Ai} - x_A) = K''_L(x_{Ae} - x_A) \quad (12.21)$$

where x_A , y_A are the mole fractions of the soluble component **A** in the liquid and gas phases, respectively.

k''_G , k''_L , K''_G , K''_L are transfer coefficients defined in terms of mole fractions by equations 12.20 and 12.21.

If m is the slope of the equilibrium curve [approximately $(y_{Ai} - y_{Ae})/(x_{Ai} - x_A)$], it can then be shown that:

$$\frac{1}{K''_G} = \frac{1}{k''_G} + \frac{m}{k''_L} \quad (12.22)$$

which is similar to equation 11.151 used for distillation.

Factors influencing the transfer coefficient

The influence of the solubility of the gas on the shape of the equilibrium curve, and the effect on the film and overall coefficients, may be seen by considering three cases in turn — very soluble, almost insoluble, and moderately soluble gases.

(a) *Very soluble gas*. Here the equilibrium curve lies close to the concentration-axis and the points E and F are very close to one another as shown in Figure 12.2. The driving force over the gas film (DE) is then approximately equal to the overall driving force (DF), so that k_G is approximately equal to K_G .

(b) *Almost insoluble gas*. Here the equilibrium curve rises very steeply so that the driving force ($C_{Ai} - C_{AL}$) (EB) in the liquid film becomes approximately equal to the overall driving force ($C_{Ae} - C_{AL}$) (AD). In this case k_L will be approximately equal to K_L .

(c) *Moderately soluble gas*. Here both films offer an appreciable resistance, and the point B at the interface must be located by drawing a line through D of slope $-(k_L/k_G) = -(P_{AG} - P_{Ai})/(C_{Ai} - C_{AL})$.

In most experimental work, the concentration at the interface cannot be measured directly, and only the overall coefficients are therefore found. To obtain values for the film coefficients, the relations between k_G , k_L and K_G are utilised as discussed previously.

12.4. DETERMINATION OF TRANSFER COEFFICIENTS

In the design of an absorption tower, the most important single factor is the value of the transfer coefficient or the height of the transfer unit. Whilst the total flowrates of the gas and liquid streams are fixed by the process, it is necessary to determine the most suitable flow per unit area through the column. The gas flow is limited by the fact that the flooding rate must not be exceeded and there will be a serious drop in performance if the liquid rate is very low. It is convenient to examine the effects of flowrates of the gas and liquid on the transfer coefficients, and also to investigate the influence of variables such as temperature, pressure, and diffusivity.

In the laboratory, wetted-wall columns have been used by a number of workers and they have proved valuable in determining the importance of the various factors, and have served as a basis from which correlations have been developed for packed towers.

12.4.1. Wetted-wall columns

In many early studies, the rate of vaporisation of liquids into an air stream was measured in a wetted-wall column, similar to that shown in Figure 12.3. Logarithmic plots of d/z_G and $Re = du\rho/\mu$ gave a series of approximately straight lines and d/z_G was proportional to $Re^{0.83}$

where: d is the diameter of tube,

z_G is the thickness of gas film,

u is the gas velocity,

ρ is the gas density,

μ is the gas viscosity, and

B is a constant.

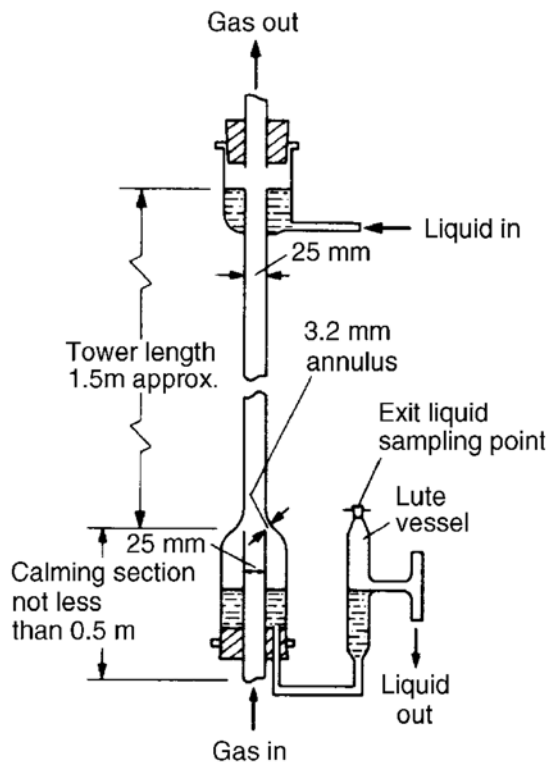


Figure 12.3. Diagram of a typical laboratory wetted-wall column

The unknown film thickness z_G may be eliminated as follows:

$$k_G = \frac{D_V P}{RT z_G P_{Bm}} \quad \text{(equation 12.9)}$$

Thus:

$$\frac{k_G RT P_{Bm}}{D_V P} = \frac{1}{z_G} = \frac{B}{d} Re^{0.83}$$

or:

$$\frac{h_D d P_{Bm}}{D_V P} = B Re^{0.83} \quad \text{(12.23)}$$

where $h_D = k_G RT$ is the mass transfer coefficient with the driving force expressed as a molar concentration difference.

GILLILAND and SHERWOOD's data⁽¹⁸⁾, expressed by equation 12.23, are shown in Figure 12.4 for a number of systems. To allow for the variation in the physical properties, the Schmidt Group Sc is introduced, and the general equation for mass transfer in a wetted-wall column is then given by:

$$\frac{h_D d P_{Bm}}{D_V P} = B' Re^{0.83} Sc^{0.44} \quad \text{(12.24)}$$

Values of B' 0.021–0.027 have been reported and a mean value of 0.023 may be taken, which means that equation 12.24 very similar to the general heat transfer equation for forced convection in tubes (Volume 1, Chapter 9). The data shown in Figure 12.4 are

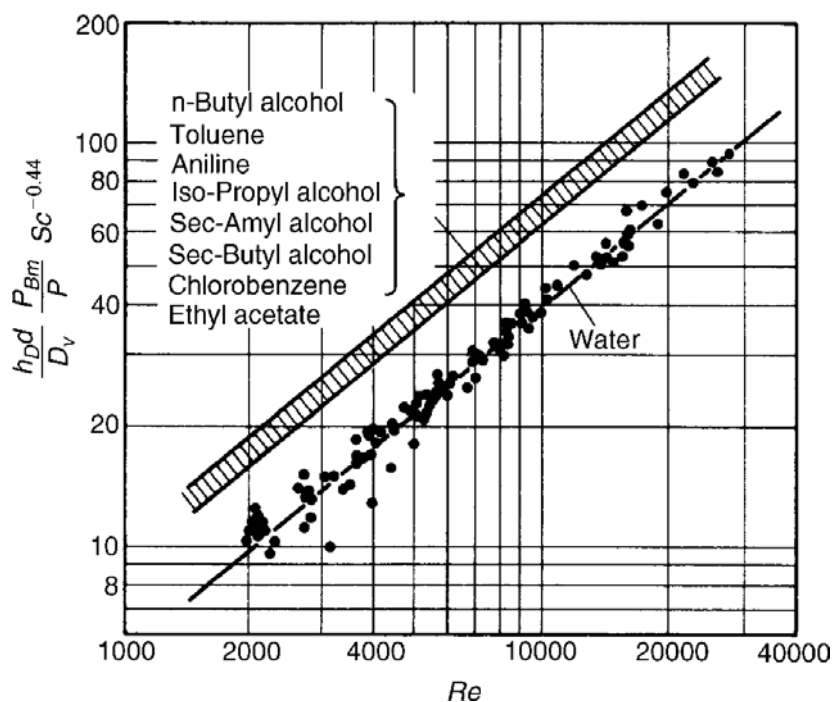


Figure 12.4. Vaporisation of liquids in a wetted-wall column

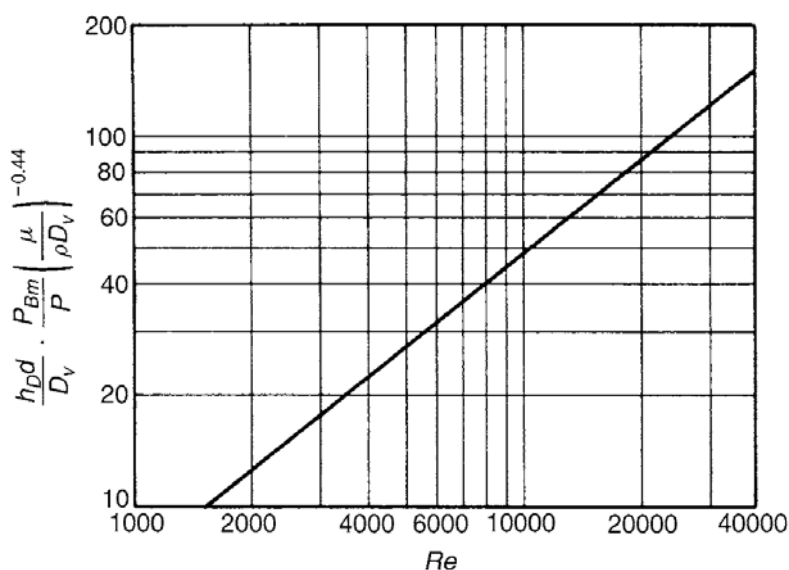


Figure 12.5. Correlation of data on the vaporisation of liquids in wetted-wall columns

replotted as $(h_{Dd}/D_V)(P_{Bm}/P)Sc^{-0.44}$ in Figure 12.5 and, in this way, they may be correlated by means of a single line.

In comparing the results of various workers, it is important to ensure that the inlet arrangements for the air are similar. Modifications of the inlet give rise to various values for the index on the Reynolds number, as found by HOLLINGS and SILVER⁽¹⁹⁾. A good calming length is necessary before the inlet to the measuring section, if the results are to be reproducible.

Equation 12.24 is frequently rearranged as:

$$\frac{h_D d}{D_V} \frac{P_{Bm}}{P} \frac{\mu}{du \rho} \left[\frac{\mu}{\rho D_V} \right]^{0.56} = B' Re^{-0.17} \left(\frac{\mu}{\rho D_V} \right)$$

or:

$$\frac{h_D}{u} \frac{P_{Bm}}{P} \left[\frac{\mu}{\rho D_V} \right]^{0.56} = B' Re^{-0.17} = j_d \quad (12.25)$$

where j_d is the j -factor for mass transfer as introduced by CHILTON and COLBURN⁽²⁰⁾ and discussed in Volume 1, Chapter 10. The main feature of this type of work is that $h_D \propto G^{0.8}$, $D_V^{0.56}$ and P/P_{Bm} . This form of relation is the basis for correlating data on packed towers.

Example 12.1

The overall liquid transfer coefficient, $K_L a$, for the absorption of SO₂ in water in a column is 0.003 kmol/s m³ (kmol/m³). By assuming an expression for the absorption of NH₃ in water at the same liquor rate and varying gas rates, derive an expression for the overall liquid film coefficient $K_L a$ for absorption of NH₃ in water in this equipment at the same water rate though with varying gas rates. The diffusivities of SO₂ and NH₃ in air at 273 K are 0.103 and 0.170 cm²/s. SO₂ dissolves in water, and Henry's constant is equal to 50 (kN/m²)/(kmol/m³). All data are expressed for the same temperature.

Solution

From equation 12.18:

$$\frac{1}{K_L a} = \frac{1}{k_L a} + \frac{1}{\mathcal{H} k_G a} = \frac{1}{0.003} = 333.3$$

For the absorption of a moderately soluble gas it is reasonable to assume that the liquid and gas phase resistances are of the same order of magnitude, assuming them to be equal.

$$\frac{1}{k_L a} = \frac{1}{\mathcal{H} k_G a} = \left(\frac{333}{2} \right) = 166.7$$

or:

$$k_L a = \mathcal{H} k_G a = 0.006 \text{ kmol/s m}^3 \text{ (kmol/m}^3\text{)}$$

Thus, for SO₂: $k_G a = 0.006/\mathcal{H} = 0.006/50 = 0.00012 \text{ kmol/s m}^3 \text{ (kN/m}^2\text{)}$

From equation 12.25: $k_G a$ is proportional to (diffusivity)^{0.56}.

Hence for NH₃ : $k_G a = 0.00012(0.17/0.103)^{0.56} = 0.00016 \text{ kmol/s m}^3 \text{ (kN/m}^2\text{)}$

For a very soluble gas such as NH₃, $k_G a \simeq K_G a$.

For NH₃ the liquid-film resistance will be small, and:

$$k_G a = K_G a = \underline{\underline{0.00016 \text{ kmol/s m}^3 \text{ (kN/m}^2\text{)}}$$

In early work on wetted-wall columns, MORRIS and JACKSON⁽²¹⁾ represented the experimental data for the mass transfer coefficient for the gas film h_D in a form similar to equation 12.25, though with slightly different indices, to give:

$$\frac{h_D}{u} = 0.04 \left[\frac{ud\rho}{\mu} \right]^{-0.25} \left[\frac{\mu}{\rho D_V} \right]^{-0.5} \left[\frac{P}{P_{Bm}} \right] \quad (12.26)$$

The velocity u of the gas is strictly the velocity relative to the surface of the falling liquid film, though little error is introduced if it is taken as the superficial velocity in the column.

Compounding of film coefficients

Assuming k_G is approximately proportional to $G^{0.8}$, equation 12.17 may be rearranged to give:

$$\frac{1}{K_G} = \frac{1}{k_G} + \frac{\mathcal{H}}{k_L} = \frac{1}{\psi u^{0.8}} + \frac{\mathcal{H}}{k_L} \quad (12.27)$$

If k_L is assumed to be independent of the gas velocity, then a plot of $1/K_G$ against $1/u^{0.8}$ will give a straight line with a positive intercept on the vertical axis representing the liquid film resistance \mathcal{H}/k_L , as shown for ammonia and for sulphur dioxide in Figure 12.6. It may be seen that in each case a straight line is obtained. The lines for ammonia pass almost through the origin showing that the liquid film resistance is very small, although the line for sulphur dioxide gives a large intercept on the vertical axis, indicating a high value of the liquid film resistance.

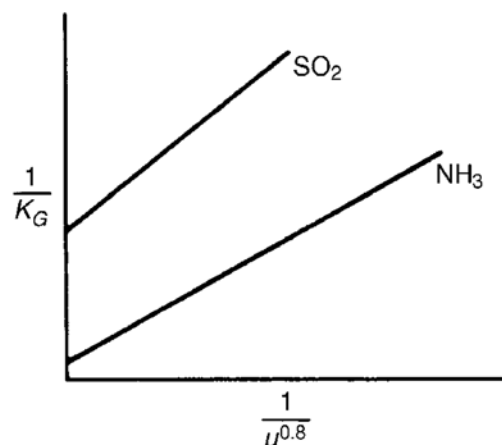


Figure 12.6. Plot of $1/K_G$ versus $1/u^{0.8}$ for ammonia and for sulphur dioxide

For a constant value of Re , the film thickness z_G should be independent of temperature, since $\mu/\rho D_V$ is almost independent of temperature. k_G will then vary as \sqrt{T} , because

$D_V \propto T^{3/2}$ and $k_G \propto D_V/T$. This is somewhat difficult to test accurately since the diffusivity in the liquid phase also depends on temperature. Thus, the data for sulphur dioxide, shown in Figure 12.6, qualitatively support the theory for different temperatures, although the increase in value of k_L masks the influence of temperature on k_G .

Example 12.2

A wetted-wall column is used for absorbing sulphur dioxide from air by means of a caustic soda solution. At an air flow of $2 \text{ kg/m}^2\text{s}$, corresponding to a Reynolds number of 5160, the friction factor $R/\rho u^2$ is 0.0200.

Calculate the mass transfer coefficient in $\text{kg SO}_2/\text{s m}^2 (\text{kN/m}^2)$ under these conditions if the tower is at atmospheric pressure. At the temperature of absorption the following values may be used:

The diffusion coefficient for $\text{SO}_2 = 0.116 \times 10^{-4} \text{ m}^2/\text{s}$, the viscosity of gas = 0.018 mNs/m^2 , and the density of gas stream = 1.154 kg/m^3 .

Solution

For wetted-wall columns, the data are correlated by:

$$\left(\frac{h_d}{u}\right) \left(\frac{P_{Bm}}{P}\right) \left(\frac{\mu}{\rho D}\right)^{0.56} = B' Re^{-0.17} = j_d \quad (\text{equation 12.25})$$

From Volume 1, Chapter 10: $j_d \simeq R/\rho u^2$

In this problem: $G' = 2.0 \text{ kg/m}^2\text{s}$, $Re = 5160$ and $R/\rho u^2 = 0.020$

$$D = 0.116 \times 10^{-4} \text{ m}^2/\text{s}, \quad \mu = 1.8 \times 10^{-5} \text{ Ns/m}^2, \quad \text{and } \rho = 1.154 \text{ kg/m}^3$$

Substituting these values gives:

$$\left(\frac{\mu}{\rho D}\right)^{0.56} = \left(\frac{1.8 \times 10^{-5}}{1.154 \times 0.116 \times 10^{-4}}\right)^{0.56} = 1.18$$

Thus:
$$\left(\frac{h_d}{u}\right) \left(\frac{P_{Bm}}{P}\right) = (0.020/1.18) = 0.0169$$

$$G' = \rho u = 2.0 \text{ kg/m}^2\text{s}$$

and:
$$u = (2.0/1.154) = 1.73 \text{ m/s}$$

Thus:
$$h_d(P_{Bm}/P) = (0.0169 \times 1.73) = 0.0293$$

d may be obtained from $d = Re\mu/\rho u = 0.046 \text{ m}$ (46 mm), which is the same order of size of wetted-wall column as that which was originally used in the research work.

$$k_G = \left(\frac{h_d}{RT}\right) \left(\frac{P_{Bm}}{P}\right)$$

$R = 8314 \text{ m}^3(\text{N/m}^2)/\text{K kmol}$ and T will be taken as 298 K, and hence:

$$k_G = [0.0293/(8314 \times 298)] = 1.18 \times 10^{-8} \text{ kmol/m}^2\text{s}(\text{N/m}^2)$$

$$= \underline{\underline{7.56 \times 10^{-4} \text{ kg SO}_2/\text{m}^2\text{s}(\text{kN/m}^2)}}$$

12.4.2. Coefficients in packed towers

The majority of published data on transfer coefficients in packed towers are for rather small laboratory units, and there is still some uncertainty in extending the data for use in industrial units. One of the great difficulties in correlating the performance of packed towers is the problem of assessing the effective wetted area for interphase transfer. It is convenient to consider separately conditions where the gas-film is controlling, and then where the liquid film is controlling. The general method of expressing results is based on that used for wetted-wall columns.

Gas-film controlled processes

The absorption of ammonia in water has been extensively studied by a number of workers. KOWALKE *et al.*⁽²²⁾ used a tower of 0.4 m internal diameter with a packing 1.2 m deep, and expressed their results as:

$$K_G a = \alpha G'^{0.8} \quad (12.28)$$

where K_G is expressed in $\text{kmol/s m}^2 (\text{kN/m}^2)$ and a is the interfacial surface per unit volume of tower (m^2/m^3). Thus $K_G a$ is a transfer coefficient based on unit volume of tower. G' is in kg/s m^2 , and varies with the nature of the packing and the liquid rate. It was noted that α increased with L' for values up to 1.1 kg/s m^2 , after which further increase gave no significant increase in $K_G a$. It was thought that the initial increase in the coefficient was occasioned by a more effective wetting of the packing. On increasing the liquid rate so that the column approached flooding conditions, it was found that $K_G a$ decreased. Other measurements by BORDEN and SQUIRES⁽²³⁾ and NORMAN⁽²⁴⁾ confirm the applicability of equation 12.28.

FELLINGER⁽²⁵⁾ used a 450 mm diameter column with downcomers and risers in an attempt to avoid the problem of determining any entrance or exit effects. Some of the results for H_{OG} are shown in Table 12.2, taken from Perry's Chemical Engineers' Handbook⁽²⁶⁾. Further discussion on the use of transfer units is included in Section 12.8.8 and in Chapter 11.

Table 12.2. Height of the transfer unit H_{OG} in metres

Raschig rings size (mm)	G' ($\text{kg/m}^2\text{s}$)	H_{OG} ($L' = 0.65 \text{ kg/m}^2\text{s}$)	H_{OG} ($L' = 1.95 \text{ kg/m}^2\text{s}$)
9.5	0.26	0.37	0.23
	0.78	0.60	0.32
25	0.26	0.40	0.22
	0.78	0.64	0.34
50	0.26	0.60	0.34
	0.78	1.04	0.58

MOLSTAD *et al.*⁽²⁷⁾ also measured the absorption of ammonia in water using a tower of 384 mm side packed with wood grids, or with rings or saddles, and obtained $K_G a$ by direct experiment. The value of $k_G a$ was then calculated from the following relation based on equation 12.17:

$$\frac{1}{K_G a} = \frac{1}{k_G a} + \frac{\mathcal{H}}{k_L a} \quad (12.29)$$

The simplest method of representing data for gas-film coefficients is to relate the Sherwood number $[(h_D d / D_V)(P_{Bm} / P)]$ to the Reynolds number (Re) and the Schmidt number $(\mu / \rho D_V)$. The indices used vary between investigators though VAN KREVELEN and HOFTIJZER⁽²⁸⁾ have given the following expression, which is claimed to be valid over a wide range of Reynolds numbers:

$$\frac{h_D d}{D_V} \frac{P_{Bm}}{P} = 0.2 Re^{0.8} \left(\frac{\mu}{\rho D_V} \right)^{0.33} \quad (12.30)$$

Later work suggests that 0.11 is a more realistic value for the coefficient.

SEMMELEBAUER⁽²⁹⁾ has recommended the following correlation for $100 < (Re)_G < 10,000$ and $0.01 \text{ m} < d_p < 0.05 \text{ m}$:

$$(Sh)_G = \beta (Re)_G^{0.59} (Sc)_G^{0.33} \quad (12.31)$$

where: $\beta = 0.69$ for Raschig rings and 0.86 for Berl saddles,

$$(Sh)_G = h_D d_p / D_G,$$

$$(Re)_G = G' d_p / \mu_G,$$

$$(Sc)_G = \mu_G / \rho_G D_G, \text{ and}$$

$$d_p = \text{packing size.}$$

Processes controlled by liquid-film resistance

The absorption of carbon dioxide, oxygen, and hydrogen in water are three examples in which most, if not all, of the resistance to transfer lies in the liquid phase. SHERWOOD and HOLLOWAY⁽³⁰⁾ measured values of $k_L a$ for these systems using a tower of 500 mm diameter packed with 37 mm rings. The results were expressed in the form:

$$\frac{k_L a}{D_L} = \beta \left[\frac{L'}{\mu_L} \right]^{0.75} \left[\frac{\mu_L}{\rho_L D_L} \right]^{0.50} \quad (12.32)$$

It may be noted that this equation has no term for characteristic length on the right-hand side and therefore it is not a dimensionally consistent equation. If values of $k_L a$ are plotted against value L' on logarithmic scales as shown in Figure 12.7, a slope of about 0.75 is obtained for values of L' 0.5–20 kg/s m². Beyond this value of L' , it was found that $k_L a$ tended to fall because the loading point for the column was reached. These values of $k_L a$ were found to be affected by the gas rate. Subsequently, COOPER *et al.*⁽³¹⁾ established that, at the high liquid rates and low gas rates used in practice, the transfer rates were much lower than given by equation 12.32. This was believed to be due to maldistribution at gas velocities as low as 0.03 m/s. The results of COOPER *et al.*⁽³¹⁾ and SHERWOOD and

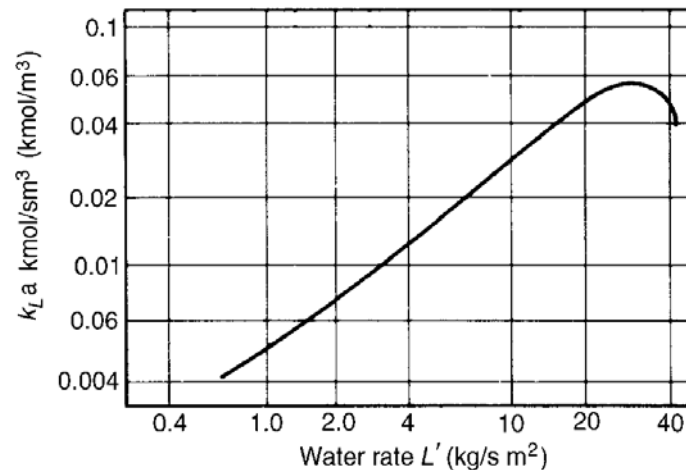


Figure 12.7. Variation of liquid-film coefficient with liquid flow for the absorption of oxygen in water

HOLLOWAY⁽³⁰⁾ are compared in Figure 12.8, where the height of the transfer unit H_{OL} is plotted against the liquid rate for various gas velocities.

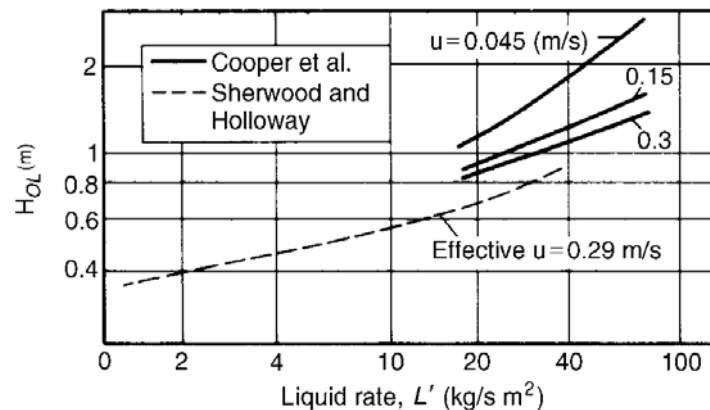


Figure 12.8. Effect of liquid rate on height of transfer unit H_{OL} . Comparison of the results of Sherwood and Holloway⁽³⁰⁾, and Cooper *et al.*⁽³¹⁾

In an equation similar to equation 12.31, SEMMELBAUER⁽²⁹⁾ produced the following correlation for the liquid film mass transfer coefficient k_L for $3 < Re_L < 3000$ and $0.01 \text{ m} < d_p < 0.05 \text{ m}$:

$$(Sh)_L = \beta' (Re)_L^{0.59} (Sc)_L^{0.5} (d_p^3 g \rho_L^2 / \mu_L^2)^{0.17} \quad (12.33)$$

where: $\beta' = 0.32$ and 0.25 for Raschig rings and Berl saddles, respectively.

$$(Sh)_L = k_L d_p / D_L,$$

$$(Re)_L = L' d_p / \mu_L, \text{ and}$$

$$(Sc)_L = \mu_L / \rho_L D_L.$$

NONHEBEL⁽³²⁾ emphasises that values of the individual film mass transfer coefficients obtained from this equation must be used with caution when designing large-scale towers and appropriately large safety factors should be incorporated.

12.4.3. Coefficients in spray towers

It is difficult to compare the performance of various spray towers since the type of spray distributor used influences the results. Data from HIXSON and SCOTT⁽³³⁾ and others show that $K_G a$ varies as $G'^{0.8}$, and is also affected by the liquid rate. More reliable data with spray columns might be expected if the liquid were introduced in the form of individual drops through a single jet into a tube full of gas. Unfortunately the drops tend to alter in size and shape and it is not possible to get the true interfacial area very accurately. This has been investigated by WHITMAN *et al.*⁽³⁴⁾, who found that k_G for the absorption of ammonia in water was about $0.035 \text{ kmol/s m}^2 \text{ (N/m}^2\text{)}$, compared with 0.00025 for the absorption of carbon dioxide in water.

Some values obtained by PIGFORD and PYLE⁽³⁵⁾ for the height of a transfer unit H_L for the stripping of oxygen from water are shown in Figure 12.9. For short heights, the efficiency of the spray chamber approximates closely to that of a packed tower although, for heights greater than 1.2 m, the efficiency of the spray tower drops off rather rapidly. Whilst it might be possible to obtain a very large active interface by producing small drops, in practice it is impossible to prevent these coalescing, and hence the effective interfacial surface falls off with height, and spray towers are not used extensively.

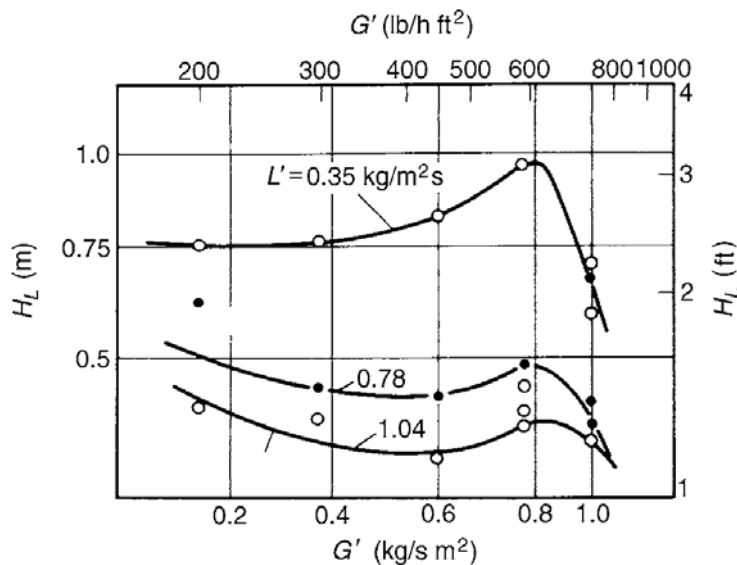


Figure 12.9. Height of the transfer unit H_L for stripping of oxygen from water in a spray tower

12.5. ABSORPTION ASSOCIATED WITH CHEMICAL REACTION

In the instances so far considered, the process of absorption of the gas in the liquid has been entirely a physical one. There are, however, a number of cases in which the gas, on absorption, reacts chemically with a component of the liquid phase⁽³⁶⁾. The topic of mass transfer accompanied by chemical reaction is treated in detail in Volume 1, Chapter 10.

In the absorption of carbon dioxide by caustic soda, the carbon dioxide reacts directly with the caustic soda and the process of mass transfer is thus made much more complicated. Again, when carbon dioxide is absorbed in an ethanolamine solution, there is direct chemical reaction between the amine and the gas. In such processes the conditions in the gas phase are similar to those already discussed, though in the liquid phase there is a liquid film followed by a reaction zone. The process of diffusion and chemical reaction may still be represented by an extension of the film theory by a method due to HATTA⁽³⁷⁾. In the case considered, the chemical reaction is irreversible and of the type in which a solute gas **A** is absorbed from a mixture by a substance **B** in the liquid phase, which combines with **A** according to the equation $\mathbf{A} + \mathbf{B} \rightarrow \mathbf{AB}$. As the gas approaches the liquid interface, it dissolves and reacts at once with **B**. The new product **AB**, thus formed, diffuses towards the main body of the liquid. The concentration of **B** at the interface falls; this results in diffusion of **B** from the bulk of the liquid phase to the interface. Since the chemical reaction is rapid, **B** is removed very quickly, so that it is necessary for the gas **A** to diffuse through part of the liquid film before meeting **B**. There is thus a zone of reaction between **A** and **B** which moves away from the gas-liquid interface, taking up some position towards the bulk of the liquid. The final position of this reaction zone will be such that the rate of diffusion of **A** from the gas-liquid interface is equal to the rate of diffusion of **B** from the main body of the liquid. When this condition has been reached, the concentrations of **A**, **B**, and **AB** may be indicated as shown in Figure 12.10, where the concentrations are shown as ordinates and the positions of a plane relative to the interface as abscissae. In this Figure, the plane of the interface between gas and liquid is shown by U, the reaction zone by R, and the outer boundary of liquid film by S. Then **A** diffuses through the gas film as a result of the driving force $(P_{AG} - P_{Ai})$ and diffuses to the reaction zone as a result of the driving force C_{Ai} in the liquid phase. The component **B** diffuses from the main body of the liquid to the reaction zone under a driving force q , and the non-volatile product **AB** diffuses back to the main bulk of the liquid under a driving force $(m - n)$.

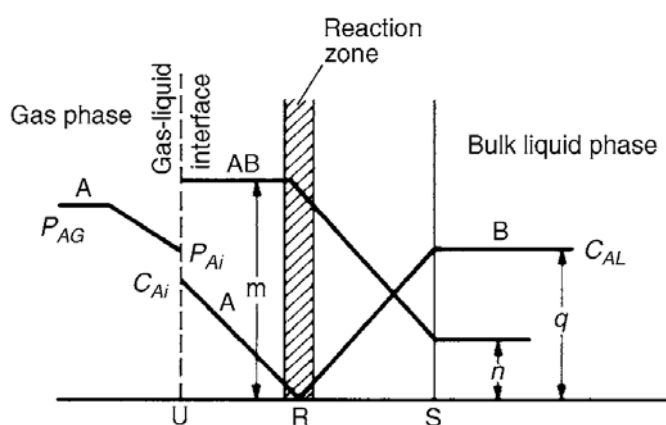


Figure 12.10. Concentration profile for absorption with chemical reaction

The difference between a physical absorption, and one in which a chemical reaction occurs, can also be shown by Figures 12.11a and 12.11b, taken from a paper by VAN KREVELEN and HOFTIJZER⁽²⁸⁾. Figure 12.11a shows the normal concentration profile for

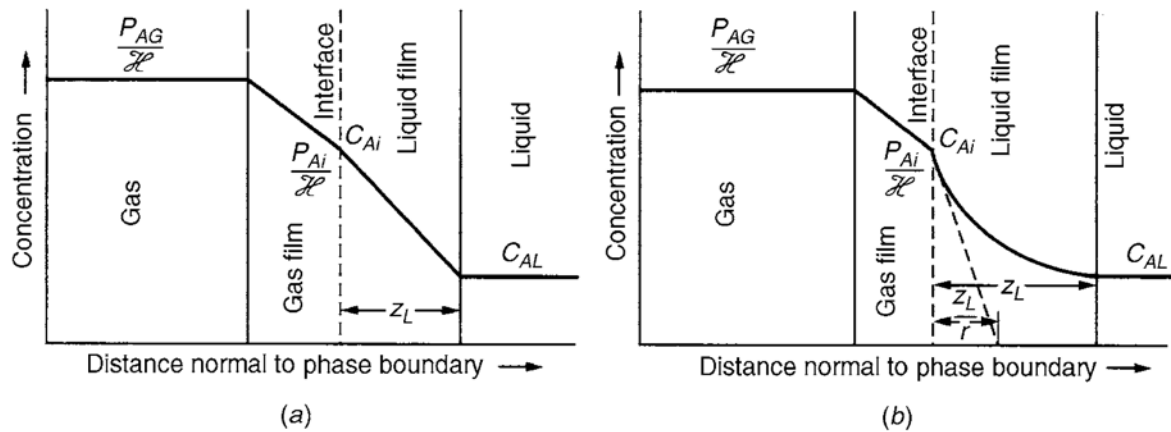


Figure 12.11. Concentration profiles for absorption (a) without chemical reaction, (b) with chemical reaction. The scales for concentration in the two phases are not the same and are chosen so that P_{Ai}/\mathcal{H} in the gas phase and C_{Ai} for the liquid phase are at the same position in the diagrams

physical absorption whilst Figure 12.11b shows the profile modified by the chemical reaction. For transfer in the gas phase:

$$N'_A = k_G(P_{AG} - P_{Ai}) \quad (12.34)$$

and in the liquid phase:

$$N'_A = k_L(C_{Ai} - C_{AL}) \quad (12.35)$$

The effect of the chemical reaction is to accelerate the removal of **A** from the interface, and supposing that it is now r times as great then:

$$N''_A = rk_L(C_{Ai} - C_{AL}) \quad (12.36)$$

In Figure 12.11a, the concentration profile through the liquid film of thickness z_L is represented by a straight line such that $k_L = D_L/z_L$. In b, component **A** is removed by chemical reaction, so that the concentration profile is curved. The dotted line gives the concentration profile if, for the same rate of absorption, **A** were removed only by diffusion. The effective diffusion path is $1/r$ times the total film thickness z_L .

Thus:

$$N''_A = \frac{rD_L}{z_L}(C_{Ai} - C_{AL}) = rk_L(C_{Ai} - C_{AL}) \quad (12.37)$$

VAN KREVELEN and HOFTYZER⁽²⁸⁾ showed that the factor r may be related to C_{Ai} , D_L , k_L , to the concentration of **B** in the bulk liquid C_{BL} , and to the second-order reaction rate constant k_2 for the absorption of CO_2 in alkaline solutions. Their relationship is shown in Figure 12.12, in which r , that is $N''_A/k_L C_{Ai}$, is plotted against $(k_2 D_L C_{BL})^{1/2}/k_L$ for various values of C_{BL}/iC_{Ai} , where i is the number of kmol of **B** combining with 1 kmol of **A**.

Figure 12.2 illustrates three conditions:

- (a) If k_2 is very small, $r \simeq 1$, and conditions are those of physical absorption.
- (b) If k_2 is very large, $r \simeq C_{BL}/iC_{Ai}$, and the rate of the process is determined by the transport of **B** towards the phase boundary.

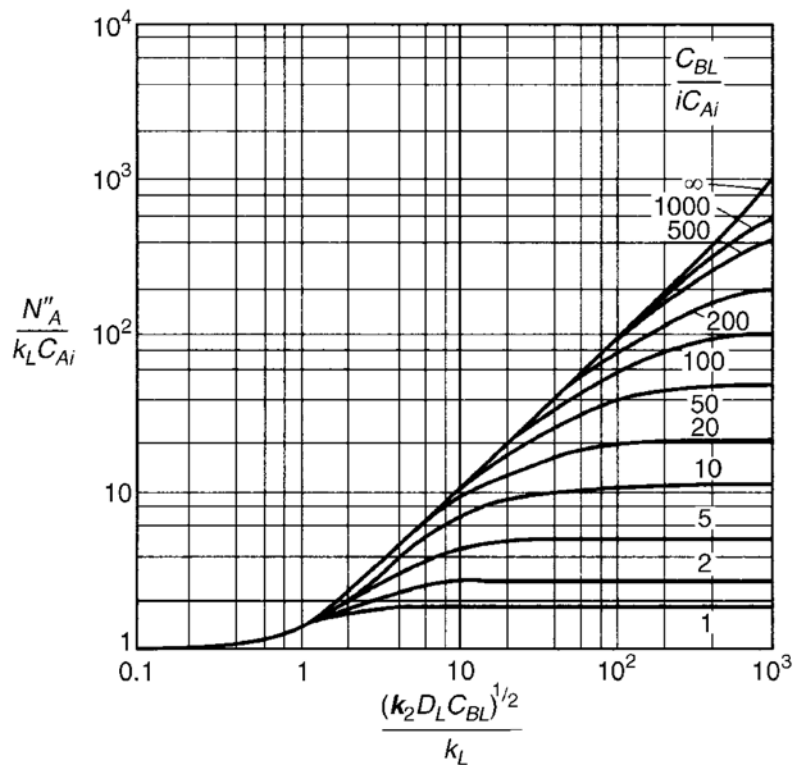


Figure 12.12. $N''_A/k_L C_{Ai}$ versus $(k_2 D_L C_{BL})^{1/2}/k_L$ for various values of C_{BL}/iC_{Ai}

- (c) At moderate values of k_2 , $r \approx (j D_L C_{BL})^{1/2}/k_L$, and the rate of the process is determined by the rate of the chemical reaction.

Thus, from equation 12.37:

$$N''_A = k_L(C_{Ai} - C_{AL}) \frac{(k_2 D_L C_{BL})^{1/2}}{k_L} = (C_{Ai} - C_{AL})(k_2 D_L C_{BL})^{1/2} \quad (12.38)$$

and the controlling parameter is now k_2 .

The results of this work have been confirmed by NIJSING, HENDRIKSZ, and KRAMERS⁽³⁸⁾.

As an illustration of combined absorption and chemical reaction, the results of TEPE and DODGE⁽³⁹⁾ on the absorption of carbon dioxide by sodium hydroxide solution may be considered. A 150 mm diameter tower filled to a depth of 915 mm with 12.5 mm carbon Raschig rings was used. Some of the results are indicated in Figure 12.13. $K_G a$ increases rapidly with increasing sodium hydroxide concentration up to a value of about 2 kmol/m³. Changes in the gas rate were found to have negligible effect on $K_G a$, indicating that the major resistance to absorption was in the liquid phase. The influence of the liquid rate was rather low, and was proportional to $L^{0.28}$. It may be assumed that, in this case, the final rate of the process is controlled by the resistance to diffusion in the liquid, by the rate of the chemical reaction, or by both together.

CRYDER and MALONEY⁽⁴⁰⁾ presented data on the absorption of carbon dioxide in diethanolamine solution, using a 200 mm tower filled with 20 mm rings, and some of their data are shown in Figure 12.14. The coefficient $K_G a$ is found to be independent of

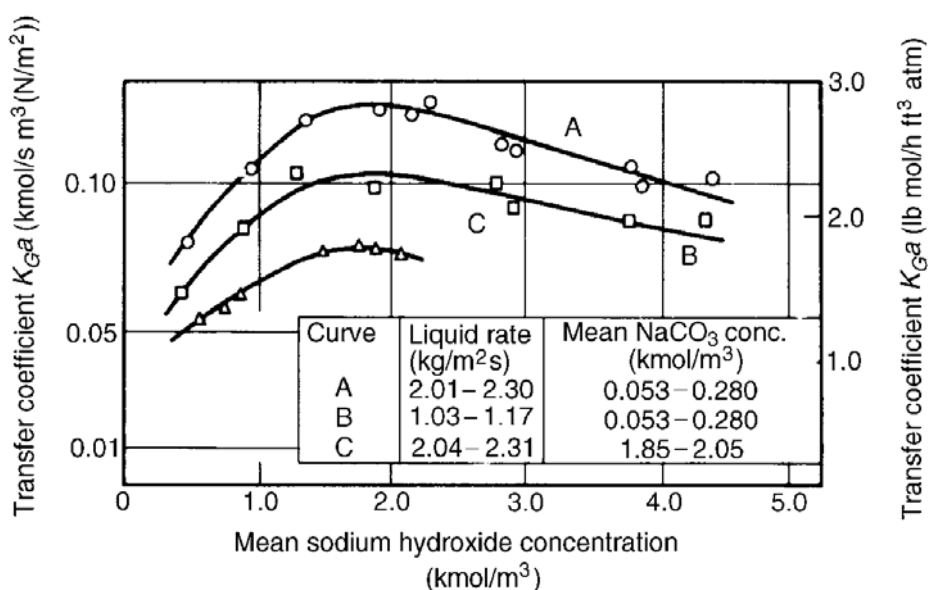


Figure 12.13. Absorption of carbon dioxide in sodium hydroxide solution $G' = 0.24-0.25 \text{ kg/m}^2\text{s}$, temperature = 298 K

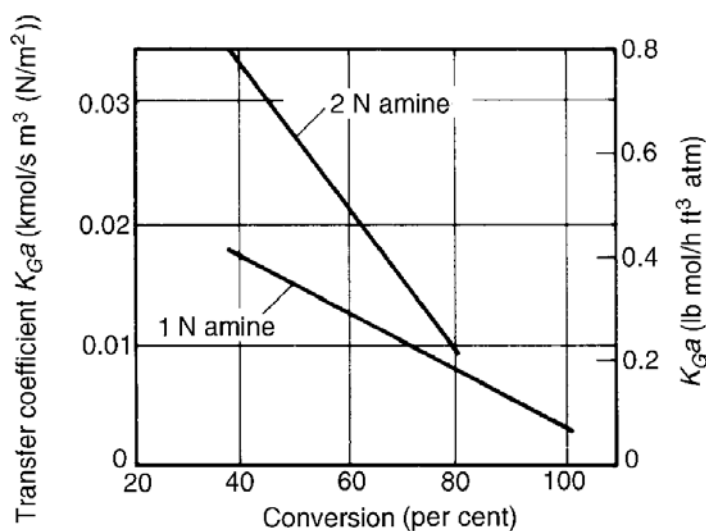


Figure 12.14. Absorption of carbon dioxide in diethanolamine solutions. Liquid rate = $1.85 \text{ kg/m}^2\text{s}$

the gas rate but to increase with the liquid rate, as expected in a process controlled by the resistance in the liquid phase.

It is difficult to deduce the size of tower required for an absorption combined with a chemical reaction, and a laboratory scale experiment should be carried out in all cases. STEPHENS and MORRIS⁽⁴¹⁾ have used a small disc-type tower illustrated in Figure 12.15 for preliminary experiments of this kind. It was found that a simple wetted-wall column was unsatisfactory where chemical reactions took place. In this unit a series of discs, supported by means of a wire, was arranged one on top of the other as shown.

The absorption of carbon dioxide into aqueous amine solutions has been investigated by DANCKWERTS and MCNEIL⁽⁴²⁾ using a stirred cell. It was found that the reaction proceeded

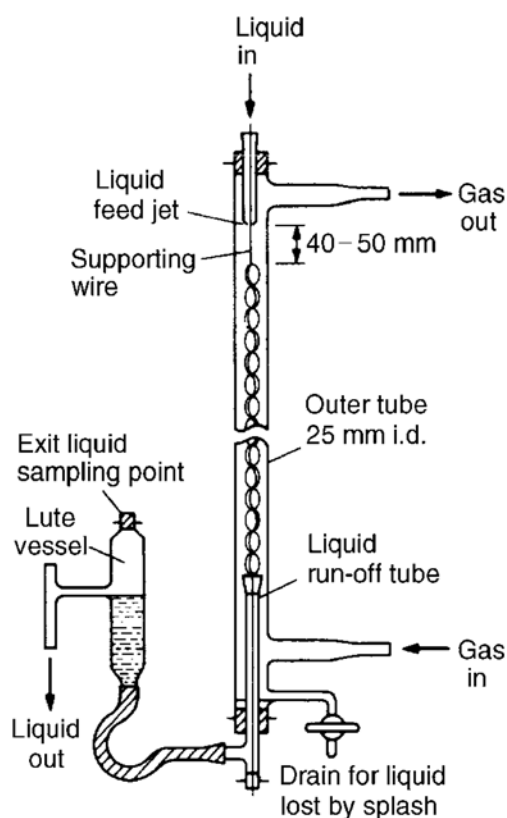


Figure 12.15. Small disc-tower for absorption tests

in two stages: first a fast reaction to give amine carbamate, and secondly a slow reaction in the bulk of the liquid in which the carbamate was partially hydrolysed to bicarbonate. The use of sodium arsenite as catalyst considerably accelerated this second reaction, showing that the overall capacity of an absorber could be substantially increased by a suitable catalyst.

A comprehensive review of work on the absorption of carbon dioxide by alkaline solutions has been carried out by DANCKWERTS and SHARMA⁽⁴³⁾ who applied results of research to the design of industrial scale equipment. Subsequently, SAHAY and SHARMA⁽⁴⁴⁾ showed that the mass transfer coefficient may be correlated with the gas and liquid rates and the gas and liquid compositions by:

$$K_G a = \text{const. } L'^{a_1} G'^{a_2} \exp(a_3 F' + a_4 y) \quad (12.39)$$

where: a_1, a_2, a_3, a_4 are experimentally determined constants,
 F' = fractional conversion of the liquid, and
 y = mole fraction of CO_2 in the gas.

ECKERT⁽⁴⁵⁾, by using the same reaction, determined the mass transfer performance of packings in terms of $K_G a$ as:

$$K_G a = \frac{N}{V(\Delta P_A)_{\text{lm}}} \quad (12.40)$$

where: N = number of moles of CO_2 absorbed,
 V = packed volume, and
 $(\Delta P_A)_{\text{lm}}$ = log mean driving force.

Data obtained from this work are limited by the conditions under which they were obtained. It is both difficult and dangerous to extrapolate over the entire range of conditions encountered on a full-scale plant.

12.6. ABSORPTION ACCOMPANIED BY THE LIBERATION OF HEAT

In some absorption processes, especially where a chemical reaction occurs, there is a liberation of heat. This generally gives rise to an increase in the temperature of the liquid, with the result that the position of the equilibrium curve is adversely affected.

In the case of plate columns, a heat balance may be performed over each plate and the resulting temperature determined. For adiabatic operation, where no heat is removed from the system, the temperature of the streams leaving the absorber will be higher than those entering, due to the heat of solution. This rise in temperature lowers the solubility of the solute gas so that a large value of L_m/G_m and a larger number of trays will be required than for isothermal operation.

For packed columns, the temperature rise will affect the equilibrium curve, and differential equations for heat and mass transfer, together with heat and mass balances, must be integrated numerically. An example of this procedure is given in Volume 1, Chapter 13, for the case of water cooling. For gas absorption under non-isothermal conditions, reference may be made to specialist texts^(46,47) for a detailed description of the methods available. As an approximation, it is sometimes assumed that all the heat evolved is taken up by the liquid, and that temperature rise of the gas may be neglected. This method gives an overestimate of the rise in temperature of the liquid and results in the design of a tower which is taller than necessary. Figure 12.16 shows the effect of the temperature rise on the equilibrium curve for an adiabatic absorption process of ammonia in water. If the amount of heat liberated is very large, it may be necessary to cool the liquid. This is most conveniently done in a plate column, either with heat exchangers connected between consecutive plates, or with cooling coils on the plate, as shown in Figure 12.17.

The overall heat transfer coefficient between the gas-liquid dispersion on the tray and the cooling medium in the tubes is dependent upon the gas velocity, as pointed out by POLL and SMITH⁽⁴⁸⁾, but is usually in the range 500–2000 W/m² K.

With packed towers it is considerably more difficult to arrange for cooling, and it is usually necessary to remove the liquid stream at intervals down the column and to cool externally. COGGAN and BOURNE⁽⁴⁹⁾ have presented a computer programme to enable the economic decision to be made between an adiabatic absorption tower, or a smaller isothermal column with interstage cooling.

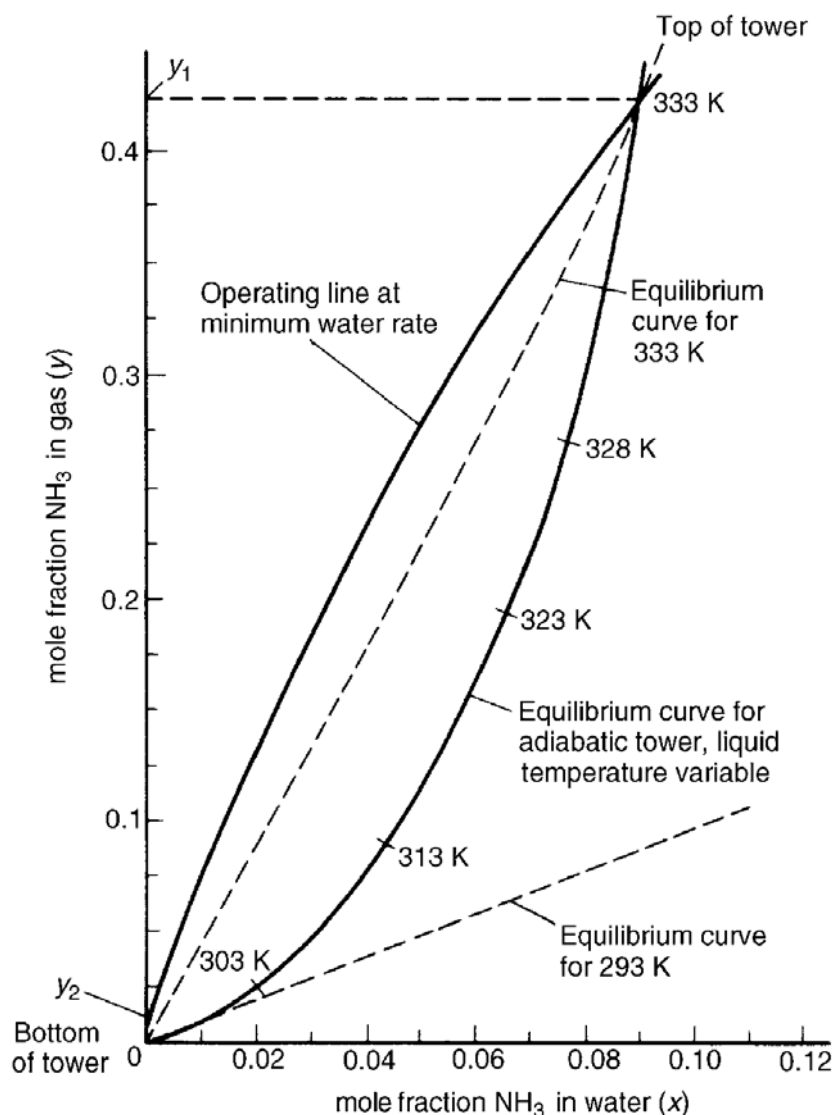


Figure 12.16. Equilibrium curve modified to allow for the heat of solution of the solute⁽⁴⁶⁾

12.7. PACKED TOWERS FOR GAS ABSORPTION

From the analysis given already of the diffusional nature of absorption, one of the outstanding requirements is to provide as large an interfacial area of contact as possible between the phases. For this purpose, columns similar to those used for distillation are suitable. However, whereas distillation columns are usually tall and thin absorption columns are more likely to be short and fat. In addition, equipment may be used in which gas is passed into a liquid which is agitated by a stirrer. A few special forms of units have also been used, although it is the packed column which is most frequently used for gas absorption applications.

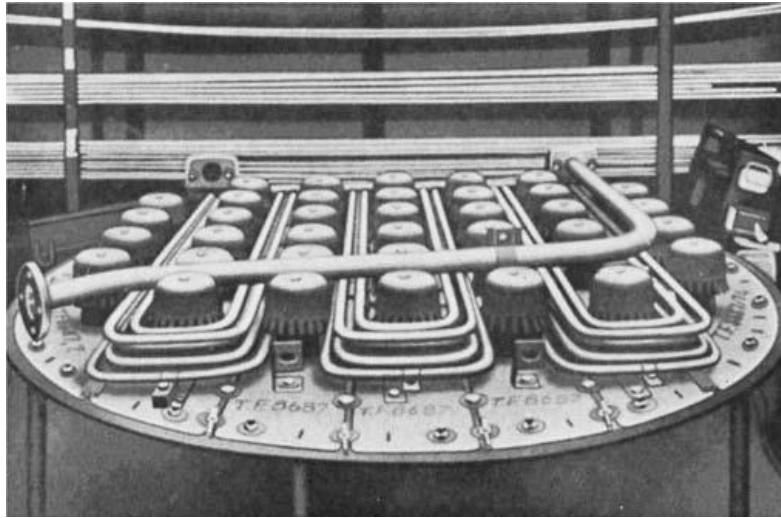


Figure 12.17. Glitsch “truss type” bubble-tray in stainless steel for a 1.9 m absorption column

12.7.1. Construction

The essential features of a packed column, as discussed in Chapter 4, are the shell, the arrangements for the gas and liquid inlets and outlets and the packing with its necessary supporting and redistributing systems. Reference may be made to Chapter 4 and to Volume 6 for details of these aspects, whilst this section is largely concerned with the determination of the height of packing for a particular duty. In installations where the gas is fed from a previous stage of a process where it is under pressure, there is no need to use a blower for the transfer of the gas through the column. When this is not the case, a simple blower is commonly used, and such blowers have been described in Volume 1, Chapter 8. The pressure drop across the column may be calculated by the methods presented in Chapter 4 of this volume and the blower sized accordingly. A pressure drop exceeding 30 mm of water per metre of packing is said to improve gas distribution though process conditions may not permit a figure as high as this. The packed height should not normally exceed 6 m in any section of the tower and for some packings a much lower height must be used.

In the design of an absorption tower it is necessary to take into account the characteristics of the packing elements and the flow behaviour discussed in Chapter 4, together with the considerations given in the following sections concerning the performance of columns under operating conditions.

12.7.2. Mass transfer coefficients and specific area in packed towers

Traditional methods of assessing the capacity of tower packings, which involve the use of the specific surface area S and the voidage e , developed from the fact that these

properties could be readily defined and measured for a packed bed of granular material such as granite, limestone, and coke which were some of the earliest forms of tower packings. The values of S and e enabled a reasonable prediction of hydraulic performance to be made. With the introduction of Raschig rings and other specially shaped packings, it was necessary to introduce a basis for comparing their relative efficiencies. Although the commonly published values of specific surface area S provide a reasonable basis of comparison, papers such as that by SHULMAN *et al.*⁽⁵⁰⁾ showed that the total area offered by Raschig rings was not used, and varied considerably with hydraulic loading.

Further evidence of the importance of the wetted fraction of the total area came with the introduction of the Pall type ring. A Pall ring having the same surface area as a Raschig ring is up to 60 per cent more efficient, though many still argue the relative merits of packings purely on the basis of surface area.

The selection of a tower packing is based on its hydraulic capacity, which determines the required cross-sectional area of the tower, and the efficiency, $K_G a$ typically, which governs the packing height. Here a is the area of surface per unit volume of column and is therefore equal to $S(1 - e)$. Table 12.3⁽⁵¹⁾ shows the capacity of the commonly available tower packings relative to 25 mm Raschig rings, for which a considerable amount of information is published in the literature. The table lists the packings in order of relative efficiency, $K_G a$, evaluated at the same approach to the hydraulic capacity limit determined by flooding in each case.

12.7.3. Capacity of packed towers

The drop in pressure for the flow of gas and liquid over packings is discussed in Chapter 4. It is important to note that, during operation, the tower does not reach flooding conditions. In addition, every effort should be made to have as high a liquid rate as possible, in order to attain satisfactory wetting of the packing.

With low liquid rates, the whole of the surface of the packing is not completely wetted. This may be seen very readily by allowing a coloured liquid to flow over packing contained in a glass tube. From the flow patterns, it is obvious how little of the surface is wetted until the rate is quite high. This difficulty of wetting can sometimes be overcome by having considerable recirculation of the liquid over the tower, although in other cases, such as vacuum distillation, poor wetting will have to be accepted because of the low volume of liquid available. In selecting a packing, it is desirable to choose the form which will give as near complete wetting as possible. The minimum liquid rate below which the packing will no longer perform satisfactorily is known as the minimum wetting rate, discussed in Chapter 4.

The following treatment is a particular application of the more general approach adopted in Volume 1, Chapter 10.

Figure 12.18 illustrates the conditions that occur during the steady operation of a countercurrent gas–liquid absorption tower. It is convenient to express the concentration of the streams in terms of moles of solute gas per mole of inert gas in the gas phase, and as moles of solute gas per mole of solute free liquid in the liquid phase. The actual area of interface between the two phases is not known, and the term a is introduced as the interfacial area per unit volume of the column. On this basis the general equation, 12.13,

Table 12.3. Capacity of commonly available packings relative to 25 mm Raschig rings⁽⁵¹⁾

Relative K_{Ga}	Raschig rings	Traditional saddles	Pall rings	Ceramic Pall rings	Ceramic cascade mini ring ⁽³⁾	Super Intalox ⁽¹⁾ saddles	Hypak ⁽¹⁾	Tellerettes ⁽²⁾	Cascade mini-ring ⁽³⁾
Materials available for this relative K_{Ga}	Ceramic	Ceramic Plastic (P)	Metal (M)	Ceramic	Ceramic	Ceramic Plastic	Metal	Plastic	Metal (M) Plastic (P)
0.6-0.7	75 mm								
0.7-0.8	50 mm								
0.8-0.9	37 mm								
0.9-1.0	25 mm								
1.0-1.1	12 mm	75 mm	87 mm		No. 5	No. 3	No. 3	Size L	
1.1-1.2		50 mm	50 mm				No. 3		
1.2-1.3		37 mm	50 mm		No. 3	No. 2	No. 2		
1.3-1.4		25 mm	37 mm						
1.4-1.5				50 mm					No. 4 (M)
1.5-1.6			25 mm	37 mm					No. 3 (P)
1.6-1.7			25 mm	25 mm	No. 2	No. 1	No. 1	Size S	No. 3 (M)
1.7-1.8			16 mm						No. 2 (P)
1.8-1.9									No. 2 (M)
1.9-2.0									No. 1 (P)
2.0-2.1									
2.1-2.2									No. 1 (M)

Note:

Relative K_{Ga} valid for all systems controlled by mass transfer coefficient (K_G) and wetted area (a) per unit volume of column. Some variation should be expected when liquid *reaction* rate is controlling (not liquid *diffusion* rate). In these cases liquid hold-up becomes more important. In general a packing having high liquid hold-up which is clearly greater than that in the falling film has poor capacity.

Gas capacity before hydraulic limit (flooding) relative to 25 mm Raschig rings (also approx. the reciprocal of tower cross-sectional area relative to 25 mm Raschig rings for the same pressure drop throughout loading range). All relative capacity figures are valid for the same liquid to gas mass rate ratio:

- (1) Trade Mark of Norton Company, U.S.A. (Hydronyl U.K.).
- (2) Trade Mark of Ceilcote Company.
- (3) Trade Mark of Mass Transfer Ltd. (& Inc.).

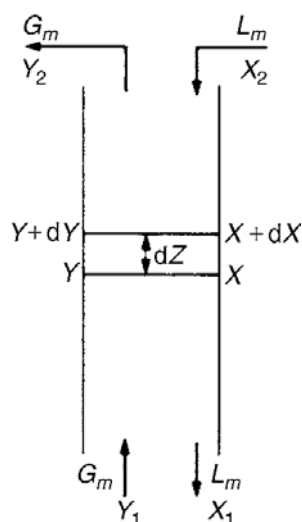


Figure 12.18. Countercurrent absorption tower

for mass transfer can be written as:

$$\begin{aligned} N'_A A dZ a &= k_G a (P_{AG} - P_{Ai}) A dZ \\ &= k_L a (C_{Ai} - C_{AL}) A dZ \end{aligned} \quad (12.41)$$

where: N'_A = kmol of solute absorbed per unit time and unit interfacial area,
 a = surface area of interface per unit volume of column,
 A = cross-sectional area of column, and
 Z = height of packed section.

$$\text{The interfacial area for transfer} = a dV = a A dZ \quad (12.42)$$

12.7.4. Height of column based on conditions in the gas film

If G_m = moles of inert gas/(unit time) (unit cross-section of tower),
 L_m = moles of solute-free liquor/(unit time) (unit cross-section of tower),
 Y = moles of solute gas **A**/mole of inert gas **B** in gas phase, and
 X = moles of solute **A**/mole of inert solvent in liquid phase.

and at any plane at which the molar ratios of the diffusing material in the gas and liquid phases are Y and X , then over a small height dZ , the moles of gas leaving the gas phase will equal the moles taken up by the liquid.

$$\text{Thus:} \quad A G_m dY = A L_m dX \quad (12.43)$$

$$\text{But:} \quad G_m A dY = N'_A (a dV) = k_G a (P_{Ai} - P_{AG}) A dZ \quad (12.44)$$

It may be noted that, in a gas absorption process, gas and liquid concentrations will decrease in the upwards direction and both dX and dY will be negative.

Since:

$$P_{AG} = \frac{Y}{1+Y}P$$

$$G_m dY = k_G a P \left[\frac{Y_i}{1+Y_i} - \frac{Y}{1+Y} \right] dZ$$

$$= k_G a P \left[\frac{Y_i - Y}{(1+Y)(1+Y_i)} \right] dZ$$

Hence the height of column Z required to achieve a change in Y from Y_1 at the bottom to Y_2 at the top of the column is given by:

$$\int_0^Z dZ = Z = \frac{G_m}{k_G a P} \int_{Y_1}^{Y_2} \frac{(1+Y)(1+Y_i) dY}{Y_i - Y} \quad (12.45)$$

which for dilute mixtures may be written as:

$$Z = \frac{G_m}{k_G a P} \int_{Y_1}^{Y_2} \frac{dY}{Y_i - Y} \quad (12.46)$$

In this analysis it has been assumed that k_G is a constant throughout the column, and provided the concentration changes are not too large this will be reasonably true.

12.7.5. Height of column based on conditions in liquid film

A similar analysis may be made in terms of the liquid film. Thus from equations 12.41 and 12.42:

$$AL_m dX = k_L a (C_{Ai} - C_{AL}) A dZ \quad (12.47)$$

where the concentrations C are in terms of moles of solute per unit volume of liquor. If $C_T = (\text{moles of solute} + \text{solvent}) / (\text{volume of liquid})$, then:

$$\frac{C_A}{C_T - C_A} = \frac{\text{moles of solute}}{\text{moles of solvent}} = X$$

whence:

$$C_A = \frac{X}{1+X} C_T \quad (12.48)$$

The transfer equation (12.47) may now be written as:

$$L_m dX = k_L a C_T \left[\frac{X}{1+X} - \frac{X_i}{1+X_i} \right] dZ$$

$$= k_L a C_T \left[\frac{X - X_i}{(1+X_i)(1+X)} \right] dZ$$

Thus:

$$\int_0^Z dZ = Z = \frac{L_m}{k_L a C_T} \int_{X_1}^{X_2} \frac{(1+X_i)(1+X) dX}{X - X_i} \quad (12.49)$$

and for dilute concentrations this gives:

$$Z = \frac{L_m}{k_L a C_T} \int_{X_1}^{X_2} \frac{dX}{X - X_i} \quad (12.50)$$

where C_T and k_L have been taken as constant over the column.

12.7.6. Height based on overall coefficients

If the driving force based on the gas concentration is written as $(Y - Y_e)$ and the overall gas transfer coefficient as K_G , then the height of the tower for dilute concentrations becomes:

$$Z = \frac{G_m}{K_G a P} \int_{Y_1}^{Y_2} \frac{dY}{Y_e - Y} \quad (12.51)$$

or in terms of the liquor concentration as:

$$Z = \frac{L_m}{K_L a C_T} \int_{X_1}^{X_2} \frac{dX}{X - X_e} \quad (12.52)$$

Equations for dilute concentrations

As the mole fraction is approximately equal to the molar ratio at dilute concentrations then considering the gas film:

$$Z = \frac{G_m}{K_G a P} \int_{Y_1}^{Y_2} \frac{dY}{Y_e - Y} = \frac{G_m}{K_G a P} \int_{y_1}^{y_2} \frac{dy}{y_e - y} \quad (12.53)$$

and considering the liquid film:

$$Z = \frac{L_m}{K_L a C_T} \int_{X_1}^{X_2} \frac{dX}{X - X_e} = \frac{L_m}{K_L a C_T} \int_{x_1}^{x_2} \frac{dx}{x - x_e} \quad (12.54)$$

12.7.7. The operating line and graphical integration for the height of a column

Taking a material balance on the solute from the bottom of the column to any plane where the mole ratios are Y and X gives for unit area of cross-section:

$$G_m(Y_1 - Y) = L_m(X_1 - X) \quad (12.55)$$

or:
$$Y_1 - Y = \frac{L_m}{G_m}(X_1 - X) \quad (12.56)$$

This is the equation of a straight line of slope L_m/G_m , which passes through the point (X_1, Y_1) . It may be seen by making a material balance over the whole column that the same line passes through the point (X_2, Y_2) . This line, known as the operating line,

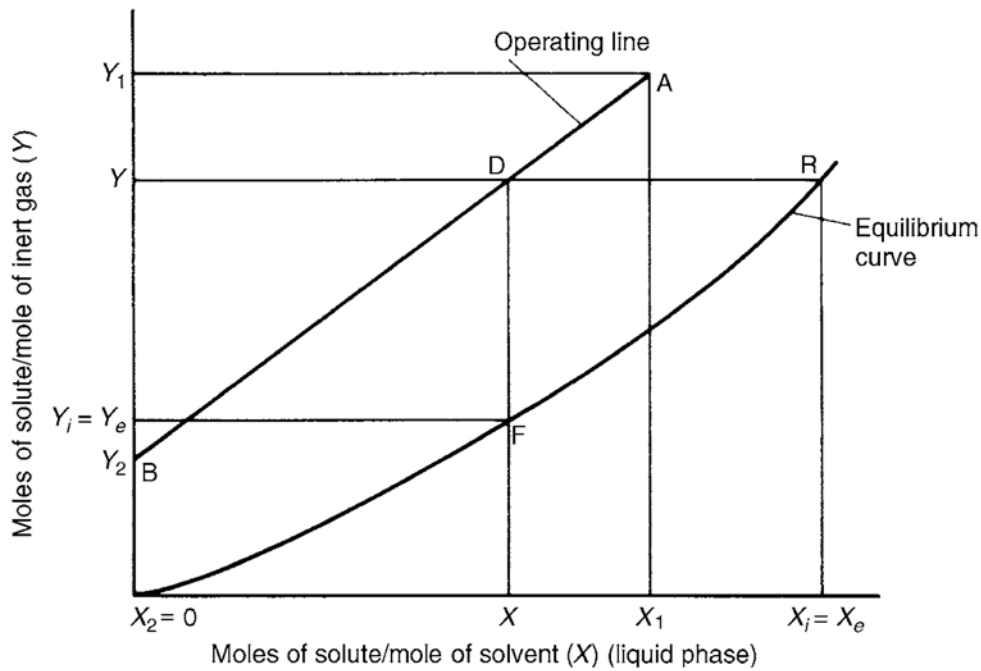


Figure 12.19. Driving force in gas and liquid-film controlled processes. The Figure shows the operating line BDA and the equilibrium curve FR

represents the conditions at any point in the column. It is similar to the operating line used in Chapter 11. Figure 12.19 illustrates typical conditions for the case of moist air and sulphuric acid or caustic soda solution, where the main resistance lies in the gas phase.

The equilibrium curve is represented by the line FR, and the operating line is given by AB, A corresponding to the concentrations at the bottom of the column and B to those at the top of the column. D represents the condition of the bulk of the liquid and gas at any point in the column, and has coordinates X and Y. Then, if the gas film is controlling the process, Y_i equals Y_e , and is given by a point F on the equilibrium curve, with coordinates X and Y_i . The driving force causing transfer is then given by the distance DF. It is therefore possible to evaluate the expression:

$$\int_{Y_1}^{Y_2} \frac{dY}{Y_i - Y}$$

by selecting values of Y, reading off from the Figure the corresponding values of Y_i , and thus calculating $1/(Y_i - Y)$. It may be noted that, for gas absorption, $Y > Y_i$ and $Y_i - Y$ and dY in the integral are both negative.

If the liquid film controls the process, X_i equals X_e and the driving force $X_i - X$ is given in Figure 12.19 by the line DR. The evaluation of the integral:

$$\int_{X_1}^{X_2} \frac{dX}{X - X_i}$$

may be effected in the same way as for the gas film.

Special case when equilibrium curve is a straight line

If over the range of concentrations considered the equilibrium curve is a straight line, it is permissible to use a mean value of the driving force over the column. For dilute concentrations, over a small height dZ of column, the absorption is given by:

$$N'_A A a dZ = G_m A dy = K_G a A P (y_e - y) dZ \quad (12.57)$$

If: $y_e = mx + c$ (12.58)

then: $y_{e2} = mx_2 + c$

and: $y_{e1} = mx_1 + c$

so that: $m = \frac{y_{e1} - y_{e2}}{x_1 - x_2}$ (12.59)

Further, taking a material balance over the lower portion of the columns gives:

$$L_m(x_1 - x) = G_m(y_1 - y)$$

and: $x = x_1 - \frac{G_m}{L_m}(y_1 - y)$ (12.60)

From equation 12.57:

$$\int_0^Z \frac{K_G a P}{G_m} dZ = \int_{y_1}^{y_2} \frac{dy}{y_e - y} \quad (12.61)$$

$$= \int_{y_1}^{y_2} \frac{dy}{m[x_1 + (G_m/L_m)(y - y_1)] + c - y}$$

(from equations 12.58 and 12.60)

$$= \frac{1}{1 - (mG_m/L_m)} \ln \frac{mx_1 + c - y_1}{y_2 - m[x_1 + (G_m/L_m)(y_2 - y_1)] - c}$$

$$= \frac{1}{1 - \frac{y_{e1} - y_{e2}}{x_1 - x_2} \cdot \frac{x_1 - x_2}{y_1 - y_2}} \ln \frac{y_{e1} - y_1}{y_{e1} - y_{e2} - \left(\frac{y_{e1} - y_{e2}}{x_1 - x_2} \frac{x_1 - x_2}{y_1 - y_2} y_1 - y_2 \right)}$$

(from equations 12.58, 12.59, and 12.60)

$$= \frac{y_1 - y_2}{(y - y_e)_1 - (y - y_e)_2} \ln \frac{(y - y_e)_1}{(y - y_e)_2}$$

$$= \frac{y_1 - y_2}{(y - y_e)_{lm}}$$

where $(y - y_e)_{lm}$ is the logarithmic mean value of $y - y_e$.

Substituting in equation 12.61:

$$\frac{K_G a P}{G_m} Z = \frac{y_1 - y_2}{(y - y_e)_{lm}}$$

Thus: $aAZN'_A = G_m(y_1 - y_2)A = K_G a A P (y - y_e)_{lm} Z$ (12.62)

and in terms of mole ratios:

$$aAZN'_A = G_m(Y_1 - Y_2)A = K_G aAP(Y - Y_e)_{lm}Z \quad (12.63)$$

Thus, the logarithmic mean of the driving forces at the top and the bottom of the column may be used.

For concentrated solutions:

$$aAZN'_A = G_m(Y_1 - Y_2)A = K_G aA\phi P(Y - Y_e)_{lm}Z \quad (12.64)$$

It is necessary to introduce the factor ϕ since Y is not directly proportional to P . The value of ϕ may be found from the relation:

$$\phi Y = \frac{Y}{1 + Y} \quad (12.65)$$

from which $\phi = 1/(1 + Y)$. Although the value of ϕ will change slightly over the column, a mean value will generally be acceptable.

It is of interest to note from Figure 12.20, that, as long as the ratio k_L/k_G remains constant (that is, if the slope of DE is constant), then the ratio of DQ, the driving force through the gas phase, divided by DF, the driving force assuming all the resistance to be in the gas phase, will be a constant. Thus, the use of the driving force DF is satisfactory even if the resistance does not lie wholly in the gas phase. The coefficient k_G on this basis is not an accurate value for the gas-film coefficient, although is proportional to it. It follows that, if the equilibrium curve is straight, either the gas-film or the liquid-film coefficient may be used. This simplification is of considerable value.

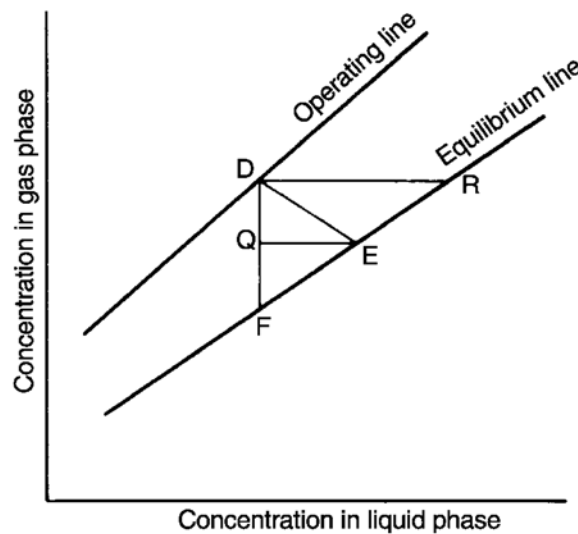


Figure 12.20. Driving force when equilibrium curve is a straight line

12.7.8. Capacity of tower in terms of partial pressures for high concentrations

A material balance taken between the bottom of the column and some plane where the partial pressure in the gas phase is P_{AG} and the concentration in the liquid is X gives:

$$L_m(X_1 - X) = G_m \left[\frac{P_{AG1}}{P - P_{AG1}} - \frac{P_{AG}}{P - P_{AG}} \right] \quad (12.66)$$

Over a small height of the column dZ , therefore:

$$\begin{aligned} -L_m dX &= \frac{-G_m P}{(P - P_{AG})^2} dP_{AG} = k_G a (P_{AG} - P_{Ai}) dZ & (12.67) \\ &= \frac{k_G P_{Bm}}{P} a P \frac{(P_{AG} - P_{Ai})}{P_{Bm}} dZ \\ &= k'_G a P \frac{(P_{AG} - P_{Ai})}{P_{Bm}} dZ \end{aligned}$$

Thus:
$$\int_0^Z dZ = G_m \int_{P_{AG1}}^{P_{AG2}} \frac{-P_{Bm} dP_{AG}}{k'_G a (P - P_{AG})^2 (P_{AG} - P_{Ai})} \quad (12.68)$$

The advantage of using k'_G instead of k_G is that k'_G is independent of concentration, although this equation is almost unmanageable in practice. If a substantial amount of the gas is absorbed from a concentrated mixture, k'_G will still change as a result of a reduced gas velocity, although it is independent of concentration.

12.7.9. The transfer unit

The group $\int (dy/y_e - y)$, which is used in Chapter 11, has been defined by CHILTON and COLBURN⁽⁵²⁾ as the number of overall gas transfer units N_{OG} . The concept of the transfer unit is also introduced in Volume 1, Chapter 10. The application of this group to the countercurrent conditions in the absorption tower is now considered.

Over a small height dZ , the partial pressure of the diffusing component **A** will change by an amount dP_{AG} . Then the moles of **A** transferred are given by:

$$(\text{change in mole fraction}) \times (\text{total moles of gas})$$

Therefore:
$$K_G a (P_{AG} - P_{Ae}) dZ = \frac{-dP_{AG}}{P} G'_m \quad (12.69)$$

(for dilute concentrations)

Thus:
$$\int_{P_{AG2}}^{P_{AG1}} \frac{dP_{AG}}{P_{Ae} - P_{AG}} = \int_0^Z \frac{K_G a P}{G'_m} dZ \quad (12.70)$$

or in terms of mole fractions:

$$N_{OG} = \int_{y_1}^{y_2} \frac{dy}{y_e - y} = \int_0^Z K_G a \frac{P}{G'_m} dZ = K_G a \frac{P}{G'_m} Z \quad (12.71)$$

The number of overall gas transfer units N_{OG} is an integrated value of the change in composition per unit driving force, and therefore represents the difficulty of the separation.

In many cases in gas absorption, $(y - y_e)$ is very small at the top of the column, and consequently $1/(y - y_e)$ is very much greater at the top than at the bottom of the column. Thus, equation 12.77 may lead to the use of an integral which is difficult to evaluate graphically because of the very steep slope of the curve.

Now:
$$N_{OG} = \int_{y_1}^{y_2} \frac{dy}{y_e - y} = \int_{y_1}^{y_2} \frac{y d(\ln y)}{y_e - y} \quad (12.72)$$

In these circumstances, the new form of the integral is much more readily evaluated, as pointed out by RACKETT⁽⁵³⁾.

As in Chapter 11 (equation 11.140), equation 12.77 may be written as:

$$N_{OG} = \frac{\text{Height of column}}{\text{Height of transfer unit}} = \frac{Z}{H_{OG}} \quad (12.73)$$

The height of the overall gas transfer unit is then $H_{OG} = \frac{G'_m}{PK_G a}$ (12.74)

If the driving force is taken over the gas-film only, the height of a gas-film transfer unit $H_G = G'_m/Pk_G a$ is obtained. Similarly for the liquid film, the height of the overall liquid-phase transfer unit H_{OL} is given by:

$$H_{OL} = \frac{L'_m}{K_L a C_T} \quad (12.75)$$

The height of the liquid-film transfer unit is given by:

$$H_L = \frac{L'_m}{k_L a C_T} \quad (12.76)$$

where C_T is the mean molar density of the liquid.

In this analysis, it is assumed that the total number of moles of gas and liquid remain the same. This is true in absorption only when a small change in concentration takes place. With distillation, the total number of moles of gas and liquid does remain more nearly constant so that no difficulty then arises. In Chapter 11, the following relationships between individual and overall heights of transfer units are obtained and methods of obtaining the values of H_G and H_L are discussed:

$$H_{OG} = H_G + \frac{mG'_m}{L'_m} H_L \quad (\text{equation 11.148})$$

$$H_{OL} = H_L + \frac{L'_m}{mG'_m} H_G \quad (\text{equation 11.149})$$

For absorption duties, SEMMELBAUER⁽²⁹⁾ presented the following equations to evaluate H_G and H_L for Raschig rings and Berl saddles:

$$H_G = \beta \left[\frac{G'^{0.41} \mu_G^{0.26} \mu_L^{0.46} \sigma^{0.5}}{L'^{0.46} \rho_G^{0.67} \rho_L^{0.5} D_G^{0.67} d_p^{0.05}} \right] \quad (12.77)$$

$$\mathbf{H}_L = \beta \left[\frac{\mu_L^{0.88} \sigma^{0.5}}{L^{0.05} \rho_L^{1.33} D_L^{0.5} d_p^{0.55}} \right] \quad (12.78)$$

where $\beta = 30$ for Raschig rings and $\beta = 21$ for Berl saddles respectively, and L' and G' are mass flowrates per unit area.

The limits of validity and the units for the terms in equations 12.77 and 12.78 are given in Table 12.4.

Table 12.4. Range of application of equations 12.77 and 12.78

L'	0.1–10	kg/m ² s	μ_L	0.2–2	mN s/m ²
G'	0.1–1.0	kg/m ² s	μ_G	0.005–0.03	mN s/m ²
d_p	0.006–0.06	m	σ	(20–200) $\times 10^{-3}$	J/m ²
ρ_L	600–1400	kg/m ³	T	273–373	K
ρ_G	0.4–4	kg/m ³	d/d_p	2.5–25	—
D_L	(3–30 $\times 10^{-10}$)	m ² /s	h_p/d_p	10–100	—
D_G	(3–90 $\times 10^{-6}$)	m ² /s			

For a range of packings, MORRIS and JACKSON⁽²¹⁾ have presented values of the heights of the individual film transfer-units as shown in Table 12.5. For Pall rings and Intalox saddles, the nomographs in Figures 12.21 and 12.22⁽⁵⁴⁾ may be used though Figure 12.22 must not be used to estimate \mathbf{H}_L for distillation applications. Table 11.6 gives the value as a function of size and type of packing. It is, however, satisfactory for absorption and stripping duties.

Concentrated solutions

With concentrated solutions, allowance must be made for the change in the total number of moles flowing, because the molar flow will decrease up the column if the amount of absorption is large.

COLBURN⁽⁵⁵⁾ has shown that, under these conditions, the number of transfer units is given by:

$$\mathbf{N}_{OG} = \int_{y_1}^{y_2} \frac{dy}{y_e - y} \frac{(1 - y)_{\text{lm}}}{1 - y} \quad (12.79)$$

where $(1 - y)_{\text{lm}}$ is the logarithmic mean of $(1 - y)$ and $(1 - y_i)$.

Example 12.3

An acetone–air mixture containing 0.015 mole fraction of acetone has the mole fraction reduced to 1 per cent of this value by countercurrent absorption with water in a packed tower. The gas flowrate G' is 1 kg/m²s of air and the water enters at 1.6 kg/m²s. For this system, Henry's law holds and $y_e = 1.75x$, where y_e is the mole fraction of acetone in the vapour in equilibrium with a mole fraction x in the liquid. How many overall transfer units are required?

Table 12.5. Height of a transfer unit for various packings⁽⁵⁴⁾

Material	Size (mm)			Height of a transfer unit (m)	
	Pitch	Height	Thickness	H_G	H_L
Grids					
Plain grids					
Metal	25	25	1.6	1	0.5
	25	50	1.6	1.2	0.6
Wood	25	25	6.4	0.9	0.5
	25	50	6.4	1.2	0.6
Serrated grids					
Wood	100	100	13	6.8	0.7
	50	50	9.5	1.8	0.6
	38	38	4.8	1.6	0.6
Solid material	nominal size				
Coke	75			0.7	0.9
	38			0.25	0.8
	25			0.2	0.7
Quartz	50			0.5	0.8
	25			0.16	0.8
	Diameter	Height	Thickness		
Stacked Raschig rings					
Stoneware	100	100	9.5	1.8	0.7
	75	75	9.5	1.1	0.6
	75	75	6.4	1.4	0.6
	50	50	6.4	0.7	0.6
	50	50	4.8	0.8	0.6
Random Raschig rings					
Metal	50	50	1.6	0.5	0.6
	25	25	1.6	0.2	0.5
	13	13	0.8	0.1	0.5
Stoneware	75	75	9.5	0.8	0.7
	50	50	6.4	0.5	0.6
	50	50	4.8	0.5	0.6
	38	38	4.8	0.3	0.6
	25	25	2.5	0.2	0.5
	19	19	2.5	0.15	0.5
	13	13	1.6	0.1	0.5
Carbon	50	50	6.4	0.5	0.6
	25	25	4.8	0.2	0.5
	13	13	3.2	0.1	0.5

Solution

As the system is dilute, mole fractions are approximately equal to mole ratios.

At the bottom of the tower: $y_1 = 0.015$, $G' = 1.0 \text{ kg/m}^2\text{s}$, x_1 is unknown

At the top of the tower: $y_2 = 0.00015$, $x_2 = 0$ and $L' = 1.6 \text{ kg/m}^2\text{s}$

Thus: $L_m = (1.6/18) = 0.0889 \text{ kmol/m}^2\text{s}$

and: $G_m = (1.0/29) = 0.0345 \text{ kmol/m}^2\text{s}$

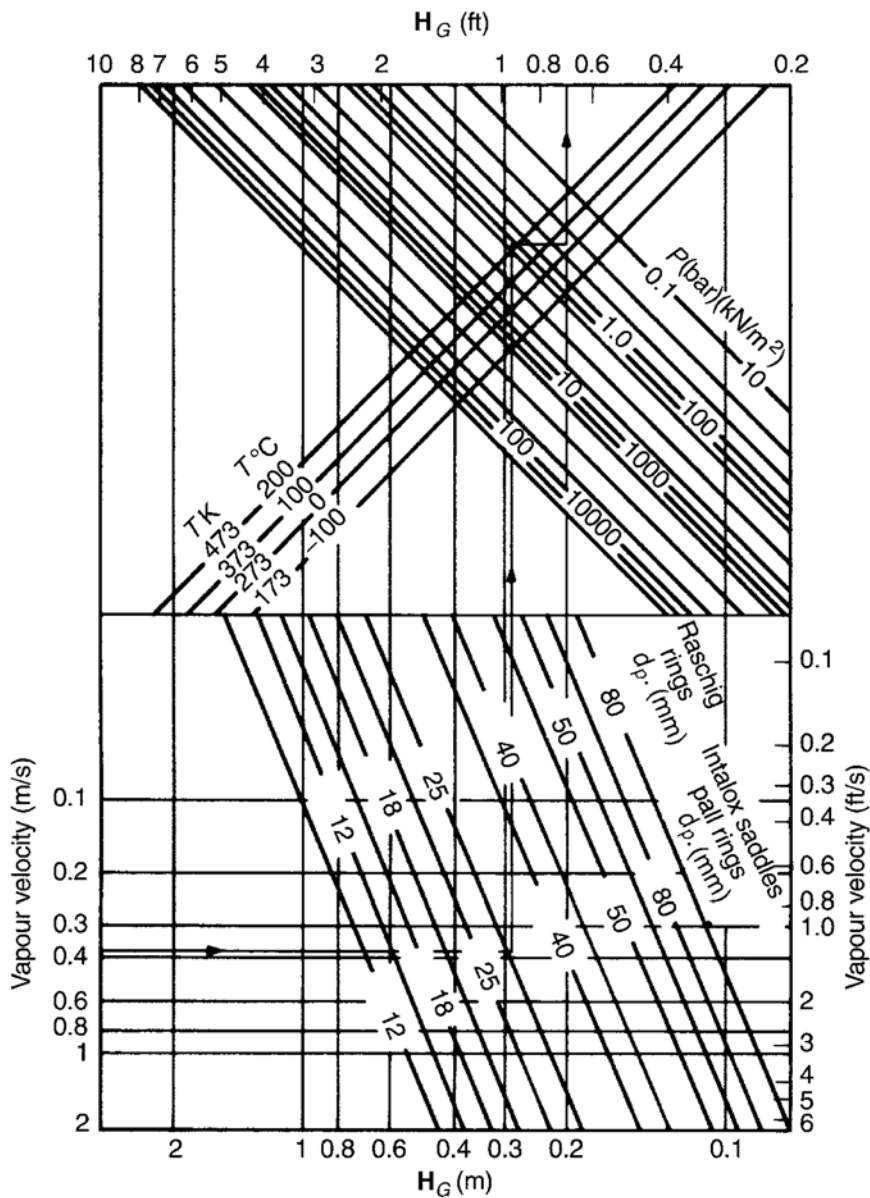


Figure 12.21. Nomograph for the estimation of the height of a gas-phase transfer unit⁽⁵⁴⁾

An overall mass balance gives:

$$G_m(y_1 - y_2) = L_m(x_1 - x_2)$$

or: $0.0345(0.015 - 0.00015) = 0.0889(x_1 - 0)$ and $x_1 = 0.00576$

Thus: $y_{e1} = (1.75 \times 0.00576) = 0.0101$

The number of overall transfer units is defined by:

$$N_{OG} = \int_{y_2}^{y_1} \frac{dy}{y - y_e} = \frac{y_1 - y_2}{(y - y_e)_{lm}} \quad \text{(from equations 12.60 and 12.70)}$$

Top driving force, $(y_2 - y_{e2}) = 0.00015$ since $x_2 = 0$.

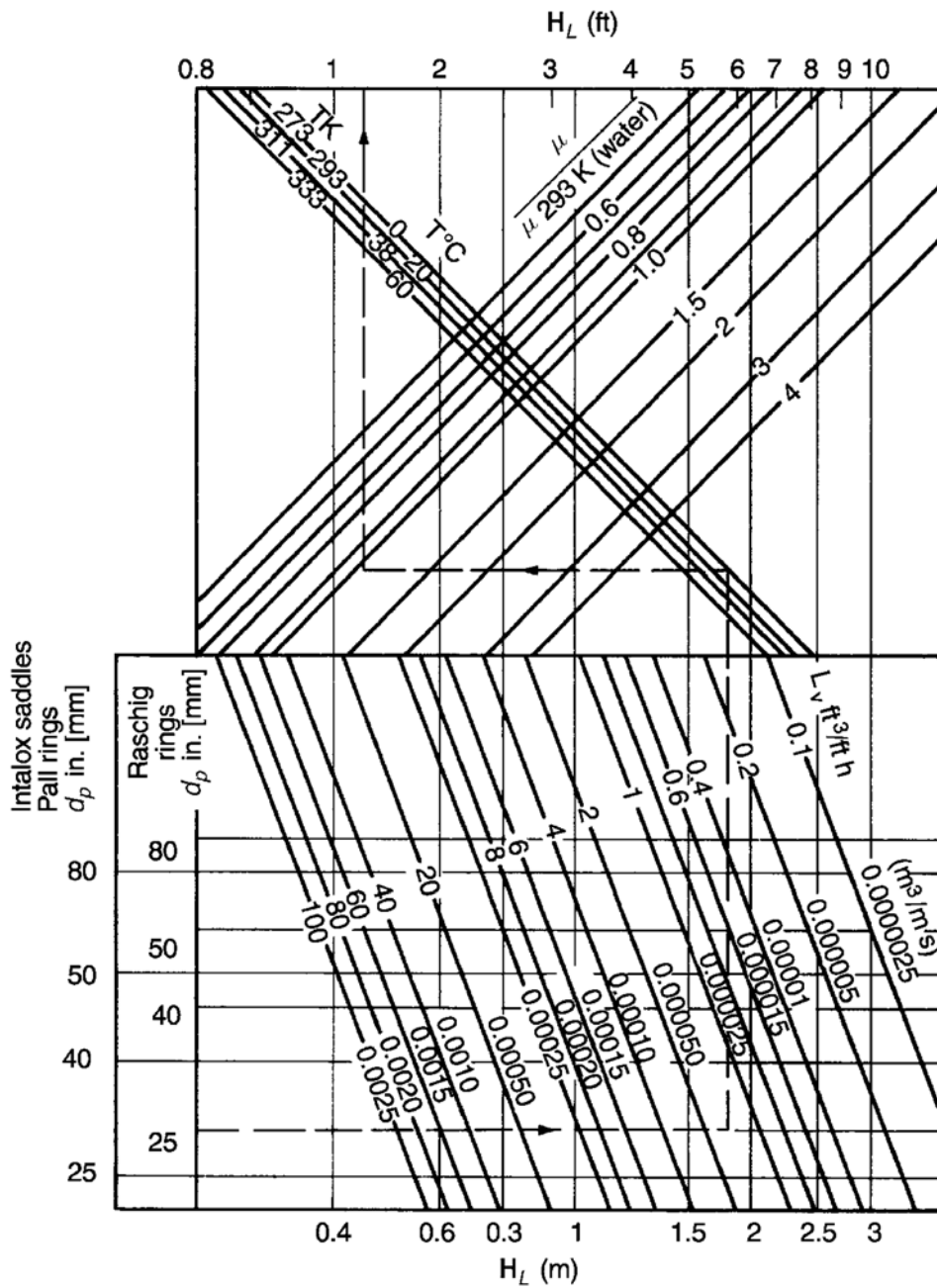


Figure 12.22. Nomograph for the estimation of the height of a liquid-phase transfer unit⁽⁵⁴⁾

Bottom driving force, $= (y_1 - y_{e1}) = (0.015 - 0.0101) = 0.0049$.

Thus: $(y - y_e)_{lm} = (0.0049 - 0.00015) / \ln(0.0049/0.00015) = 0.00136$

and: $N_{OG} = (0.015 - 0.00015) / 0.00136 = \underline{\underline{10.92}}$

Also: $N_{OL} = N_{OG} \times mG_m / L_m$
 $= (10.92 \times 1.75 \times 0.0345) / 0.0889 = \underline{\underline{7.42}}$

12.7.10. The importance of liquid and gas flowrates and the slope of the equilibrium curve

For a packed tower operating with dilute concentrations, since $x \simeq X_1$ and $y \simeq Y_1$, then:

$$G'_m(y_1 - y_2) = L'_m(x_1 - x_2) \quad (12.80)$$

where, as before, x and y are the mole fractions of solute in the liquid and gas phases, and G'_m and L'_m are the gas and liquid molar flowrates per unit area on a solute free basis.

A material balance between the top and some plane where the mole fractions are x , y gives:

$$G'_m(y - y_2) = L'_m(x - x_2) \quad (12.81)$$

If the entering solvent is free from solute, then $x_2 = 0$ and:

$$x = \frac{G'_m}{L'_m}(y - y_2) \quad (12.82)$$

But the number of overall transfer units is given by:

$$N_{OG} = \int_{y_1}^{y_2} \frac{dy}{y_e - y}$$

For dilute concentrations, Henry's law holds and $y_e = mx$. Therefore:

Thus:

$$\begin{aligned} N_{OG} &= \int_{y_1}^{y_2} \frac{dy}{\frac{mG'_m}{L'_m}(y - y_2) - y} \\ &= \int_{y_1}^{y_2} \frac{dy}{\left[\frac{mG'_m}{L'_m} - 1\right]y - \frac{mG'_m}{L'_m}y_2} \end{aligned}$$

and:

$$N_{OG} = \frac{1}{1 - \frac{mG'_m}{L'_m}} \ln \left[\left(1 - \frac{mG'_m}{L'_m}\right) \frac{y_1}{y_2} + \frac{mG'_m}{L'_m} \right] \quad (12.83)$$

COLBURN⁽⁵⁵⁾ has shown that this equation may usefully be plotted as shown in Figure 12.23 which is taken from his paper. In this plot the number of transfer units N_{OG} is shown for values of y_1/y_2 using mG'_m/L'_m as a parameter and it may be seen that the greater mG'_m/L'_m , the greater is the value of N_{OG} for a given ratio of y_1/y_2 . From equation 12.82:

$$\frac{L'_m}{G'_m} = \frac{y_1 - y_2}{x_1} = \frac{y_1 - y_2}{y_{e1}/m}$$

Thus:

$$\frac{mG'_m}{L'_m} = \frac{y_{e1}}{y_1 - y_2}$$

where y_{e1} is the value of y in equilibrium with x_1 .

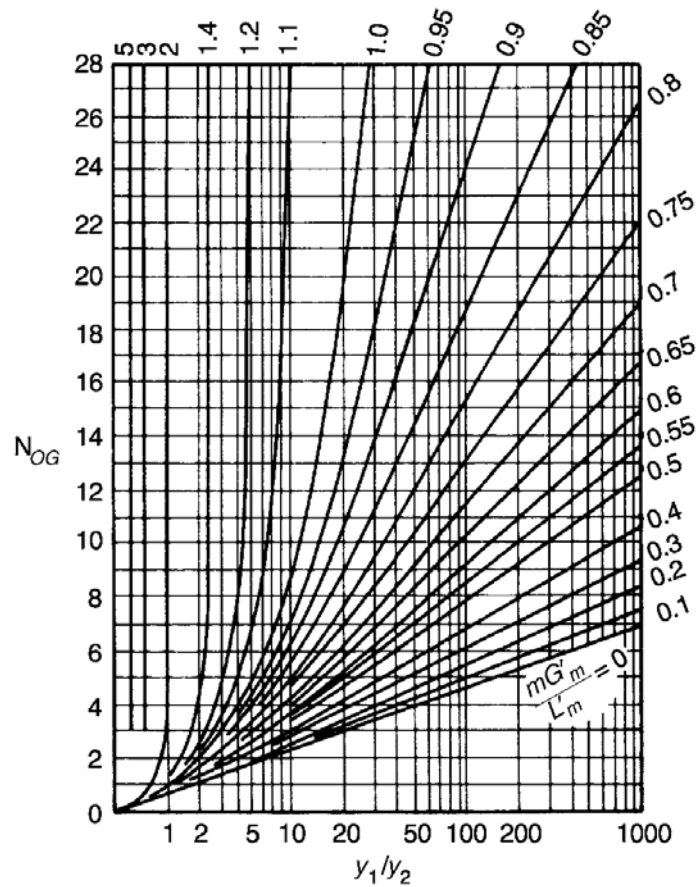


Figure 12.23. Number of transfer units N_{OG} as a function of y_1, y_2 , with mG'_m/L'_m as parameter

On this basis, the lower the value of mG'_m/L'_m , the lower will be y_{e1} , and hence the weaker the exit liquid. Colburn has suggested that the economic range for mG'_m/L'_m is 0.7–0.8. If the value of H_{OG} is known, the quickest way of obtaining a good indication of the required height of the column is by using Figure 12.23.

Example 12.4

Gas, from a petroleum distillation column, has a concentration of H_2S reduced from 0.03 (kmol H_2S /kmol of inert hydrocarbon gas) to 1 per cent of this value by scrubbing with a triethanolamine–water solvent in a countercurrent tower, operating at 300 K and atmospheric pressure. The equilibrium relation for the solution may be taken as $Y_e = 2X$.

The solvent enters the tower free of H_2S and leaves containing 0.013 kmol of H_2S /kmol of solvent. If the flow of inert gas is 0.015 kmol/s m^2 of tower cross-section, calculate:

- (a) the height of the absorber necessary, and
- (b) the number of transfer units N_{OG} required.

The overall coefficient for absorption $K''_G a$ may be taken as 0.04 kmol/s m^3 (unit mole fraction driving force).

Solution

$$\text{Driving force at top of column} = (Y_2 - Y_{2e}) = 0.0003$$

$$\text{Driving force at bottom of column} = (Y_1 - Y_{1e}) = (0.03 - 0.026) = 0.004$$

$$\begin{aligned} \text{Logarithmic mean driving force} &= \frac{(0.004 - 0.0003)}{\ln\left(\frac{0.004}{0.0003}\right)} \\ &= 0.00143 \end{aligned}$$

$$\text{From equation 12.62: } G'_m(Y_1 - Y_2)S = K_G a P (Y - Y_e)_{\text{lm}} S Z$$

$$\text{That is: } G'_m(Y_1 - Y_2) = K'_G a (Y - Y_e)_{\text{lm}} Z$$

$$\text{Thus: } 0.015(0.03 - 0.0003) = 0.04 \times 0.00143 Z$$

$$\text{and: } Z = \frac{0.000446}{0.0000572} = 7.79 = 7.8 \text{ m (say)}$$

$$\begin{aligned} \text{Height of transfer unit } H_{OG} &= \frac{G'_m}{K'_G a} \\ &= \frac{0.015}{0.04} = 0.375 \text{ m} \end{aligned}$$

$$\text{Number of transfer units } N_{OG} = \frac{7.79}{0.375} = 20.7 = \underline{\underline{21}} \text{ (say)}$$

Example 12.5

Ammonia is to be removed from a 10 per cent ammonia–air mixture by countercurrent scrubbing with water in a packed tower at 293 K so that 99 per cent of the ammonia is removed when working at a total pressure of 101.3 kN/m².

If the gas rate is 0.95 kg/m²s of tower cross-section and the liquid rate is 0.65 kg/m²s, find the necessary height of the tower if the absorption coefficient $K_G a = 0.001 \text{ kmol/m}^3\text{s(kN/m}^2)$ partial pressure difference. The equilibrium data are:

kmol NH ₃ /kmol water:	0.021	0.031	0.042	0.053	0.079	0.106	0.159
Partial pressure NH ₃ :							
(mm Hg)	12.0	18.2	24.9	31.7	50.0	69.6	114.0
(kN/m ²)	1.6	2.4	3.3	4.2	6.7	9.3	15.2

Solution

If the compositions of the gas are given as per cent by volume, at the bottom of the tower are $y_1 = 0.10$ and $Y_1 = 0.10/(1 - 0.10) = 0.111$.

At the top of the tower: $y_2 = 0.001 \simeq Y_2$.

Mass flowrate of gas = 0.95 kg/m²s.

Mass per cent of air = $[0.9 \times 29 / (0.1 \times 17 + 0.9 \times 29)] \times 100 = 93.8$.

Thus: mass flowrate of air = $(0.938 \times 0.95) = 0.891 \text{ kg/m}^2\text{s}$

and: $G'_m = (0.891/29) = 0.0307 \text{ kmol/m}^2\text{s}$

$L'_m = 0.65/18 = 0.036 \text{ kmol/m}^2\text{s}$

A mass balance between a plane in the tower where the compositions are X and Y and the top of the tower gives:

$$G'_m(Y - Y_2) = L'_m(X - X_2)$$

But: $X_2 = 0$

Thus: $0.0307(Y - 0.001) = 0.036X$, or $Y = 1.173X + 0.001$

This is the equation of the operating line in terms of mole ratios.

The given equilibrium data may be converted to the same basis since:

$$P_G = yP = \frac{YP}{1 + Y}$$

and: $Y = \frac{P_G}{P - P_G}$

Using these equations, the following data are obtained:

kmol NH ₃ /kmol H ₂ O	0.021	0.031	0.042	0.053	0.079	0.106	0.159
Partial pressure P_G (mm)	12	18.2	24.9	31.7	50.0	69.6	114.0
$P - P_G = 760 - P_G$ (mm)	748	741.8	735.1	728.3	710	690.4	646
$Y = P_G / (P - P_G)$	0.016	0.0245	0.0339	0.0435	0.0704	0.101	0.176

These data are plotted in Figure 12.24.

From a mass balance over the column, the height Z is given by:

$$Z = \frac{G'_m}{k_G a P} \int_{Y_2}^{Y_1} \frac{(1 + Y)(1 + Y_i)}{(Y - Y_i)} dY \quad \text{(equation 12.45)}$$

Figure 12.25 may be used to evaluate the integral as follows:

Y	Y_i	$(1 + Y)(1 + Y_i)$	$\frac{(1 + Y)(1 + Y_i)}{(Y - Y_i)}$
0.111	0.089	1.21	55.0
0.10	0.078	1.185	53.8
0.08	0.059	1.14	54.3
0.06	0.042	1.11	61.4
0.04	0.027	1.067	82.0
0.02	0.013	1.035	148
0.01	0.006	1.016	254
0.005	0.0026	1.010	421
0.001	0	1.0	1000

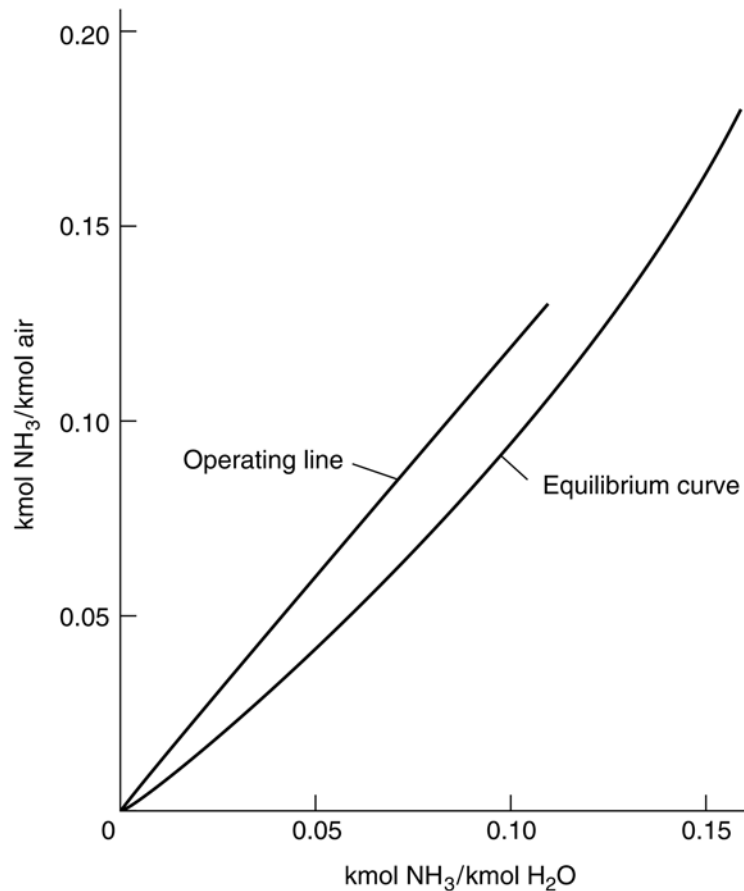


Figure 12.24. Operating and equilibrium lines for Example 12.5

The area under the curve in Figure 12.25 is 12.6. For a very soluble gas $k_{G^a} \simeq k_{G^a}$ so that:

$$Z = \frac{0.0307}{(0.001 \times 101.3)} \times 12.6 = \underline{\underline{3.82 \text{ m}}}$$

If the equilibrium line is assumed to be straight, then:

$$G'_m(Y_2 - Y_1) = K_G a Z \Delta P_{lm}$$

Top driving force = $\Delta Y_2 = 0.022$. Bottom driving force = $\Delta Y_1 = 0.001$.

Thus: $\Delta Y_{lm} = 0.0068$, $\Delta P_{lm} = 0.688 \text{ kN/m}^2$

and: $Z = \frac{(0.0307 \times 0.11)}{(0.001 \times 0.688)} = \underline{\underline{4.91 \text{ m}}}$.

12.8. PLATE TOWERS FOR GAS ABSORPTION

Bubble-cap columns or sieve trays, of similar construction to those described in Chapter 11 on distillation, are sometimes used for gas absorption, particularly when the load is more than can be handled in a packed tower of about 1 m diameter and when there is any

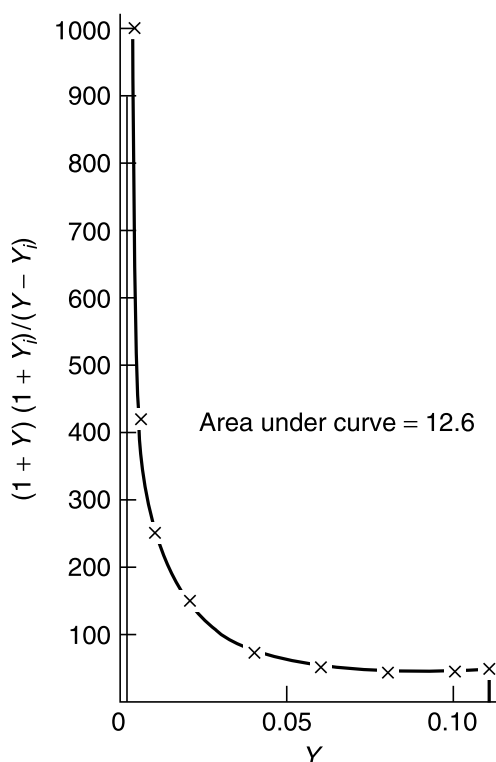


Figure 12.25. Determination of column height for Example 12.5

likelihood of deposition of solids which would quickly choke a packing. Plate towers are particularly useful when the liquid rate is sufficient to flood a packed tower. Since the ratio of liquid rate to gas rate is greater than with distillation, the slot area will be rather less and the downcomers rather larger. On the whole, plate efficiencies have been found to be less than with the distillation equipment, and to range from 20 to 80 per cent.

The plate column is a common type of equipment for large installations, although when the diameter of the column is less than 2 m, packed columns are more often used. For the handling of very corrosive fluids, packed columns are frequently preferred for larger units. The essential arrangement of such a unit is shown in Figure 12.26, where:

- L'_m is the molar rate of flow per unit area of solute free liquid,
- G'_m is the molar rate of flow per unit area of inert gas,
- n refers to the plate numbered from the bottom upwards (and suffix n refers to material leaving plate n),
- x is the mole fraction of the absorbed component in the liquid,
- y is the mole fraction of the absorbed component in the gas, and
- s is the total number of plates in the column.

It may be assumed that dilute solutions are used so that mole fractions and mole ratios are approximately equal. Each plate is taken as an “ideal” unit, so that the gas leaving of composition y_n is in equilibrium with the liquid of composition x_n leaving the plate.

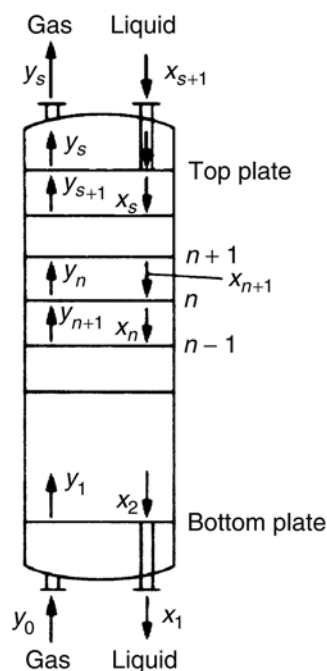


Figure 12.26. Plate tower—nomenclature for fluid streams

A material balance for the absorbed component from the bottom to a plane above plate n gives:

$$G'_m y_n + L'_m x_1 = G'_m y_0 + L'_m x_{n+1} \quad (12.84)$$

or:

$$y_n = \frac{L'_m}{G'_m} x_{n+1} + y_0 - \frac{L'_m}{G'_m} x_1 \quad (12.85)$$

This is the equation of a straight line of slope L'_m/G'_m , relating the composition of the gas entering a plate to the liquid leaving the plate, and is known as the *operating line*. As shown in Figure 12.27, such a line passes through two points B(x_{s+1}, y_s) and A(x_1, y_0), representing the terminal concentrations in the column. The equilibrium curve is shown in this figure as PQR.

Point A represents conditions at the bottom of the tower. The gas rising from the bottom plate is in equilibrium with a liquid of concentration x_1 and is shown as point 3 on the operating line. Then point 4 indicates the concentration of the liquid on the second plate from the bottom. In this way steps may be drawn to point B, giving the gas y_s rising from the top plate and the liquid x_{s+1} entering the top of the absorber.

12.8.1. Number of plates by use of absorption factor

If the equilibrium curve can be represented by the relation $y_e = mx$, then the number of plates required for a given degree of absorption can conveniently be found by a method due to KREMSER⁽⁵⁶⁾ and SOUDERS and BROWN⁽⁵⁷⁾. The same treatment is applicable for concentrated solutions provided concentrations are expressed as mole ratios, and if the equilibrium curve can be represented approximately by $Y_e = mX$.

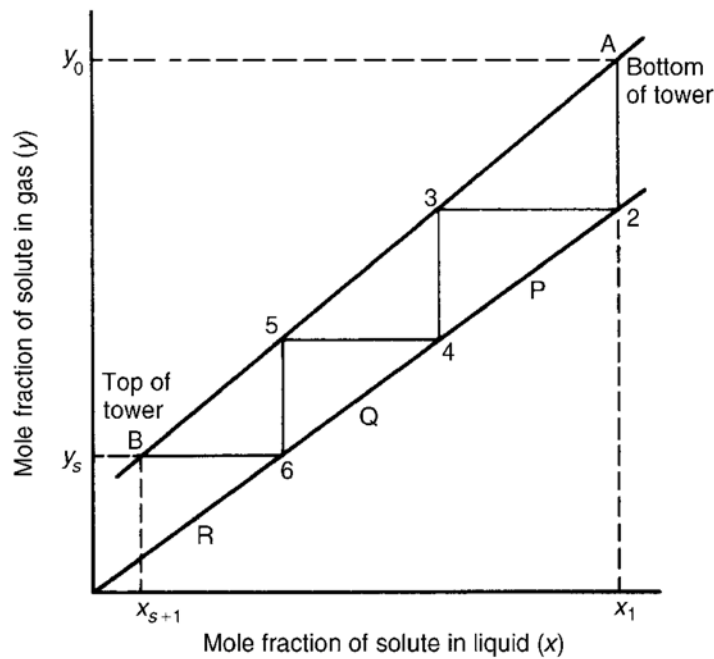


Figure 12.27. Diagrammatic representation of changes in a plate column

A material balance over plate n gives:

$$L'_m(x_n - x_{n+1}) = G'_m(y_{n-1} - y_n) \tag{12.86}$$

For an ideal plate, $y_n = mx_n$;

and:

$$\frac{L'_m}{mG'_m}(y_n - y_{n+1}) = y_{n-1} - y_n \tag{12.87}$$

This group L'_m/mG'_m , which will be taken as constant, is called the *absorption factor* \mathcal{A} .

Thus:

$$y_n = \frac{y_{n-1} + \mathcal{A}y_{n+1}}{1 + \mathcal{A}} \tag{12.88}$$

Applying this relation to the bottom plate and taking y_0 as the mole fraction of absorbed component in the gas entering the column, then:

$$y_1 = \frac{y_0 + \mathcal{A}y_2}{1 + \mathcal{A}}$$

And for the second plate from the bottom:

$$\begin{aligned} y_2 &= \frac{y_1 + \mathcal{A}y_3}{1 + \mathcal{A}} \\ &= \frac{\mathcal{A}(1 + \mathcal{A})y_3 + \mathcal{A}y_2 + y_0}{(1 + \mathcal{A})^2} \end{aligned}$$

Simplifying:

$$y_2 = \frac{y_0(1 + \mathcal{A}) + \mathcal{A}^2 y_3}{\mathcal{A}^2 + \mathcal{A} + 1}$$

And for the third plate from the bottom:

$$y_3 = \frac{y_0(1 + \mathcal{A} + \mathcal{A}^2) + \mathcal{A}^3 y_4}{\mathcal{A}^3 + \mathcal{A}^2 + \mathcal{A} + 1}$$

which may be written as:

$$\begin{aligned} y_3 &= \frac{[(\mathcal{A}^3 - 1)/(\mathcal{A} - 1)]y_0 + \mathcal{A}^3 y_4}{(\mathcal{A}^4 - 1)/(\mathcal{A} - 1)} \\ &= \frac{(\mathcal{A}^3 - 1)y_0 + \mathcal{A}^3(\mathcal{A} - 1)y_4}{\mathcal{A}^4 - 1} \end{aligned}$$

Proceeding thus until plate n is reached:

$$y_n = \frac{(\mathcal{A}^n - 1)y_0 + \mathcal{A}^n(\mathcal{A} - 1)y_{n+1}}{\mathcal{A}^{n+1} - 1}$$

$$y_0 = \frac{(\mathcal{A}^{n+1} - 1)y_n - \mathcal{A}^n(\mathcal{A} - 1)y_{n+1}}{\mathcal{A}^n - 1}$$

Thus:

$$y_0 - y_n = \frac{(\mathcal{A}^{n+1} - \mathcal{A}^n)y_n - \mathcal{A}^n(\mathcal{A} - 1)y_{n+1}}{\mathcal{A}^n - 1}$$

and:

$$y_0 - y_{n+1} = \frac{(\mathcal{A}^{n+1} - 1)y_n - (\mathcal{A}^{n+1} - 1)y_{n+1}}{\mathcal{A}^n - 1}$$

Dividing:

$$\begin{aligned} \frac{y_0 - y_n}{y_0 - y_{n+1}} &= \frac{(\mathcal{A}^{n+1} - \mathcal{A}^n)y_n - \mathcal{A}^n(\mathcal{A} - 1)y_{n+1}}{(\mathcal{A}^{n+1} - 1)y_n - (\mathcal{A}^{n+1} - 1)y_{n+1}} \\ &= \frac{\mathcal{A}^n(\mathcal{A} - 1)(y_n - y_{n+1})}{(\mathcal{A}^{n+1} - 1)(y_n - y_{n+1})} \end{aligned}$$

or:

$$\frac{y_0 - y_n}{y_0 - y_{n+1}} = \frac{\mathcal{A}^{n+1} - \mathcal{A}}{\mathcal{A}^{n+1} - 1}$$

Applying this relation over the whole column and putting $n = s$ gives:

$(y_0 - y_s)$ = actual change in composition of gas, and

$(y_0 - y_{s+1})$ = maximum possible change in composition of gas, that is if the gas leaving the absorber is in equilibrium with the entering liquid (or $y_s = mx_{s+1}$).

Then:

$$\frac{y_0 - y_s}{y_0 - mx_{s+1}} = \frac{(L'_m/mG'_m)^{s+1} - (L'_m/mG'_m)}{(L'_m/mG'_m)^{s+1} - 1} \quad (12.89)$$

This equation is conveniently represented, as suggested by SOUDERS and BROWN⁽⁵⁷⁾, by Figure 12.28, and it is easy to use such a diagram to determine the number of plates required.

A high degree of absorption can be obtained, either by using a large number of plates, or by using a high absorption factor L'_m/mG'_m . Since m is fixed by the system, this means that L'_m/G'_m must be large if a high degree of absorption is to be obtained, although this

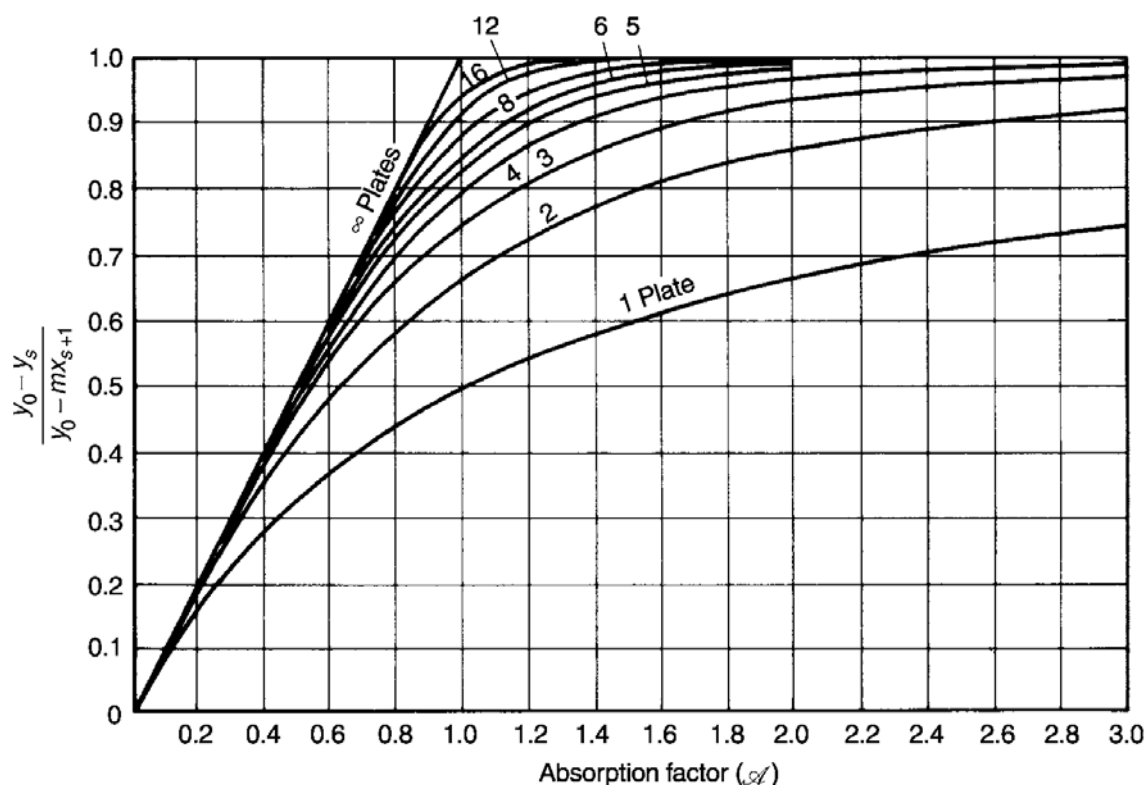


Figure 12.28. Graphical representation of the effect of the absorption factor and the number of plates on the degree of absorption

will result in a low value of x for the liquid leaving at the bottom. This problem is to some extent met by recirculating the liquid over the tower, although the advantages of a countercurrent flow system are then lost. A value of mG'_m/L'_m of about 0.7–0.8 is probably the most economic, that is L'_m/mG'_m 1.3.

It is important to note that, if L'_m/mG'_m is less than 1, then a very large number of plates are required to achieve a high recovery, and even an infinite number will not give complete recovery. L'_m/mG'_m is the ratio of the slope of the operating line L_m/G_m to the slope of the equilibrium curve m , so that if $L'_m/G'_m < m$, or $L'_m/mG'_m < 1$, then the operating line will never cut the equilibrium curve and the gas leaving the top of the column will not therefore reach equilibrium with the entering liquid.

12.8.2. Tray types for absorption

It has already been noted that trays which are suitable for distillation may be used for absorption duties though in general lower efficiencies will be obtained. In Chapter 11, the design of trays for common contacting devices is considered and the methods presented in that chapter are generally applicable. The most commonly used tray types are shown in Figure 11.50a with the crossflow tray being the most popular.

At high liquid flowrates, the liquid gradient on the tray can become excessive and lead to poor vapour distribution across the plate. This problem may be overcome by the shortening of the liquid flow-path as in the case of the double-pass and cascade trays. The whole design process is discussed in Volume 6.

Example 12.6

A bubble-cap column with 30 plates is to be used to remove n-pentane from a solvent oil by means of steam stripping. The inlet oil contains 6 kmol of n-pentane per 100 kmol of pure oil and it is desired to reduce the solute content to 0.1 kmol per 100 kmol of solvent. Assuming isothermal operation and an overall plate efficiency of 30 per cent, find the specific steam consumption, that is the kmol of steam required per kmol of solvent oil treated, and the ratio of the specific and minimum steam consumptions. How many plates would be required if this ratio were 2.0?

The equilibrium relation for the system may be taken as $Y_e = 3.0X$, where Y_e and X are expressed in mole ratios of pentane in the gas and liquid phases respectively.

Solution

Number of theoretical plates = $(30 \times 0.3) = 9$.

At the bottom of the tower:

$$\text{Flowrate of steam} = G'_m \text{ (kmol/m}^2\text{s)}$$

$$\text{Mole ratio of pentane in steam} = Y_1, \text{ and}$$

$$\text{Mole ratio of pentane in oil} = X_1 = 0.001$$

At the top of the tower:

$$\text{exit steam composition} = Y_2, \text{ inlet oil composition} = X_2 = 0.06,$$

$$\text{flowrate of oil} = L'_m \text{ (kmol/m}^2\text{s)}$$

The minimum steam consumption occurs when the exit steam stream is in equilibrium with the inlet oil, that is when:

$$Y_{e2} = (0.06 \times 3) = 0.18$$

$$L'_{\min}(X_2 - X_1) = G'_{\min}(Y_2 - Y_1)$$

If $Y_1 = 0$, that is the inlet steam is pentane-free, then:

$$L'_{\min}(0.06 - 0.001) = (G'_{\min} \times 0.18)$$

and:

$$(G'/L')_{\min} = (0.06 - 0.001)/0.18 = 0.328$$

The operating line may be fixed by trial and error as it passes through the point (0.001, 0), and 9 theoretical plates are required for the separation. Thus it is a matter of selecting the operating line which, with 9 steps, will give $X_2 = 0.001$ when $X_1 = 0.06$. This is tedious but possible, and the problem may be better solved analytically since the equilibrium line is straight.

Use may be made of the absorption factor method where

$$\frac{Y_1 - Y_2}{Y_1 - mX_2} = \frac{\mathcal{A}^{N+1} - \mathcal{A}}{\mathcal{A}^{N+1} - 1} \quad (\text{equation 12.89})$$

where \mathcal{A} is the absorption factor = L'_m/mG'_m and N is the number of theoretical plates.

The corresponding expression for a stripping operation is:

$$\frac{X_2 - X_1}{X_2 - Y_1/m} = \frac{(1/\mathcal{A})^{N+1} - (1/\mathcal{A})}{(1/\mathcal{A})^{N+1} - 1}$$

In this problem, $N = 9$, $X_2 = 0.06$, $X_1 = 0.001$, and $Y_1 = 0$

Thus:
$$\frac{(0.06 - 0.001)}{0.06} = 0.983 = \frac{(1/\mathcal{A})^{10} - (1/\mathcal{A})}{(1/\mathcal{A})^{10} - 1}$$
 from which $(1/\mathcal{A}) = 1.37$

Thus:
$$\frac{mG'_m}{L'_m} = 1.37, \frac{G'_m}{L'_m} = \frac{1.37}{3} = 0.457$$

and:
$$\frac{\text{actual } G'_m/L'_m}{\text{minimum } G'_m/L'_m} = \frac{0.457}{0.328} = \underline{\underline{1.39}}$$

If $(\text{actual } G'_m/L'_m)/(\text{min } G'_m/L'_m) = 2$, actual $G'_m/L'_m = 0.656$.

Thus:
$$1/\mathcal{A} = mG'_m/L'_m = 1.968$$

and:
$$0.983 = \frac{(1.968)^{N+1} - 1.968}{(1.968)^{N+1} - 1}$$
 from which $N = 4.9$

The actual number of plates = $(4.9/0.3) = 16.3$ (say 17).

12.9. OTHER EQUIPMENT FOR GAS ABSORPTION

12.9.1. The use of vessels with agitators

A gas may be dissolved in a liquid by dispersing it through holes in a pipe immersed in the liquid which is stirred with some form of agitator, as shown in Figure 12.29. Although this type of equipment will give only one theoretical stage per unit, but it often provides a useful method of saturating a liquid with a gas. COOPER *et al.*⁽⁵⁸⁾ have studied the absorption of oxygen from air in an aqueous solution of sodium sulphite using simple vessels of 0.15 to 2.44 m diameter fitted with four simple baffles. Air was just below the agitator which was a vaned-disc or flat-paddle. It was found that the absorption coefficient $K_G a$ varied almost directly with \mathbf{P}_V , the power input per unit volume. For constant values of \mathbf{P}_V , the following relation was obtained:

$$K_G a \propto u_s^{0.67} \tag{12.90}$$

where u_s is the superficial gas velocity based on the volume of gas at inlet and the cross-section of tank. A general correlation was obtained by plotting $K_G a/u_s^{0.67}$ against the power input per unit volume \mathbf{P}_V , as shown in Figure 12.30 taken from this investigation. AYERST and HERBERT⁽⁵⁹⁾ have given some data on the use of this type of unit for the absorption of carbon dioxide into ammoniacal solutions.

The interfacial area, a , was the subject of an investigation by WESTERTERP *et al.*⁽⁶⁰⁾ though the correlations proposed are complex. Maximum values of a are about $1000 \text{ m}^2/\text{m}^3$. Further work on the interfacial area in agitated vessels has been reviewed and summarised by SRIDAR and POTTER⁽⁶¹⁾ who found that the correlation of CALDERBANK⁽⁶²⁾ was applicable for most situations. Calderbank proposed that, for pure

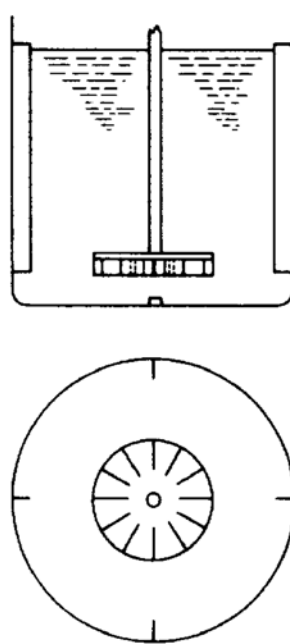


Figure 12.29. Vessel fitted with vaned-disc agitator

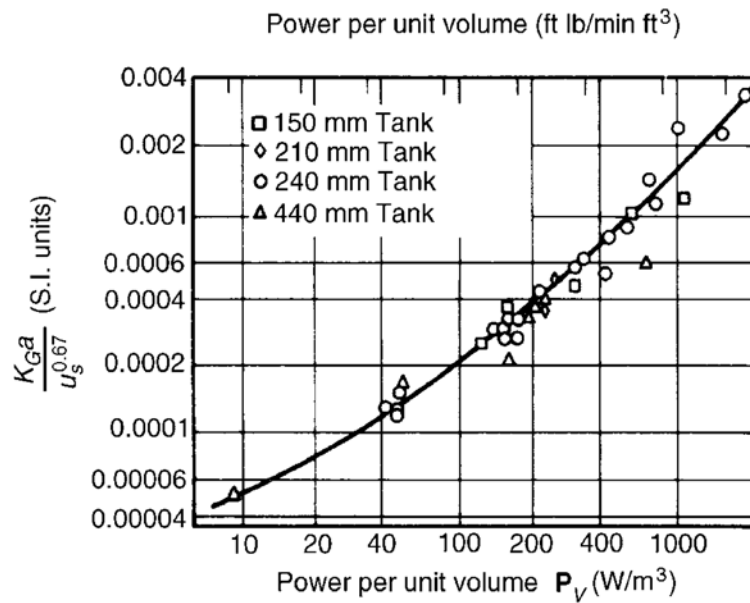


Figure 12.30. General correlation of data for a vessel (height = diameter) with vaned-disc agitator

liquids, the specific interfacial area, that is the surface area per unit volume of aerated suspension, $a(\text{m}^2/\text{m}^3)$ is given by:

$$a = 24,200 (\mathbf{P}_V)^{0.4} \left(\frac{\rho_L^{0.2}}{\sigma^{0.6}} \right) \left(\frac{u_s}{u_0} \right)^{0.5} \quad (12.91)$$

where surface aeration is negligible, that is when:

$$\left(\frac{N'd_t^2\rho_L}{\mu_L}\right)^{0.7}\left(\frac{N'd_i}{u_s}\right)^{0.3} < 25,000 \quad (12.92)$$

When the surface aeration is significant, then the interfacial area is:

$$\frac{a'}{a} = 10^{-4} \left\{ \left[\left(\frac{N'd_t^2\rho_L}{\mu_L}\right)^{0.7}\left(\frac{N'd_i}{u_s}\right)^{0.2} \right] - 25,000 \right\} \quad (12.93)$$

In these equations, a' is the specific interfacial area for a significant degree of surface aeration (m^2/m^3), \mathbf{P}_V is the agitator power per unit volume of vessel (W/m^3), ρ_L is the liquid density, σ is the surface tension (N/m), u_s is the superficial gas velocity (m/s), u_0 is the terminal bubble-rise velocity (m/s), N' is the impeller speed (Hz), d_i is the impeller diameter (m), d_t is the tank diameter (m), μ_L is the liquid viscosity (Ns/m^2) and d_0 is the Sauter mean bubble diameter defined in Chapter 1, Section 1.2.4.

The effects of gas hold-up and bubble diameter have also been studied by Sridhar and Potter and, again, the correlations obtained by Calderbank are recommended.

The liquid-phase mass transfer coefficient, k_L , in agitated vessels has been measured and data correlated by several workers. SIDEMAN *et al.*⁽⁶³⁾ and VALENTIN⁽⁶⁴⁾ have presented reviews of the early work and more recent work has been published by YAGI and YOSHIDA⁽⁶⁵⁾, ZLOKARNIK⁽⁶⁶⁾, VAN'T RIET⁽⁶⁷⁾ and HOKER, LANGER and UDO⁽⁶⁸⁾. For small bubbles (< 2.5 mm diameter) produced in well-agitated vessels, CALDERBANK⁽⁶²⁾ suggests the following correlation for bubbles in agitated electrolytes:

$$k_L = 0.31 \left(\frac{\Delta\rho\mu_L g}{\rho_L^2}\right)^{1/3} (Sc)^{-2/3} \quad (12.94)$$

where: $\Delta\rho$ = density difference between gas and liquid,
 ρ_L, μ_L = density and viscosity of the liquid, and
 Sc = Schmidt number for transport in the liquid.

JOSHI and SHARMA⁽⁶⁹⁾ and FUKADA *et al.*⁽⁷⁰⁾ have investigated the performance of vessels with multiple impellers on horizontal shafts.

Several investigations have been carried out into the power requirements for agitation of aerated liquids including those of YUNG *et al.*⁽⁷¹⁾ and LUONG and VOLESKY⁽⁷²⁾ and it is generally concluded that the power required is less for an aerated system than for a non-aerated system.

Although, as described by BJERLE *et al.*⁽⁷³⁾, liquid jet-type absorbers are also used, one relatively recent application of mass transfer in agitated tanks with chemical reaction is the absorption of pollutants from flue gases and, in particular, the scrubbing of sulphur dioxide by a slurry containing fine limestone particles. In this case, the concentration of sulphur dioxide is usually very low and the mechanism of the absorption is complicated due to the presence of solids in the liquid phase where the rate of solid dissolution may significantly affect the absorption rate.

Studies on the dissolution of solids in the liquid phase include that of HIXSON and BAUM⁽⁷⁴⁾ whose correlation of data in terms of Reynolds, Sherwood and Schmidt numbers, discussed in detail in Section 10.2 in connection with mass transfer during leaching, is one of the most frequently used methods for calculating the mass transfer coefficient for the solid dissolution.

Further work on the absorption of sulphur dioxide by UCHIDA *et al.*⁽⁷⁵⁾ has shown that the absorption rate changes with the surface area of the limestone particles which in turn varies with the size and the number of particles, and that the rate of dissolution plays a very important role on the absorption. It was further found the absorption rate does not vary significantly with temperature and that the reactions involved may be considered as being instantaneous.

12.9.2. The centrifugal absorber

In an attempt to obtain the benefits of repeated spray formations, a centrifugal type absorber has been developed from the ideas of Piazza for a still head. The principle of the unit is shown in Figure 12.31. A set of stationary concentric rings intermeshes with a second set of rings attached to a rotating plate. Liquid fed to the centre of the plate

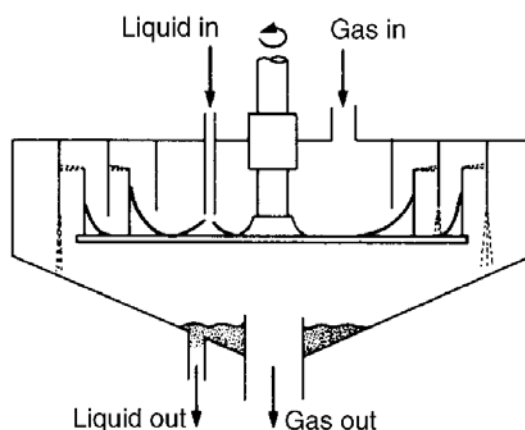


Figure 12.31. The centrifugal absorber

is carried up the first ring, splashes over to the baffle and falls into the through between the rings. It then runs up the second ring and in a similar way passes from ring to ring through the unit. The gas stream can be introduced at the top to give cocurrent flow, or at the bottom if countercurrent flow is desired. Some of the features of this unit are discussed by AHMED⁽⁷⁶⁾ who found that the depth of the ring was not very important and that most of the transfer took place as the gas mixed with the liquid spray leaving the top of the rings. CHAMBERS and WALL⁽⁷⁷⁾ have given some particulars of the performance of the 510 mm diameter unit shown in Figure 12.32, for the absorption of carbon dioxide

from air containing 10–15 per cent of carbon dioxide, using mono-ethanolamine solution. Some values of absorption rates are given in Table 12.7.

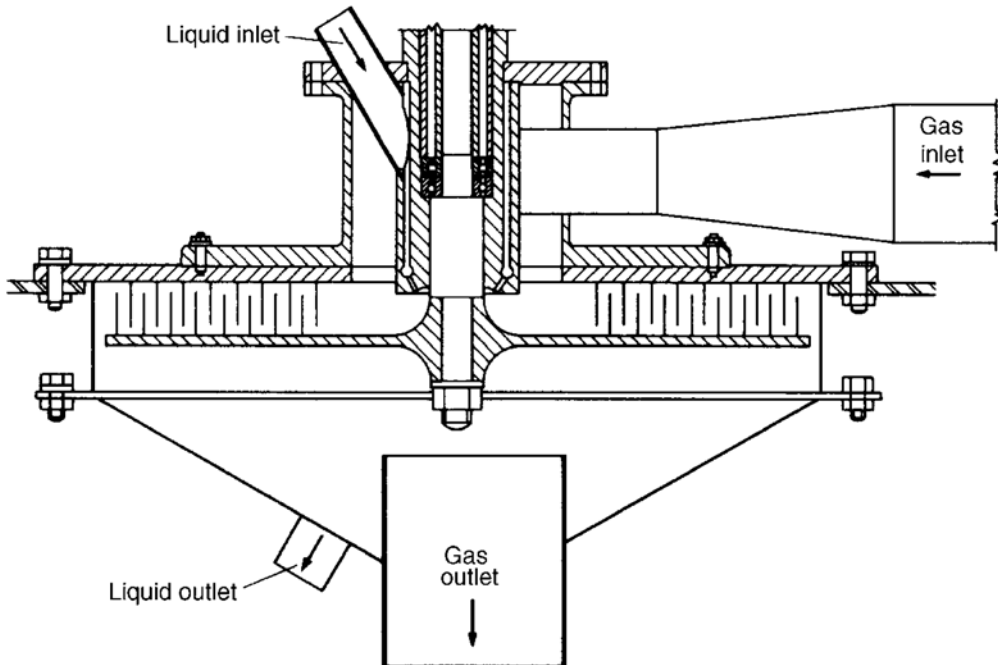


Figure 12.32. Details of a 510 mm diameter centrifugal absorber

Table 12.7. Results for absorption in a 510 mm diameter absorber

Gas flow (m ³ /s)	Liquid flow (m ³ /s)	per cent CO ₂ in gas		Absorption rate (kg/s)
		in	out	
0.016	1.07 × 10 ⁻⁴	16.3	2.3	0.0044
0.024	1.07 × 10 ⁻⁴	15.8	4.5	0.0055
0.031	1.07 × 10 ⁻⁴	14.3	6.6	0.0051
0.039	1.07 × 10 ⁻⁴	16.3	8.7	0.0065

12.9.3. Spray towers

In the spray tower, the gas enters at the bottom and the liquid is introduced through a series of sprays at the top. The performance of these units is generally rather poor, because the droplets tend to coalesce after they have fallen through a few metres, and the interfacial surface is thereby seriously reduced. Although there is considerable turbulence in the gas phase, there is little circulation of the liquid within the drops, and the resistance of the equivalent liquid film tends to be high. Spray towers are therefore useful only where the

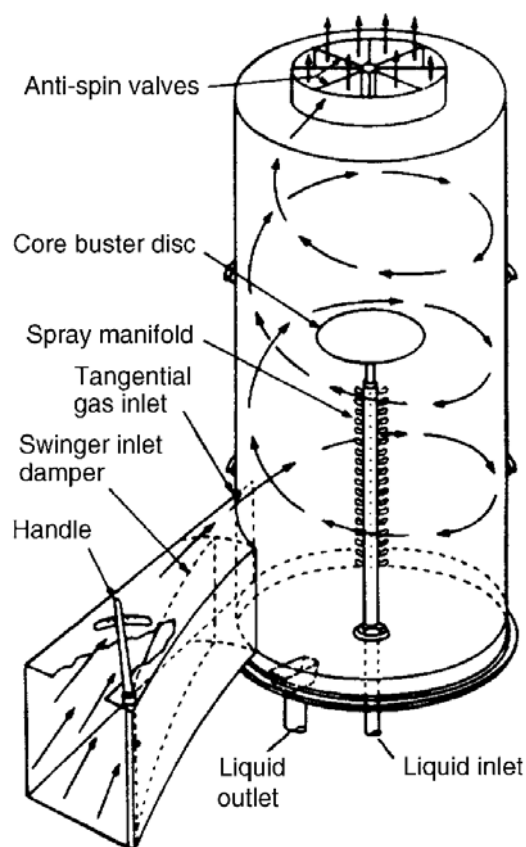


Figure 12.33. Centrifugal spray tower⁽⁷⁸⁾

main resistance to mass transfer lies within the gas phase, and have consequently been used with moderate success for the absorption of ammonia in water. They are also used as air humidifiers, in which case the whole of the resistance lies within the gas phase.

Centrifugal spray tower

Figure 12.33, taken from the work of KLEINSCHMIDT and ANTHONY⁽⁷⁸⁾, illustrates a spray tower in which the gas stream enters tangentially, so that the liquid drops are subjected to centrifugal force before they are taken out of the gas stream at the top.

12.10. FURTHER READING

- HOBBLER, T.: *Mass Transfer and Absorbers* (Pergamon Press, Oxford, 1966).
 MCCABE, W. L., SMITH, J. C. and HARRIOTT, P.: *Unit Operations of Chemical Engineering*, 4th edn. (McGraw-Hill, New York, 1984).
 NORMAN, W. S.: *Absorption, Distillation and Cooling Towers* (Longmans, London, 1961).
 SHERWOOD, T. K. and PIGFORD, R. L.: *Absorption and Extraction* (McGraw-Hill Book Co., New York, 1952).
 SHERWOOD, T. K., PIGFORD, R. L. and WILKE, C. R.: *Mass Transfer* (McGraw-Hill Book Co., New York, 1975).
 SMITH, B. D.: *Design of Equilibrium Stage Processes* (McGraw-Hill Book Co., New York, 1963).
 TREYBAL, R. E.: *Mass Transfer Operations*, 3rd edn. (McGraw-Hill Book Co., New York, 1980).

- WANKAT, P. C.: *Equilibrium Staged Separations: Separations for Chemical Engineers* (Elsevier, New York, 1988).
- ZARZYCKI, R. and CHACUK, A.: *Absorption. Fundamentals and Applications* (Pergamon Press, Oxford, 1993).
- ZENZ, F. A.: *Design of Gas Absorption Towers*. In SCHWEITZER, P. A.: *Handbook of Separation Techniques for Chemical Engineers* 2nd edn. (McGraw Hill, New York, 1988).

12.11. REFERENCES

1. WHITMAN, W. G.: *Chem. Met. Eng.* **29** (1923) 147. The two-film theory of absorption.
2. HIGBIE, R.: *Trans. Am. Inst. Chem. Eng.* **31** (1935) 365. The rate of absorption of pure gas into a still liquid during short periods of exposure.
3. DANCKWERTS, P. V.: *Ind. Eng. Chem.* **43** (1951) 1460. Significance of liquid-film coefficients in gas absorption.
4. DANCKWERTS, P. V. and KENNEDY, A. M.: *Trans. Inst. Chem. Eng.* **32** (1954) S49. Kinetics of liquid-film processes in gas absorption.
5. LYNN, S., STRAATEMEIER, J. R., and KRAMERS, H.: *Chem. Eng. Sci.* **4** (1955) 49, 58, 63. Absorption studies in the light of the penetration theory. I. Long wetted-wall columns. II. Absorption by short wetted-wall columns. III. Absorption by wetted-spheres, singly and in columns.
6. DAVIDSON, J. F., CULLEN, E. J., HANSON, D., and ROBERTS, D.: *Trans. Inst. Chem. Eng.* **37** (1959) 122. The hold-up and liquid film coefficient of packed towers. Part I. Behaviour of a string of spheres.
7. DAVIDSON, J. F.: *Trans. Inst. Chem. Eng.* **37** (1959) 131. The hold-up and liquid film coefficient of packed towers. Part II: Statistical models of the random packing.
8. DANCKWERTS, P. V. and KENNEDY, A. M.: *Chem. Eng. Sci.* **8** (1958) 201. The kinetics of absorption of carbon dioxide into neutral and alkaline solutions.
9. ROBERTS, D. and DANCKWERTS, P. V.: *Chem. Eng. Sci.* **17** (1962) 961. Kinetics of CO₂ absorption in alkaline solutions. I. Transient absorption rates and catalysis by arsenite.
10. DANCKWERTS, P. V., KENNEDY, A. M. and ROBERTS, D.: *Chem. Eng. Sci.* **18** (1963) 63. Kinetics of CO₂ absorption in alkaline solutions. II. Absorption in a packed column and tests of surface renewal models.
11. CULLEN, E. J. and DAVIDSON, J. F.: *Trans. Faraday Soc.* **53** (1957) 113. Absorption of gases in liquid jets.
12. RAIMONDI, P. and TOOR, H. L.: *A.I.Ch.E.Jl.* **5** (1959) 86. Interfacial resistance in gas absorption.
13. STERNLING, C. V. and SCRIVEN, L. E.: *A.I.Ch.E.Jl.* **5** (1959) 514. Interfacial turbulence: Hydrodynamic instability and the Marangoni effect.
14. GOODRIDGE, F. and ROBB, I. D.: *Ind. Eng. Chem. Fundamentals* **4** (1965) 49. Mechanism of interfacial resistance.
15. GARNER, F. H. and KENDRICK, P.: *Trans. Inst. Chem. Eng.* **37** (1959) 155. Mass transfer to drops of liquid suspended in a gas stream. Part I—A wind tunnel for the study of individual liquid drops.
16. GARNER, F. H. and LANE, J. J.: *Trans. Inst. Chem. Eng.* **37** (1959) 162. Mass transfer to drops of liquid suspended in a gas stream. Part II: Experimental work and results.
17. KING, C. J.: *A.I.Ch.E.Jl.* **10** (1964) 671. The additivity of individual phase resistances in mass transfer operations.
18. GILLIAND, E. R. and SHERWOOD, T. K.: *Ind. Eng. Chem.* **26** (1934) 516. Diffusion of vapours into air streams.
19. HOLLINGS, H. and SILVER, L.: *Trans. Inst. Chem. Eng.* **12** (1934) 49. The washing of gas.
20. CHILTON, T. H. and COLBURN, A. P.: *Ind. Eng. Chem.* **26** (1934) 1183. Mass transfer (absorption) coefficients—prediction from data on heat transfer and fluid friction.
21. MORRIS, G. A. and JACKSON, J.: *Absorption Towers* (Butterworths, London, 1953).
22. KOWALKE, O. L., HOUGEN, O. A., and WATSON, K. M.: *Bull. Univ. Wisconsin Eng. Sta. Ser. No. 68* (1925). Transfer coefficients of ammonia in absorption towers.
23. BORDEN, H. M. and SQUIRES, W.: Massachusetts Institute of Technology, S. M. thesis (1937). Absorption of ammonia in a ring-packed tower. (cited in Reference 30).
24. NORMAN, W. S.: *Trans. Inst. Chem. Eng.* **29** (1951) 226. The performance of grid-packed towers.
25. FELLINGER, L. L.: Massachusetts Institute of Technology. D.Sc. thesis (1941). Absorption of ammonia by water and acids in various standard packings.
26. PERRY, R. H., GREEN, D. W., and MALONEY, J. O. (eds.): *Perry's Chemical Engineers' Handbook*. 7th edn. (McGraw-Hill Book Company, New York, 1997).
27. MOLSTAD, M. C., MCKINNEY, J. F. and ABBEY, R. G.: *Trans. Am. Inst. Chem. Eng.* **39** (1943) 605. Performance of drip-point grid tower packings, III. Gas-film mass transfer coefficients: additional liquid-film mass transfer coefficients.

28. VAN KREVELEN, D. W. and HOFTHIJZER, P. J.: *Rec. Trav. Chim* **67** (1948) 563. Kinetics of gas-liquid reactions. Part I. General theory.
29. SEMMELBAUER, R.: *Chem. Eng. Sci.* **22** (1967) 1237. Die Berechnung der Schütthöhe bei Absorptionsvorgängen in Füllkörperkolonnen. (Calculation of the height of packing in packed towers.)
30. SHERWOOD, T. K. and HOLLOWAY, F. A. L.: *Trans. Am. Inst. Chem. Eng.* **36** (1940). Performance of packed towers. 21—Experimental studies of absorption and desorption, 39, 181—liquid film data for several packings.
31. COOPER, C. M., CHRISTL, R. J., and PEERY, L. C.: *Trans. Am. Inst. Chem. Eng.* **37** (1941) 979. Packed tower performance at high liquor rates—The effect of gas and liquor rates upon performance in a tower packed with two-inch rings.
32. NONHEBEL, G.: *Gas Purification Processes for Air Pollution Control*, 2nd edn. (Newnes-Butterworth, London, 1972).
33. HIXSON, A. W. and SCOTT, C. E.: *Ind. Eng. Chem.* **27** (1935) 307. Absorption of gases in spray towers.
34. WHITMAN, W. G., LONG, L., and WANG, H. Y.: *Ind. Eng. Chem.* **18** (1926) 363. Absorption of gases by a liquid drop.
35. PIGFORD, R. L. and PYLE, C.: *Ind. Eng. Chem.* **43** (1951) 1649. Performance characteristics of spray-type absorption equipment.
36. NORMAN, W. S.: *Absorption, Distillation and Cooling Towers* (Longmans, London, 1961).
37. HATTA, S.: *Tech. Repts. Tohoku Imp. Univ.* **10** (1932) 119. On the absorption velocity of gases by liquids. II. Theoretical considerations of gas absorption due to chemical reaction.
38. NIJSING, R. A. T. O., HENDRIKSZ, R. H. and KRAMERS, H.: *Chem. Eng. Sci.* **10** (1959) 88. Absorption of CO₂ in jets and falling films of electrolyte solutions, with and without chemical reaction.
39. TEPE, J. B. and DODGE, B. F.: *Trans. Am. Inst. Chem. Eng.* **39** (1943) 255. Absorption of carbon dioxide by sodium hydroxide solutions in a packed column.
40. CRYDER, D. S. and MALONEY, J. O.: *Trans. Am. Inst. Chem. Eng.* **37** (1941) 827. The rate of absorption of carbon dioxide in diethanolamine solutions.
41. STEPHENS, E. J. and MORRIS, G. A.: *Chem. Eng. Prog.* **47** (1951) 232. Determination of liquid-film absorption coefficients. A new type of column and its application to problems of absorption in presence of chemical reaction.
42. DANCKWERTS, P. V. and MCNEIL, K. M.: *Trans. Inst. Chem. Eng.* **45** (1967) 32. The absorption of carbon dioxide into aqueous amine solutions and the effects of catalysis.
43. DANCKWERTS, P. V. and SHARMA, M. M.: *Chem. Engr. London* No. **202** (Oct. 1966) CE244. The absorption of carbon dioxide into solutions of alkalis and amines (with some notes on hydrogen sulphide and carbonyl sulphide).
44. SAHAY, B. N. and SHARMA, M. M.: *Chem. Eng. Sci.* **28** (1973) 41. Effective interfacial areas and liquid and gas side mass transfer coefficients in a packed column.
45. ECKERT, J. S.: *Chem. Eng.* **82** (14 April 1975) 70. How tower packings behave.
46. SHERWOOD, T. K., PIGFORD, R. L., and WILKE, C. R.: *Mass Transfer* (McGraw-Hill Book Company, New York, 1980).
47. TREYBAL, R. E.: *Mass Transfer Operations*, 3rd edn. (McGraw-Hill Book Co., New York, 1980).
48. POLL, A. and SMITH, W.: *Chem. Engng.* **71** (26 Oct. 1964) 111. Froth contact heat exchanger.
49. COGGAN, C. G. and BOURNE, J. R.: *Trans. I. Chem. E.* **47** (1969) T96, T160. The design of gas absorbers with heat effects.
50. SHULMAN, H. L., ULLRICH, C. F., PROULX, A. Z. and ZIMMERMAN, J. O.: *A.I.Ch.E.Jl.* **1** (1955) 2, 253. Interfacial areas—gas and liquid phase mass transfer rates.
51. EASTHAM, I. E.: Private communication (1977).
52. CHILTON, T. H. and COLBURN, A. P.: *Ind. Eng. Chem.* **27** (1935) 255. Distillation and absorption in packed columns.
53. RACKETT, H. G.: *Chem. Eng. Albany* **71** (21 Dec. 1964) 108. Modified graphical integration for determining transfer units.
54. Norton Chemical Process Products Div., Box 350, Akron, Ohio; Hydronyl Ltd., King St., Fenton, Stoke-on-Trent, U.K.
55. COLBURN, A. P.: *Trans. Am. Inst. Chem. Eng.* **35** (1939) 211. The simplified calculation of diffusional processes. General consideration of two-film resistances.
56. KREMSER, A.: *Nat. Petroleum News* **22** (21 May 1930) 43. Theoretical analysis of absorption processes.
57. SOUDERS, M. and BROWN, G. G.: *Ind. Eng. Chem.* **24** (1932) 519. Fundamental design of high pressure equipment involving paraffin hydrocarbons. IV. Fundamental design of absorbing and stripping columns for complex vapours.
58. COOPER, C. M., FERNSTROM, G. A., and MILLERS, S. A.: *Ind. Eng. Chem.* **36** (1944) 504. Performance of agitated gas-liquid contactors.
59. AYERST, R. R. and HERBERT, L. S.: *Trans. Inst. Chem. Eng.* **32** (1954) S68. A study of the absorption of carbon dioxide in ammonia solutions in agitated vessels.

60. WESTERTERP, K. R., VAN DIERENDONCK, L. L. and DE KRAA, J. R.: *Chem. Eng. Sci.* **18** (1963) 157. Interfacial areas in agitated gas-liquid contactors.
61. SRIDHAR, T. and POTTER, O.E: *Chem. Eng. Sci.* **35** (1980) 683. Interfacial areas in gas-liquid stirred vessels.
62. CALDERBANK, P. H.: *Chem. Engnr.* No. 212 (Oct. 1967) CE 209. Gas absorption from bubbles.
63. SIDEMAN, S. O., HORTACSU, O., and FULTON, J. W.: *Ind. Eng. Chem.* **58** (July 1966) 32. Mass transfer in gas-liquid contacting systems.
63. SIDEMAN, S. O., HORTACSU, O. and FULTON, J. W.: *Ind. Eng. Chem.* **58** (July 1966) 32. Mass transfer in gas-liquid contacting systems.
64. VALENTIN, F. H. H.: *Brit. Chem. Eng.* **12** (1967) 1213. Mass transfer in agitated tanks.
65. YAGI, H. and YOSHIDA, F.: *Ind. Eng. Chem. Proc. Des. Dev.* **14** (1975) 488. Gas absorption by Newtonian and non-Newtonian fluids in sparged agitation vessels.
66. ZLOKARNIK, M.: *Adv. Biochem. Eng.* **8** (1978) 133. Sorption characteristics for gas-liquid contacting in mixing vessels.
67. VAN'T RIET, K.: *Ind. Eng. Chem. Proc. Des. Dev.* **18** (1979) 357. Review of measuring methods and results in nonviscous gas-liquid mass transfer in stirred tanks.
68. HÖCKER, H., LANGER, G. and UDO, W.: *Germ. Chem. Eng.* **4** (1981) 51. Mass transfer in aerated Newtonian and non-Newtonian liquids.
69. JOSHI, J. B. and SHARMA, M. M.: *Can. J. Chem. Eng.* **54** (1976) 460. Mass transfer characteristics of horizontal agitated contactors.
70. FUKUDA, H., IDOGAWA, K., IKEDA, K. and ENDOH, K.: *J. Chem. Eng. Japan* **13** (1980) 298. Volumetric gas-phase mass transfer coefficients in baffled horizontal stirred tanks.
71. YUNG, C. H., WONG, C. W. and CHANG, C. L.: *Can. J. Chem. Eng.* **59** (1979) 672. Gas holdup and aerated power consumption in mechanically stirred tanks.
72. LUONG, H. T. and VOLESKY, B.: *AIChE Jl.* **25** (1970) 893. Mechanical power requirements of gas-liquid agitated systems.
73. BJERLE, I., BENGTSOON, S. and FÄRNKVIST, K.: *Chem. Eng. Sci.* **27** (1972) 1853. Absorption of SO₂ in CaCO₃-slurry in a laminar jet absorber.
74. HIXSON, A. W. and BAUM, S. J.: *Ind. Eng. Chem.* **33** (1941) 478. Agitation: heat and mass transfer coefficients in liquid-solid systems.
75. UCHIDA, S., MORIGUCHI, H., MAEJIMA, H., KOIDE, K. and KAGEYAMA, S.: *Can. J. Chem. Eng.* **56** (1978) 690. Absorption of sulphur dioxide into limestone slurry in a stirred tank reactor.
76. AHMED, N.: University of London, Ph.D. thesis (1949). Design of gas scrubber based upon thin films and sprays.
77. CHAMBERS, H. H. and WALL, R. C.: *Trans. Inst. Chem. Eng.* **32** (1954) S96. Some factors affecting the design of centrifugal gas absorbers.
78. KLEINSCHMIDT, R. V. and ANTHONY, A. W.: *Trans. Am. Soc. Mech. Eng.* **63** (1941) 349. Recent development of Pease-Anthony gas scrubber.

12.12. NOMENCLATURE

		Units in SI System	Dimensions in M, L, T θ
A	Cross-sectional area of column	m ²	L ²
\mathcal{A}	Absorption factor	—	—
a	Surface area of interface per unit volume of column	m ² /m ³	L ⁻¹
$a_1, a_2 \dots$	Constants in equation 12.39	—	—
a'	Specific surface area (equation 12.94)	m ⁻¹	L ⁻¹
B	A constant in equation 12.23	—	—
B'	A constant in equation 12.24	—	—
C	Molar concentration	kmol/m ³	NL ⁻³
C_A, C_B	Molar concentrations of A, B	kmol/m ³	NL ⁻³
C_{AL}, C_{BL}	Molar concentrations of A, B in bulk of liquid phase	kmol/m ³	NL ⁻³
C_{Ae}	Molar concentration of A in liquid phase in equilibrium with partial pressure P_{AG} in gas phase	kmol/m ³	NL ⁻³

		Units in SI System	Dimensions in M, L, T θ
C_{Ai}	Molar concentration of A at interface	kmol/m ³	NL ⁻³
C_{AL}	Molar concentration of A in bulk of liquid	kmol/m ³	NL ⁻³
C_T	Total molar concentration	kmol/m ³	NL ⁻³
c	Constant term in equation of equilibrium line	—	—
c_G	Gas mixture constant $(\rho_r/\mu_r)^{0.25}/$ $(D_{Vr})^{0.5}$ in cgs units	[(cm ² /s) ^{-3/4}]	L ^{-3/2} T ^{3/4}
D_L	Liquid phase diffusivity	m ² /s	L ² T ⁻¹
D_V	Vapour phase diffusivity	m ² /s	L ² T ⁻¹
d	Column diameter	m	L
d_i	Impeller diameter	m	L
d_0	Sauter mean diameter	m	L
d_p	Packing size	m	L
d_t	Tank diameter	m	L
e	Voidage	—	—
F'	Fractional conversion (equation 12.39)	—	—
f	Fraction of surface renewed per unit time	s ⁻¹	T ⁻¹
G'_m	Molar rate of flow of inert gas per unit cross-section	kmol/m ² s	NL ⁻² T ⁻¹
G'	Gas flowrate (mass) per unit cross-section	kg/m ² s	ML ⁻² T ⁻¹
H	Height of transfer unit	m	L
h	Heat transfer coefficient	W/m ² K	MT ⁻³ θ ⁻¹
h_D	Mass transfer coefficient (D_V/z_G)	m/s	LT ⁻¹
h_p	Height of packing	m	L
\mathcal{H}	Henry's constant	(N/m ²)/(kmol/m ³)	MN ⁻¹ L ² T ⁻²
i	Number of mole of B reacting with 1 mole of A	—	—
j_d	j -factor for mass transfer	—	—
K_G	Overall gas-phase transfer coefficient	s/m	L ⁻¹ T
K_L	Overall liquid-phase transfer coefficient	m/s	LT ⁻¹
K''_G	Overall gas-phase transfer coefficient in terms of mole fractions	kmol/m ² s	NL ⁻² T ⁻¹
K''_L	Overall liquid-phase transfer coefficient in terms of mole fractions	kmol/m ² s	NL ⁻² T ⁻¹
k	Thermal conductivity	W/m K	MLT ⁻³ θ ⁻¹
k_G	Gas-film transfer coefficient $(D_V P/RT z_G P_{Bm})$	s/m	L ⁻¹ T
k'_G	Gas-film transfer coefficient $(D_V/RT z_G)$	s/m	L ⁻¹ T
k''_G	Gas-film transfer coefficient in terms of mole fractions	kmol/m ² s	NL ⁻² T ⁻¹
k_L	Liquid-film transfer coefficient	m/s	LT ⁻¹
k''_L	Liquid-film transfer coefficient in terms of mole fractions	kmol/m ² s	NL ⁻² T ⁻¹
k_2	Reaction rate constant for second-order reaction	m ³ /kmols	N ⁻¹ L ³ T ⁻¹
L'_m	Molar rate of flow of solute-free liquor per unit cross-section	kmol/s m ²	NL ⁻² T ⁻¹
L_v	Volumetric liquid rate	m ³ /s	L ³ T ⁻¹
L'	Liquid flowrate (mass) per unit cross-section	kg/s m ²	ML ⁻² T ⁻¹
m	Slope of equilibrium line	—	—
N_A, N_B	Molar rate of diffusion of A, B per unit area	kmol/s m ²	NL ⁻² T ⁻¹

		Units in SI System	Dimensions in M, L, T θ
N'_A, N'_B	Molar rate of absorption of A, B per unit area	kmol/s m ²	NL ⁻² T ⁻¹
N''_A	Molar rate of absorption of A per unit area with chemical reaction	kmol/s m ²	NL ⁻² T ⁻¹
N	Number of transfer units	—	—
N'	Impeller speed	s ⁻¹ , (Hz)	T ⁻¹
<i>n</i>	Number of plates from bottom	—	—
<i>P</i>	Total pressure	N/m ²	ML ⁻¹ T ⁻²
P_A, P_B	Partial pressures of A and B	N/m ²	ML ⁻¹ T ⁻²
P_{Bm}	Logarithmic mean value of P_B	N/m ²	ML ⁻¹ T ⁻²
P_{Ae}	Partial pressure of A in equilibrium with concentration C_{AL} in liquid phase	N/m ²	ML ⁻¹ T ⁻²
P_{AG}	Partial pressure of A in bulk of gas phase	N/m ²	ML ⁻¹ T ⁻²
P_{Ai}	Partial pressure of A at interface	N/m ²	ML ⁻¹ T ⁻²
ΔP_{Alm}	Log mean driving force for A	N/m ²	ML ⁻¹ T ⁻²
P_V	Power input per unit volume	W/m ³	ML ⁻¹ T ⁻³
R	Universal gas constant	J/kmol K	NM ⁻¹ L ² T ⁻² θ ⁻¹
<i>r</i>	Ratio of effective film thickness for absorption without and with chemical reaction	—	—
<i>S</i>	Specific surface of packing	m ⁻¹	L ⁻¹
<i>s</i>	Total number of plates in column	—	—
<i>T</i>	Absolute temperature	K	θ
<i>t</i>	Time	s	T
<i>u</i>	Gas velocity	m/s	LT ⁻¹
u_0	Terminal rise velocity	m/s	LT ⁻¹
u_s	Superficial gas velocity (based on inlet conditions)	m/s	LT ⁻¹
<i>V</i>	Volume of packed section of column	m ³	L ³
<i>X</i>	Moles of solute gas A per mole of solvent in liquid phase	—	—
<i>x</i>	Mole fraction of A in liquid phase	—	—
<i>Y</i>	Molar ratio of solute gas A to inert gas B in gas phase	—	—
<i>y</i>	Mole fraction of A in gas phase	—	—
<i>Z</i>	Height of packed column	m	L
<i>z</i>	Distance of direction of mass transfer	m	L
z_G	Thickness of gas film	m	L
z_L	Thickness of liquid film	m	L
α	A coefficient in equation 12.28	s ^{1.8} /kg ^{0.8} m ^{0.4}	M ^{-0.8} L ^{-0.4} T ^{1.8}
β	A coefficient	1/m ^{1.25}	L ^{-5/4}
β'	A coefficient (equation 12.33)	—	—
μ	Viscosity of gas	Ns/m ²	ML ⁻¹ T ⁻¹
μ_L	Viscosity of liquid	Ns/m ²	ML ⁻¹ T ⁻¹
ρ	Density of gas	kg/m ³	ML ⁻³
ρ_L	Density of liquid	kg/m ³	ML ⁻³
σ	Surface tension	J/m ²	MT ⁻²
ϕ	Correction factor for concentrated solutions	—	—
<i>Ga</i>	Galileo number	—	—
<i>Pr</i>	Prandtl number	—	—
<i>Re</i>	Reynolds number	—	—
<i>Sc</i>	Schmidt number	—	—
<i>Sh</i>	Sherwood number	—	—

Units in
SI SystemDimensions
in **M, L, T θ**

Suffixes

1	denotes conditions at bottom of packed column, or at plane 1
2	denotes conditions at top of packed column, or at plane 2
<i>A</i>	denotes soluble gas
<i>B</i>	denotes insoluble gas
<i>e</i>	denotes equilibrium value
<i>f</i>	denotes film value
<i>i</i>	denotes value at interface
<i>G</i>	denotes gas phase
<i>L</i>	denotes liquid phase
lm	denotes logarithmic mean value
<i>n</i>	denotes values on plate <i>n</i>
<i>r</i>	denotes reference state (293 K, 101.3 kN/m ²)

LG, OG, L, OL refer to gas film, overall gas, liquid film, and overall liquid transfer units