

# Absorption (Chemical Engineering)

**James R. Fair**

*University of Texas at Austin*

**Henry Z. Kister**

*Fluor-Daniel Corp.*

- I. Absorption in Practice
- II. Principles of Absorption
- III. Models for Absorption Equipment
- IV. Absorber Design

## GLOSSARY

**Absorption factor** Ratio of liquid to gas flow rate divided by the slope of the equilibrium curve.

**Films** Regions on the liquid and gas sides of the interface in which fluid motion is considered slow and through which material is transported by molecular diffusion alone.

**Gas solubility** Quantity of gas dissolved in a given quantity of solvent at equilibrium conditions.

**Hatta number** Ratio of the maximum conversion of reacting components into products in the liquid film to the maximum diffusion transport through the liquid film.

**Height of a transfer unit** Vertical height of a contactor required to give a concentration change equivalent to one transfer unit.

**Ideal stage** Hypothetical device in which gas and liquid are perfectly mixed, are contacted for a sufficiently long

period of time so that equilibrium is obtained, and are then separated.

**Inerts** Gas components that are not absorbed by the liquid.

**Interface** Surface separating the liquid from the gas. Equilibrium is assumed to exist at this surface.

**LPG** Liquefied petroleum gas.

**Lean gas** Gas leaving the absorber, containing the inerts and little or no solute.

**Lean solvent** Solvent entering the absorber, containing little or no solute.

**Mass transfer coefficient** Quantity describing the rate of mass transfer per unit interfacial area per unit concentration difference across the interface.

**Number of transfer units** Parameter that relates the change in concentration to the average driving force. It is a measure of the ease of separation by absorption.

**Operating line** Line on the  $y$ - $x$  diagram that represents the locus of all the points obeying the component material balance.

**Rich gas** Gas entering the absorber, containing both the inerts and solutes.

**Rich solvent** Solvent leaving the absorber, which contains solute removed from the feed gas.

**Slope of equilibrium curve** Ratio of the change of the solute concentration in the gas to a given change in solute concentration in the liquid when the solvent and solute are at equilibrium and when solute concentrations are expressed as mole fractions.

**Solute(s)** Component(s) absorbed from the gas by the liquid

**Solvent** Dissolving liquid used in an absorption process.

**Stripping (or desorption)** Process in which the absorbed gas is removed from the solution.

**$y$ - $x$  diagram** Plot in which the solute mole fraction in the gas is plotted against the solute mole fraction in the liquid.

**ABSORPTION** is a unit operation in which a gas mixture is contacted with a suitable liquid for the purpose of preferentially dissolving one or more of the constituents of the gas. These constituents are thus removed or partially removed from the gas into the liquid. The dissolved constituents may either form a physical solution with the liquid or react chemically with the liquid. The dissolved constituents are termed *solutes*, while the dissolving liquid is termed the *solvent*. When the concentration of solute in the feed gas is low, the process is often called *scrubbing*.

The inverse operation, called stripping, desorption, or regeneration, is employed when it is desirable to remove the solutes from the solvent in order to recover the solutes or the solvent or both.

## I. ABSORPTION IN PRACTICE

### A. Commercial Application

Absorption is practiced for the following purposes:

1. Gas purification, for example, removal of pollutants from a gas stream.
2. Production of solutions, for example, absorption of hydrogen chloride gas in water to form hydrochloric acid.
3. Product recovery, for example, absorption of liquified petroleum gases (LPG) and gas olines from natural gas.
4. Drying, for example, absorption of water vapor from a natural gas mixture.

Some common commercial applications of absorption are listed in [Table I](#).

### B. Choice of Solvent for Absorption

If the main purpose of absorption is to generate a specific solution, as in the manufacture of hydrochloric acid, the solvent is specified by the nature of the product. For all other purposes, there is some choice in selecting the absorption liquid. The main solvent selection criteria are as follows:

1. Gas solubility. Generally, the greater the solubility of the solute in the solvent, the easier it is to absorb the gas, reducing the quantity of solvent and the equipment size needed for the separation. Often, a solvent that is chemically similar to the solute or that reacts chemically with the solute will provide high gas solubility.
2. Solvent selectivity. A high selectivity of the solvent to the desired solutes compared with its selectivity to other components of the gas mixture lowers the quantity of undesirable components dissolved. Application of a solvent of higher selectivity reduces the cost of downstream processing, which is often required to separate out the undesirable components.
3. Volatility. The gas leaving the absorber is saturated with the solvent. The more volatile the solvent is, the greater are the solvent losses; alternatively, the more expensive are the down-stream solvent separation facilities required to reduce the losses.
4. Effects on product and environment. For example, toxic solvents are unsuitable for food processing; noxious solvents are unsuitable when the gas leaving the absorber is vented to the atmosphere.
5. Chemical stability. Unstable solvents may be difficult to regenerate or may lead to excessive losses due to decomposition.
6. Cost and availability. The less expensive is the solvent, the lower is the cost of solvent losses. Water is the least expensive and most plentiful solvent.
7. Others. Noncorrosiveness, low viscosity, nonflammability, and low freezing point are often desirable properties.

### C. Absorption Processes

Absorption is usually carried out in a countercurrent tower, through which liquid descends and gas ascends. The tower may be fitted with trays, filled with packing, or fitted with sprays or other internals. These internals provide the surface area required for gas-liquid contact.

A schematic flow diagram of the absorption-stripping process is shown in [Fig. 1](#). Lean solvent enters at the top

**TABLE I Common Commercial Applications of Gas Absorption**

Type of plant	Feed gas	Solutes	Solvent	Commercial purpose	Stripping practice
Refineries, natural gas plants, petrochemical plants, coal processing plants, hydrogen plants	Refinery gas, natural gas, LPG towns gas, coal gas, hydrogen reformer gas	Hydrogen sulfide, carbon dioxide, mercaptans	Ethanolamines, alkaline solutions, potassium carbonate	Gas purification for downstream processing or to achieve product specifications	Stripping practiced when using ethanolamines or carbonate for the purpose of solvent recovery and recycle; stripping normally not practiced when using an alkaline solution
Combustion plants	Combustion gases	Sulfur dioxide	Water, alkaline solutions	Pollutant removal	Stripping normally not practiced
Natural gas plants	Natural gas	Water	Glycol	Gas drying for further processing or to achieve product specification	Stripping practiced for solvent recovery
Refineries, natural gas plants, petrochemical plants	Gas stream containing mostly hydrogen, methane, and light gases as well as some LPG and gasolines	LPG, gasolines	Kerosene, diesel, gas oil, other refinery oils	Product recovery of LPG, gasolines	Stripping practiced for LPG and gasoline recovery
Coke ovens	Coke oven gas	Benzene, toluene	Heavy oil	By-product recovery	Stripping practiced to recover the by-product
Sulfuric acid	Sulfur trioxide mixed with oxygen and nitrogen	Sulfur trioxide	Sulfuric acid, oleum	Sulfuric acid manufacture	Stripping not practiced
Nitric acid	Nitrogen dioxide mixed with nitrogen oxide, oxygen, nitrogen	Nitrous oxides	Nitric acid, water	Nitric acid manufacture	Stripping not practiced
Carbon dioxide	Combustion gases, kiln gases	Carbon dioxide	Carbonate, bicarbonate solution	Carbon dioxide production	Stripping practiced to recover carbon dioxide
Hydrochloric acid	By-products of chlorination reaction	Hydrogen chloride	Hydrochloric acid, water	Hydrochloric acid production	Stripping not practiced
Soda ash (sodium carbonate), mineral processing	Combustion gases, lime-kiln gases	Carbon dioxide	Ammonia solution	Ammonium bicarbonate production, ammonium carbonate production	Stripping not practiced
Soda ash (sodium carbonate), mineral processing	Waste gases, ammonia makeup	Ammonia	Brine solution	Production of ammonium hydroxide for ammonium bicarbonate production	Stripping not practiced
Hydrogen cyanide	Tail gases, ammonia, hydrogen cyanide	Ammonia	Sulfuric acid	Ammonia removal while producing ammonium sulfate by-product	Stripping not practiced
Hydrogen cyanide, acrylonitrile	Hydrogen cyanide, tail gases, acrylonitrile	Hydrogen cyanide acrylonitrile	Water	Separation of hydrogen cyanide and acrylonitrile from tail gases	Stripping is practiced to recover hydrogen cyanide and acrylonitrile from water

*Continues*

TABLE I (continued)

Type of plant	Feed gas	Solutes	Solvent	Commercial purpose	Stripping practice
Ethylene oxide, glycol	Reactor effluent	Ethylene oxide	Water	Ethylene oxide recovery	Stripping is practiced to recover ethylene oxide from the solution
Ketones from alcohol	Hydrogen, ketones	Ketones	Water	Ketone–hydrogen separation	Stripping is practiced to recover ketones from the solution
Maleic anhydride	Reactor effluent	Maleic anhydride separation	Water	Maleic anhydride from reactor gases	Stripping is practiced to remove water from the maleic acid formed in the absorption process, converting it back to maleic anhydride
Isoprene	Reactor effluent	Isoprene, C <sub>4</sub> 's, C <sub>5</sub> 's	Heavy oil	Separation of C <sub>4</sub> 's, C <sub>5</sub> 's, and isoprene from light gases	Stripping is practiced to recover the solute and regenerate the oil for recycling to the absorbent
Urea	Reactor effluent	CO <sub>2</sub> , NH <sub>3</sub>	Water	Formation of ammonium carbonate solution, which is recycled to the reactor	Stripping not practiced

of the absorber and flows downward through the internals. Rich gas enters at the bottom of the absorber and flows upward through the internals. The liquid and gas are contacted at the absorber internals, and the solute is absorbed by the solvent. Overhead product from the absorber is the

solute-free lean gas, and bottom product is the rich solvent, which contains the absorbed solute. The rich solvent then flows to the stripper where the solute is stripped from the rich solvent, this operation being at a higher temperature and/or lower pressure than maintained in the absorber. The solute leaves the stripper as the overhead product, and the solute-free lean solvent leaves the stripper bottom and is recycled to the absorber.

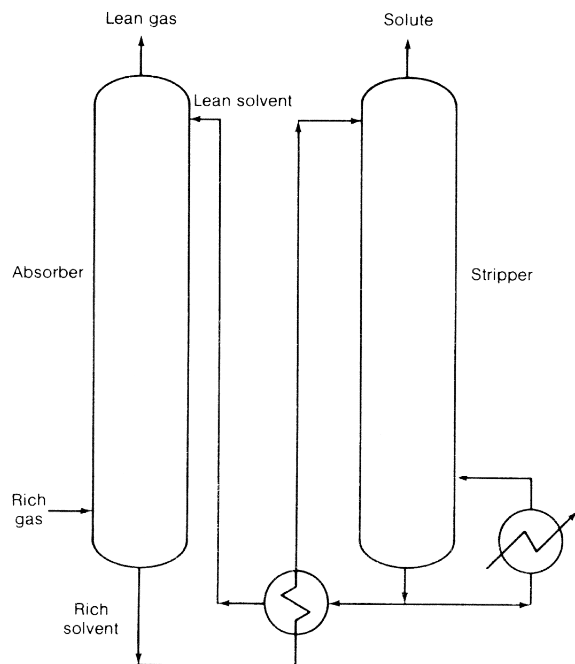


FIGURE 1 Typical schematic absorber–stripper flow diagram.

## II. PRINCIPLES OF ABSORPTION

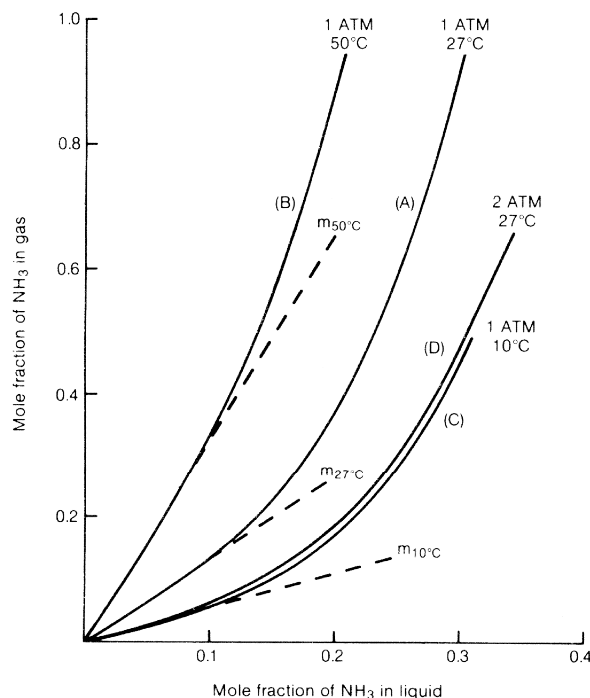
The important fundamental physical principles in absorption are solubility and mass transfer. When a chemical reaction is involved, the principles of reaction equilibria and reaction kinetics are also important.

### A. Gas Solubility

At equilibrium, the fugacity of a component in the gas is equal to the fugacity of the same component in the liquid. This thermodynamic criterion defines the relationship between the equilibrium concentration of a component in the gas and its concentration in the liquid. The quantity of gas dissolved in a given quantity of solvent at equilibrium conditions is often referred to as the gas solubility.

Gas solubility data are available from handbooks and various compendia and often show solubility as a function of gas composition, temperature and pressure. A typical





**FIGURE 2** Solubility data for  $\text{NH}_3$  absorption from air using  $\text{H}_2\text{O}$ . [Data from Perry, R. H., ed. (1985). "Chemical Engineer's Handbook," McGraw-Hill, New York.]

graphical presentation is shown in Fig. 2, where gas composition of a given solute is plotted against liquid composition of the same solute, at equilibrium. Compositions can be represented in various units, such as mole fraction, mole ratio, partial pressure (gas). Figure 2 shows the effect of temperature and pressure on solubility. Solubility is also dependent on whether the solute reacts chemically with the solvent as well as on the nature and amounts of other solutes present.

The equilibrium curve is often approximated linearly,

$$y_A = mx_A \quad (1a)$$

where  $m$  is a constant at a given temperature and pressure. This expression is often valid at low concentrations (Fig. 2).

For a solution that is thermodynamically ideal,  $m$  is given by "Raoult's law"

$$m = p^{vap} P \quad (1b)$$

or the ratio of vapor pressure to total pressure. When the gas composition is expressed as partial pressure, the Henry's law coefficient for a given solute is

$$H = p/x \quad (1c)$$

or

$$m = H/P \quad (1d)$$

Henry's law is usually a reasonable approximation at low and moderate concentrations, at constant temperature, and at relatively low pressures (generally less than 5 atm; however, the law may be obeyed at higher pressure at low solubilities).

If a gas mixture containing several components is in equilibrium with a liquid, Henry's law applies separately so long as the liquid is dilute in all the components. If a component is almost insoluble in the liquid, for example, air in water, it has a very high Henry's law constant and a high value of  $m$  in Eq. (1). Such a component is absorbed in negligible quantities or by the liquid, and it is often referred to as an *inert component*. The nature and type of the inert component have little effect on the equilibrium curve.

Equilibrium data for absorption are usually available in the literature in three forms:

1. Solubility data, expressed either as mole percent, mass percent, or Henry's law constants
2. Pure-component vapor pressure data
3. Equilibrium distribution coefficients ( $K$  values)

To define fully the solubility of a component in a liquid, it is necessary to state the temperature, the partial pressure of the solute in the gas, the concentration of the solute in the liquid, and generally also the pressure.

When gas solubility data are lacking or are unavailable at the desired temperature, they can be estimated using available models. The method of Prausnitz and Shair (1961), which is based on regular solution theory and thus has the limitations of that theory. The applicability of regular solution theory is covered in detail by Hildebrand *et al.* (1970). A more recent model, now widely used, is UNIFAC, which is based on structural contributions of the solute and solvent molecular species. This model is described by Fredenslund *et al.* (1977) and extensive tabulations of equilibrium data, based on UNIFAC, have been published by Hwang *et al.* (1992) for aqueous systems where the solute concentrations are low and the solutions depart markedly from thermodynamic equilibrium.

Perhaps the best source of information on estimating gas solubility is the book by Reid *et al.* (1987), which not only lists the various solubility models but also compares them with a database of experimental measurements.

## B. Mass Transfer Principles

The principles of mass transfer determine the rate at which the equilibrium is established, that is, the rate at which the solute is transferred into the solvent.

For a system in equilibrium, no net transfer of material occurs between the phases. When a system is not in equilibrium, diffusion of material between the phases will occur so as to bring the system closer to equilibrium. The departure from equilibrium provides the driving force for diffusion of material between the phases.

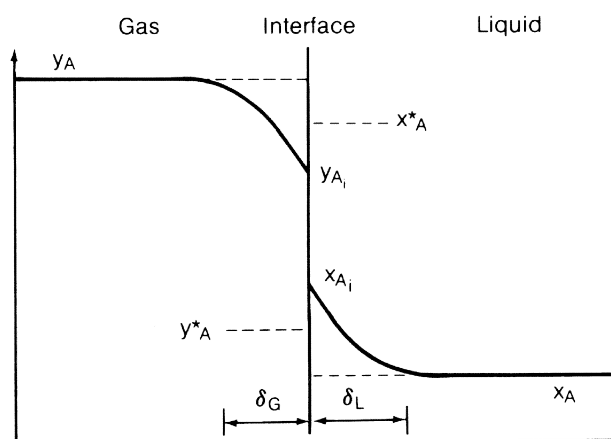
The rate of diffusion can be described by the film theory, the penetration theory, or a combination of the two. The most popular description is in terms of a two-film theory. Accordingly, there exists a stable interface separating the gas and liquid. A certain distance from the interface, large fluid motions exist; and these distribute the material rapidly and equally, so that no concentration gradients develop. Next to the interface, however, there are regions in which the fluid motion is slow; in these regions, termed *films*, material is transferred by diffusion alone. At the interface, material is transferred instantaneously, so that the gas and liquid are in equilibrium at the interface. The rate-governing step in absorption is therefore the rate of diffusion in the gas and liquid films adjacent to the interface. The concentration gradient in both phases are illustrated in Fig. 3. Note that  $y_{Ai}$  may be higher or lower than  $x_{Ai}$ , depending on the equilibrium curve (e.g., Fig. 2); however,  $y_{Ai}$  is always lower than  $y_A$ , and  $x_{Ai}$  is always higher than  $x_A$ , or no mass transfer will occur.

### 1. Dilute Solutions

Applying the diffusion equations to each film and approximating the concentration gradient linearly yields an expression for the mass transfer rates across the films,

$$N_A = k_G(y_A - y_{Ai}) = k_L(x_{Ai} - x_A) \quad (2)$$

This equation states that, for each phase, the rate of mass transfer is proportional to the difference between the bulk concentration and the concentration at the gas-liquid interface.



**FIGURE 3** Concentration profiles in the vapor and liquid phases near an interface.

terface. Here  $k_G$  and  $k_L$  are the mass transfer coefficients, and their reciprocals,  $1/k_G$  and  $1/k_L$  are measures of the resistance to mass transfer in the gas and liquid phases, respectively. Note that the rate of mass transfer in the gas film is equal to that in the liquid film; otherwise, material will accumulate at the interface.

The concentration difference in the gas can be expressed in terms of partial pressures instead of mole fractions, while that in the liquid can be expressed in moles per unit volume. In such cases, an equation similar to Eq. (2) will result. Mole fraction units, however, are generally preferred because they lead to gas mass transfer coefficients that are independent of pressure.

It is convenient to express the mass transfer rate in terms of a hypothetical bulk-gas  $y_A^*$ , which is in equilibrium with the bulk concentration of the liquid phase, that is,

$$N_A = K_{OG}(y_A - y_A^*) \quad (3)$$

If the equilibrium curve is linear, as described by Eq. (1), or can be linearly approximated over the relevant concentration range, with an average slope  $m$  such that

$$m = (y_A - y_A^*) / (x_A^* - x_A) \quad (4)$$

then Eqs. (2)–(4) can be combined to express  $K_{OG}$  in terms of  $k_G$  and  $k_L$ , as follows:

$$\frac{1}{K_{OG}} = \frac{1}{k_G} + \frac{m}{k_L} \quad (5)$$

Equation (5) states that the overall resistance to mass transfer is equal to the sum of the mass transfer resistances in each of the phases.

The use of overall coefficients is convenient because it eliminates the need to calculate interface concentrations. Note that, theoretically, this approach is valid only when a linear approximation can be used to describe the equilibrium curve over the relevant concentration range. Figure 4 illustrates the application of this concept on an  $x$ - $y$  diagram.

For most applications it is not possible to quantify the interfacial area available for mass transfer. For this reason, data are commonly presented in terms of mass transfer coefficients based on a unit volume of the apparatus. Such volumetric coefficients are denoted  $k_G a$ ,  $k_L a$  and  $K_{OG} a$ , where  $a$  is the interfacial area per unit volume of the apparatus.

If most of the resistance is known to be concentrated in one of the phases, the resistance in the other phase can often be neglected and Eq. (5) simplified. For instance, when hydrogen chloride is absorbed in water, most of the resistance occurs in the gas phase, and  $K_{OG} \approx k_G$ . When oxygen is absorbed in water, most of the resistance occurs in the liquid phase, and  $K_{OG} \approx k_L / m$ .

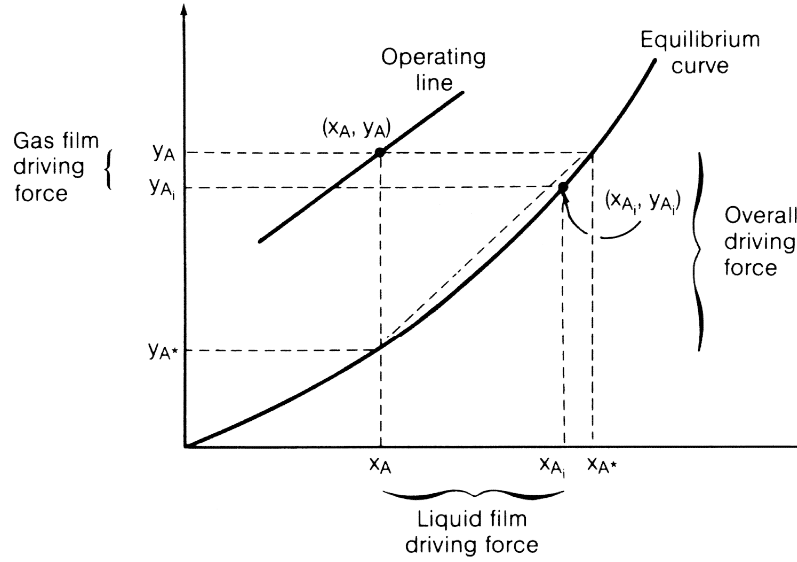


FIGURE 4 Absorption driving forces in terms of the  $x$ - $y$  diagram.

## 2. Concentrated Solutions

Equation (2), derived for dilute solutions, is valid when the flow of solute from the gas to the gas film is balanced by an equal flow of the inert component from the film to the gas; similarly, it requires that the flow of solute from the liquid film to the solvent be balanced by an equal flow of solvent from the liquid into the liquid film. This is a good approximation when both the gas and the liquid are dilute solutions. If either or both are concentrated solutions, the flow of gas out of the film, or the flow of liquid into the film, may contain a significant quantity of solute. These solute flows counteract the diffusion process, thus increasing the effective resistance to diffusion.

The equations used to describe concentrated solutions are derived in texts by [Sherwood \*et al.\* \(1975\)](#), [Hobler \(1966\)](#), and [Hines and Maddox \(1985\)](#). These reduce to Eqs. (2) and (3) when applied to dilute solutions. These equations are as follows:

$$N_A = k'_G(y_A - y_{Ai})/y_{BM} = k'_L(x_{Ai} - x_A)/x_{BM} = K'_{OG}(y_A - y_A^*)/y_{BM}^*, \quad (6a)$$

where

$$y_{BM} = \frac{(1 - y_A) - (1 - y_{Ai})}{\ln[(1 - y_A)/(1 - y_{Ai})]} \quad (6b)$$

$$x_{BM} = \frac{(1 - x_A) - (1 - x_{Ai})}{\ln[(1 - x_A)/(1 - x_{Ai})]} \quad (6c)$$

$$y_{BM}^* = \frac{(1 - y_A) - (1 - y_A^*)}{\ln[(1 - y_A)/(1 - y_A^*)]} \quad (6d)$$

The terms subscripted BM describe the log-mean solvent or log-mean inert gas concentration difference between the bulk fluid and the interface [Eqs. (6b) and (6c)] or between the bulk fluid and the equilibrium values [Eq. (6d)].

Equation (6a) is analogous to Eqs. (2) and (3). Comparison of these shows that, in concentrated solutions, the concentration-independent coefficients of Eqs. (2) and (3) are replaced by concentration-dependent coefficients in Eq. (6a) such that

$$k_G = k'_G y_{BM} \quad (7a)$$

$$k_L = k'_L x_{BM} \quad (7b)$$

$$K_{OG} = K'_{OG} y_{BM}^* \quad (7c)$$

## 3. Multicomponent Absorption

The principles involved in multicomponent absorption are similar to those discussed for concentrated solutions. [Wilke \(1950\)](#) developed a set of equations similar to Eq. (6a) to represent this case,

$$N_A = k''_G(y_A - y_{Ai})/y_{fm} = k''_L(x_{Ai} - x_A)/x_{fm} = K''_{OG}(y_A - y_A^*)/y_{fm}^*, \quad (8a)$$

where

$$y_{fm} = \frac{(1 - t_A y_A) - (1 - t_A y_{Ai})}{\ln[(1 - t_A y_A)/(1 - t_A y_{Ai})]} \quad (8b)$$

$$x_{fm} = \frac{(1 - t_A x_A) - (1 - t_A x_{Ai})}{\ln[(1 - t_A x_A)/(1 - t_A x_{Ai})]} \quad (8c)$$

$$y_{fm}^* = \frac{(1 - t_A y_A) - (1 - t_A y_A^*)}{\ln[(1 - t_A y_A)/(1 - t_A y_A^*)]} \quad (8d)$$

$$t_A = \frac{N_A + N_B + N_C + \dots}{N_A} \quad (8e)$$

In a manner similar to the concentrated solutions case, the coefficients in Eqs. (8) can be expressed in terms of the concentration-independent coefficients using relationships similar to those of Eqs. (7), that is,

$$k_G = k_G'' y_{fm} \quad (9a)$$

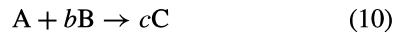
$$k_L = k_L'' x_{fm} \quad (9b)$$

$$K_{OG} = K_{OG}'' y_{fm}^* \quad (9c)$$

#### 4. Absorption with Chemical Reaction

When the solute is absorbed into a solution containing a reagent that chemically reacts with it, the concentration profile shown in Fig. 3 becomes dependent on the kinetics of the reaction and the concentration of the reacting reagent in the liquid.

Figure 5 shows concentration profiles that commonly occur when solute A undergoes an irreversible second-order reaction with component B, dissolved in the liquid, to give product C,



The rate equation is

$$r_A = k_2 C_A C_B; \quad r_B = br_A \quad (11)$$

Figure 5 shows that a fast reaction takes place only in the liquid film. In such instances, the dominant mass transfer mechanism is physical absorption and the diffusion model above is applicable, but the resistance to mass transfer in the liquid phase is lower because of the reaction. On the other hand, a slow reaction occurs in the bulk of the liquid, and its rate has little dependence on the resistances to diffusion in either the gas or liquid film. Here the dominant mass transfer mechanism is that of chemical reaction; therefore, this case is considered part of chemical reaction technology, as distinct from absorption technology.

The Hatta number  $Ha$  is a dimensionless group used to characterize the speed of reaction in relation to the diffusional resistance to mass transfer,

$$Ha = \frac{\text{max. possible conversion in the liquid film}}{\text{max. diffusional transport through the liquid film}} \\ = \frac{D_A k_2 C_{B0}}{(k_L^\circ)^2} \quad (12)$$

When  $Ha \gg 1$ , all the reaction occurs in the film, and the process is that of absorption with chemical reaction. As in the case of absorption with no reaction, the main consideration is to provide sufficient surface area for diffusion. On the other hand, when  $Ha \ll 1$ , all the reaction occurs

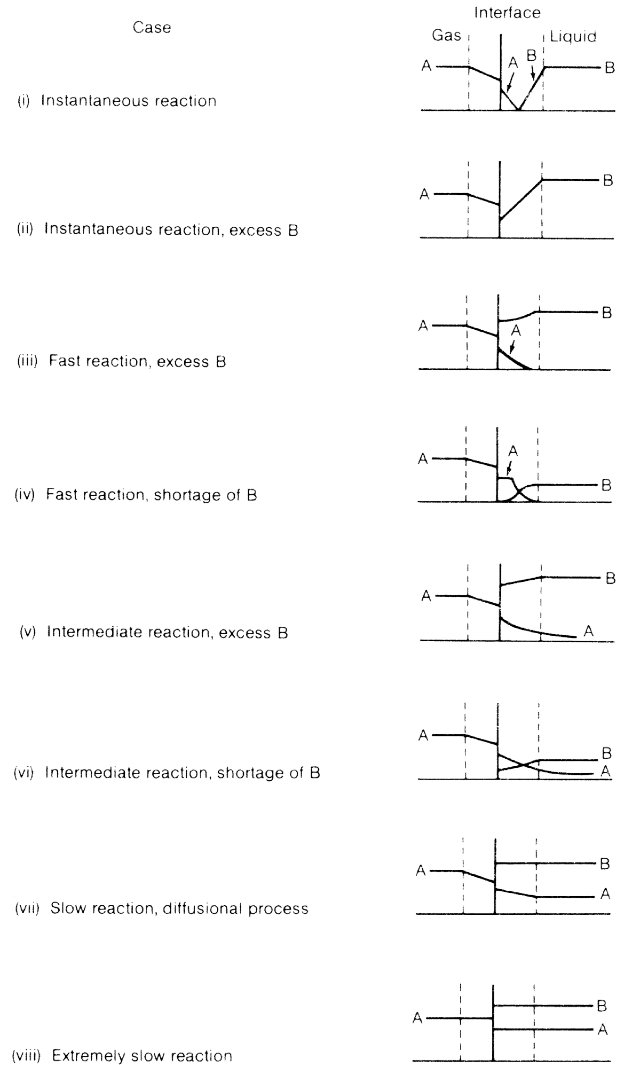


FIGURE 5 Vapor- and liquid-phase concentration profiles near an interface for absorption with chemical reaction.

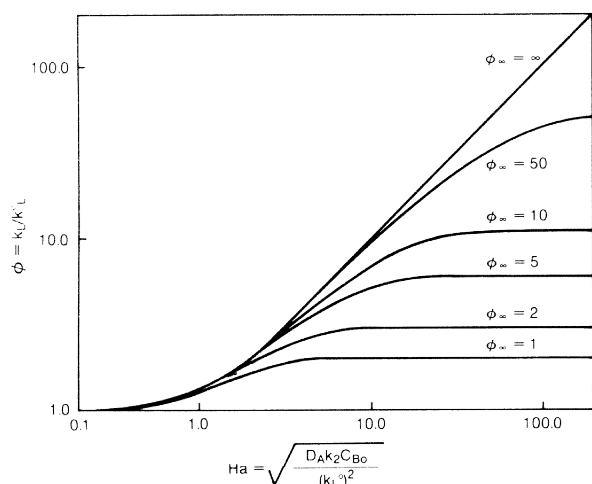
in the bulk of the liquid, and the contactor behaves as a reactor, not an absorber. Here, the main consideration is providing sufficient liquid holdup for the reaction to take place.

The effect of chemical reaction on rate of absorption is described in terms of an *enhancement factor*  $\phi$  which is used as a multiplier:

$$\phi = k_L / k_L^\circ,$$

where  $k_L^\circ$  is the physical mass transfer coefficient.

The enhancement factor can be evaluated from equations originally developed by Van Krevelen and Hoftijzer (1948). A convenient chart based on the equations is shown in Fig. 6. The parameter for the curves is  $\phi_x - 1$ , where  $\phi_x$  is the enhancement factor as  $Ha$  approaches



**FIGURE 6** Effect of chemical reaction on liquid-phase mass transfer coefficient (assumes bimolecular irreversible reaction). [Data based on Van Krevelen, D. W., and Hoftijzer, P. J. (1948). *Rec. Trav. Chim.* **67**, 563.]

infinity, i.e., when all the solute reacts in the film. Values of  $\phi_x$  were originally based on two-film theory, but a more recent refinement described in Perry's Handbook (Fair, 1997) enables one to make the evaluation in terms of penetration theory, as follows:

$$\phi_\infty = \sqrt{\frac{D_A}{D_B}} + \sqrt{\frac{D_B}{D_A} \left( \frac{C_{B0}}{bC_{Ai}} \right)} \quad (14a)$$

The upper curve of Fig. 6 represents a pseudo-first-order reaction, at which the concentration of B is the same in the film as in the bulk of the liquid. For values of  $Ha$  greater than 3,  $k_L$  for pseudo-first-order reactions is given by

$$k_L = \sqrt{k_2 C_{B0} D_A} \quad (14b)$$

This discussion applies to an irreversible second-order reaction. For reversible reactions the relationships are more complex and are discussed in the texts by Sherwood *et al.* (1975) and by Danckwerts (1970).

### III. MODELS FOR ABSORPTION EQUIPMENT

The principles discussed in Section II describe the equilibrium and mass transfer behavior at a given point. In actual plant equipment, because of the transfer of solute from the gas to the solvent, concentrations change from point to point as the gas and liquid travel through the equipment. These changes cause variations in equilibrium concentrations, driving forces, and mass transfer resistances. The point relationships can be translated into equipment mass

transfer behavior with the aid of material and heat balances. In order to apply these balances, the equipment must be described in terms of a mathematical model.

In this section, the equations are presented for the common types of contactors: differential contactors and stage-wise contactors. The equations are developed for the case of steady-state, countercurrent contacting of liquid and gas with negligible heat effects, with a single-component absorption. Some discussion of extensions to other situations follows.

#### A. Differential Contactors

##### 1. Material Balances

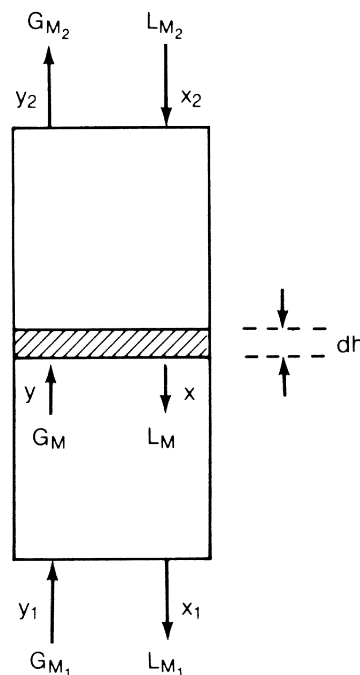
Differential contactors include packed towers, spray towers, and falling-film absorbers, and are often called *counterflow contactors*. In such devices gas and liquid flow more or less continuously as they move through the equipment.

A material balance over a contactor slice (Fig. 7) gives

$$dG_M = N_A a dh \quad (15a)$$

Similarly, a component balance over the same slice gives

$$d(G_M y) = y dG_M + G_M dy = N_A a dh \quad (15b)$$



**FIGURE 7** Material balance for a differential contactor.

Combining these to eliminate  $dG_M$  and integrating gives

$$h_T = \int_{y_2}^{y_1} \frac{G_M dy}{N_A a(1-y)}, \quad (15c)$$

Substituting Eq. (6a) for  $N_A$  gives

$$h_T = \int_{y_2}^{y_1} \frac{G_M y_{BM} dy}{k'_G a(1-y)(y-y_i)} \quad (15d)$$

The group  $G_M/k'_G a$  is independent of concentration and can be taken out of the integral, giving

$$h_T = \left( \frac{G_M}{k'_G a} \right) \int_{y_2}^{y_1} \frac{y_{BM} dy}{(1-y)(y-y_i)} = H_G N_G \quad (16a)$$

Here  $N_G$  is dimensionless and is referred to as the number of gas-phase transfer units;  $H_G$  has the dimension of length or height and is referred to as the height of a gas-phase transfer unit.

Here  $N_G$  is dimensionless and is called the *number of gas-phase transfer units*;  $H_G$  has the dimension of length or height and is referred to as the *height of a gas-phase transfer unit*. As shown in Eq. (16a), the required height of the packed bed  $h_T$  is the product of  $H_G$  and  $N_G$ .

A similar derivation can be carried out in terms of liquid concentrations and flows, giving

$$h_T = H_L N_L = \frac{L_M}{k'_L a} \int_{x_2}^{x_1} \frac{x_{BM} dx}{(1-x)(x_i-x)} \quad (16b)$$

A derivation similar to the preceding one but in terms of the overall mass transfer coefficient  $K'_{OG}$  [Eq. (6)] gives

$$h_T = \frac{G_M}{K'_{OG} a} \int_{y_2}^{y_1} \frac{y_{BM}^* dy}{(1-y)(y-y^*)} = H_{OG} N_{OG} \quad (16c)$$

where

$$H_{OG} = G_M / K'_{OG} a \quad (16d)$$

and

$$N_{OG} = \int_{y_2}^{y_1} \frac{y_{BM}^* dy}{(1-y)(y-y^*)} \quad (16e)$$

Equation (16c) is of great practical interest. It is the basis for computing the required packed height for a given separation, and takes into account mass transfer resistances on both sides of the interface. Also, it avoids the need to calculate the interfacial concentrations required for Eqs. (16a) and (16b).

The  $N_{OG}$  in Eq. (16e) is termed the *overall number of transfer units*. It is dimensionless and is the ratio of the change of bulk-phase concentration to the average concentration driving force. It is essentially a measure of the ease of separation. The  $H_{OG}$  in Eq. (16d) is termed the *overall height of a transfer unit*. It has the dimension of length and defines the vertical height of contactor required to provide a change of concentration equivalent to one transfer unit.

It is therefore a measure of the efficiency of contacting provided by the particular device used in the tower.

Mass transfer data are often expressed in terms of  $H_G$  and  $H_L$ , and these are used to obtain the value of  $H_{OG}$ . The relationship between  $H_{OG}$ ,  $H_G$ , and  $H_L$  is obtained by substituting the expressions for  $H_G$ ,  $H_L$ , and  $H_{OG}$  in Eqs. (16a)–(16c), together with Eqs. (7a)–(7c), in Eq. (5) to give

$$H_{OG} = \frac{y_{BM}}{y_{BM}^*} H_G + \frac{m G_M}{L_M} \frac{x_{BM}}{y_{BM}^*} H_L \quad (17)$$

## 2. Dilute Systems

For dilute systems, the  $x_{BM}$ ,  $y_{BM}$ , and  $1-y$  terms approach unity, and Eqs. (16e) and (17) can be rewritten

$$N_{OG} = \int_{y_2}^{y_1} \frac{dy}{y-y^*} \quad (18a)$$

$$H_{OG} = H_G + m \frac{G_M}{L_M} H_L \quad (18b)$$

When Henry's law is valid [Eq. (1c)], Eq. (18a) can be analytically integrated; alternatively, the graphical form shown in Fig. 8 can be used for evaluating  $N_{OG}$ . Expressions for cases in which the equilibrium curve cannot be linearly approximated are available in several texts, such as Hines and Maddox (1985). Figure 8 shows that the number of transfer units increases with the ratio  $m G_M / L_M$ . When this ratio increases above unity, the number of transfer units, and therefore column height, rapidly increase;

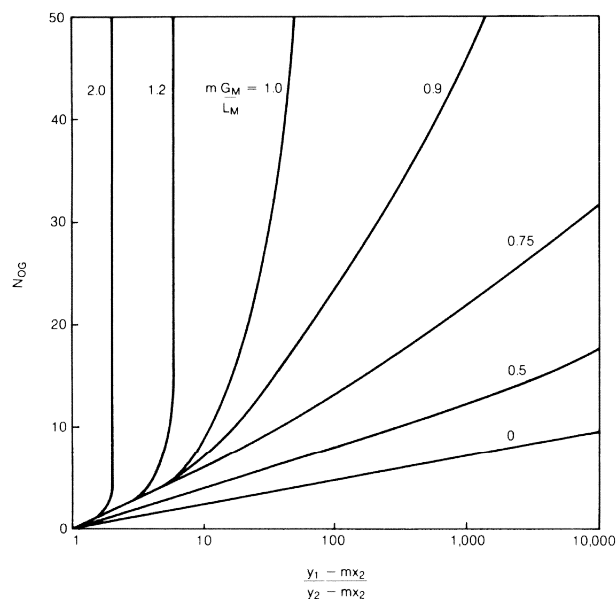


FIGURE 8 Number of overall gas-phase transfer units at constant  $m G_M / L_M$ .



in such case, a large column height is required to achieve a reasonable level of absorption.

### 3. Multicomponent Absorption

The above derivations can be extended to multicomponent absorption, making use of Eqs. (8) as described by Hobler and by Sherwood *et al.* (1975) and giving

$$N_{OG} = \int_{y_2}^{y_1} \left( \frac{y_{fm}}{1 - ty} \right) \left( \frac{dy}{y - y^*} \right) \quad (19a)$$

and

$$H_{OG} = \frac{G_M}{K''_{OG} a y_{fm}^*}, \quad (19b)$$

where  $y_{fm}^*$  and  $t$  are given by Eqs. (8d) and (8e), respectively.

## B. Stagewise Contactors

Tray columns and sometimes also packed and spray columns are described in terms of a stage model. An ideal or theoretical stage is hypothetical device in which the gas and liquid are perfectly mixed, contacted for a sufficiently long period of time so that equilibrium is attained, and then separated. The gas leaving the stage is therefore in equilibrium with the liquid leaving the stage. In practice, complete equilibrium can never be attained, since infinite contact time is required to achieve equilibrium. A factor used to account for this nonideality is stage efficiency.

### 1. Material Balances

An absorber is often modeled as a device that contains a finite number of ideal stages (Fig. 9), with countercurrent flow of vapor and liquid. As the gas rises from stage to stage, it contains less and less of the solute, which is transferred to the solvent.

A material balance can be written for envelope 1 in Fig. 9.

$$L_{M,0}x_0 + G_{M,n}y_n = L_{M,n-1}x_{n-1} + G_{M,1}y_1 \quad (20)$$

The equation can be expressed in terms of the flows entering the absorber, that is, the solute-free solvent entering at the top, and the rich gas, such that

$$y' = yG_M/G'_M \quad (21a)$$

$$x' = xL_M/L'_M \quad (21b)$$

Substituting Eqs. (21a) and (21b) in Eq. (20) gives

$$L'_{M,0}x'_0 + G'_{M,n}y'_n = L'_{M,n-1}x'_{n-1} + G'_{M,1}y'_1 \quad (22)$$

Since the feed flows  $G'_M$  and  $L'_M$  do not change throughout the contactor,

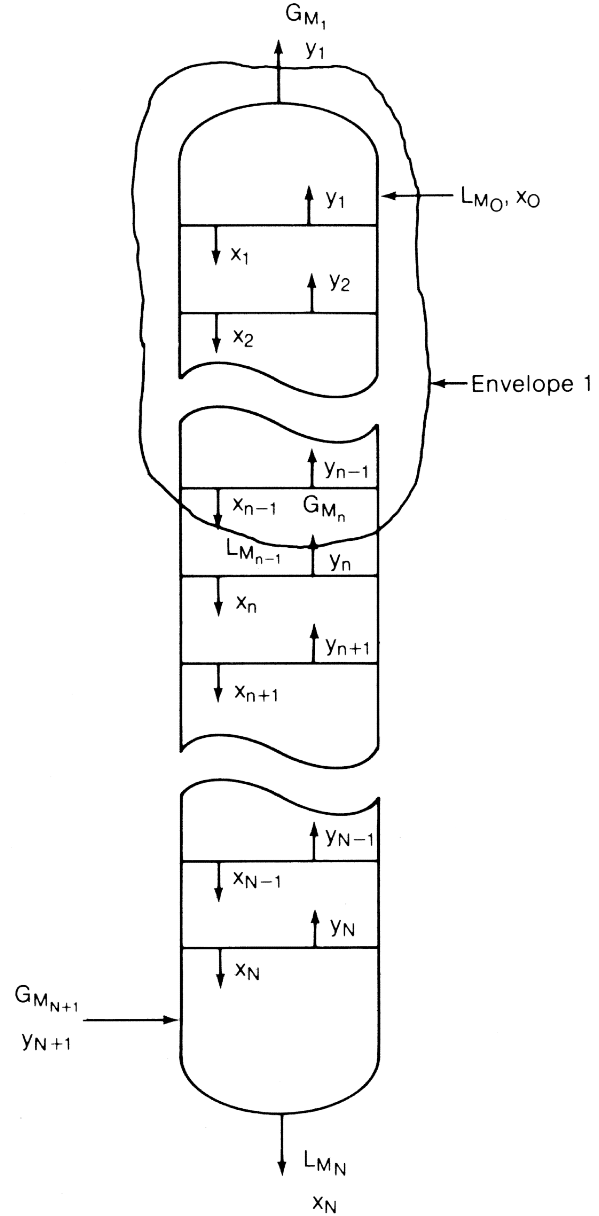


FIGURE 9 Schematic diagram of a stagewise absorber.

$$L'_{M,0} = L'_{M,1} = L'_{M,2} = \dots = L'_{M,n-1} \\ = L'_{M,n} = \dots = L'_{M,N} \quad (23a)$$

$$G'_{M,0} = G'_{M,1} = G'_{M,2} = \dots = G'_{M,n-1} \\ = G'_{M,n} = \dots = G'_{M,N} \quad (23b)$$

and Eq. (22) can be simplified to give

$$y'_n = \frac{L'_M}{G'_M} x'_{n-1} + \left( y'_1 - \frac{L'_M}{G'_M} x'_0 \right) \quad (24)$$

Equation (24) is an equation of a straight line when plotted on  $x - y$  coordinates, with a slope of  $L'_M/G'_M$  and an intercept of  $y'_1 - L'_M x'_0/G'_M$ . This line is often referred to as the *operating line* and is the locus of all the points that obey the stage material balance given by Eq. (20).

For dilute solutions,  $L'_M \approx L_M$ ,  $G'_M \approx G_M$ ,  $x' \approx x$ , and  $y' \approx y$ , Eq. (24) simplifies to

$$y_n = \frac{L_M}{G_M} x_{n-1} + \left( y_1 - \frac{L_M}{G_M} x_0 \right) \quad (25)$$

## 2. Graphical Method

The operating line can be plotted as a straight line on  $y'$  versus  $x'$  coordinates. The equilibrium curve can also be plotted on the same coordinates (Fig. 10). Each point on the operating line obeys the stage material balance given by Eq. (24); the  $(x'_n, y'_{n+1})$  values of a point on this line give the compositions of the liquid leaving and vapor entering stage  $n$ . Each point on the equilibrium curve, given by  $y'_n = f(x'_n)$ , obeys the equilibrium relationship at stage  $n$ ;  $y'_n$  is in equilibrium with  $x'_n$ .

To obtain the number of ideal stages in the contactor, one starts by plotting the point  $y'_{N+1}$  (which is the feed composition of the gas) on the operating line; this defines  $x'_N$ . This corresponds to solving the material balance given by Eq. (24) to determine  $x'_N$ . Next, one draws a vertical line from the point  $(x'_N, y'_{N+1})$  to the equilibrium curve; this defines  $y'_N$ . This corresponds to solving the equilibrium relationship to determine  $y'_N$ . From the point

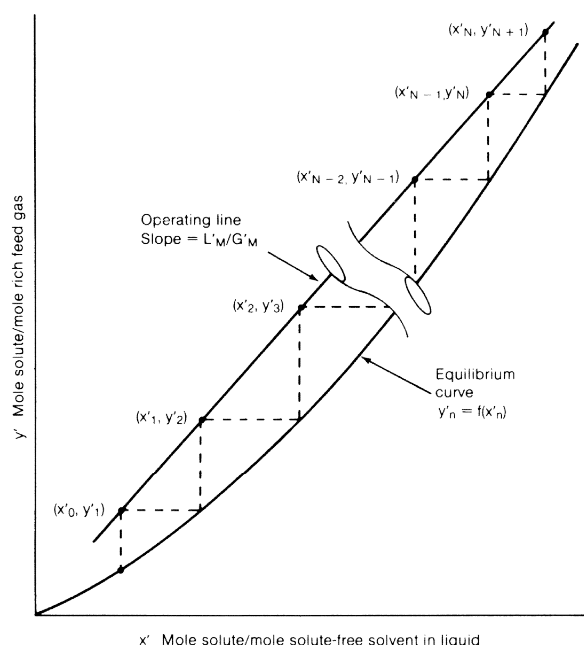


FIGURE 10 Graphic method for stagewise contactors.

$(x'_N, y'_N)$ , one draws a horizontal line back to the operating line, thus solving the material balance to obtain  $x'_{N-1}$ . The process is then continued until the top liquid composition  $x'_0$  is reached (Fig. 10). Each step shown on the diagram represents one ideal stage; the number of ideal stages is counted from the diagram.

Often, slightly different coordinates are used for  $y-x$  plotting. Instead of plotting  $y'$  against  $x'$ , one can plot  $y$  against  $x$ ; in this case, the operating line will be curved. At other times,  $y'$  can be expressed in terms of moles of solute per mole of gas leaving the absorber. The construction described above is similar in all these cases.

Rousseau and Staton (1988) extended the  $y-x$  diagram plot to situations where a single component  $A$  is absorbed into the liquid and instantaneously reacts with a reactive species in the liquid. In this plot, the  $x$  axis is the fraction of reactive species in the liquid that has reacted with solute  $A$ , while the  $y$  axis is the ratio of moles of solute  $A$  in the gas to moles of solute-free gas. The equilibrium curve is derived using both Henry's Law constant and the reaction equilibrium constant.

## 3. Minimum Solvent Rate

When the operating line and the equilibrium curve intersect, an infinite number of stages is required to achieve the separation (Fig. 11). The intersection point is called the *pinch point* and may occur at the bottom (Fig. 11a), at the top (Fig. 11b), or at a tangent point (Fig. 11c). The solvent rate leading to this intersection is the minimum solvent flow required to absorb the specified amount of solute.

Since the top and bottom pinch points shown in Fig. 11 represent intersections of operating and equilibrium lines, they may be predicted analytically. At the top, the lean gas and the lean solvent are in equilibrium, i.e.,  $y_1 = Kx_0$  (Fig. 9). Similarly, at the bottom the rich gas and the rich solvent are in equilibrium, i.e.,  $y_{N+1} = mKx_N$ . By material balance, the minimum solvent rate can be calculated. Frequently, the pinch occurs at the bottom.

The actual solvent rate specified for the separation must exceed the minimum solvent rate, or an infinite number of stages will be required. For a contactor with a finite number of stages, this means that the separation will not be achieved unless actual solvent rate exceeds the minimum. The higher the solvent rate specified, the greater is the distance between the operating line and the equilibrium curve, and the smaller is the number of stages required.

## 4. Absorption Factors

For each stage, an absorption factor can be defined by

$$A_n = (L_M/mG_M)_n \quad (26)$$



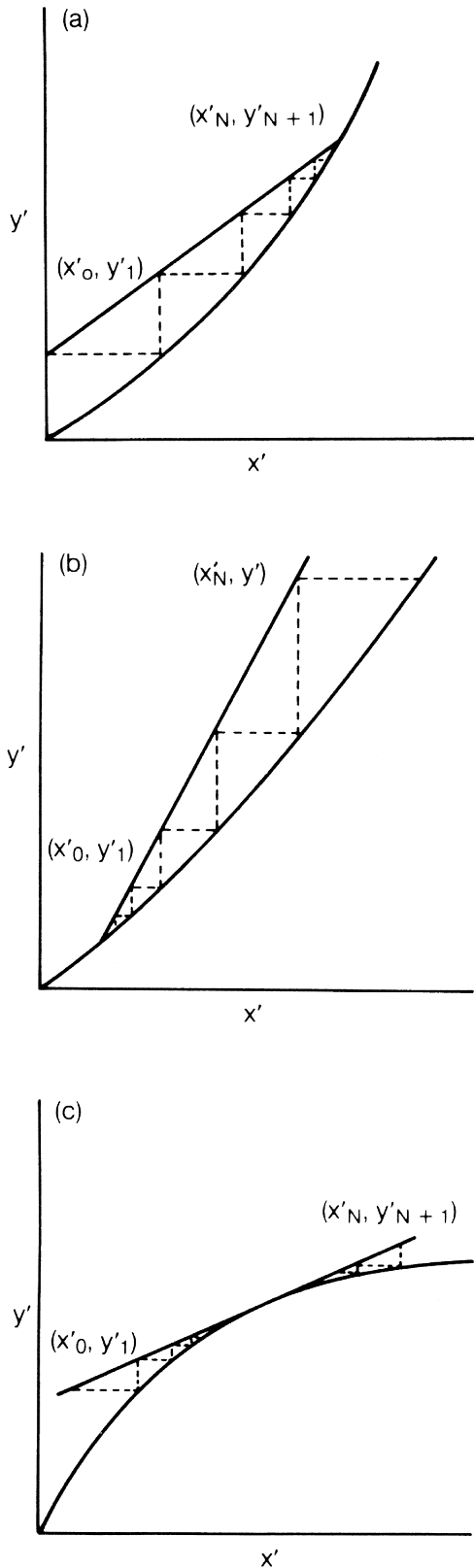


FIGURE 11 Graphic illustrations of minimum solvent rate. (a) Pinch at the bottom, (b) pinch at the top, (c) tangent pinch.

This absorption factor is the ratio of slope of the operating line to that of the equilibrium curve. When the absorption factor is lower than unity, the pinch is located near the bottom of the column (Fig. 11a); when it is higher than unity, the pinch is located near the top of the column (Fig. 11b).

For a dilute gas, and when the equilibrium curve can be approximated by a linear relationship passing through the origin, Eq. (25) is applicable, and an average absorption factor  $A$  can be applied to describe the contactor. Under these conditions, an analytical solution of the material balance equation and the equilibrium relationship is possible, giving the Kremser equation:

$$\frac{y_{N+1} - y_1}{y_{N+1} - mx_0} = \frac{A^{N+1} - A}{A^{N+1} - 1} \quad (A \neq 1)$$

$$= N/(N + 1) \quad (A = 1) \quad (27)$$

The left-hand side of Eq. (27) is in principle the ratio of the change of composition of the gas through the contactor to the change that would have occurred had the gas come to equilibrium with the liquid entering the column.

For concentrated gases, the absorption factor varies from stage to stage. In many cases Eq. (27) can be used with an effective average absorption factor and the mole ratio concentration  $y'$ :

$$\frac{y'_{N+1} - y'_1}{y'_{N+1} - y'_0} = \frac{A_{\text{ave}}^{N+1} - A_{\text{ave}}}{A_{\text{ave}}^{N+1} - 1} \quad (A_{\text{ave}} \neq 1)$$

$$= \frac{N}{N + 1} \quad (A_{\text{ave}} = 1) \quad (28)$$

The value for  $A_{\text{ave}}$  is often defined using Eq. (29), with  $m_{\text{ave}}$  evaluated at the average column temperature:

$$A_{\text{ave}} = L'_M / G'_M m_{\text{ave}} \quad (29)$$

If the absorption is multicomponent, the average equilibrium constant  $m_{\text{ave}}$  is determined for each of the solute components at the average temperature and pressure of the absorber, and a separate absorption factor  $A_{\text{ave}}$  is defined for each component. These absorption factors can be used in Eq. (28) to define the absorbed fraction of the component.

Horton and Franklin (1940) used the average absorption factor approach in analyzing a number of absorbers in the petroleum industry. Edmister (1943) extended the Horton and Franklin concept, retaining the Kremser equation form and making use of several empirical factors. He used an effective absorption factor  $A_e$  and a modified absorption factor  $A'$ , given by

$$A_e = \sqrt{A_N(A_1 + 1) + 0.25} - 0.5 \quad (30a)$$

$$A' = \frac{A_N(A_1 + 1)}{A_N + 1} \quad (30b)$$

Using these definitions, Eq. (28) becomes

$$\frac{y'_{N+1} - y'_1}{y'_{N+1}} = \frac{A_e^{N+1} - A_e}{A_e^{N+1} - 1} \left( 1 - \frac{L'_M x'_0}{A' G'_M y'_{N+1}} \right) \quad (30c)$$

Hines and Maddox (1985) found that the Edmister method gives a close approximation to observed or rigorously computed concentration gradients in many multicomponent absorbers.

## 5. Other Procedures

Graphical procedures such as those described above can also be extended to multicomponent absorption. This subject is discussed in detail by Sherwood (1975).

A method suitable for computer calculations, which carries out tray-by-tray mass, component, and heat balances was first developed by Sujata (1961). In this method, the liquid and vapor flow rates and the temperature profile are assumed and used to calculate an absorption factor for each stage [Eq. (26)]. A component balance is written for each stage in terms of the component flows and absorption factors. The component balances are solved by matrix techniques to give component flows for each stage. Energy balances are then solved to obtain a new temperature profile. The total vapor and liquid flow profiles are found by summing the individual component flows. The calculation is then repeated with the updated temperatures and flows in a trial-and-error manner, until convergence is reached. There are several variations of the above procedure. Some of the popular ones are discussed in Wankat's text. Some rigorous distillation methods have also been extended to absorption.

## C. Rate Models

Traditionally, absorbers and strippers were described as stagewise contactors. Krishnamurthy and Taylor developed a new rate (nonequilibrium stage) approach for modeling absorbers and strippers. This approach describes an absorber as a sequence of nonequilibrium stages. Each stage represents a real tray in the case of a tray tower or a section of a continuous contacting device such as a packed column. For each nonequilibrium stage, the mass, component, and energy balance equations for each phase are solved simultaneously, together with the mass and energy transfer rate equations, reaction rate equations, and the interface equilibrium equations. Computation of stage efficiencies is thus avoided altogether and is, in effect, substituted by the rate equations.

Although the rate model can be applied to any separation, it has become most popular in absorption and stripping. Reported case studies demonstrated that, in at least some situations, a rate model can more closely approxi-

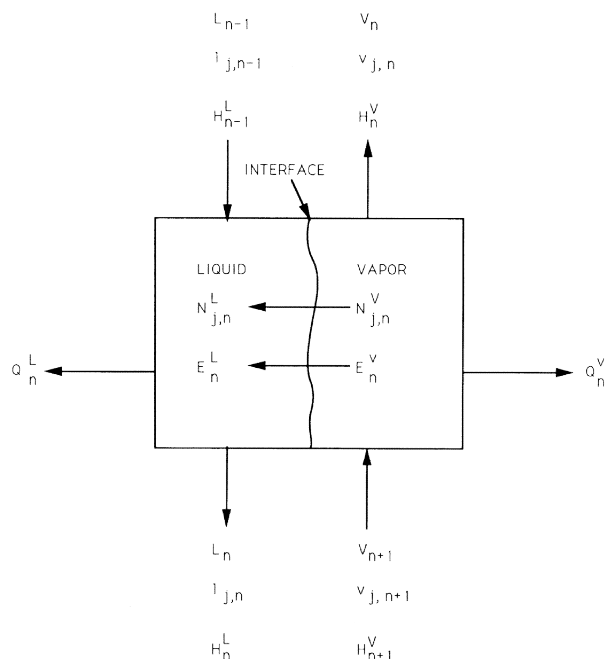


FIGURE 12 Schematic diagram of a nonequilibrium stage  $n$ .

mate absorber performance than can an equilibrium stage model. The success of rate models in absorption is largely a result of the difficulty of reliably predicting stage efficiencies in absorbers. The presence of many components, low stage efficiencies, significant heat effects, and chemical reactions are commonly encountered in absorbers and difficult to accommodate in stage efficiency prediction.

Figure 12 is a schematic diagram of a nonequilibrium stage  $n$  in an absorber. The equations applying to this stage are described below. A more detailed description is given by Krishnamurthy and Taylor.

Component balances for component  $j$  on stage  $n$  are given in Eqs. (31a–c) for the vapor phase, the liquid phase, and the interface, respectively:

$$v_{j,n} - v_{j,n+1} + N_{j,n}^V = 0, \quad (31a)$$

$$l_{j,n} - l_{j,n-1} - N_{j,n}^L = 0, \quad (31b)$$

$$N_{j,n}^V = N_{j,n}^L \quad (31c)$$

Energy balances on stage  $n$  are given in Eqs. (32a–c) for the vapor phase, the liquid phase, and the interface, respectively:

$$V_n H_n^V - V_{n+1} H_{n+1}^V + Q_n^V + E_n^V = 0 \quad (32a)$$

$$L_n H_n^L - L_{n-1} H_{n-1}^L + Q_n^L - E_n^L = 0 \quad (32b)$$

$$E_n^V = E_n^L \quad (32c)$$

The interface equilibrium is written at the interface.

$$y_{j,n}^I = m_{j,n}^I x_{j,n}^I \quad (33)$$

In Eq. (31c),  $N_{j,n}^V$  and  $N_{j,n}^L$  are the mass transfer rates. These are calculated from multicomponent mass transfer equations. The equations used take into account the mass transfer coefficients and interfacial areas generated in the specific contactor, reaction rates, heat effects, and any interactions among the above processes.

The above equations, including those describing the mass transfer rates on each stage, are solved simultaneously for all stages. Solution of these nonlinear equations is complex and usually requires a computer. Newton's numerical convergence technique, or a variant of it, is considered to be most effective in solving these equations.

#### D. Heat Effects in Absorption

When absorption liberates a considerable quantity of heat, and if a large quantity of solute is absorbed, the solution temperature rises. This reduces the solubility of the solute in the liquid, thus counteracting absorption.

The temperature rise can be evaluated from the quantity of heat liberated, which in turn is a function of the change in liquid composition. An equilibrium curve that takes into account the temperature variations through the absorber is shown in Fig. 13, corresponding to a bulk temperature rise from  $T_2$  to  $T_1$  as the bulk liquid composition changes from  $x_2$  to  $x_1$ . The location of the curves depends on which resistance controls, because the equilibrium relationship is obeyed at the interface and not at the bulk.

Work on absorption with large heat effects indicates that the temperature inside an absorber often goes through a

maximum when the solvent is volatile (e.g., ammonia-water absorption).

When solute is absorbed rapidly, the rate of heat liberation is largest near the bottom of the absorber, causing the equilibrium curve to bend upward at the solute-rich end, while remaining relatively unaffected by the heat of solution at the lean end of the absorber. This may sometimes lead to a pinched condition at the rich end of the absorber. When this type of pinching is a concern, it is customary either to provide cooling coils inside the absorber or to divert a liquid stream through an external cooler and then return it to the next lower tray in the column.

Other than the heat of solution, heat effects that may influence absorber performance are solvent vaporization, sensible heat exchange between the gas and the liquid, and loss of sensible heat due to cooling coils or atmospheric cooling. Detailed discussion on heat effects in absorption is presented in the text by [Sherwood et al. \(1975\)](#).

### IV. ABSORBER DESIGN

Absorber design is normally carried out in three phases: process design, column sizing, and hydraulic design. In the process design phase, the main system parameters (e.g., solvent selection, operating pressure and temperature, solvent rate, theoretical number of stages, type of contactor) are determined. In the column-sizing phase, the height, diameter, and sizes of the main internals such as downcomers, packings, and tray spacing are determined. Finally, the hydraulic design phase defines all the sizes, dimensions, and layouts of column inlets, outlets, and the multitude of internals used in the column.

#### A. Process Design

The following steps are followed in column process design:

1. Specification of the separation. A separation is specified by defining column feed flow rate and composition, overhead solute concentration (alternatively, solute recovery), and the concentration of solute (if any) in the lean solvent. If the purpose of absorption is to generate a specific solution, as in acid manufacture, the solution concentration completes the separation specification. For all other purposes, one specifying variable (e.g., rich solvent concentration or solvent flow rate) remains to be specified and is usually set by optimization as outlined below.
2. Selection of solvent and solute recovery process. This was discussed in Section I.
3. Setting the operating pressure. A higher pressure favors the gas solubility and decreases the diameter of

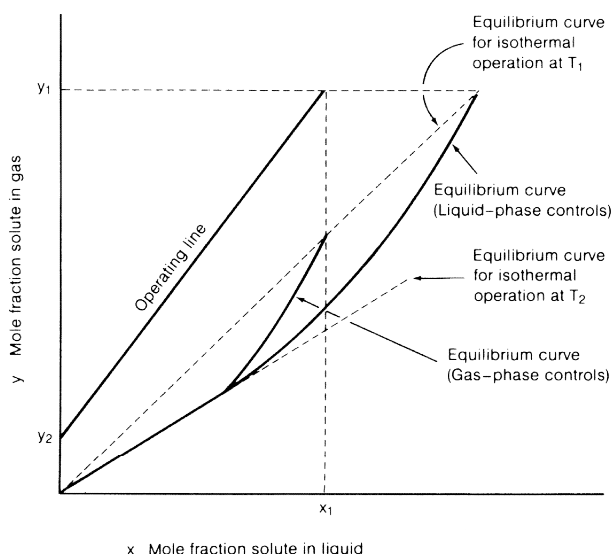


FIGURE 13 Effect of heat liberation on the equilibrium curve.

the contactor. However, the cost of attaining the pressure must be considered. Off-gas scrubbers, for example, process large quantities of gas which is then discharged to the atmosphere; in such a case, the absorber pressure is set at near atmospheric, and the cost of moving the gas through the contactor (due to pressure drop) may govern the decision on operating pressure.

4. Determining solvent circulation rate. If the purpose of absorption is to generate a solution of a specific concentration, the circulation rate is a fixed function of this concentration. For all other purposes, this circulation rate is determined by optimization. As circulation rate is increased, the absorption factor  $L_M/(mG_M)$  increases, as does the distance from the operating line to the equilibrium curve. This leads to a shorter and therefore cheaper column. On the other hand, the higher the circulation rate, the greater is the cost of separating the solute from the solvent and the larger is the diameter of the absorber. Many studies have shown that the optimum circulation rate is about 40% greater than the minimum solvent rate.

5. Selection of contactor type. Tray and packed columns are most common; other types are generally used only for special services.

The main factors favoring packed columns are (1) very corrosive applications, where plastic or ceramic packings are favored over trays, which are almost always constructed of metal; (2) low pressure drop requirement, which is easier to achieve with packings than with trays; (3) small-diameter columns, because trays require access for inspection and maintenance; and (4) foaming systems, which are easier to handle in packed towers.

The main factors favoring tray columns are (1) presence of solids (packings have a greater tendency to trap solids and to suffer from the resulting blockage and channeling), (2) very high or very low liquid rates (trays are more suitable to handle these than packings, except for structured packings, which are also capable of handling very low liquid rates), (3) slow reaction rate processes (trays can provide a greater liquid holdup and therefore more residence time), (4) complexities such as cooling coils or intercoolers, which are easier to incorporate into tray columns, and (5) column weight (tray columns are generally lighter and easier to support).

6. The number of theoretical stages, or transfer units, is calculated using a mathematical model of the type described in Section III. At this stage, it is necessary to allow for any heat effects; if these are significant, coiling coils or intercoolers may be required.

## B. Column Sizing

In this section, the main types of absorption equipment (packed columns and tray columns) are described, and

the main considerations in their design and sizing are discussed.

### 1. Packed Columns

A typical arrangement (Fig. 14) consists of a vertical tower containing one or more beds of packings. The descending liquid is distributed over the packing layers, forming liquid films that flow along the surfaces of the particles, thus exposing a large surface area for gas-liquid contact. The solute is transferred from the gas to the liquid across this surface. The type and size of packings may be the same throughout the column or may differ from bed to bed.

The characteristics considered most desirable for good packing performance are a high surface area, a uniform distribution of liquid, and a low resistance to gas flow.

Two types of packings are common: random packings, which are discrete pieces of packings randomly dumped into the column, and structured packings, which are layers of systematically arranged packings, mostly corrugated sheets or wire mesh. Structured packings provide uniform channels for gas and liquid flow, a more even distribution, and greater surface area for the same resistance to gas flow. In general, they tend to lead a more efficient operation but are also more expensive. For absorbers, random packings are more popular, with structured packings being justified only when pressure drop and efficiency demands are unusually high.

Common types of random packings are shown in Fig. 15. The packings shown in Fig. 15a-c have been largely superseded by the packings shown in Fig. 15d-h.

Table II shows several common random packings and compares them on two bases: (1) surface area per unit volume, the larger area providing more opportunity for mass transfer, and (2) packing factor, a measure of throughput capacity and pressure drop, the lower is value, the higher the capacity and the lower the pressure drop. The table shows that as packing size increases, capacity rises while efficiency decreases. The table includes two packings fabricated from plastic (usually polypropylene); this material of construction is resistant to corrosion and is light weight. Plastic packing applications range from sulfuric acid absorbers to off-gas scrubbers and stripping columns.

Table II shows that, as packing size increases, packing capacity rises while packing efficiency decreases. It also shows that both capacity and efficiency are greater for Pall rings and Intalox® saddles than for Raschig rings and Berl saddles.

The data in Table II are approximate, because the geometry of each packing varies slightly from one manufacturer to another. Usually, the type of data shown in this table is provided by each manufacturing company for

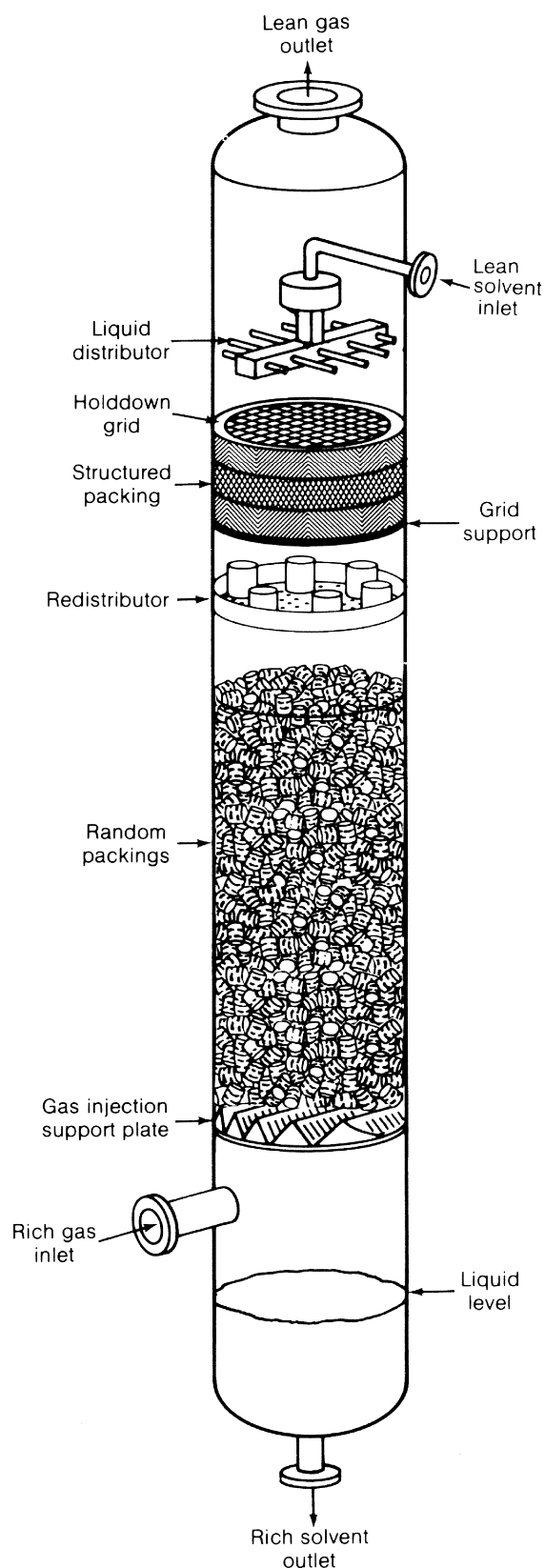


FIGURE 14 Packed column.

its own packings. Perry's text contains a more extensive tabulation of packing factors.

Maximum capacity of a packed bed is usually limited by the onset of flooding. During normal operation, gas flows up while liquid drains freely along the packing surfaces. As gas rate is increased, it begins to interfere with free draining, causing some liquid accumulation in the bed. When this interference is so high that liquid fills the tower, the column is said to be flooded.

The condition of flooding is predicted from generalized charts such as that in Fig. 16. The abscissa shows a scale of a dimensionless term called the flow parameter. This parameter represents a ratio of the kinetic energy of the liquid to the kinetic energy of the gas; thus very low values of the parameter are associated with low-pressure absorbers where the volumetric ratio of gas to liquid may be very high. The ordinate scale shows values of a capacity parameter, generalized through the packing factor (Table II)

Each curve in Fig. 16 represents a constant pressure drop value. Packed absorbers are usually sized to give a pressure drop of 0.25 to 0.50 in. H<sub>2</sub>O per foot (200–400 N/m<sup>2</sup> per meter) of packed depth. Figure 16 is used to determine the column cross sectional area to achieve

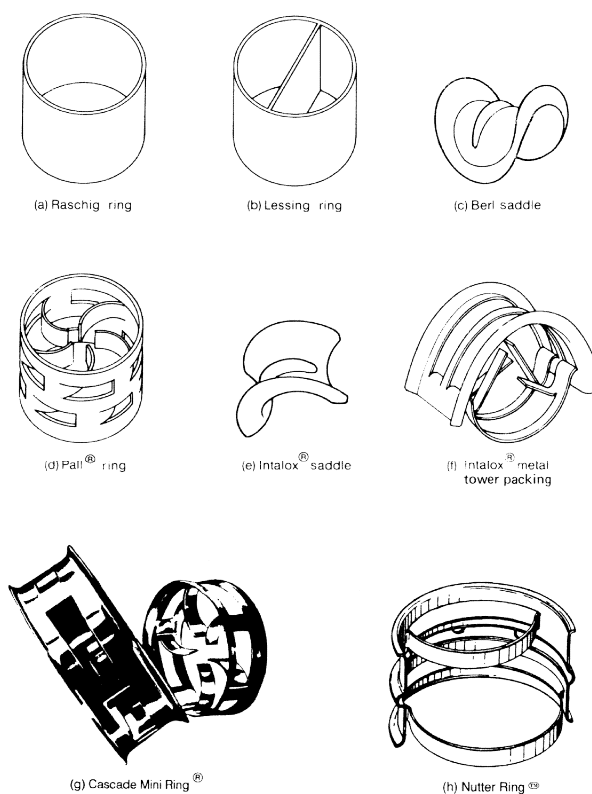


FIGURE 15 Common types of random packings (Parts e and f, courtesy of Norton Co.; part g, courtesy of Glitsch, Inc.; part h, courtesy of Nutter Engineering Corp.)



**TABLE II Characteristics of Random Packings<sup>a</sup>**

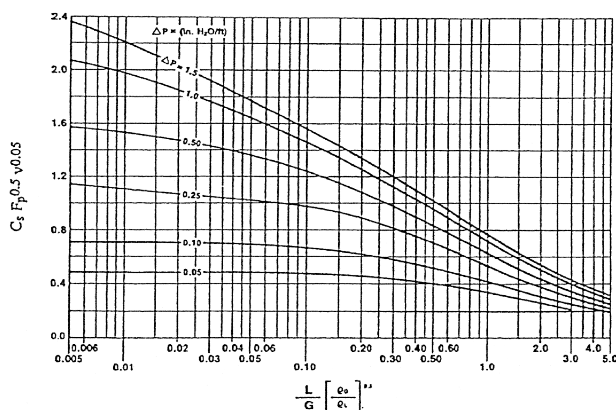
Nominal size (mm)	Surface area (m <sup>2</sup> /m <sup>3</sup> )					Packing factor (m <sup>-1</sup> )				
	25	38	50	75	90	25	38	50	75	90
Type										
Raschig ring (metal)	185	130	95	66	—	450	270	187	105	—
Pall ring (metal)	205	130	115	—	92	157	92	66	—	53
Intalox <sup>®</sup> Metal Tower Packing	—	—	—	—	—	135	82	52	43	—
Raschig ring (ceramic)	190	120	92	62	—	510	310	215	120	—
Berl saddle (ceramic)	250	150	105	—	—	360	215	150	—	—
Intalox <sup>®</sup> saddle (ceramic)	255	195	118	92	—	320	170	130	70	—
Intalox <sup>®</sup> saddle (plastic)	206	—	108	88	—	105	—	69	50	—
Pall ring (plastic)	205	130	100	—	85	170	105	82	—	52

<sup>a</sup> (From Perry, R. H. (ed.) (1985). "Chemical Engineer's Handbook," 6th ed., McGraw-Hill, New York.)

this pressure drop at the design and liquid loads. Pressure drops of 1.5–1.7 in. H<sub>2</sub>O per foot are representative of incipient flooding and values this high are to be avoided.

Packed height is determined from the relationships in Section III. Application of these relationships requires knowledge of the liquid and gas mass transfer coefficients. It is best to obtain these from experimental data on the system if available, but caution is required when extending such data to column design, because mass transfer coefficients depend on packing geometry, liquid and gas distribution, physical properties, and gas and liquid loads, and these may vary from one contactor to another.

In the absence of experimental data, mass transfer coefficients (and hence heights of transfer units) can be estimated by generalized models. A popular and easy to use correlation for random packings is that of Bolles and Fair (1982). The earlier correlations of Onda *et al.* (1968) and Bolles and Fair are also useful for random packings.



**FIGURE 16** Generalized pressure drop correlation of Strigle<sup>21</sup>.  $C_s$  = flow parameter =  $U_s[\rho_g/(\rho_L - \rho_g)]^{0.5}$ , ft/s.  $F_p$  = packing factor, ft<sup>-1</sup>,  $\nu$  = kinematic viscosity of liquid, centipoises/specific gravity.

For structured packings the correlation of Rocha *et al.* (1996) has been well validated for a number of packings tested in larger equipment. Even if experimental data are available, one must be cautious in applying data taken in small laboratory columns to designs of large commercial contactors.

## 2. Tray Columns

A typical arrangement (Fig. 17) consists of a vertical tower fitted with horizontal plates or trays, on which liquid and gas are contacted. Each tray is equipped with gas passages, which may be perforations in the tray floor or other devices such as valves or bubble caps that disperse the rising gas into the liquid layer. The liquid layer on the tray is maintained by the outlet weir. Liquid descends from each tray to the tray below via a downcomer.

Liquid enters the column and flows across the top tray, where it contacts the rising gas to form a froth, emulsion, or spray-type dispersion (Fig. 18). It then overflows the weir into the downcomer, which separates gas from the liquid, and carries liquid by gravity to the tray below. The liquid then flows across the next tray, and the process is repeated. Liquid is thus contacted with gas in a stagewise manner.

Two types of trays are most common: sieve trays and valve trays. A sieve tray is a simple perforated plate. Gas issues from the perforations to give a multiorifice effect; liquid is prevented from descending the perforations or "weeping" by the upward motion of the gas. At low gas flow rates, the upward gas motion may be insufficient to prevent weeping.

In valve trays, the perforations are equipped with valve units (Fig. 19). At high gas rates, the gas force opens the valves, thus providing area for gas flow. At low gas rates, there is insufficient force to keep many of the valves open, and these close, preventing the liquid from weeping. Sieve and valve trays show comparable capacity, efficiency, and

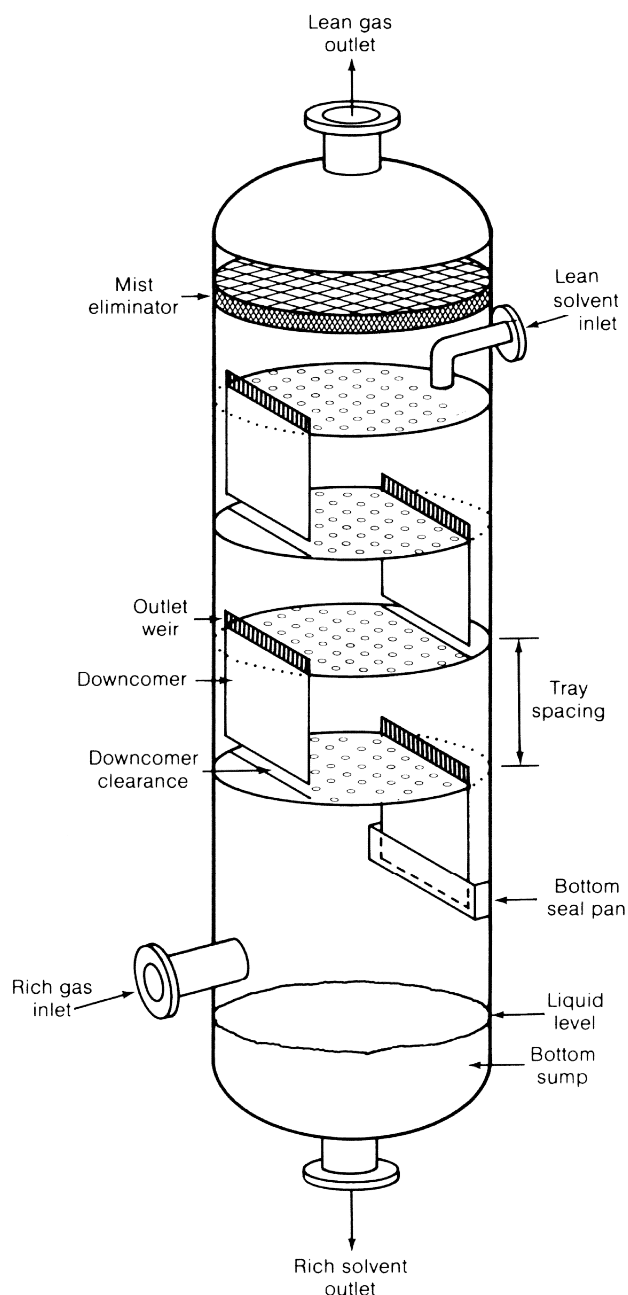


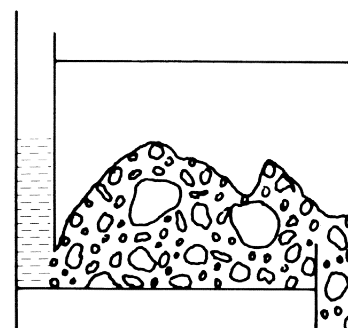
FIGURE 17 Tray column.

other performance characteristics at high gas rates; but valve trays weep less and therefore perform better at low gas rates.

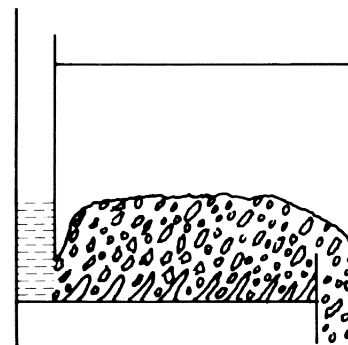
A third type of tray, once commonly employed but currently used only for special applications, is the bubble-cap tray. Its design and operation are discussed by [Bolles \(1963\)](#).

The maximum capacity of a tray column is usually limited by the onset of flooding, which occurs when liquid excessively accumulates inside the column. Flooding is

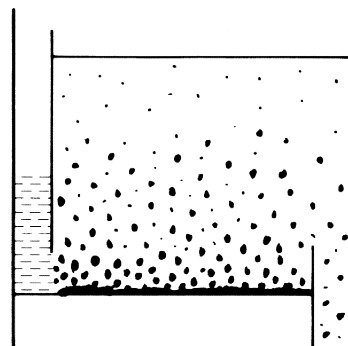
caused by massive liquid carryover from tray to tray (entrainment flood) or when liquid backup in the downcomer reaches the tray above (downcomer backup flood) or when the downcomer is unable to handle the total quantity of descending liquid (downcomer choke flood). At low liquid rates and high gas velocities, entrainment flooding is the most common limit. At high liquid flow rates and low gas velocities (e.g., high pressure operation), downcomer backup and downcomer choke flood are the most common limits.



(a) Froth



(b) Emulsion



(c) Spray

FIGURE 18 Types of dispersion on an absorption tray.





residence times and because of extensive gas backmixing. Their application is limited to easy absorption duties (one or two theoretical stages), usually in systems where the controlling resistance to mass transfer is in the gas phase. Column capacity is usually limited by liquid droplet entrainment from the top.

Spray absorbers are advantageous where low pressure drops are critical and where the gas may contain some solids, such as in the absorption of  $\text{SO}_2$  from coal-fired boiler exhaust gases.

**Falling-film absorbers.** These are usually vertical heat exchangers with the cooling medium in the shell and the absorption taking place in the tubes. The solvent flows downward, while the gas may enter either at the bottom (countercurrent flow) or at the top (cocurrent flow).

Mass transfer in falling-film absorbers is strongly dependent on the gas velocity in the tubes, the liquid and gas distribution, and the tube surface conditions. The maximum capacity of falling-film absorbers is normally restricted either by flooding or by pressure drop. Another important limit in these absorbers is film breakup. If heat flux is excessive, dry areas may form at the tube wall and reduce mass transfer.

Falling-film absorbers make continuous heat removal possible and are therefore extensively used in applications where the heat released during absorption is high, such as in the absorption of hydrogen chloride to form hydrochloric acid.

**Stirred tanks.** These are mechanically stirred vessels, which are advantageous when absorption is accompanied by a slow liquid-phase chemical reaction. As discussed earlier (Section II), this application is considered a chemical reactor rather than an absorber. Stirred tanks provide high liquid residence times but are limited to low gas flow rates.

**Bubble columns.** These are columns full of liquid into which gas is introduced by a perforated pipe or a sparger. Bubble columns are used for applications similar to stirred tanks, but their contact efficiency is lower.

**Venturi scrubbers.** In a venturi scrubber, a liquid jet issues from a nozzle. The jet induces cocurrent gas flow into the throat of the jet. Mass transfer takes place between the gas and the atomized liquid downstream of the nozzle. Mass transfer is usually poor and depends on the throat velocity or pressure drop, the liquid/gas ratio, and the liquid atomization pattern. Because of the cocurrent nature of contacting, the maximum solute removal does not exceed a single theoretical stage. Venturi scrubbers are used primarily for separation of fine particulate matter or

fine liquid mist from a gas stream. They are often also used for simultaneously absorbing certain components from the gas stream, but because of their poor mass transfer are effective only when these components are highly soluble in the liquid. Common applications are scrubbing incinerator fumes and sulfuric and phosphoric acid mists.

**Wet scrubbers.** These are devices in which a liquid spray contacts a gas stream, primarily for the purpose of removing fine solid particles or liquid mists from the gas. In this process, the liquid spray simultaneously absorbs soluble components from the gas. The sprays are generated by a variety of mechanical devices.

## C. Hydraulic Design

This design phase determines the types, dimensions, location, and orientation of the multitude of internals used in absorption columns. It usually leads to refinements to the column design and sizing and, most important, is critical for ensuring trouble-free operation.

### 1. Packed Columns

The most important aspects of packed-column internals and their design are outlined in the following paragraphs.

Packed-tower efficiency and turndown are strongly dependent on the quality of initial *liquid distribution*. Uneven distribution may cause local variations in the liquid/gas ratio, localized pinch conditions, and reduced vapor-liquid contact. Figure 14 shows two common liquid distributor types, the ladder type (shown as the top distributor) and the orifice type (shown as the redistributor). The ladder type is a horizontal header of pipes, which are perforated on the underside. The orifice type is a flat perforated plate equipped with round or rectangular risers for gas passage. Other common types of distributors are a header equipped with spray nozzles (spray distributor) and a header of horizontal channels, with V notches cut in the vertical walls of the channels (notched-trough distributor).

Ladder and spray distributors rely on pressure for their action. They provide a large gas flow area but a somewhat limited liquid flow area; they are light and cheap but are sensitive to corrosion, erosion, and to a certain extent plugging. They are most suitable for high gas/liquid ratio applications.

Orifice and notched-trough distributors rely on gravity for their action. They provide a large liquid flow area; the notched-trough distributor also provides a large gas flow area. They are more robust and expensive than pressure distributors and are sensitive to levelness. The orifice distributor is most sensitive to plugging, while the notched-trough is the least sensitive to plugging, corrosion,

or erosion. The orifice distributor has the potential to generate a distribution pattern superior to most others, but its application is often restricted to clean fluids where the gas/liquid ratio is not high. The notched-trough distributor is often considered the most reliable distributor, although the quality of distribution may be somewhat inferior than that of the orifice or ladder distributors.

*Liquid redistributors* are installed at frequent intervals in a packed column to remix the liquid, thus counteracting the propagation of maldistribution effects and the natural tendency of liquid to migrate toward the wall. A common design practice is to redistribute the liquid every 20 ft (6–7 m).

*Redistributor design* is similar to gravity distributor design. The orifice type is most popular (Fig. 14). A notched-through type requires a liquid collection device above it to feed the liquid onto the distributor. Often, the gas risers are equipped with caps to prevent liquid from dropping through the gas spaces.

*Liquid collectors* are installed when liquid must be collected for redistribution or drawoff (e.g., for external cooling). The common device used is a chimney tray, which is similar to an orifice redistributor, but without perforations. Another common device is the Chevron-type collector, which is a series of Chevron blades, with liquid being collected at the bottom of the blades.

*Packing supports* have to support the packed bed physically, while incorporating a large free area to permit free passage of gas and liquid. Grid supports are common, especially in nonmetallic applications. Gas injection supports (Fig. 14) are usually preferred; these provide separate passages for the gas and liquid and a large free area.

*Holddown plates and bed limiters* are grids or wire screens with openings small enough to prevent migration of packing particles. They prevent bed fluidization, which may cause breakage of ceramic and carbon packings or entrainment of metal or plastic packings with the gas.

## 2. Tray Columns

The most important features of tray column internals and their designs are outlined in the following paragraphs.

*Liquid inlets.* Liquid enters the top tray via a hole in the column shell, often discharging against a vertical baffle or weir, or via a short, down-bending pipe (Fig. 17), or via a distributor. Restriction, excessive liquid velocities, and interference with tray action must be avoided, as these may lead to excessive entrainment, premature flooding, and even structural damage. Disperser units (e.g., perforations, valves) must be absent in the liquid entrance area (Fig. 17) or excessive weeping may result.

*Gas inlets.* Gas must enter above the bottom liquid level or, if bubbled through the liquid, through a well-designed sparger. Commonly, no sparger is used; in such cases, the feed nozzle should be located at least 12 in. (0.3 m) above the liquid level. Impingement on the liquid level, seal pan overflow, and instrument nozzles must be avoided. Failure to follow these guidelines may result in premature flooding, excessive entrainment, and in some cases mechanical damage to the trays.

*Bottom liquid outlets.* Sufficient residence time must be provided in the bottom of the column to separate any entrained gas from the leaving liquid. Gas in the bottom outlet may also result from vortexing or from forthing caused by liquid dropping from the bottom tray (a “waterfall pool” effect). Vortex breakers are commonly used, and liquid-drop height is often restricted. Inadequate gas separation may lead to bottom pump cavitation or vapor choking the outlet line.

*Intermediate liquid outlets.* Liquid may be withdrawn using a chimney tray or from a downcomer. A chimney tray is a flat, unperforated plate with vapor risers. It permits total withdrawal of liquid; a downcomer drawoff permits only partial withdrawal because some weeping occurs through the tray. A downcomer drawoff may contain some entrained gas, which must be separated downstream or allowed for in downstream equipment design.

*Gas outlets.* Sufficient liquid disentrainment from the overhead gas is usually required. This may be achieved by providing sufficient vertical height above the top tray, installation of mist eliminators, or providing external knockout facilities downstream of the column.

*Tray layout.* The preliminary tray and downcomer layout is prepared in the column-sizing phase and refined during the hydraulic design phase. In addition to the parameters previously set, such parameters as hole diameter or the type of valve unit are determined.

Smaller hole diameters usually enhance efficiency and capacity but are also more sensitive to corrosion and plugging. Holes smaller than  $\frac{3}{16}$  in. (5 mm) are uncommon because they require an expensive manufacturing technique. Half-inch (13-mm) holes are common when corrosion or plugging is expected.

The best type of valve unit depends on the corrosive and fouling tendencies of the service, as some valve units tend to pop out of their seats in corrosive services, while others tend to stick to their seats in fouling services.

Other parameters such as level tolerance, tray supports, drainage, weir shape and type are also determined in this phase.

**Downcomers layout.** Usually, segmental downcomers are used, in which the downcomer area extends from the weir to the column wall (Fig. 17), but other designs are not uncommon. The design must consider downcomer hydraulics as well as mechanical and structural factors.

The need for positively sealing the downcomer is determined in this phase. This could be achieved by installing an inlet weir, which is a weir installed at the tray inlet to keep the downcomer outlet immersed in liquid. A similar device, which extends below the tray floor, is a seal pan (Fig. 17). Both devices provide positive assurance against vapor rising up the downcomer, but they may also trap solids and dirt and cause blockage. A seal pan must always be used in the downcomer from the bottom tray; otherwise there is nothing to prevent vapor from rising up the bottom downcomer.

## NOMENCLATURE

A	Component A	$F_p$	Packing factor, $\text{ft}^{-1}$ ( $\text{m}^{-1}$ )
A	Absorption factor, $L_M/(mG_M)$ , dimensionless	$G$	Gas flow rate (Fig. 16 only), $\text{lb}/(\text{s ft}^2)$ ( $\text{kg}/(\text{s m}^2)$ )
$a$	Effective interfacial mass transfer area per unit volume, $\text{ft}^2/\text{ft}^3$ ( $\text{m}^2/\text{m}^3$ )	$G$	Gas flow rate, $\text{lb}/\text{h}$ ( $\text{kg}/\text{h}$ )
$A'$	Modified absorption factor, given by Eq. (31b)	$g_c$	Conversion factor, $32.2$ ( $\text{lb ft}/(\text{lbf s}^2)$ ) ( $1.0(\text{kg m})/(\text{N s}^2)$ )
$A_e$	Effective absorption factor, given by Eq. (31a)	$G_M$	Molar gas-phase mass velocity, $\text{lb mol}/(\text{h ft}^2)$ [ $\text{kmol}/(\text{s m}^2)$ ]
B	Component B	$G'_M$	Molar gas-phase mass velocity of rich gas, $\text{lb mol}/(\text{h ft}^2)$ [ $\text{kmol}/(\text{s m}^2)$ ]
$b$	Number of moles of component B reacting with 1 mole of component A	$H$	Enthalpy, $\text{Btu}/\text{lb mole}$ ( $\text{kJ}/\text{kmol}$ ) (Fig. 12 and Eq. (33) only)
C	Component C	$H$	Henry's Law constant, $\text{atm}$ ( $\text{kPa}$ )
$c$	Number of moles of component C produced when 1 mole of component A reacts with $b$ moles of component B	$h$	Height parameter for packed towers, $\text{ft}$ ( $\text{m}$ )
$C_A$	Concentration of reactant A in the liquid, $\text{lb mole}/\text{ft}^3$ ( $\text{kg mole}/\text{m}^3$ )	$H_a$	Hatta number, defined by Eq. (12), dimensionless
$C_B$	Concentration of reactant B in the liquid, $\text{lb mole}/\text{ft}^3$ ( $\text{kg mole}/\text{m}^3$ )	$H_G$	Height of a transfer unit based on gas-phase resistance, $\text{ft}$ ( $\text{m}$ )
$C_{B_0}$	Concentration of reactant B in the bulk liquid, $\text{lb mole}/\text{ft}^3$ ( $\text{kg mole}/\text{m}^3$ )	$H_L$	Height of a transfer unit based on liquid-phase resistance, $\text{ft}$ ( $\text{m}$ )
$C_{SB}$	Flooding capacity parameter, given in Fig. 20, $\text{ft}/\text{s}$ ( $\text{m}/\text{s}$ )	$H_{OG}$	Height of an overall gas-phase mass-transfer unit, $\text{ft}$ ( $\text{m}$ )
$D_A$	Diffusion coefficient of component A in the liquid phase, $\text{ft}^2/\text{h}$ ( $\text{m}^2/\text{s}$ )	$h_T$	Contact height, $\text{ft}$ ( $\text{m}$ )
$D_B$	Diffusion coefficient of component B in the liquid phase, $\text{ft}^2/\text{h}$ ( $\text{m}^2/\text{s}$ )	$k_2$	Second order reaction rate constant, $\text{ft}^3/(\text{h lb mol})$ [ $\text{m}^3/(\text{s kmol})$ ]
$E$	Energy transfer rate across interface, $\text{Btu}/\text{h}$ ( $\text{kJ}/\text{s}$ )	$k_G$	Gas-phase mass-transfer coefficient for dilute systems, $\text{lb mol}/(\text{h ft}^2 \text{ mole fraction solute})$ ( $\text{kmol}/(\text{s m}^2 \text{ mole fraction solute})$ )
$F_{iv}$	Flow parameter, $(L/G) \sqrt{\rho_G/\rho_L}$ , dimensionless	$k'_G$	Gas-phase mass-transfer coefficient for concentrated systems, same units as $k_G$
		$k''_G$	Gas-phase mass transfer coefficient for multicomponent systems, same units as $k_G$
		$k_L$	Liquid-phase mass-transfer coefficient for dilute systems, same units as $k_G$
		$k'_L$	Liquid-phase mass-transfer coefficient for concentrated systems, same units as $k_G$
		$k''_L$	Liquid-phase mass transfer coefficient for multicomponent systems, same units as $k_G$
		$k_L^o$	Liquid-phase mass-transfer coefficient for pure physical absorption (no reaction), same units as $k_G$
		$K_{OG}$	Overall gas-phase mass-transfer coefficient for dilute systems, same units as $k_G$
		$K'_{OG}$	Overall gas-phase mass-transfer coefficient for concentrated systems, same units as $k_G$
		$K''_{OG}$	Overall gas-phase mass transfer coefficient for multicomponent systems, same as units as $k_G$
		$L$	Liquid flow rate (Fig. 16 only), $\text{lb}/(\text{s ft}^2)$ ( $\text{kg}/(\text{s m}^2)$ )

$L$	Liquid flow rate, lb/h (kg/h)	$x_{\text{fm}}$	Film factor, given by Eq. (8b)
$L$	Liquid flow rate, lb mole/h (kmol/s) (Fig. 12 and Eq. 33 only)	$y$	Mole fraction solute (in bulk-gas phase, unless otherwise subscripted)
$l$	Liquid component flow rate, lb mole/h (kmol/s)	$y'$	Mole solute in gas per mole of rich gas entering the absorber
$L_M$	Molar liquid-phase mass velocity, lb mol/(h ft <sup>2</sup> ) [kmol/(s m <sup>2</sup> )]	$y_A$	Mole fraction solute A (in bulk-gas phase, unless otherwise subscripted)
$L'_M$	Molar solute-free solvent mass velocity, lb mol/(h ft <sup>2</sup> ) [kmol/(s m <sup>2</sup> )]	$y^*$	Mole fraction solute in bulk-gas in equilibrium with solute concentration in bulk-liquid
$m$	Slope of equilibrium curve $= dy^*/dx$ , dimensionless	$y_A^*$	Mole fraction solute in bulk-gas in equilibrium with solute concentration in bulk-liquid
$N$	Number of stages in a stagewise contactor	$y_{\text{BM}}$	Logarithmic-mean inert-gas concentration between bulk-gas and interface, defined by Eq. (6a)
$N$	Mass transfer rate across interface, lb mole/h (kmol/s) (Fig. 12 and Eq. (32) only)	$y_{\text{BM}}^*$	Logarithmic-mean inert-gas concentration between bulk-gas and value in equilibrium with bulk-liquid
$N_A$	Molar flow rate of solute A per unit interfacial area, lb mol/(h ft <sup>2</sup> ) [kmol/(s m <sup>2</sup> )]	$y_{\text{fm}}$	Film factor, given by Eq. (8b)
$N_B, N_C, \dots$	As $N_A$ , but with respect to solute B, C, ...	$y_{\text{fm}}^*$	Film factor, given by Eq. (8d)
$N_G$	Number of gas-phase mass-transfer units, dimensionless	$\delta$	Film thickness, ft (m)
$N_L$	Number of liquid-phase mass-transfer units, dimensionless	$\mu$	Liquid viscosity, cP [kg/(s m)]
$N_{\text{OG}}$	Number of overall gas-phase mass-transfer units, dimensionless	$\rho$	Density, lb/ft <sup>3</sup> (kg/m <sup>3</sup> )
$P$	Pressure, atm (kPa)	$\sigma$	Surface tension, dyn/cm (N/m)
$p$	Solute partial pressure in bulk gas, atm (kPa)	$\phi$	Ratio $k_L/k_L^0$ , reaction enhancement factor, dimensionless
$Q$	Heat removal rate, Btu/h (kJ/s)	$\phi_\infty$	Ratio $k_L/k_L^0$ when $H_a = \infty$ , dimensionless
$r_A$	Volumetric reaction rate of component A, lb mol/(h ft <sup>3</sup> ) [kmol/(s m <sup>3</sup> )]	$\psi$	Ratio of water to liquid density, dimensionless
$r_B$	Volumetric reaction rate of component B, lb mol/(h ft <sup>3</sup> ) [kmol/(s m <sup>3</sup> )]	Subscripts	
$T$	Temperature, °F (°C)	0	Liquid inlet to stage contactor
$t$	Parameter defined by Eq. (8d), indicating degree of counter-diffusion	1	Column bottom (differential contactor)
$U_{\text{nf}}$	Vapor velocity, based on tray area less the area at the bottom of the downcomer, at the flood point, ft/s (m/s)	1	Stage 1 (Top stage in a stagewise contactor)
$U_s$	Superficial vapor velocity, ft/s	2	Column top (differential contactor)
$V$	Gas flow rate, lb mole/h (kmol/s)	2	Stage 2 (Stagewise contactor)
$v$	Gas component flow rate, lb mole/h (kmol/s)	A	Component A
$x$	Mole fraction solute (in bulk-liquid phase unless otherwise subscripted)	ave	Average for the column
$x'$	Mole solute in liquid per mole of solute-free solvent entering absorber	B	Component B
$x_A$	Mole fraction solute A (in bulk-liquid phase, unless otherwise subscripted)	C	Component C
$x_A^*$	Mole fraction solute in bulk-liquid inequilibrium with solute concentration in bulk-gas	G	Gas phase
$x_{\text{BM}}$	Logarithmic-mean inert-solvent concentration between bulk liquid and interface, given by Eq. (6b)	i	Interface
		j	Component j
		L	Liquid phase
		N	Stage N (bottom stage in a stagewise contactor)
		n	Stage n
		Superscripts	
		I	At the interface
		L	Liquid
		v	Vapor (or gas)