GAS-LIQUID SEPARATION PROCESSES

Our intention is to separate one or more components in gas phase by using liquid phase. The gas phase must be brought into intimate contact with liquid phase. The solutes, i.e. components to be separated, can diffuse from gas phase to the liquid phase.

The absorption of ammonia from air by the liquid water and SO₂ removal from flue gases by means of alkaline solution are two general examples of gas-liquid separation processes.

The following diagram represents the feature of gas-liquid separation processes.

![Figure 3.1: Schematic diagram of general gas-liquid separation unit](image)

This process involves mass transfer of solute (define as alphabet A) through a stagnant, nondiffusing gas (define as alphabet B) into a stagnant liquid (define as alphabet C).

If mixing time is long enough, the solute in each of two phases are essentially at equilibrium. In order to predict the concentration of the solute in these two phases, experimental equilibrium data must be available. Note that, most of equilibrium data are available in moles fraction.

Henry’s law is a linear expression of equilibrium data between partial pressure of solute, $p_A$ in gas phase and fraction of solute, $x_A$ in liquid phase, at low concentration of solute, $y_A$ in gas phase entering the gas-liquid separation unit, i.e. must be less than 10 mol%.
Henry’s law equation is as follows:

\[ p_A = H x_A \]

where \( H \) is the Henry’s law constant in atm/mole fraction. Henry’s constants for some common gases are given in Appendix A3 (page 988). If both sides are divided by total pressure, \( P \) in atm,

\[ y_A = H' x_A \]

where \( H' \) is the Henry’s law constant in mole fraction of gas/mole fraction of liquid.

### 3.1 Plate Absorption Tower

In this section, you will be required to determine number of theoretical or ideal stages (or trays) needed to accomplish the separation and to obtain the desired concentration of solute.

Also note that, for common absorption tower (or column) gas phase is essentially insoluble in liquid phase, and liquid phase does not vaporize to gas phase. Only solute diffuses and redistributes itself between these two phases. Both liquid phase and gas phase leaving the stage are at equilibrium.

To conserve use of the liquid phase stream entering the absorption tower and to get more concentrated product, countercurrent multiple-stage contacting is generally used. The following figure represents the schematic diagram of multiple-stage absorption tower.

![Figure 3.2: Schematic diagram of plate absorption tower](image)

Total overall balance:

\[ L_0 + V_{n+1} = L_n + V_1 \]

Component balance:

\[ L_0(x_o) + V_{n+1}(y_{n+1}) = L_n(x_n) + V_1(y_1) \]
Solving $y_{n+1}$ gives the operating line equation having a slope of $L_n/V_{n+1}$:

$$y_{n+1} = \frac{L_0x_o}{V_{n+1}} + \frac{V_1y_1 - L_0x_o}{V_{n+1}}$$

Component balance (using inert flow rate of $L'$ and $V'$):

$$L'\left(\frac{x_o}{1-x_o}\right) + V'\left(\frac{y_{n+1}}{1-y_{n+1}}\right) = L'\left(\frac{x_n}{1-x_n}\right) + V'\left(\frac{y_1}{1-y_1}\right)$$

Two Operating line coordinates are:

$$(x_o, y_1) \text{ and } (x_n, y_{n+1})$$

3.1.1 Plate Absorption Tower for Dilute Gas Mixture

Dilute gas mixture is defined as concentration of solute in entering gas stream ($y_{n+1}$) is less than 10 mol%. For this condition, both operating and equilibrium lines are assumed to be a perfect straight line.

The slope of operating line ($L_n/V_{n+1}$) is almost constant for low concentration, which means there is no significant change in $L_n$ and $V_{n+1}$ throughout the process. On the other hand, the equilibrium line abides by linear relation of Henry’s law for low concentration. You can simply plot two coordinates of equilibrium data, which in the range of $y$-coordinates of operating line in order to build a straight equilibrium line.

3.1.2 Plate Absorption Tower for Concentrated Gas Mixture

This concentrated gas mixture indicates that concentration of solute in gas phase entering the column ($y_{n+1}$) is greater or equal to 10 mol%. To plot the operating line, you need another two intermediate coordinates of operating line.

Step size:

$$Step \ size = \frac{y_{n+1} - y_1}{3}$$
Intermediate y-coordinate:

\[(y_1 + \text{step size}) \text{ and } (y_1 + 2 \times \text{step size})\]

Intermediate x-coordinate:

\[
L \left( \frac{x_o}{1 - x_o} \right) + V' \left( \frac{y_{n+1}}{1 - y_{n+1}} \right) = L' \left( \frac{x_n}{1 - x_n} \right) + V' \left( \frac{y_1}{1 - y_1} \right)
\]

To plot the equilibrium line, you also need at least 4 coordinates within the range of y-coordinates of operating line. Pick up the nearest value of these y-coordinates and intermediate y-coordinates from the equilibrium data. Also note that the equilibrium line will intersect at origin.

### 3.1.3 Number of Stages for Plate Absorption Tower

**Method to step off number of stages:**

1. Draw a horizontal line from \((x_o, y_1)\) through the equilibrium line.
2. At the intersection of this horizontal line and equilibrium line, draw a vertical line through the operating line.
3. Repeat the above procedures until the horizontal line exceed the x_n line at x-axis.
4. Number of theoretical or ideal stages is equal to the number of perfect horizontal lines drawn, plus the length ratio of horizontal line close to the operating line to the whole length of horizontal line, where x_n line divides the horizontal line into two.

**Number of actual stages:**

Number of actual stages (or trays)  
= \[\text{Number of theoretical or ideal trays/Efficiency}\]

**Analytical Kremser equation to calculate number of theoretical stages:**

\[
N = \frac{\ln \left[ \frac{y_{n+1} - m_i x_o}{y_1 - m_i x_o} \left( 1 - \frac{1}{A} \right) + \frac{1}{A} \right]}{\ln A}
\]
where \( A = \sqrt{A_1 A_n} \), \( A_i = \frac{L_o}{m_i V_i} \), \( A_n = \frac{L_o}{m_n V_{n+1}} \) and

\( m_1 = \text{slope of the equilibrium line at point } y_1 \)
\( m_n = \text{slope of the equilibrium line at point } x_n \)

Note: For low solute concentration \( (y_{n+1} < 10 \text{ mol\%}) \), \( m_n = m_1 = \text{Slope of the equilibrium line}. \)

### 3.2 Mass Transfer Between Phases

In gas-liquid separation processes, mass transfer of solute A from gas phase is by convection and through a liquid phase also by convection. Concentration gradient must occur in order for mass transfer to take place. At the interface, equilibrium exists in most cases.

Figure 3.3: Concentration profile of solute A diffusing through two phases

Two phases are present together with an interface between these two phases. The concentration of solute in the bulk gas phase, \( y_{AG} \) decreases to \( y_{Ai} \) at the interface and the solute concentration in liquid starts at \( x_{Ai} \) at the interface before falls to \( x_{AL} \) in the bulk liquid phase. There would be no resistance to transfer across the interface, so \( y_{Ai} \) is in equilibrium with \( x_{Ai} \).

Most of gas-liquid separation processes involve mass transfer of solute A through a stagnant, non-diffusing gas B into a stagnant liquid C. Therefore, the discussion of film and overall mass-transfer coefficients will be more emphasized on diffusion of solute though stagnant or non-diffusing gas.
3.2.1 Film Mass-Transfer Coefficients

1. Diffusion of solute through stagnant or nondiffusing gas

Flux:  
\[
N_A = \frac{k'_y}{(1 - y_A)_{im}} (y_{AG} - y_{Ai}) = \frac{k'_x}{(1 - x_A)_{im}} (x_{Ai} - x_{AL})
\]

\(k'_y\) = Film gas phase mass-transfer coefficient in kg mol/s.m².mol fraction  
\(k'_x\) = Film liquid phase mass-transfer coefficient in kg mol/s.m².mol fraction

\[
(1 - y_A)_{im} = \frac{(1 - y_{Ai}) - (1 - y_{AG})}{\ln \left(\frac{1 - y_{Ai}}{1 - y_{AG}}\right)} \quad \text{and} \quad (1 - x_A)_{im} = \frac{(1 - x_{AL}) - (1 - x_{Ai})}{\ln \left(\frac{1 - x_{AL}}{1 - x_{Ai}}\right)}
\]

Rearrange the flux equation:

\[
\text{Slope} = \frac{-k'_x}{k'_y} \frac{(1 - x_A)_{im}}{(1 - y_A)_{im}} = \frac{y_{AG} - y_{Ai}}{x_{AL} - x_{Ai}}
\]

This slope is used to determine the actual value of \(x_{Ai}\) and \(y_{Ai}\).

2. Equimolar Counterdiffusion

Simply omit \((1-x_A)_{im}\) and \((1-y_A)_{im}\) terms in flux and slope equations from the earlier section.
3.2.2 Overall Mass-Transfer Coefficients

1. Diffusion of solute through stagnant or nondiffusing gas

Flux:

\[
N_A = \left[ \frac{K'_y}{(1 - y_A)_m} \right] y'_A - y'_A \left[ \frac{K'_x}{(1 - x_A)_m} \right] x'_A - x'_A 
\]

\( K'_y \) = Overall gas phase mass-transfer coefficient in kg mol/s m\(^2\).mol fraction
\( K'_x \) = Overall liquid phase mass-transfer coefficient in kg mol/s m\(^2\).mol fraction

\[
(1 - y_A)_m = \frac{(1 - y_A^*) - (1 - y_{AG})}{\ln \left( \frac{(1 - y_A^*)}{(1 - y_{AG})} \right)} \quad \text{and} \quad (1 - x_A)_m = \frac{(1 - x_{AL}) - (1 - x_A^*)}{\ln \left( \frac{(1 - x_{AL})}{(1 - x_A^*)} \right)}
\]

Total resistance based on the overall driving force in gas and liquid phases:

\[
\frac{1}{K'_x/(1 - x_A)_m} = \frac{1}{k'_y/(1 - y_A)_m} \frac{1}{k'_x/(1 - x_A)_m} + \frac{m'}{k'_y/(1 - y_A)_m} \quad \text{and} \quad \frac{1}{K'_x/(1 - x_A)_m} = \frac{1}{m''k'_y/(1 - y_A)_m} \frac{1}{k'_y/(1 - x_A)_m} \frac{1}{k'_y/(1 - x_A)_m} + 1
\]

where \( m' = \frac{y_{AI} - y_A^*}{x_{AI} - x_{AL}} \) and \( m'' = \frac{y_{AG} - y_{AI}}{x_{AI}^* - x_{AI}} \)

If \( m' \) is very small, the major resistance is in the gas phase. Thus the solute is very soluble in the liquid phase. On the other hand, if \( m'' \) is very large, the major resistance is in the liquid phase. Thus, the solute is very insoluble in the liquid phase. Turbulence should be increased in liquid phase to increase rates of mass transfer. Generally, when the major resistance of mass transfer is in the gas phase, the overall mass transfer coefficient \( K'_y \) and film coefficient \( k'_y \) is used.
2. Equimolar Counterdiffusion

Simply omit $(1-x_A)_{im}, (1-y_A)_{im}, (1-x_A)_m$ and $(1-y_A)_m$ terms in flux and total resistance equations from the earlier section.

Note: The value of flux, $N_A$ is remained the same, regardless either using film or overall mass transfer coefficients.

3.3 Continuous Humidification Processes

Warm water is brought into intimate contact with unsaturated air. Warm water enters the top and air enters at the bottom. Packing provides large interfacial areas of contact.

The water cannot be cooled below the wet bulb temperature. Total flow of water and gas is usually assumed to be constant throughout the tower.

$k_G a$ is gas phase volumetric mass transfer coefficient, where ‘a’ is defined as $m^2$ of interfacial area per $m^3$ volume of packed section, having a unit of kg mol/s.m$^3$.Pa (lb mol/h.ft$^3$.atm).
Figure 3.5: Schematic diagram of packed cooling tower

Both total flow of gas (G) and liquid (L) is in kg/s.m², and temperature for both gas and liquid streams is in °C. $H_1$ and $H_2$ represent the humidity exists in entering gas and exit gas respectively. $H_y$ refers to the enthalpy of gas streams.

The Humidity:

Determine $H_1$ using the information of wet bulb temperature and dry bulb temperature or, dry bulb temperature and percentage of humidity.

The Enthalpy:

\[
H_y = (1.005 + 1.88H_1) \times 10^3 (T_{G1} - 0) + 2.501 \times 10^6 H_1 \quad \text{(SI Unit)}
\]
\[
H_y = (0.24 + 0.45H_1)(T_{G1} - 32) + 1075.4H_1 \quad \text{(English Unit)}
\]

Heat Balance:

\[
G(H_y - H_{y1}) = Lc_L(T_{L2} - T_{L1})
\]

$c_L = 4.187 \times 10^3$ J/kg.K (SI Unit)
$c_L = 1.00$ btu/lbm.°F (English Unit)

Operating line coordinates:

$(T_{L1}, H_{y1})$ and $(T_{L2}, H_{y2})$
Rearrange the Heat Balance:

\[
\frac{Lc}{G} = \frac{H_{y2} - H_{y1}}{T_{L2} - T_{L1}} = \text{Operating line slope}
\]

Tower Height:

\[
z = \frac{G}{M_b k_G a P} \int_{y_1}^{y_2} \frac{dH_y}{H_{yi} - H_y} = H_G N_G \quad \text{and,} \quad z = \frac{G}{M_b K_G a P} \int_{y_1}^{y_2} \frac{dH_y}{H_{yi} - H_y} = H_{OG} N_{OG}
\]

\[M_B = \text{Molecular weight of air} = 29\]
\[P = \text{Pressure in Pa (SI Unit) or atm (English Unit)}\]
\[K_G a = \text{Overall mass transfer coefficient in kg mol/s.m}^3. \text{ Pa (lb mol/h.ft}^3.\text{atm)}\]
\[H_G \text{ and } H_{OG} \text{ are height of transfer unit using film and overall mass transfer coefficients respectively.}\]
\[N_G \text{ and } N_{OG} \text{ are number of transfer unit using film and overall mass transfer coefficients correspondingly.}\]

Interface values using film mass transfer coefficient:

\[-h_L a \frac{M_b k_G a P}{k_G a M_b P} = \frac{H_{yi} - H_y}{T_i - T_L} = \text{Slope of line connecting a point at the operating line to the equilibrium line. A point where this line and equilibrium line intersect is the interface value. } \]

\[h_L a \text{ is the liquid-phase volumetric heat-transfer coefficient in W/m}^3.\text{K.}\]

Values of \(H^* y\) using overall mass transfer coefficient:

Simply draw a straight vertical line from any point at the operating line through the equilibrium line. The point where this line intersect at the equilibrium line is the \(H^* y\) value.

Compute the integral terms:

\[
\int_{H_{yi}}^{H_{y2}} \frac{dH_y}{H_{yi} - H_y} \quad \text{and,} \quad \int_{H_{yi}}^{H_{y2}} \frac{dH_y}{H_{yi} - H_y}
\]

can be solved using simple trapezoidal rule.

Temperature approach:

\[\text{Temperature approach} = T_{L1} - T_{\text{wet bulb}}\]
Minimum value of air flow, $G_{\text{min}}$:

A line connecting $(T_{L1}, y_{1})$ coordinates at the operating line to $(T_{L2}, H^{*}y_{2})$ coordinates at the equilibrium line and having a slope of $Lc/G_{\text{min}}$. This line must not intersect (cut) the equilibrium line. If the equilibrium line is quite curved, this line will become tangent to the equilibrium line. The actual gas flow rate is in the order of 1.3 to 1.5 times $G_{\text{min}}$.

Temperature and humidity of air stream in tower:

$$\frac{H_{y2} - H_{y1}}{T_{G2} - T_{G1}} = \frac{H_{y1} - H_{y1}}{T_{i1} - T_{G1}} \quad \text{and,} \quad \frac{H_{y2} - H_{y1}}{T_{G2} - T_{G1}} = \frac{H_{y2} - H_{y2}}{T_{i2} - T_{G2}}$$

Flooding point and pressure drop:

In a given packed tower with specified type and size of packing, and with definite flow of liquid, there is an upper limit to the rate of gas flow, also known as flooding velocity. At certain gas flow rate, the gas starts to hinder the liquid downflow, and local accumulations or pools of liquid start to appear in the packing. The pressure drop of the gas starts to rise at faster rate. This is known as loading point. As the gas flow rate is increased, the liquid holdup or accumulation increases. At the flooding point, the liquid can no longer flow down through the packing and blown out with the gas. Thus, the tower cannot operate any longer.

### 3.4 Packed Absorption Tower

![Diagram of packed absorption tower](image_url)
Two Operating line coordinates are:

\((x_1, y_1)\) and \((x_2, y_2)\)

Operating line equation:

\[ L' \left( \frac{x_2}{1-x_2} \right) + V' \left( \frac{y_1}{1-y_1} \right) = L' \left( \frac{x_1}{1-x_1} \right) + V' \left( \frac{y_2}{1-y_2} \right) \]

Minimum value of liquid flow, \(L_{\text{min}}\):

A line connecting \((x_2, y_2)\) coordinates at the operating line and \((x_{1\text{max}}, y_1)\) coordinates at the equilibrium line. \(L_{\text{min}}\) can be solved using total balance calculation as follows;

\[ L'_{\text{min}} \left( \frac{x_2}{1-x_2} \right) + V' \left( \frac{y_1}{1-y_1} \right) = L'_{\text{min}} \left( \frac{x_{1\text{max}}}{1-x_{1\text{max}}} \right) + V' \left( \frac{y_2}{1-y_2} \right) \]

Design Equation to calculate height of packed absorption tower:

\[ z = \int_{y_2}^{y_1} \frac{Vdy}{k_y' aS} \left( -y - y_{i\text{m}} \right) \]

\[ z = \int_{x_2}^{x_1} \frac{Ldx}{k_x' aS} \left( -x - x_{i\text{m}} \right) \]

\[ z = \int_{y_2}^{y_1} \frac{Vdy}{K_y' aS} \left( -y - y_{*\text{m}} \right) \]

\[ z = \int_{x_2}^{x_1} \frac{Ldx}{K_x' aS} \left( -x - x_{*\text{m}} \right) \]

\(S = \) Cross sectional area of packed tower

\(k_y'a, k_x'a = \) Volumetric film mass transfer coefficient in kg mol/s.m\(^3\).mol fraction

\(K_y'a, K_x'a = \) Volumetric overall mass transfer coefficient in kg mol/s.m\(^3\).mol fraction

Note: For low concentration of solute in gas phase, these coefficients are essentially constant. However, for concentrated solute in gas phase, these coefficients may vary with total flows and usually given in empirical correlation.
Interface values using film mass transfer coefficient:

\[-k_y \frac{a}{(1-x)_{im}} = \text{Slope of line connecting a point at the operating line to the equilibrium line. A point where this line and equilibrium line intersect is the interface value. Use the coordinates (x_1, y_1) for first trial. For low concentration of gas mixture, this value is sufficient enough, and second trial is not necessary. Furthermore, this slope is constant throughout the operating line for low concentration of gas mixture. For high concentration of gas mixture, you can still use any point at the operating line as preliminary estimate and second iteration also not necessary. However, the slope is no longer a constant value throughout the operating line.}\]

Values of y* and x* using overall mass transfer coefficient:

Simply draw a straight vertical line from any point at the operating line through the equilibrium line. The point where this line intersect at the equilibrium line is the y* value. Or, draw a horizontal straight line from any point at the operating line through the equilibrium line. The point where this line intersect at the equilibrium line is the x* value.

3.4.1 Packed Absorption Tower for Dilute Gas Mixture

Concentration of solute in entering gas stream (y_1) must less than 10 mol%. For this condition, both operating and equilibrium lines are assumed to be a perfect straight line. You can simply plot two coordinates of equilibrium data, which in the range of y-coordinates of operating line in order to build a straight equilibrium line.

Simplified design equation:

\[z = \frac{V_{av}}{k_y a S} \left( \phi_1 - y_2 \right) = H_G N_G, \quad z = \frac{L_{av}}{k_x a S} \left( \phi_1 - x_2 \right) = H_L N_L, \]
\[z = \frac{V_{av}}{K_y a S} \left( \phi_1 - y^* \right) = H_{OG} N_{OG}, \quad z = \frac{L_{av}}{K_x a S} \left( \phi_1 - x^* \right) = H_{OL} N_{OL}, \]

\(V_{av}\) and \(L_{av}\) refer to the average molar flow rate values of gas and liquid respectively.

\(H_G\) and \(H_{OG}\) are height of transfer unit using film and overall mass transfer coefficients respectively.
N\text{G} and N\text{OG} are number of transfer unit using film and overall mass transfer coefficients correspondingly.

Where; 
\[
(y - y_i)_m = \frac{(y_1 - y_{1i}) - (y_2 - y_{2i})}{\ln[(y_1 - y_{1i})/(y_2 - y_{2i})]}, \quad (y - y^*_i)_m = \frac{(y_1 - y^*_{1i}) - (y_2 - y^*_{2i})}{\ln[(y_1 - y^*_{1i})/(y_2 - y^*_{2i})]}
\]

\[
(x_i - x)_m = \frac{(x_{1i} - x_i) - (x_{2i} - x)}{\ln[(x_{1i} - x_i)/(x_{2i} - x)]} \quad \text{and} \quad (x^*_i - x)_m = \frac{(x^*_{1i} - x_i) - (x^*_{2i} - x)}{\ln[(x^*_{1i} - x_i)/(x^*_{2i} - x)]}
\]

Other relations:

\[
\frac{1}{K'_y a/(1-y)_m} = \frac{1}{k'_y a/(1-y)_m} + \frac{m'}{k'_y a/(1-x)_m}
\]

\[
\frac{1}{K'_x a/(1-x)_m} = \frac{1}{m'k'_y a/(1-y)_m} + \frac{1}{k'_x a/(1-x)_m}
\]

Where; 
\[
(1-y)_m = \frac{(1-y_{1i}) - (1-y_i)}{\ln[(1-y_{1i})/(1-y_i)]} \quad \text{and} \quad (1-x)_m = \frac{(1-x_i) - (1-x_{1i})}{\ln[(1-x_i)/(1-x_{1i})]}
\]

\[
(1-y^*_m) = \frac{(1-y^*_{1i}) - (1-y^*_i)}{\ln[(1-y^*_{1i})/(1-y^*_i)]} \quad \text{and} \quad (1-x^*_m) = \frac{(1-x^*_i) - (1-x^*_{1i})}{\ln[(1-x^*_i)/(1-x^*_{1i})]}
\]

Or;

\[
H_{OG} = H_G + (m V_{av}/L_{av}) H_L \quad H_{OL} = H_L + (L_{av}/m V_{av}) H_G
\]

For low concentration of solute in gas phase, m=m'=m"=slope of the equilibrium line.

3.4.2 Packed Absorption Tower for Concentrated Gas Mixture
Concentration of solute in gas phase entering the column $y_1 \geq 10$ mol%. To plot the operating line, you need another two intermediate coordinates of operating line.

Step size:

$$\text{Step size} = \frac{y_1 - y_2}{3}$$

Intermediate y-coordinate:

$$(y_2 + \text{step size}) \text{ and } (y_2 + 2 \times \text{step size})$$

Intermediate x-coordinate:

$$L' \left( \frac{x_2}{1-x_2} \right) + V \left( \frac{y_1}{1-y_1} \right) = L' \left( \frac{x_1}{1-x_1} \right) + V' \left( \frac{y_2}{1-y_2} \right)$$

To plot the equilibrium line, you also need at least 4 coordinates within the range of y-coordinates of operating line. Pick up the nearest value of these y-coordinates and intermediate y-coordinates from the equilibrium data. Again, please note that the equilibrium line will intersect at origin.

All values (except cross sectional area, $S$) in the design equation to calculate height of absorption tower may vary if concentrated solute being used. Therefore, the simplified design equations are no longer valid. You need to integrate these equations thoroughly. You can solve height of absorption tower, $z$ using either graphical or numerical method.

Graphical method:

Plot $\frac{V}{k' \ aS (1-y)(1-y_y)}$ versus $y$. Area under curve of this plot is equal to height of absorption tower. Similar plot can be applied for equations using $k' \ a$, $K' \ a$ and $K'' \ a$. At least 4 coordinates are required to construct the plot.

Numerical method:

The most preferable method is Simpson’s 3/8 rule, as follows;
\[ z = \int_{y_3}^{y_2} f(y) dy = \frac{3}{8} h \left[ (y_0) + 3f(y_1) + 3f(y_2) + f(y_3) \right] \quad \text{and} \quad h = \frac{y_3 - y_0}{3} \]

Number of transfer unit for concentrated gas mixture:

\[ N_G = \int_{y_2}^{y_1} (1-y)_{im} \frac{dy}{y_i - y} \quad , \quad N_L = \int_{y_2}^{y_1} (1-x)_{im} \frac{dx}{x_i - x} \quad \text{and} \quad N_{OG} = \int_{y_2}^{y_1} (1-y)_{im} \frac{dy}{y_i - y} \]

3.5 Predicting Mass Transfer Film Coefficient

The following equations are using SI units.

\[ H_G = \left( \frac{0.226}{f_p} \right) \left( \frac{N_{Sc.G}}{0.66} \right)^{0.5} \left( \frac{G_x}{6.782} \right)^{-0.5} \left( \frac{G_y}{0.678} \right)^{0.35} \]

\[ H_L = \left( \frac{0.357}{f_p} \right) \left( \frac{N_{Sc.L}}{372} \right)^{0.5} \left( \frac{G_x}{\mu} \right)^{0.3} \left( \frac{6.782}{0.8937 \times 10^{-3}} \right) \]

\[ N_{Sc} = \text{Schmidt number} = \frac{\mu}{\rho D_{AB}} \]

\[ D_{AB} = \text{Diffusivity} \]

For gas; \( D_{AB} \propto T^{-1.75} \) (Please refer table 6.2-1, text book)

For liquid; \( D_{AB} \propto T/\mu \) (Please refer table 6.3-1, text book)

\( F_p = \text{Relative mass transfer coefficient} \) (Please refer table 10.6-1, text book)

\( G_x \) and \( G_y \) are kg total/s.m² of liquid and gas respectively.