

Mass Transfer-I

Gas Absorption (Continue...)



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Gas Absorption (Continue...)

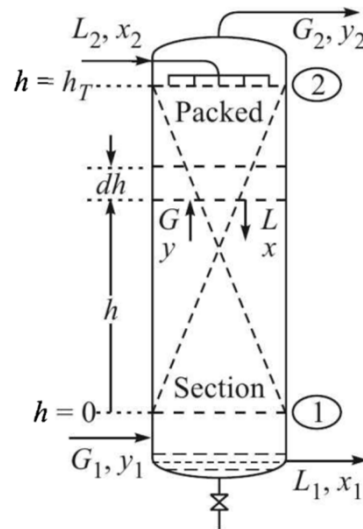
EXAMPLE (*Packed tower design using individual coefficients*) A gas mixture containing 10 mol% SO_2 and 90 mol% air at 1 atm total pressure and 30°C is to be scrubbed with water to remove 97% of the SO_2 in a tower packed with 25 mm ceramic Raschig rings. The feed gas rate is 1500 kg per hour. Calculate (a) the minimum liquid rate, (b) the tower diameter if the liquid rate is 1.25 times the minimum and the tower operates at 70% of the flooding velocity, and (c) the packed height.

For the calculation of the tower diameter, the liquid may be assumed to have properties like water. Equilibrium data for SO_2 -water system at 30°C may be taken from Perry's Handbook. The Colburn-Drew volumetric mass transfer coefficients at the given conditions are:

$$k'_x \bar{a} = 1.25 \text{ kmol}/(\text{m}^3)(\text{s})(\Delta x) \quad \text{and} \quad k'_y \bar{a} = 0.075 \text{ kmol}/(\text{m}^3)(\text{s})(\Delta y).$$

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Average molecular weight of the feed gas (10% SO₂, 90% air) = (0.10)(64) + (0.90)(28.8)
= 32.3

Total feed gas rate = 1500 kg/h = 1500/32.3 = 46.41 kmol/h = G_1

Feed concentration, $y_1 = 0.1$ (mole fraction); $Y_1 = 0.1/0.9 = 0.111$ (in mole ratio unit)

Feed gas rate on solute-free basis = $G_s = G_1(1 - y_1) = (46.41)(1 - 0.10) = 41.77$ kmol/h

SO₂ entering = $G_1 y_1 = (46.41)(0.10) = 4.641$ kmol/h; 97% of this SO₂ is absorbed.

Therefore, SO₂ absorbed = (4.641)(0.97) = 4.502 kmol/h; SO₂ leaving = 0.14 kmol/h;

Concentration: $Y_2 = 0.14/G_s = 0.14/41.77 = 0.00335$

$$y_2 = Y_2/(1 + Y_2) = 0.00335/1.00335 \approx 0.00335$$

The calculated equilibrium data in mole fraction unit (30°C, 1 atm total pressure) are given below:

10 ⁴ x	0	0.562	1.403	2.8	4.22	8.42	14.03	19.65	27.9
10 ³ y	0	0.790	2.23	6.19	10.65	25.9	47.3	68.5	104

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(a) Determination of the minimum liquid rate

The equilibrium (x, y) data are converted to mole ratio units (X, Y) and plotted on the X–Y plane (not shown here; this part is left as an exercise). The curve is slightly convex downwards. So the operating line corresponding to the minimum liquid rate *will not touch* the equilibrium line. It will rather meet the equilibrium line at the point having an ordinate $Y_1(0.111)$. This is the pinch point having the abscissa = $(X_1)_{\max} = 0.00272$. The feed water is SO₂-free. Therefore, $X_2 = 0$.

By an overall material balance,

$$G_s(Y_1 - Y_2) = (L_s)_{\min}(X_{1,\max} - X_2)$$

$$\Rightarrow (41.77)(0.111 - 0.00335) = (L_s)_{\min}(0.00272 - 0), \text{ i.e. } (L_s)_{\min} = \boxed{1655 \text{ mol/h}}$$

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(b) Determination of the tower diameter

The molecular weight of the solvent (water) = 18; $(L_s)_{\min} = (1655)(18) = 29,790 \text{ kg/h}$.

Take the actual liquid rate as 1.25 times the minimum. The feed liquid is *solute-free*.

The actual feed liquid rate (1.25 times the minimum),

$$L_s = (1655)(1.25) = 2069 \text{ kmol/h} (= 37,237 \text{ kg/h})$$

The liquid rate at the bottom of the tower,

$$L_1 = L_s + \text{SO}_2 \text{ absorbed} = 37,237 + (4.502)(64) = 37,525 \text{ kg/h}$$

The flooding velocity should be calculated at a point of the tower where the liquid and the gas

flow rates are maximum. In an absorption tower, it occurs at the bottom; in a stripping tower, it occurs at the top. In the given case, the flow rates at the bottom of the tower are taken.

Total pressure in the tower = 1 atm (the pressure drop across the tower is neglected);
temperature = 303 K

Following the procedure of Example 5.2, the tower diameter can be calculated. At 70% of the flooding velocity, tower cross-section = 0.781 m^2 ; diameter = 1 m; and gas mass flow rate, $G' = 1921 \text{ kg/m}^2 \cdot \text{h}$. Determination of these values is left as an exercise.

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(c) Determination of the packed height

$$\text{Overall material balance, } G_s \left(\frac{y_1}{1-y_1} - \frac{y_2}{1-y_2} \right) = L_s \left(\frac{x_1}{1-x_1} - \frac{x_2}{1-x_2} \right)$$

$$G_s = 41.77 \text{ kmol/h; } L_s = 2069 \text{ kmol/h; } y_1 = 0.10; y_2 = 0.00335; x_2 = 0$$

$$\Rightarrow x_1 = 0.00217 = \text{concentration of SO}_2 \text{ in the exit solution}$$

$$X_1 = \frac{x_1}{1-x_1} = 0.002175$$

Individual gas- and liquid-phase mass transfer coefficients are given. The following equations are to be used to determine the packed height.

$$h = H_{tG} N_{tG}; \quad H_{tG} = \frac{G'}{k'_y \bar{a}} \quad \text{and} \quad N_{tG} = \int_{y_2}^{y_1} \frac{(1-y)_{iM}}{(1-y)(y-y_i)} dy = \int_{y_2}^{y_1} f(y) dy \quad (i)$$

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In order to evaluate the integral (graphically or numerically), it is required to determine the interfacial concentration on the gas-side, i.e. y_i . We follow the procedure described in Section 6.4.1.

The equilibrium data are plotted on the x - y plane. The operating line as given below is also plotted on the same graph [Figure 6.7(a)].

$$G_s \left(\frac{y_1}{1-y_1} - \frac{y}{1-y} \right) = L_s \left(\frac{x_1}{1-x_1} - \frac{x}{1-x} \right)$$

$$G_s = 41.77, L_s = 2069, y_1 = 0.1, x_1 = 0.00217$$

In order to determine the interfacial gas-phase concentration y_i at a particular section of the column where the bulk concentrations are (x, y) , we proceed as follows:

(i) Locate any point $P(x, y)$ on the operating line; (ii) draw a line of slope $-k'_x \bar{a}/k'_y \bar{a}$ ($= -1.25/0.075 = -16.67$) from (x, y) to meet the equilibrium line at the point M_1 which gives the first approximation of the interfacial concentrations (x_i, y_i) . Because in the given system, mass transfer occurs by diffusion of A through non-diffusing B , the true value of (x_i, y_i) can be

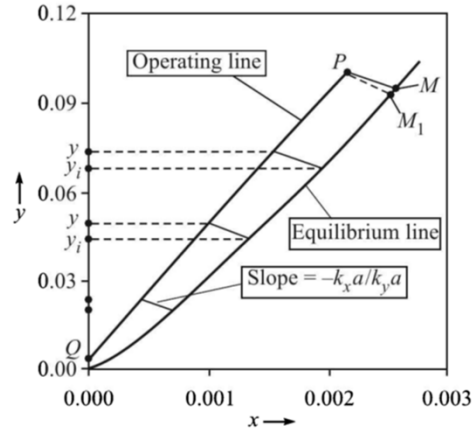


Figure 6.7(a) Determination of interfacial concentrations.

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obtained by drawing a line of slope $k_x \bar{a}/k_y \bar{a}$ from the point (x, y) on the operating line to meet the equilibrium line at M_1 . Now

$$-\frac{k_x \bar{a}}{k_y \bar{a}} = -\frac{(k'_x \bar{a}/(1-x)_{iM})}{(k'_y \bar{a}/(1-y)_{iM})} = -\left(\frac{1.25}{0.075}\right)\left(\frac{(1-y)_{iM}}{(1-x)_{iM}}\right) = -16.67\left(\frac{(1-y)_{iM}}{(1-x)_{iM}}\right)$$

Using the values of interfacial concentrations at the point M_1 , the log-mean concentrations $(1-x)_{iM}$ and $(1-y)_{iM}$ can be calculated. From this we get a better estimate of $-k_x \bar{a}/k_y \bar{a}$. A line having this recalculated slope drawn from the point P meets the equilibrium line at M that gives a better estimate of (x_i, y_i) . Since we have chosen the upper terminal point $P(0.00217, 0.1)$ on the operating line and drawn a line of slope -16.67 , we reach the point $M_1(0.00255, 0.0937)$ on the equilibrium line. We have

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$$(1 - y)_{iM} = \frac{(1 - 0.0937) - (1 - 0.1)}{\ln \frac{1 - 0.0937}{1 - 0.1}} = 0.903$$

$$(1 - x)_{iM} = \frac{(1 - 0.00217) - (1 - 0.00255)}{\ln \frac{1 - 0.00217}{1 - 0.00255}} = 0.9976$$

$$\Rightarrow -\frac{k_x \bar{a}}{k_y \bar{a}} = -(16.67)(0.903/0.9976) = -15.1$$

Draw a line of slope -15.1 through P to reach M (0.00256, 0.0947) on the equilibrium line [see Figure 6.7(a)]. The point M gives a fairly accurate estimate of the interfacial concentration (x_i, y_i) . No further trial is necessary.

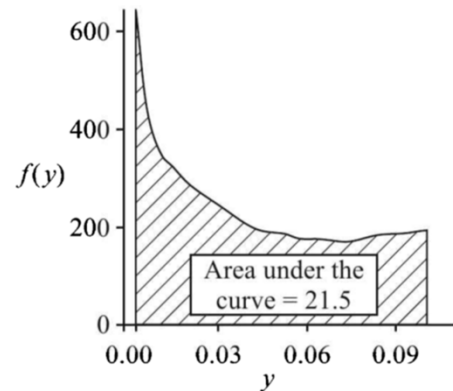


Figure 6.7(b) Graphical integration to determine N_{IG} .

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In a similar way, the interfacial gas-phase concentrations y_i are determined for a set of values of the bulk concentrations y . The values of the integrand, $f(y)$, are now calculated and plotted for a set of values of y .

y	0.00335	0.00887	0.0237	0.0427	0.0569	0.0734	0.1
y_i	0.00178	0.00633	0.02	0.0376	0.0513	0.068	0.047
$f(y)$	640	395	273	198	180	170	190

The area under the graph [Figure 6.7(b)] from $y_2 = 0.00335$ to $y_1 = 0.1$ is 21.5. Therefore,

$$\text{Number of gas-phase transfer units, } N_{IG} = 21.5$$

The height of a gas-phase transfer unit,

$$H_{IG} = \frac{G'}{k_y \bar{a} (1 - y)_{LM}} = \frac{G'}{k'_y \bar{a}}$$

or

$$k'_y = (0.075)(3600) = 270 \text{ kmol/(h)(m}^2\text{)(}\Delta y\text{)}$$

Use the average gas flow rate to calculate H_{IG} .

$$\text{Tower cross-section} = 0.781 \text{ m}^2$$

$$G'_1 = G_1/0.781 = 46.41/0.781 = 59.42 \text{ kmol/h} \cdot \text{m}^2$$

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The rate of mass transfer over a height h of a packed section may be written as

$$\text{Rate of mass transfer} = K_G \bar{a} P h (y - y^*)_{\text{av}}$$

where $(y - y^*)$ is the average driving force over the packed height h . If the rate of mass transfer on a tray [given in Eq. (6.41)] is the same as that achieved over a height h of a packed tower [given by Eq. (6.42)] are equal (i.e. $h = \text{HETP}$),

$$K_G \bar{a} P h (y - y^*)_{\text{av}} = G'(y_{n+2} - y_{n+1})$$

$$h = \frac{G'(y_{n+2} - y_{n+1})}{K_G \bar{a} P (y - y^*)_{\text{av}}}$$

It is reasonable to take $(y - y^*)_{\text{av}}$ as the log-mean of the overall gas-phase driving forces at the points P and Q (see Figure 6.9). Therefore,

$$(y - y^*)_{\text{av}} = \frac{(y - y^*)_P - (y - y^*)_Q}{\ln \left[\frac{(y - y^*)_P}{(y - y^*)_Q} \right]} = \frac{(y_{n+2} - y_{n+1}) - (y_{n+1} - y_n)}{\ln \left[\frac{y_{n+2} - y_{n+1}}{y_{n+1} - y_n} \right]} \quad (6.44)$$

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$$\frac{y_{n+1} - y_n}{y_{n+2} - y_{n+1}} = \frac{(y_{n+1} - y_n)/(x_{n+1} - x_n)}{(y_{n+2} - y_{n+1})/(x_{n+1} - x_n)} = \frac{\text{slope of the equilibrium line}}{\text{slope of the operating line}} = \frac{m}{L'/G'}$$

$$\Rightarrow \frac{mG'}{L'} - 1 = \frac{y_{n+1} - y_n}{y_{n+2} - y_{n+1}} - 1 = \frac{(y_{n+1} - y_n) - (y_{n+2} - y_{n+1})}{y_{n+2} - y_{n+1}}$$

Comparing Eqs. (6.44), (6.45) and (6.46),

$$(y - y^*)_{\text{av}} = \frac{\left(\frac{mG'}{L'} - 1 \right) (y_{n+2} - y_{n+1})}{\ln \left(\frac{mG'}{L'} \right)}$$

Substituting for $(y - y^*)_{\text{av}}$ from Eq. (6.47) in Eq. (6.43),

$$h = \frac{G'}{K_G \bar{a} P} \frac{\ln (mG'/L')}{(mG'/L') - 1} = H_{\text{IOG}} \frac{\ln (mG'/L')}{(mG'/L') - 1}$$

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But, h = the packed height required to achieve the same separation as on the n th tray = HETP.

$$\Rightarrow \text{HETP} = H_{\text{tOG}} \frac{\ln(mG'/L')}{(mG'/L') - 1} = H_{\text{tOG}} \frac{\ln(\bar{S})}{\bar{S} - 1} \quad (6.49)$$

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References



ETH

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Mass Transfer

Theories for Mass Transfer Coefficients

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EXTRACTION

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