Non-isothermal operation

Many absorbers and strippers deal with dilute gas mixtures and liquids therefore

in these cases operation is assume to be isothermal. However, absorption operations are usually exothermic, and when large quantities of solute gas are absorbed to form concentrated solutions, the temperature effects cannot be ignored.

If by absorption the temperature of the liquid is raised to considerable extent, the equilibrium solubility of the solute will be appreciably reduced and the capacity of the absorber decreased (or large flow rates of liquid will be required/cooling coils can be installed).

For stripping, an endothermic action, the temperature tends to fall.

Consider the tray tower shown in figure. If Q_T is the heat removed per unit time from the entire tower. An enthalpy balance for the entire tower is

 $L_0 H_{L0} + G_{N_p+1} H_{G, N_p+1} = L_{N_p} H_{L, N_p} + G_1 H_{G1} + Q_T$



Where H represent in each case the molal enthalpy of the stream at its particular concentration and condition.

Thus, the molal enthalpy of a liquid solution, temperature t_L , composition x mole fraction solute can be obtained

$$H_L = C_L(t_L - t_0) + \Delta H_S$$

First term on the right represents the sensible heat and the second the molal enthalpy of mixing, or integral heat of solution, at the prevailing concentration and at the base temperature t_0 , per mole of solution.

$$H_G = C_{pG, \text{inert}} (t_G - t_0)(1 - y) + y [C_{pG, \text{solute}} (t_G - t_0) + \lambda_0]$$

$$H_L = C_{pL, \text{inert}} (t_L - t_0)(1 - x) + x [C_{pL, \text{solute}} (t_L - t_0)]$$

For adiabatic operation Q_T of above equation is zero and the temperature of the streams leaving an absorber will generally be higher than the entering temperatures owing to the heat of solution.

The rise in temperature causes a decrease in solute solubility, which in turn results larger minimum L/G and a larger number of trays than isothermal operation.

The design of such absorbers may be done numerically, calculating tray by tray from the bottom to the top.

Thus, total and solute balances up to tray n, as shown in figure by the envelope are

$$L_n + G_{N_p+1} = L_{N_p} + G_{n+1}$$
$$L_n x_n + G_{N_p+1} y_{N_p+1} = L_{N_p} x_{N_p} + G_{n+1} y_{n+1}$$

From which L_n and x_n are computed. An enthalpy balance

$$L_n H_{L,n} + G_{N_p+1} H_{G,N_p+1} = L_{N_p} H_{L,N_p} + G_{n+1} H_{G,n+1}$$

- From which the temperature of stream L_n can be obtained. Stream G_n is then at the same temperature as L_n and in composition equilibrium with it.
- Above equations are then applied to tray n-1, and so forth.
- Since only the temperatures of the entering streams L_0 and G_{N_P+1} are known.
- Estimate the temperature t₁ of the gas G₁ (which is the same as the top tray temperature)
- Use overall enthalpy balance equation to compute the temperature of the liquid leaving at the bottom of the tower.
- The estimate is checked when the calculations reach the top tray, and if necessary the entire computation is repeated. ABHISMEK NUMARICH, JDRA

Real Tray and Tray Efficiency

For a given absorber or stripper, the tray efficiency as a function of fluid compositions and temperature as they vary from one end of the tower to the other.

Usually it is sufficient to make such computations at only three or four locations and then proceed as in figure. The dotted line is drawn between equilibrium curve and operating line at a fractional vertical distance from the operating line equal to the prevailing Murphree gas efficiency.

Thus the value of E_{MGE} for the bottom tray is the ratio of the lengths of lines, AB/AC.

The dotted line represents the real effluent compositions from the trays, it is used instead of the equilibrium curve to

 $\mathbf{E}_{o} =$



x

 $\log(1/A)$

complete the tray construction, which now provides the number of real trays.

The overall tray efficiency can be computed and the number of real trays can be determined analytically; $\frac{\text{equilibrium trays}}{1 + E_{MGE}(1/A - 1)}$

real trays

Continuous-Contact Equipment

Countercurrent packed and spray towers operate in a different manner from plate towers in that the fluids are in contact continuously in their path through the tower, rather than intermittently.

Thus, in a packed tower the liquid and gas compositions change continuously with height of packing.

Height Equivalent to an Equilibrium Stage (Theoretical Plate)

In the designing of packed towers, number of theoretical trays or plates required for given change in concentration is computed. This is then multiplied by a quantity, the height equivalent to a theoretical tray or plate (HETP) to give the required height of packing to do the same job.

HETP varies, not only with the type and size of the packing but also very strongly with flow rates of each fluid and for every system with concentration as well.



Absorption of one component

Consider a packed tower of unit cross-section as shown in figure.

The total effective interfacial surface of mass transfer is S per unit cross section. This is conveniently described as the product of a specific interfacial surface, surface per volume of packing, by the packed volume Z volume per unit tower cross section, or height.

In the differential volume dZ, the interface surface is

dS = a dZ

The quantity of solute A in the gas passing the differential section of the tower is Gy mol/(area)(time), and the rate of mass transfer is therefore d(Gy) mol A/(differential volume)(time).





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

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

Substituting in equation [1], rearranging and integrating give

$$Z = \int_0^Z dZ = \int_{y_2}^{y_1} \frac{G \, dy}{F_G a (1-y) \ln[(1-y_i)/(1-y)]}$$
[2]

It is more convenient to write, $y - y_i = [(1 - y_i) - (1 - y)]$ The numerator and denominator of equation [2] can be multiplied by the right and left hand sides with this and obtain

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

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

$$Z = \frac{G}{F_G a} \int_{y_2}^{y_1} \frac{(1-y)_{iM} \, dy}{(1-y)(y-y_i)} \approx H_{tG} \cdot N_{tG}$$

Where \boldsymbol{H}_{tG} is the height of a gas transfer unit and \boldsymbol{N}_{tG} is the number of gas transfer units

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

In terms of other individual mass transfer coefficients.

 N_{tG} is simplified further by substituting the arithmetic average instead of logarithmic average of (1-y)_{iM}

Hence,

$$(1-y)_{iM} = \frac{(1-y_i) - (1-y)}{\ln\left[\frac{(1-y_i)}{(1-y)}\right]} \approx \frac{(1-y_i) + (1-y)}{2}$$
$$N_{tG} = \int_{y_2}^{y_1} \frac{(1-y)_{iM}}{[(1-y)(y-y_i)]} dy = \int_{y_2}^{y_1} \frac{dy}{(y-y_i)} + \frac{1}{2} \ln\frac{(1-y_2)}{(1-y_1)}$$

Similarly, when the above mentioned relations have been applied for liquid compositions, we obtain

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

Where H_{tL} is the height of liquid transfer unit, N_{tL} is the number of liquid transfer units and $(1-x)_{iM}$ is logarithmic mean of (1-x) and $(1-x_i)$ On simplification, we get

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

And

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

All the above equation can be used to determine the height of the tower.



- With the known quantities, H_{tG} or H_{tL} can be easily determined.
- However, N_{tG} and N_{tL} can be determined only through the graphical method. For this, plot of $1/(y-y_i)$ against y is drawn and the area under the curve will give N_{tG} .
- The values of y and y_i can be evaluated by drawing a line between equilibrium curve and operating line with the slope (-k_xa/k_ya) where y and y_i points of intersection of this line on operating line and equilibrium curve respectively.

Strippers

The same relationships apply as for absorption. The driving forces $y-y_i$ and x_i-x which appear 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.



The overall transfer units

In some cases where the equilibrium curve is straight and the ratio of mass transfer coefficient is constant, it is more convenient to make use of overall mass transfer coefficient. The height of the tower can be expressed in such case as

$$Z = N_{loG} \cdot H_{loG}$$

$$N_{loG} = \int_{y_2}^{y_1} \frac{(1-y)_{\bullet M} \, dy}{(1-y)(y-y^{\bullet})} \approx \int_{y_2}^{y_1} \frac{dy}{(y-y^{\bullet})} + \frac{1}{2} \ln \frac{(1-y_2)}{(1-y_1)}$$

$$H_{loG} = \frac{G}{F_{OG}a} = \frac{G}{K_y a(1-y)_{\bullet M}} = \frac{G}{K_G a P_t (1-y)_{\bullet M}}$$

$$N_{loL} = \int_{x_2}^{x_1} \frac{dx}{(x^{\bullet}-x)} + \frac{1}{2} \ln \frac{(1-x_1)}{(1-x_2)}$$

$$H_{loL} = \frac{L}{F_{oL}a} = \frac{L}{K_x a(1-x)_{\bullet M}}$$

Dilute solution

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

$$N_{toG} = \int_{y_2}^{y_1} \frac{dy}{(y - y^*)}$$
 or $N_{toL} = \int_{x_2}^{x_1} \frac{dx}{(x^* - x)}$

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

$$y^* = mx + r$$

The operating line can be considered as a straight line as well

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

So that the driving force $y - y^*$ is then linear in x

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

Where q, r and s are constant.

Therefore equation becomes

$$N_{tOG} = \frac{L}{G} \int_{x_2}^{x_1} \frac{dx}{qx+s} = \frac{L}{Gq} \ln \frac{(y-y^*)_1}{(y-y^*)_2} = \frac{y_1 - y_2}{\frac{(y-y^*)_1 - (y-y^*)_2}{\ln[(y-y^*)_1/(y-y^*)_2]}}$$

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

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}

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

Dilute solutions, Henry's Law

If Henry's law applies [constant r of equation = 0], by elimination of x between equilibrium and operating line equation and substitution of y^* in first equation there result for absorbers

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

Where A = L/mG, as before. For strippers, the corresponding expression in terms of N_{tOL} is similar

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

These are shown in convenient graphical form in figure shown in next slide.

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 developed in mass transfer coefficient chapter. Thus equation, with

$$\frac{1}{F_{OG}} = \frac{1}{F_G} \frac{(1 - y_A)_{iM}}{(1 - y_A)_{\cdot M}} + \frac{m'(1 - x_A)_{iM}}{F_L(1 - y_A)_{\cdot M}}$$

m' = m" = m = const, ca be written

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



Whence, by definition of the heights of transfer units,

$$H_{iOG} = H_{iG} \frac{(1-y)_{iM}}{(1-y)_{*M}} + \frac{mG}{L} H_{iL} \frac{(1-x)_{iM}}{(1-y)_{*M}}$$

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

$$H_{iOG} = H_{iG} + \frac{mG}{L} H_{iL} \frac{(1-x)_{iM}}{(1-y)_{*M}}$$

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

$$\frac{1}{F_{OL}} = \frac{1}{m''F_G} \frac{(1-y_A)_{iM}}{(1-x_A)_{M}} + \frac{1}{F_L} \frac{(1-x_A)_{iM}}{(1-x_A)_{M}}$$

Yields

$$H_{IOL} = H_{IL} \frac{(1-x)_{iM}}{(1-x)_{*M}} + \frac{L}{mG} H_{IG} \frac{(1-y)_{iM}}{(1-x)_{*M}}$$

And if mass transfer resistance is essentially all in the liquid, $H_{IOL} = H_{IL} \frac{(1-x)_{iM}}{(1-x)_{*M}} + \frac{L}{mG} H_{IG} \frac{(1-y)_{iM}}{(1-x)_{*M}}$