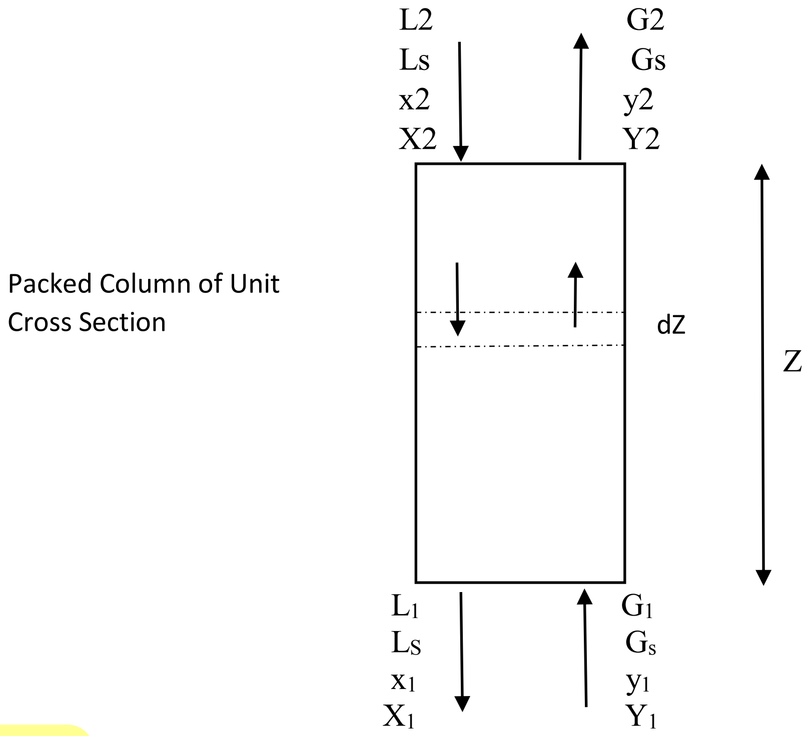
Transfer Units: Application to Gas Absorption



المطلوب هو Z

Definitions:

S: total effective interfacial surface for mass transfer per unit cross section (dimensionless)

Z: Packed volume OR packed height for a tower of unit cross section (Dimension L)

= volume of packing per unit cross section

a: Specific interfacial area = surface per unit cross section / packed volume (Dimension L^{-1})

therefore, by definition: S = a Z → dimensionless.

dZ: Differential Volume → per unit cross section

dS: Differential interfacial area

Consider the differential volume dZ:

Interfacial area is: dS = a dZ

The net flux of solute A → absorbed compound. passing through the differential surface is given by:

$$\left[\frac{\overset{\text{or mass}}{\text{Mole}}}{\text{Area}}\right] N_A = \frac{dm}{dS}; \quad N_A = \frac{dm}{a dZ}$$

m = mole of A / area / time ; m = G y (both G and y vary inside the tower)

$$dm = d(Gy)$$

$$dm = d\left(\frac{G_S y}{(1-y)}\right) \quad ; \quad dm = G_s d\left(\frac{y}{(1-y)}\right) \quad ; dm = \frac{G_S}{(1-y)^2} dy \quad ; \quad dm = \frac{G}{(1-y)} dy$$

$$\therefore \quad N_A = \frac{G dy}{(1-y)\, a\, dZ}$$

→ integration this will get Ⓩ

It can be seen that d(Gy) can be related to the mass transfer coefficients and the driving forces at this section of the column.

Molar density

$$F = \frac{C\,D}{\Delta z}$$

Diffusivity

distance

In terms of local coefficients, in the general case the flux of component A (N_A) is:

$$N_A = \frac{N_A}{\sum N} F_G \ln \frac{N_A/\sum N - y_{A,i}}{N_A/\sum N - y_{A,G}}\ \text{[redacted]} = \frac{N_A}{\sum N} F_L \ln \frac{N_A/\sum N - x_{A,L}}{N_A/\sum N - x_{A,i}}$$

Where F_G and F_L are the gas and liquid phase mass transfer coefficients for substance A. They have dimensions of : mole/L^2 time

For binary mixture the net flux $\sum N = N_A + N_B$; N_A and N_B are molar fluxes A and B

For absorption $N_B = 0$ (A diffusing through stagnant B) and hence $\frac{N_A}{\sum N} = 1$

$$Z = \overset{HTU}{\left(\frac{G}{K_G \cdot P_T a}\right)} \overset{NTU}{\frac{y_1 - y_2}{(y - y_e)_{ln}}}$$

$$(y-y_e)_{ln} = \frac{(y_1-y_{e1})-(y_2-y_{e2})}{\ln\left[\frac{(y_1-y_{e1})}{(y_2-y_{e2})}\right]}$$

$$\therefore N_A = F_G \ln\left(\frac{1-y_i}{1-y}\right)$$

G mass flux (kg/m².s)

KG => Mass transfer coefficient

PT => Total pressure

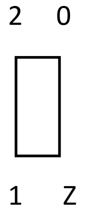
a => interfacial Area ($\frac{m^2}{m^3}$) per unit volume of packing

$$\therefore F_G \ln\left(\frac{1-y_i}{1-y}\right) = \frac{G\,dy}{(1-y)\,a\,dZ}$$

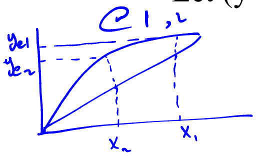
y1, y2 => mole ratio of solute at location 1 & 2

ye2 ye1 -> equilibrium mole ratio @ 1, 2

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



Let (y-y$_i$) = (1-y$_i$) – (1-y)



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

$y_{e1} = m\,x_1 + c$

$y_{e2} = m\,x_2 + c$

$$Z = \int_{y_2}^{y_1} \frac{G}{F_G\,a} \cdot \frac{(1-y)_{iM}\,dy}{\underbrace{(y-y_i)}_{\text{Driving force at any section}}(1-y)}\ ;\ (1-y)_{iM} = \log mean\ of\ (1-y_i) and\ (1-y)$$

$$\boxed{m = \frac{y_{e2}-y_{e1}}{x_2 - x_1}}$$

y_i ≡ composition at the interface

- $\frac{G}{F_G\,a}$ has units of height and can be determined experimentally. It is called the ***height of a transfer unit***.

بتعبّر عن درجة التغيّر
ودونه تركيز

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

كل تغيّر بالبسط
يعوّضه
بالمقام / [illegible]
ال ratio ثابتة

G and *(F$_G$.a)* are not constant but they seem to compensate for each other so that:

***H_{tg}* ≈ constant**

- The ratio of the terms including y is dimensionless and when integrated gives the number of transfer units to make up Z.

كم قفزة
due to [illegible]
mass transfer

$$N_{tG} = \int_{y_2}^{y_1} \frac{(1-y)_{iM}\,dy}{(y-y_i)(1-y)}$$ This integral is evaluated graphically or numerically

$$Z = H_{tG} \,.\, N_{tG}$$

Transfer Units in Terms of Liquid Concentrations:

$$Z = \int_0^Z dZ = \int_{x_2}^{x_1} \frac{L}{F_L\, a} \cdot \frac{dx}{(1-x)\ln[(1-x)/(1-x_i)]}$$

$$Z = \int_{x_2}^{x_1} \frac{L}{F_L\, a} \cdot \frac{(1-x)_{iM}\, dx}{(1-x)(x_i-x)}$$

Based on Liquid

not equal

$$\boxed{Z = H_{tL} \,.\, N_{tL}} \Rightarrow \text{this is equl to}$$

$$Z = H_{tG} \cdot N_{tG}.$$

The same relations apply for strippers.

Simplifications:

1. log mean ≅ Arithmetic Average

$$(1-y)_{iM} \cong \frac{(1-y_i)+(1-y)}{2}$$

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

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

2. For dilute gases $(1-y_2) \cong (1-y_1)$

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

3. If we consider $(y-y_i) \cong$ constant = average driving force in the column, then it can be seen that:

$$N_{tG} = \int_{y_2}^{y_1} \frac{1}{(y-y_i)_{Average}}\, dy = \left[\frac{y_1-y_2}{(y-y_i)_{Average}}\right]$$

Which describes the number of times the driving force must divide into the change in gas concentration. This is a measure of the difficulty of the mass transfer operation and in an integrated form is called the number of transfer units.

CONTINUOUS CONTACT EQUIPMENT – PACKED TOWERS:

Main features:

- Fluids are in continuous contact in their path through the column.
- Liquid and gas compositions change continuously.
- Operating line describes compositions of liquid and gas phases in all sections.

Packed tower versus plate columns

a) Advantages of packed towers:

1- Generally are smaller in size and are less expensive for small diameters (2-3m).
2- Easily constructed even in corrosion resistant materials such as glass, ceramics.
3- Give lower pressure drop → useful for vacuum operations.
4- Liquid hold up is generally lower.

b) Disadvantages:

1- Difficult to clean without complete dismantling.
2- Channeling problems, therefore liquid redistributors are usually fitted every three meters and packing size should be $\langle \frac{1}{8}$ column diameter.
3- Heavier than plate columns.
4- If plant is operated intermittently, the alternate expansion and contraction of the shell may crush the packing.

Design of packed towers

The same operating line equation as for plate towers can be used to determine stream concentrations at any height in the column and the minimum L/G ratio. The main problem with the design of packed towers is the height of the tower. Various concepts have been used:

- HETP
- HTU

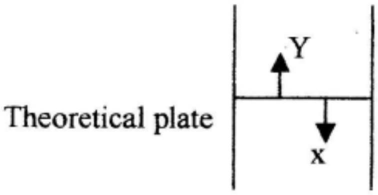
1- Height Equivalent to a Theoretical Plate (HETP)

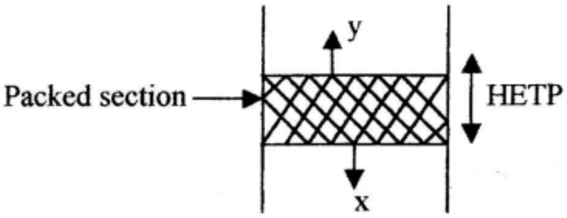
Simple method, which ignores the basic difference between the operation of plate towers and packed towers.

Height of tower = No. of theoretical plates × HETP

No. of theoretical plates same as for plate towers.

HETP: height of packing such that the liquid leaving the bottom of the section is in equilibrium with the gas leaving the top of the section.





HETP must be experimentally determined on pilot plants.

It varies with:

- type and size of packing
- flow rates of each fluid (very strong dependence)
- system and concentration

For these reasons this method is not recommended.

2- Transfer units

Z = height of transfer unit × no. of transfer units

height of transfer unit	no. of transfer units
Determined from flow rates and transfer coefficients	from concentration changes in the column. (this term describes the ease or difficulty of the operation)

Starting with a differential material balance over a differential volume, together with Ficks law, one can obtain the height of the column by the integration of the resultant equation.