

Section - 1

① The driving force in the mole ratio unit, γ , between two point 1 and 2 can be written as,

$$\begin{aligned}\gamma_{A1} - \gamma_{A2} &= \frac{P_{A1}}{P - P_{A1}} - \frac{P_{A2}}{P - P_{A2}} \\ &= \frac{PP_{A1} - P_{A1}P_{A2} - PP_{A2} + P_{A1}P_{A2}}{(P - P_{A1})(P - P_{A2})} \\ &= \frac{P(P_{A1} - P_{A2})}{P_{B1} P_{B2}}\end{aligned}$$

Since, $P_{B1} = P - P_{A1}$, $P_{B2} = P - P_{A2}$

$$P_{A1} - P_{A2} = (\gamma_{A1} - \gamma_{A2}) \frac{P_{B1} P_{B2}}{P}$$

Now, we have

$$\begin{aligned}N_A &= k_{G_1} (P_{A1} - P_{A2}) = \frac{D_{AB} P}{RT \delta P_{BLN}} (P_{A1} - P_{A2}) \\ &= \frac{D_{AB} P}{RT \delta P_{BLN}} (\gamma_{A1} - \gamma_{A2}) \left(\frac{P_{B1} P_{B2}}{P} \right) \\ &= k_y (\gamma_{A1} - \gamma_{A2})\end{aligned}$$

$$\therefore k_y = \frac{D_{AB} P_{B1} P_{B2}}{RT \delta P_{BLN}} = k_{G_1} \left(\frac{P_{B1} P_{B2}}{P} \right)$$

$$\therefore k_y = k_{G_1} \left(\frac{P_{B1} P_{B2}}{P} \right)$$

A similar relation exists between k'_G and k'_y (in the case of equimolar countercurrent diffusion) applies

② In this case, diffusion of A (alcohol vapour) occurs through non-diffusing air (B)

a) Given, $P = 1.2 \text{ bar} = \frac{1.2}{1.013} = 1.185 \text{ atm}$
 $= 900.3 \text{ mm of Hg}$

$$R = 0.08317 \text{ m}^3 \text{ bar/kmol-K}$$

$$T = 300 \text{ K}$$

(i) WRT, $k_y = P k_G = 2.4 \times 10^{-6} \frac{\text{kmol}}{\text{m}^2 \text{s mmHg}} \times 900.3 \text{ mm Hg}$
 $\therefore k_y = 2.161 \times 10^{-3} \text{ kmol/sm}^2 \Delta y$

(ii) At the surface of drop,

$$P_{A1} = \text{vapour pressure of alcohol at } 300\text{K} = 0.0877 \text{ bar}$$

$$\text{i.e., } P_{B1} = P - P_{A1} = 1.2 - 0.0877$$

$$= 1.1123 \text{ bar} = 834.5 \text{ mm Hg}$$

$$k_y = k_G \frac{P_{B1} P_{B2}}{P} = 2.4 \times 10^{-6} \times \frac{834.5 \times 900.3}{900.3}$$

$$\therefore k_y = 2.003 \times 10^{-3} \text{ kmol/sm}^2 \Delta y$$

(iii) $k_e = k_G RT = 2.4 \times 10^{-6} \times \frac{760}{1.013} \times 0.08317 \times 300 \text{ m/s}$
 $\Rightarrow k_e = 0.0449 \text{ m/s}$

$$F_G = k_G P_{BLM} = 2.4 \times 10^{-6} \frac{P_{B1} - P_{B2}}{\ln(P_{B1}/P_{B2})}$$

$$= 2.4 \times 10^{-6} \times \frac{900.3 - 834.5}{\ln(\frac{900.3}{834.5})}$$

$$\therefore F_G = 2.081 \times 10^{-3} \text{ kmol/m}^2 \text{s}$$

$$\begin{aligned}
 b) \text{ i) } k_{G_1} &= 2.4 \times 10^{-6} \text{ kmol/m}^2\text{s (mm Hg)} \\
 &= 2.4 \times 10^{-6} \frac{\text{kmol}}{\text{m}^2\text{s mm Hg}} \times \frac{2.046 \text{ lb mol}}{\text{kmol}} \times \frac{1 \text{ m}^2}{10.464 \text{ ft}^2} \times \frac{51.7 \text{ mm Hg}}{\text{psi}}
 \end{aligned}$$

$$\therefore k_{G_1} = 2.541 \times 10^{-5} \text{ lb mol/s ft}^2 \text{ psi}$$

$$\text{ii) Similarly, } k_{G_1} = 1.345 \text{ lb mol/ft}^2\text{h atm}$$

$$\text{iii) Similarly, } k_{G_1} = 4.55 \times 10^{-2} \text{ lb mol/ft}^2\text{h inch Hg}$$

Thickness of stagnant film:

$$\delta = \frac{D_{AB} P}{k_{G_1} R T \ln P_{BH}}$$

The diffusivity value is given at 1 atm and 0°C. At 1.2 bar and 300 K, the diffusivity is,

$$\begin{aligned}
 D_{AB} &= 0.102 \left(\frac{1.013}{1.185} \right) \left(\frac{300}{275} \right)^{1.75} \\
 &= 0.103 \text{ cm}^2/\text{s} \\
 &= 1.03 \times 10^{-5} \text{ m}^2/\text{s}
 \end{aligned}$$

$$P_{BH} = \frac{P_{B2} - P_{B1}}{\ln(P_{B2}/P_{B1})} = \frac{1.2 - 1.1123}{\ln(1.2/1.1123)} = 1.1556 \text{ bar}$$

$$\begin{aligned}
 k_{G_1} &= 2.4 \times 10^{-6} \text{ kmol/m}^2\text{s (mm Hg)} \\
 &= 2.4 \times 10^{-6} \times \frac{760}{1.013} = 1.8 \times 10^{-3} \text{ kmol/m}^2\text{s bar}
 \end{aligned}$$

$$\begin{aligned}
 \therefore \delta &= \frac{1.03 \times 10^{-5}}{1.8 \times 10^{-3} \times 0.08317 \times 300} \times \frac{1}{1.1556} \\
 &= 2.4 \times 10^{-4} \text{ m}
 \end{aligned}$$

$$\therefore \delta = 0.24 \text{ mm}$$

- ③ Naphthalene vapour (A) is transported through non-diffusing air (B). The mass transfer coefficient can be obtained from the given relation,

$$Sh = 2 + 0.6 (Re)^{0.5} (Sc)^{0.33}$$

The Re and thus Sh depend upon radius of the ball, even the H.T.C depends on radius of ball which decreases with time. If r is the radius of naphthalene ball at any time t,

$$Re = \frac{2r \rho v}{\eta} = r \frac{(0)(1.1)(5)}{1.92 \times 10^{-5}}$$

$$\therefore Re = 5.73 \times 10^5 r$$

The maximum value of Re = 2864

$$Sc = \frac{L}{\rho D_{AB}} = \frac{1.92 \times 10^5}{1.1 \times 6.92 \times 10^{-6}} = 2.522$$

$$Sh = 2 + 0.6(Re)^{0.5} (Sc)^{0.33} \\ = 2 + 618 \sqrt{r} - ①$$

Given, $T = 45^\circ C = 318 K$, since v.p of naphthalene is small,
 $P/P_{BH} \approx 1$

$$\text{Then, } Sh = \frac{k_A P_{BH} RT 2r}{\rho D_{AB}} = \frac{k_A (0.08317)(318)(2r)}{6.92 \times 10^{-6}} \\ \therefore Sh = 7.644 \times 10^6 r k_A - ②$$

From eqn ①, ②,

$$2 + 618 \sqrt{r} = 7.644 \times 10^6 r k_A$$

$$\therefore k_A = \frac{2.616 \times 10^{-7}}{r} + \frac{8.085 \times 10^{-5}}{\sqrt{r}}$$

Now we have to relate the rate of change of size of ball with mass transfer rate. At time t , rate of sublimation can be expressed as

$$-\frac{d}{dt} \left(\frac{4}{3} \pi r^3 \frac{\rho}{M} \right) = (4\pi r^2) k_G (P_{Av} - P_{Ab})$$

$$\text{i.e. } -\frac{dr}{dt} = \left(\frac{M}{\rho} \right) k_G P_{Av}$$

$\rho \rightarrow$ Density of solid naphthalene at $45^\circ C = 1140 \text{ kg/m}^3$

$M \rightarrow$ Mol. wt = 128

$$P_{Av} \rightarrow \text{V.P of naphthalene at } 45^\circ C = 0.8654 \text{ mm Hg} \\ = 0.001153 \text{ bar}$$

$P_{Ab} \rightarrow$ P.P of naphthalene in bulk air = 0

$$-\frac{dr}{dt} = \left(\frac{128}{1140} \right) \left(\frac{2 \cdot 616 \times 10^{-7}}{r} + \frac{8 \cdot 085 \times 10^{-5}}{\sqrt{r}} \right) (0.001153) \\ = \frac{3.3867 \times 10^{-11}}{r} + \frac{1.0967 \times 10^{-8}}{\sqrt{r}}$$

The above eqn is integrated from $r_i = 0.5 \text{ cm}$ to $r_{i/2} = 0.25 \text{ cm}$ to find req. time,

$$t = - \int_{0.005}^{0.0025} \frac{dr}{\frac{3.3867 \times 10^{-11}}{r} + \frac{1.0967 \times 10^{-8}}{\sqrt{r}}} = - \int_{0.005}^{0.0025} \frac{r dr}{3.3867 \times 10^{-11} + 1.0967 \times 10^{-8} \sqrt{r}}$$

The integral form, $I = \int \frac{r dr}{a + b\sqrt{r}}$; using $\sqrt{r} = x$
 $\Rightarrow dr = 2x dx$

$$I = 2 \int \frac{x^3 dx}{a + b^2 x^2}, \quad \xi_1 = \alpha + \beta x \Rightarrow x = \frac{\xi_1 - \alpha}{\beta}, \quad dx = \frac{d\xi_1}{\beta}$$

$$I = \frac{2}{\beta^4} \left[\frac{\xi_1^3}{3} - \frac{3 \xi_1^2 \alpha}{2} + 3 \xi_1 \alpha^2 - \alpha^3 \ln \xi_1 \right]$$

$$\text{Putting the limits, } I = 13820 \text{ s} \\ = 3.84 \text{ hr}$$

- ④ Let at any time t , the mass of undissolved solid in the vessel be m kg. Mass of solid already dissolved = $20 - m$
 Mass of soln. = $500 + (20 - m)$, Volume = $(520 - m)/1000 \text{ m}^3$
 Inst. conc. of soln., $C = (20 - m)/(520 - m)/1000 \text{ kg/m}^3$
 Solubility, $C_s = 80 \text{ kg/m}^3$

Now we express the quantities k_L and area a (total area of particles at same time, t) in terms of m .

$$d_0 = 0.6 \text{ mm} \quad (\text{Initial dia of a particle})$$

$$\text{No. of particles} = 6.574 \times 10^7$$

$$\text{Mass of a single particle at time } t = m/6.574 \times 10^7$$

$$\text{Inst. diameter } (dp) = 2.211 \times 10^{-4} (m)^{1/3}$$

$$\text{Area of single particle} = \pi dp^2 = (2.211 \times 10^{-4} m^{1/3})^2 \pi$$

$$\text{Total area of particles} = (10.1) m^{2/3}$$

$$Sc = \frac{\mu}{\rho D} = 653$$

$$\text{Inst. Re number, } Re_s = Ndp \frac{\rho}{\mu} = 0.09976 \text{ m}^{2/3}$$

$$Sh = 2 + 0.44 (0.09976 \text{ m}^{2/3})^{0.504} (653)^{0.385}$$

$$= 2 + 1.67 \text{ m}^{0.336} = \frac{k_L dp}{D}$$

$$\therefore k_L = (6.784 \times 10^{-6}) (2 + 1.67 \text{ m}^{0.336}) \text{ m}^{-1/3}$$

$$\text{The rate of dissolution of crystall, } -\frac{dm}{dt} = k_L a (C_s - C)$$

$$k_L a (C_s - C) = [6.784 \times 10^{-6} (2 + 1.67 \text{ m}^{0.336}) \text{ m}^{-1/3}] (10.1 \text{ m}^{2/3})$$

$$\left(80 - \frac{(20 - m)(1000)}{520 - m} \right) = \phi(m)$$

$$-\frac{dm}{dt} = \phi(m) \Rightarrow -\int_{20}^0 \frac{dm}{\phi(m)} = \int_0^t dt$$

on solving the integral, $t = 15 \text{ min.}$

⑤ A liquid jet has a rod-like flow; it has a flat velocity profile. In other words, the velocity is uniform over a cross-section of the jet.

Let us use the following notation:

$r \rightarrow$ radius of the jet

$v \rightarrow$ uniform velocity of jet

$Q \rightarrow$ liquid flow rate ($= \pi r^2 v$) = $1.32 \times 10^{-5} \text{ m}^3/\text{s}$

$L \rightarrow$ length of the jet = 0.05 m

$t_c \rightarrow$ contact time = $4v$

Molecular wt. of $\text{H}_2\text{S} = 34$

The solute gas gets dissolved at the surface of the jet and penetrates into it by unsteady state diffusion. The situation fits in the penetration theory. Accordingly, the avg. MTC and the gas-liquid area of contact area,

$$k_{L,av} = 2 \sqrt{\frac{D_{AB}}{\pi t_c}} = 2 \sqrt{\frac{D_{AB} v}{\pi L}} ; \text{ area of contact} = 2\pi r L \quad \text{---(1)}$$

$$\text{Rate of absorption of } \text{H}_2\text{S} = 2\pi r L (k_{L,av}) (C_{Ai} - C_{Ab}) \quad \text{---(2)}$$

$C_{Ab} \rightarrow$ H_2S conc. in the bulk of jet

$C_{Ai} \rightarrow$ Interfacial conc. of H_2S

Using Henry's law, $C_{Ai} = 1.03 \times 0.1136 = 0.117 \text{ kmol/m}^3$

$C_{Ab} = 0$ (Because of small contact time, the depth of penetration of solute in the jet will be small and the bulk liquid free from H_2S)

$$\begin{aligned} \text{The given rate of absorption of } \text{H}_2\text{S} &= 4.42 \times 10^{-9} \text{ g/s} \\ &= 1.3 \times 10^{-8} \text{ kmol/s} \end{aligned}$$

Using (1), (2) and given rate of absorption,

$$1.3 \times 10^{-8} = 2\pi L \sqrt{\frac{Q}{\pi v}} 2 \sqrt{\frac{D_{AB} v}{\pi L}} (0.117 - 0) = 4 \sqrt{D_{AB} (1.32 \times 10^{-5})(0.05)(1.17)}$$

$$\therefore D_{AB} = 1.77 \times 10^{-9} \text{ m}^2/\text{s}$$

⑥ contact time of a liquid element with a gas bubble, $t_c = d_b/v_b = 0.05s$

$$\text{Mass transfer co-eff, } k_L = 2 \left(\frac{D_{AB}}{\pi t_c} \right)^{1/2} = 0.0236 \text{ cm/s}$$

$$\begin{aligned} \text{Residence time of a single bubble in the liq} &= \text{liq depth / bubble ris rel.} \\ &= 0.3/0.2 = 1.5s \end{aligned}$$

$$\text{Vol. of a bubble} = 0.5236 \text{ cm}^3$$

$$\text{Area of a bubble} = 3.1416 \text{ cm}^2$$

$$\begin{aligned} \text{Solubility of CO}_2 \text{ in water (from Henry's law)} &= 1.493 \times 10^{-5} \text{ gmol/cm}^3 \\ (\text{P}_{\text{CO}_2} \text{ is approximately taken as } 0.5 \text{ atm}) \end{aligned}$$

$$\text{Conc. of CO}_2 \text{ in the bulk water, } C_b = 0$$

$$\begin{aligned} \text{Amount of CO}_2 \text{ absorbed from a single bubble during its residence} \\ \text{time of } 1.5s &= (k_L) (\text{area of bubble}) (C_s - C_b) (1.5s) \\ &= 1.66 \times 10^{-6} \text{ gmol} \end{aligned}$$

$$\text{Avg. no. of bubbles formed per min} = 15/0.5236 = 28.65 \text{ min}^{-1}$$

$$\begin{aligned} \text{Rate of absorption of CO}_2 \text{ from bubbles per min} &= 1.66 \times 10^{-6} \times 28.65 \\ &= 4.757 \times 10^{-5} \text{ gmol/min} \end{aligned}$$

⑦ The solubility of CO₂ at the experimental temp. is given by Henry's law, P = 1640 x*

$$\text{At } P = 2 \text{ atm}, x^* = 2/1640 = 0.00122$$

$$\begin{aligned} \text{Mol. wt. of soln.} &= (44)(0.00122) + 18(1 - 0.00122) \\ &\approx 18.03 \approx 18 \end{aligned}$$

$$\begin{aligned} \text{Moles of soln. per m}^3 &= (99.7 \text{ kg/m}^3) (18 \text{ kg/mol}) \\ &= 55.4 \text{ kmol/m}^3 \end{aligned}$$

$$\begin{aligned} \text{Moles of CO}_2 \text{ per m}^3 \text{ soln.}, C_s &= 55.4 \times 0.00122 \\ &\approx 0.0676 \text{ kmol/m}^3 \end{aligned}$$

Conc. of carbonated CO₂, leaving the vessel,

$$C = 2.3 \text{ g CO}_2/\text{L} = 0.0323 \text{ kmol/m}^3$$

This is also the conc. of CO₂ in liq. in the vessel (since it's well mixed)

$$\begin{aligned} \text{Vol. of rate of input of water} &= \text{Vol. rate of output of soln.} \\ &= 1 \text{ L/min} = 1.667 \times 10^{-5} \text{ m}^3/\text{s} \end{aligned}$$

The inlet water is CO_2 -free. So, at steady state rate of absorption of CO_2 in the vessel = $1.667 \times 10^{-5} \times 0.0523$
 $= 8.718 \times 10^{-7} \text{ kmol/s}$

$V \rightarrow$ volume of gas-liqu. dispersion in the vessel

$$V = 8 \text{ L} = 0.008 \text{ m}^3$$

$\bar{a} \Rightarrow$ specific interfacial area of contact b/w the dispersed gas and the liqu. ($= 80 \text{ m}^2/\text{m}^3$)

$k_L \rightarrow$ Mass Transfer Co-efficient

$$\begin{aligned} \text{Rate of absorption of } \text{CO}_2 \text{ at steady state} &= V \bar{a} k_L (C_s - C) \\ &= 8.718 \times 10^{-7} \text{ kmol/s} \\ \Rightarrow (0.008)(80)(k_L)(0.0676 - 0.0523) &= 8.718 \times 10^{-7} \\ \text{when, } k_L &= 8.903 \times 10^{-5} \text{ m/s} \end{aligned}$$

a) If film theory is applicable, film thickness,

$$\delta = \frac{D_{AB}}{k_L} = \frac{1.92 \times 10^{-9}}{8.903 \times 10^{-5}} = 0.0216 \text{ mm}$$

b) If the penetration theory is applicable, the contact time,

$$t_c = \frac{4 D_{AB}}{\pi k_L^2} = \frac{4 \times 1.92 \times 10^{-9}}{\pi \times (8.903 \times 10^{-5})^2} = 0.308 \text{ s}$$

c) If surface renewal theory is applicable, the fractional rate of surface renewal,

$$\delta = \frac{k_L^2}{D_{AB}} = \frac{(8.903 \times 10^{-5})^2}{1.92 \times 10^{-9}} = 4.13 \text{ s}^{-1}$$

This means that the surface is renewed 4.13 times per second on avg.

⑧ $c_b \rightarrow$ conc. of benzoic acid in bulk liq.

$c_{b_1} = 0$ (At the inlet valve)

$c_s \rightarrow$ conc. of liquid in contact with surface of sphere $= 3.01 \text{ kg/m}^3$

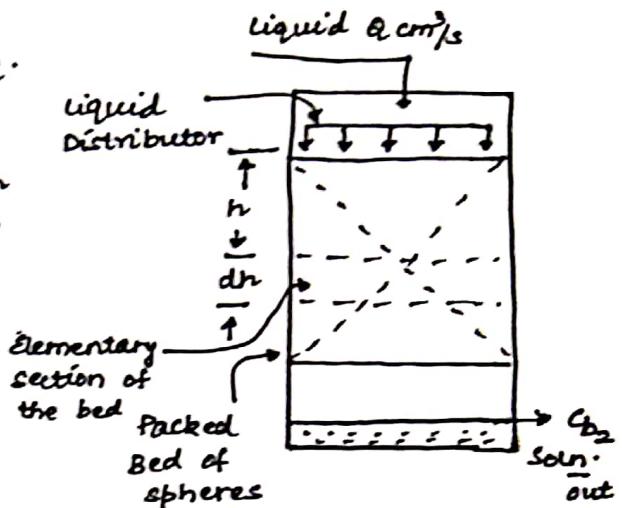
Driving force along the bed will be

$$c_s - c_b$$

$Q \rightarrow$ Volumetric Flow rate

$a \rightarrow$ cross sectional area of bed

$$Q/a = 2.2 \text{ cm/s}$$



~ volumetric H.T.C

$$Q \frac{dc_b}{dh} = (k_L \bar{a}) (Adh) (c_s - c_b)$$

$$\int_{c_{b_1}}^{c_{b_2}} \frac{dc_b}{c_s - c_b} = k_L \bar{a} \frac{a}{Q} \int_0^h dh \Rightarrow \ln \frac{c_s - c_{b_1}}{c_s - c_{b_2}} = k_L \bar{a} \frac{a}{Q} h$$

(1)

Diameter of sphere, $d_p = 0.8 \text{ cm}$

Kinematic visc., $\nu = 0.0095 \text{ cm}^2/\text{s}$

Superficial liq vel, $v_0 = 2.2 \text{ cm/s}$

$$D_{AB} = 10^{-5} \text{ cm}^2/\text{s}$$

Bed voidage, $\epsilon = 0.4$

Using the relation, $E_{fD} = 0.25 Re^{-0.31}$

$$Re = \frac{dp v_0}{\nu} = 185$$

$$Sc = \frac{\nu}{D_{AB}} = 950$$

$$E_{fD} = \epsilon \frac{Sh}{Re Sc^{1/3}} = 0.25 (185)^{-0.31} = 0.0495$$

$$\therefore Sh = 225$$

$$k_L = Sh \left(\frac{D_{AB}}{dp} \right) = 225 \left(\frac{10^{-5}}{0.8} \right) = 0.00281 \text{ cm/s}$$

$$\text{superficial area of contact, } \bar{a} = \frac{6(1-\epsilon)}{dp} = 4.5 \text{ cm}^2/\text{cm}^3$$

Substituting these values in eqn ①,

$$c_{b_2} = 1 \text{ kg/m}^3$$

Q) Given,

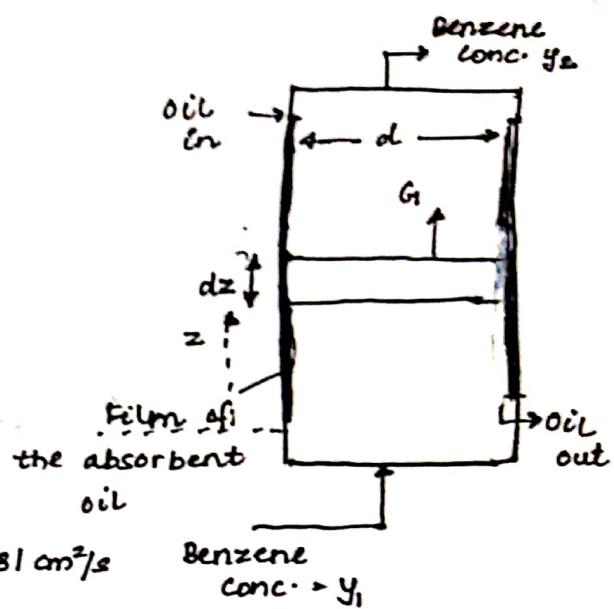
$$i.d \text{ of column} = 3.5 \text{ cm}$$

$$h \text{ of wetted section} = 3 \text{ m}$$

$$\text{Visc. of } N_2 \text{ (At } 30^\circ\text{C)} = 0.018 \text{ cP}$$

$$\text{Visc. of Air (At } 25^\circ\text{C)} = 0.0183 \text{ cP}$$

$$\begin{aligned} \text{Diffusivity of Benzene in } N_2 \text{ at } 30^\circ\text{C} \\ = 0.0973 \text{ cm}^2/\text{s} \end{aligned}$$



$$\text{Diffusivity of } NH_3 \text{ in air at } 25^\circ\text{C} = 0.281 \text{ cm}^2/\text{s}$$

$$\text{Benzene conc. } y_1$$

Since the conc. of benzene in the gas is small, the change of flow rate over the column remains small. So calculations can be done on the basis of avg. gas flow.
Column factor (j_p) varies as $Re^{-0.23}$

Consider an elementary section of column of thickness dz at height z from bottom as shown.

y → Local conc. of benzene in bulk gas

y_i → Interfacial conc. at the same section

G_1 → Molar gas flow rate

d → diameter of the column.

$$\text{The elementary gas liquid contact} = (\pi d)(dz)$$

If the gas conc. changes by dy over the section, a differential mass balance can be written by,

$$-G_1 dy = (\pi d)(dz) k_y (y - y_i)$$

Assuming that the gas flow remains reasonably constant over the column. Integrating from $z=0$, $y=y_i = 0.02$ to $z=L=3\text{m}$,

$$y = y_2 = 0.0052 \text{ (Top)}$$

$$k_y = \frac{G_1 \ln(y_1/y_2)}{\pi d L}$$

$$\text{Avg. mol wt. of gas at the bottom} = (0.02)(78) + (0.98)(28) \\ = 29$$

Similarly, Avg. mol wt. of gas at the top = 28.2

$$\text{Avg. gas flow rate at the bottom} = (13500/29) \\ = 465.5 \text{ kmol/m}^2\text{-hr}$$

$$\text{Avg. gas flow rate at the top} = 462 \text{ kmol/m}^2\text{h}$$

$$\text{Area of cross section of the tube (0.035m dia)} = 9.62 \times 10^{-4} \text{ m}^2$$

Actual flow rate through the tube,

$$G_1 = (9.62 \times 10^{-4})(462) \\ = 0.4448 \text{ kmol/hr}$$

$$\text{Then, } k_y = 1.786 \text{ kmol/hm}^2(\Delta y)$$

System 1 \rightarrow N₂-benzen system

System 2 \rightarrow Air - NH₃ system

$$Re_1 = 7140, Re_2 = 8630$$

$$Sc_1 = 1.5, Sc_2 = 0.675$$

$$\frac{j_{D_1}}{j_{D_2}} = \left(\frac{Re_1}{Re_2} \right)^{-0.23}$$

$$\Rightarrow \frac{Sh_1}{Sh_2} = \left(\frac{Re_1}{Re_2} \right)^{0.77} \left(\frac{Sc_1}{Sc_2} \right)^{0.33}$$

Taking P_{atm} \approx P in sh, $k_y = 4.25 \text{ kmol/hm}^2(\Delta y)$ for system 2 at the given ~~figur~~, on solving, we get

(10) Bulk conc. of A in the gas phase, $y_b = 0.04$

Eq. relation, $y = 1.2x$

The corresponding eq. liquid-phase conc.

$$x_b^* = \frac{y_b}{1.2} = \frac{0.04}{1.2} = 0.033$$

The overall driving force,

on gas-phase basis, $y_b - y_b^* = 0.04 - 1.2x_b = 0.01$

on liquid-phase basis, $x_b^* - x_b = 0.0083$

The actual conc. of solute in bulk liq, $x_b = 0.025$ which is less than x_b^* . Thus, transport of solute A will occur from gas phase to liquid phase.

a) $k_y = 7.2 \text{ kmol/hm}^2(\Delta y)$, $k_x = 4.6 \text{ kmol/hm}^2(\Delta x)$

At steady state, the local flux is,

$$N_A = k_y(y_b - y_i) = k_x(x_i - x_b) \Rightarrow 7.2(0.04 - y_i) = 4.6(x_i - 0.025) \quad \text{---(1)}$$
$$\therefore y_i = 1.2x_i \quad \text{---(2)}$$

Solving the above two eqn: $x_i = 0.03044$, $y_i = 0.03653$

b) Overall gas-phase coefficient,

$$\frac{1}{k_y} = \frac{1}{k_y} + \frac{m}{k_x} = \frac{1}{7.2} + \frac{1.2}{4.6} = 0.3998$$

$$\therefore k_y = 2.506 \text{ kmol/hm}^2(\Delta y)$$

Similarly, overall liq. phase coefficient,

$$\frac{1}{k_x} = \frac{1}{m k_y} + \frac{1}{k_x} = \frac{1}{1.2 \times 7.2} + \frac{1}{4.6} = 0.331$$

$$\therefore k_x = 3.002 \text{ kmol/hm}^2(\Delta x)$$

c) The local mass flux,

$$N_A = k_y(y_b - y_i) = 7.2(0.04 - 0.03653)$$

$$\therefore N_A = 0.025 \text{ kmol/hm}^2$$

The controlling resistance:

Fraction of total mass transfer resistance in gas phase

$$= \left(\frac{1}{k_y}\right) \left(\frac{1}{k_{y'}}\right) = 0.347$$

Fraction of total mass transfer resistance in liq. phase

$$= \left(\frac{1}{k_x}\right) \left(\frac{1}{k_{x'}}\right) = 0.653$$

so the mass transfer resistances are of comparable magnitude and neither of them can be said controlling

- ⑤ If there is no interfacial resistance, the overall MTC, K_y is $25.3 \text{ kmol}/\text{nm}^2\text{dy}$.

$$\text{Theoretical flux} = K_y (Y_b - Y_b') = 25.3 [0.08 - (0.8)(0.03)]$$

This is more than the actual flux of $1.2 \text{ kmol}/\text{nm}^2$. so there is interfacial resistance to mass transfer.

The actual overall co-eff is,

$$(K_y)_{\text{true}} = \frac{1.2}{[0.08 - (0.8)(0.03)]} = 21.43 \text{ kmol}/\text{nm}^2\text{dy}$$

$$\therefore \text{Interfacial resistance} = \frac{1}{(K_y)_{\text{true}}} - \frac{1}{K_y} = 0.0074 \text{ nm}^2\text{dy}/\text{kmol}$$

$$⑫ P = 1.45 \text{ bar} \quad (\text{total pressure})$$

Partial pressure of solute A in the bulk gas at given section of apparatus, $P_b = 1.45 \times 0.065 = 0.09125 \text{ bar}$, $T = 303 \text{ K}$

$$k_y = \frac{P}{RT} k_c = \frac{1.45 \times 90.3}{0.08314 \times 303} = 5.196 \text{ kmol/m}^2 \cdot \text{h}(\Delta y)$$

Interfacial conc. of liq. side, $x_i = 0.00201$,

$$P_i = (3.318 \times 10^4) (0.00201) = 0.0613$$

The eq. relation is, $P = 3.318 \times 10^4 x^*$,

$$y = \frac{P}{P_i} = 30.5 x^*$$

$$\text{a) Flux of the solute, } N_A = k_y (y_b - y_i) = (5.196) (0.065 - 0.0613) \\ = 0.01922 \text{ kmol/h m}^2$$

$$\text{b) Gas phase resistance} = 1/k_y = 1/5.196 = 0.1924 \text{ hm}^2 \Delta y / \text{kmol} \\ \text{(13.6% of total resistance)}$$

$$\text{Total mass transfer resistance} = 0.1924 / 0.136 = 1.415$$

$$\text{Liquid-phase resistance} = (1.415) (1 - 0.136) = 1.2223 = m/k_x; k_x = 30.5 \\ \therefore k_x = 24.95 \text{ kmol/h m}^2 \Delta x$$

$$\text{The local flux, } N_A = 0.01922 = k_x (x_i - x_b) = (24.95) (0.00201 - x_b) \\ \therefore x_b = 0.00124$$

c) Overall liquid phase MTC,

$$\frac{1}{K_x} = \frac{1}{k_x} + \frac{1}{m k_y} = \frac{1}{24.95} + \frac{1}{(30.5)(5.196)} \\ = 0.04639$$

$$\therefore K_x = 21.55 \text{ kmol/h m}^2 (\Delta x)$$

d) Individual gas phase driving force, $\Delta y = y_b - y_b^*$

$$= 0.065 - 0.0613$$

$$= 0.0084 \text{ mole fraction}$$

In terms of difference of partial pressure, $\Delta p = P \Delta y$

$$= 0.00536$$

Overall gas phase driving force, $\Delta y = y_b - y_b^*$

$$= y_b - m x_b$$

$$= 0.0272$$

(13) Gas flow rate (9.2% SO_2 and 85.5% air) = 1.8 L/min at 20°C, 1 atm

$$Y_{\text{in}} = 0.142 / (1 - 0.142) = 0.165$$

$$Y_{\text{out}} = 0.11 / (1 - 0.11) = 0.1236$$

Air flow rate (SO_2 free basis) = 0.0535 gmol/min

Rate of absorption = $0.0535 (0.165 - 0.1236) = 0.002215 \text{ gmol/min}$

Water flow rate = 27 mL/min = 1.5 gmol/min

SO_2 conc. in the effluent from cell = 1.476×10^{-3} mole fraction at steady state in the cell (since the liquid is well mixed, its conc. in the cell and at the cell exit are same)

a) The bulk concentrations of liquid and gas phases,

$$y_b = 1.476 \times 10^{-3}; \quad y_b^* = 0.11$$

Equ. relation: $y = mx$, $m = 31.3$ (given).

$$\text{Then, } y_b^* = mx_b = (31.3)(0.001476) = 0.0462$$

Rate of absorption at steady state = $0.002215 \text{ gmol/min}$

$$= 1.329 \times 10^{-9} \text{ kmol/hr}$$

Area of mass transfer = 0.00315 m^2

Thus,

$$\begin{aligned} 1.329 \times 10^{-9} &= k_y (0.00315) (y_b - y_b^*) \\ &= k_y (0.00315) (0.11 - 0.0462) \\ \therefore k_y &= 0.6613 \text{ kmol/nm}^2(\Delta y) \end{aligned}$$

$$\text{Wkt, } \frac{1}{k_y} = \frac{1}{k_x} + \frac{m}{k_x} \Rightarrow \frac{1}{0.6613} = \frac{1}{3.8} + \frac{31.3}{k_x}$$

$$\therefore k_x = 25.06 \text{ kmol/nm}^2(\Delta x)$$

using Table 3.1, $k_L = K_L/c$

since the soln. is dilute, the total molar conc. of water, i.e.

$$c = 55.5 \text{ kmol/m}^3$$

$$K_L = \frac{K_x}{55.5} = \frac{25.06}{55.5} = 0.4518 \text{ m/hr}$$

$$= 0.0125 \text{ cm/s}$$

The surface renewal rate,

$$s = \frac{k_L^2}{D_{AB}} = \frac{(0.0125)^2}{1.51 \times 10^{-5}} = 10.4 \text{ per second}$$

b) If the outlet gas conc. is 10% at the same inlet gas flow rate, absorption rate = 0.002883 mol/min. If the req. water flow rate is Q ml/min,

$$0.002883 \approx Q \frac{x_b}{18} = \frac{Q x_b}{18} \quad \dots \textcircled{1}$$

$$\text{WKT, } 0.002883 \times 10^{-3} \times 60 = k_y (0.00815) (y_b - y_b^*) ; y_b = 0.1 \text{ (given)}$$

$$\text{using these, } x_b = \frac{y_b^*}{m} = 5.42 \times 10^{-4}$$

$$\Rightarrow Q = (5.42 \times 10^{-4})^{-1} \times 18 \times 0.002883$$

$$\therefore Q = 95.8 \text{ ml/min}$$

(14) Computation of the eq. data in terms of x, y :

This is shown for one data set, $p = 52 \text{ mm Hg}$ and liq. conc. is 0.7 g SO_2 per 100g water.

The total pressure, $P = 4.5 \text{ bar} = (4.5) (760/1.013) \text{ mm Hg}$

$$y = \frac{52 \text{ mm Hg}}{(4.5)(760/1.013)} = 0.0154$$

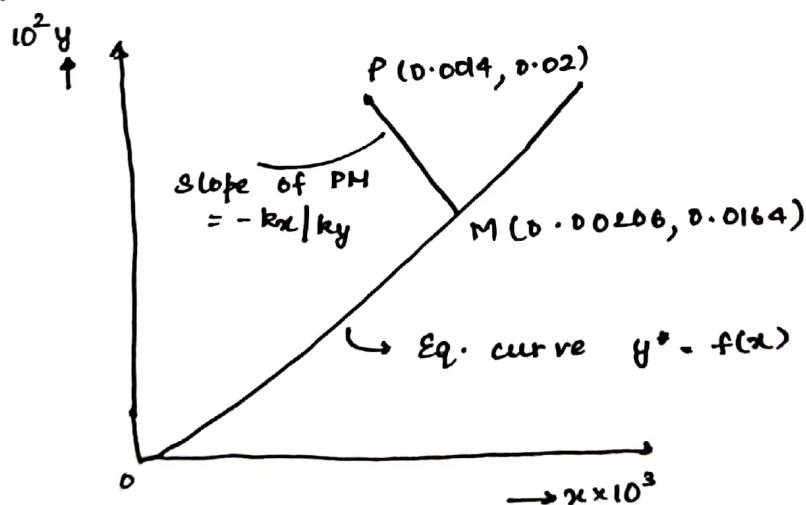
$$\text{Mol. wt of } \text{SO}_2 = 64 ; \quad x = \frac{0.7/64}{0.7/64 + 100/18} = 0.001965$$

a) The calculated eq. data is,

| | | | | | | |
|----------|--------|--------|-------|-------|-------|-------|
| $10^3 x$ | 0.0562 | 0.1403 | 0.280 | 0.422 | 0.564 | 0.842 |
| P, mm Hg | 0.6 | 1.7 | 4.7 | 8.1 | 11.8 | 19.7 |
| $10^2 y$ | 0.0178 | 0.0504 | 0.189 | 0.24 | 0.35 | 0.584 |

| | | | | | |
|----------|-------|-------|------|------|------|
| $10^3 x$ | 1.403 | 1.963 | 2.79 | 3.10 | 4.14 |
| P, mm Hg | 36 | 52 | 79 | 84 | 104 |
| $10^2 y$ | 1.07 | 1.54 | 2.39 | | |

This data is plotted and the eq. curve is shown. The bulk conc. of two phases at given locations are: $x_b = 0.0014$, $y_b = 0.02$. Locating the point P(0.0014, 0.02) on x-y plane. A line of slope LPH $= -k_x/k_y = -5.33$ is drawn through P to meet the curve at M. The point M gives the interfacial conc. of the phases, $x_i = 0.00206$ and $y_i = 0.0164$.



b) The overall co-efficients : In the region near interfacial conc., the eq. line is almost linear having slope $m = 5.65$

$$\frac{1}{k_y} = \frac{1}{k_y} + \frac{m}{k_x} = \frac{1}{15} + \frac{5.65}{80} = 0.0873$$

$$\therefore k_y = 7.28 \text{ kmol/h m}^2 (\Delta y)$$

$$\frac{1}{k_x} = \frac{1}{m k_y} + \frac{1}{k_x} = \frac{1}{5.65 \times 15} + \frac{1}{80} = 0.0243$$

$$\therefore k_x = 4.11 \text{ kmol/m}^2 \text{ h} (\Delta x)$$

$$\text{Absorption flux, } N_A = k_y (y_b - y_i) = 15 (0.02 - 0.0164) \\ = 0.054 \text{ kmol/m}^2\text{hr}$$

c) Mass Transfer co-efficients, k_x' and k_y'

$$k_y = \frac{k_y'}{(1-y)_M}, \quad k_x' = \frac{k_x}{(1-x)_M} \quad M \text{ means log}$$

since x, y are small, $(1-y)_M \approx 1$, $(1-x)_M \approx 1$,

$$\text{Thus, } k_x' = k_x, \quad k_y' = k_y$$

(B) Feed conc. = 12%.

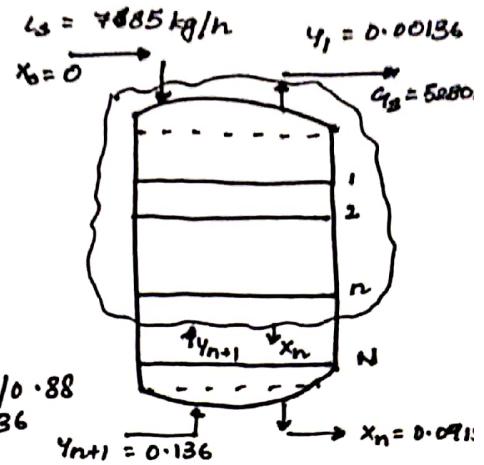
$$\text{Feed rate on solute free basis, } G_{ls} = (6000)(1 - 0.12) \\ = 5280 \text{ kg}$$

$$\text{Mass of solute entering} = 6000 \times 0.12 = 720 \text{ kg}$$

$$\text{Fraction of solute removed} = 0.99$$

$$\text{Using mass ratio unit, feed conc., } Y_{N+1} = 0.12/0.88 \\ = 0.136$$

$$\text{Exit conc., } Y_1 = 0.00136$$



The entering solvent doesn't have any C, i.e. $x_0 = 0$. Solvent input rate, $L_s = 7685 \text{ kg/hr}$, Material balance over the column:

$$G_{ls} (Y_{N+1} - Y_1) = L_s (x_N - x_0)$$

$$5280 (0.136 - 0.00136) = 7685 (x_1 - 0) \Rightarrow x_1 = 0.0915$$

Eqn. of operating line,

$$G_{ls} (Y_{N+1} - Y_1) = L_s (x_N - x_0)$$

$$\Rightarrow Y_{N+1} = 1.455 x_N + 0.00136$$

This is the operating line which passes through terminal points, Top (0, 0.00136) and Bottom (0.0915, 0.136)

The eq. line, $Y = \alpha X = 1.32 X$ ($X = 1.32$) is drawn on x-Y plane.

The number of ideal trays = 23.8

Actual no. of trays = $23.8/0.4 = 59.5$ (60)

In the given problem both eq. and operating lines are linear.

Using the Kremser eqn., $(\bar{A} = L_s / \alpha G_{ls} = 7685 / 1.32 \times 5280 = 1.103)$

$$N = \frac{\log \left[\left(\frac{Y_{N+1} - \alpha x_0}{Y_1 - \alpha x_0} \right) \left(1 - \frac{1}{\bar{A}} \right) + \frac{1}{\bar{A}} \right]}{\log \bar{A}}$$

$$= \log \left[\left(\frac{0.136 - 0}{0.00136 - 0} \right) \left(1 - \frac{1}{1.103} \right) + \frac{1}{1.103} \right] / \log (1.103)$$

$$= 23.7$$

No. of real trays = $23.7 / 0.4 = 59$

$$\bar{A} = 1.103, \quad \frac{Y_1 - \alpha x_0}{Y_{N+1} - \alpha x_0} = \frac{0.00136 - 0}{0.136} = 0.01, \text{ from charts}$$

the number at ideal stage $N = 23.5$

(12) a) The given data is to be converted,

$$x = \frac{\text{kg solute}}{\text{kg solute-free clay}} \quad y = \frac{\text{kg solute}}{\text{kg solute-free soln.}}$$

Mass of clay taken = 60g

Mass of soln. taken = 1000g (100g solute + 900g H₂O)

∴ solute A in soln. after eq. = (900 / (1 - 0.0693)) - 900 = 67g

Mass of A absorbed = 33g

$$x = \frac{33}{60} = 0.55; \quad y = \frac{67}{900} = 0.0744$$

Calculated Data,

$$\begin{array}{cccccccccc} x = 0 & 0.667 & 0.592 & 0.55 & 0.46 & 0.407 & 0.326 & 0.238 & 0.189 \\ y = 0 & 0.1 & 0.0847 & 0.0744 & 0.06 & 0.05 & 0.035 & 0.025 & 0.015 \end{array}$$

a) On fitting the above data, we get, $y = 0.173 \times 1.36$

$$\alpha = 0.173, \quad \beta = 1.36$$

b) Mass of adsorbent used, at eq., $x = 90/M$, $y = 10/900 = 0.011^1$

Using these values in Freundlich's Isotherm,

$$M = 678 \text{ kg}$$

c) Two stage contact use = $\frac{6.78 \times 10^2}{2} = 339 \text{ kg}$

$$C_s = 900 \text{ kg}$$

$$L_s = 339 \text{ kg}$$

$$\text{Slope} = - L_s/C_s = - 0.376$$

$$Y_0 = 100/900 = 0.11$$

$$X_0 = 0$$

$$X_2 = 0.006$$

Mass of solute leaving second stage

$$= 900 \times 0.006$$

$$= 5.4 \text{ kg}$$

Section - 2

① Diffusivity of naphthalene vapour in air = $6 \times 10^{-6} \text{ m}^2/\text{s}$

Kinematic viscosity of air = $1.5 \times 10^{-5} \text{ m}^2/\text{s}$

concentration of naphthalene = 10^{-5} kmol/mL

a) The average mass transfer co-efficient is, (over the flat plate),

For correlation for heat transfer,

$$Nu = 0.664 Re^{1/2} Pr^{1/3}$$

similarly, relation for mass transfer,

$$Sh = 0.664 Re^{1/2} Sc^{1/3}$$

$Sh \rightarrow$ Sherwood Number, $Sh = \frac{kL}{D_{AB}}$

$Re_2 \rightarrow$ Reynolds Number, $Re_2 = \frac{\rho v L}{\mu}$

$Sc \rightarrow$ Schmidt Number, $Sc = \frac{\mu}{\rho D_{AB}}$

$K \rightarrow$ Overall MTC, $L \rightarrow$ length of sheet

$D_{AB} \rightarrow$ diffusivity of A in B, $\mu \rightarrow$ viscosity of air

$v \rightarrow$ velocity of air, $\rho \rightarrow$ density of air

$$\frac{K (0.2)}{6 \times 10^{-6}} = 0.664 \left(\frac{0.2 \times 20}{1.5 \times 10^{-5}} \right)^{1/2} \left(\frac{1.5 \times 10^{-5}}{6 \times 10^{-6}} \right)^{1/3}$$

$$\therefore K = 0.014 \text{ sec}^{-1}$$

$$\text{Rate of loss of naphthalene} = K (C_{A0} - C_{Ai})$$

$$= 0.014 (10^{-5} - 0)$$

$$= 1.4024 \times 10^{-7} \text{ kmol/m}^2 \text{ sec}$$

$$\begin{aligned} \text{Rate of loss per meter width} &= (1.4024 \times 10^{-7}) (0.2) \\ &= 0.8048 \times 10^{-7} \text{ kmol/m sec} \\ &= 0.101 \text{ gmol/m hr} \end{aligned}$$

$$② \text{ WKT, } Nu = 0.43 + 0.532 Re^{0.0} Pr^{0.81}$$

$$d = 6 \text{ mm}$$

$$v = 3 \text{ m/s}$$

$$T_s = 43^\circ\text{C}$$

$$\rho_v = 400 \text{ mm Hg}$$

$$= 53.32 \text{ kN/m}^2$$

$$T_0 = 60^\circ\text{C}$$

Air properties
at $T = 41.5^\circ\text{C}$
 $c_p = 1.008 \text{ kJ/kg K}$

$$\mu = 1.963 \times 10^{-5} \text{ Ns/m}^2$$

$$K = 0.03 \text{ W/mK}$$

$$R_p = \frac{\rho v d}{\mu} = \frac{0.006 \times 3 \times 1.25}{1.963 \times 10^{-5}} = 1146.2047$$

$$Pr = \frac{c_p \mu}{K} = \frac{1.008 \times 1.963 \times 10^{-5}}{0.03} = 6.59 \times 10^{-4}$$

$$Nu = \frac{hd}{K} = 0.43 + 0.532 (1146.2047)^{0.5} (6.59 \times 10^{-4})^{0.31}$$

$$h \frac{(6 \times 10^{-3})}{0.03} = 2.29 \rightarrow h = 11.45 \text{ W/m}^2\text{K}$$

Now, rate of sublimation is estimated by assuming rate of heat transferred by convection equals to rate of sublimation

$$\therefore N_A = Q_A$$

$$= h(T_s - T_0)$$

$$Q_A = 114.5 \cdot 60 - 43$$

$$= 194.65 \text{ W/m}^2 = \text{Heat transferred}$$

$$\text{Rate of sublimation} = \frac{dm}{dt} = \sqrt{\frac{H}{2\pi RT}} (P_{\infty} - P)$$

$$\frac{dm}{dt} = \sqrt{\frac{302}{2 \times \pi \times 8.314 \times 353.15}} (101.325 - 53.32) \times 10^3$$

$$\therefore \frac{dm}{dt} = 6822.9383 \text{ mol/m}^2\text{s}$$

$$③ D = 3 \text{ cm}$$

$$\omega = 20 \text{ rpm} = \frac{20}{60} \text{ rps} = \frac{1}{3} \text{ rps} = \frac{2\pi}{3} \text{ radian per sec.}$$

$$T = 25^\circ C = 298 \text{ K}$$

$$D_B = 10^{-9} \text{ cm}^2/\text{s} = 10^{-9} \text{ m}^2/\text{s}$$

$$\rho = 0.003 \text{ g/cm}^3 = 3 \text{ kg/m}^3$$

$$Sh = 0.62 Re^{1/2} Sc^{1/3} - ①$$

$$v_{\text{water}} = \frac{10^{-8}}{10^3} = 10^{-11} \text{ m}^2/\text{s}$$

$$Sc = \frac{v}{D_B} = \frac{10^{-6}}{10^{-9}} = 1000$$

$$n = \frac{K_c + D}{D_{AB}} = 0.62 \left(\frac{D^2 \omega \rho}{4} \right)^{1/2} \left(\frac{\mu}{\rho D_{AB}} \right)^{1/3}$$

$$\rho = 1 \text{ gm/cm}^3, \quad \mu = 1 \text{ cP}, = 0.01 \text{ g/cm-sec}$$

$$K_c = 0.62 D_{AB} \left(\frac{\omega \rho}{4} \right)^{1/2} \left(\frac{\mu}{\rho D_{AB}} \right)^{1/3}$$

$$= 8.973 \times 10^{-4} \text{ cm/s}$$

$$N_A = 8.973 \times 10^{-4} (0.003) = 2.692 \times 10^{-6} \text{ g/cm}^2 \text{ s}$$

$$\begin{aligned} \text{Rate} &= N_A \cdot S = N_A (2\pi r^2) \\ &= 2.692 \times 10^{-6} (2\pi) (1.6)^2 \end{aligned}$$

$$\therefore \text{Rate} = 8.805 \times 10^{-5} \text{ g/s}$$

④ For methanol in air at 298 K,

$$D_{\text{methanol-air}} = \frac{1.641}{1.068 \times 10^6} = 1.62 \times 10^{-8} \text{ m}^2/\text{s}$$

$$v_{\text{air}} = 1.563 \times 10^{-5} \text{ m}^2/\text{s}$$

$$Sc = \frac{v}{D_{\text{AB}}} = \frac{1.563 \times 10^{-5}}{1.62 \times 10^{-8}} = 0.968$$

⑤ Dittus Boelter eqn:

$$Nu = \frac{hD}{K} = 0.023 Re^{0.8} Pr^{1/3}$$

The corresponding eqn for NTC is given by Sherwood Number and Pr is changed to Schmidt Number.

Using the analogy between heat transfer and mass transfer,

$$Sc = \frac{k_c D}{D_{\text{AB}}} = 0.023 Re^{0.8} (Sc)^{1/3}$$

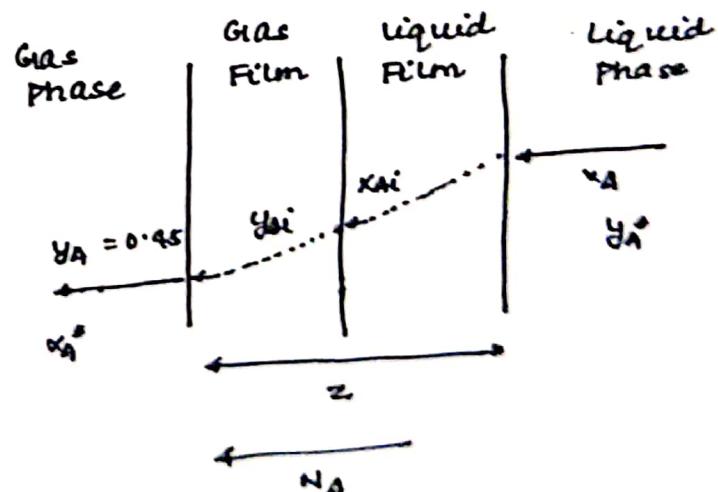
This is the Dittus Boelter eqn for mass transfer.

$$y_A^* = 0.75 x_A$$

$$K_A = 0.9$$

$$y_A = 0.45$$

$$k_y = 0.02716 \text{ kmol/m}^2\text{s}$$



a) Molar Flux of A,

$$N_A = k_y (y_A^* - y_A)$$

$$\text{where, } \frac{y_A^*}{1/k_y} = 0.7 \rightarrow 1/k_y = 0.7/y_A^*$$

$$\therefore k_y = 0.7 \times 0.02716 \\ = 0.019012 \text{ kmol/m}^2\text{s}$$

$$N_A = 0.019012 (0.75 \times 0.9 - 0.45) \\ = 4.278 \times 10^{-3} \text{ kmol/m}^2\text{s}$$

$$b) N_A = k_y (y_{AC} - y_A) \\ = k_y (x_A - x_{Ai})$$

$$y_{AC} = \frac{N_A}{k_y} + y_A \\ = \frac{4.278 \times 10^{-3}}{0.02416} + 0.45$$

$$\therefore y_{AC} = 0.6075$$

$$\text{WKT, } y_{AC} = 0.6075,$$

$$\frac{1}{k_y} = \frac{1}{k_x} + \frac{m}{R_x} \Rightarrow \frac{1}{0.019012} = \frac{1}{0.02416} + \frac{0.75}{k_x}$$

$$\text{on solving, } k_x = 0.04753 \text{ kmol/m}^2\text{s}$$

$$\text{Now, } x_{AC} = x_A - \frac{N_A}{R_x} \\ x_{AC} = 0.9 - \frac{4.278 \times 10^{-3}}{0.04753} \Rightarrow x_{AC} = 0.81$$

② overall MTC

$$\frac{1}{K_x} = \frac{1}{m k_y} + \frac{1}{k_x} \Rightarrow \frac{1}{K_x} = \frac{1}{0.75 \times 0.02416} + \frac{1}{0.04753}$$

$$\text{on solving, } K_x = 0.01426 \text{ kmol/m}^2\text{s}$$

↳ overall mass transfer co-eff for liq phase.

③ Mass Transfer co-eff of A = $K_A = 10.2 \text{ lbmol/n-ft}^2$.

Mass Transfer co-eff of B = $\therefore = \therefore$

Mass Transfer Resistance of A = $1/K_A = 0.09794 \text{ h-ft}^2/\text{lb mol}$

Mass Transfer Resistance of B = $1/K_B = 0.2298 \text{ h-ft}^2/\text{lb mol}$

since resistance of phase B is higher, so phase B resistance is a mass transfer controlling resistance.

⑧ Convert K_G in terms of K_x ,

$$K_x = K_G P = 2.75 \times 10^{-6} \times 101.3 \\ = 2.786 \times 10^{-4}$$

For a gas-phase resistance that accounts for 85% of resistance,

$$K_y / 0.85 = k_y \Rightarrow K_y = 0.85 k_y$$

wRT, $\frac{1}{k_y} = \frac{1}{K_y} + \frac{m}{K_x} - \textcircled{1}$

$$m = 1.64$$

using this in eqn ①, $K_x = 3.05 \times 10^{-4}$

To estimate the ammonia flux and the interfacial conc. at this particular point,

$$y^* = m x(A, C) = 1.64 \times 1.15 \times 10^{-3} \\ = 1.886 \times 10^{-3}$$

$$N_A = k_y (y_{Ai} - y_A^*) = 2.768 \times 10^{-4} (0.08 - 0.001886) \\ = 2.18 \times 10^{-5}$$

$$y_{Ai} = y_{AGi} - \frac{N_A}{k_y} \\ = 0.08 - \frac{2.18 \times 10^{-5}}{3.28 \times 10^{-4}} = 0.01362$$

$$\pi_{AC} = \frac{y_{Ai}}{m} = \frac{0.01362}{1.64} = 8.305 \times 10^{-3}$$