GAS LIQUID OPERATIONS

Mass Transfer II

- The operations include humidification and dehumidification, gas absorption and desorption, and distillation in its various forms, all have common requirement that a gas and a liquid phase comes in contact for the purpose of a diffusional interchange between them.
- The rate of mass transfer, in the equipment used for gas liquid operations, is directly dependent upon the interfacial surface exposed between the phases, and the nature and degree of dispersion of one fluid in the other.
- The equipment can be broadly classified according to whether its principal action is to disperse the gas or the liquid, although in many devices both phases become dispersed.
- The devices in which the gas phase is dispersed into bubbles or foams are sparged and agitated vessels and the various types of tray towers.
- Tray towers are the most important because they produce counter-current multistage contact, whereas simple contactors have many applications.

Sparged Vessels (Bubble Columns)

A sparger is a device for introducing a stream of gas in the form of small bubbles into a liquid.

If the vessel diameter is small, the sparger, located at the bottom of the vessel. For vessel of diameter greater than roughly 0.3 m, it is better to use several orifices for introducing the gas to ensure better gas distribution.

Orifice hole 1.5 to 3 mm dia

Porous plates made of ceramics, plastics, or sintered metals are also used.

It can provide the gentlest of agitation. There is no standardization of the depth of liquid; indeed, very deep tanks, 15 m or more may be advantageous.

Gas-Bubble Diameter

The size of gas bubbles depends upon the rate of flow through the orifices, the orifice diameter, the fluid properties, and the extent of turbulence prevailing in the liquid.

Very slow gas flow rate $Q_{Go} < [20(\sigma d_o g_c)^5/(g\ \Delta\rho)^2\ \rho_L]^{1/6}$ for waterlike liquids, the diameter can be computed by equating the bouyant force on the immersed bubble, $(\pi/6)d_P^3\Delta\rho g/g_c$

This tends to lift the bubble away from the orifice, to the force $\pi d_o \sigma$ due to surface tension, which tends to retain it at the orifice. This provides

$$d_p = \left(\frac{6 d_o \sigma g_c}{g \, \Delta\rho} \right)^{1/3}$$ orifice diameter up to 10 mm

For large liquid viscosities up to 1 kg/m s (1000 cp)

$$d_p = 2.312 \left(\frac{\mu_L Q_{Go}}{\rho_L g} \right)^{1/4}$$

Intermediate flow rates $Q_{Go} > [20(\sigma d_o g_c)^5/(g\ \Delta\rho)^2\ \rho_L]^{1/6}$ but $Re_o < 2100$ these bubbles are larger than those described above , although still fairly uniform, and they form chain rather than separately. For air-water

$$d_p = 0.0287 d_o^{1/2}\ Re_o^{1/3}$$

Where d_p and d_o are in meter and $Re_o = d_o V_o \rho_G/\mu_G = 4wo/\pi\ d_o \mu_G$. For other gases and liquids

$$d_p = \left(\frac{72 \rho_L}{\pi^2 g\ \Delta\rho} \right)^{1/5} Q_{Go}^{0.4}$$

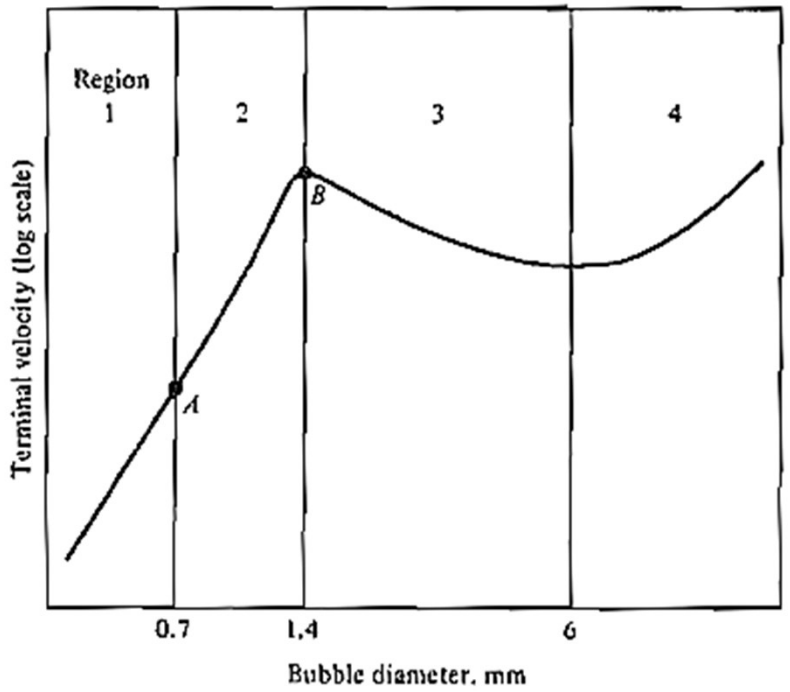
Large gas rates Reo = 10000 to 50000 jets of gas which rise from the orifice break into bubbles at some distance from the orifice. The bubbles are smaller than those described above and non-uniform in size. For air-water and orifice diameters 0.4 to 1.6 mm

$$d_p = 0.0071\ \mathrm{Re}_o^{-0.05}$$ d_p in meters

For the transition range (Re_o = 2100 to 10000) there is no correlation of data. This can be approximated by straight line on log-log scale between the points given by d_p at Re_o 2100 and 10000.

Rising Velocity (Terminal Velocity) of Single Bubbles

The steady state rising velocity of single gas bubbles, which occurs when buoyant force equals the drag force on the bubbles, varies with the bubble diameter as shown in figure.



Region 1, d_p < 0.7 mm The bubbles are spherical, and they behave like rigid spheres, for which the terminal velocity is given by Stokes' law

$$V_t = \frac{g d_p^2 \Delta\rho}{18 \mu_L}$$

Region 2, 0.7 mm < d_p < 1.4 mm The gas within the bubbles circulates, so that he surface velocity is not zero. Therefore, the bubble rises faster than rigid spheres of he same diameter. There is no correlation of data; V_t may be estimated from the figure.

Region 3 (1.4 mm < d_p < 6 mm) and 4 (d_p > 6 mm) The bubbles are no longer spherical and in rising follow a zigzag or helical path.
For region 4 the bubbles have a spherically shaped cap. In both these regions for liquids of low viscosity.

$$V_t = \sqrt{\frac{2\sigma g_c}{d_p \rho_L} + \frac{g d_p}{2}}$$

Gas Holdup

Gas holdup, $\varphi_{G,}$ is defined as the volume fraction of the gas liquid mixture in the vessel occupied by the gas.

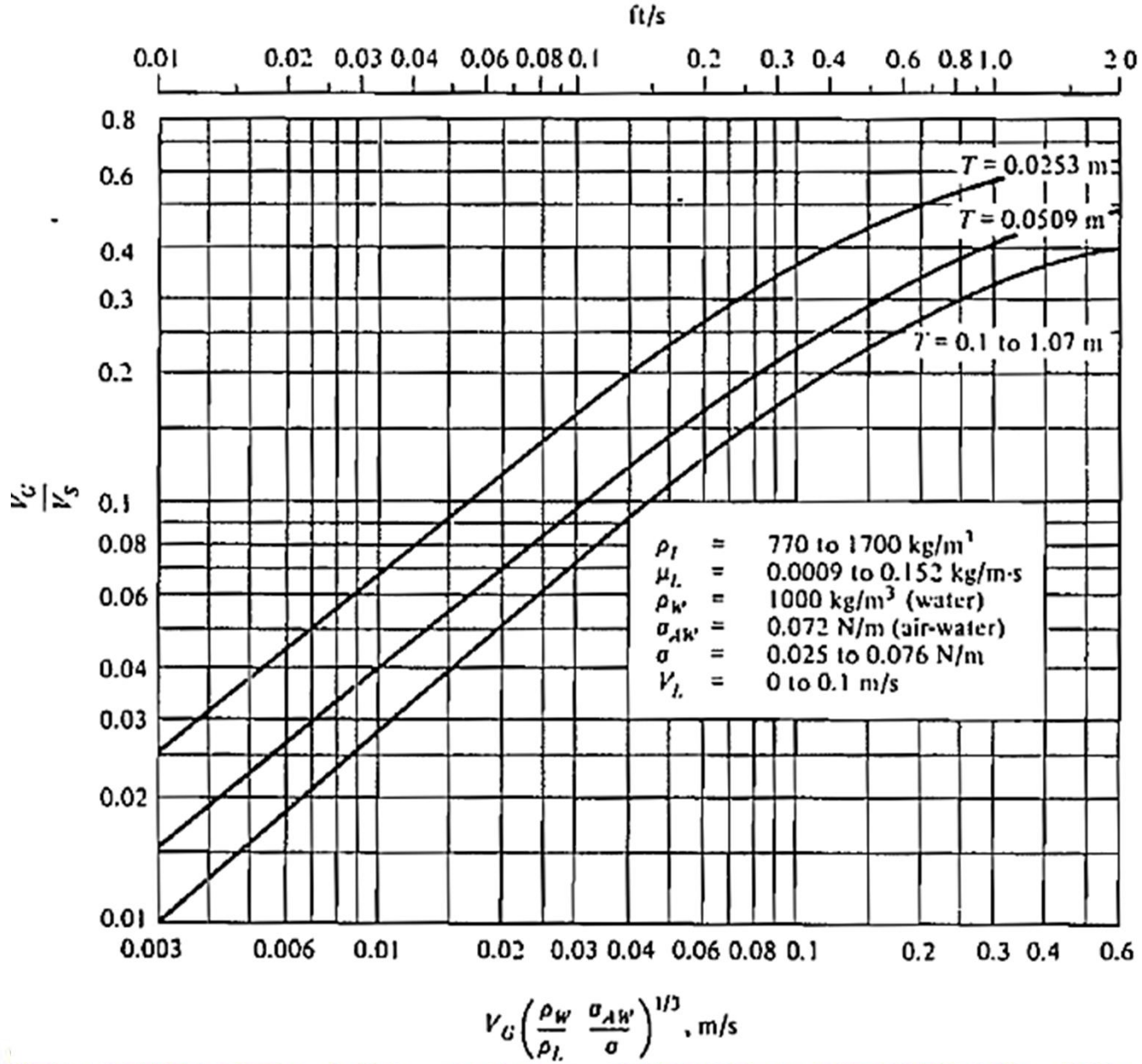
If the superficial gas velocity, defined as the volume rate of gas flow divided by the cross-sectional area of the vessel, is V_G, then V_G/φ_G can be taken as the true gas velocity relative to the vessel walls.

If the liquid flows upward cocurrently with the gas, at a velocity relative to the vessel walls $V_L/(1- \varphi_G)$. The relative velocity of gas and liquid, or slip velocity, is

$$V_S = \frac{V_G}{\varphi_G} - \frac{V_L}{1 - \varphi_G}$$

Equation will also give the slip velocity for counter-current flow of liquid, liquid flow is assigned a negative sign.

The holdup for sparged vessels are correlated through slip velocity, shown in figure. This is satisfactory for no liquid flow ($V_L = 0$), for cocurrent liquid flow up to $V_L = 0.1$ m/s and for small counter-current liquid flow.


ft/s
0.01 0.02 0.03 0.04 0.06 0.08 0.1 0.2 0.3 0.4 0.6 0.8 1.0 2.0
0.8 0.6 0.4 0.3 0.2 0.1 0.08 0.06 0.04 0.03 0.02 0.01
V_G/V_S
T = 0.0253 m
T = 0.0509 m
T = 0.1 to 1.07 m
ρ_l = 770 to 1700 kg/m³
μ_l = 0.0009 to 0.152 kg/m·s
ρ_w = 1000 kg/m³ (water)
σ_{AW} = 0.072 N/m (air-water)
σ = 0.025 to 0.076 N/m
V_l = 0 to 0.1 m/s
0.003 0.006 0.01 0.02 0.03 0.04 0.06 0.08 0.1 0.2 0.3 0.4 0.6
$V_G\left(\frac{\rho_W}{\rho_l}\frac{\sigma_{AW}}{\sigma}\right)^{1/3}$, m/s

Specific Interfacial Area

If unit volume of gas liquid mixture contains a gas volume φ_G made up of n bubbles of diameter d_p then $n = \varphi_G/(\pi d_p^3/6)$. If the interfacial area in the unit volume is a, then $n = a/\pi d_p^2$.

Equating the two expression for n provides the specific area

$$a = \frac{6\varphi_G}{d_p}$$

Thus, for air –water, and in the ranges φ_G = 0.1 to 0.4 and $V_L/(1-\varphi_G)$ = 0.15 to 15 m/s, the bubble size is approximated by

$$d_p = \frac{2.344 \times 10^{-3}}{[V_L/(1-\varphi_G)]^{0.67}}$$

Where d_p and V_L in meters.

Mass transfer

In practically all the gas-bubble liquid systems, the liquid phase mass transfer resistance is strongly controlling, and gas-phase coefficients are not needed.

The liquid-phase coefficients, to within about 15%, are correlated by

$$\mathrm{Sh}_L = \frac{F_L d_p}{cD_L} = 2 + b'\,\mathrm{Re}_G^{0.779}\,\mathrm{Sc}_L^{0.546}\left(\frac{d_p g^{1/3}}{D_L^{2/3}}\right)^{0.116}$$

Where $$b' = \begin{cases} 0.061 & \text{single gas bubbles} \\ 0.0187 & \text{swarms of bubbles} \end{cases}$$

The gas Reynolds number must be calculated with the slip velocity: $Re_G = d_p V_s \rho_L / \mu_L$.

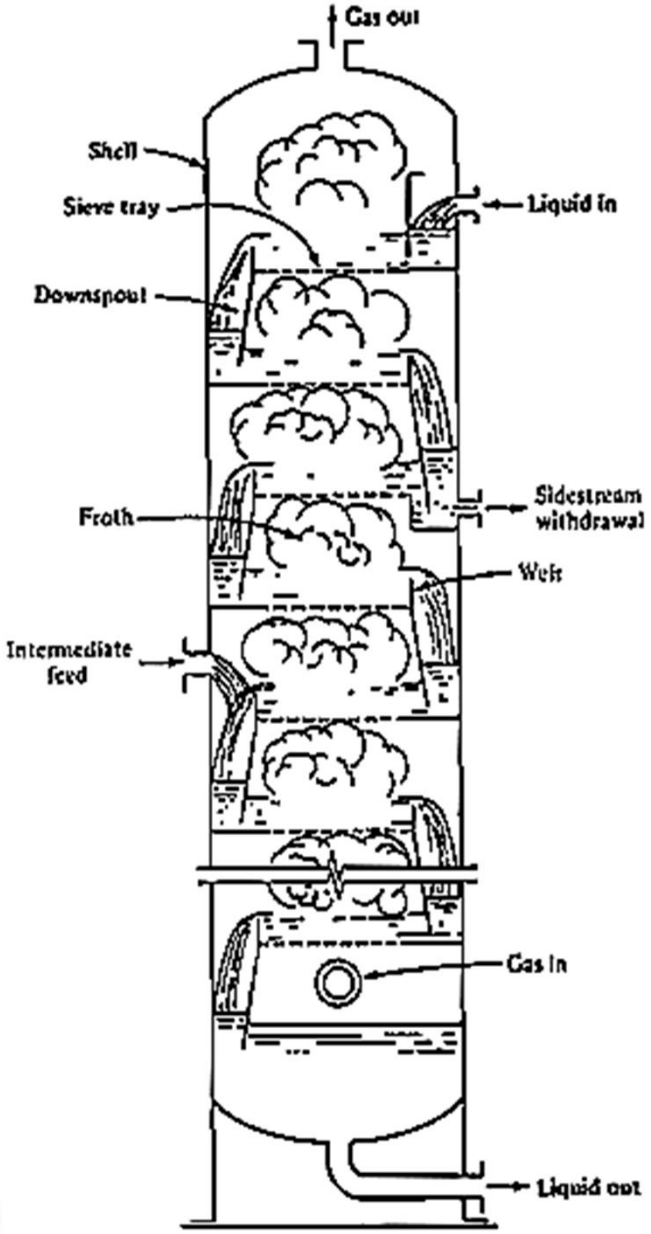
For single bubbles, $\varphi_G = 0$ and $V_S = V_t$. In most cases the liquid will be stirred well enough to permit solute concentrations in the liquid to be considered uniform throughout the vessel.

Tray Towers

Tray towers are vertical cylinders in which the liquid and gas are contacted in stepwise fashion on trays or plates , as shown schematically in figure.

The liquid moving downward by gravity and gases moving upward through the liquid to form a froth, gases disengages from the froth, and passes on to the next tray above.

The overall effect is a multiple countercurrent contact of gas and liquid.



- Each tray of the tower act as a stage, in order to provide a longer contact time, the liquid pool on each tray should be deep so that the gas bubbles will require relatively a longer time to rise through the column of the liquid.
- When the gas velocity is relatively high, it is dispersed very thoroughly into the liquid, which in turn is agitated into froth. This provides large interfacial areas.
- However, these lead to certain operational difficulties like entrainment of droplets of liquid in the rising gas stream and a high pressure drop for the gas flowing through the trays leads to high pumping cost and hence a higher operating cost.
- Sometimes a higher pressure drop also leads to a condition of ***flooding*** in which there will be gradual build-up of liquid in each tray and ultimately fill the entire space between the trays.
- For high gas velocities tend to foam excessively and may lead to a condition of ***priming***. In such case the foam is present in the space between trays and there is a great deal of liquid getting entrained with the gas.
- If the liquid rates are too low, the gas rising through the opening of the tray may push the liquid away, a phenomenon called ***coning*** resulting in poor gas liquid contact.

- When the gas rate is too low, much of the liquid may rain down through the opening of tray, called ***weeping***.
- At very low gas rates, none of the liquid reaches the downspouts and this is known as ***dumping***.

General Characteristics of Tray Towers

Shell and Trays: The tower may be made of any number of materials depending upon the corrosion conditions such as glass, glass-lined metal, impervious carbon, plastics, even wood but most frequently metal is used.

The trays are usually made of sheet metals, of special alloys if necessary, the thickness governed by the anticipated corrosion rate.

The tray must be stiffened and supported and must be fastened to the shell to prevent movement owing to surge of gas, with allowance of thermal expansion.

Tray spacing is chosen on the basis of expediency in construction, maintenance cost, flooding and entrainment.

Tower diameter should be sufficiently large to handle the gas and liquid rates under satisfactory operating conditions. It can also be decreased by the use of increases tray spacing.

The liquid is drawn to next lower tray by means of downcomers or downspouts. These may be circular pipes or portion of the tower cross-section set aside for liquid flow by vertical plates.

The leg of the downcomer will normally dip in the liquid in the next lower tray, which prevents short-circuiting of gas.

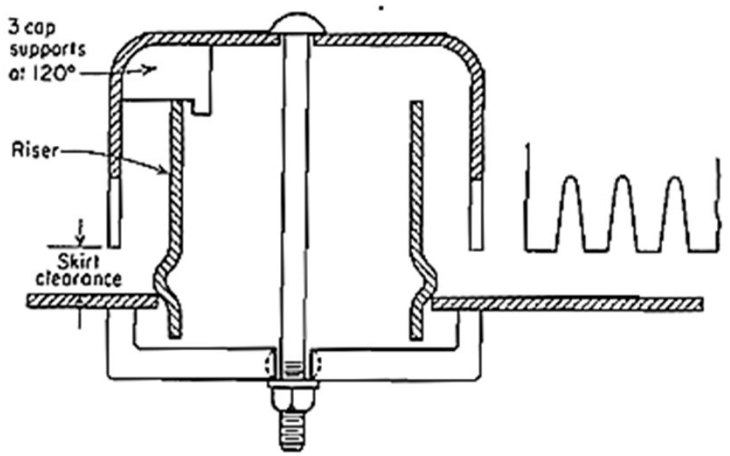
The depth of liquid on the tray required for gas contacting is maintained by overflow weir, which may or may not be a continuation of the downspout plate.

The straight weir are common, V-notch weirs and circular weirs are also used. Weir length varies from 60 to 80% of tower diameter.

Types of Trays

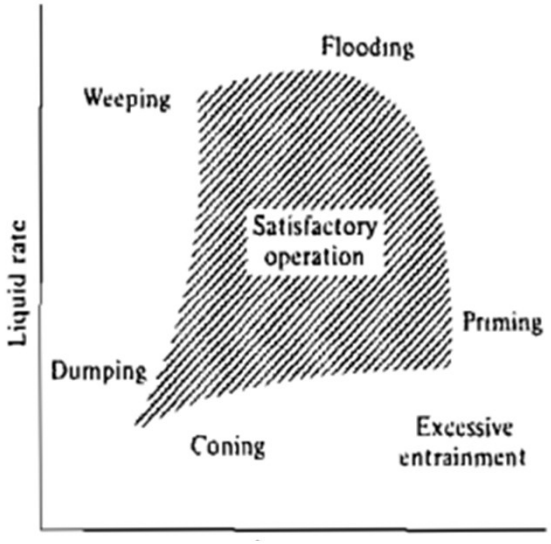
Bubble Cap Trays

In these trays, chimneys or riser lead the gas through the tray and underneath cap surrounding the risers. The gas passes through a series of slots cut into the rim or skirt of each cap. The liquid depth is such that the caps are fully covered by them.



Sieve Trays

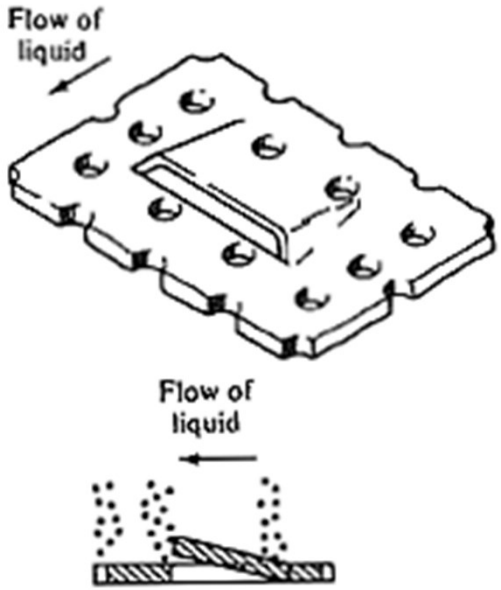
These are trays with perforations and gas flows through them. The gas dispersed by the perforations, expands the liquid into a turbulent froth and results in providing enormous interfacial area for mass transfer. These trays are subject to flooding because of backup of liquid in the downspouts or excessive entrainment. In comparison to bubble caps these are quite simple and are also cost effective.



Proprietary Trays

Linde Trays

These are slotted trays which show an alteration in the perforation pattern to influence the flow of liquid. The slots distributed all over the tray, not only reduce the hydraulic gradient in large trays but also influence the direction of liquid flow and eliminate stagnant areas. Thus, the efficiency of these trays are very high.

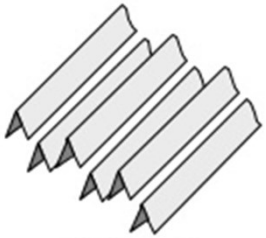


Valve Trays

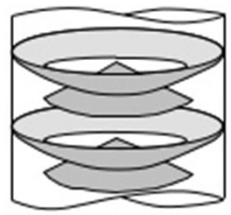
These are sieve trays with large variable openings for gas flow. The perforations are covered with movable caps, which rise as the gas flow rate increases. Though the gas pressure drop is low, it is higher then sieve trays. Due to small openings the tendency to weep is also reduced.

Counter-flow Trays

In this liquid and vapor flow counter-currently through the same openings. Downspouts is absent in these trays. They are more suited for vacuum distillation as the pressure drop is low.



Shed deck



Disc and Donut tray

Ripple Tray™

Plate Efficiency

The designer is concerned with real contacting stages, not the theoretical equilibrium stage assumed for convenience in the mathematical analysis of multistage processes.

Equilibrium will rarely be attained in a real stage. The concept of a stage efficiency is used to link the performance of practical contacting stages to the theoretical equilibrium stage.

Three principal definitions of efficiency are used:

- **Murphree plate efficiency** (Murphree, 1925), defined in terms of the vapor compositions by

$$E_{mV} = \frac{y_n - y_{n-1}}{y_e - y_{n-1}}$$

where y_e is the composition of the vapor that would be in equilibrium with the liquid leaving the plate.

The Murphree plate efficiency is the ratio of the actual separation achieved to that which would be achieved in an equilibrium stage.

In this definition of efficiency, the liquid and the vapor stream are taken to be perfectly mixed; the compositions in equation are the average composition values for the streams.

- **Point efficiency** (Murphree point efficiency).

If the vapor and liquid compositions are taken at a point on the plate, equation gives the local or point efficiency, E_{mv}.

- **Overall column efficiency**.

This is sometimes confusingly referred to as the overall plate efficiency.

$$E_o = \frac{\text{number of ideal stages}}{\text{number of real stages}}$$

An estimate of the overall column efficiency will be needed when the design method used gives an estimate of the number of ideal stages required for the separation.

In some methods, the Murphree plate efficiencies can be incorporated into the procedure for calculating the number of stages and the number of real stages determined directly.

For the idealized situation in which the operating and equilibrium lines are straight, the overall column efficiency and the Murphree plate efficiency are related by an equation derived by Lewis (1936):

$$E_0 = \frac{\log\left[1 + E_{mV}\left(\frac{mV}{L} - 1\right)\right]}{\log\left(\frac{mV}{L}\right)}$$

where

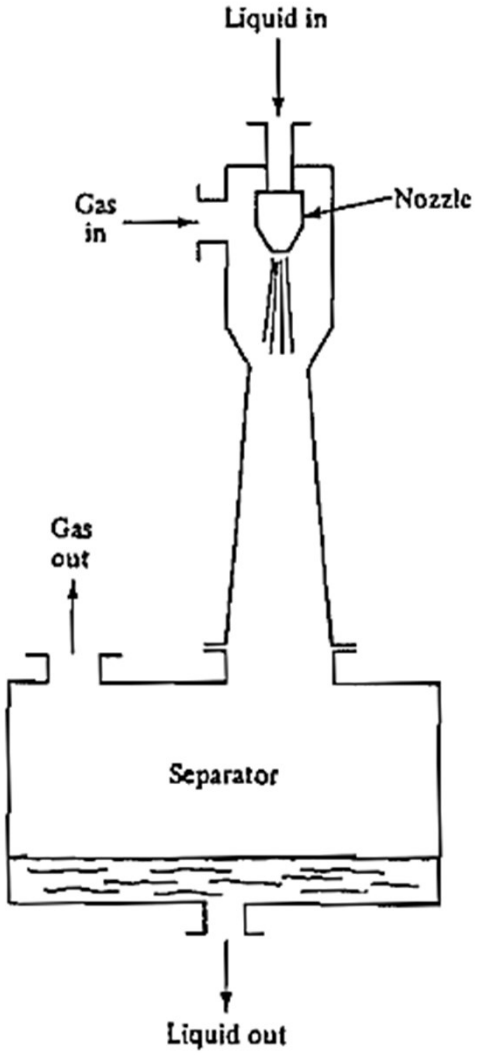
m = slope of the equilibrium line; V = molar flow rate of the vapor;

L = molar flow rate of the liquid.

Equation is not of much practical use in distillation, as the slopes of the operating and equilibrium lines will vary throughout the column. It can be used by dividing the column into sections and calculating the slopes over each section.

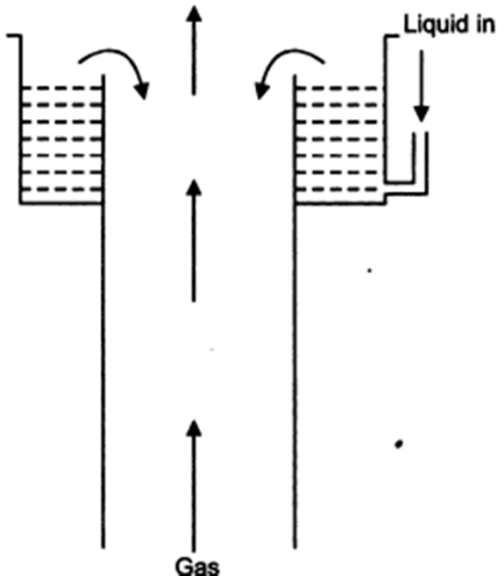
Venturi Scrubber

This is similar to ejectors. Here a stream of absorbing liquid sprayed in the convergent duct section draws the gas into the throat of a venturi. These devices will be useful when the liquid contains suspended solids, which may plug the plate/packed towers. The pressure drop is also low.



Wetted-Wall Towers

In these towers as shown in figure, a thin film of liquid flows down the inner wall of the empty vertical tube with the gas flowing cocurrently or counter-currently. Generally the flow of gas is counter-current to liquid flow. These are normally used for the measurement of mass transfer coefficient.

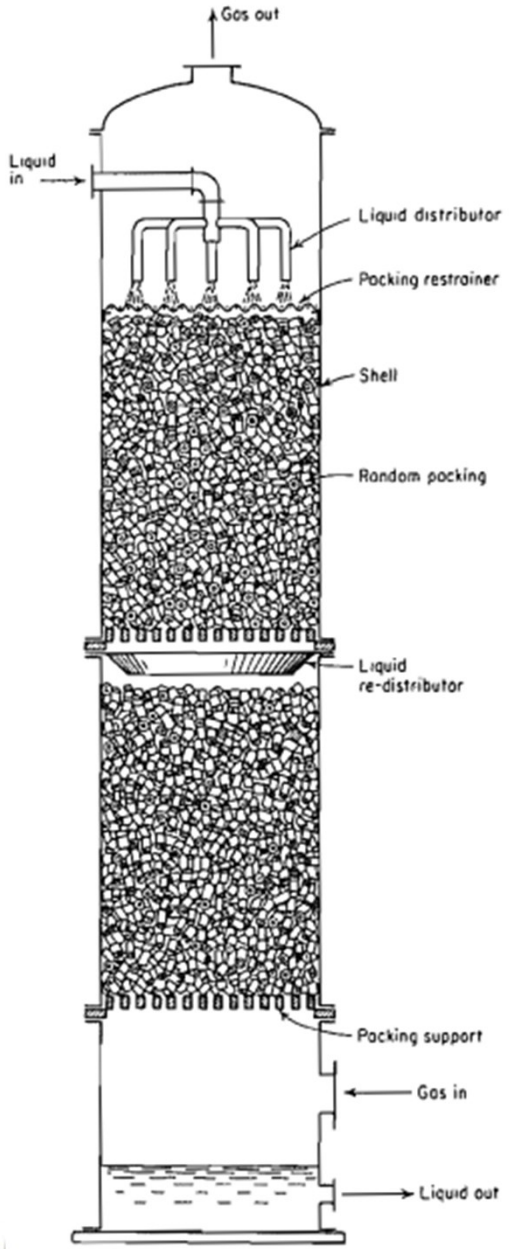


Spray Towers and Spray Chambers

In these units the liquid is sprayed into a gas stream by means of nozzle as fine droplets. The flow may be counter-current as in vertical towers with the liquid flowing downward and gas upward. It can also be cocurrent as in the case of horizontal spray chambers. Their main feature is the low pressure drop for gas. However, it suffers from the disadvantage of high pumping cost for liquids as it has to flow out through fine nozzles and also a very high entrainment of liquids as it has to flow out through fine nozzles and also a very high entrainment of liquids with gas, which necessitates the use of mist eliminators.

Packed Towers

These are towers filled with packing and are used for continuous contact of liquid and gas either cocurrently or counter-currently. The presence of packing gives enormous gas liquid contact area. The liquid is distributed over the packing and trickles down through the packed bed. A typical tower is shown in figure.



Characteristics of Packings

1. Should provide large interfacial surface between liquid and gas.
2. Should possess desirable fluid flow characteristics like low pressure drop for gas and good enough to give high value mass transfer coefficient.
3. Chemically inert to the fluid being processed.
4. Should have good structural strength to permit easy handling and installation.
5. Cost effective.

Types of Packings

There are two types of packings

- Random or dumped packing
- Regular or stacked packing

Random Packing

In this type, packing are simply allowed to fall randomly. Earlier these were materials like broken stone, gravel or coke.

However, due to their poor surface characteristics, they are now replaced by regular materials like Raschig rings, Berl saddles, Pall rings etc.

These are made of ceramics, metals or plastics. The material of construction for these depends on the nature of fluid being handled.

Ceramics are good except when alkalis and hydrofluoric acid are being used. Metals are good except in oxidising and corrosive atmospheres. Plastics deteriorate in presence of organic solvents and also at high temperatures.

Random packing offer large specific surface and hence large pressure drop. However, with larger packing sizes of the cost per unit volume is less.

Packings in the range of 25 mm to 50 mm are used for gas rates of 0.25 m^3/s and 50 mm or larger are used for gas rates of 1 m^3/s.
Some of the commonly used packings are shown in figure.

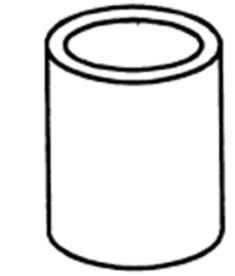
Regular Packing

In these packings there is an organised manner in which the packings are arranged in the tower.

The main feature of this is the low pressure for gas and higher fluid flow rates compared to random packings.

Stacked packings like stacked Raschig rings, wood grids and woven wire screens are some of the examples for regular packings.

Some of the commonly used regular packings are shown in figure.

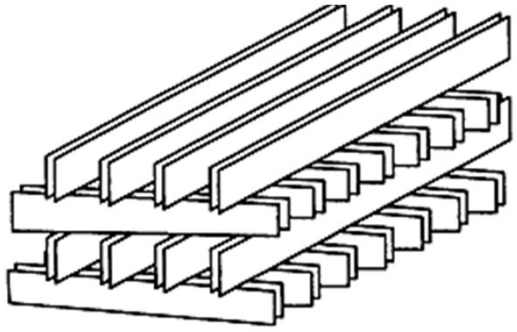


(a) Raschig rings

(b) Partition rings

(c) Berl saddle

(d) Pall ring



Wood grids

Shell

Tower shell is made of wood, metal, stoneware, acid proof brick, glass, plastic and metal lined with glass or plastic used as material of construction depending on the corrosive nature of the liquid or gas. They are generally circular in cross-section. In most of the instances it is made of metal because of their strength and ease of operation.

Packing Supports

The packing material is normally supported in the tower by means of supports. The objective of these supports is not only that they should carry the weight of packings but also ensure a proper flow of gas and liquid with minimum restriction.

They are also made of different materials like metals, ceramics and plastics.

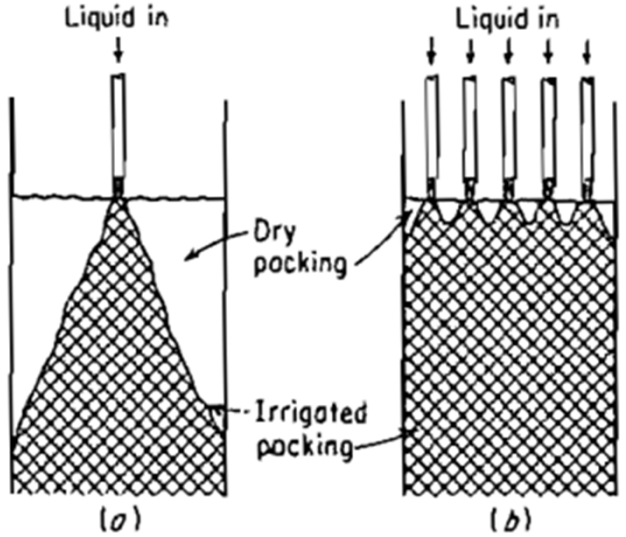
Packing Restrainer

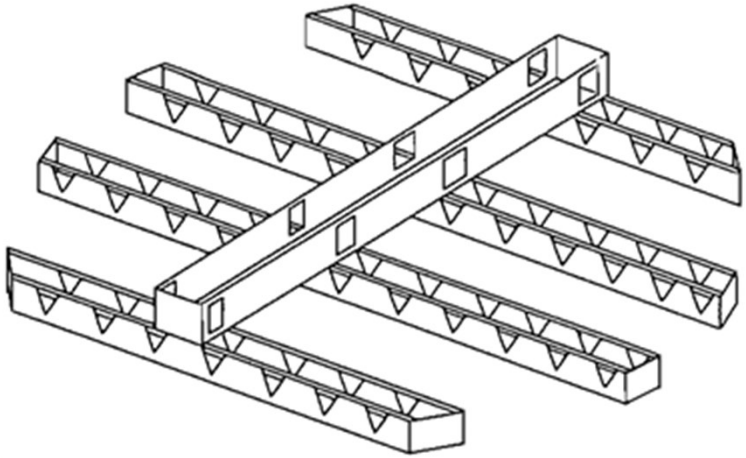
In order to prevent the lifting of packings, restrainer are provided at the top of the packing materials especially of ceramics, heavy bar plates resting freely on top of the packing may be used.

For plastics and other light weight packings, restrainer is attached to the tower.

Liquid Distributors

The liquid distribution at the top of the tower must be uniform so that the wetting of packing is uniform. A uniformly wetted packing is essential to have effectiveness in mass transfer. With non-uniform distribution of liquid one has dry packing which is ineffective for mass transfer. A ring of perforated pipe can be used in small towers. For larger diameters it is necessary to provide a number of liquid introduction points so that distribution is uniform.





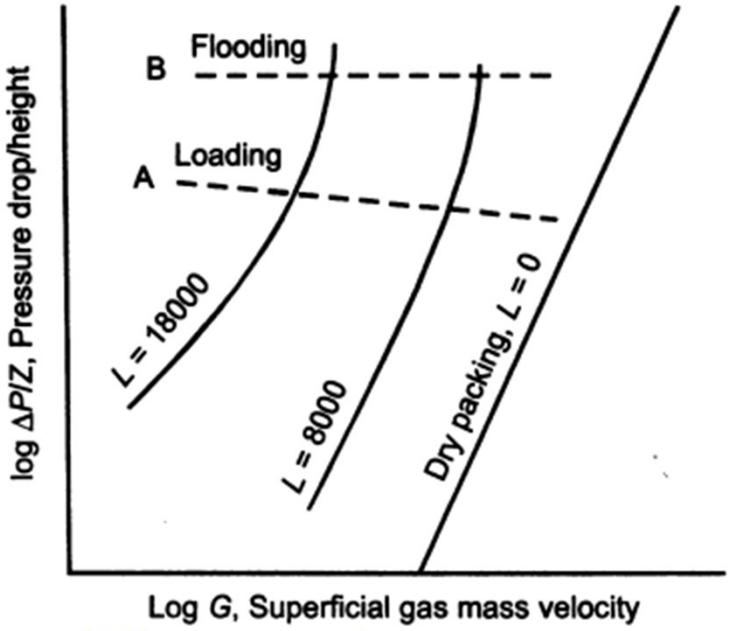
Loading

Pressure drop in a packed bed is basically due to fluid friction. As the gas flow rate is increased, the pressure drop per unit length of packing increases. Pressure drop is low when the packing is dry. With increase in liquid flow rate, pressure drop increases as it reduce the space available for gas flow.

When the packing is gradually wetted with a constant flow of liquid, initially there is a linear relationship between pressure drop and gas flow rate and is parallel to that of dry packing as shown in figure.

The linear line becomes steeper at moderate gas velocities since the gas flow retards the down flowing liquid resulting in an increase in liquid hold up.

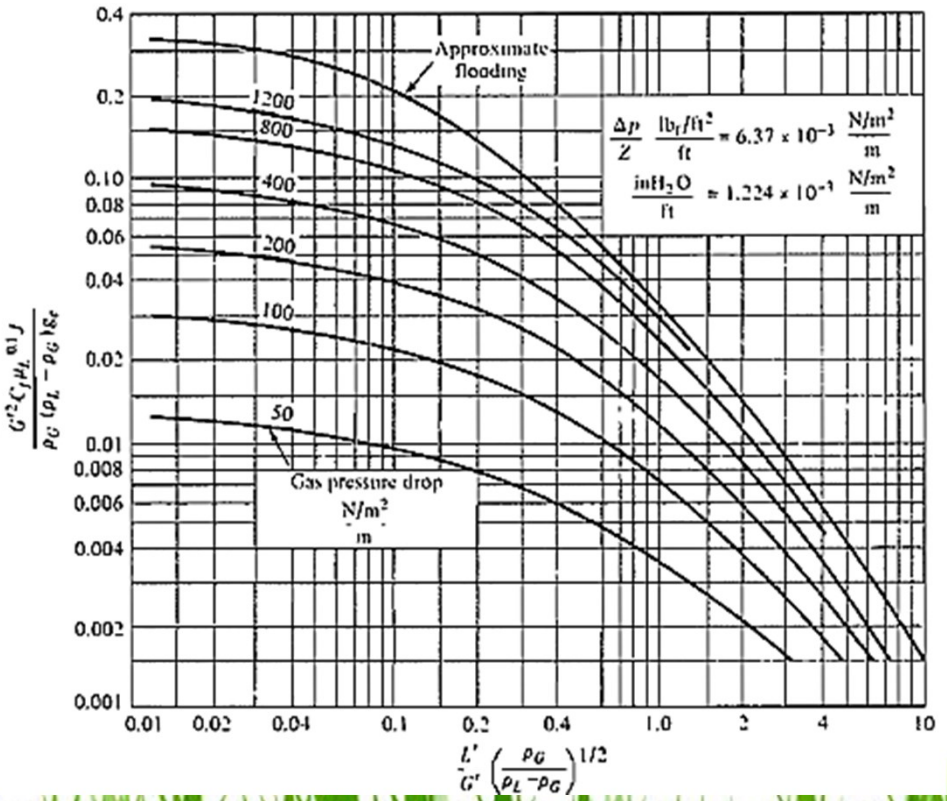
The point at which the liquid holdup start to increase, as indicated by a change in slope of the pressure drop-gas flow rate relationship is called the loading point.



Flooding

With further increase in velocity of gas (beyond loading point) the pressure drop increases rapidly and pressure drop-gas flow rate relationship becomes almost vertical. At some portions of the column, the liquid becomes the continuous phase and the flooding point is said to be reached, and the accumulation of liquid is rapid and the entire column may be filled with liquid.

Hence, while a bed is being operated, the gas velocity must be lesser than the flooding velocity and as flooding is approached, most or the entire packing surface is wetted, maximizing the gas-liquid contact area.



We must choose a velocity lower than the flooding velocity. This will lead to a larger column diameter.

The flooding velocity depends on the type and size of packing, liquid velocity and properties of liquid and gas.

Pressure Drop for Single-Phase Flow

The pressure drop suffered by a single phase fluid in flowing through a bed of packed solids such as spheres, cylinders, gravel, sand, etc., when it alone fills the voids in the bed, is reasonably well correlated by the Ergun equation

$$\frac{\Delta p}{Z}\frac{g_c\varepsilon^3 d_p\rho_g}{(1-\varepsilon)G'^2} = \frac{150(1-\varepsilon)}{\mathrm{Re}} + 1.75$$

Friction factor

Friction factor due to laminar flow

Friction factor due to turbulent flow

This is applicable equally well to flow of gases and liquids. There is a gradual transition from one type of flow to the other as a result of the diverse character of the void spaces, the two term of the equation changing their relative importance as the flow rate changes.

$$\mathrm{Re} = d_p G'/\mu$$

d_p is the effective diameter of the particles

$$d_p = \frac{6(1-\varepsilon)}{a_p}$$

a_p specific surface area

For flow of gases at G' greater than about 0.7 kg/m2 s, the first term on the right of equation is negligible. For a specific type and size of manufactured tower packing, equation can then be simplified to the empirical expression

$$\frac{\Delta p}{Z} = C_D \frac{G'^2}{\rho_G}$$

C_D is the characteristics of packing listed in table.

Liquid Holdup

Holdup refers to the liquid retained in the tower as films wetting the packing and as pools caught in the gaps between packing particles. It is found that the total holdup ϕ_{Lt} is made up of two parts,

$$\phi_{Lt} = \phi_{L0} + \phi_{Ls}$$

Where ϕ_{Ls} is the static and ϕ_{L0} the operating, or moving, holdup, each expressed as volume liquid/packed volume.

The static holdup is liquid retained as pool in protected interstices in the packing, largely stagnant, and only slowly replaced by fresh liquid.

Mass Transfer

Shulman and coworkers established the nature of the area free mass transfer coefficient kG, for Raschig rings and berl saddles, by passing gases through beds filled with packings made with naphthalene.

The gas coefficient is given by

$$\frac{F_G\,\mathrm{Sc}_G^{2/3}}{G} = \frac{k_G p_{\mathrm{B.M}}\,\mathrm{Sc}_G^{2/3}}{G} = 1.195\left[\frac{d_s G'}{\mu_G(1-\varepsilon_{Lo})}\right]^{-0.36}$$

Where ε_{Lo}, the operating void space in the packing, is given by

$$\varepsilon_{Lo} = \varepsilon - \phi_{Lt}$$

d_s is the diameter of a sphere of the same surface as a single packing particle(not the same as dp). The fluid properties should be evaluated at average conditions between interface and bulk gas.

The liquid coefficient is given by

$$\frac{k_L d_s}{D_L} = 25.1\left(\frac{d_s L'}{\mu_L}\right)^{0.45}\mathrm{Sc}_L^{0.5}$$