

# ABSORPTION AND STRIPPING



*P. Chattopadhyay*



*Asian Books Private Limited*

# Absorption & Stripping

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# Preface

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**Absorption & Stripping** are essentially two very important unit operations frequently encountered in both CPIs (Chemical Process Industries) and PCIs (Petrochemical Industries). In many plants, absorption & stripping operate in conjunction with distillation—the oldest unit operation that emerged from alchemists' laboratory centuries back. Yet surprisingly there is quite a few titles exist in the market. Of course, I must admit that there are some excellent texts still available on **absorption and stripping**. They're old ones but very good ones. They provide sound theoretical backup to these unit operations. It is here that this present title bears good similarities to them.

However, there is a basic difference between those erstwhile texts and the present one: It is the *industrial approach*. This one banks heavily on industry & focuses its major concern on the industrial application of absorption and stripping inasmuch as *all unit operations must find their ultimate application in industries*. It gives a detail survey of Tower Internals, Design of Absorbers & Strippers, Typical Industrial Absorbers & Strippers, Revamping of Absorbers & Strippers, Cost Estimation of Absorption Towers. Author's two-&-a-half decades of cumulative experience as Assistant Process Engineer & as Senior Process Engineer involved in the operation & troubleshooting of Mass Transfer Equipment (particularly, fractionators, absorbers, strippers, reactors and heat exchangers) has provided the book industrial design concept & many practical tips. Proprietary design data of tower internals have been planted into the book to expand its coverage. As such the book *may* reward the reader with a sense that the book is a complete one—from sound theoretical base to concrete industrial design.

Obviously, the theoretical bases for these design procedures had been developed many years ago. These theories are available in most academic texts the reader may encounter in chemical engineering courses on mass transfer. Unfortunately, the direct application of these theoretical concepts in many practical situations ends up with inaccurate sizing of absorbers & strippers of industrial scale. This is due to lack of physical & chemical constants and mostly because the data in academic texts are based on laboratory columns and pilot plants operating near atmospheric pressure. In sharp contrast, some large-dia industrial columns must operate at high pressures or with foaming systems that are dirty.

Very often than not industrial absorption & stripping columns are fretted with the nagging problems of corrosion, side reactions, foaming, packing degradation, and the like. And that renders actual plant (or pilot-plant) operating data invaluable adjuncts to a theoretical design. As such operating data have been given due emphasis and inducted wherever possible.

The first two chapters provide the necessary fundamentals & theoretical development of absorbers & strippers. Adequate numerical examples have been dished out to enable the reader to get a good grip of the topics.

Design of all gas-liquid contacting columns begins with the hydraulics of operation. So is this one. **Hydraulics** of all the three basic tray-columns as well as of packed towers have been explained to the minutest details. Discussed also are the factors & parameters that influence the hydraulics of packed towers. This is followed by basic concepts of design of **Tray Towers** and **Packed Towers**. Adequate numerical examples have been plugged in. Two chapters (CH-4 & CH-5) deal exclusively with **design**.

Packings come almost inevitably with Absorption & Stripping. So little wonder why they'll occupy a special position in this book. So the author has devoted one whole chapter (CH-6) on packing.

Equally important are tower internals without which the packing's functions are seriously impaired. Each & every such tower internals has been discussed in comprehensive detail (CH-7).

Finally, the last three chapters on absorption & stripping of **industrial importance, revamping of absorbers & strippers & cost estimation of absorption towers** are a pleasant excursion to the domain of large commercial absorbers & strippers. Design consideration, design guidelines & operation of important industrial absorption have been discussed at length. The author believes that the title will come in good stead to the students of Chemical Engineering and Applied Chemistry as well as Process Engineers and Designers of CPIs and PCIs. Any shortcoming of the book lies entirely on the shoulder of the author.

2nd January, 2007

P. Chattopadhyay

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## NOTATIONS USED IN THE BOOK

BDF	Bottom Driving Force	mtc	mass transfer coefficient
BTEX	Benzene, Toluene, Ethylbenzene & Xylene	NCH	Non-condensable Hydrocarbon
BTMS	Bottoms	NG	Natural Gas
$\textcircled{d}$	Circulated	NTU	Number of Transfer Units
$\textcircled{ng}$	Circulating	OL	Operating Line
$\textcircled{n}$	Circulation	Op. line	Operating Line
Col	Column	OVHD	Overhead
CPI	Chemical Process Industries	Press	Pressure
CTC	Carbon Tetra Chloride	qty	quantity
DCA	1:2-Dichloroethane	scm	Standard cubic meter
DCE	1:1-Dichloroethylene	SG	Sour Gas
DEP	Diethanolpiperazine	sol <sup>n</sup>	solution
DIPA	Di-isopropanolamine	TCA	1:1:1-Trichloroethane
EPA	Environmental Protection Agency	TCE	Trichloroethylene
HAP	Hazardous Air Pollutants	TDF	Top Driving Force
HC	Hydrocarbon	TEG	Triethylene Glycol
HCB	Hexachlorobenzene	Temp	Temperature
HE	Heat Exchanger	THEED	Trihydroxyethylene diamine
htc	heat transfer coefficient	VLE	Vapor Liquid Equilibrium
HTU	Height of a Transfer Unit	VOCs	Volatile Organic Compounds
h/up	hold up		
k\$	kilo-dollar [ 1k\$ = US\$1000 ]		
l/up	lined up		
MDF	Mean Driving Force (usually log-mean-driving force)		
MGD	Million Gallons per Day		
MMBtu	Million Btu		
MMs.ft <sup>3</sup>	Million Standard Cubic foot required referred to 15°C/100kPa		

# Absorption

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**Absorption** is a gas-liq mass transfer operation in which a component is transferred from the gas phase to the liquid. And the rate of absorption is determined by the rate of molecular diffusion that largely controls this interphase mass transfer.

The component which is absorbed is called **solute** and in which it is absorbed is called **solvent**.

Generally, the solute enters the column in a gas introduced at the bottom of a column while the solvent is fed to the top as liquid. The solute is more or less soluble in the solvent while the carrier gas is either insoluble or slightly soluble in the liquid phase. The absorbed gas and solvent leave at the bottom and the unabsorbed components plus some of the liquid vaporized into the gas phase leave as gas from the top.

The absorbed solute may form a simple solution in the liquid phase or it may react chemically with a component in the liquid phase. Therefore, the absorption processes are conveniently divided into two main groups :

- **Physical Absorption** – in which the process is solely physical and is limited to the formation of solution of the gas in the liquid, *e.g.*, absorption of ammonia by water from an air-ammonia mixture; similarly liquid methanol at low temperature absorbs  $\text{CO}_2$  and  $\text{H}_2\text{S}$  at high pressure and forms their solution.
- **Chemical Absorption** — in which absorption follows incipient chemical reaction, *e.g.*, absorption of  $\text{CO}_2$  in hot alkali and absorption of  $\text{NO}_x$  in water.

Gas absorption is a major unit operation for the selective removal of one or more components from a gas mixture by a suitable liquid forming a solutions of the gases upon absorption. The solvent is regenerated from the solution by a process called **Desorption**.

**Desorption** (or **Stripping**) is the just reverse of gas absorption. The rich solution, *i.e.*, the solvent loaded with absorbed solute (or solutes) is charged to the regeneration tower (*i.e.*, stripping column) at the top and the stripping stream (usually saturated steam) is introduced at the bottom. Upon gas-liq contact, mass transfer occurs in the opposite direction, *i.e.*, physical transfer of solute from the liq phase to the gas phase. The lean solution (*i.e.*, the solution stripped off much of its gas load) recovered from the bottom of the column is recycled to the absorption tower to ensure continuous operation.

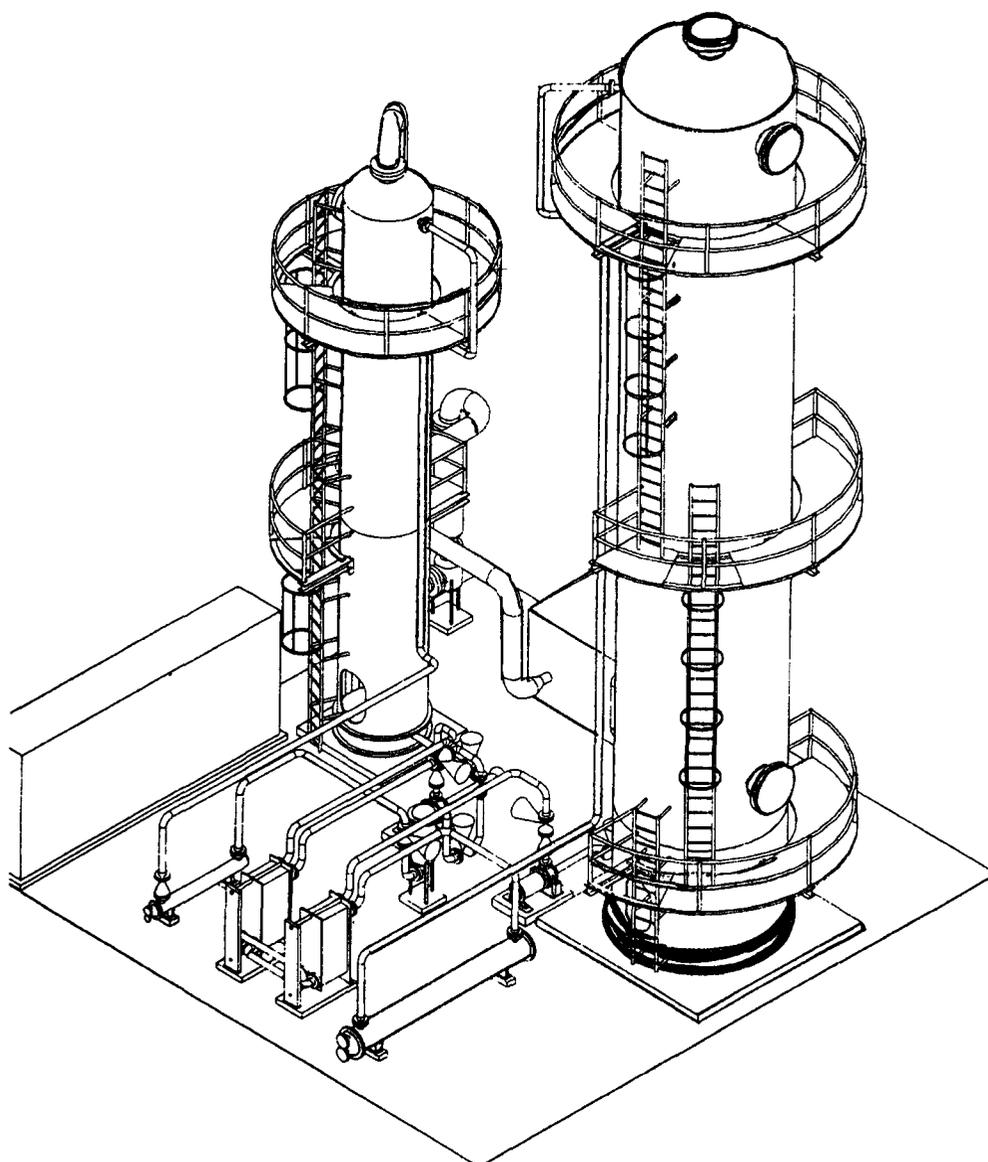
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This is incidental to the absorption operation

Absorption and desorption are traditional thermal separation processes. A complete absorption process comprises an absorber and a desorber **Figure 1.1**.

In the absorber the scrubbing liquor (solvent) is charged with the gaseous component (components) to be removed, and in the desorber (regenerator) it is regenerated or freed from the substances dissolved in it. The stripped solvent is pumped back to the absorption tower to complete the cycle.

**Absorption plant with absorption and regeneration column.**



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**Fig. 1.1. Absorption Tower Hooked up with a Regeneration Column completes  
A Commercial Absorption Process.**

Source : Absorption Technology — Sulzer Chemtech

Be it absorption or desorption—the intimate gas-liq contact is essential in both cases and as such the effectiveness of the equipment is, by and large, determined by the success with which it promotes contact between the two phases.

**Absorption differs from distillation in the following aspects :**

- Stripping vapor is generated in the distillation column by the partial evaporation of the liquid which is therefore at its boiling point, whereas in absorption the liquid is always well below its boiling point.
- Distillation is characterized by simultaneous molecular diffusion in both directions and in ideal systems equimolar counter-diffusion occurs across the gas-liq phase boundaries of these two contacting streams. But in gas absorption, diffusion is chiefly unidirectional — the solute molecules are diffusing into the liquid while the movement in the reverse direction is practically very small.
- The ratio of the liq flowrate to the gas flowrate is considerably greater in absorption than in distillation with the effect that the layout of trays is different in these two cases.
- The primary objective of absorption is only solute recovery or solute removal while distillation involves separation of solutes from each other to any important extent.

## **1.1. APPLICATIONS**

Gas absorption technology finds its commercial application in the following fields :

### **The Gas Industry**

- Gas dehydration
- Removal of CO<sub>2</sub> and H<sub>2</sub>S
- Selective absorption of H<sub>2</sub>S

### **Refineries**

- Hydrocarbon absorbers for lean oil etc.)
- H<sub>2</sub>S absorbers (MEA, DSA, etc.)
- Various types of stripping columns
- Sour water strippers

### **The Petrochemical Industry**

- Synthesis gas processing
- Gas saturation
- Ethylene oxide absorption
- Acrylonitrile absorption

### **The Chemical Industry**

- Synthesis gas processing (CO<sub>2</sub> removal, Saturation)
- Chlorine drying
- HCl and ammonia absorption
- Absorption of nitrous gases

**The Cellulose Industry**

- Sulfur dioxide absorption
- Chlorine dioxide absorption
- Flue gas scrubbing with sulfur recovery.

**Food Processing**

- Stripping various components producing odours (*i.e.*, deodorization)
- Processing fatty acids
- Hexane absorption and stripping

**The Metal & Packaging Industries**

- Absorption of triethylamine (in foundries)
- Absorption of lube & cooling oils
- Absorption of nitrous gases
- Absorption and recovery of solvent vapors

**Exhaust Air Scrubbing**

- Removal of acid components (wet-and dry-scrubbing of  $\text{SO}_x$  &  $\text{NO}_x$ )
- Removal of base components
- Removal & recovery of organic solvents

**Wastewater/sewage Treatment and Pollution Control**

- Airstripping of chlorinated hydrocarbons
- Desorption & recovery of ammonia
- Effluent neutralization
- Deaeration of seawater.

**1.2. GAS-LIQ EQUILIBRIUM : CONDITIONS OF**

The liq and the gas phases, when brought into contact, tend to reach equilibrium. The rate at which a gaseous component from a feed gas mixture will dissolve in an absorbent liquid depends upon the departure from equilibrium which exists.

The solubility of any gas in a liquid [defined as the resulting concentration of the dissolved gas in the liquid at the prevailing pressure and temperature when the equilibrium is established] is influenced by the temperature and pressure in a manner described by van't Hoff's law of dynamic equilibrium :

- *at a fixed temperature, the solubility concentration will increase with pressure*
- *if, on the other hand, temperature of a gas-liq system in equilibrium is raised, that change will occur which will absorb heat.* Frequently, the dissolution of a gas in a liq results in an evolution of heat and it follows, therefore, that in most cases the solubility of a gas decreases with increasing temperature.

If the concentration of solute (dissolved gas) in the liquid phase is small and the solute forms a simple solution, Henry's Law applies :

$$p^* = H \cdot x \quad \dots(1.1)$$

where,  $p^*$  = partial pressure of solute in the gas phase over the liquid in equilibrium with the gas, Pa.

It is also the vapor pressure of solute as the latter is very little soluble in liquid phase & resides mostly in gas phase.

$H$  = Henry's Law constant, Pa/mol fraction

$x$  = mol fraction of solution in liq phase.

This equilibrium relationship is valid for dilute solutions of most gases and over a wide range for some gases.

The partial pressure of solute in the gas phase is a function of the gas composition :

$$p = y \cdot P \quad \dots(1.2)$$

where,  $p$  = partial pressure of solute in gas phase, Pa

$y$  = mol fraction of solute in gas phase

$P$  = total system pressure, Pa

Combining these Equations (1.1) and (1.2) we get :

$$y^* = \frac{H \cdot x}{P} \quad \dots(1.3)$$

where,  $y^*$  = equilibrium mol fraction of solute in gas phase.

Since, Partial Pressure = Mol Fraction  $\times$  Total Pressure

$$p^* = y^* \cdot P$$

Eqn. (1.3) is the expression for the vap-phase concentration of solute in equilibrium with the liq phase.

Now, the physical transfer of solute from the gas phase to the liquid (*i.e.*, absorption process) will occur whenever the partial pressure of solute in gas phase (Eqn. 1.2) exceeds the vapor pressure of solute above liq phase (Eqn. 1.1).

If the temperature of the liq phase is gradually increased to its boiling point, its vapor pressure will approach the system pressure. Thus, at the solvent (liq phase) boiling temperature, the solubility of the solute is reduced to zero\*. The vapor pressure of the solute gas also increases with increasing temperature. Therefore, the Henry's Law constant increases with the rising of liq-phase temperature [Eqn. 1.1]. Now, as per Eqn. 1.3, the solubility of a gaseous solute in the liquid phase, at constant gas composition and pressure, is inversely proportional to the Henry's Law constant. Therefore, with the rise of liq-phase temperature, the concentration of solute in liquid phase decreases.

Eqn. 1.1 enables us to calculate the vapor pressure of the solute only for low concentrations of the solute in the liquid phase. Should this equation be applied at higher concentrations of the solute in the liquid phase, the value of Henry's Law constant,  $H$ , must be modified. In case the system pressure exceeds 1000 kPa, a correction factor to account for the pressure effect may be introduced.

---

\*This is the basic principle of desorption. Also it explains why the temperature of absorbent (liq solvent) is kept well below its boiling temperature in the absorption tower.

### 1.3. DRIVING FORCE

Absorption is the physical transfer of solute from the gas phase to the liquid phase. It is a diffusional mass transfer operation that occurs across the gas-liq interface. Since the solute is diffusing from the gas phase into the liquid, there must be a concentration gradient in the direction of mass transfer within each phase. Of course, it is the difference of chemical potential of solute in the gas phase and in the liquid phase that acts as the real driving force of absorption and determines the rate of this interphase mass transfer.

For any mass transfer operation between two gases in contact, it is the departure from the state of dynamic equilibrium of the two phases that generates the driving force. This driving force is measured by the difference between the chemical potentials ( $\mu_y - \mu_x$ ) of the transferring component (solute) at equal temperature and pressure of the phases. When the phases are in equilibrium, the driving force is nil, so

	$\mu_y - \mu_x = 0$
<i>i.e.,</i>	$\mu_y = \mu_x$
<b>For Absorption :</b>	$\mu_y > \mu_x$
<b>For Desorption :</b>	$\mu_x > \mu_y$

Now substituting for  $\mu_x$  the potential of the equilibrium gaseous phase  $\mu_y^*$  equal to it, the driving force becomes :

$$\mu_y - \mu_x = \mu_y - \mu_y^* \quad \dots(1.3)$$

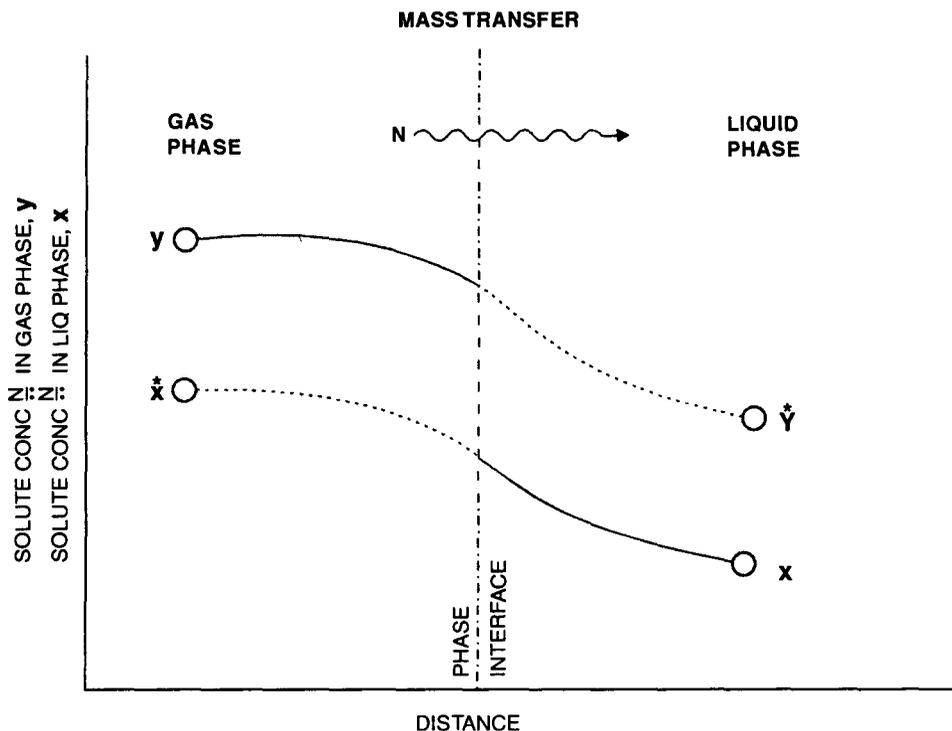


Fig. 1.3.1. Mass Transfer Mechanism.

Similarly, substituting for the chemical potential  $\mu_y$  the potential of the equilibrium liquid phase  $\mu_x^*$  equal to it, we get :

$$\mu_y - \mu_x = \mu_x^* - \mu_x \quad \dots(1.4)$$

The pair of Eqns. 1.3 and 1.4 lends us two different ways to express the difference between the chemical potentials  $\mu_y - \mu_x$  :

- with respect to gas phase
- with respect to gas phase

It must be remembered that the gas phase with chemical potential  $\mu_y^*$  and solute concentration  $y^*$  is a hypothetical phase, so is the liquid phase with the chemical potential  $\mu_x^*$  and solute concentration  $x^*$  — in a real process of mass transfer they are absent (**Figure 1.3.1**)

In practical calculations, the chemical potentials are not used : instead they are replaced by the concentration terms ( $x, y$  etc.,) which are simpler quantities and they are easy to determine. They can also be used to characterize the deviation of the gas and liquid phases from equilibrium. However, unlike  $(\mu_y - \mu_x)$ , the difference between the concentration  $(y - x)$  never equal zero upon equilibrium and hence it cannot be the measure of the deviation of the phases from the equilibrium state, *i.e.*, cannot be the driving force of a mass transfer process. Therefore, for the sake of practical calculation, the driving force of any mass transfer process — the deviation of the system from the state of equilibrium — is expressed as the difference :

## ABSORPTION PROCESS

$y - y^*$  = the driving force with respect to the gas phase

$x^* - x$  = the driving force with respect to the liquid phase

Forasmuch as the concentration can be expressed in different units, the driving force of any mass transfer process, accordingly, can have different units (**Table 1.1**)

**Table 1.1. Driving Force of Absorption**

Phase	The Driving Force	
	Expression	Unit
Gas	$\Delta p = p - p^*$	mm Hg or Pa
Gas	$\Delta y = y - y^*$	mol fraction
Gas	$\Delta Y = Y - Y^*$ $= \frac{y}{1-y} - \frac{y^*}{1-y^*}$	mol ratio
Liquid	$\Delta x = x^* - x$	mol fractions
Liquid	$\Delta X = X^* - X$ $= \frac{x^*}{1-x^*} - \frac{x}{1-x}$	mol ratio

### 1.3.1. Mean Driving Force

The general equation of absorption when the driving force is expressed with respect to gaseous phase as  $\Delta y = y - y^*$  is

$$N = K_{G,y} \cdot A \cdot \Delta y_m \quad \dots(1.5)$$

and when the driving force is expressed with respect to liquid phase as  $\Delta x = x^* - x$ , it becomes

$$N = K_{L,x} \cdot A \cdot \Delta x_m \quad \dots(1.6)$$

These two general expressions of mass transfer results from its complete analogy to heat transfer.

$N$  = molar flowrate of solute from gas phase to liquid phase, kmol/h

$A$  = area of mass transfer surface,  $m^2$

$K_{G,y}$  = overall liq phase mass transfer coefficient related to the driving force  $\Delta y$ , kmol/ $(m^2 \cdot h)$  or, kmol/ $(m^2 \cdot h \cdot kmol/kmol)$

$K_{G,y}$  = overall gas phase mass transfer coefficient related to the driving force  $\Delta y$ , kmol/ $(m^2 \cdot h)$  or, kmol/ $(m^2 \cdot h \cdot kmol/kmol)$

$\Delta y_m$  &  $\Delta x_m$  = mean driving forces of the entire process with respect to gas phase and liquid phase respectively

The Equation 1.5 can be expressed with  $\Delta y_m$  replaced by  $\Delta Y_m$ ,  $\Delta c_{y_m}$  and  $\Delta p_m$  :

$$N = K_{G,y} \cdot A \cdot \Delta Y_m \quad \dots(1.5A)$$

$$N = K_{G,c} \cdot A \cdot \Delta c_{y_m} \quad \dots(1.5B)$$

$$N = K_{G,p} \cdot A \cdot \Delta p_m \quad \dots(1.5C)$$

Likewise, Eqn. 1.6 can be expressed with  $\Delta x_m$  replaced by  $\Delta X_m$  and  $\Delta c_{x_m}$  :

$$N = K_{L,x} \cdot A \cdot \Delta X_m \quad \dots(1.6A)$$

$$N = K_{L,c} \cdot A \cdot \Delta c_{x_m} \quad \dots(1.6B)$$

If the rate of flow of the component being absorbed is expressed in kg/h, the general equation of mass transfer becomes :

$$M' = K_{G,\bar{y}} \cdot A \cdot \Delta \bar{Y}_m \quad \dots(1.7)$$

where,  $M'$  = mass flowrate of diffusing solute, kg/h

$K_{G,\bar{y}}$  = overall gas phase mass transfer coefficient, kg/ $(m^2 \cdot h)$

$A$  = surface area of mass transfer in the absorber,  $m^2$

$\Delta \bar{Y}_m$  = mean driving force

$$= \frac{\Delta \bar{Y}_b - \Delta \bar{Y}_t}{\ln \left[ \frac{\Delta \bar{Y}_b}{\Delta \bar{Y}_t} \right]} \quad \dots(1.8)$$

$\Delta \bar{Y}_b = \bar{Y}_b - \bar{Y}_b^*$  = terminal driving force at the bottom of the absorber when  $\bar{X} = \bar{X}_b$

$\Delta \bar{Y}_t = \bar{Y}_t - \bar{Y}_t^*$  = terminal driving force at the top of the absorber when  $\bar{X} = \bar{X}_t$

where  $\bar{X}$  = mass ratio of solute in liq phase

$$= \left[ \frac{\text{kg of Solute}}{\text{kg of Remaining Components}} \right] \text{ in liq phase}$$

$\bar{Y}$  = mass ratio of solute in gas phase

$$= \left[ \frac{\text{kg of Solute}}{\text{kg of Remaining Components}} \right] \text{ in gas phase}$$

**Case – (I)** In case

$$\frac{1}{2} \leq \left[ \frac{\Delta \bar{Y}_b}{\Delta \bar{Y}_t} \right] \leq 2$$

use the simpler formula :

$$\Delta \bar{Y}_m = \frac{1}{2} [\Delta \bar{Y}_b + \Delta \bar{Y}_t] \quad \dots(1.9)$$

to calculate the mean driving force in an absorber.

**Case – (II)** In case the VLE (vapor-liq equilibrium) line is not straight, then

$$\Delta \bar{Y}_m = \frac{\Delta \bar{Y}_b - \bar{Y}_t}{\int_{\bar{Y}_t}^{\bar{Y}_b} \frac{d\bar{Y}}{\bar{Y} - \bar{Y}^*}} \quad \dots(1.10)$$

The value of the integral  $\int_{\bar{Y}_t}^{\bar{Y}_b} \frac{d\bar{Y}}{\bar{Y} - \bar{Y}^*}$  can be obtained either by the method of graphical integration

or by graphical construction.

Frequently the driving force is expressed in units of pressure while calculating absorbers. The pressure driving force (*i.e.*, pressure difference) must be determined at the bottom ( $\Delta p_b$ ) as well as at the top ( $\Delta p_t$ ) of the absorber.

For a gas stream containing a low concentration of solute, the ratio of liq to vapor flowrate (**L/G**) is almost constant and the operating line of the absorber is straight. Now if absorption accompanies negligible heat of solution, the pressure driving force, under these conditions, is the logarithmic mean of the driving forces at the bottom and top of the column :

$$\Delta p_{lm} = \frac{\Delta p_b - \Delta p_t}{\ln \left[ \frac{\Delta p_b}{\Delta p_t} \right]} \quad \dots(1.11)$$

and the mass transfer equation becomes :

$$N = K_{G,p} \cdot A \cdot \Delta p_{lm} \quad \dots(1.12)$$

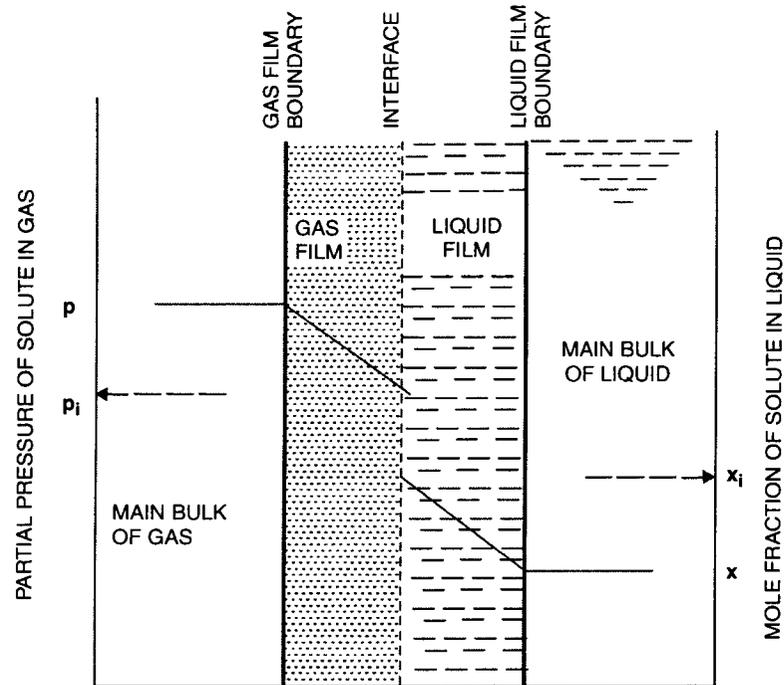
where,  $N$  = rate of solute transfer from gas to liq phase, kmol/h

$K_{G,p}$  = overall gas phase mass transfer coefficient in terms of partial pressure, kmol/(m<sup>2</sup>. h. Pa)

## 1.4. ABSORPTION MECHANISM

The most widely accepted theory to explain gas-liq mass transfer operations is the **double-film theory** of Lewis and Whitman (W. K. Lewis and W. G. Whitman — *Industrial and Engineering Chemistry, Vol. 16 (1924)* :

The boundary between the gas phase and liquid phase in contact is presumed to be composed of two films — a gas film and a liq film — separated by an interface **Figure 1.4.1**. The gas film is adjacent to the main bulk of gas and the liq film to the main bulk of liq.



**Fig. 1.4.1. Double-Film Theory of Mass Transfer Mechanism in Absorption.**

Diffusional resistances reside only in the fluids themselves and there is no resistance to solute transfer across the interface separating the phases. Therefore, the solute concentration in the gas film at the interface is assumed to be in equilibrium with the solute concentration in the liquid film at the interface.

Concentration gradient exists in both films and the flow in both these films is assumed to be laminar or stagnant. However, no concentration gradient exists in the main body of both the gas and liq phases because of prevailing turbulences that thoroughly mix up either phase.

The driving force causing solute transfer in the gas phase :

$$p - p_i \equiv DE$$

**Fig. 1.4**

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

$$x_i - x \equiv BE$$

**Fig. 1.4**

In complete analogy to heat transfer wherein the flow of heat is equal to the product of a heat transfer coefficient, a transfer surface area and a driving force, the rate of mass transfer from the main body of gas phase thru the gas film in absorption is given by the Eqn.

$$N = k_{G,P} \cdot A \cdot (p - p_i) \quad \dots(1.13)$$

where,  $k_{G,P}$  = gas film transfer coefficient in terms of partial pressure,  $\text{kmol}/(\text{h} \cdot \text{m}^2 \cdot \text{Pa})$

$p - p_i$  = difference between the partial pressure of solute in the main-body gas phase and that in the gas film at the interface. It is the driving force for mass transfer across the gas film.

- A = area of mass transfer surface. In absorber design calculations it is taken equal to column cross-sectional area,  $m^2$ .

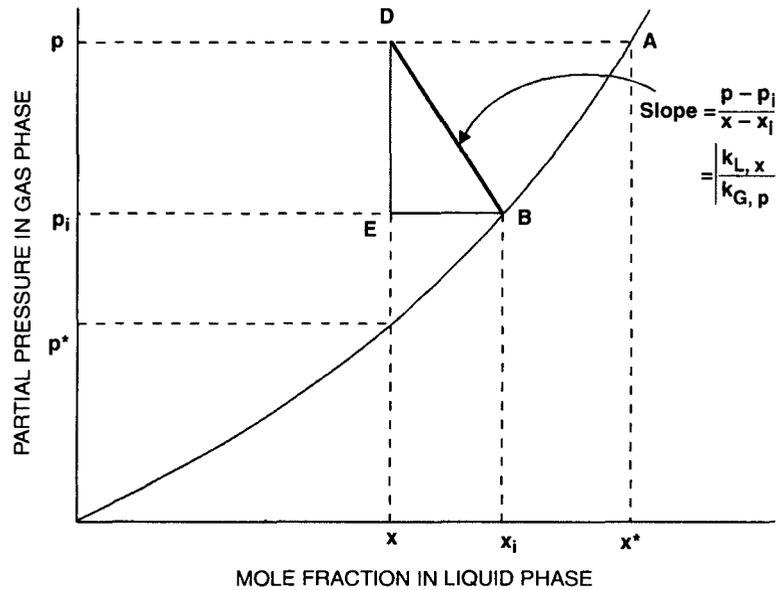


Fig. 1.4.2. Driving Forces in Gas & Liq Phases in the Course of Absorption.

Similarly the rate of solute transfer across the liq film to the main bulk of liq phase is given by the expression :

$$N = k_{L,x} \cdot A \cdot (x_i - x) \quad \dots(1.14)$$

where,  $k_{L,x}$  = liq film transfer coefficient in terms of mol fraction,  $kmol/(h.m^2.kmol/kmol)$

$x_i - x$  = concentration difference across the liq film. It is the driving force across the liq film.

***In the steady-state process of absorption, the rate of transfer of material thru the gas film equals the rate of transfer of the material thru the liq film*** and with that the general equation of mass transfer may be represented as :

$$N = k_{G,p} \cdot A \cdot (p - p_i) = k_{L,x} \cdot A \cdot (x_i - x)$$

$$\frac{k_{L,x}}{k_{G,p}} = \frac{p - p_i}{x_i - x} = \text{ratio of driving forces in gas phase to liquid phase} \quad \dots(1.15)$$

The value of  $x_i$  and  $p_i$  are difficult to determine in practical cases. Hence overall terms have been adopted for calculative purposes.

For absorption involving highly soluble solutes, the driving force usually is the partial pressure of the solute in the gas phase minus the vapor pressure of the solute above the liquid phase, *i.e.*, the driving force is  $p - p^*$  [or  $p - H \cdot x$  where Henry's Law is applicable]. The greater this driving force, the faster will be the rate of mass transfer. Thus the overall mass transfer analogous equation for the gas phase is :

$$N = K_{G,p} \cdot A \cdot (p - p^*) \quad \dots(1.16)$$

where,  $K_{G,p}$  = overall gas-phase mass transfer coefficient in terms of partial pressure,  $kmol/(h.m^2.Pa)$

Likewise the driving force in liquid phase for the diffusion of highly soluble solutes during absorption is  $x^* - x$  [Or,  $p/H - x$  where Henry's Law applies]. Therefore, for practical purposes the rate of mass transfer in liquid phase in absorption process is :

$$N = K_{L,X} \cdot A \cdot (x^* - x) \quad \dots(1.17)$$

where,  $K_{L,X}$  = overall liq-phase mass transfer coefficient in terms of mol fraction, kmol/(h.m<sup>2</sup>.mol/mol)

For steady-state absorption,

$$N = K_{G,P} \cdot A \cdot (p - p^*) = K_{L,X} \cdot A \cdot (x^* - x)$$

$$\text{or,} \quad \frac{K_{G,P}}{K_{L,X}} = \frac{x^* - x}{p - p^*} \quad \dots(1.18)$$

If the partial pressure terms are replaced by concentration terms  $c$ , the Eqn. 1.16 would become

$$N = K_{G,C} \cdot A \cdot (c - c^*) \quad \dots(1.19)$$

whereupon the Eqn. 1.18 would take the shape of :

$$\frac{K_{G,C}}{K_{L,X}} = \frac{x^* - x}{c - c^*} \quad \dots(1.20)$$

#### 1.4.1. Overall and Film Transfer Coefficients : Interrelationships

The rate of mass transfer of solute in gas phase :

$$\text{Film Transfer} \quad N = K_{G,Y} \cdot A \cdot (y - y_i) \quad \dots(1.21)$$

$$\text{Overall Transfer} \quad N = K_{G,Y} \cdot A \cdot (y - y^*) \quad \dots(1.22)$$

The rate of mass transfer of solute in liq phase :

$$\text{Film Transfer} \quad N = k_{L,X} \cdot A \cdot (x_i - x) \quad \dots(1.14)$$

$$\text{Overall Transfer} \quad N = K_{L,X} \cdot A \cdot (x^* - x) \quad \dots(1.17)$$

For steady-state absorption :

$$\frac{N}{A} = k_{G,Y} (y - y_i) = K_{G,Y} (y - y^*) = k_{L,X} (x_i - x) = K_{L,X} (x^* - x) \quad \dots(1.23)$$

Now, combining Eqns. 1.21 & 1.22 we get :

$$\frac{1}{K_{G,Y}} = \frac{1}{k_{G,Y}} \cdot \left[ \frac{y - y^*}{y - y_i} \right] \quad \dots(1.24)$$

And combining Eqns. (1.14) and (1.7) we get :

$$\frac{1}{K_{L,X}} = \frac{1}{k_{L,X}} \cdot \left[ \frac{x^* - x}{x_i - x} \right] \quad \dots(1.25)$$

Expanding Eqn. 1.24 results :

$$\frac{1}{K_{G,Y}} = \frac{1}{k_{G,Y}} \cdot \left[ \frac{y - y_i}{y - y_i} \right] + \frac{1}{k_{G,Y}} \cdot \left[ \frac{y_i - y^*}{y - y_i} \right] = \frac{1}{k_{G,Y}} + \frac{1}{k_{G,Y}} \cdot \left[ \frac{y_i - y^*}{y - y_i} \right]$$

But from Eqn. 1.23 :

$$\frac{1}{k_{G,Y}} = \frac{1}{k_{L,X}} \cdot \left[ \frac{y - y_i}{x_i - x} \right]$$

$$\therefore \frac{1}{K_{G,Y}} = \frac{1}{k_{G,Y}} + \frac{1}{k_{L,X}} \cdot \left[ \frac{y_i - y^*}{x_i - x} \right] \quad \dots(1.26)$$

From the geometry of Figure 1.5, slope of the chord  $CM$

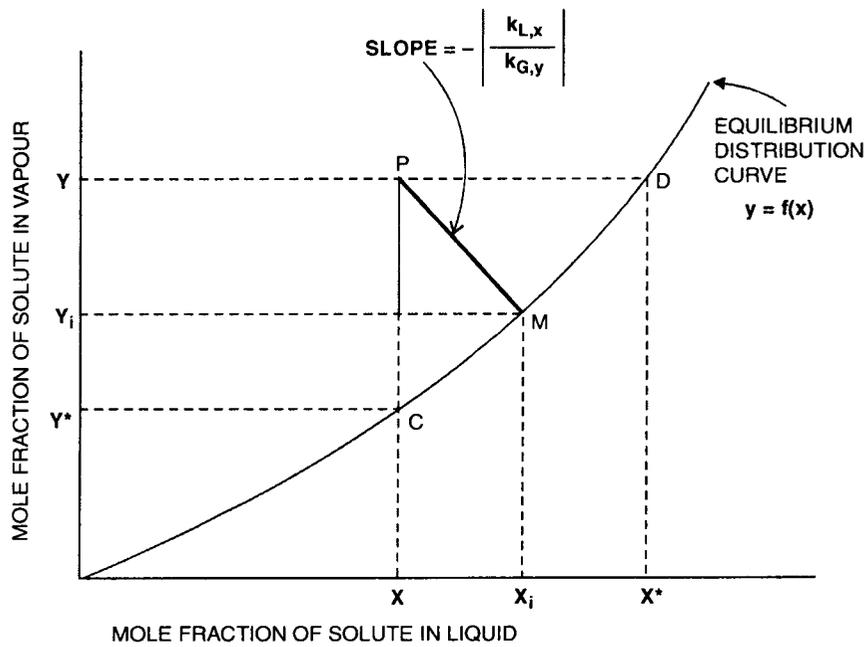


Fig. 1.5. *Equilibrium Distribution Curve*

$$m = \left[ \frac{y_i - y^*}{x_i - x} \right]$$

Therefore, Eqn. 1.26 reduces to

$$\frac{1}{K_{G,y}} = \frac{1}{k_{G,y}} + \frac{m}{k_{L,x}} \quad \dots(1.27)$$

Similarly, from Eqn. 1.25 we get :

$$\begin{aligned} \frac{1}{K_{L,X}} &= \frac{1}{k_{L,X}} \cdot \left[ \frac{x^* - x_i}{x_i - x} \right] + \frac{1}{k_{L,X}} \cdot \left[ \frac{x_i - x}{x_i - x} \right] \\ &= \frac{1}{k_{L,X}} \cdot \left[ \frac{x^* - x_i}{x_i - x} \right] + \frac{1}{k_{L,X}} \end{aligned}$$