GAS Absorption

Gas absorption is an operation in which a gas mixture is contacted with a liquid for the purposes of preferentially dissolving one or more components of the gas and to provide a solution of them in the liquid. For example, the gas from by-product coke oven is washed with water to remove ammonia and again with an oil to remove benzene and toluene vapors.

Equilibrium Solubility of gases in liquids :-

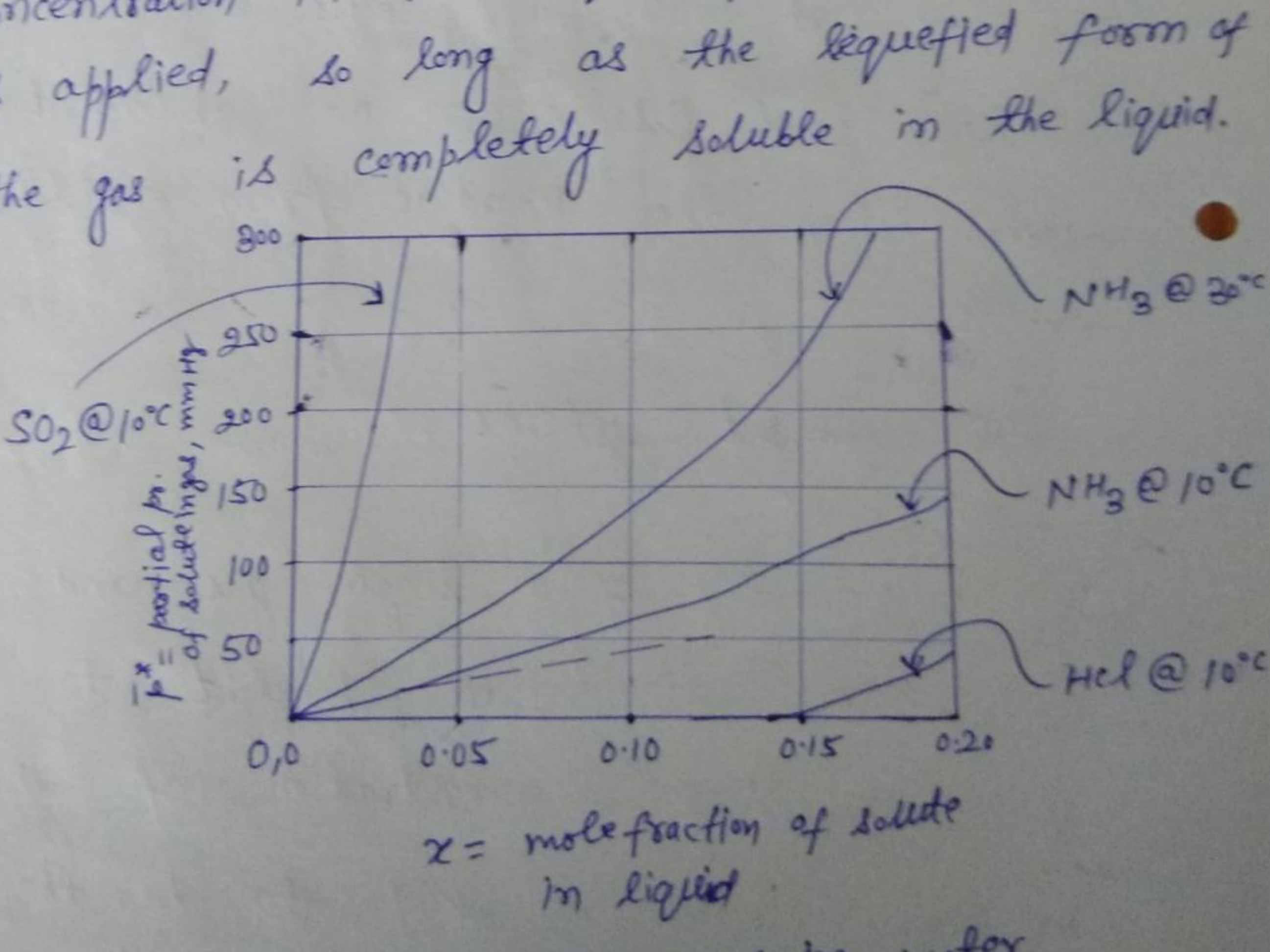
The rate at which a gaseous constituent of a mixture will dissolve in an absorbent liquid depends upon the departure from equilibrium which exists.

Two component systems :-

If a quantity of a single gas and a relatively nonvolatile liquid are brought to equilibrium, the resulting concentration of dissolved gas in the liquid is said to be the gas solubility at the prevailing temperature and pressure.

At fixed temperature, the solubility concentration will increase with pressure.

Different gases and liquids yield separate solubility curves, which must be ordinarily determined experimentally for each system. If the equilibrium pressure of a gas at a given liquid concentration is high, the gas is said to be relatively insoluble in the liquid, while if it is low, the solubility is said to be high. But these are relative matters only, for it is possible to produce any ultimate gas concentration in the liquid if sufficient pressure is applied, so long as the liquefied form of the gas is completely soluble in the liquid.



Solubilities of gases in water.

From the equilibrium solubility curves, it is obvious that HCl is more soluble than NH_3 in water, as at specified concentration (x), the equilibrium partial pressure in case of HCl is lower than in the case of NH_3. In most of the cases, the solubility of a gas decreases with increasing temperature as it is clear from the curves for NH_3 at two different temperatures 10°C and 30°C.

Multicomponent systems :-

If a mixture of gases is brought into contact with a liquid, under certain conditions the equilibrium solubilities of each gas will be independent of the others, provided, however, that the equilibrium is described in terms of the partial pressures in the gas mixture. If all but one of the components are substantially insoluble, their concentrations in the liquid will be so small that they cannot influence the solubility of the relatively soluble component. For example, the solubility of NH_3

in water when NH_3 is diluted with air, since air is so insoluble in water, the curves represented in above figure for NH_3 solubility will also describe for the present NH_3 + air - water system, provided that the ordinate of the plot is considered as the partial pressure of NH_3 in the gas mixture.

If several components are soluble, the generalization will be applicable only if the solute gases are indifferent to the nature of the liquid, which will be the case only for ideal solutions. For example a mixture of propane and butane gases will dissolve in a nonvolatile paraffin oil independently since the solutions that result are substantially ideal. On the other hand, the solubility of NH_3 in water can be expected to be influenced by the presence of methylamine, since the resulting solutions of these gases are not ideal. The solubility of a gas will also be influenced by the presence of a nonvolatile solute in the liquid, such as a salt in water solution, when such solutions are nonideal.

Ideal liquid solutions :-

When the liquid phase can be considered ideal, we can compute the equilibrium partial pressure of a gas from the solution without resort to experimental determination.

There are four significant characteristics of Ideal solutions, all interrelated :-

(a) The average intermolecular forces of attraction and repulsion in the solution are unchanged on mixing the constituents.

(b) The volume of the solution varies linearly with composition.

(c) There is neither absorption nor evolution of heat in mixing the constituents. For gases dissolving in liquids, however, this criterion should not include the heat of condensation of the gas to the liquid state.

(d) The total vapor pressure of the solution varies linearly with composition expressed as mole fractions.

In reality there are no ideal solutions, and actual mixtures only approach ideality as a limit. Adjacent or nearly adjacent members of a homologous series of organic compounds form nearly ideal solutions. For example, solutions of benzene in toluene, of ethyl and propyl alcohols, or of the paraffin hydrocarbon gases in paraffin oil can ordinarily be considered as ideal solutions.

when the gas mixture in equilibrium with an ideal liquid solution also follows the ideal-gas law, the partial pressure of a solute gas A can be computed using Raoult's law.

$$\bar{p}_A = P_A x_A$$

where, $\bar{p}_A$ is partial pressure, P_A is vapor pressure at same temperature, and x_A is mole fraction of solute in liquid phase.

The nature of the solvent liquid does not enter into consideration except insofar as it establishes the ideality of the solution, and it follows that the solubility of a particular gas in ideal solution in any solvent is always the same.

Nonideal liquid solutions :-

For liquid solutions which are not ideal, Raoults law cannot be applied for such solutions. For such solutions Henry's law can be applied if the solute concentration is very low (for dilute solutions).

$$p_A = m x_A \quad \text{(Henry's law)}$$

where, m is a constant.

Choice of solvent for absorption :-

Following properties are considered for the choice of an appropriate solvent :-

(a) Gas solubility :- The gas solubility should be high, thus increasing the rate of absorption and decreasing the quantity of solvent required.

that the solubility of a particular gas in ideal solution in any solvent is always the same.

Nonideal liquid solutions :-

For liquid solutions which are not ideal, Raoults law cannot be applied for such solutions. For such solutions Henry's law can be applied if the solute concentration is very low (for dilute solutions).

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where, m is a constant.

Choice of solvent for absorption :-

Following properties are considered for the choice of an appropriate solvent :-

(a) Gas solubility :- The gas solubility should be high, thus increasing the rate of absorption and decreasing the quantity of solvent required.

(b) Volatility :- The solvent should have a low vapor pressure since the gas leaving an absorption operation is ordinarily saturated with the solvent and much may thereby be lost.

(c) Corrosiveness :- The materials of construction required for the equipment should not be unusual or expensive.

(d) Cost : The solvent should be inexpensive, so that losses are not costly, and should be readily available.

(e) viscosity :- Low viscosity is preferred for reasons of rapid absorption rates, improved flooding characteristics in absorption towers, low pressure drops on pumping, and good heat-transfer characteristics.

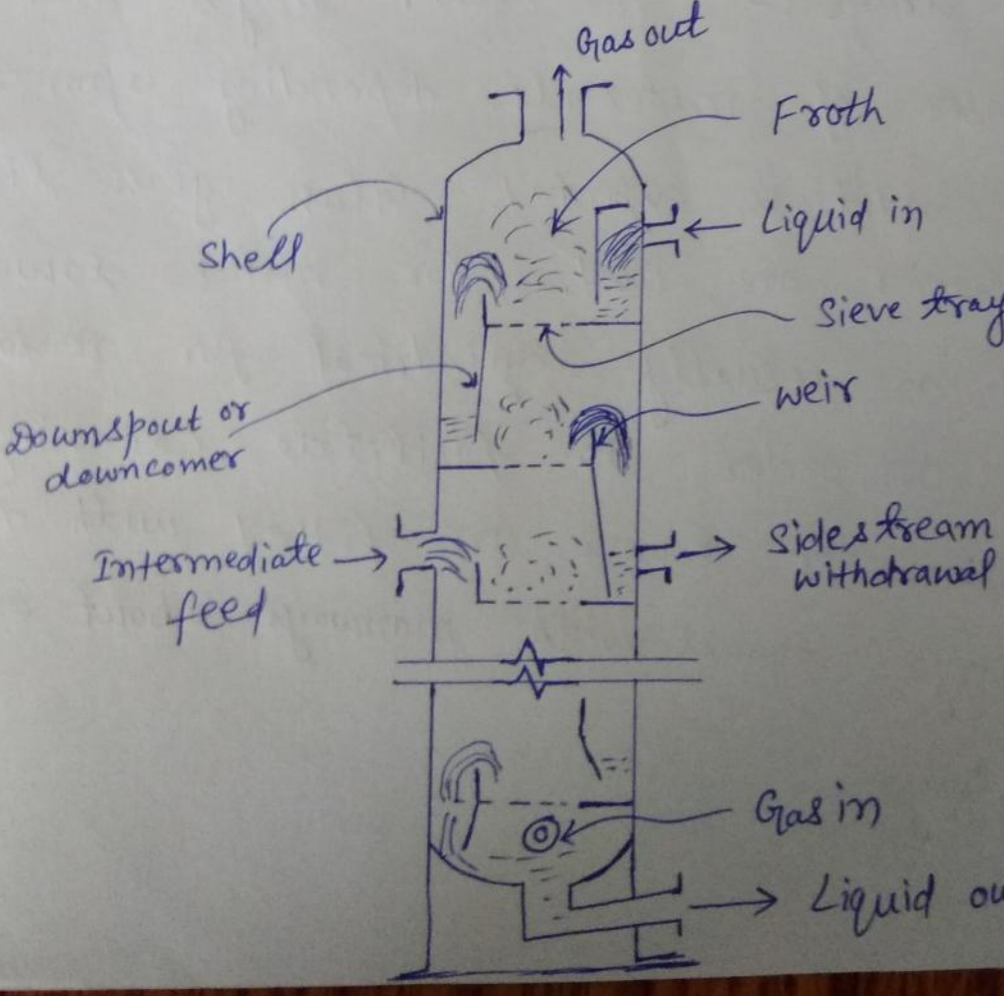
(f) Miscellaneous :- The solvent if possible should be nontoxic, nonflammable, and chemically stable and should have a low freezing point.

Equipment for gas- liquid contact :-

For better contact of gas and liquid various equipments are used such as mechanically agitated vessels, multistage towers, tray towers and packed columns. For the gas absorption tray towers and packed columns are frequently used.

Tray Towers :-

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



The liquid enters at the top and flows downward by gravity. on the way, it flows across each tray and through a downcomer to the tray below. The gas passes upward through openings of one sort or another in the tray, then bubbled through the liquid to form a froth, disengages from the froth, and passes on the next tray above. Each tray of the tower is a stage, since on the tray the fluids are brought into intimat contact, interphase diffusion occurs, and fluids are separated.

General characteristics of tray towers:-

Shell and trays :- The tower may be made of any number of materials, depending upon the corrosion conditions expected. Glass, glass-lined metal, plastics are used. For metal towers, the shell are usually cylindrical for reasons of cost. In order to facilitate cleaning, small diameter towers are fitted with hand holes, large towers with manways about every tenth tray

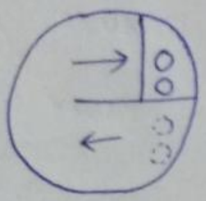
The trays are usually made of metal sheets, of special alloys if necessary. the thickness governed by the anticipated corrosion rate. The trays must be stiffened and supported and must be fastened to the shell to prevent movement owing to surges of gas, with allowance for thermal expansion. Large trays must be fitted with manways so that a person can climb from one tray to another during repair and cleaning.

Tray spacing :- For special cases where tower height is an important consideration, specing of 0.15 m have been used. For all except the smallest tower diameters, 0.50 m would seem to be a more workable minimum from the point of view of cleaning the trays. The tray specing generally varies in the range of 0.15 m to 0.90 m, depending on the tower diameters.

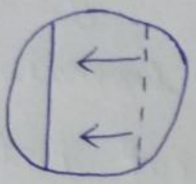
Downspouts :- The liquid is led from one tray to the next by means of downspouts or downcomers. These maybe circular pipes or preferably portions of tower cross section set aside for liquid flow by vertical plates. The downspout must be brought close enough to the tray below to seal into the liquid on that tray, thus preventing gas from rising up the downspout to short-circuit the tray above.

Weirs:- The depth of liquid on the tray required for gas contacting is maintained by an overflow weir. Straight weirs are most common. In order to ensure reasonably uniform distribution of liquid flow on a single-pass tray, a weir length of from 60 to 80% of the tower diameter is used.

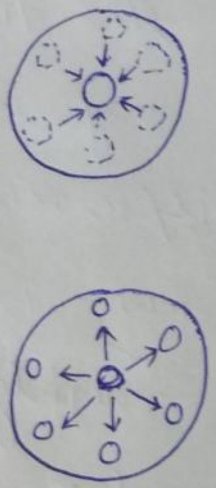
Liquid flow:- Several of the schemes commonly used for directing the liquid flow on trays are shown in below figure. Reverse flow can be used for relatively small towers, but by far the most common arrangement is the single pass cross-flow (fig- b). For large-diameter towers radial or split flow can be used, although every attempt is usually made to use the cross-flow tray because of its lower cost.



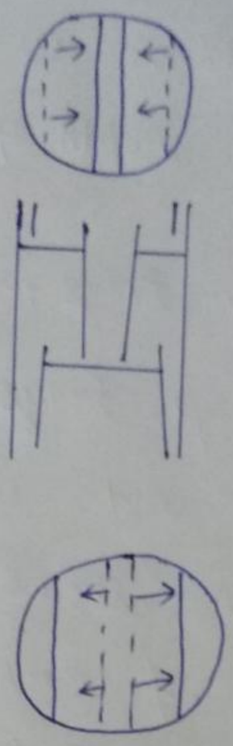
(a) Reverse flow



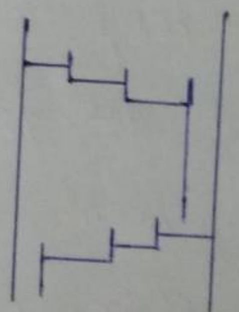
(b) crossflow



(c) Radial flow



(d) split flow



(e) cascade trays

Tray arrangments. Arrows show direction of liquid flow.

Liquid depth :- liquid depth should not ordinarily be less than 50 mm (2 in), to ensure good froth formation. Depth of 100 mm is a more common.

Bubble-cap trays :-

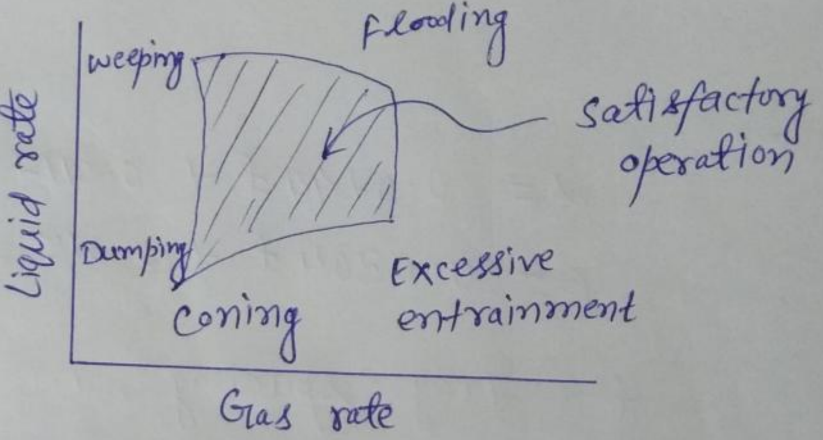
Bubble-Cap trays risers lead the gas through the tray and underneath caps surmounting the risers. A series of slots is cut into the rim or skirt of each cap, and the gas passes through them to contact the liquid which flows past the caps. Bubble caps offer the distinct advantage of being able to handle very wide ranges of liquid and gas flow rates satisfactorily. They have now been abandoned for new installations because of their cost, which is roughly double that for sieve, counterflow, and valve trays.

Sieve trays :-

These trays have been known almost as long as bubble cap trays, but they fell out of favor during the first half of nineteenth century. Their low cost, however, has now made them the most important of tray devices.

Tower diameter :-

The tower diameter and consequently its cross-sectional area must be sufficiently large to handle the gas and liquid rates within the region of satisfactory operation of below figure.



For a given type of tray at flooding, the superficial velocity of the gas V_F (volumetric rate of gas flow Q per net cross section for flow A_n) is related to fluid densities by

$$V_F = C_F \left(\frac{\rho_L - \rho_G}{\rho_G} \right)^{1/2}$$

The net cross section A_n is the tower cross section A_t minus the area taken up by the downcomer (A_d) in the case of a cross-flow tray.

C_F is an empirical constant, the value of which depends on the tray design.

For sieve trays :-

$$C_F = \left[\alpha \log \frac{1}{(L'/G')(\rho_G/\rho_L)^{0.5}} + \beta \right] \left(\frac{\sigma}{0.020} \right)^{0.2}$$

where,

$$\left.\begin{aligned} \alpha &= 0.0744\,t + 0.01173 \\ \beta &= 0.0304\,t + 0.015 \\ t &= \text{Tray spacing, in m} \end{aligned}\right\} \quad \text{— (A)}$$

(A) is valid for $A_o/A_a > 0.1$ and $\frac{L'}{G'}\left(\frac{\rho_G}{\rho_L}\right)^{0.5}$ in between $0.01 - 0.1$.

A_o = hole area, A_a = active area.

$$\frac{A_o}{A_a} = \frac{\text{hole area}}{\text{active area}} = 0.907 \left(\frac{d_o}{p'} \right)^2$$

p' : ~~Tr~~ Triangular pitch (2.5-5 times of hole diameter)

when, $\frac{A_o}{A_a} < 0.1$, multiply α and β by $5\frac{A_o}{A_a} + 0.5$

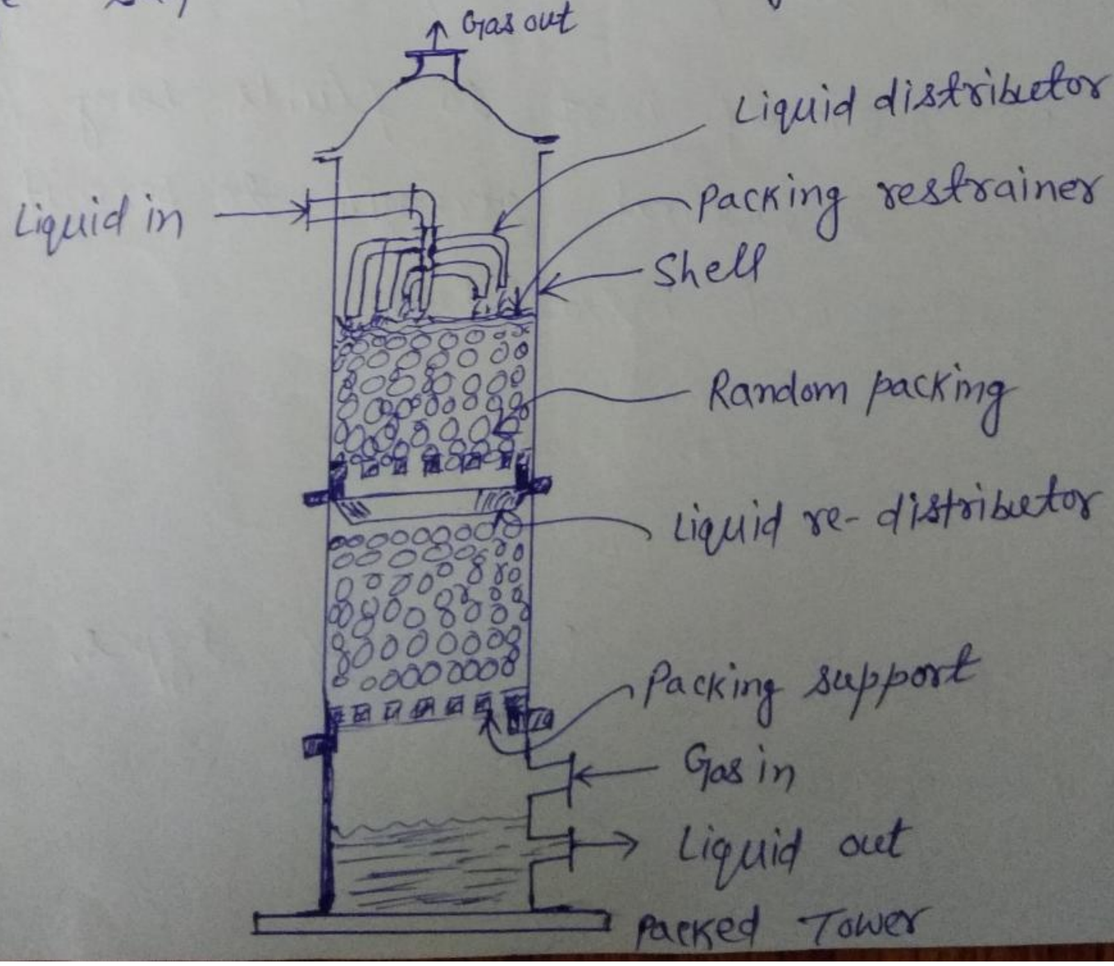
σ = Surface tension

L' & G' = Liquid and gas flow rates, respectively.

$$\frac{L'}{G'}\left(\frac{\rho_G}{\rho_L}\right)^{0.5} = 0.1, \quad \text{if} \quad \frac{L'}{G'}\left(\frac{\rho_G}{\rho_L}\right)^{0.5} < 0.1$$

Packed Towers:-

Packed towers, used for continuous contact of liquid and gas in both counter current and cocurrent flow, are vertical columns which have been filled with packing or devices of large surface, as shown in below figure. The liquid is distributed over, and trickles down through, the packed bed, exposing a large surface to contact the gas.



Packed Tower

Packing :-

The tower packing, or fill, should offer the following characteristics :-

(a) provide for large interfacial surface between liquid and gas.

(b) possess desirable fluid-flow characteristics. This ordinarily means that the fraction void volume ϵ in the packed bed should be large. The packing must permit passage of large volumes of fluid through small tower cross sections without loading or flooding and with the low pressure drop for the gas.

(c) Be chemically inert to fluids being processed.

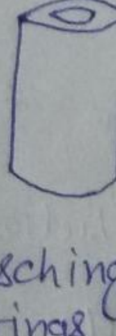
(d) Have structural strength to permit easy handling and installation.

(e) Represent low cost.

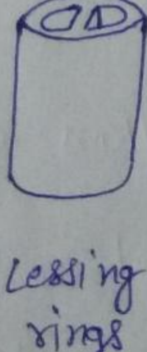
Packing are of two major types, random and regular.

Random packings :-

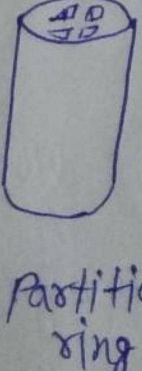
Random packings are simply dumped into the tower during the installation and allowed to fall at random. Random packings most frequently used at present are manufactured. Raschig rings are hollow cylinders, as shown in figure ranging from 6 to 100 mm (1/4 to 4 in) or more. They may be made of chemical stoneware or porcelain, which are useful in contact with most liquids except alkalies and hytrofluoric acid; of carbon, which is useful except strongly oxidizing atmospheres, of metals; or of plastics. Lessing rings and others with internal partitions are less frequently used. The saddle-shaped packings, Berl and intalox saddles, and variants of them, are available in sizes from 6 to 75 mm (1/4 to 3 in), made of chemical stoneware or plastic. Pall rings, also known as flexirings, cas-cade rings, and as a variant, Hy-pak are available in metal and plastic.



Rasching rings



Lessing rings



Partition ring

Berl seddle

Regular Packings:-

These are of great variety. The counterflow trays are considered a form of regular packing. The regular packings offer the advantages of low pressure drop for the gas and greater possible fluid flow rates, usually at the expense of more costly installation than random packing. Knitted or otherwise woven wire screen, rolled as a fabric into cylinders (Neo-Kloss) or other metal gauzelike arrangements provide a large interfacial surface of contacted liquid and gas and a very low-gas pressure drop, especially useful for vacuum distillation.

Tower shells:-

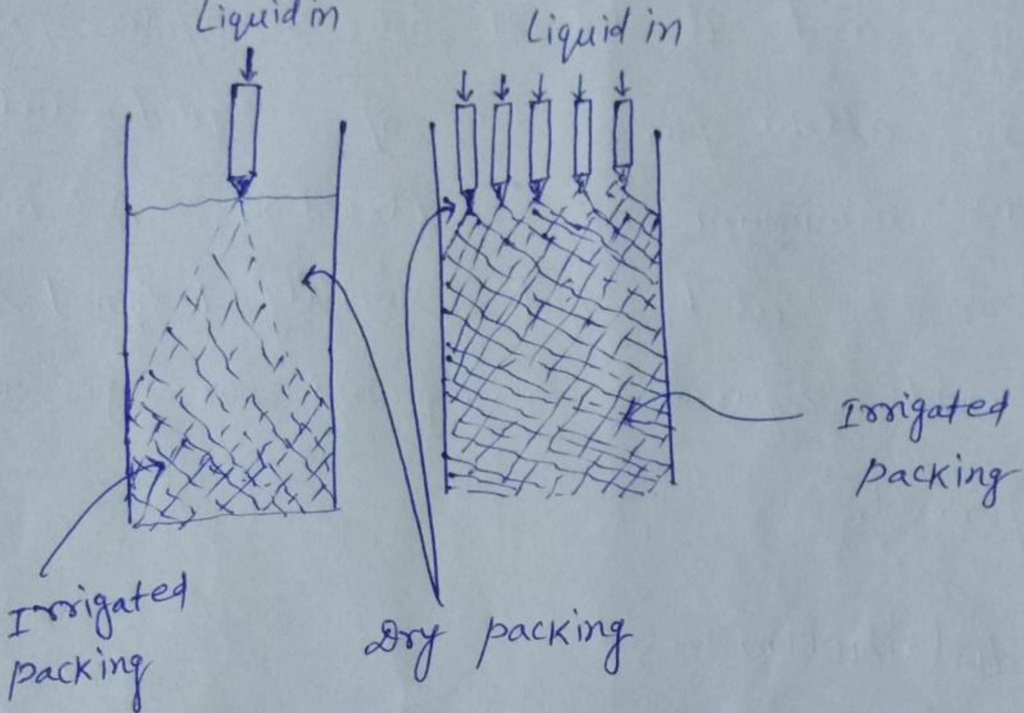
These may be of wood, metal, chemical stoneware, acidproof brick, glass, plastic, plastic- or glass-lined metal, or other material depending upon the corrosion conditions. For ease of construction and strength they are usually circular in cross section.

Packing supports:-

An open space at the bottom of the tower is necessary for ensuring good distribution of the gas into the packing. Consequently the packing must be supported above the open space. The support must, of course, be sufficiently strong to carry the weight of a reasonable height of packing, and it must have ample free area to allow for flow of liquid and gas with a minimum of restriction. A bar grid can be used, but specially designed supports which provide separate passageways for gas and liquid are preferred.

Liquid distribution :-

The importance of adequate initial distribution of liquid at the top of the packing is indicated in below figure. Dry packing is of course completely ineffective for mass transfer, and various devices are used for liquid distribution. Spray nozzles generally result in too much entrainment of liquid in the gas to be useful. For large diameters, a distributor of the type "weir- trough liquid distributor" can be used. It is generally considered necessary to provide at least five points of introduction of liquid for each 0.1 m^2 (1 ft^2) of tower cross section for large towers ($d \geq 1.2$ m) and a greater number for smaller diameters.



Random packing size and liquid redistribution:-

In the case of random packings, the packing density, i.e. the number of packing pieces per unit volume, is ordinarily less in the immediate vicinity of the tower walls, and this leads to a tendency of the liquid to segregate towards the walls and the gas to flow in the center of the tower. It is recommended that, if possible, the ratio $dp/T = 1:15$ may provide better distribution and thus channeling is less pronounced. It is customary to provide for redistribution of the liquid at intervals varying from 3 to 10 times the tower diameter, but at least every 6 or 7 m.

Packing restrainers :-

These are necessary when gas velocities are high, and they are generally desirable to guard against lifting of packing during a sudden gas surge. Heavy screens or bars may be used. For heavy ceramic packing, heavy bar plates resting freely on the top of the packing may be used. For plastics and other leightweight packings, the restrainer is attached to the tower shell.

Entrainment Eliminators :-

Especially at high gas velocities, the gas leaving the top of the packing may carry off droplets of liquid as a mist. This can be removed by mist eliminators, through which the gas must pass, installed above the liquid inlet. A layer of mesh especially Knitted with 98 to 99 percent voids, roughly 100 mm thick, will collect virtually all mist particles. other types of eliminators include cyclones and venetian- blind arrangements. A meter of dry random packing is very effective.

Difficulties in tower operation :-

Under normal operating conditions, an average liquid depth is maintained on a tray. High pressure drop may lead directly to a condition of flooding. With a large pressure difference in the space between trays, the level of liquid leaving a tray at relatively low pressure and entering one of high pressure must necessarily assume an elevated position in the downspouts. As the pressure difference is increased due to increased rate of flow of either gas or liquid, the level in the downspout will rise further to permit the liquid to enter the lower tray. Ultimately the liquid level may reach that on the tray above. Further increase in either flow rate then aggravates the condition rapidly, and the liquid will fill the entire space between the trays. The tower is then flooded, the tray efficiency falls to a low value, the flow of gas is erratic, and liquid may be forced out of the exit pipe at the top of the tower.

Priming:- For liquid-gas combinations which tend to foam excessively, high gas velocities may lead to a condition of priming, which is also an inoperative situation. Here the foam persists throughout the space between trays, and a great deal of liquid is carried by the gas from one tray to the tray above. This is an exaggerated condition of entrainment. The liquid so carried recirculates between trays, and the added liquid-handling load increases the gas pressure drop sufficiently to lead to flooding.

We can summerize these opposing tendencies as follows. Great depths of liquid on the trays lead to high tray efficiencies through long contact time but also to high pressure drop per tray. High gas velocities, within limits, provide good vapor-liquid contact through excellence of dispersion but lead to excessive entrainment and high pressure drop.

Coning :- If liquid rate are too low, the gas rising through the openings of the tray may push the liquid away coning and contact of the gas and liquid is poor.

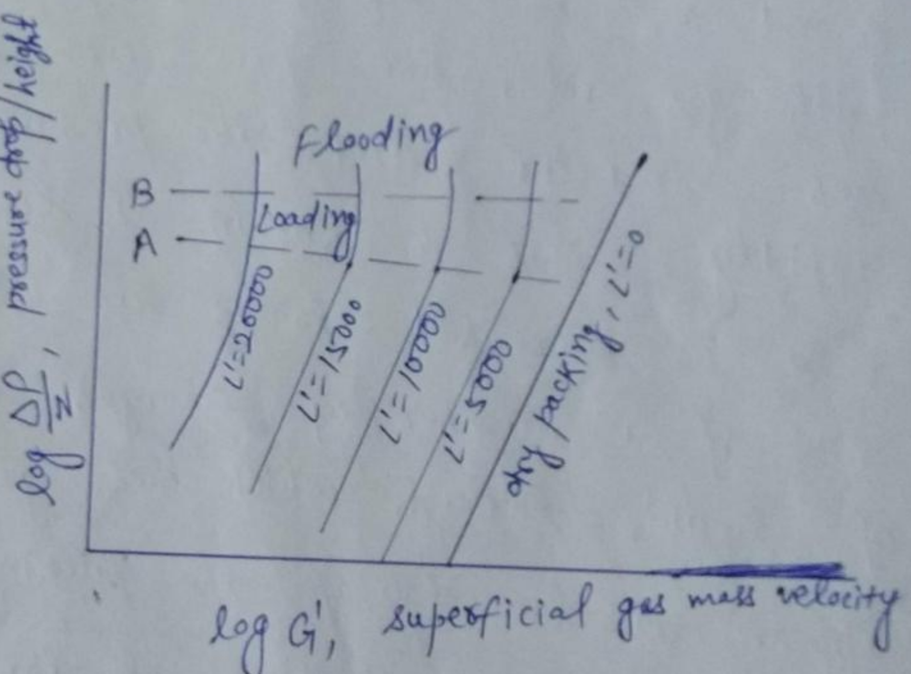
weeping :- If the gas rate is too low, much of liquid may rain down through the openings of the trays, thus failing to obtain the benefit of complete flow over the tray, and at very low gas rates, non of liquid reaches the downspouts, and leads to a condition called dumping.

Countercurrent flow of liquid and gas through packing :-

For most random packings, the pressure drop suffered by the gas is influenced by the gas and liquid flow rates as shown in below figure The slope of the line for dry packing is usually in the range 1·8 to 2, indicating turbulent flow for most practical gas velocities.

At a fixed gas velocity, the gas pressure drop increases with increased liquid rate, principally because of the reduced free cross section available for flow of gas resulting from the presence of the liquid. In the region of figure below A, the liquid holdup, i.e. the quantity of liquid contained in the packed bed, is reasonably constant with changing gas velocity, although it increases with liquid rate. In the region between A and B, the liquid holdup increases rapidly with gas rate, the free area for gas flow becomes smaller, and the pressure drop rises more rapidly. This is known as loading. As the gas rate is increased to B at fixed liquid rate, one of a number of changes occurs : (a) a layer of liquid, through which the gas bubbles, may appear at the top of the packing, (b) liquid may fill the tower, starting at the bottom or at any intermediate restriction such as a packing support, so that there is a change from gas-continuous liquid-dispersed to liquid-continuous gas-dispersed (inversion), or, (c) slugs of foam may rise rapidly upward through the packing. At the same time, entrainment of liquid by the effluent gas increases rapidly, and the tower is flooded.

The gas pressure drop then increases very rapidly. The gas pressure drop varies gradually in the region A to B of the figure. The initial loading and flooding are frequently determined by the change in slope of the pressure-drop curves rather than through any visible effect. It is not practical to operate a tower in a flooded condition, most towers operate just below, or in the lower part of the loading region.



Flooding and loading :-

Flooding conditions in random packings depend upon the method of packing (wet or dry) and settling of the packing. The upper curve of below figure correlates the flooding data for most random packings reasonably well. The limit of loading cannot readily be correlated.

Typically, absorbers and strippers are designed for gas pressure drops of 200 to 400 N/m² per meter of packed depth, atmospheric pressure fractionators from 400 to 600 (N/m²)/m, and vacuum stills for 8 to 40 (N/m²)/m. values of C_f which characterize the packings are available in literature and changes with changing manufacturing procedures. flooding velocities for regular or stacked packings will generally be considerably greater than for random packing.

Tray towers versus packed towers :–

The following may be useful in considering a choice between the two major types of towers.

(a) Gas pressure drop :- Packed towers will ordinarily require a smaller pressure drop. This is especially important for vacuum distillation.

(b) Liquid holdup :- Packed towers will provide a substantially smaller liquid holdup. This is important where liquid deterioration occurs with high temperatures and short holding times are essential. It is also important in obtaining sharp separations in batch distillation.

(c) Liquid-gas ratio :- very low values of this ratio are best handled in tray towers. High values are best handled in packed towers.

(d) Liquid cooling:- Cooling coils are more readily built into tray towers, and liquid can more readily be removed from trays, to be passed through coolers and returned, than from packed towers.

(e) Side streams:- These are more readily removed from tray towers.

(f) Foaming systems:- packed towers operate with less bubbling of gas through the liquid and are the more suitable.

(g) Corrosion:- Packed towers ~~of~~ for difficult corrosion problems are likely to be less costly.

(h) Solids present:- Neither type of tower is very satisfactory. Agitated vessels and venturi scrubbers are best but provide only a single stage. If multistage countercurrent action is required, it is best to remove the solids first. Dust in the gas can be removed by a venturi scrubber at the bottom of a tower. Liquids can be filtered or otherwise clarified before entering a tower.

(i) Cleaning:- Frequent cleaning is easier with tray towers.

(j) Large temperature fluctuations:- Fragile packings & (ceramic, graphite) tend to be crushed. Trays or metal packings are satisfactory.

(k) Floor loading:- Plastic packed towers are lighter in weight than tray towers, which in turn are lighter than ceramic or metal packed towers. In any event, floor loading should be designed for accidental complete filling of the tower with liquid.

(l) Cost:- If there is no overriding consideration, cost is the major factor to be taken into account.

Absorption : one component transferred; Material balances :-

countercurrent flow :-

Below figure shows a countercurrent tower which may be either a packed or spray tower, filled with bubble cap trays or of any internal construction to bring about liquid-gas contact. The gas stream at any point in the tower consists of G total mole/(area of tower cross section) (time), made up of diffusing solute A of mole fraction y, partial pressure $\bar{p}_A$, or mole ratio Y, and nondiffusing, essentially insoluble gas G_s mole/(area)(time).

The relationship between these is

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

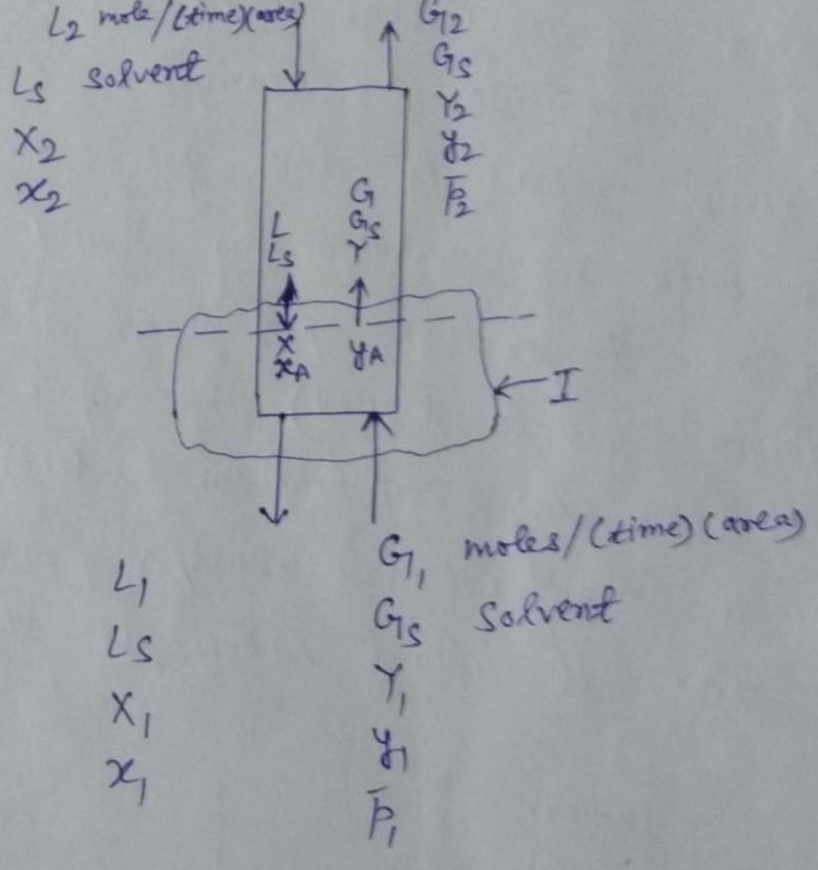
$$G_s = G(1-y) = \frac{G}{1+Y}$$

Similarly the liquid stream consists of L Total mol/(area) (time), containing x mole fraction soluble gas, or mole ratio X, and essentially nonvolatile solvent L_s mole/(area)(time).

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

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

Since the solvent gas and solvent liquid are essentially unchanged in quantity as they pass through the tower, it is convenient to express the material balance in terms of these.

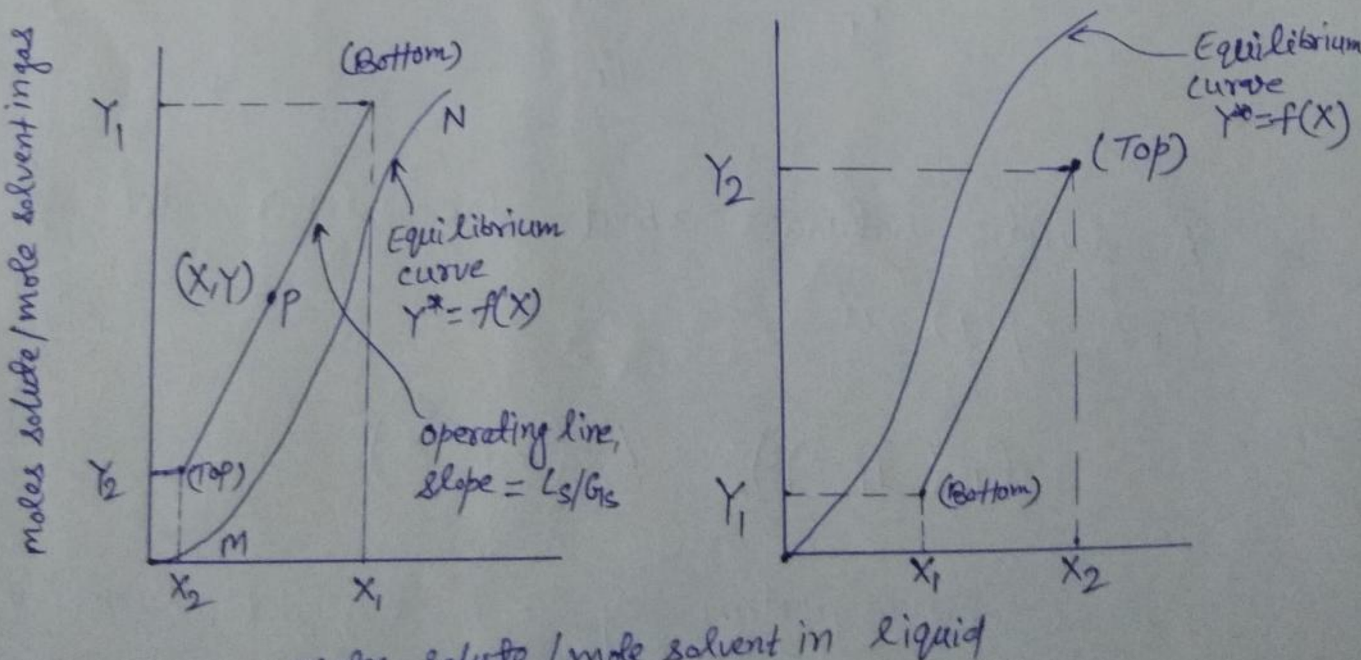


A solute balance about the lower part of the tower (envelop I) is

$$G_s(Y_1 - Y) = L_s(X_1 - X)$$

This is the equation of a straight line (the operating line) on X, Y coordinates, of slope L_s/G_s, which passes through (X_1, Y_1). Substitution of X_2 and Y_2 for X and Y shows the line to pass through (X_2, Y_2) as shown in below figure for an absorber. This line indicates the relationship between the liquid and gas concentration at any level in the tower, as at point P.

The equilibrium-solubility data for the solute gas in the solvent liquid can also be plotted in terms of these concentrations units on the same diagram, as curve MN, for example. Each point on this curve represents the gas concentration in equilibrium with the corresponding liquid at its local concentration and temperature. For an absorber (mass transfer from gas to liquid). the operating line always lies above the equilibrium-solubility curve, while for a stripper (mass transfer from liquid to gas) the line is always below as shown in figure.

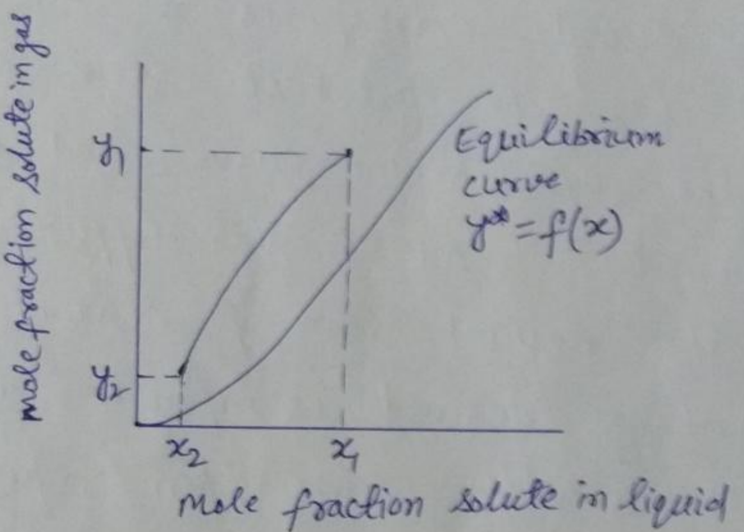


Absorber

Stripper

The operating line is straight only when plotted in terms of the mole-ratio units. In terms of mole fractions or partial pressures the line is curved, as shown in below figure for an absorber. The equation of the line is then

$$G_S\left(\frac{y_1}{1-y_1}-\frac{y}{1-y}\right)=G_S\left(\frac{\bar{p}_1}{P_t-\bar{p}_1}-\frac{\bar{p}}{P_t-\bar{p}}\right)=L_S\left(\frac{x_1}{1-x_1}-\frac{x}{1-x}\right)$$



The total pressure P_t at any point can ordinarily be considered constant throughout the tower for this purpose.

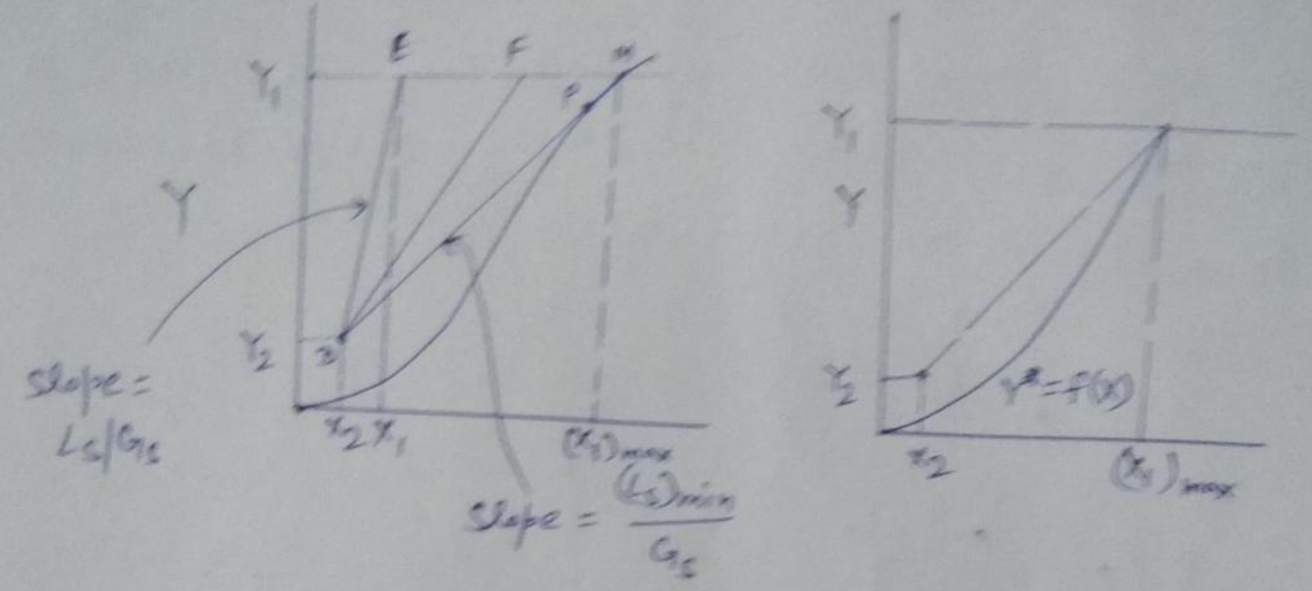
Minimum liquid-gas ratio for absorbers:—

Minimum liquid- gas ratio for absorbers:— the quantity of gas to be
In the design of absorbers, the terminal concentrations Y_1 and Y_2,
treated G or G_S, the entering liquid X_2 are
and the composition of the entering liquid X_2 are
ordinarily fixed by process requirements, but the quantity
of liquid to be used is subject to choice. The operat-
ing line as shown in below figure must pass through
point D and must end at ordinate Y_1. If such
a quantity of liquid is used to give operating

line QE, the exit liquid will have the composition x_1. If less liquid is used, the exit-liquid composition will clearly be greater, as at point F, but since the driving forces for diffusion are less, the absorption is more difficult. The time of contact between gas and liquid must then be greater, and the absorber must be correspondingly taller. The minimum liquid which can be used corresponds to the operating line QM, which has the greatest slope for any line touching the equilibrium curve and is tangent to the curve at P. At P the diffusional driving force is zero, the required time of contact for the concentration change desired is infinite, and an infinitely tall tower results. This then represents the limiting liquid-gas ratio.

The equilibrium curve is frequently concave upward, and the minimum liquid-gas ratio then corresponds to an exit-liquid concentration in equilibrium with the entering gas.

This principles also apply to strippers, where an operating line which anywhere toughes the equilibrium curve represents a maximum ratio of liquid to gas and a maximum exit-gas concentration.



Minimum liquid-gas ratio, absorption

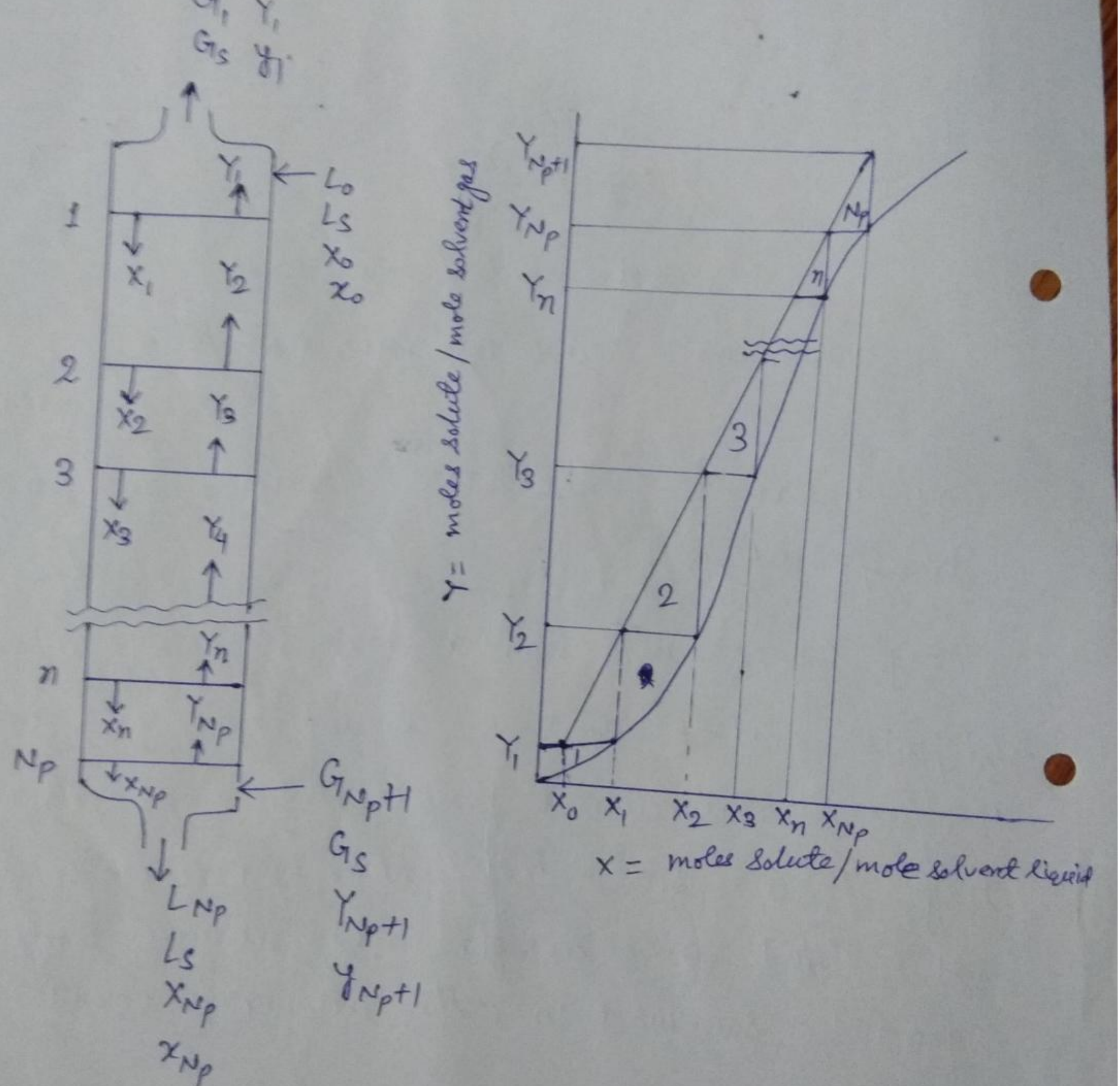
Cocurrent flow

Countercurrent multistage operation; one component transferred :-

Tray towers and similar devices bring about stepwise contact of the liquid and gas and are therefore countercurrent multistage cascades. on each tray of a sieve-tray tower, for example, the gas and liquid are brought into intimate contact and separated somewhat in the manner of countercurrent cascade.

A theoretical, or ideal tray is defined as one where the average composition of all the gas leaving the tray is in equilibrium with the average composition of all the liquid leaving the tray.

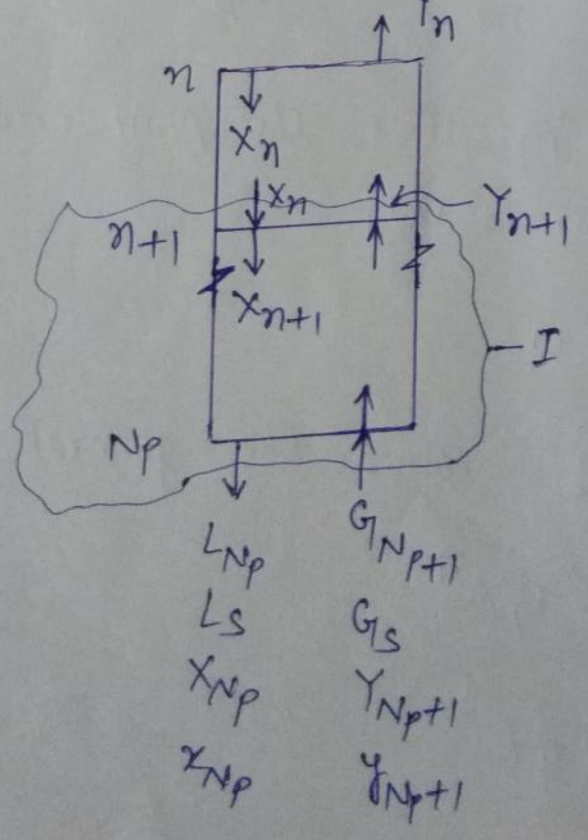
The number of ideal tray required to bring about a given change in composition of the liquid or the gas, for either absorbers or strippers, can then be determined graphically.

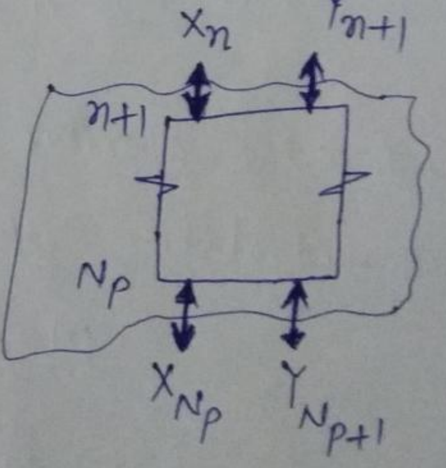


Tray absorber

Dilute gas mixtures:-

When both operating line and equilibrium curve can be considered straight, the number of ideal trays can be determined without recourse to graphical methods. This will frequently be the case for relatively dilute gas and liquid mixtures. If the quantity of gas absorbed is small, the total flow of liquid entering and leaving the absorber remains substantially constant, $L_0 \approx L_{N_p} \approx L$ total mol/(area)(time), and similarly the total flow of gas is substantially constant at G total mole/(area)(time). An operating line plotted in terms of mole fractions will then be substantially straight. Kremser equations can be applied to calculate the required ideal trays for given specifications.





Liquid - down
gas - up

From above figure, a solute balance for stages $n+1$ through N_p is

$$G_s\left(Y_{n+1} - Y_{N_p+1}\right) = L_s\left(X_n - X_{N_p}\right)$$

If A, absorption factor, $A = \frac{L_s}{mG_s}$, then

$$X_{n+1} - AX_n = \frac{Y_{N_p+1}}{m} - AX_{N_p} \qquad \text{——— (1)}$$

where, $m = \frac{Y_{n+1}}{X_{n+1}}$, slope of equilibrium curve

This is a linear first-order finite-difference equation, whose solution is handled much like that of ordinary differential equations. Thus putting it in operator form gives

$$(D - A)\,X_n = \frac{Y_{N_p+1}}{m} - AX_{N_p} \qquad \text{——— (2)}$$

where the operator D indicates the finite difference. The characteristic equation is then

$$M - A = 0$$

from which $M = A$. Hence the general solution is

$$X_n = C_1 A^n$$

with C_1 a constant.

Since the right-hand side of equation (2) is a constant, the particular solution is $X = C_2$, where C_2 is a constant.

Substituting this into the original finite-difference equation (1) provides

$$C_2 - A C_2 = \frac{Y_{N_p+1}}{m} - A X_{N_p}$$

from which

$$C_2 = \frac{Y_{N_p+1}/m - A X_{N_p}}{1-A}$$

The complete solution is therefor

$$X_n = C_1 A^n + \frac{Y_{N_p+1}/m - A X_{N_p}}{1-A}$$

To determine C_1, we set $n=0$

$$C_1 = X_0 - \frac{Y_{N_p+1}/m - A X_{N_p}}{1-A}$$

and therefore

$$X_n = \left(X_0 - \frac{Y_{N_p+1}/m - A X_{N_p}}{1-A} \right) A^n + \frac{Y_{N_p+1}/m - A X_{N_p}}{1-A}$$

This result is useful to get the concentration X_n at any stage in the cascade, Knowing the terminal concentrations. Putting $n = N_p$ and rearranging provide the very useful forms which follow:-

For stripping :-

$A \neq 1$

$$\frac{X_0 - X_{N_p}}{X_0 - Y_{N_p+1}/m} = \frac{\left(\frac{1}{A}\right)^{N_p+1} - 1/A}{\left(\frac{1}{A}\right)^{N_p+1} - 1}$$

$$N_p = \frac{\log\left[\frac{X_0 - Y_{N_p+1}/m}{X_{N_p} - Y_{N_p+1}/m}(1-A) + A\right]}{\log(1/A)}$$

$A = 1$

$$\frac{X_0 - X_{N_p}}{X_0 - Y_{N_p+1}/m} = \frac{N_p}{N_p+1}$$

$$N_p = \frac{X_0 - X_{N_p}}{X_{N_p} - Y_{N_p+1}/m}$$

For absorption :-

$A \neq 1$ $$\frac{Y_{N_p+1} - Y_1}{Y_{N_p+1} - mX_0} = \frac{A^{N_p+1} - A}{A^{N_p+1} - 1}$$

$$N_p = \frac{\log\left[\frac{Y_{N_p+1} - mX_0}{Y_1 - mX_0}\left(1 - \frac{1}{A}\right) + \frac{1}{A}\right]}{\log A}$$

$A = 1$ $$\frac{Y_{N_p+1} - Y_1}{Y_{N_p+1} - mX_0} = \frac{N_p}{N_p+1}, \quad N_p = \frac{Y_{N_p+1} - Y_1}{Y_1 - mX_0}$$

These are called the Kremser- Brown- Souders (or simply Kremser) equations. Small variations in A from one end of the tower to the other due to changing L/G as a result of absorption or stripping or to change in gas solubility with concentration or temperature can be roughly allowed for by using the geometric average of the values of A at top and bottom. For large variations, either more elaborated corrections for A, graphical computations, or tray-to-tray numerical calculations must be used.

Kremser equations for dilute solutions can be written in forms of mole fractions by replacing mole ratios by mole fractions,

Absorption :-

$$\frac{y_{N_p+1} - y_1}{y_{N_p+1} - mx_0} = \frac{A^{N_p+1} - A}{A^{N_p+1} - 1}$$

$$N_p = \frac{\log\left[\frac{y_{N_p+1} - mx_0}{y_1 - mx_0}\left(1-\frac{1}{A}\right) + \frac{1}{A}\right]}{\log A}$$

Stripping:-

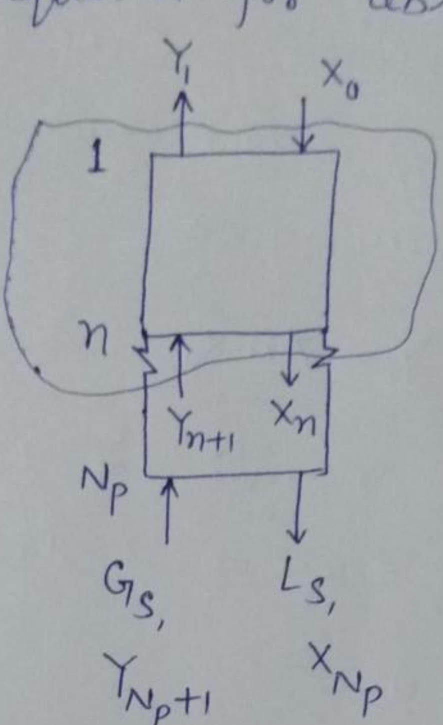
$$\frac{x_0 - x_{N_p}}{x_0 - y_{N_p+1}/m} = \frac{S^{N_p+1} - S}{S^{N_p+1} - 1}$$

$$N_p = \frac{\log\left[\frac{x_0 - y_{N_p+1}/m}{x_{N_p} - y_{N_p+1}/m}\left(1-\frac{1}{S}\right) + \frac{1}{S}\right]}{\log S}$$

where $A =$ absorption factor $= \frac{L}{mG}$

$S =$ stripping factor $= \frac{mG}{L} = \frac{1}{A}$

Kremser equation for absorption tower :-



From above figure, a solute balance for stages 1 through n is

$$G_s(-Y_1 + Y_{n+1}) = L_s(X_n - X_0)$$

$$(Y_{n+1} - Y_1) = -\frac{L_s}{G_s}(X_0 - X_n)$$

$$Y_{n+1} - Y_1 = -\frac{L_s}{mG_s}(mX_0 - mX_n)$$

$$A = \frac{L_s}{mG_s}$$

$$Y_{n+1} - Y_1 = -AmX_0 + AmX_n$$

$$\because Y_n = mX_n$$

$$Y_{n+1} - Y_1 = AY_n - AmX_0$$

$$Y_{n+1} - AY_n = Y_1 - AmX_0 \quad \text{—— (1)}$$

This is a linear first order finite-difference equation, whose solution is handled much like that of ordinary differential equations. Thus putting it in operator form gives

$$(D - A) Y_n = Y_1 - Amx_0 \quad \text{—} \quad (2)$$

where the operator D indicates the finite difference. The characteristic equation is then

$$M - A = 0$$

from which $M = A$, Hence the general solution is

$$\dot{Y}_n = C_1 A^n \qquad \text{with } C_1 \text{ a constant}$$

Since the right-hand side of equation (2) is a constant, the particular solution is $Y = C_2$, where C_2 is a constant.

Substituting this into the original finite-difference equation (1) provides

$$C_2 - AC_2 = Y_1 - Amx_0$$

$$C_2 = \frac{Y_1 - Amx_0}{(1 - A)}$$

The complete solution is therefore,

$$Y_n = C_1 A^n + \frac{Y_1 - AmX_0}{(1-A)}$$

To determine C_1 we set $n = 0$

$$C_1 = Y_0 - \frac{Y_1 - AmX_0}{1-A}$$

$\because Y_0 = mX_0$

$$C_1 = \frac{mX_0 - AmX_0 - Y_1 + AmX_0}{(1-A)}$$

$$\therefore C_1 = \frac{mX_0 - Y_1}{(1-A)}$$

$$\therefore Y_n = \left[\frac{mX_0 - Y_1}{(1-A)}\right] A^n + \frac{Y_1 - AmX_0}{(1-A)}$$

This result is useful to get the concentration Y_n at any stage in the cascade, knowing the terminal concentrations. Putting $n = (N_p + 1)$ and rearranging provide the very useful forms which is :-

$$Y_{N_p+1} = \left(\frac{mX_0 - Y_1}{1-A}\right) A^{N_p+1} + \frac{Y_1 - AmX_0}{(1-A)}$$

$$A^{N_p+1} = \frac{Y_{N_p+1} - \dfrac{Y_1 - AmX_0}{1-A}}{\dfrac{mX_0 - Y_1}{1-A}}$$

$$A^{N_p+1} = \frac{Y_{N_p+1} - AY_{N_p+1} - Y_1 + AmX_0}{mX_0 - Y_1}$$

$$A^{N_p+1} = \frac{(Y_{N_p+1} - Y_1) - A(Y_{N_p+1} - mX_0)}{mX_0 - Y_1}$$

$$A^{N_p+1} = \frac{(Y_{N_p+1} - Y_1) - A(Y_{N_p+1} - mX_0)}{mX_0 - Y_1}$$

Let

$Y_{N_p+1} - Y_1 = \gamma$

$Y_{N_p+1} - mX_0 = S$

Then

$mX_0 - Y_1 = \gamma - S$

Calculate γ/s

$$A^{N_p+1} = \frac{(Y_{N_p+1} - Y_1) - A(Y_{N_p+1} - mX_0)}{(Y_{N_p+1} - Y_1) - (Y_{N_p+1} - mX_0)}$$

$$A^{N_p+1}(Y_{N_p+1} - Y_1) - A^{N_p+1}(Y_{N_p+1} - mX_0) = (Y_{N_p+1} - Y_1) - A(Y_{N_p+1} - mX_0)$$

$$(Y_{N_p+1} - Y_1)(A^{N_p+1} - 1) = (Y_{N_p+1} - mX_0)(A^{N_p+1} - A)$$

$$\therefore \quad \frac{Y_{N_p+1} - Y_1}{Y_{N_p+1} - mX_0} = \frac{A^{N_p+1} - A}{A^{N_p+1} - 1} \quad \text{— (3)}$$

Now calculate $\frac{S}{S-r}$ and then substitute the values

Now from equation (3), rearranging provide

$$1-\left(\frac{Y_{N_p+1}-Y_1}{Y_{N_p+1}-mX_0}\right) = 1-\left(\frac{A^{N_p+1}-A}{A^{N_p+1}-1}\right)$$

$$\frac{Y_{N_p+1}-mX_0-Y_{N_p+1}+Y_1}{Y_{N_p+1}-mX_0} = \frac{A^{N_p+1}-1-A^{N_p+1}+A}{A^{N_p+1}-1}$$

$$\frac{Y_1-mX_0}{Y_{N_p+1}-mX_0} = \frac{A-1}{A^{N_p+1}-1}$$

or, $$\frac{Y_{N_p+1}-mX_0}{Y_1-mX_0} = \frac{A^{N_p+1}-1}{A-1}$$

Now, $$A^{N_p+1} = \left(\frac{Y_{N_p+1}-mX_0}{Y_1-mX_0}\right)(A-1)+1$$

$$A^{N_p} = \left(\frac{Y_{N_p+1}-mX_0}{Y_1-mX_0}\right)\left(1-\frac{1}{A}\right)+\frac{1}{A}$$

Therefore,

$$N_p = \frac{\log\left[\frac{Y_{N_p+1} - mX_0}{Y_1 - mX_0}\left(1-\frac{1}{A}\right)+\frac{1}{A}\right]}{\log A}$$

For $A=1$ (operating line and equilibrium line are parallel)

from equation (3)

$$\frac{Y_{N_p+1} - Y_1}{Y_{N_p+1} - mX_0} = \frac{A^{N_p+1} - A}{A^{N_p+1} - 1} \quad — (4)$$

R.H.S of equation (4)

$$\lim_{A\to 1} \frac{A^{N_p+1} - A}{A^{N_p+1} - 1} = \lim_{A\to 1} \frac{(N_p+1)A^{N_p} - 1}{(N_p+1)A^{N_p}}$$

(From L Hospital rule)

$$= \lim_{A\to 1} \frac{(N_p+1)A^{N_p}\left(1 - \frac{1}{(N_p+1)A^{N_p}}\right)}{(N_p+1)A^{N_p}}$$

$$= \lim_{A\to 1} 1 - \frac{1}{(N_p+1)A^{N_p}} = 1 - \frac{1}{(N_p+1)}$$

$$= \frac{N_p+1-1}{N_p+1} = \frac{N_p}{N_p+1}$$

Hence, from equation (4), for A = 1

$$\frac{Y_{N_p+1} - Y_1}{Y_{N_p+1} - mX_0} = \frac{N_p}{N_p+1}$$

$$\frac{Y_{N_p+1} - mX_0}{Y_{N_p+1} - Y_1} = \frac{N_p+1}{N_p} = 1 + \frac{1}{N_p}$$

$$\therefore \quad \frac{1}{N_p} = \frac{Y_{N_p+1} - mX_0}{Y_{N_p+1} - Y_1} - 1$$

$$\frac{1}{N_p} = \frac{Y_{N_p+1} - mX_0 - Y_{N_p+1} + Y_1}{Y_{N_p+1} - Y_1}$$

$$\therefore \quad N_p = \frac{Y_{N_p+1} - Y_1}{Y_1 - mX_0}$$

These are called the Kremser - Brown - Souders (or simply Kremser) equations.

For the case of stripper, the final expression is :-

$$X_n = \left(X_0 - \frac{Y_{N_p+1}/m - AX_{N_p}}{1-A} \right) A^n + \frac{Y_{N_p+1}/m - AX_{N_p}}{1-A}$$

Put $n = N_p$ gives,

$$X_{N_p} = \left(X_0 - \frac{Y_{N_p+1}/m - AX_{N_p}}{1-A} \right) A^{N_p} + \frac{Y_{N_p+1}/m - AX_{N_p}}{1-A}$$

$$\therefore A^{N_p} = \frac{X_{N_p} - \dfrac{Y_{N_p+1}/m - AX_{N_p}}{1-A}}{X_0 - \dfrac{Y_{N_p+1}/m - AX_{N_p}}{1-A}}$$

$$A^{N_p} = \frac{X_{N_p} - Y_{N_p+1}/m}{X_0 - AX_0 - Y_{N_p+1}/m + AX_{N_p}}$$

$$A^{N_p} = \frac{(X_0 - Y_{N_p+1}/m) - (X_0 - X_{N_p})}{\underbrace{(X_0 - Y_{N_p+1}/m)}_{s} - A\underbrace{(X_0 - X_{N_p})}_{r}}$$

$\frac{r}{s} = ?$

$$A^{N_p} = \frac{s-r}{s-Ar} \quad , \quad s - Ar = \frac{s-r}{A^{N_p}}$$

$$s - Ar = \left(\frac{1}{A}\right)^{N_p} s - \left(\frac{1}{A}\right)^{N_p} r$$

$$r\left(A-\left(\frac{1}{A}\right)^{N_p}\right) = s\left(1-\left(\frac{1}{A}\right)^{N_p}\right)$$

$$\frac{r}{S} = \frac{1-\left(\frac{1}{A}\right)^{N_p}}{A-\left(\frac{1}{A}\right)^{N_p}}$$

or $$\frac{r}{S} = \frac{\left(\frac{1}{A}\right)^{N_p+1}-\frac{1}{A}}{\left(\frac{1}{A}\right)^{N_p+1}-1}$$

$$\therefore \frac{x_0-x_{N_p}}{x_0-Y_{N_p+1}/m} = \frac{\left(\frac{1}{A}\right)^{N_p+1}-\frac{1}{A}}{\left(\frac{1}{A}\right)^{N_p+1}-1} \quad -(5)$$

Now, reduce equation (5) to calculate N_p

as $\frac{S}{S-r}$

$$\because \frac{r}{S} = \frac{\left(\frac{1}{A}\right)^{N_p+1}-\frac{1}{A}}{\left(\frac{1}{A}\right)^{N_p+1}-1}$$

$$1-\frac{r}{S} = 1-\left\{\frac{\left(\frac{1}{A}\right)^{N_p+1}-\frac{1}{A}}{\left(\frac{1}{A}\right)^{N_p+1}-1}\right\}$$

$$\frac{S-r}{S} = \frac{\left(\frac{1}{A}\right)^{N_p+1}-1-\left(\frac{1}{A}\right)^{N_p+1}+\frac{1}{A}}{\left(\frac{1}{A}\right)^{N_p+1}-1}$$

$$\therefore \frac{S}{S-\gamma} = \frac{\left(\frac{1}{A}\right)^{N_p+1} - 1}{\frac{1}{A} - 1}$$

$$\therefore \left(\frac{1}{A}\right)^{N_p+1} = \left(\frac{S}{S-\gamma}\right)\left(\frac{1}{A} - 1\right) + 1$$

$$(N_p+1) \log\left(\frac{1}{A}\right) = \log\left[\frac{S}{(S-\gamma)}\left(\frac{1}{A} - 1\right) + 1\right]$$

$$\therefore N_p + 1 = \frac{\log\left\{\left(\frac{S}{S-\gamma}\right)\left(\frac{1}{A} - 1\right) + 1\right\}}{\log(1/A)}$$

$$N_p = \frac{\log\left\{\left(\frac{S}{S-\gamma}\right)\left(\frac{1}{A} - 1\right) + 1\right\}}{\log(1/A)} - 1$$

$$N_p = \frac{\log\left\{\left(\frac{S}{S-\gamma}\right)\left(\frac{1}{A} - 1\right) + 1\right\} - \log(1/A)}{\log(1/A)}$$

$$N_p = \frac{\log\left[\left(\frac{S}{S-\gamma}\right)(1-A) + A\right]}{\log(1/A)}$$

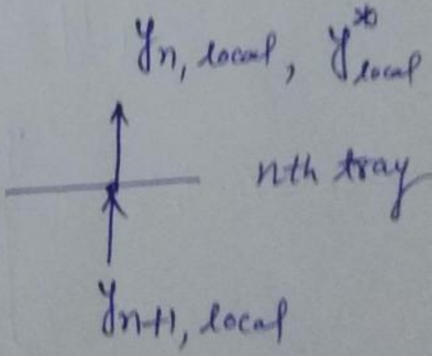
$$\therefore N_p = \frac{\log\left[\frac{X_0 - Y_{N_p+1}/m}{X_{N_p} - Y_{N_p+1}/m}(1-A) + A\right]}{\log(1/A)}$$

Tray Efficiency :-

Tray efficiency is the fractional approach to an equilibrium stage which is attained by a real tray. Since the conditions at various locations on the tray may differ, we begin by considering the local, or point, efficiency of mass transfer at a particular place on the tray surface.

Point efficiency :- Point efficiency on a particular location of tray is defined based on the changes of vapor phase composition, assuming that the local liquid concentration x_{local} is constant in the vertical direction. The point efficiency is then defined by

$$E_{OG} = \frac{y_{n,local} - y_{n+1,local}}{y^*_{local} - y_{n+1,local}} \quad \text{—(1)}$$



Here, y^*_{local} is the concentration in equilibrium with x_{local}, and above equation (1), then represents the change in gas concentration which actually occurs as a fraction of that which would occur if equilibrium were established.

The subscript G signifies that gas concentrations are used, and the O emphasizes that E_{OG} is a measure of the overall resistance to mass transfer for both phases.

Consider that the gas rises at a rate G mole/(area)(time). Let the interfacial surface between gas and liquid be a area/volume of liquid-gas foam. As the gas rises a differential height dh_L, the area of contact is $a\,dh_L$ per unit area of tray. If, while of concentration y, it undergoes a concentration change dy in this height, and if the total quantity of gas remains essentially constant, the rate of solute transfer is $G\,dy$:

$$G\,dy = K_y\,(a\,dh_L)\,(y^*_{local} - y)$$

Then,

$$\int_{y_{n+1,\,local}}^{y_{n,\,local}} \frac{dy}{y^*_{local} - y} = \int_0^{h_L} \frac{K_y\,a\,dh_L}{G}$$

Since y^*_{local} is constant for constant x_{local},

$$-\ln \frac{y^*_{local} - y_{n,\,local}}{y^*_{local} - y_{n+1,\,local}} = -\ln\left(1 - \frac{y_{n,\,local} - y_{n+1,\,local}}{y^*_{local} - y_{n+1,\,local}}\right)$$

$$= -\ln(1 - E_{OG}) = \frac{K_y a h_L}{G}$$

Therefore

$$E_{OG} = 1 - e^{-\frac{K_y a h_L}{G}} = 1 - e^{N_{tOG}}$$

The exponent of e is simplified to N_{tOG}, the number of overall gas-transfer units. Just as K_y contains both gas and liquid resistance to mass transfer, so also is N_{tOG} made up of the transfer units for the gas N_{tG} and those for the liquid N_{tL}. These can be combined as below:-

$$\frac{1}{N_{tOG}} = \frac{1}{N_{tG}} + \frac{mG}{L} \frac{1}{N_{tL}}$$

The terms on the right represent, respectively, the gas and liquid mass-transfer resistances, which must be obtained experimentally.

Murphree tray efficiency:- The bulk-average concentrations of all the local pencils of gas are y_{n+1} and y_n. The Murphee efficieny of the entire tray is then

$$E_{MG} = \frac{y_n - y_{n+1}}{y_n^* - y_{n+1}}$$

where y_n^* is the value in the equilibrium with the leaving liquid of concentration x_n.

The relationship between E_{MG} and E_{OG} can then be derived by integrating the local E_{OG}'s over the surface of the tray. Clearly, if all the gas entering were uniformly mixed and fed uniformly to the entire tray cross section, and if the mechanical contacting of gas and liquid were everywhere uniform, the uniformity of ~~liquid~~ concentration of exit gas $y_{n+1,\text{local}}$ would then depend on the uniformity of liquid concentration on the tray. Liquid on the tray is splashed about by the action of the gas, some of it even being thrown backward in the direction from which it enters the tray (back mixing). The two extreme cases which might be visualized are :-

(a) Liquid completely back-mixed, everywhere of uniform concentration x_n. In this case

$$E_{MG} = E_{OG}$$

(b) Liquid in plug flow, with no mixing, each particle remaining on the tray for the same length of time. In this case, it has been shown that

$$E_{MG} = \frac{L}{mG}\left(e^{E_{OG}\, mG/L} - 1\right)$$

and $E_{MG} > E_{OG}$.

In the more likely intermediate case, the transport of solute by the mixing process can be described in terms of an eddy diffusivity D_E, whereupon

$$\frac{E_{MG}}{E_{OG}} = \frac{1-e^{-(\eta+Pe)}}{(\eta+Pe)\left[1+(\eta+Pe)/\eta\right]} + \frac{e^{\eta}-1}{\eta\left[1+\eta/(\eta+Pe)\right]}$$

where $$\eta = \frac{Pe}{2}\left[\left(1+\frac{4mG\,E_{OG}}{L\,Pe}\right)^{0.5}-1\right]$$

and $$Pe = \frac{Z^2}{D_E\,\theta_L}$$

Here θ_L is the liquid residence time on tray and Z the length of liquid travel. Pe is a Peclet number, as can be seen better by writing is as $(z/D_E)(z/\theta_L)$, where z/θ_L becomes the average liquid velocity. $Pe = 0$ corresponds to complete backmixing ($D_E = \infty$), while $Pe = \infty$ corresponds to plug flow ($D_E = 0$). Large value of Pe result when mixing is not extensive and for large values of z (large tower diameters). Although point efficiencies cannot exceed unity, it is possible for Murfree efficiencies.

Entrainment :- A further correction is required for the damage done by entrainment. Entrainment represents a foam of back mixing, which acts to destroy the concentration changes produced by the trays. It can be shown that the Murphree efficiency corrected for entrainment is

$$E_{MGE} = \frac{E_{MG}}{1 + E_{MG}\left[\frac{E}{1-E}\right]}$$

Overall tray efficiency :- Another method of describing the performance of a tray tower is through the overall tray efficiency,

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

when the Murphree efficiency is constant for all trays, and under conditions such that the operating line and equilibrium curves are straight (Henry's law, isothermal operation, dilute solutions), the overall tray efficiency can be computed and the number of real trays can be determined analytically:-

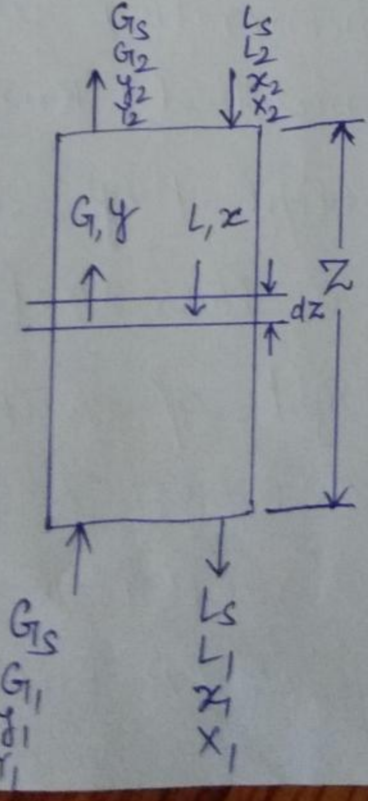
$$E_0 = \frac{\text{equilibrium trays}}{\text{real trays}} = \frac{\log\left[1 + E_{MGE}\left(1/A - 1\right)\right]}{\log\left(1/A\right)}$$

Height Equivalent to an Equilibrium Stage (Theoretical plate) :-

A simple method for designing packed towers, ignores the differences between stagewise and continuous contact. In this method the number of theoretical trays or plates required for a given change in concentration is computed by either graphical or analytical (Kremser equations) methods. This is then multiplied by a quantity, the height equivalent to a theoretical tray (HETP) to give the required height of packing to do the same job. The HETP must be an experimently determined quantity characteristic for each packing. owing to variation of HETP with type and size of packing, and liquid and gas flow rates; and for every system with concentration, so that the enormous amount of data would have to be accumulated to permit utilization of the method, and the method has now largely been abandoned.

Continuous-contact Equipment, Absorption of one component :-

Consider a packed tower of unit cross section, as shown in below figure. The total effective interfacial surface for mass transfer, as a result of spreading of the liquid in a film over the packing, is S per unit tower cross section. This is conveniently described as the product of a specific interfacial surface, surface per unit volume of packing, by the packed volume Z volume per unit tower cross section, or height.



Packed tower

In the differential volume dZ, the interfecial surface is

$$dS = a\,dZ$$

The quantity of solute A in the gas passing the differential section of the tower under consideration is Gy mole/(area)(time), and the rate of mass transfer is therefore $d(Gy)$ mole/(time)(area), this is intrinsically negative in the case of absorption. Since $N_B = 0$, and $N_A/(N_A + N_B) = 1$, application of below equation (1) provides

$$N_A = \frac{N_A}{\Sigma N} F_G \ln \frac{N_A/\Sigma N - y_{A,i}}{N_A/\Sigma N - y_{A,G}} = \frac{N_A}{\Sigma N} F_L \ln \frac{N_A/\Sigma N - x_{A,L}}{N_A/\Sigma N - x_{A,i}} \quad \text{—(1)}$$

$$N_A = F_G \ln \frac{1-y_i}{1-y} = \frac{-d(Gy)}{a\,dZ} \quad \text{— (2)}$$

Both G and y vary from one end of the tower to the other, but G_s, the solvent gas which is essentially insoluble, does not. Therefore,

$$d(Gy) = d\left(\frac{G_s y}{1-y}\right) = \frac{G_s\,dy}{(1-y)^2} = \frac{G\,dy}{1-y}$$

Substituting the value of $d(Gy)$ in equation (2), rearranging, and integrating gives

$$Z = \int_0^Z dZ = \int_{y_2}^{y_1} \frac{G\,dy}{F_G a\,(1-y)\ln[(1-y_i)/(1-y)]} \qquad -(3)$$

The value of y_i can be found by the methods for calculating interface concentrations.

For $N_A/\Sigma N = 1$

$$\frac{1-y_i}{1-y} = \left(\frac{1-x}{1-x_i}\right)^{F_L/F_G} = \left(\frac{1-x}{1-x_i}\right)^{F_L a/F_G a} \qquad -(4)$$

For any value of (x, y) on the operating curve plotted in terms of mole fractions, a curve of x_i v.s y_i from equation (4) is plotted to determine the intersection with the equilibrium curve. This provide local y and y_i for use in equation (3). Equation (3) can then be integrated graphically after plotting the integrand as ordinate vs. y as abscissa.

However, it is more customary to proceed as follows:-

since $y - y_i = (1-y_i) - (1-y)$

$$(1-y)_{iM} = \frac{(1-y_i) - (1-y)}{\ln\left(\frac{1-y_i}{1-y}\right)}$$

where, $(1-y)_{iM}$ is the logarithmic mean of $(1-y_i)$ and $(1-y)$.

Therefore,

$$\ln\left(\frac{1-y_i}{1-y_t}\right) = \frac{(1-y_i)-(1-y)}{(1-y)_{iM}}$$

substitute in equation (3) gives

$$Z = \int_{y_2}^{y_1} \frac{G\,(1-y)_{iM}\,dy}{F_G\,a\,(1-y)\,(y-y_i)} \quad — \quad (5)$$

when we define a height of a gas transfer unit H_{tG} as

$$H_{tG} = \frac{G}{F_G\,a} = \frac{G}{k_y\,a(1-y)_{iM}} = \frac{G}{k_G\,a P_t\,(1-y)_{iM}}$$

equation (5) becomes

$$Z = \int_{y_2}^{y_1} H_{tG}\,\frac{(1-y)_{iM}\,dy}{(1-y)\,(y-y_i)} \simeq H_{tG}\int_{y_2}^{y_1} \frac{(1-y)_{iM}\,dy}{(1-y)(y-y_i)} = H_{tG}N_{tG} \quad —(6)$$

Here advantage is taken of the fact that the ratio $G/F_G a = H_{tG}$ is very much more constant than either G or $F_G a$ and in many cases may be considered constant within the accuracy of the available data.

$$N_{tG} = \int_{y_2}^{y_1} \frac{(1-y)_{iM}\,dy}{(1-y)\,(y-y_i)}$$ is called the number of

gas transfer units. It contains only the y terms. This is a measure of the difficulty of the absorption.

H_{tG} is then the packing height providing one gas transfer unit.

Equation (6) can be further simplified by substituting the arithmetic average for the logarithmic average $(1-y)_{iM}$

$$(1-y)_{iM} = \frac{(1-y_i)-(1-y)}{\ln[(1-y_i)/(1-y)]} \simeq \frac{(1-y_i)+(1-y)}{2}$$

which involves little error, N_{tG} then becomes

$$N_{tG} = \int_{y_2}^{y_1} \frac{dy[(1-y_i)+(1-y)]}{2(1-y)(y-y_i)}$$

$$\because \frac{(1-y_i)+(1-y)}{2} = \frac{2(1-y)-(1-y)+(1-y_i)}{2}$$

$$= (1-y)+\frac{1}{2}(y-y_i)$$

$$N_{tG} = \int_{y_2}^{y_1} \frac{[(1-y)+\frac{1}{2}(y-y_i)]\,dy}{(1-y)(y-y_i)}$$

$$N_{tG} = \int_{y_2}^{y_1} \frac{(1-y)\,dy}{(1-y)(y-y_i)} + \frac{1}{2}\int_{y_2}^{y_1} \frac{(y-y_1)\,dy}{(1-y)(y-y_i)}$$

$$N_{tG} = \int_{y_2}^{y_1} \frac{dy}{(y-y_i)} + \frac{1}{2}\int_{y_2}^{y_1} \frac{dy}{1-y}$$

Therefore,

$$N_{tG} = \int_{y_2}^{y_1} \frac{dy}{(y-y_i)} + \frac{1}{2}\ln\left(\frac{1-y_2}{1-y_1}\right) \quad \text{— (7)}$$

which makes for simpler graphical integration. A plot of $1/(y-y_i)$ vs. y for the graphical integration of equation (7) often covers awkwardly large ranges of the ordinate. This can be avoided by replacing dy by its equal $y\,d\ln y$, so that

$$N_{tG} = 2.303\int_{\log y_2}^{\log y_1} \frac{y}{(y-y_i)}\,d\log y + 1.152\log\left(\frac{1-y_2}{1-y_1}\right) \quad \text{—(8)}$$

For dilute solutions, the second term on the right of equations (7) and (8) is negligible, $F_G a \approx k_y a$, and y_i can be obtained by plotting a line of slope $-k_x a/k_y a$ from point (x, y) on the operating line to intersection with the equilibrium curve.

The above relationships all have their counterparts in terms of liquid concentrations, derived in exactly the same way

$$Z = \int_{x_2}^{x_1} \frac{L\,dx}{F_L a(1-x)\ln\left[(1-x)/(1-x_i)\right]} = \int_{x_2}^{x_1} \frac{L\,(1-x)_{iM}\,dx}{F_L a\,(1-x)(x_i-x)}$$

$$Z = \int_{x_2}^{x_1} H_{tL}\,\frac{(1-x)_{iM}\,dx}{(1-x)(x_i-x)} \approx H_{tL}\int_{x_2}^{x_1} \frac{(1-x)_{iM}\,dx}{(1-x)(x_i-x)} = H_{tL}N_{tL}$$

$$H_{tL} = \frac{L}{F_L a} = \frac{L}{k_x a\,(1-x)_{iM}} = \frac{L}{k_L a c\,(1-x)_{iM}}$$

$$N_{tL} = \int_{x_2}^{x_1} \frac{dx}{x_i - x} + \frac{1}{2}\ln\left(\frac{1-x_1}{1-x_2}\right)$$

where, H_{tL} = height of liquid transfer unit
N_{tL} = number of liquid transfer unit
$(1-x)_{iM}$ = logarithmic mean of $(1-x)$ and $(1-x_i)$.

For strippers, same relationships apply as for absorption. The driving forces $(y-y_i)$ and (x_i-x) which appears in the above equations are then negative, but since for strippers $x_2 > x_1$ and $y_2 > y_1$, the result is a positive Z as before.

Overall coefficients and transfer units ($m = \frac{dy_i}{dx_i} =$ const) :-

For cases where the equilibrium distribution curve is straight and the ratio of mass-transfer coefficients is constant, the overall mass-transfer coefficients are convenient. The expressions for the height of packing can then be written

$$Z = N_{tOG} H_{tOG}$$

$$N_{tOG} = \int_{y_2}^{y_1} \frac{(1-y)_{*m}\, dy}{(1-y)(y-y^*)}$$

$$N_{tOG} = \int_{y_2}^{y_1} \frac{dy}{y-y^*} + \frac{1}{2} \ln \frac{1-y_2}{1-y_1}$$

$$N_{tOG} = \int_{Y_2}^{Y_1} \frac{dY}{Y-Y^*} + \frac{1}{2} \ln \frac{1+Y_2}{1+Y_1}$$

$$H_{tOG} = \frac{G}{F_{OG}\, a} = \frac{G}{K_y a (1-y)_{*m}} = \frac{G}{K_G a P_t (1-y)_{*m}}$$

Here y^* (or Y^*) is the solute concentration in the gas corresponding to equilibrium with the bulk liquid concentration x (or X), so that $y-y^*$ (or $Y-Y^*$) is simply the vertical distance between operating line and equilibrium curve. $(1-y)_{*M}$ is the logarithmic average of $(1-y)$ and $(1-y^*)$. These methods are convenient since ~~the operating line on X, Y coordinates is straight~~ interfacial concentrations need not be obtained. N_{tOG} is the number of overall gas transfer units, H_{tOG} the height of an overall gas transfer unit

For cases where the principal mass-transfer resistance lies within the liquid, it is more convenient to use

$$Z = N_{tOL} H_{tOL}$$

$$N_{tOL} = \int_{x_2}^{x_1} \frac{(1-x)_{*m}\, dx}{(1-x)(x^*-x)}$$

$$N_{tOL} = \int_{x_2}^{x_1} \frac{dx}{(x^*-x)} + \frac{1}{2} \ln\left(\frac{1-x_1}{1-x_2}\right)$$

$$N_{tOL} = \int_{X_2}^{X_1} \frac{dX}{X^*-X} + \frac{1}{2} \ln\left(\frac{1+X_1}{1+X_2}\right)$$

$$H_{tOL} = \frac{L}{F_{OL}\, a} = \frac{L}{K_x a (1-x)_{*M}} = \frac{L}{K_L a c (1-x)_{*M}}$$

Dilute Solutions :-

The computation of the number of transfer units for dilute mixtures can be greatly simplified. When the gas mixture is dilute, for example, the second term of the definition of N_{toG} becomes entirely negligible and can be discarded

$$N_{toG} = \int_{y_2}^{y_1} \frac{dy}{y-y^*}$$

If the equilibrium curve in terms of mole fractions is linear over the range of compositions x_1 to x_2, then

$$y^* = mx + r \qquad \text{—— (1)}$$

If the solutions are dilute, the operating line can be considered as a straight. line as well

$$y = \frac{L}{G}(x - x_2) + y_2 \qquad \text{—— (2)}$$

substracting equation (1) from equation (2) gives,

$$y - y^* = \left(\frac{L}{G} - m\right)x - \frac{Lx_2}{G} + y_2 - r$$

$$y - y^* = qx + s$$

where, $q, r,$ and s are constants. Therefore,

$$N_{toG} = \frac{L}{G}\int_{x_2}^{x_1} \frac{dx}{qx+s} = \frac{L}{Gq} \ln \frac{(y-y^*)_1}{(y-y^*)_2} \quad -③$$

Since $dy = \frac{L}{G}dx$ from eq. (2)

$$N_{toG} = \frac{y_1 - y_2}{\dfrac{(y-y^*)_1 - (y-y^*)_2}{\ln\left[(y-y^*)_1/(y-y^*)_2\right]}}$$

$$N_{toG} = \frac{y_1 - y_2}{(y-y^*)_M}$$

$$\because \frac{L}{G} = \frac{y_2 - y_1}{x_2 - x_1}, \text{ and } q = \frac{(y-y^*)_1 - (y-y^*)_2}{x_1 - x_2}$$

$$\text{Therefore, } \frac{L}{Gq} = \frac{y_1 - y_2}{(y-y^*)_1 - (y-y^*)_2}$$

substituting in equation ③, we get

$$N_{toG} = \frac{y_1 - y_2}{(y-y^*)_1 - (y-y^*)_2} \ln \frac{(y-y^*)_1}{(y-y^*)_2}$$

$$\therefore \quad N_{toG} = \frac{y_1 - y_2}{(y-y^*)_M}$$

where, $(y-y^*)_M$ is the logarithmic average of the concentration difference at the ends of the tower. This equation is sometimes used in the familiar rate form obtained by substituting the definition of N_{toG}

since,

$$Z = H_{toG} N_{toG}$$

$$Z = \frac{G}{K_G a P_t (1-y)_{*M}} \cdot N_{toG}$$

For dilute solution,

$$Z = \frac{G}{K_G a P_t} \cdot \frac{y_1 - y_2}{(y-y^*)_M}$$

Therefore,

$$G(y_1 - y_2) = K_G a Z P_t (y-y^*)_M$$

Dilute solutions, Henry's law :-

If Henry's law applies r of equation (1), will be zero, substituting $x = y^*/m$ in equation ②

yields

$$y = \frac{L}{G}\left(\frac{y^*}{m} - x_2\right) + y_2$$

or,

$$y = \frac{L}{G}\left(\frac{y^*}{m} - x_2\right) + y_2$$

~~$\frac{mG}{L}y = y^* - \frac{mG}{L}x_2 + y_2$~~

$$\frac{mGy}{L} = y^* - mx_2 + \frac{mG}{L}y_2$$

$$\therefore (y - y^*) = \left(1 - \frac{mG}{L}\right)y + \frac{mG}{L}y_2 - mx_2$$

$$\therefore N_{tOG} = \int_{y_2}^{y_1} \frac{dy}{y - y^*} = \int_{y_2}^{y_1} \frac{dy}{\left(1 - \frac{mG}{L}\right)y + \frac{mG}{L}y_2 - mx_2}$$

or,

$$N_{tOG} = \frac{1}{\left(1 - \frac{mG}{L}\right)} \ln\left[\frac{\left(1 - \frac{mG}{L}\right)y_1 + \frac{mG}{L}y_2 - mx_2}{\left(1 - \frac{mG}{L}\right)y_2 + \frac{mG}{L}y_2 - mx_2}\right]$$

or,

$$N_{tOG} = \frac{\ln\left[\left(\frac{y_1 - mx_2}{y_2 - mx_2}\right)\left(1 - \frac{mG}{L}\right) + \frac{mG}{L}\right]}{\left(1 - \frac{mG}{L}\right)}$$

or

$$N_{tOG} = \frac{\ln\left[\left(\frac{y_1 - mx_2}{y_2 - mx_2}\right)\left(1 - \frac{1}{A}\right) + \frac{1}{A}\right]}{1 - \frac{1}{A}}$$

where, $A = \frac{L}{mG}$

For strippers, the corresponding expression in terms of N_{toL} is similar

$$N_{toL} = \frac{\ln\left[\left(\frac{x_2 - y_1/m}{x_1 - y_1/m}\right)(1-A) + A\right]}{1-A}$$

—o—

Prove $N_{toG} = \int_{y_2}^{y_1} \frac{dy}{\left(1-\frac{mG}{L}\right)y + \frac{mG}{L}y_2 - mx_2}$ = final expressions.

$$= \frac{1}{\left(1-\frac{mG}{L}\right)} \int_{y_2}^{y_1} \frac{dy}{y + \frac{\frac{mG}{L}}{\left(1-\frac{mG}{L}\right)}y_2 - \frac{mx_2}{\left(1-\frac{mG}{L}\right)}}$$

$$= \frac{1}{\left(1-\frac{mG}{L}\right)} \ln\left[\frac{y_1 + \frac{mGy_2/L}{\left(1-\frac{mG}{L}\right)} - \frac{mx_2}{\left(1-\frac{mG}{L}\right)}}{y_2 + \frac{mGy_2/L}{\left(1-\frac{mG}{L}\right)} - \frac{mx_2}{\left(1-\frac{mG}{L}\right)}}\right]$$

$$= \frac{1}{\left(1-\frac{mG}{L}\right)} \ln\left[\frac{y_1\left(1-\frac{mG}{L}\right) + \frac{mG}{L}y_2 - mx_2}{y_2\left(1-\frac{mG}{L}\right) + \frac{mG}{L}y_2 - mx_2}\right]$$

$$= \frac{1}{\left(1-\frac{mG}{L}\right)} \ln\left[\frac{(y_1 - mx_2) - y_1\frac{mG}{L} + \frac{mGy_2}{L}}{(y_2 - mx_2) - y_2\frac{mG}{L} + \frac{mGy_2}{L}}\right]$$

$$= \frac{1}{\left(1-\frac{mG}{L}\right)} \ln\left[\frac{(y_1 - mx_2) - \frac{mG}{L}(y_1 - y_2)}{(y_2 - mx_2)}\right]$$

$$= \frac{1}{\left(1-\frac{mG}{L}\right)} \ln\left[\left(\frac{y_1 - mx_2}{y_2 - mx_2}\right) - \frac{mG}{L}\left(\frac{y_1 - y_2}{y_2 - mx_2}\right)\right]$$

$$= \frac{1}{\left(1-\frac{mG}{L}\right)} \ln\left[\left(\frac{y_1 - mx_2}{y_2 - mx_2}\right) - \frac{mG}{L}\left(\frac{y_1 - mx_2}{y_2 - mx_2} - \frac{y_2 - mx_2}{y_2 - mx_2}\right)\right]$$

$$= \frac{1}{\left(1-\frac{mG}{L}\right)} \ln\left[\left(\frac{y_1 - mx_2}{y_2 - mx_2}\right) - \frac{mG}{L}\left(\frac{y_1 - mx_2}{y_2 - mx_2} - 1\right)\right]$$

$$= \frac{1}{\left(1-\frac{mG}{L}\right)} \ln\left[\left(\frac{y_1 - mx_2}{y_2 - mx_2}\right)\left(1 - \frac{mG}{L}\right) + \frac{mG}{L}\right]$$

or $$N_{tOG} = \frac{\ln\left[\left(\frac{y_1 - mx_2}{y_2 - mx_2}\right)\left(1 - \frac{1}{A}\right) + \frac{1}{A}\right]}{\left(1 - \frac{1}{A}\right)}$$

Where, $A = \frac{L}{mG}$

Proved.

Overall Heights of transfer units :-

when overall numbers of transfer units are appropriate, the overall heights of transfer units can be synthesized from those for the individual phases through the relationships with overall mass transfer $m' = m'' = m$ = constant for coefficients and can be written as :-

$$\frac{G}{F_{OG}a} = \frac{G(1-y)_{iM}}{F_G a\,(1-y)_{*M}} + \frac{mG}{L}\,\frac{L}{F_L a}\,\frac{(1-x)_{iM}}{(1-y)_{*M}}$$

whence, by definition of the heights of transfer units,

$$H_{tOG} = H_{tG}\,\frac{(1-y)_{iM}}{(1-y)_{*M}} + \frac{mG}{L}\,H_{tL}\,\frac{(1-x)_{iM}}{(1-y)_{*M}}$$

If the mass-transfer resistance is essentially all in the gas, $y_i \approx y^*$, and

$$H_{tOG} = H_{tG} + \frac{mG}{L}\,H_{tL}\,\frac{(1-x)_{iM}}{(1-y)_{*M}}$$

and for dilute solutions, the concentration ratio of the last equation can be dropped. In similar fashion

$$H_{tOL} = H_{tL}\,\frac{(1-x)_{iM}}{(1-x)_{*M}} + \frac{L}{mG}\,H_{tG}\,\frac{(1-y)_{iM}}{(1-x)_{*M}}$$

and if the mass-transfer resistance is essentially all in the liquid,

$$H_{tOL} = H_{tL} + \frac{L}{mG} H_{tG} \frac{(1-y)_{iM}}{(1-x)_{*M}}$$

The concentration ratio of the last equation can be dropped for dilute solutions.

HUMIDIFICATION

- **Humidity** H is the mass of vapour carried by a unit mass of vapour free gas.
- $H = \frac{M_A p_A}{M_B(P-p_A)}$
- p_A =partial pressure of the vapour , M_A= molecular weight of vapour
- M_B= molecular weight of vas
-
- Relationship between **mole fraction** and humidity
- $y = \frac{H/M_A}{1/M_B+H/M_A}$
- **Saturated gas** is gas in which the vapour is in equilibrium with the liquid at the gas temperature.
- $H_S = \frac{M_A P_A'}{M_B(P-P_A')}$
- $P_A' = vapour\ pressure\ of\ the\ liquid$

- **Relative Humidity**, H_R, is defined as the ratio of the partial pressure of the vaour to the vapour pressure of the liquid at the gas temperature. It is usually expressed as a percentage basis, 100% meaning saturated gas and 0% means vapour free gas.

$$H_R = 100\frac{p_A}{P'_A}$$

- **Percentage humidity, H_A**, is the ratio of the actual humidity H to the saturation humidity H_S, gas temperature:

$$H_A = 100\frac{H}{H_s} = 100\frac{p_A/(P - p_A)}{P'_A/(P - P'_A)} = H_R\frac{P - P'_A}{P - p_A}$$

- **Humid Heat** c_s is the heat energy necessary to increase the temperature of 1 g of gas plus whatever vapour it may contain by 1^oC.

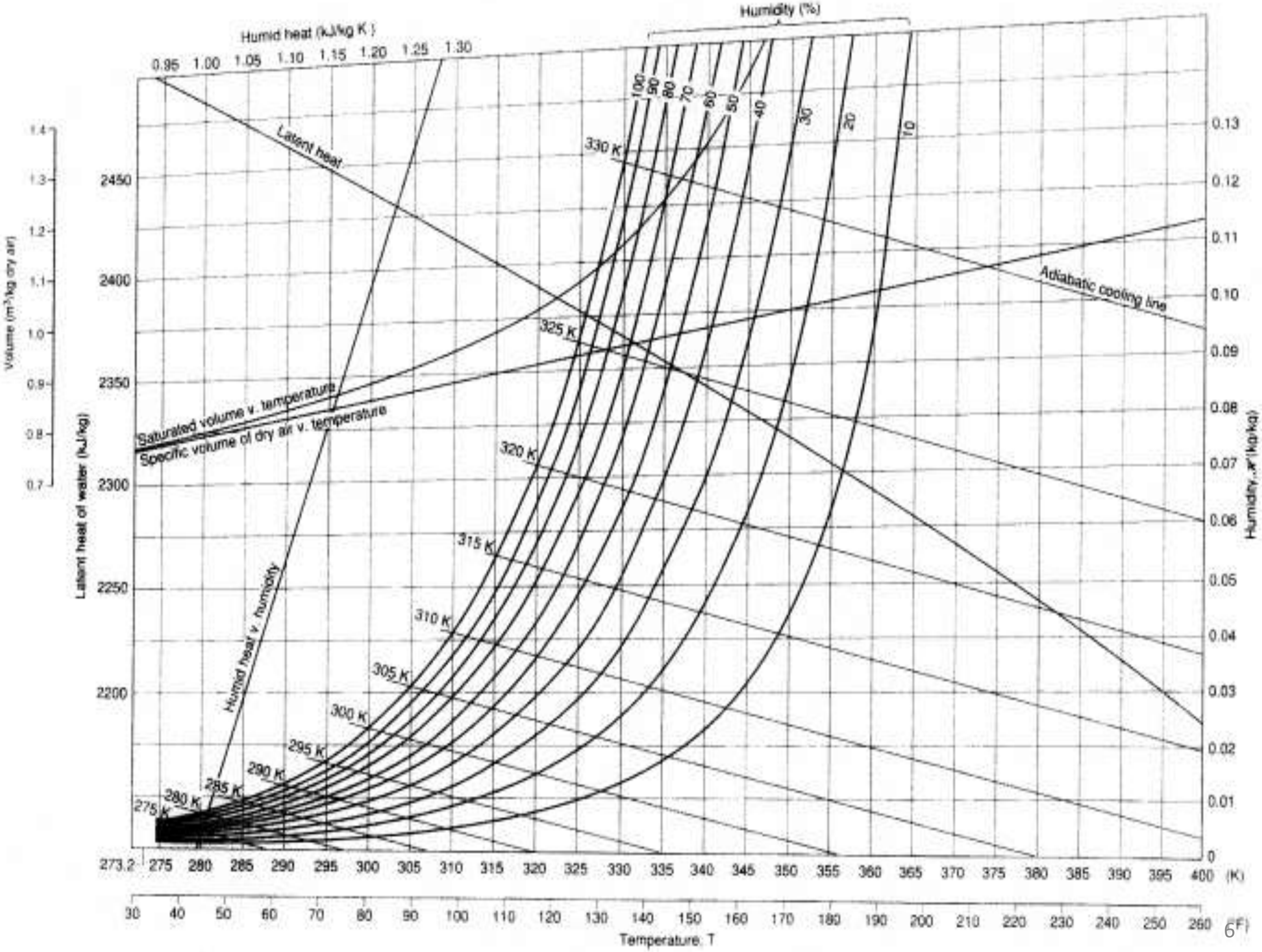
$$c_S = c_{pB} + c_{pA}H$$

- **Humid volume** v_H is the total volume of a unit mass of vapour free gas plus the accompanying vapour at 1 atm and gas temperature.

$$v_H = \frac{0.0224T}{273}\left(\frac{1}{M_B} + \frac{H}{M_A}\right)$$

- v_H is in cubic meters per gram and T is in ^oK
- **Dew point** is the temperature to which a vapour gas mixture must be cooled at constant humidity to become saturated.

- **Total enthalpy** is the enthalpy of a unit mass of gas plus accompanying vapour.
- $H_y = c_{pB}(T - T_0) + H\lambda_0 + c_{pA}H(T - T_0)$
- $H_y = c_S(T - T_0) + H\lambda_0$


Humidity (%)
Humid heat (kJ/kg K)
0.95
1.00
1.05
1.10
1.15
1.20
1.25
1.30
100
90
80
70
60
50
40
30
20
10
Latent heat
330 K
325 K
320 K
315 K
310 K
305 K
300 K
295 K
290 K
285 K
280 K
275 K
Adiabatic cooling line
Saturated volume v. temperature
Specific volume of dry air v. temperature
Humid heat v. humidity
Volume (m³/kg dry air)
1.4
1.3
1.2
1.1
1.0
0.9
0.8
0.7
Latent heat of water (kJ/kg)
2450
2400
2350
2300
2250
2200
Humidity, ℋ (kg/kg)
0.13
0.12
0.11
0.10
0.09
0.08
0.07
0.06
0.05
0.04
0.03
0.02
0.01
0
273.2
275
280
285
290
295
300
305
310
315
320
325
330
335
340
345
350
355
360
365
370
375
380
385
390
395
400
(K)
30
40
50
60
70
80
90
100
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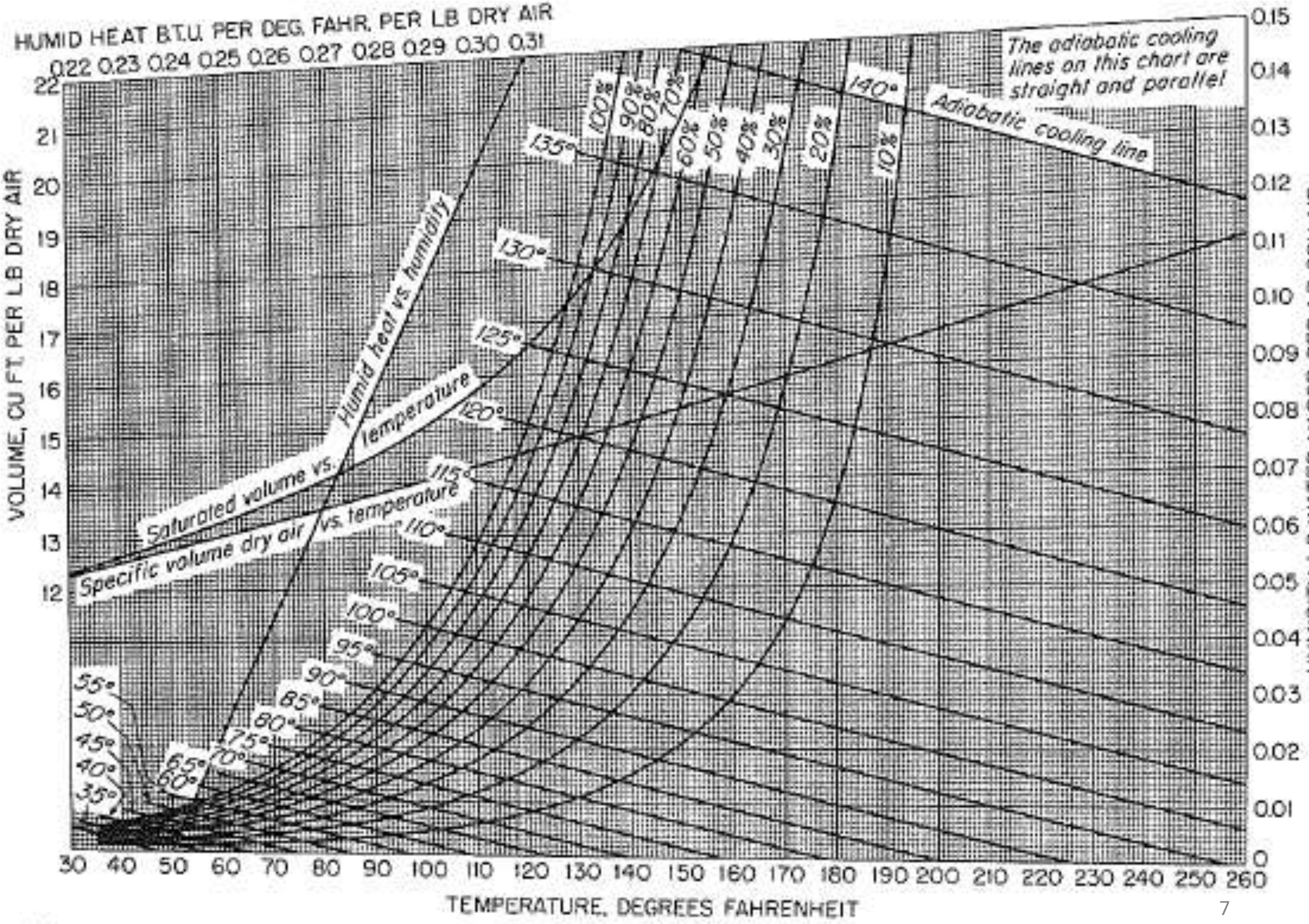
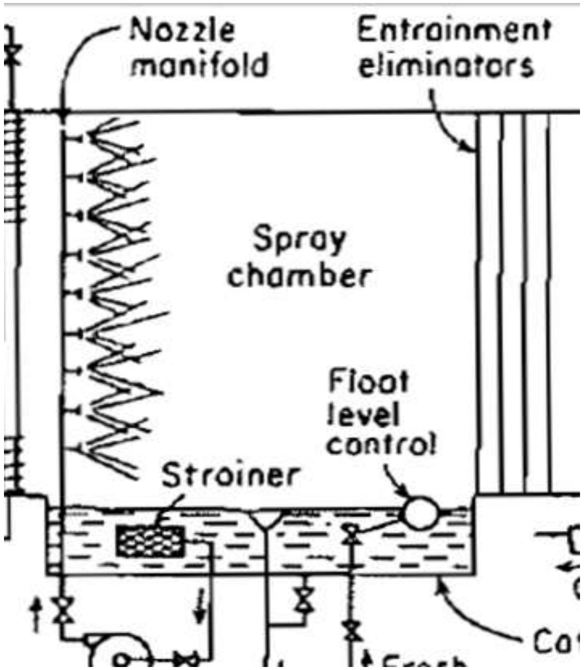
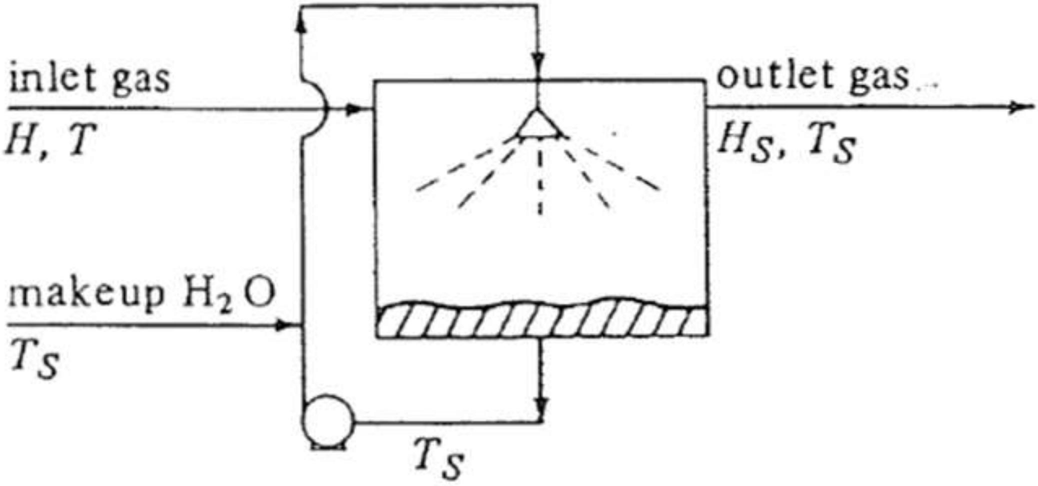
260
(°F)
Temperature, T


HUMID HEAT B.T.U. PER DEG. FAHR. PER LB DRY AIR
0.22 0.23 0.24 0.25 0.26 0.27 0.28 0.29 0.30 0.31
VOLUME, CU FT PER LB DRY AIR
22
21
20
19
18
17
16
15
14
13
12
0.15
0.14
0.13
0.12
0.11
0.10
0.09
0.08
0.07
0.06
0.05
0.04
0.03
0.02
0.01
0
30 40 50 60 70 80 90 100 110 120 130 140 150 160 170 180 190 200 210 220 230 240 250 260
TEMPERATURE, DEGREES FAHRENHEIT
The adiabatic cooling lines on this chart are straight and parallel
Adiabatic cooling line
Humid heat vs. humidity
Saturated volume vs. temperature
Specific volume dry air vs. temperature
100%
90%
80%
70%
60%
50%
40%
30%
20%
10%
140°
135°
130°
125°
120°
115°
110°
105°
100°
95°
90°
85°
80°
75°
70°
65°
60°
55°
50°
45°
40°
35°

- The temperature and dew point of the air entering a certain dryer are 65.6 and 15.6oC respectively.
- [vapour pressure = 25.1 kPa, atmospheric pressure = 101.35kPa]
- Determine:
  1. Humidity
  2. Mole fraction of moisture in air
  3. Partial pressure of moisture in air
  4. Relative humidity and % saturation
  5. Adiabatic saturation temperature/wet bulb temperature
  6. Humid heat
  7. Saturated volume and humid volume

ADIABATIC SATURATION





- If air is sprayed into aa stream of gas in an insulated pipe or spray chamber, it is cooled and humidified. If not all the water evaporates and there is sufficient time for the gas to come to equilibrium with water the exit temperature [also that of the remaining water] is called the **adiabatic saturation temperature** Ts

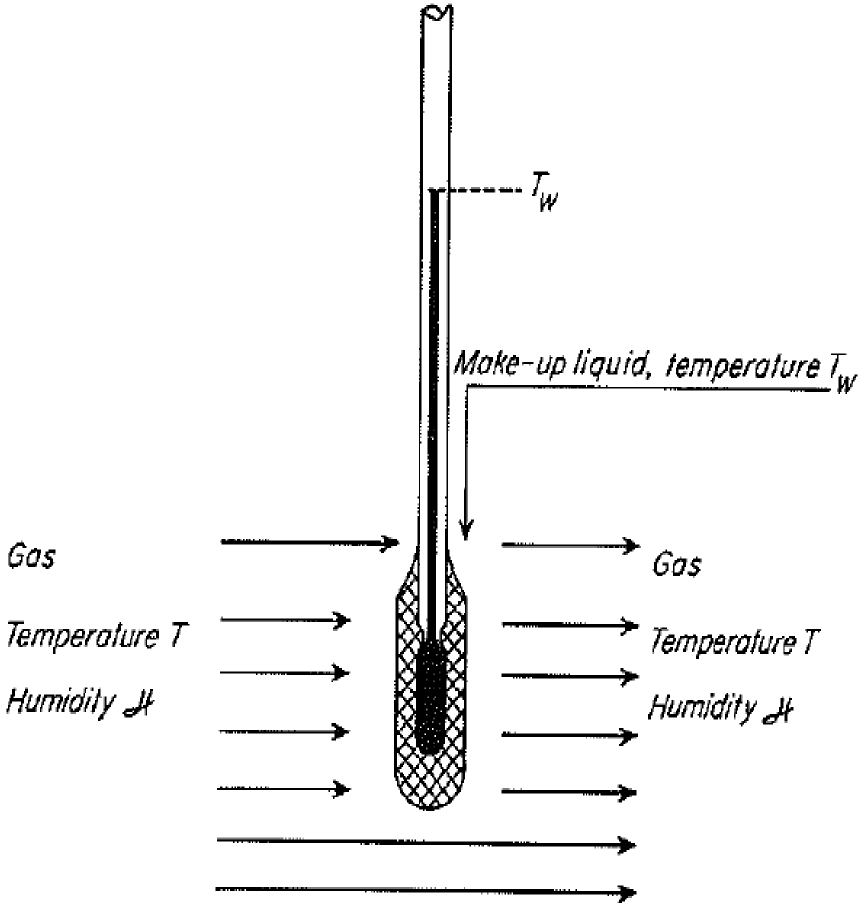
Adiatic saturation temperature, T_s

- $c_{pB}(T - T_s) + c_{pA}H(T - T_s) = (H_s - H)\lambda_s$

$$\frac{H_s - H}{T - T_s} = \frac{c_s}{\lambda_s} = \frac{c_{pB} + c_{pA}H}{\lambda_s}$$

- $T_s = Adiabatic\ saturation\ temperature$
- Ts can be determined by **trial and error** only as Hs and λs are function of Ts
- Adiabatic cooling curve is drawn from a particular, Ts, when these parameters can be fixed.

Wet bulb temperature



- Wet bulb temperature is the steady state , non-equilibrium temperature reached by a small mass of liquid exposed under adiabatic conditions to a continuous stream of gas

MEASUREMENT

- Wick must be completely wet
- Velocity of air at least 5m/s
- Make up liquid is supplied

WET BULB TEMPERATURE

- At steady state
- Heat transfer from gas to liquid = rate of vaprization x sum of latent heat of vaporization and sensible heat of vapour
- Rate of heat transfer to liquid,

$$q = M_A N_A \left[\lambda_w + c_{pA}(T - T_w)\right] = h_y(T - T_i)A$$

- NA = molal rate of vaporization, moles/time
- λw=latent heat of liquid at wet bulb temperature Tw,
- h_y= heat transfer coefficient between gas and surface of liquid,
- Ti= temperature at interface = Tw, A= surface area of liquid

- $N_A = \frac{k_y}{(1-y)_{lm}}(y_i - y)A = \frac{k_y}{(1-y)_{lm}}\left(\frac{H_w/M_A}{1/M_B+H_w/M_A} - \frac{H/M_A}{1/M_B+H/M_A}\right)A$
- *Where,* y_i *= mole fraciton of vap. at interface, y=mole fraction of vapin air stream,* k_y *=MTC,* $(1-y)_{lm}$ *= one way diffusion factor*

$$M_A N_A\left[\lambda_w + c_{pA}(T - T_w)\right] = h_y(T - T_i)A$$

$$h_y(T - T_w)A = \frac{k_y}{(1-y)_{lm}} A\left(\frac{H_w}{1/M_B+H_w/M_A} - \frac{H}{1/M_B+H/M_A}\right)\left[\lambda_w + c_{pA}(T - T_w)\right]$$

- Assumptions:

1. $(1-y)_{lm} \cong 1$
2. Neglecting Sensible heat, $c_{pA}(T - T_w)$
3. Neglecting H_w/M_A and H/M_A in comparison to $1/M_B$

$$h_y(T - T_w) = k_y(H_w M_B - H\,M_B)[\lambda_w]$$

EQUATION OF PSYCROMETRIC LINE

$$h_y(T - T_w) = M_B k_y \lambda_w (H_w - H)$$

$$\boldsymbol{\frac{H - H_w}{T - T_w} = -\frac{h_y}{M_B k_y \lambda_w}}$$

Psychrometric line and Lewis equation

$$\frac{H - H_w}{T - T_w} = -\frac{h_y}{M_B k_y \lambda_w}$$

- Psychrometric line intersects 100% line at Tw, and has a slope of

$$-\frac{h_y}{M_B k_y \lambda_w}$$

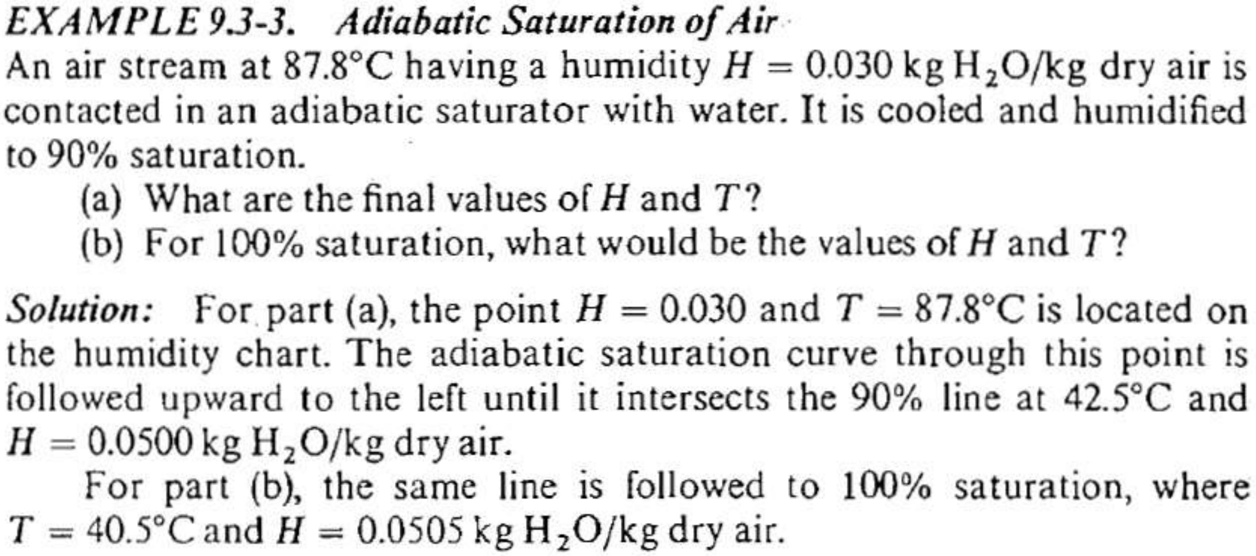
- For water – air system: $\frac{H_S - H}{T - T_S} = \frac{c_S}{\lambda_S}$

$$\frac{h_y}{M_B k_y} \cong c_S$$ - **Lewis relation**

- Comparing Psychrometric line and Adiabatic cooling line

$$\frac{H - H_w}{T - T_w} = -\frac{h_y}{M_B k_y \lambda_w} \text{ and } \frac{H_S - H}{T - T_S} = \frac{c_S}{\lambda_S}$$

- **Hence adiabatic line and psychrometric lines are identical for air- water vapour system.**



EXAMPLE 9.3-3. Adiabatic Saturation of Air

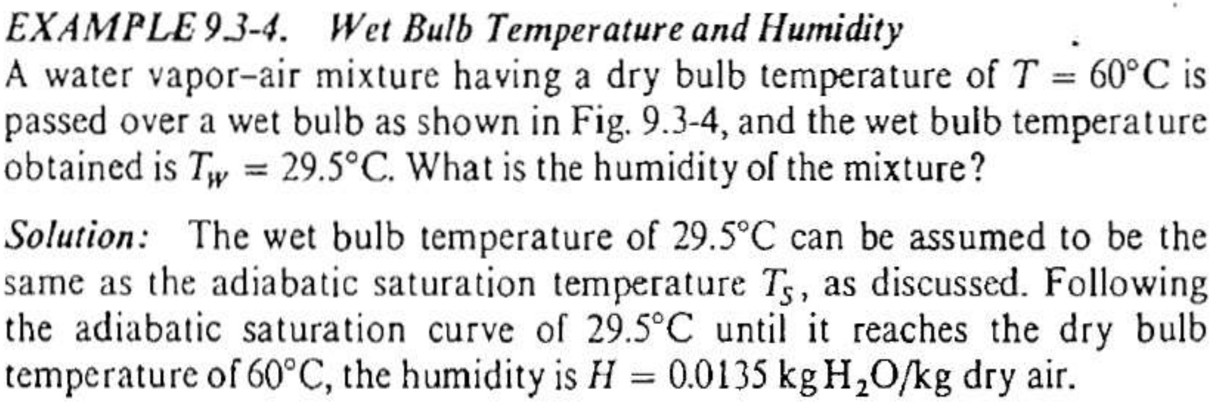
An air stream at 87.8°C having a humidity $H = 0.030$ kg H_2O/kg dry air is contacted in an adiabatic saturator with water. It is cooled and humidified to 90% saturation.

(a) What are the final values of H and T?

(b) For 100% saturation, what would be the values of H and T?

Solution: For part (a), the point $H = 0.030$ and $T = 87.8$°C is located on the humidity chart. The adiabatic saturation curve through this point is followed upward to the left until it intersects the 90% line at 42.5°C and $H = 0.0500$ kg H_2O/kg dry air.

For part (b), the same line is followed to 100% saturation, where $T = 40.5$°C and $H = 0.0505$ kg H_2O/kg dry air.



EXAMPLE 9.3-4. Wet Bulb Temperature and Humidity

A water vapor–air mixture having a dry bulb temperature of $T = 60$°C is passed over a wet bulb as shown in Fig. 9.3-4, and the wet bulb temperature obtained is $T_W = 29.5$°C. What is the humidity of the mixture?

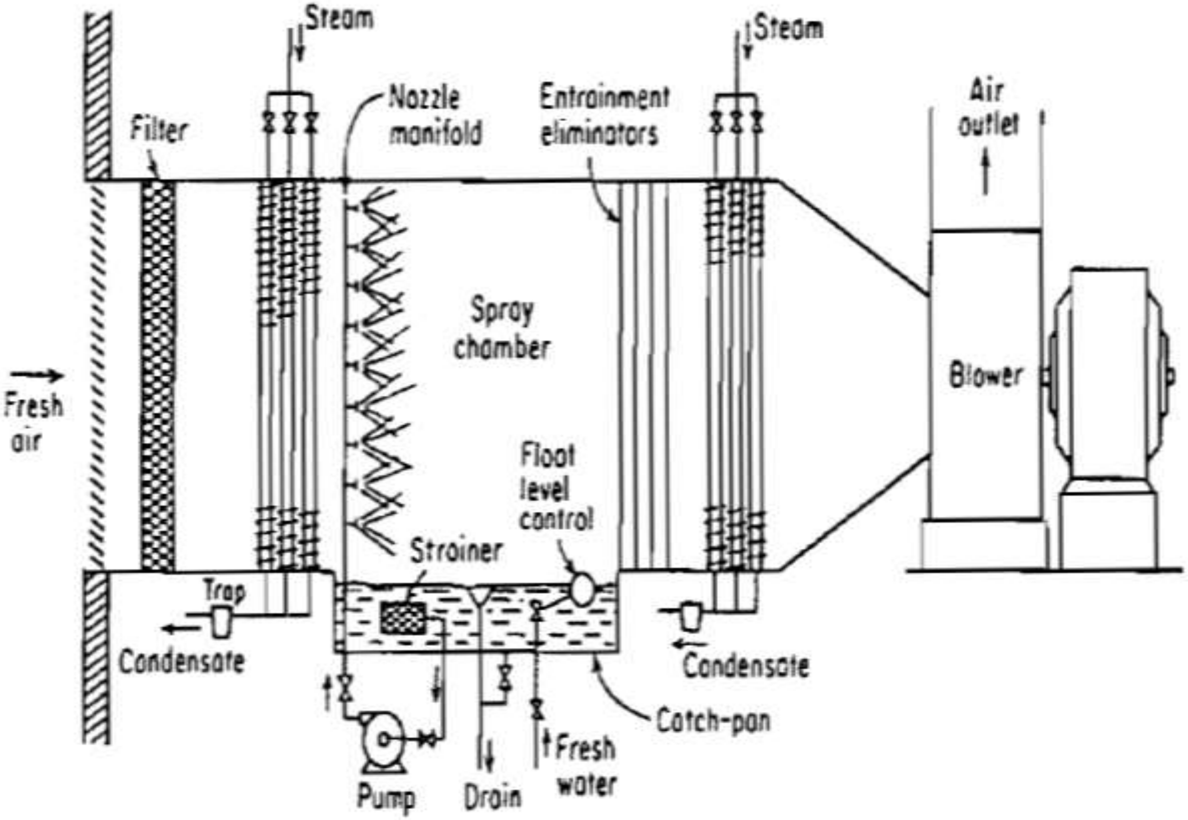
Solution: The wet bulb temperature of 29.5°C can be assumed to be the same as the adiabatic saturation temperature T_S, as discussed. Following the adiabatic saturation curve of 29.5°C until it reaches the dry bulb temperature of 60°C, the humidity is $H = 0.0135$ kg H_2O/kg dry air.

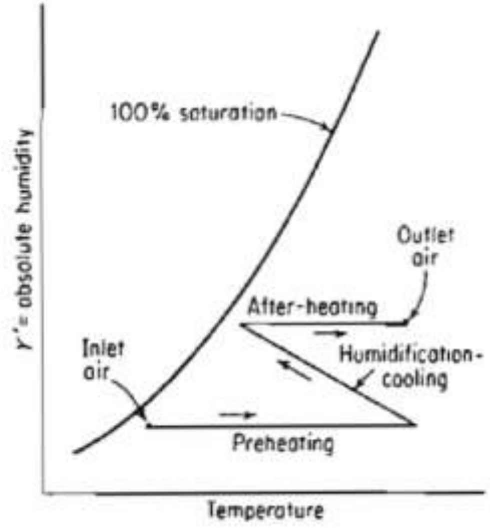
MEASUREMENT OF HUMIDITY

1. Dew point method- Polished gold disc cooled by peltier effect
2. Psychrometric method

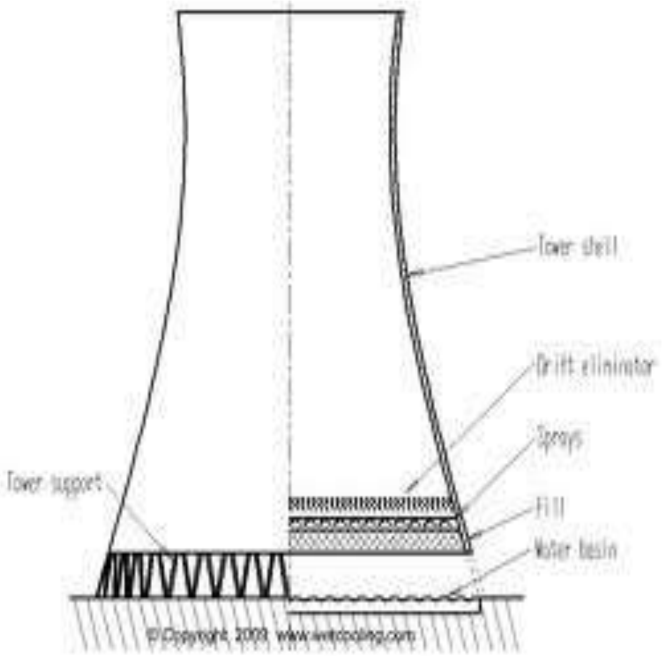
   Wet Bulb Dry bulb thermometer
3. Direct chemical method – absorption by phosphorous pentoxide, sulphuric acid
4. Hair Hygrometer- Measurement of length of hair/fibre
5. Measurement of heat of absorption on to a surface
6. Piezoelectric hygrometers- quartz crystal coated with hygroscopic material
7. Capacitance meters – electrical capacitance
8. Observation of colour change –cobaltous chloride

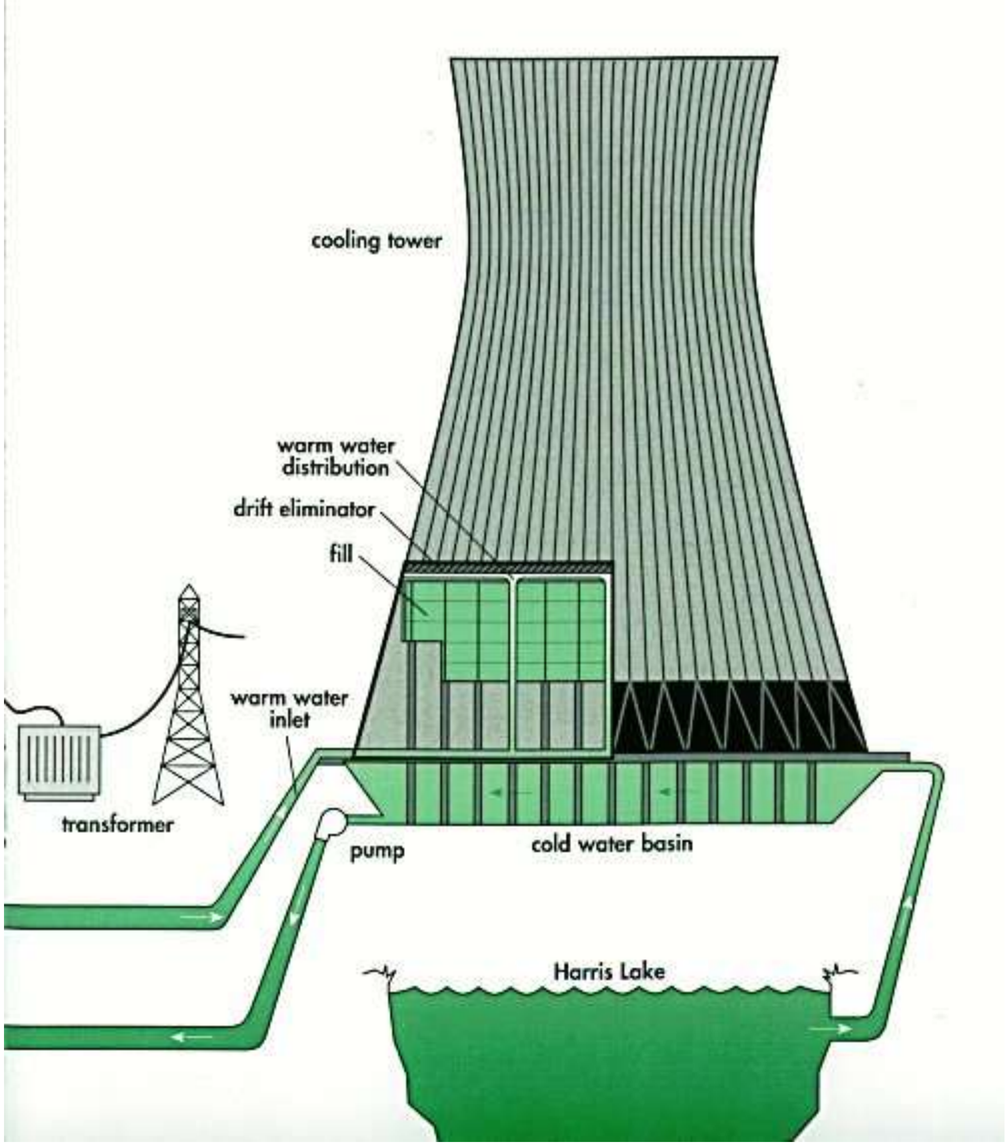
AIR CONDITIONING

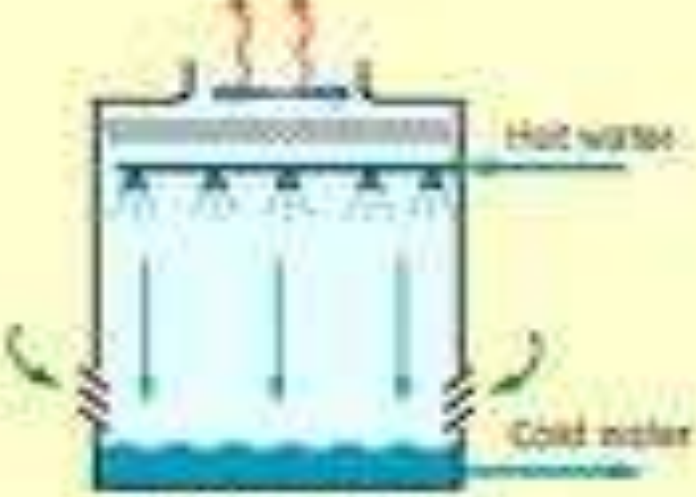




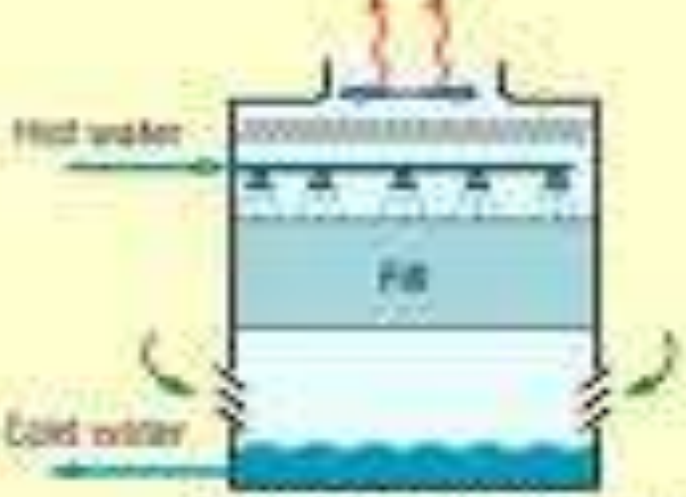
COOLING TOWERS


Tower shell
Drift eliminator
Sprays
Fill
Water basin
Tower support
©Copyright 2009 www.wetcooling.com

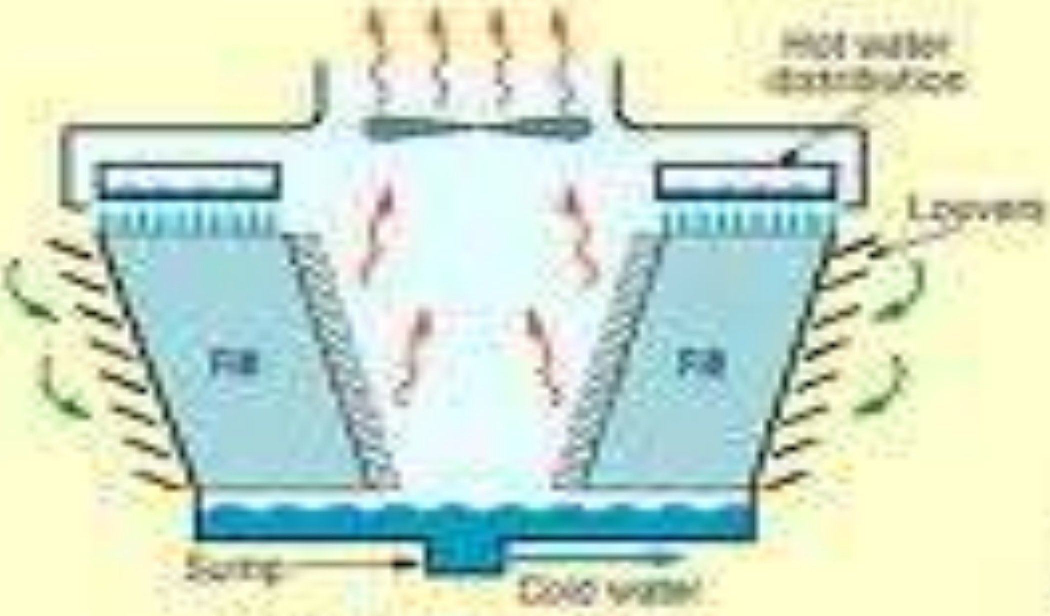

cooling tower
warm water distribution
drift eliminator
fill
warm water inlet
transformer
pump
cold water basin
Harris Lake



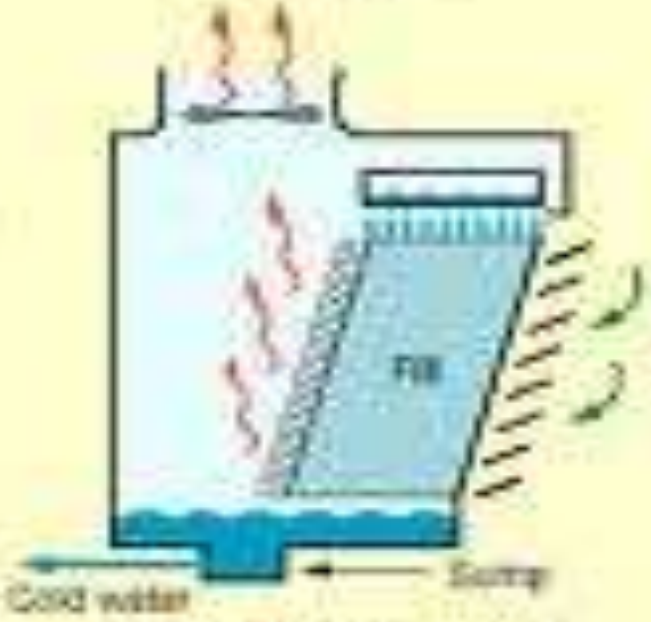
Induced draft counterflow tower



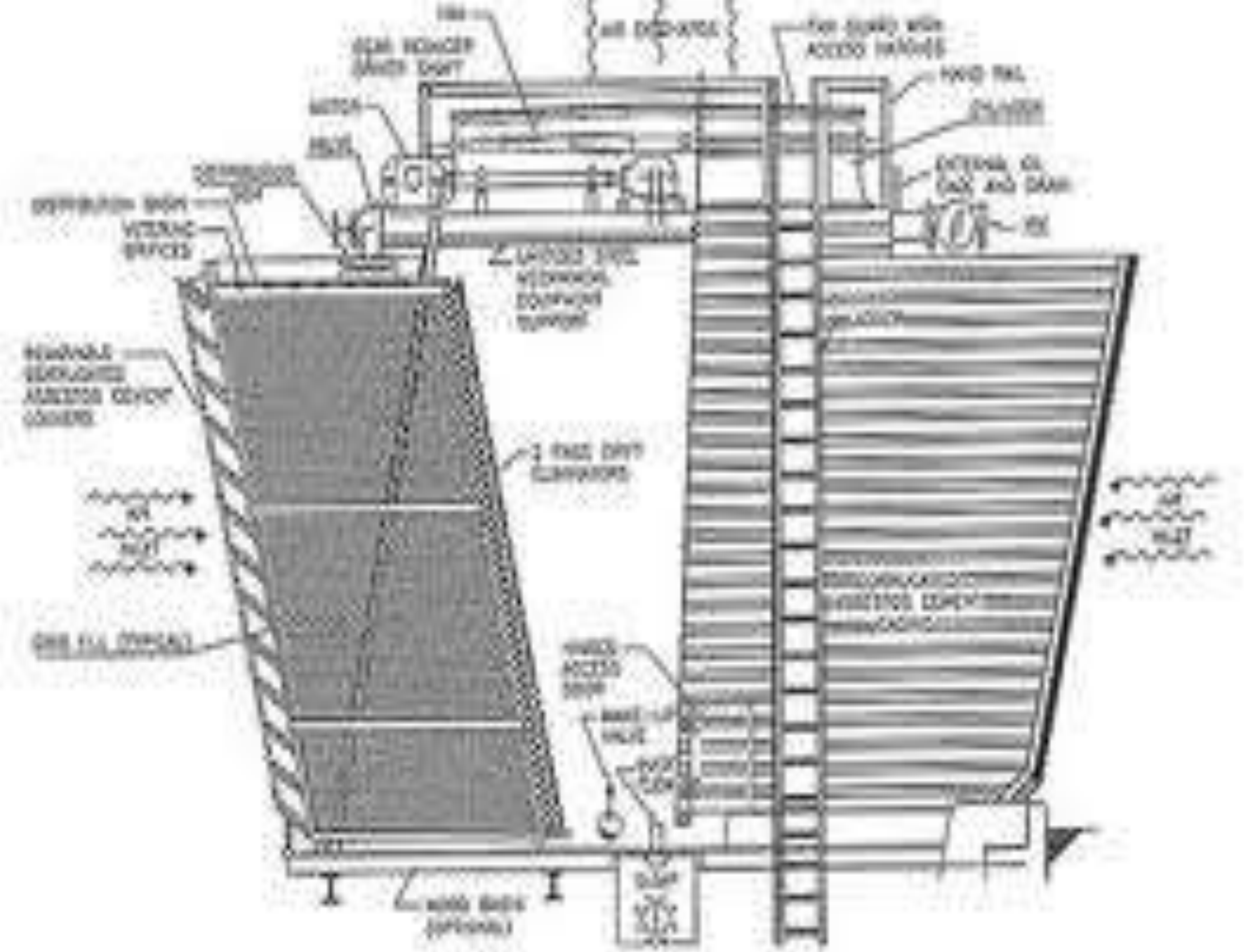
Induced draft counterflow tower with fill



Induced draft, double-flow crossflow tower



Induced draft, single-flow crossflow tower


FAN
AIR DISCHARGE
FAN GUARD WITH ACCESS HATCHES
GEAR REDUCER DRIVE SHAFT
HAND RAIL
MOTOR
VALVE
EXTERNAL OIL GAGE AND DRAIN
DISTRIBUTION BASIN
GALVANIZED STEEL MECHANICAL EQUIPMENT SUPPORT
AIR INLET
AIR INLET
MAKE-UP VALVE
SUMP


ENVIRO TECH
ENVIRO TECH.
9871034100


ENVIRO TECH.
9871034100.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Cross Corrugated Fill

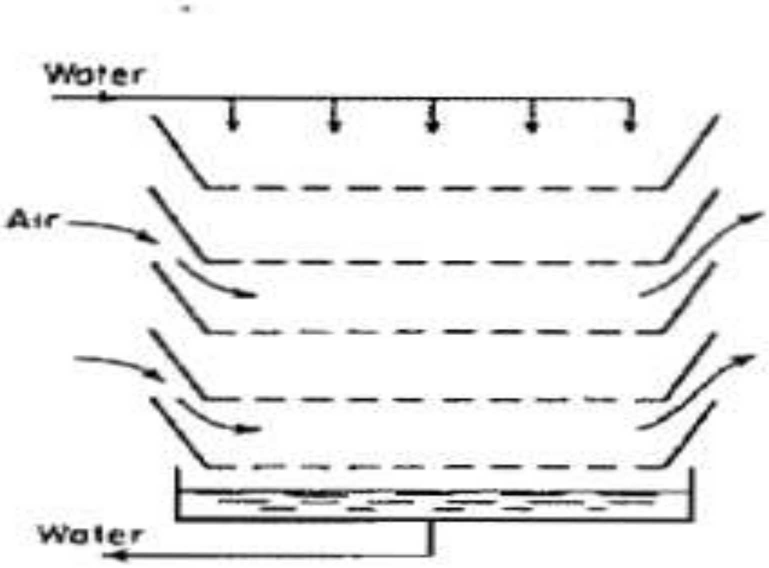
- Very efficient
- Can foul
- Max surface area

Vertical Offset Fill

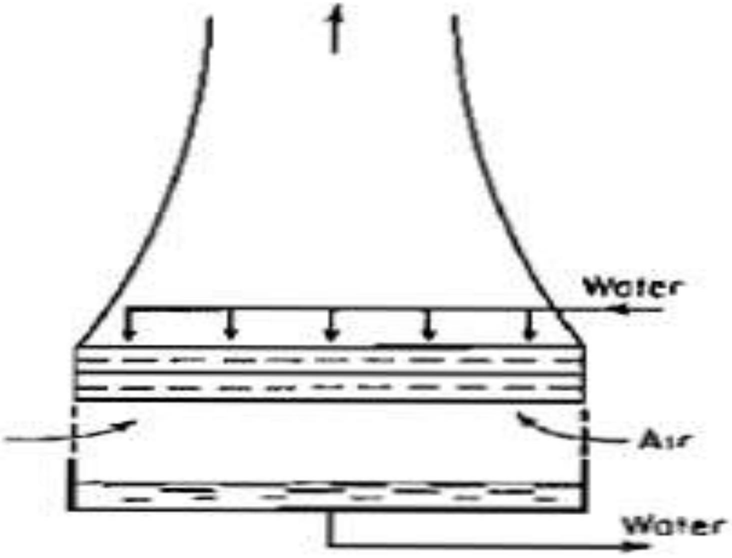
- Efficient
- Less fouling
- Max surface area

Vertical Fill

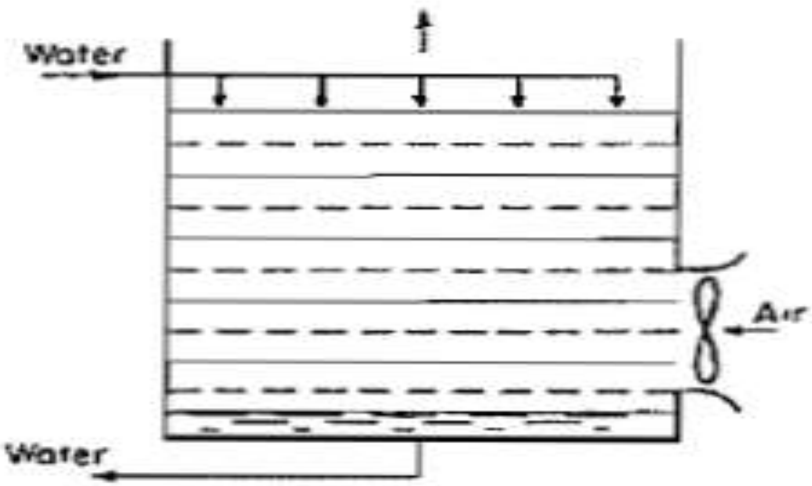
- Good for dirty water
- Low fouling
- Low pressure drop



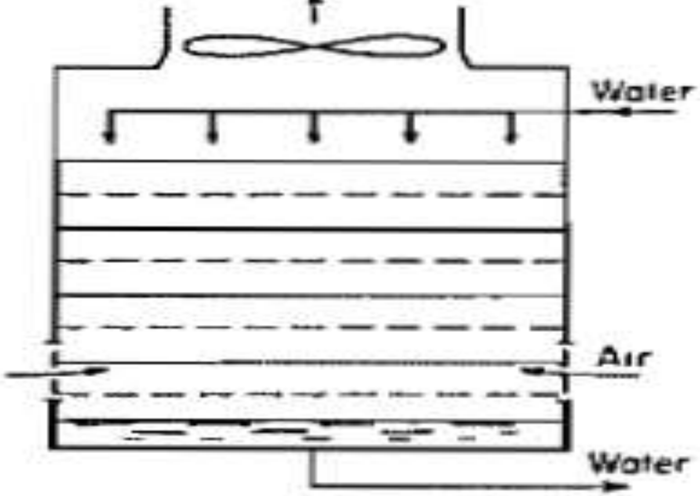
(a) Atmospheric



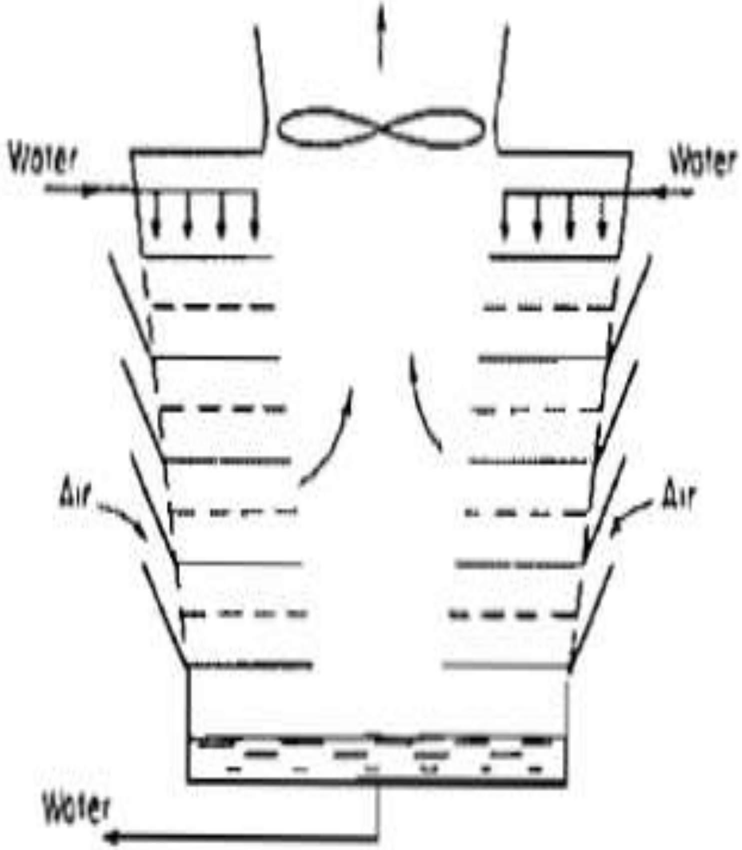
(b) Natural draft



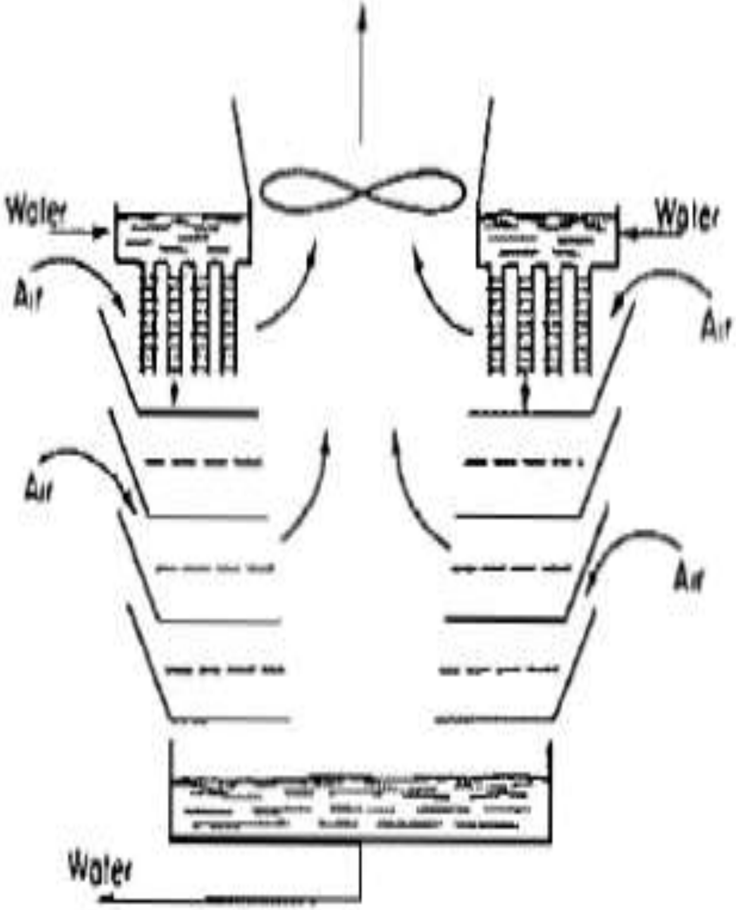
(c) Forced draft



(d) Countercurrent induced draft

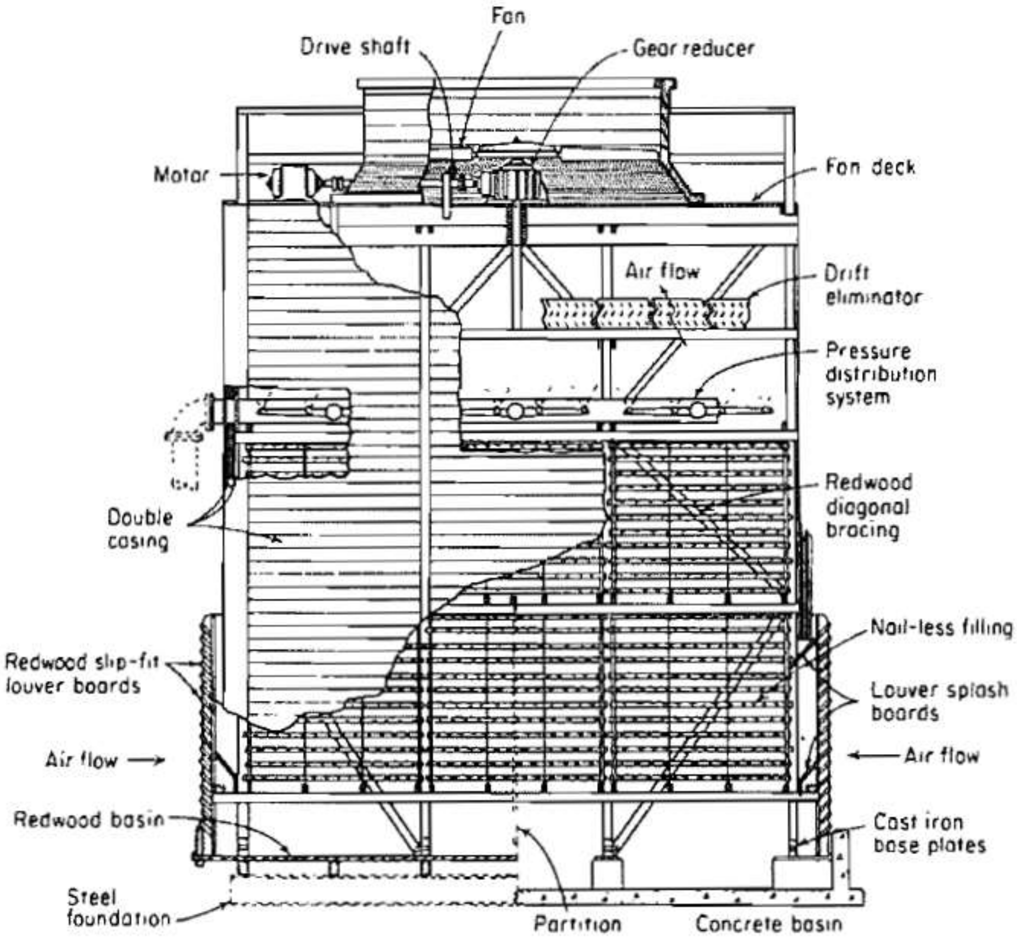


(e) Cross-flow induced draft

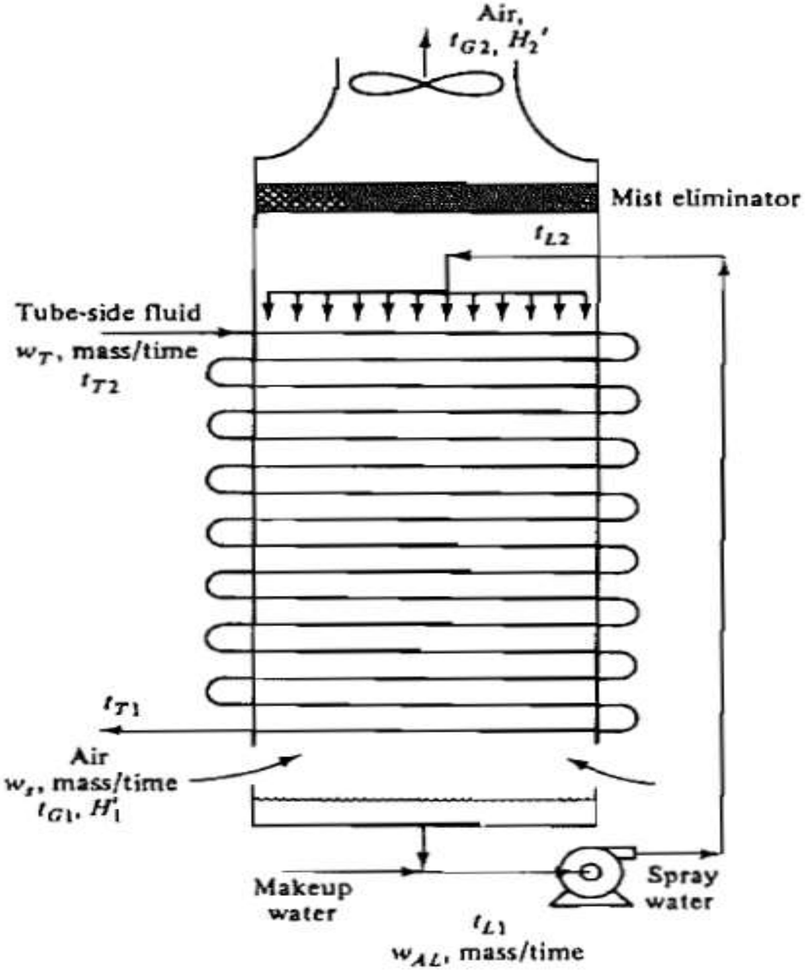


(f) Wet/dry tower

INDUCED DRAFT COOLING TOWER



Evaporative water cooler



Tube arrangement in evaporative coolers