Improvement of Gas Absorption Unit in Laboratory

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Abstract

This research was an improvement of gas absorption unit within the laboratory of Chemical Engineering Department. The packed-bed gas absorption is a crucial unit operation for learning the mass transfer. The aim is to foster chemical engineering students' understanding in principles and operating condition within absorption column. Hence, it has been designed into three parts including: 1) improvement the carbon removal efficiency using water, 2) the study on pressure drop indicating the operating condition within absorption column and 3) the study on parameters influencing overall mass transfer coefficient. The gas absorption column improvement caused an increase in carbon removal efficiency from 0.005% to 0.424% at the liquid flow rate as 800 lit/h, gas flow rate of as 20 lit/h and carbon dioxide concentration as 25%mol. In this study, at normal operating condition found that pressure drop was a straight parallel to dry pressure drop. An increase in gas velocity affecting rapidly increased in pressure drop. That was a working limitation of gas absorption column. The parameters affecting the overall mass transfer coefficient including air flow rate (800 and 100 lit/h), liquid flow rate (15 and 20 lit/h) and inlet carbon dioxide concentration (15 and 20 %mol) were studied. The results demonstrated that an increasing liquid flow rate caused an increase in overall mass transfer coefficient. However, increase in the gas flow rate or inlet carbon dioxide concentration affected the overall mass transfer coefficient decreasing due to carbon dioxide absorbed process is controlled by the liquid-film.

Keywords: Absorption column, Pressure drop, Mass transfer coefficient

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บทคัดย่อ

โครงงานนี้เป็นการปรับปรุงการทดลองปฏิบัติการเฉพาะหน่วยการดูดซึมก๊าซภายในห้องปฏิบัติการของภาควิชาวิศวกรรมเคมี การดูดซึมก๊าซด้วยหอวัสดุบรรจุเป็นการปฏิบัติการเฉพาะหน่วยที่สำคัญในการศึกษาการถ่ายโอนมวล เพื่อให้นักศึกษาวิศวกรรมเคมีได้เข้าใจถึงหลักการ และสภาวะการทำงานภายในหอดูดซึม จึงได้ออกแบบการทดลองเป็น 3 ส่วน คือ การเพิ่มประสิทธิภาพการกำจัดคาร์บอนไดออกไซด์ด้วยน้ำของหอดูดซึมเดิม การศึกษาค่าความดันลดซึ่งสามารถบ่งบอกสภาวะการทำงานภายในหอดูดซึม และการศึกษาตัวแปรที่ส่งผลต่อค่าสัมประสิทธิ์การถ่ายโอนมวล จากการศึกษาพบว่าที่สภาวะการทำงานปกติ เมื่อเพิ่มอัตราเร็วก๊าซ ค่าความดันลดมีลักษณะการเพิ่มขึ้นเป็นเส้นตรงขนานกับค่าความดันลดขณะหอแห้ง และเมื่อเพิ่มอัตราเร็วก๊าซถึงจุดท่วม ค่าความดันลดเพิ่มขึ้นอย่างรวดเร็ว นั่นคือจุดที่แสดงถึงข้อจำกัดการทำงานของหอดูดซึมก๊าซ ในส่วนของตัวแปรที่ส่งผลต่อค่าสัมประสิทธิ์การถ่ายโอนมวลสาร ได้ทำการศึกษาที่อัตราการไหลของอากาศที่ 800 และ 1,000 ลิตรต่อชั่วโมง อัตราการไหลของน้ำที่ 15 และ 20 ลิตรต่อชั่วโมง และความเข้มข้นของคาร์บอนไดออกไซด์ขาเข้าที่ 15 และ 20 เปอร์เซ็นต์โดยโมล ซึ่งผลการศึกษาพบว่าเมื่ออัตราการไหลของน้ำเพิ่มขึ้น ส่งผลให้ค่าสัมประสิทธิ์การถ่ายโอนมวลเพิ่มมากขึ้น แต่เมื่อเพิ่มอัตราการไหลของก๊าซหรือเพิ่มความเข้มข้นของคาร์บอนไดออกไซด์ขาเข้า จะทำให้ค่าสัมประสิทธิ์การถ่ายโอนมวลลดน้อยลง เนื่องจากกระบวนการดูดซึมคาร์บอนไดออกไซด์ด้วยน้ำเป็นกระบวนการควบคุมโดยชั้นฟิล์มของของเหลว และเมื่อได้มีปรับปรุงหอดูดซึมแก๊ส ทำให้ได้ประสิทธิภาพการกำจัดคาร์บอนไดออกไซด์เพิ่มขึ้นจาก 0.005% เป็น 0.424% ที่สภาวะอัตราการไหลของน้ำ 800 ลิตรต่อชั่วโมง อัตราการไหลของน้ำ 20 ลิตรต่อชั่วโมง และความเข้มข้นคาร์บอนไดออกไซด์ขาเข้า 25% โดยโมล

คำสำคัญ: หอดูดซึม, ความดันลด, สัมประสิทธิ์การถ่ายโอนมวล

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Nomenclature

a	Mass transfer area per unit volume of packed bed
a_h	Specific hydraulic area of packing
c	Bulk average concentration of A in the liquid
C	Concentration
C_a	Concentration of a species in the aqueous phase
C_h	Characteristic of packing depends on size and material
d_p	Packing size
D_{AB}	Diffusivity coefficient of A in B
D_T	Column diameter
f	Fraction of flooding
F_{LV}	Liquid-to-gas kinetic energy ratio
F_p	Packing factor
g	Gravitational constant
G	Gas molar velocity
h_L	Specific liquid holdup
H_A	Henry’s constant
H_{OG}	Overall height of a gas transfer unit (HTU)
J_A	Molecular flux
k	Mass transfer coefficients, Experimental coefficient for various packing
K	Overall mass transfer coefficient
L	Flow rate of solvent
m	Slope of the equilibrium line
M_V	Molecular weight of mixed gas
N_{Re_L}	Liquid Reynold number
N_{Fr_L}	Liquid Froude number
N_A	Mass transfer flux
N_{OG}	Overall number of gas transfer units (NTU)
p	Partial pressure
ΔP_T	Total pressure drop
ΔP	Pressure drop

ΔP_{flood}	Pressure drop at flooding
r	Mass-transfer rate per unit volume of packed bed
S	Inside cross-sectional area of the column
u	Superficial velocity
$u_{V,f}$	Gas velocity at flooding point based on cross sectional area of column
V	Volume, Flow rate
x	Mole fraction x of solute in liquid phase
x*	Fictitious liquid mole fraction in equilibrium with the mole fraction
y*	Fictitious vapor mole fraction in equilibrium with the mole fraction
y	Mole fraction y in vapor phase
z_T	Packed column height in z direction

Greek latters

δ	Thickness of film
ρ	Density
μ	Dynamic viscosity
ν	Kinematic viscosity
ρ_V	Gas density
ε	Void fraction

Subscripts

in	Inlet
out	Outlet
v,y	Gas phase
l,x	Liquid phase
d	dry
w	wet
I	Interface

CHAPTER I

INTRODUCTION

1.1 Background

Gas absorption is a one of momentous unit operation in the industry, which employed to eliminate solutes such as impurities, contaminants, pollutants, or prevent catalyst poisons from the carrier gas by contacting with flow of solvent. Solute in a gas carrier transfers into a film gas due to the difference in concentration. At the film interface between two phases, physical equilibrium exists. In fact, a soluble solute is dissolved into a solvent corresponding to the Henry's law which described as the amount of dissolved gas is proportional to its partial pressure in the gas phase.

Absorption column or absorber is classified into many types, for example, plate columns, which also separated according to mode of flow as crossflow and counterflow, packed columns installed with structure packing or random packing. And other types such as spray tower, bubble column and centrifugal contactor. However, the choice of device is most often between a trayed and a packed column[1].

In the industries, for example, initial raw natural gas produced at a wellhead needed the cleaning process, then enters the other parts of natural gas processing. Those gas generally contains 2-20% by volume of hydrogen sulfide (H_2S) and 2-70% by volume CO_2, are commonly known as a sour gas. CO_2 are eliminated until meet the specifications and becomes a sweetens gas. The treated gas generally contains CO_2 less than 2% by volume. This process using an amine ($-NH_2$) absorption in acid gas removal (AGR) and CO_2 absorption process. Additionally, ethylene oxide (EO) production also utilizes the gas absorber where water absorbs the EO to separate EO as a product from the reactor product stream. After that, the dissolved EO in water is further transported to the distillation unit[2]. Moreover, in cement production, the major emitter of CO_2 as a waste is the rotary kiln where CO_2 is released from both fuel combustion and limestone calcination[3]. Absorption process has used to remove any impurities. Due to it is claimed as the most mature and reliable choice and a successfully incorporated into the gas separation from various industrial gas mixtures for decades[4].

Principle and understanding of gas absorption are studied on the laboratory scale in order to apply with pilot plant or industrial scale. On the contrary, gas absorption unit in the laboratory of chemical engineering department has been used for long time. Some of problems may arise. Therefore, the study for improving the gas absorption unit will be done. Moreover, the operating condition in packed column could be descripted by studying changes of pressure drop. Lastly, effect of studied variables including inlet concentration of gas, flow rate of gas and flow rate of liquid on the overall mass transfer coefficient are studied.

1.2 Objectives

1.2.1 To study operating condition within packed column by changes of pressure drop

1.2.2 To study effects of gas flow rate, liquid flow rate and inlet gas concentration to the overall mass transfer coefficient between gas and liquid

1.2.3 To change and rearrange gas absorption apparatus in the laboratory and rewrite the gas absorption manual

1.3 Scopes of work

1.3.1 Measure pressure drop by regulating gas flow rate in range of 300 to 1,500 lit/h at constant liquid flow rate (in range of 40 to 55 lit/h)

1.3.2 Carry out the experiment by using gas flow rate of 800 and 1,000 lit/h, liquid flow rate of 15 and 20 lit/h and inlet CO_2 concentration of 15 and 25%mol

1.3.3 Change and rearrange gas absorption apparatus within the laboratory of department of chemical engineering

1.4 Expected Outputs

1.4.1 Completely detail of gas absorption manual

1.4.2 Enhanced efficiency of CO_2 removal using the improved gas absorption column

CHAPTER II

LITERATURE REVIEW

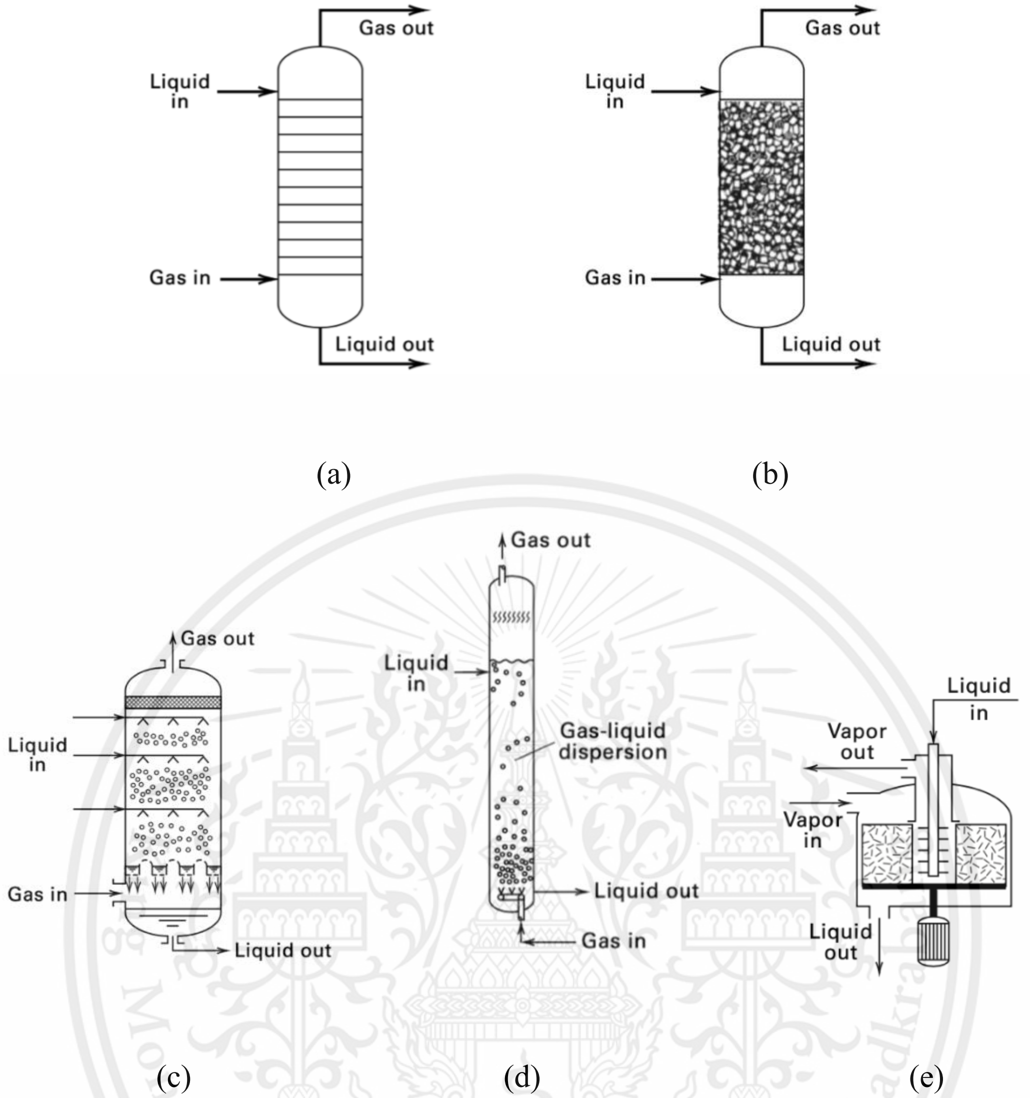
Gas absorption is a vapor-liquid separation process which consisting at lease 3 components, soluble solute (absorbate), suitable solvent (absorbent) and carrier gas by using gas solubility as a main property. Due to the difference in concentration of solute between carrier gas and solvent phase, mass transfer of solute certainly occurs. Generally, only one of these components crosses the phase boundary. For example, emitted CO_2 contained in the air is eliminated by amine absorption. When multiple stages are required, phase contacting is most carried out in different type of columns described next.

2.1 Type of absorption column and equipment[5]

Absorbers is mainly trayed towers (plate columns) and packed columns, and less often spray towers, centrifugal contactors, and bubble columns, respectively as shown in Figure 2.1. For trayed towers, countercurrent-flow of vapor and liquid are contacted on trays or plates that provide close contact of liquid with vapor to promote rapid mass transfer. Liquid solvent from upper tray flows across each tray, which becomes a froth, and then flows over an outlet weir into a downcomer. Whereas a carrier gas contained solute flows upward through opening holes, usually from the high pressure to low pressure area. As a result, in case of excellent contacting or very low viscous liquid is used, equilibrium is approached on each tray due to mass transfer by contacting between gas bubbles and continuously-flow of solvent.

In the same way for packed column, vapor flows upward through the wetted packing, contacting with the countercurrent-flow of liquid. Besides, the solvent flows downward as a film or as droplets between packing elements in order to enhance contacting area. Packed bed can be generally divided into 2 categories including random arrangement and structure packing caused difference in existed pressure drop.

For contactors, spray tower, if only one or two stages and very low-pressure drop are required, and the solute is very soluble in the liquid, spray tower is indicated for suitable absorption. In detail, vertical vessel is filled with gas whereas solvent is sprayed from each tray. Hence, mass transfer continuously occurs between a continuous gas phase and a dispersed solvent. In case of centrifugal contactor, it consists of a stationary, ringed housing, intermeshed with a ringed rotating section. Solvent is fed in



(a) (b)

(c) (d) (e)

Figure 2.1 Type of absorption columns: (a) trayed tower; (b) packed column; (c) spray towers; (d) bubble column; (e) centrifugal contactor[1]

near the center and flows outward by centrifugal force while gas flows inward by pressure driving force. Reportedly, high mass transfer rates can be achieved.

In the bubble column, vertical cylindrical vessel partially filled with solvent into which the gas is bubbled. This column should not be considered unless has a very low solubility in liquid or slow chemical reaction occurs in liquid phase which re quires large residence time. In this regard, advantages and disadvantages for each type of absorption column are shown on Table 2.1.

Table 2.1 Comparison of types of absorption column[6]

Types of absorption column	Advantages	Disadvantages
Trayed towers	1. Can handle high or low liquid flow rates cost effectively 2. Easily customized to the specific requirements such as operations requiring much heat	1. Higher pressure drop than packed column 2. Slow reaction rate processes 3. Plugging and fouling may occur
Packed column	1. Superior contact between gasses and liquid that promote high mass transfer rate 2. Low capital, operating, and maintenance cost	1. Channeling, which must be controlled by redistributing liquid 2. Cannot handle extremely high or low flow rates
Spray towers	1. Low pressure drop 2. Most effective for solutes with high liquid solubility	1. High pumping cost 2. Poor mass transfer 3. Low residence time
Centrifugal contactors	1. High mass transfer rates can be achieved 2. Favored when no headroom for tray or packed column	1. Vary low residence time 2. Low number of contact stages
Bubble columns	1. Suitable for reaction which requires large residence time 2. Little maintenance and operating cost are required	1. High vapor pressure drop 2. Low vapor throughput

2.2 Absorption in packed column

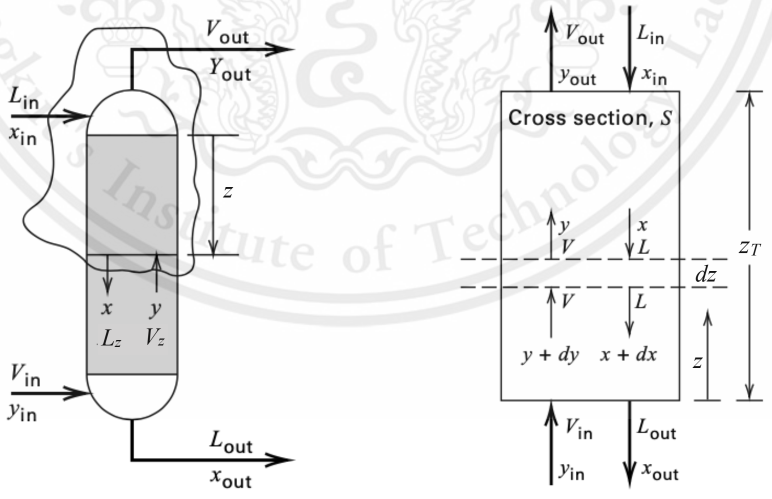
Accordingly, gas absorption in packed columns is widely utilized in several industries. Rochelle[7] claimed that chemical solvent scrubbing using mainly monoethanolamine (MEA) in packed columns is considered as the most commercially proven post-combustion capture approach. Liuta[8] also indicated that gas-liquid random packed bed columns is candidate for gas absorption due to its large capacity, high surface area to volume ratios, low pressure drop, enhanced mechanical stability and reduced costs. Consequently, understanding of absorption in packed column principle and calculation, liquid holdup which will be descripted further, limiting for absorption such as flooding, and also pressure drop are significant issues for designing the packed column absorption.

2.2.1 Packing and packed tower design[1]

The packed column consists of a cylindrical column, which equipped with a gas inlet and distributing space at the bottom, a liquid inlet and distributor at the top, gas and liquid outlets at the top and bottom, respectively.

As shown in Figure 2.2, flow rate of inlet gas (V_{in}) from the bottom of column contains solute at mole fraction y in vapor phase (y_{in}) flows upward and then purged at the top as the outlet gas flow rate (V_{out}) with the solute mole fraction y (y_{out}). Whereas flow rate of inlet solvent (L_{in}) which contains solute at mole fraction x in liquid phase (x_{in}) is fed from top and then flows downward to the bottom as flow rate of solvent outlet (L_{out}) at mole fraction x of solute (x_{out}).

Packing is instrumental in creating longer gas–liquid contact time during gas absorption process. Figure 2.3 shows some examples of commercial packing.



(a) (b)

Figure 2.2 (a) The big picture (b) Differential contact for absorption in packed columns with countercurrent flow[1]

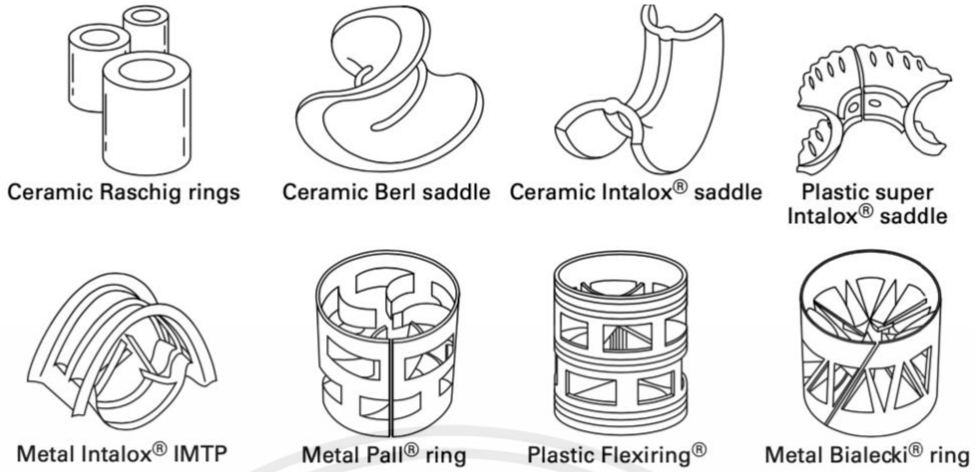


Figure 2.3 Typical random packing materials used in a packed column

As mentioned, the others key issue in packed column design is the selection of the packing type and material which directly affected to the contacting between liquid and gas, and pressure drop. In detail, packing is instrumental in creating longer gas–liquid contact time during gas absorption process. As packing size increases, mass transfer efficiency and pressure drop decrease. Therefore, an optimal packing size exists.

2.2.2 Operation method for packed bed absorption[9]

Basically, film theory is generally used to expound a process of absorption of A into liquid B takes place without vaporization of B, and there is no resistance to mass transfer of A in the gas phase. At the interface, concentration of A (c_I) directly relates with the partial pressure of A (P_I) according to solubility relation of Henry's law as shown in equation 2-1.

$$x_I = H_A y_I \qquad 2\text{-}1$$

Where H_A = Henry's constant

According to Figure 2.4, solute A diffuses pass through a liquid film, which has thickness δ, by different in concentration between c_I and the bulk average concentration

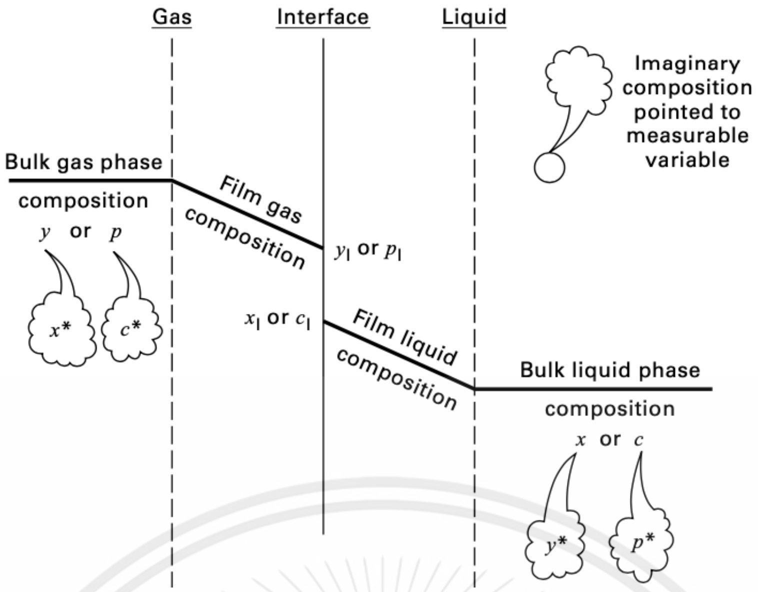


Figure 2.4 Film theory for mass transfer from fluid–fluid interface into a liquid[1]

of A in the liquid (c). Assumed that δ is very thin, hence Fick's first law gives the molecular flux shown in equation 2-2.

$$J_A = \frac{D_{AB}}{\delta}(c_I - c) = \frac{cD_{AB}}{\delta}(x_I - x) \qquad 2\text{-}2$$

Where D_{AB} = diffusivity coefficient of A in B, m/s

If the liquid phase is dilute in A, the bulk-flow effect, which expressed as a convection flux, can be neglected then molecular flux is represented as a mass transfer flux corresponding to equation 2-3.

$$N_A = \frac{D_{AB}}{\delta}(c_I - c) = \frac{cD_{AB}}{\delta}(x_I - x) \qquad 2\text{-}3$$

Equation 2-4 represented the mass transfer coefficient. Mass-fraction driving force is existed for the common combination of vapor–liquid mass transfer, which expressed in mass transfer coefficients k_x and k_y for liquid and gas phases respectively.

$$r = k_x a(x_I - x) = k_y a(y - y_I) \qquad 2\text{-}4$$

Ratio of mass transfer coefficient, which corresponding to equation 2-5, indicates the composition at the interface.

$$\frac{(y\text{-}y_I)}{(x\text{-}x_I)} = -\frac{k_x a}{k_y a} \qquad 2\text{-}5$$

As shown in Figure 2.5, slope $-k_x a/k_y a$ which is a line drown from point (x,y) on the operating line to point (x_I,y_I) on the equilibrium line determines the relative resistances of the two phases. The distance AE and AF represent the gas driving force $(y\text{-}y_I)$ and liquid driving force $(x\text{-}x_I)$ respectively.

If the resistance in gas phase is low, y_I is nearly equal to y. Hence, the resistance resides entirely in the liquid phase, mentioned liquid-film controlling process which occurs in absorption of a slightly soluble solute in the liquid phase. On the contrary, x_I is almost equal to x when the resistance in liquid phase is low, referred gas-film controlling process. This occurs in absorption of a very soluble solute in the liquid phase.

Due to the composition at the interface is difficult to measure hence overall volumetric mass-transfer coefficients are defined in terms of overall driving forces as shown in equation 2-6.

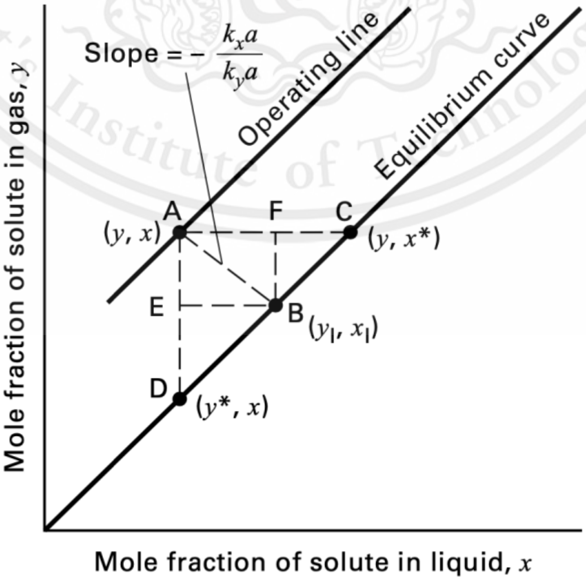


Figure 2.5 Interface composition in terms of the ratio of mass transfer coefficients.

$$r = K_x a(x^* - x) = K_y a(y - y^*) \qquad 2\text{-}6$$

where r = mass-transfer rate per unit volume of packed bed, kmol/m^3·h

K_x, K_y = overall mass transfer coefficient for liquid phase and vapor phase, kmol/m^2·h

a = mass transfer area per unit volume of packed bed, m^2/m^3

Additionally, y^* is the fictitious vapor mole fraction in equilibrium with the mole fraction while x^* is the fictitious liquid mole fraction in equilibrium with the mole fraction.

From Figure 2.2 (a), the material balance for the upper boundary becomes

$$L_{in}x_{in} + V_z y = V_{out}y_{out} + L_z x \qquad 2\text{-}7$$

where L_{in}, L_z = liquid flow rate at inlet and z position respectively, m^3/h

x_{in}, x = mole fraction of solute in liquid phase at inlet and z position

V_{out}, V_z = gas flow rate at outlet and z position respectively, m^3/h

y, y_{in} = mole fraction of solute in gas phase at inlet and z position

and a differential material balance around gas at the gas-liquid interface gives

$$K_y a(y - y^*)S\,\Delta z + GS\,y(z+\Delta z) = GS\,y(z) \qquad 2\text{-}8$$

where G = gas molar velocity, kmol/m^2·h

S = inside cross-sectional area of the column in unit of m^2

for dilute system in which the mole fractions of x and y are small and volume constant

$$K_y a(y - y^*)S\,dz = -V\,dy \qquad 2\text{-}9$$

Finally, the packed height can be calculated in term of integral form as

$$z_T = \frac{V}{K_y a S}\int_{y_{out}}^{y_{in}} \frac{dy}{y-y^*} \quad \text{2-10}$$

$$z_T = \frac{V}{K_y a S}\frac{(y_{in}-y_{out})}{(y-y^*)_M} \quad \text{2-11}$$

and

$$H_{OG} = \frac{V}{K_y a S} \quad \text{and} \quad N_{OG} = \int_{y_{out}}^{y_{in}} \frac{dy}{y-y^*} \quad \text{2-12}$$

where

$$(y-y^*)_M = \frac{(y_{in}-y^*_{in}) - (y_{out}-y^*_{out})}{\ln\left[(y_{in}-y^*_{in})/(y_{out}-y^*_{out})\right]} \quad \text{2-13}$$

$$y^*_{in} = H_A x_{out} \text{ and } y^*_{out} = H_A x_{in} \quad \text{2-14}$$

Besides, H_{OG} is represented the overall height of a gas transfer unit (HTU). The smaller the HTU, the more efficient the contacting. While N_{OG} is represented the overall number of gas transfer units (NTU), which means the larger the NTU, the greater the time or area of contact required.

2.2.3 Liquid holdup[10]

Liquid holdup is defined as the volume of liquid held per volume of the packed bed under operation conditions. In some case, such event can be summarized in three modes including static, total and operating liquid holdup. Static holdup is the amount of liquid remaining on packing that has been fully wetted and then drained. Whereas total holdup is amount of liquid on the packing under dynamic conditions. Operating holdup is measured experimentally as the difference between total and static holdup.

Data taken from Billet[11], relation of specific liquid holdup (h_L) and superficial gas velocity (u_V) was studied for air-water system. According to Figure 2.6, loading line, which occurs due to loading point for each superficial liquid velocity (u_L), is used

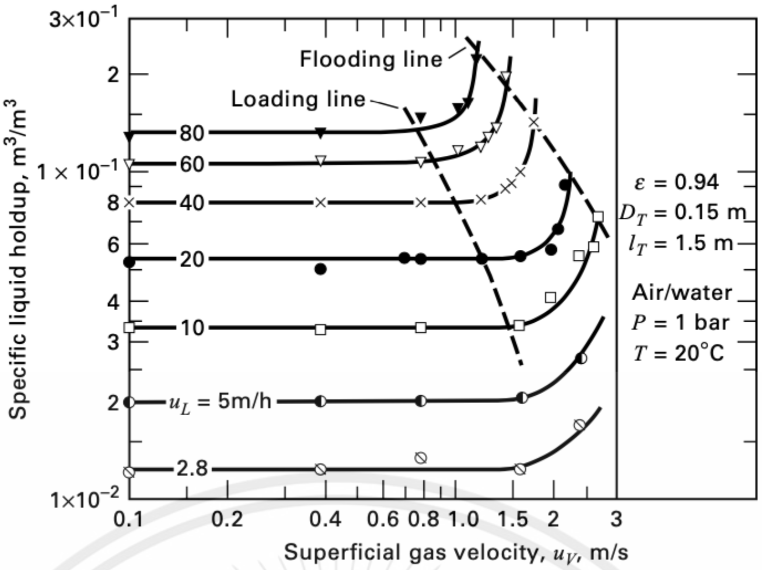


Figure 2.6 Specific liquid holdup for irrigated 25-mm metal Bialecki rings[1]

to portion a zone into preloading and loading zone. Before the loading point, specific liquid holdup independences on a superficial gas velocity, which means volume of liquid held at constant. Alternatively, when gas velocity limit is reached, downward flow of liquid is obstructed by flow of gas, and then liquid begins fill the bed, replacing gas. From this point, pressure drop rapidly increases, and liquid holdup also increases sharply, and mass-transfer efficiency decreases with increasing gas velocity.

Unless superficial gas velocity remains constant, liquid is continuous across the top of the packing and the column, referred flooded. At the flooding point, gas has a sufficient drag force to entrain the entire liquid. However, operating condition is chosen at the preloading region. A dimensionless expression for specific liquid holdup (h_L) in the preloading region depends on packing characteristics and viscosity, density, and superficial velocity of the liquid as expressed in equation 2-15.

$$h_L = \left(12\frac{N_{Fr_L}}{N_{Re_L}}\right)^{1/3} \left(\frac{a_h}{a}\right)^{2/3} \qquad \text{2-15}$$

where N_{Rr_L} is a liquid Reynold number, represented a ratio of inertial force to viscous force.

$$N_{Re_L} = \frac{u_L \rho_L}{a\mu_L} = \frac{u_L}{\nu_L} \qquad 2\text{-}16$$

whereas N_{Fr_L} is a liquid Froude number, represented a ratio of inertial force to gravitational force.

$$N_{Fr_L} = \frac{{u_L}^2 a}{g} \qquad 2\text{-}17$$

and the ratio of specific hydraulic area of packing (a_h) to specific surface area of packing (a) is given by

$$a_h/a = C_h N_{Re_L}^{0.15} N_{Fe_L}^{0.1} \text{ for } N_{Re_L} < 5 \qquad 2\text{-}18$$

$$a_h/a = 0.85\, C_h N_{Re_L}^{0.25} N_{Fe_L}^{0.1} \text{ for } N_{Re_L} \geq 5 \qquad 2\text{-}19$$

value of a and C_h are the characteristic of packing depends on size and material.

2.2.4 Flooding and pressure drop[12]

Liquid holdup, column diameter, and pressure drop are closely related. In addition, cross-sectional area of column depends on the fluid-flow limiting, which can be happened by limiting condition such as flooding. In 1954, Leva[13] used experimental data on ring and saddle packings to construct flooding correlations include lines of constant pressure drop, the resulting chart becomes known as the generalized pressure drop correlation (GPDC).

Correspond to Figure 2.7, liquid-to-gas kinetic energy ratio (F_{LV}) is received in order to calculate superficial gas velocity (u_V) at flooding point, which can be directly computed to column diameter from equation 2-20.

$$D_T = \left(\frac{4VM_V}{f\, u_{V,f}\, \pi \rho_V} \right)^{0.5} \qquad 2\text{-}20$$

where V = gas flow rate, kmol/h

M_V = molecular weight of mixed gas, kg/kmol

f = fraction of flooding (usually from 0.5 to 0.7)

ρ_V = gas density, kg/m^3

$u_{V,f}$ = gas velocity at flooding point based on cross sectional area of column, m/h

Likewise, Takahashi[12] demonstrated that the total pressure drop (ΔP_T) is the summarize of dry pressure drop (ΔP_d) and wet pressure drop (ΔP_w). Dry pressure drop is mainly originated from the friction of gas which flows upward through the void of packed bed, following the Fanning equation

$$\Delta P_d = 4f\left(\frac{1}{d_p}\right)\left(\frac{u_V}{\varepsilon}\right)^2\left(\frac{\rho_V}{2}\right) \qquad \text{2-21}$$

where f is a friction factor, which related to the Reynolds number of gas (Re_G).

$$f = 114\ Re_G^{-0.742} \text{ for } Re_G < 200 \qquad \text{2-22}$$

and

$$f = 6.85\ Re_G^{-0.216} \text{ for } Re_G \geq 200 \qquad \text{2-23}$$

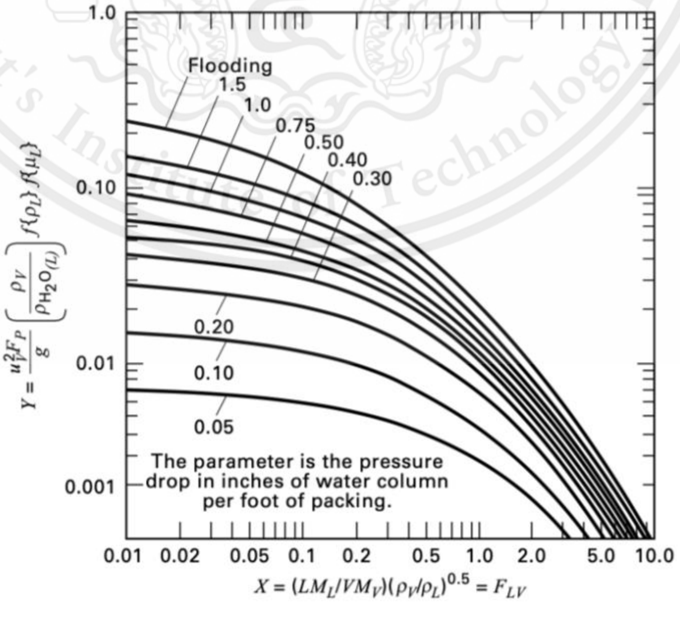


Figure 2.7 Generalized pressure-drop correlation of Leva for packed columns[1]

For wet pressure drop, the equation is experimentally expressed as

$$\Delta P_d = 4f\left(\frac{1}{d_p}\right)\left(\frac{u_V}{\varepsilon}\right)^2\left(\frac{\rho_V}{2}\right) \qquad \text{2-24}$$

Finally, total pressure drop is expressed as equation 2-25.

$$\Delta P_T = 4f\left(\frac{1}{d_p}\right)\left(\frac{u_V}{\varepsilon - h_L}\right)^2\left(\frac{\rho_V}{2}\right) + kh_L^3\left(\frac{u_V}{\varepsilon - h_L}\right)^2 \qquad \text{2-25}$$

where k = a experimental coefficient for various packing, depended on their types
d_p = packing size, m
f = fraction of flooding (usually from 0.5 to 0.7)
u_V = gas velocity at flooding point based on cross sectional area of column, m/h
h_L = liquid holdup, m^3/m^3
ε = void fraction

Pressure drop is a significant parameter used to indicate condition in the gas absorption column. General relation between pressure drop and proportion of gas volumetric flow rate to cross sectional area of column (superficial gas velocity) is shown in Figure 2.8. Dry specific pressure drop originates mainly from the friction of gas which rises through the void of the packed bed under conditions of no liquid flow. At preloading, region for difference liquid velocity, specific pressure drop is a straight line parallel to the dry specific pressure drop curve due to flow of gas does not affect to flow of liquid. After loading point, flow of gas affects to flow of liquid and liquid hold up. Increasing of specific pressure drop is sharply non-linear due to the rapid accumulation of liquid in the packing-void volume. Until liquid is completely resisted by gas, referred flooded.

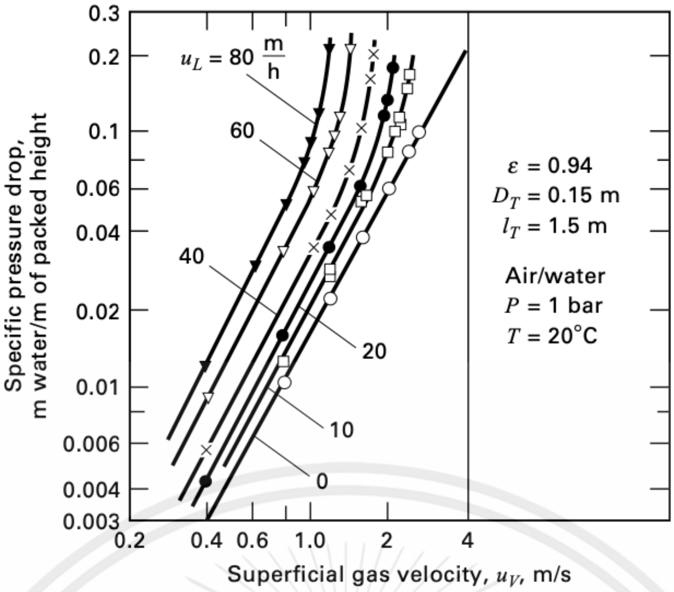


Figure 2.8 Specific pressure drop for dry and irrigated 25-mm metal Bialecki rings.[1]

2.3 Literature review

Ghaemi and Hemmati[14] studied regarding gas-phase mass transfer coefficient for piperazine (PZ), CO_2 and water system in a packed column. In addition, Buckingham pi-theorem was utilized to develop an overall dimensionless correlation. The main parameters were considered including molar gas flow, gas concentration and gas viscosity. The experiment was carried out under atmospheric pressure, CO_2 partial pressure of 18-66 kPa and temperature range of 40-100°C. In part of equipment, absorption column with 0.1 m inner diameter randomly filled with 16 mm stainless steel Pall rings was used. PZ solution and mixed gas contained N_2 and CO_2 were countercurrent fed into the absorption column and investigated with NDIR gas analyzer. The results indicated that the rising of temperature affects to the reduction of gas phase mass transfer coefficient due to increasing of CO_2 diffusivity in water. Furthermore, Schmidt number (Sc), which defined as the ratio of momentum diffusivity and mass diffusivity, is affected by CO_2 partial pressure. The mass transfer coefficient changes in the gas phase relative to the Schmidt number decrease. As the same way, Ghaemi and Hemmati confirmed that when liquid hold-up increases, there is an enhancement in the amount of gas phase mass transfer coefficient due to the increase in the gas−liquid interfacial area, because more liquid would be spread on the packing surface by rising in hold-up. Lastly, density-to-gas concentration ratio which roughly discussed, also has an upward trend with the gas phase mass transfer coefficient.

Nair and Selvi[15] researched regarding volumetric overall mass transfer coefficients for CO_2 absorption into aqueous ammonia solution using packed column filled with diameter of 8 mm ceramic Raschig ring. The experiment was carried out under up to 16 kPa partial pressure of CO_2, 61-214 $m^3/(m^2 \cdot h)$ gas flow rate, 0.75-3.06 $m^3/(m^2 \cdot h)$ liquid flow rate, 2%-16% ammonia mass fraction, and 20- 50°C temperature. The CO_2 concentration at the top of the column was determined by an infrared gas analyzer. The results showed that increasing of liquid flow rate directly affected to the volumetric overall mass transfer coefficients due to the wetted surface of the packing material for gas–liquid contact increased. The higher liquid flow rate leads to a higher liquid-side mass transfer coefficient in the case of liquid-phase controlled mass transfer. Besides, increasing in gas flow rate gave an under-expected higher volumetric overall mass transfer coefficients value excepting when the ammonia concentration is high owing to the ability to react with the liquid. Nevertheless, Aroonwilas and Veawab[16] observed that no significant effect on overall mass transfer coefficient for gas flow rate in the range of 46–97 kmol m^{-2} h^{-1} in amine-based aqueous solution. As the results of the mass transfer process which controlled by the resistance residing in the liquid phase, diffusion of solvent molecules within the liquid phase was restricted causing a constant amount of CO_2 absorbed.

Wang et al.[17] educated regarding properties of mass transfer over packing characterization and also focused on the measurement of three fundamental packing characteristics consisting effective gas-liquid contact area, gas phase and liquid phase film mass transfer coefficient. The experiment was operated in PVC column which had an inner diameter of 0.42 m and a packed height of 3.05 m. Since CO_2 mixed with air was countercurrent flew upward whereas 0.1 gmol/L of NaOH was pumped in a closed loop and distributed at the top of the column. A Horiba infrared carbon dioxide analyzer is used to evaluate CO_2 concentrations. The end results indicated that the effective area which indirectly affected by packing characteristic increased sharply with liquid flow rate and showed a temperate dependence with liquid flow rate. However, the effective area essentially independent of gas flow rate especially random packing. For diversity of packing materials, plastic random packing was compared with those in metal type. The plastic packing provided 20% less fractional area, which expressed as a ratio of effective area and specific area of packing, than the metal packings. Wang et al. concluded that this might be due to the increased hydrophobicity of polypropylene relative to stainless steel combined with the aqueous nature of the experiments.

Sovilj et ai.[18] studied regarding determination of the hydrodynamics characteristic and mass transfer within packed bed absorption column. The experiment was carried out under atmospheric pressure of 101.9 kPa, and temperature inside the absorption column of 14°C. Absorption column with 0.1 m inner diameter, 0.35 m height. The packing in the column consisted of 15 mm Raschig rings with specific surface area of 330 m^2/m^3 and the porosity of 0.70 was used. In part of experimental, water and mixed gas of air contained 8-16%mol CO_2 were countercurrent fed into the absorption column and analyzed by standard titration with a 0.1M NaOH solution. The results indicated that the overall mass transfer coefficient of the air-CO_2-water system depends on the flow rate of gas and the composition of CO_2 in the intel gas mixture. With increasing of gas flow rate, with a constant composition of CO_2 in the starting gas mixture, the overall mass transfer coefficient increases. Whereas, increasing of the amount of CO_2 in the gas mixture at a constant flow rate of fluid leads to decreasing of the overall mass transfer coefficient.

In addition, pressure drop was interested. It was found that pressure drop of air on the air velocity through the dry charge was shown to be in the form of a quadratic function. Relations of the pressure drop and the air velocity during the countercurrent flow was shown in 3 phases including 1) the area of low load, at low apparent air velocities, 2) the area of high load, at higher air velocities, and 3) the flooding area, when the liquid completely fills the beds.

CHAPTER III

RESEARCH METHODOLOGY

3.1 To change and rearrange gas absorption apparatus

1. Study the operating principle of the gas absorption unit in the laboratory
2. Inspect and analyze any defect that may occur within the column such as gas leakage
3. Resolve the defect such as reset the column and rearrange equipment
4. Check the readiness of using the gas absorption column

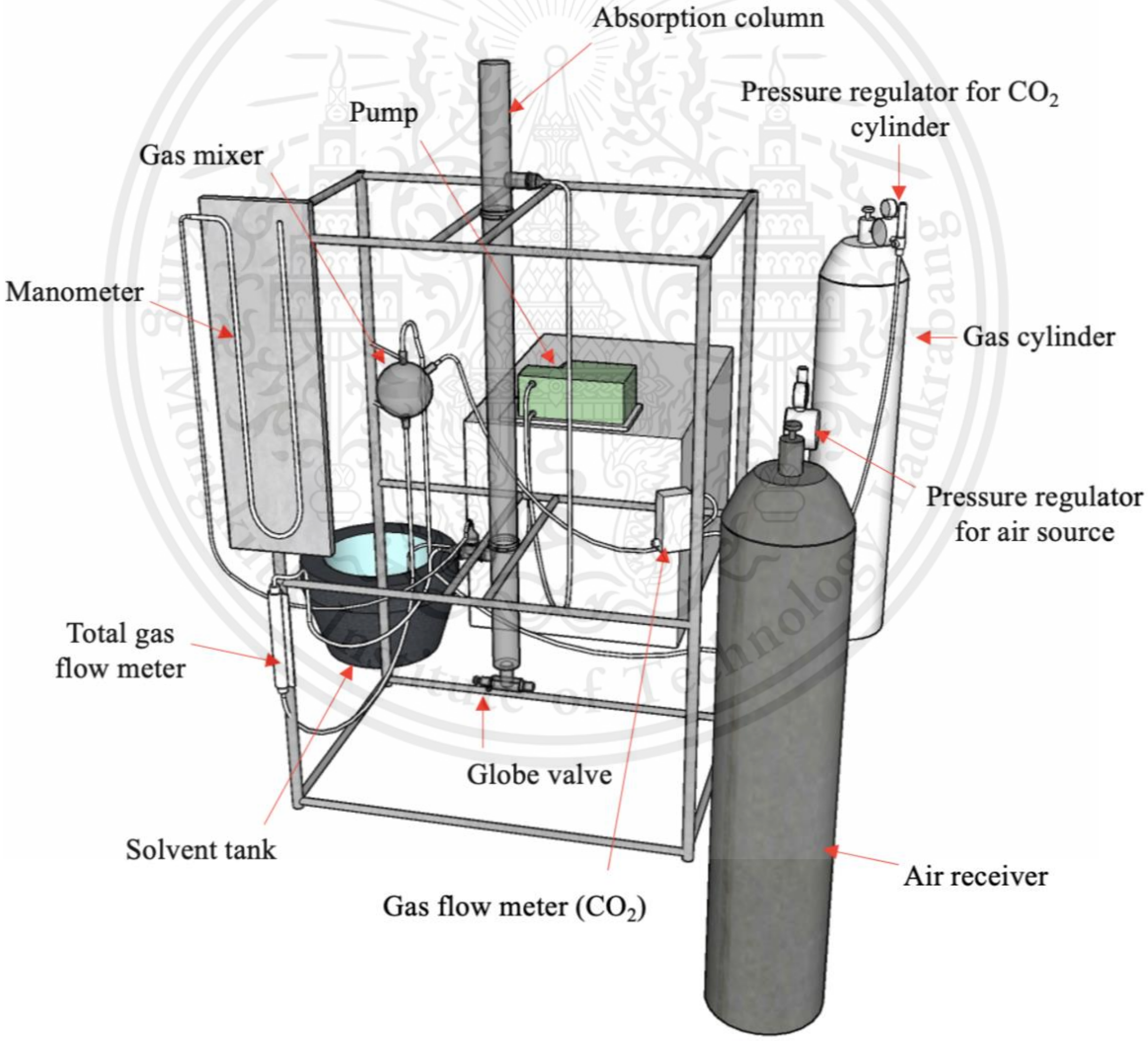


Figure 3.1 Pre-improved gas absorption unit in laboratory

3.2 To study of operating limitation

1. Check the water level in manometer as zero
2. Adjust gas flow rate to 300, 500, 700, 900, 1,100, 1,300 and 1,500 lit/h then record pressure drop
3. Adjust liquid flow rate in range of 40-55 lit/h then change gas flow rate at 300-1,500 lit/h and record pressure drop
4. Increase gas flow rate to estimate flooding point then record pressure drop
5. Construct graph regarding liquid flow rate (x axis) and pressure drop (y axis) to consider the gas flow rate at loading and flooding point

3.3 To study effects of parameters changing to the overall mass transfer coefficient

Chemicals

1. Sodium hydroxide (NaOH) solution at concentration 0.01 M
2. Phenolphthalein
3. RO Water
4. Pure CO_2

Glassware and apparatus

1. Absorption column
2. 3 of 50 ml Beaker
3. 14 of 50 ml Erlenmeyer flask
4. 14 of 50 ml sample bottles
5. 25 ml Pipette
6. Burette
7. Burette clamp
8. Funnel
9. Pipe pump
10. Water tank

Procedure

1. Check the water level in manometer as zero
2. Adjust flow rate of air and CO_2 to achieve the total gas flow rate and CO_2 concentration as set in Table 3.1

Table 3.1 Experiment plan

Condition	Liquid flow rate (lit/h)	Gas flow rate (lit/h)	CO_2 concentration (%mol)
1	15	800	15
2	15	800	25
3	15	1,000	15
4	15	1,000	25
5	20	800	15
6	20	800	25
7	20	1,000	15
8	20	1,000	25

3. Adjust speeds of pump to get the appointed liquid flow rate, which can be calculated from calibration curve in Figure 3.2

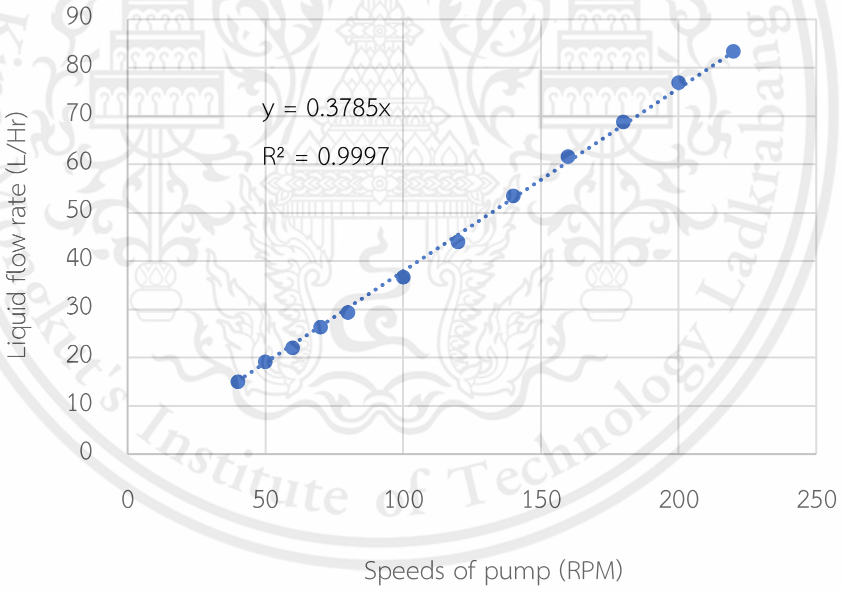


Figure 3.2 Calibration curve for speeds of pump and liquid flow rate

4. Time when liquid begins contact with packing and collect samples at the outlet at 2, 4, 6, 8, 10, 15 and 20 minutes
5. Pipet samples of 25 cm^3 and add 1-2 drops of phenolphthalein then titrate with 0.01 M NaOH solution

6. Calculate overall mass transfer coefficient (K_y) from obtained CO_2 concentration in 5.

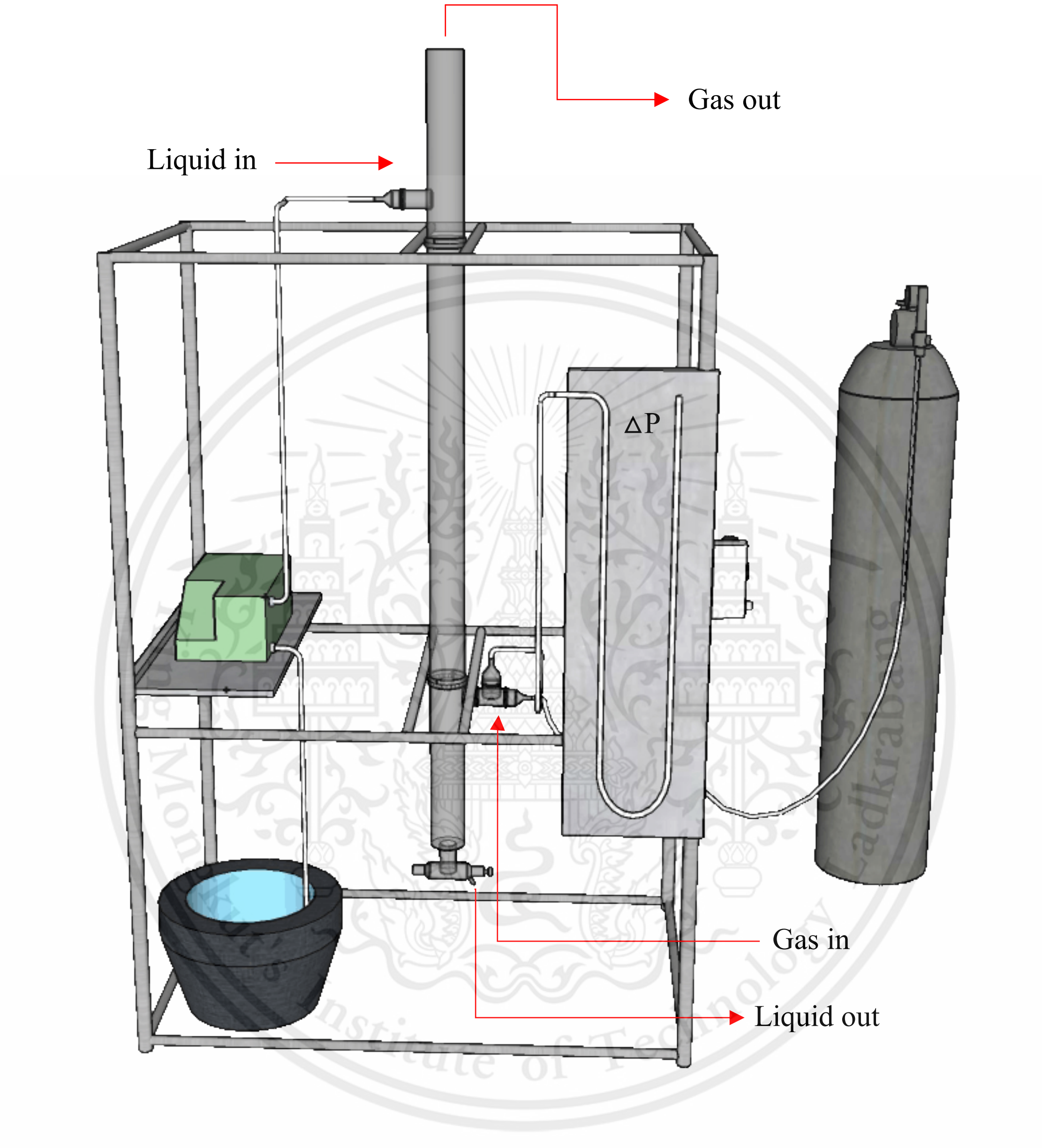


Figure 3.3 Gas absorption diagram with countercurrent flow

CHAPTER IV

RESULTS AND DISCUSSION

4.1 Improvement of CO_2 removal efficiency

From the examination, there were several defects that occurred within the column as shown in Figure 4.1. Firstly, tee fitting and check valve were installed instead of gas mixer to get a well-mixed gas at point 1. Flow of air and CO_2 capable to use a small mixing space.

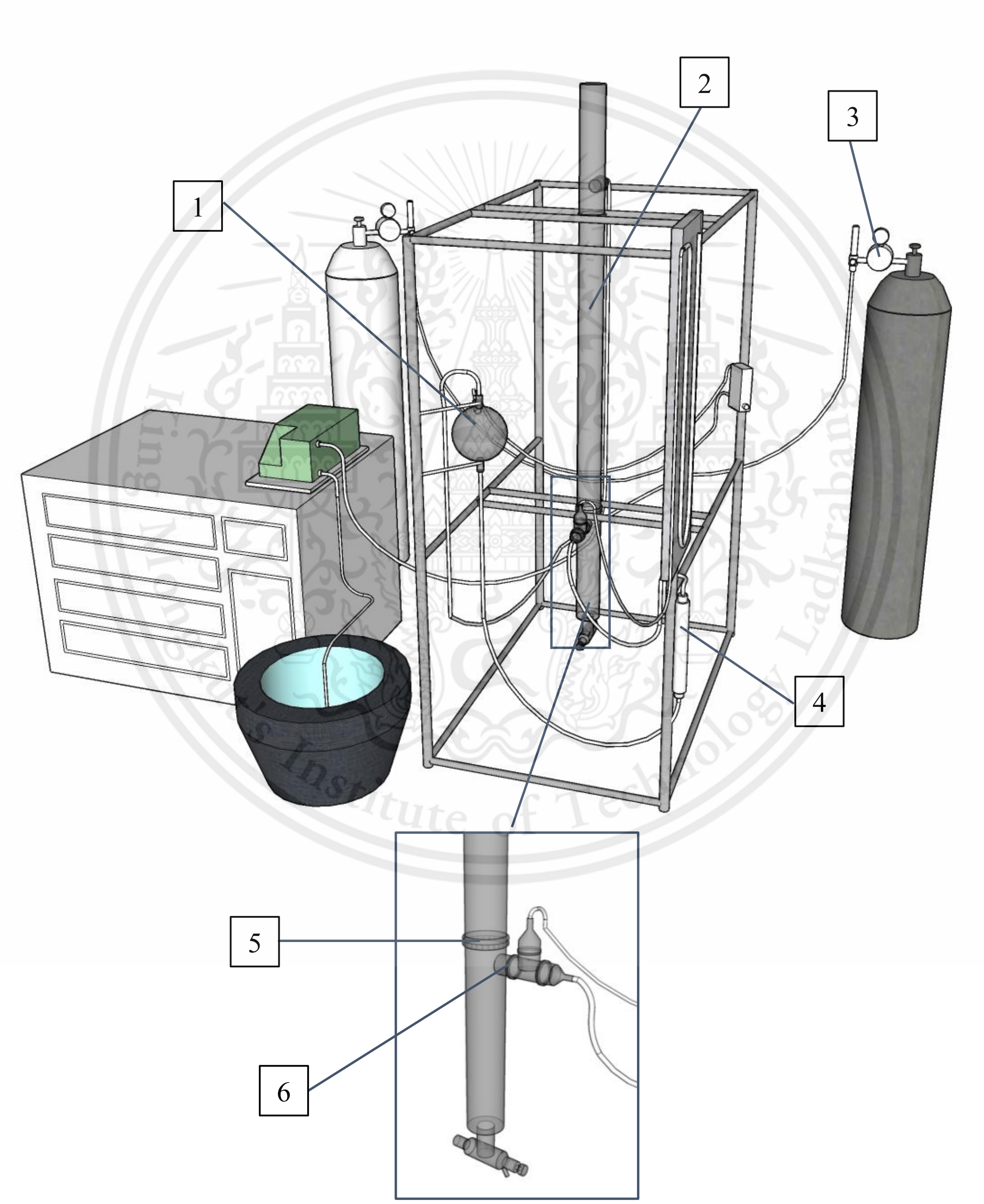


Figure 4.1 The summary defects found in pre-improved gas absorption unit

Check valves were installed on each of the individual gas streams to prevent mixing of the gases in the original source.

At point 2, packing in absorption column were cleaned to remove the dirtiness such as moss found commonly in moist shady locations. Surface area of packing was decreased by existing of the dirtiness. Commonly, the mass transfer coefficient increased with increasing of surface area due to packing is instrumental in creating longer gas–liquid contact time during CO_2 absorption process. Besides, meaningly decreasing of void fraction easily caused of flooding, or unstable condition in column.

Generally, when labs in the department were simultaneously done, flow of air often unstable and uncontrollable. Hence, pressure regulator for air receiver at point 3 was changed. Moreover, rotameter of air, which presented in point 4, was examined and found that flow reading was approximately inaccurate of 30%. For these results, rotameter with flow regulator for air was installed instead of the previous. To effortlessly usability, the control panel will be done to garter the flow regulator of carrier gas and solute.

At point 5, it was found that the joints of the gas absorption column equipment were connected by plastic gaskets, which may be the cause of gas leaks. Therefore, the resetting of the column and checking the gasket was done. Additionally, the main cause of the deviation is a gas leak found at the gas entrance connected to the packed column due to the glass connected equipment had been broken, which is point 6. Therefore, the broken equipment was changed. Prior to the improvement of the gas absorption column, there was a small amount of gas that was able to pass through the column. A significant amount of gas has leaked out, causing tiny CO_2 removal efficiency and inaccurate overall mass transfer coefficient.

Besides, manual of gas absorption unit had been amended. To prevent gas leaking out at below of the column, liquid at the bottom has to remain at the given level. Hence, the manual has to clearly specify the above requirements in the procedure.

Finally, the improved gas absorption unit, shown in Figure 4.2, was completed. For CO_2 removal efficiency comparison, the experiments were carried out under ambient condition. Data achieved from the gas absorption unit including outlet CO_2 concentration in gas and liquid phases, the overall mass transfer coefficient and CO_2 removal efficiency were expressed in Table 4.1.

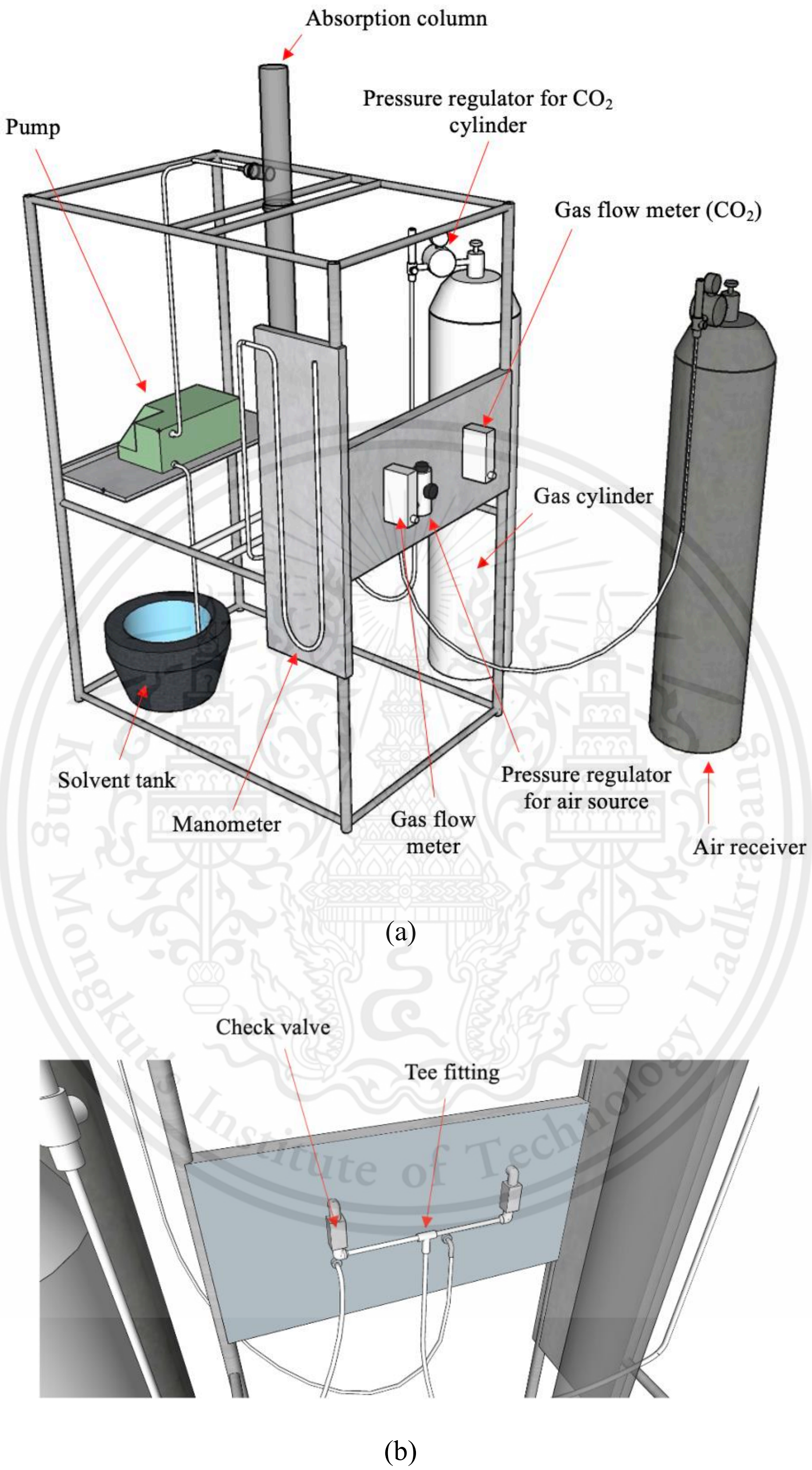


Figure 4.2 (a) Overall (b) The back of the control panel of improved gas absorption column

The improvement of CO_2 removal efficiency by rearrangement of the gas absorption component in the laboratory. The results showed in Table 4.1 presented that the CO_2 removal efficiency increased from 0.003% to 0.258% for the first condition and obviously improved likewise second and third condition. Since, outlet CO_2 concentration in gas phase were achieved from material balance as equation 2-7.

$$L_{in}x_{in} + V_{in}y_{in} = L_{out}x_{out} + V_{out}y_{out} \qquad 2\text{-}7$$

Additionally, CO_2 removal efficiency is defined as

$$\%\ CO_2\ \text{removal} = \left(\frac{y_{in} - y_{out}}{y_{in}}\right) \times 100 \qquad 4\text{-}1$$

Table 4.1 Comparison of CO_2 removal efficiency before and after improvement of gas absorption laboratory

Condition		1st	2nd	3rd
- Gas flow rate (lit/h)		800	800	1,000
- Liquid flow rate (lit/h)		15	20	20
- Inlet CO_2 concentration, y_{in} (%mol)		25	25	15
Results				
Outlet CO_2 concentration, x_{out} (M)	before	0.000018	0.000020	0.000015
	after	0.001380	0.001700	0.000940
Outlet CO_2 concentration, y_{out} (% mol)	before	24.99916	24.99875	14.99925
	after	24.93540	24.89390	14.95312
K_y ($mol/m^2 \cdot h$)	before	0.000739	0.001095	0.001370
	after	0.061866	0.103990	0.094885
CO_2 removal efficiency (%)	before	0.003	0.005	0.005
	after	0.258	0.424	0.313

4.2 Predicted operating condition in packed column by pressure drop

According to Figure 4.3. represents relation of pressure drop depends on superficial gas velocity (u_v) and superficial liquid velocity (u_L), which including dry pressure drop (a line), at liquid velocity of 27.5 m/h (b line), 31.0 m/h (c line), 34.4 m/h (d line) and 37.9 m/h (e line) respectively. The results demonstrate that pressure drop for dry packing is almost a straight line which conforms to Ergun's equation presented in equation 4-2. For identical column, pressure drop depends on gas velocities. At low superficial gas velocities, the pressure drop per unit height is proportional to the superficial vapor velocity, u_V. At high velocity, corresponding to turbulent flow, the pressure drop per unit height depends on the square of the gas velocity.

$$\frac{\Delta P}{z_T} = \frac{150\mu(1\text{-}\varepsilon)^2 u_V}{d_p{}^2\varepsilon^3} + \frac{1.75\rho(1\text{-}\varepsilon)\, u_V{}^2}{d_p\varepsilon^3} \qquad 4\text{-}2$$

Figure 4.3 shows conditions inside the absorber under changes of superficial gas velocity at given superficial liquid velocity. For instance, superficial liquid velocity of 27.5 m/h (b line) at the low superficial gas velocity (0.06-0.17 m/s), pressure drop is linear consistent with dry pressure drop. After that pressure drop non-linearly increases with increasing of superficial gas velocity, which reaches to loading zone. At high superficial gas velocity (0.19-0.21 m/s), liquid is completely hindered by flow of gas and void of packing are fully filled, called flooded. Pressure drop at flooding point is high and unstable.

Moreover, Figure 4.3 depicts the comparison of pressure drop changed by superficial gas velocity and superficial liquid velocity. Pressure drop increases with increasing of liquid flow rate and gas flow rate which conforms to equation 2-24.

$$\Delta P_d = 4f\left(\frac{1}{d_p}\right)\left(\frac{u_V}{\varepsilon}\right)^2\left(\frac{\rho_V}{2}\right) \qquad 2\text{-}24$$

The fact that a greater liquid flow rate desired a lesser gas flow rate for flooding. For instance, using of liquid volumetric flow rate at 31.0 m/h (c line) begins to flood at gas

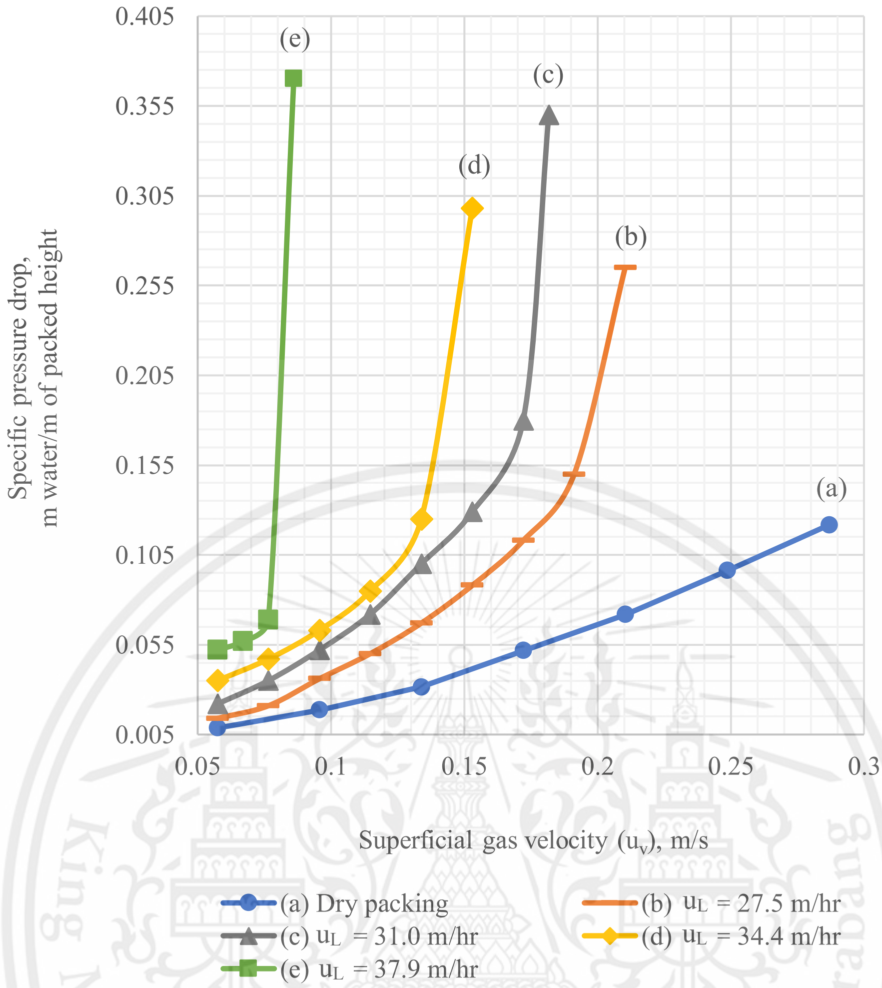


Figure 4.3 The pressure drop of gas in the countercurrent flow of gas and liquid as a function of gas velocity, with the liquid flow rate as a parameter

flow rate of 0.18 m/s while using of liquid volumetric flow rate at 37.9 m/h (e line) begins to flood at gas flow rate of 0.09 m/s. Because of liquid holdup, which defined as the amount of liquid on the surface of packing materials or in the form of liquid drops in the space of the packing materials, is high in accordance with high liquid flow rate and results in elevated pressure drop.

4.3 Factors affecting the overall mass transfer coefficient

The effects of different operating and design parameters including gas flow rate, liquid flow rate and inlet CO_2 concentration were examined. The study variables which affect to the overall mass transfer coefficient for air-CO_2-water system is further discussed in the following subsections. The results shown in Table 4.2.

Table 4.2 The overall mass transfer coefficient varied by gas and liquid flow rate and inlet CO_2 concentration

Cases	**Condition**			**Results**
	Gas flow rate (lit/h)	Liquid flow rate (lit/h)	Intel CO_2 concentration (%mol)	Mean K_y (mol/m^2·h)
1	800	15	15	$0.094 \pm 3.22\times10^{-6}$
2			25	$0.059 \pm 8.54\times10^{-6}$
3		20	15	$0.132 \pm 5.93\times10^{-6}$
4			25	$0.101 \pm 1.16\times10^{-5}$
5	1,000	15	15	$0.075 \pm 1.85\times10^{-4}$
6			25	$0.073 \pm 4.11\times10^{-6}$
7		20	15	$0.090 \pm 3.07\times10^{-5}$
8			25	$0.113 \pm 4.90\times10^{-7}$

4.3.1 Effect of liquid flow rate

The results showed in Figure 4.4 indicated that the overall mass transfer coefficient is higher when increasing of liquid flow rate. Due to increasing of liquid flow rate results in the rising of liquid holdup owing to the more liquid would be spread on the packing surfaces. As a result, the wetted surface of the packing material for gas–liquid contact increased and led to the higher overall mass transfer coefficient.

Moreover, rate of mass transfer can be increased by promoting turbulence in the controlling phase and increasing the dispersion. Due to CO_2 slightly dissolve in liquid, the studied system is classified to be liquid-film controlling process. Therefore, liquid flow rate is a significant parameter for the overall mass transfer coefficient. According to the combination of equation 2-4 to 2-6, the overall resistance can be expressed as shown in equation 4-2.

$$\frac{1}{K_y} = \frac{1}{k_y a} + \frac{m}{k_x a} \qquad 4\text{-}3$$

m is a slope of the equilibrium line, which represented to K-value for vapor-liquid equilibrium ratio. In fact that CO_2 is a slightly soluble solute in water. Consequently, m is high value then the liquid film resistance is a controlling parameter.

$$\frac{1}{k_y a} \ll \frac{m}{k_x a}$$

then

$$\frac{1}{K_y} \approx \frac{m}{k_x a}$$

Liquid film controlling means that any change in $k_x a$ has a nearly proportional effect on the overall mass transfer coefficient, whereas a change in $k_y a$ has a slight effect. Hence, rate of mass transfer can be increased by promoting turbulence in liquid phase and increasing the dispersion of liquid.

However, there is a limitation for increasing of liquid flow rate. As previous mentioned, operating condition is chosen at the preloading region. Therefore, range of

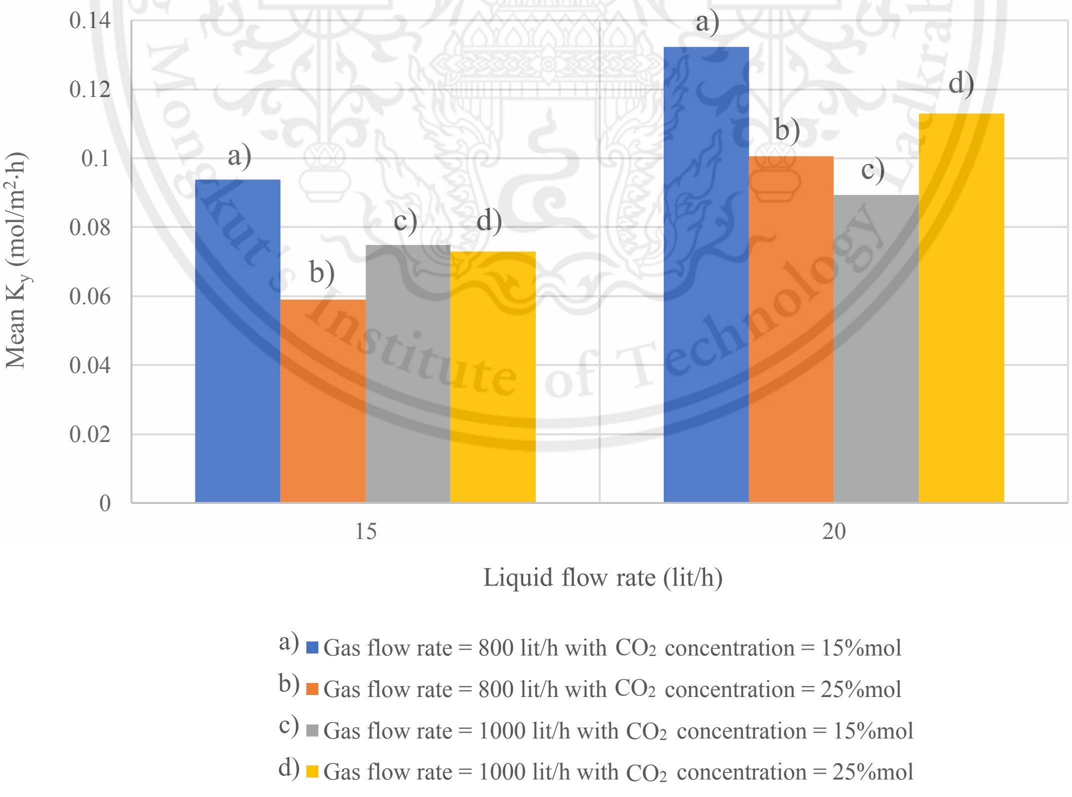


Figure 4.4 Effect of liquid flow rate to the overall mass transfer coefficient

liquid flow rate and gas flow rate have to be carefully selected in order to prevent loading and flooding operated region.

4.3.2 Effect of inlet CO_2 concentration

Trend of the overall mass transfer coefficient obviously is differing for CO_2 concentration 15 and 25%mol as presented in Figure 4.5. Increasing of CO_2 concentration cause of decreasing of the overall mass transfer coefficient. Mustafa et al.[19] reported that this may cause by the restricted diffusion between CO_2 in gas phase.

Generally, higher CO_2 content in the inlet gas resulted in a higher driving force for absorption owing to the more CO_2 molecules to transfer from gas bulk to gas-liquid interface. According to the results, the cause of this behavior is the restricted diffusion in liquid phase basically results in a constant amount of CO_2 absorbed. That is the overall mass transfer coefficient is conversely proportional to the driving force.

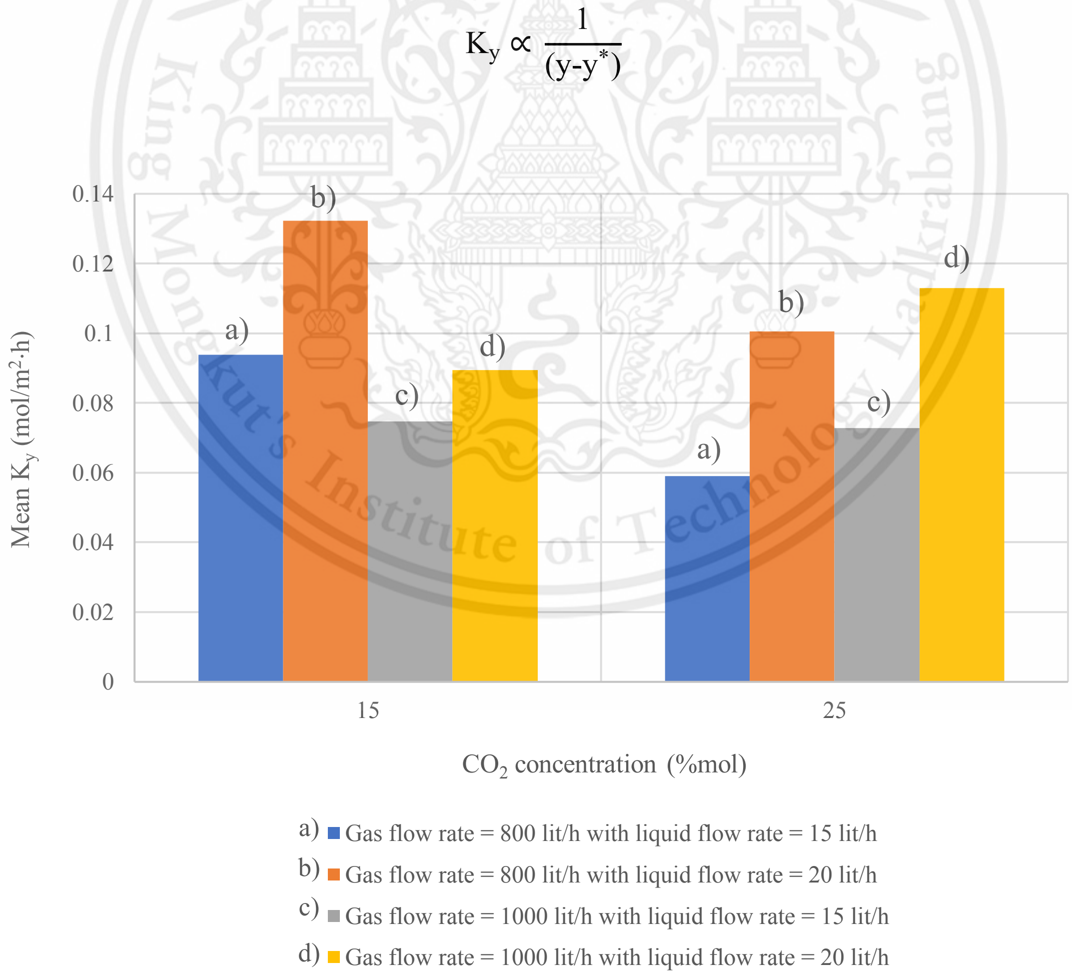


Figure 4.5 Effect of CO_2 concentration rate to the overall mass transfer coefficient

The term rate mass-transfer, r in equation 2-6 is constant hence higher driving force (y-y*) cause in decreasing of the overall mass transfer coefficient.

$$r = K_x a(x^* - x) = K_y a(y - y^*) \qquad 2\text{-}6$$

Nevertheless, the results in case of increasing of CO_2 concentration at high gas and liquid flow rate provided the higher overall mass transfer coefficient increases. It may be due to the driving force of gas flow rate and CO_2 concentration have sufficient force to enhance resistance in gas-film. Whereas higher liquid flow rate led to poorer resistance in liquid-film. Hence, rate of mass transfer increased and influenced on higher the overall mass transfer coefficient.

4.3.3 Effect of gas flow rate

According to Figure 4.6, CO_2 concentration of 15%mol, the overall mass transfer coefficient significantly reduces by increasing of volumetric gas flow rate. This may cause by decreasing of residence time obtained by divided reactor volume by the volumetric flow rate entering the reactor. The less residence time of gas phase found in high gas flow rate is designated to be a causation of adverse mass transfer process. Hence, a longer time is required for higher gas-liquid contacting time for achieving a greater mass transfer.

In addition, Figure 4.6 depicted that rising of gas flow rate results in the less overall mass transfer coefficient for CO_2 concentration at 15%mol. Tan et al.[20] deducted that the study, which controlled by the resistance residing in the liquid phase, found the restricted diffusion within liquid phase. Therefore, this causing a constant amount of CO_2 absorbed regardless of the gas load values. According to the results, the higher gas flow rate possibly could entrain part of liquid and results in reducing of liquid flow rate spreading on the packing surface.

In addition, similar to case of increasing CO_2 concentration, pressure of system is improved by rising of gas flow rate. Subsequently, partial pressure of CO_2 is increased as well. Gao et al.[21] reported that increasing inert gas flow rate can lead to the acceleration of the renewal of CO_2 in the gas-liquid interface, causing a greater bulk partial pressure.

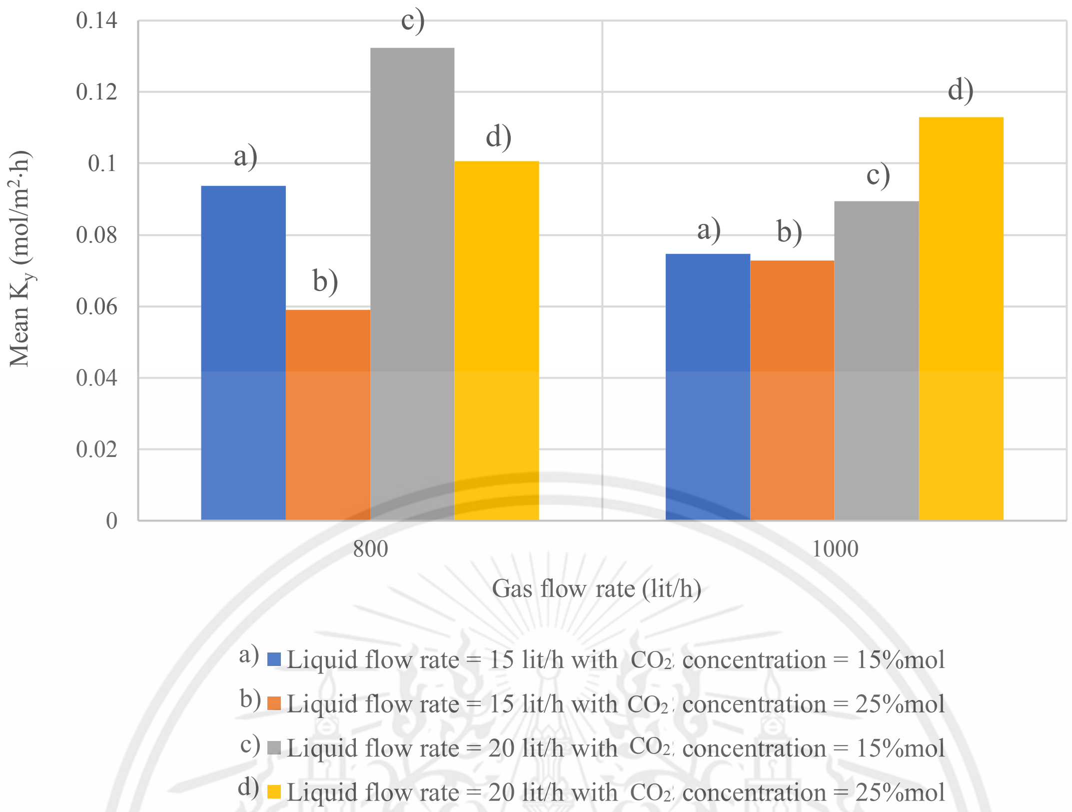


Figure 4.6 Effect of gas flow rate to the overall mass transfer coefficient

Driving force in terms of partial pressure gradient (p-p*) has an influence on the overall mass transfer coefficient. As earlier mentioned, the overall mass transfer coefficient is conversely proportional to the driving force in case of restricted diffusion in liquid phase and mass-transfer rate constant. In short, rising of gas flow rate results in the less overall mass transfer coefficient.

Nevertheless, rising of volumetric gas flow rate influences on higher the overall mass transfer coefficient for CO_2 concentration of 25%mol. Mass transfer flux is inversely proportional to thickness of gas film. The thickness of the concentration boundary layer on the gas side at the gas-liquid interface becomes thinner when gas flow rate is risen.

Additionally, at high CO_2 concentration, it may be due to the driving force of gas flow rate and CO_2 concentration have significant force to improve resistance in gas-film. Hence, rate of mass transfer increased and influenced on higher the overall mass transfer coefficient.

CHAPTER V

CONCLUSION

The study results of pressure drop varies by superficial liquid velocity and superficial gas velocity for air-water system in packed column. To summarize, pressure drop is a significant parameter used to indicate condition in the gas absorption column. Theoretically, condition is classified into three regions including preloading, loading, and flooding region. Pressure drop at preloading region is a straight line parallel to the dry pressure drop curve. Afterward, flow of gas strongly affects to flow of liquid and liquid hold up. Apparent increasing of pressure drop is sharply non-linear due to the rapid accumulation of liquid in the packing, occurs at loading region. Velocity of gas was increased until liquid is completely resisted by gas, referred flooded.

Studied effects for the overall mass transfer coefficient are including gas flow rate, liquid flow rate and inlet CO_2 concentration. Due to the studied system is a liquid film controlling hence liquid flow rate is determined to be a strongest main effect for the overall mass transfer coefficient. With increasing of liquid flow rate from 15 lit/h to 20 lit/h, the overall mass transfer coefficient obviously increased.

Differ from effect of the inlet CO_2 concentration varied by 15%mol and 20% mol and effect of gas flow rate varied by 800 lit/h and 1,000 lit/h. The studied results indicated that the overall mass transfer coefficient is reduced by increasing of CO_2 concentration and gas flow rate. The possible cause may be the restricted diffusion in liquid phase basically results in a constant amount of CO_2 absorbed. That is the overall mass transfer coefficient is conversely proportional to the driving force. Higher driving force is an imaginable cause in decreasing of the overall mass transfer coefficient.

Conclusively, the improvement of CO_2 removal efficiency by rearrangement of the gas absorption component in the laboratory. It was found that there were two main causes of the deviation including gas leak by breakage of the connected equipment and dirtiness of packing. In case of gas leak, an under-expected amount of gas was able to pass through the column causing unreliable CO_2 removal efficiency.

In case of dirtiness, moss significantly reduced surface area and void fraction of packing. Therefore, dirty packing created not enough longer gas–liquid contact time during CO_2 absorption process. This behavior affected to the low CO_2 removal efficiency and high pressure drop in absorption column.

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APPENDIX

APPENDIX A
GAS ABSORPTION MANUAL LABORATORY
GAS ABSORPTION

1. Objective

To understand and determine the overall mass transfer coefficient between gas and liquid phases in packed column as well as liquid holdup, flooding and observe an emerged pressure drop.

2. Theory

Gas absorption is a vapor-liquid separation process which consisting at lease 3 components, soluble solute (absorbate), suitable solvent (absorbent) and carrier gas by using gas solubility as a main property. Due to the difference in concentration of solute between carrier gas and solvent phase, mass transfer of solute certainly occurs. Absorbers is mainly packed columns and trayed towers (plate columns), and less often spray towers, centrifugal contactors and bubble columns, respectively. Advantage and disadvantage and configuration of each absorber are shown in Table 1 and Figure 1 as below.

Table A.1 Comparison of types of absorption column

Types of absorption column	Advantages	Disadvantages
Packed column	1. Superior contact between gasses and liquid that promote high mass transfer rate 2. Low capital, operating, and maintenance cost	1. Channeling, which must be controlled by redistributing liquid 2. Cannot handle extremely high or low flow rates
Trayed towers	1. Can handle high or low liquid flow rates cost effectively 2. Easily customized to the specific requirements such as operations requiring much heat	1. Higher pressure drop than packed column 2. Slow reaction rate processes 3. Plugging and fouling may occur

Spray towers	1. Low pressure drop 2. Most effective for solutes with high liquid solubility	1. High pumping cost 2. Poor mass transfer 3. Low residence time
Bubble columns	1. Suitable for reaction which requires large residence time 2. Low maintenance and operating cost are required	1. High vapor pressure drop 2. Low vapor throughput
Centrifugal contactors	1. High mass transfer rates can be achieved 2. Favored when no headroom for tray or packed column	1. Vary low residence time 2. Low number of contact stages

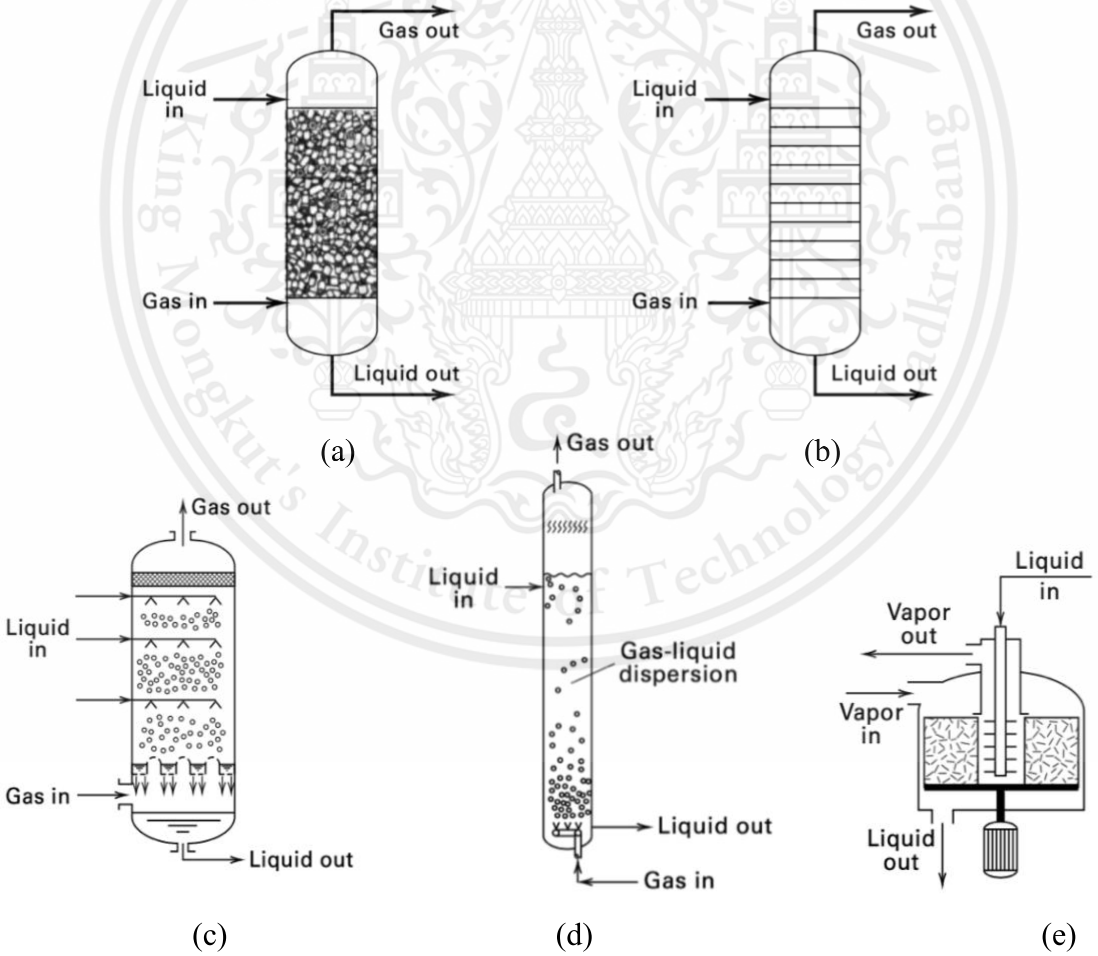


Figure A.1 Type of absorption columns: (a) packed column; (b) trayed tower; (c) spray towers; (d) bubble column; (e) centrifugal contactor[1]

Operating limitation; loading and flooding region

Operating limitation for packed column is flooding which can avoid by the control of liquid and gas flow rate. Generally, in absorption column, the difference of liquid and gas flow rates results in 3 regions, which can be occurred, preloading, loading or flooding. Liquid and gas flow rate should be operated at the preloading region where located before loading region due to the high efficiency and stability of column. Between the loading and flooding points is the loading region. Finally, if superficial gas velocity is built up excessively beyond loading point, operating condition in column is reached to flooding. Liquid holdup and specific pressure drop are a well indicator used to indicate operating region in absorption column.

Liquid holdup is defined as the volume of liquid held per volume of the packed bed under operation conditions. According to Figure 2, at preloading region, liquid holdup independences on a superficial gas velocity. Alternatively, at loading point, downward flow of liquid is obstructed by flow of gas, and then liquid begins fill the bed, replacing gas. From this point, liquid holdup increases sharply, and mass-transfer efficiency decreases. At flooding region, gas has a sufficient drag force to entrain the entire liquid. Hence, liquid is continuous across the top of the packing and the column, referred flooded.

Figure 3 shows specific pressure drop for a given liquid velocity. Dry specific pressure drop originates mainly from the friction of gas which rises through the void of the packed bed under conditions of no liquid flow. At preloading region for difference liquid velocity, specific pressure drop is a straight line parallel to the dry specific pressure drop curve. After loading point, increasing of specific pressure drop is sharply non-linear due to the rapid accumulation of liquid in the packing-void volume.

Specific pressure drop at flooding (ΔP_{flood}) in units of inches H_2O/ft of packed height can be calculated by the empirical expression [Kister & Gill, 1992].

$$\Delta P_{flood} = 0.115\ F_p^{0.7} \qquad \text{A-1}$$

where F_p which is packing factor (ft^2/ft^3) depends on types of packing.

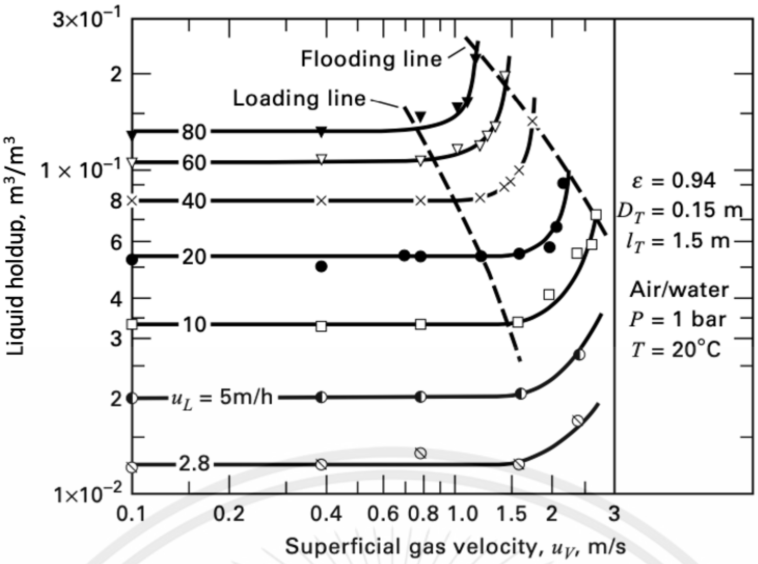


Figure A.2 Liquid holdup for irrigated 25-mm metal Bialecki rings[1]

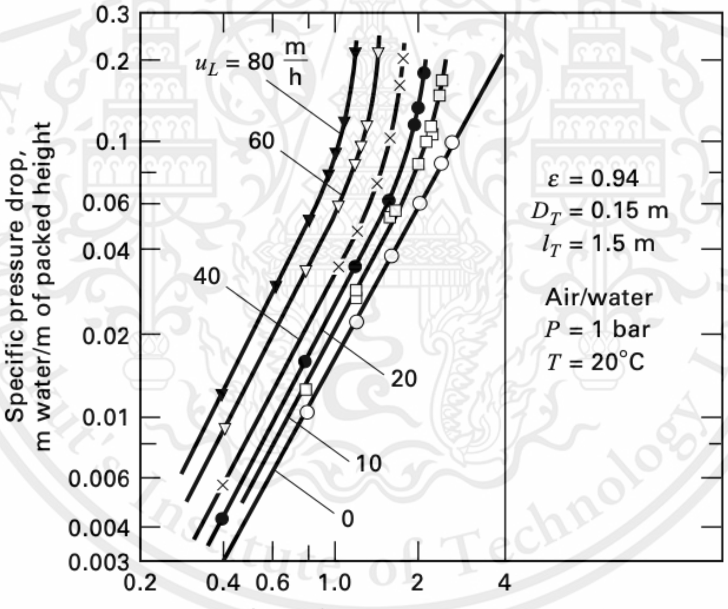


Figure A.3 Specific pressure drop for dry and irrigated 25-mm metal Bialecki rings[1]

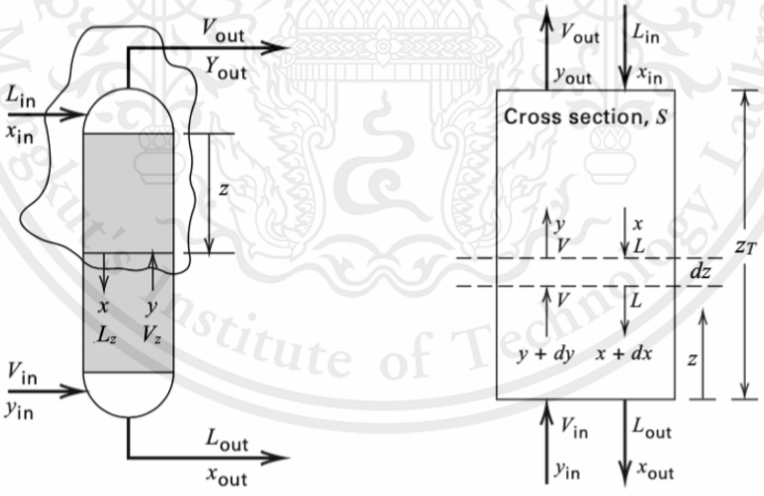
Packing

Packing is instrumental in creating longer gas–liquid contact time during gas absorption process. Type and material of packing directly affected to the contacting between liquid and gas, and also pressure drop.

Table A.2 Example of packing

Packing	Raschig rings 	Intalox saddles 	Pall rings 	
Material	Glass	Ceramic	Plastic	Plastic
Size, d_p (mm)	4	35	35	50
Packing factor, F_p (ft^2/ft^3)	1589	50	40	26
Area/volume, a (m^2/m^3)	1007.0	114.6	151.1	111.1
Void fraction, ε (m^3/m^3)	0.578	0.761	0.906	0.919

Rate-based method for packed column



(a) (b)

Figure A.4 Material balance in packed columns with countercurrent flow (a) Overall (b) Differential contact for absorption[1]

The packed-tower absorption column consists of a hollow vertical tower, usually cylindrical in cross-section, filled with a loosely fitting packing which allows easy flow of both gas and liquid streams. In Figure 4, flow rate of inlet gas (V_{in}) from the bottom of column contains solute with mole fraction y_{in} flows upward and then purged at the top as the outlet gas flow rate (V_{out}) with the solute mole fraction y_{out}. Whereas flow rate of inlet solvent (L_{in}) which contains solute with mole fraction x_{in} is fed from top and then flows downward to the bottom as flow rate of solvent outlet (L_{out}) with mole fraction x_{out}.

$$L_{in}x_{in} + V_{in}y_{in} = L_{out}x_{out} + V_{out}y_{out} \qquad \text{A-2}$$

where L_{in}, L_{out} = liquid flow rate at inlet and outlet respectively, mol/h
V_{in}, V_{out} = gas flow rate at inlet and outlet respectively, mol/h
x_{in}, x_{out} = mole fraction of solute in liquid phase at inlet and outlet
y_{in}, y_{out} = mole fraction of solute in gas phase at inlet and outlet

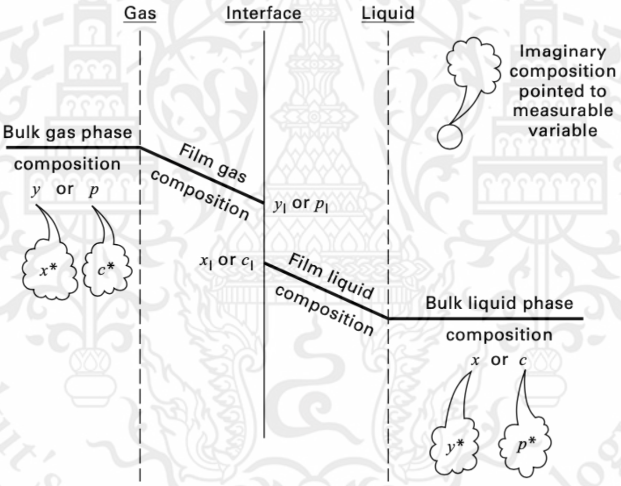


Figure A.5 Film theory for mass transfer from fluid–fluid interface into a liquid[1]

Gas-liquid equilibrium at interface

$$y_I = H\,x_I \qquad \text{A-3}$$

where H = Henry's constant

In terms of the mass transfer coefficient, mole-fraction driving force is existed for the common combination of vapor–liquid mass transfer, which expressed in mass transfer coefficients k_x and k_y for liquid and gas phases respectively.

$$r = k_x(x_I - x) = k_y(y - y_I) \qquad \text{A-4}$$

Due to the composition at the interface is difficult to measure hence overall volumetric mass-transfer coefficients are defined in terms of overall mass transfer coefficient.

$$r = K_x a\,(x^* - x) = K_y a\,(y - y^*) \qquad \text{A-5}$$

where r = mass-transfer rate per unit volume of packed bed, kmol/m^3·h

K_x, K_y = overall mass transfer coefficient for liquid phase and vapor phase respectively, kmol/m^2·h

y^* = fictitious vapor mole fraction in equilibrium with the mole fraction x, = Hx

x^* = fictitious liquid mole fraction in equilibrium with the mole fraction y, = y/H

a = mass transfer area per unit volume of packed bed, m^2/m^3

In packed column, cross section is S and the difference volume in height dz is Sdz. For dilute gas, the change in molar flow rate is neglected, and the amount absorbed in section dz is -Vdy which equals the absorption rate times the differential volume.

$$-V\,dy = K_y a\,(y - y^*)\,S dz \qquad \text{A-6}$$

Finally, the packed height can be calculated in term of integral form as

$$z_T = \frac{V}{K_y a S}\int_{y_{out}}^{y_{in}} \frac{dy}{y - y^*} \qquad \text{A-7}$$

$$z_T = \frac{V}{K_y a S}\,\frac{(y_{in} - y_{out})}{(y - y^*)_M} \qquad \text{A-8}$$

where

$$(y-y^*)_M = \frac{(y_{in}-y^*_{in}) - (y_{out}-y^*_{out})}{\ln[(y_{in}-y^*_{in})/(y_{out}-y^*_{out})]} \qquad \text{A-9}$$

$$y^*_{in} = H_A x_{out} \text{ and } y^*_{out} = H_A x_{in} \qquad \text{A-10}$$

from equation (8)

$$H_{OG} = \frac{V}{K_y aS} \quad \text{and} \quad N_{OG} = \int_{y_{out}}^{y_{in}} \frac{dy}{y-y^*} \qquad \text{A-11}$$

Besides, H_{OG} is represented the overall height of a gas transfer unit (HTU). The smaller the HTU, the more efficient the contacting. While N_{OG} is represented the overall number of gas transfer units (NTU), which means the larger the NTU, the greater the time or area of contact required.

3. Questions before the experiment

1. From Figure A.2, liquid velocity at 10 m/h is considered. What are the gas velocity and liquid holdup at loading and flooding point? Descripts relations between gas velocity and liquid holdup.
2. From Figure A.3, liquid velocity at 80 m/h is considered. What are the gas velocity and specific pressure drop at loading and flooding point? Descripts relations between gas velocity and pressure drop.
3. Data of packing is given below. Please calculate specific pressure drop at flooding and descripts relations between pressure drop and size of the packing.

Packing type	Packing factor, F_p (ft^2/ft^3)	Size, d_p (mm)
Plastic Pall rings	40	35
Plastic Pall rings	26	50

4. Why is Henry's law considered for calculating equilibrium of CO_2 concentration in gas and liquid phase?
5. From equation A-5, please write down regarding the mole flux (N_A) in unit of kmol/m^2·h in terms of overall mass transfer coefficient in liquid and gas phases.

4. Experimental

Chemicals

5. Sodium hydroxide (NaOH) solution at concentration 0.01 M
6. Phenolphthalein
7. RO Water
8. Pure CO_2

Glassware and apparatus

11. Absorption column
12. 3 of 50 ml Beaker
13. 14 of 50 ml Erlenmeyer flask
14. 14 of 50 ml sample bottles
15. 25 ml Pipette
16. Burette
17. Funnel
18. Pipe pump
19. Water tank

Part 1 To study of operating limitation

6. Check the water level in manometer as zero
7. Adjust gas flow rate to 300, 500, 700, 900, 1100, 1300 and 1500 lit/h then record pressure drop
8. Adjust liquid flow rate in range of 40-55 lit/h then change gas flow rate at 300-1,500 lit/h and record pressure drop
9. Increase gas flow rate to estimate flooding point then record pressure drop
10. Construct graph regarding liquid flow rate (x axis) and pressure drop (y axis) to consider the gas flow rate at loading and flooding point

Part 2 To study the changing of parameters which effects to overall mass transfer coefficient

Table A.3 Experiment data for each group

Group	Condition							
	1	2	3	4	5	6	7	8
1	✓		✓					
2		✓				✓		
3			✓	✓				
4						✓		✓
5	✓				✓			
6							✓	✓
7					✓		✓	
8		✓				✓		
9		✓		✓				
10	✓				✓			

Condition	Liquid flow rate (lit/h)	Gas flow rate (lit/h)	CO_2 concentration (%mol)
1	15	800	15
2	15	800	25
3	15	1,000	15
4	15	1,000	25
5	20	800	15
6	20	800	25
7	20	1,000	15
8	20	1,000	25

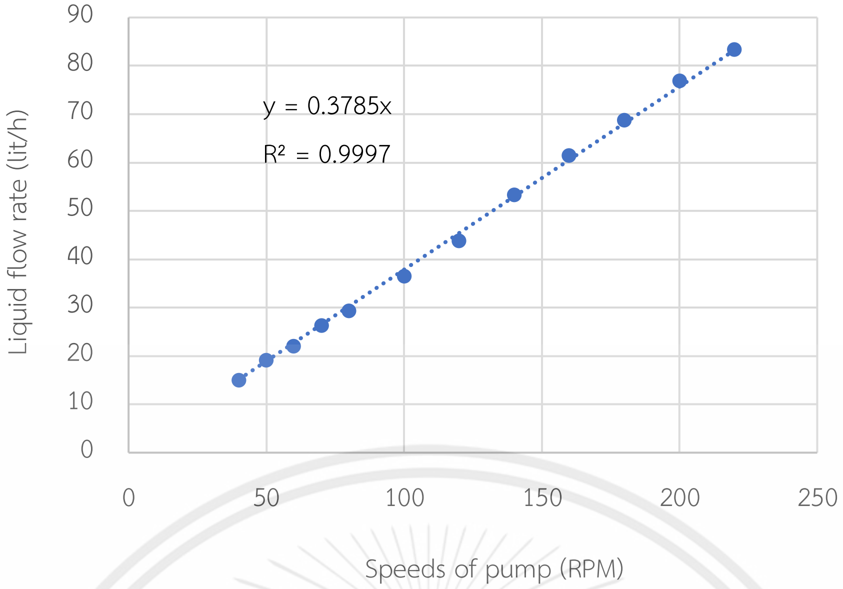


Figure A.6 Calibration curve for speeds of pump and liquid flow rate

Part 2.1 To estimate steady state time of CO_2 absorption

7. Check the water level in manometer as zero
8. Adjust flow rate of air and CO_2 to achieve the total gas flow rate and CO_2 concentration as set in Table A.3
9. Adjust speeds of pump to get the appointed liquid flow rate, which can be calculated from calibration curve in Figure A.6
10. Time when liquid begins contact with packing and collect samples at the outlet at 2, 4, 6, 8, 10, 15 and 20 minutes
11. Pipet samples of 25 cm^3 and add 1-2 drops of phenolphthalein then titrate with 0.01 M NaOH solution to estimate period of time before reach the steady state

Part 2.2 Acid-Base and its dissociation calculation for estimating CO_2 concentration in water

CO_2 in water

$$CO_2\ (g) + H_2O\ (l) \leftrightarrow H_2CO_3\ (aq)$$

$$H_2CO_3\ (aq) \leftrightarrow HCO_3^-\ (aq) + H^+\ (aq)$$

$$HCO_3^-\ (aq) \leftrightarrow CO_3^{2-}\ (aq) + H^+\ (aq)$$

H_2CO_3 with NaOH

$$2NaOH\ (aq) + H_2CO_3\ (aq) \rightarrow Na_2CO_3\ (aq) + 2H_2O\ (l)$$

Part 2.3 Calculation of overall mass transfer coefficient

12. Calculate overall mass transfer coefficient (K_y) from obtained CO_2 concentration in part 2
13. Student can explain impact of the changing of parameters which effects to overall mass transfer coefficient

The specific information for calculation

Suggest unit for equation (2): $V_{in} = V_{out}$ (mol/h), $L_{in} = L_{out}$ (lit/h), y (mol_{CO2}/mol), x (mol_{CO2}/lit)

Henry's constant of CO_2 (H_{CO2}) = 29.41 lit_{sol} atm/mol

Inside diameter of absorption column = 4.3 cm

Height of absorption column = 1 m

Area/volume of packing material (a) = 10.07 cm^2/cm^3

Packing factor (F_p) = 1589 ft^{-1}

5. References

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APPENDIX B
EXPERIMENTAL DATA

B.1 Based data for gas absorption column

Tojuvong and Wanaprasert[22] who have studied the CO_2 absorption and used this gas absorption column before provided an information for detail of the column and packing as shown in Table B.1.

Table B.1 Based data for gas absorption column

Inner diameter of absorption column	4.3	cm
Height of packed column	1	m
Type of packing	Glass Raschig ring	
Outer diameter of packing	4.0	mm
Inner diameter of packing	3.1	mm
Height of packing	4.1	mm
Total volume of column	1,379.6	cm^3
Approximate total number of packing	17,881	
Void fraction	0.5781	
Area/volume of packing	10.07	cm^3/cm^2
Packing factor	1,589	ft^{-1}

B.2 Experimental data

B.2.1 Data of measured pressure drop depends on volumetric liquid flow rate and volumetric gas flow rate.

Table B.2 Pressure drop (cm H_2O) varied by volumetric liquid flow rate and volumetric gas flow rate

Liquid flow rate (lit/h)	Gas flow rate (lit/h)												
	300	350	400	450	500	600	700	800	900	1,000	1,100	1,300	1,500
Dry column	0.9	▨	▨	▨	1.9	▨	3.2	▨	5.2	▨	7.2	9.6	12.2
40	1.4	▨	2.1	▨	3.6	5.0	6.7	8.8	11.3	15.0	26.5*	▨	▨
45	2.2	▨	3.5	▨	5.2	7.2	10.0	12.9	18.0	35.0*	▨	▨	▨
50	3.5	▨	4.7	▨	6.3	8.5	12.5	29.8*	▨	▨	▨	▨	▨
55	5.2	5.7	6.9	37.0*	▨	▨	▨	▨	▨	▨	▨	▨	▨

* means pressure drop at that point is approximate value due to flooding.

▨ means pressure drop at that condition was not measured.

B.2.2 Data of used sodium hydroxide solution (NaOH) 0.01 M titrated with carbonic acid (H_2CO_3) obtained from CO_2 absorbed by water at collected time.

Table B.3 Experimental plan

Condition	Liquid flow rate (lit/h)	Gas flow rate (lit/h)	CO_2 concentration (%mol)
1	15	800	15
2			25
3		1,000	15
4			25
5	20	800	15
6			25
7		1,000	15
8			25

Table B.4 Experimental data at condition 1

Time (minute)	Volume of NaOH (ml)		Outlet CO_2 concentration in water, x_{out} (M)	
	1st	2nd	1st	2nd
2	3.1	5.7	0.00062	0.00114
4	6.0	6.4	0.00120	0.00128
6	8.3	7.8	0.00166	0.00156
8	7.4	6.3	0.00148	0.00126
10	8.7	6.3	0.00174	0.00126
15	7.6	7.0	0.00152	0.00140
20	6.9	6.3	0.00138	0.00126

Table B.5 Experimental data at condition 2

Time (minute)	Volume of NaOH (ml)		Outlet CO_2 concentration in water, x_{out} (M)	
	1st	2nd	1st	2nd
2	3.1	5.7	0.00062	0.00114
4	6.0	6.4	0.00120	0.00128
6	8.3	7.8	0.00166	0.00156
8	7.4	6.3	0.00148	0.00126
10	8.7	6.3	0.00174	0.00126
15	7.6	7.0	0.00152	0.00140
20	6.9	6.3	0.00138	0.00126

Table B.6 Experimental data at condition 3

Time (minute)	Volume of NaOH (ml)		Outlet CO_2 concentration in water, x_{out} (M)	
	1st	2nd	1st	2nd
2	6	3.5	0.0012	0.00070
4	6.4	3.5	0.00128	0.00070
6	5.5	3.8	0.00110	0.00076
8	5.9	3.9	0.00118	0.00078
10	5.8	4.0	0.00116	0.00080
15	5.6	4.1	0.00112	0.00082
20	5.7	4.1	0.00114	0.00082

Table B.7 Experimental data at condition 4

Time (minute)	Volume of NaOH (ml)		Outlet CO_2 concentration in water, x_{out} (M)	
	1st	2nd	1st	2nd
2	8.7	7.4	0.00174	0.00148
4	7.1	8.2	0.00142	0.00164
6	6.0	8.4	0.00120	0.00168
8	5.2	7.3	0.00104	0.00146
10	6.4	7.3	0.00128	0.00146
15	6.9	7.4	0.00138	0.00148
20	8.2	7.8	0.00164	0.00156

Table B.8 Experimental data at condition 5

Time (minute)	Volume of NaOH (ml)		Outlet CO_2 concentration in water, x_{out} (M)	
	1st	2nd	1st	2nd
2	5.8	5.0	0.00116	0.0010
4	6.0	5.8	0.0012	0.00116
6	6.1	6.1	0.00122	0.00122
8	6.5	6.6	0.00130	0.00132
10	6.1	6.3	0.00122	0.00126
15	6.0	6.4	0.0012	0.00128
20	6.2	6.4	0.00124	0.00128

Table B.9 Experimental data at condition 6

Time (minute)	Volume of NaOH (ml)		Outlet CO_2 concentration in water, x_{out} (M)	
	1st	2nd	1st	2nd
2	8.1	7.1	0.00162	0.00142
4	6.6	8.4	0.00132	0.00168
6	6.5	8.2	0.00130	0.00164
8	6.8	8.1	0.00136	0.00162
10	8.8	8.1	0.00176	0.00162
15	8.0	8.5	0.00160	0.00170
20	8.1	7.1	0.00162	0.00142

Table B.10 Experimental data at condition 7

Time (minute)	Volume of NaOH (ml)		Outlet CO_2 concentration in water, x_{out} (M)	
	1st	2nd	1st	2nd
2	5.1	3.8	0.00102	0.00076
4	4.8	3.4	0.00096	0.00068
6	3.8	3.8	0.00076	0.00076
8	5.4	3.7	0.00108	0.00074
10	4.4	4.8	0.00088	0.00096
15	4.3	3.5	0.00086	0.00070b
20	4.7	4.2	0.00094	0.00084

Table B.11 Experimental data at condition 8

Time (minute)	Volume of NaOH (ml)		Outlet CO_2 concentration in water, x_{out} (M)	
	1st	2nd	1st	2nd
2	5.2	11.4	0.00104	0.00228
4	6.6	11.7	0.00132	0.00234
6	7.7	11.9	0.00154	0.00238
8	8.6	10.4	0.00172	0.00208
10	9.2	9.6	0.00184	0.00192
15	9.3	9.1	0.00186	0.00182
20	9.2	9.1	0.00184	0.00182

B.2.2 Data of used sodium hydroxide solution (NaOH) titrated with carbonic acid (H_2CO_3) obtained from CO_2 absorbed by water in absorption column before improvement.

Table B.12 Experimental data obtained from previous absorption column for comparison.

Condition	Time (minute)	Volume of NaOH (ml)	Outlet CO_2 concentration in water, x_{out} (M)
Gas flow rate = 800 lit/h Liquid flow rate = 15 lit/h Inlet CO_2 concentration = 25%mol NaOH concentration = 0.001 M	2	1.2	0.000024
	4	0.8	0.000016
	6	0.9	0.000018
	8	0.8	0.000016
	10	1.2	0.000024
	15	1.0	0.000020
	20	0.9	0.000018

Table B.12 (Cont.) Experimental data obtained from previous absorption column for comparison.

Condition	Time (minute)	Volume of NaOH (ml)	Outlet CO_2 concentration in water, x_{out} (M)
Gas flow rate = 800 lit/h Liquid flow rate = 20 lit/h Inlet CO_2 concentration = 25%mol NaOH concentration = 0.001 M	2	0.9	0.000018
	4	0.9	0.000018
	6	0.5	0.000010
	8	0.7	0.000014
	10	0.8	0.000016
	15	0.5	0.000010
	20	1.0	0.000020
Gas flow rate = 1,000 lit/h Liquid flow rate = 20 lit/h Inlet CO_2 concentration = 15%mol NaOH concentration = 0.0005 M	2	1.8	0.000018
	4	1.7	0.000017
	6	1.6	0.000016
	8	1.8	0.000018
	10	1.5	0.000015
	15	1.5	0.000015
	20	1.5	0.000015

APPENDIX C

CALCULATION

Calculation of the overall mass transfer coefficient (K_y)

Assumption

1. The absorption process of CO_2 in gas absorbed by water as a solvent is non-reaction process.
2. Liquid flow rate and gas flow rate were assumed to be constant throughout the experiments.
3. CO_2 did not exist in the inlet of liquid (x_{in} = 0 M).
4. The whole experiments were done under the same temperature.

Calculation condition

Volumetric gas flow rate	800	lit/h
Volumetric liquid flow rate	15	lit/h
Inlet CO_2 concentration	25	%v/v

Acid-base calculation

At equivalence point, concentration of liquid can be calculated because it is the point which equivalent amounts of the reactants and products have reacted.

$$\text{moles acid} = \text{moles base} \qquad \text{C.1}$$

Then, moles were transformed to concentration (C) in molarity of a solution multiplied by the volume (V). Since, a and b represented number of H^+ and OH^- in acid and base respectively.

$$a\, C_A V_A = b\, C_B V_B \qquad \text{C.2}$$

$$C_A = \frac{b\, C_B V_B}{a\, V_A} \qquad \text{C.3}$$

Outlet concentration in liquid phase (x_{out}) in unit of molarity became

$$x_{out} = \frac{(0.01\ \text{M})(6.9\ \text{ml})}{(2)(25\ \text{ml})} = 0.00138\ \text{M}$$

To find outlet concentration in gas phase (y_{out}), mole balance equation can be used.

$$L_{in}x_{in} + V_{in}y_{in} = L_{out}x_{out} + V_{out}y_{out} \qquad \text{C.4}$$

Inlet concentration in liquid phase (x_{in}) was canceled due to pure solvent was used. Hence, y_{out} became

$$y_{out} = \frac{L_{out}x_{out} - V_{in}y_{in}}{V_{out}} \qquad \text{C.5}$$

V_{in} and V_{out} represented total mole flow rate of gas. Volumetric flow rate was converted to moles flow rate by properties of gas at 30°C shown in Table C.1

Table C.1 Properties of gas at 30°C

Density of air	1.1649	g/l
Density of CO2	1.7475	g/l
Molecular weight of air	29	g/mol
Molecular weight of CO_2	44	g/mol

Therefore,

$$V_{in} = V_{out} = \text{Total volumetric flow rate} \times \left[\frac{(1\text{-}y_{in})D_{air}}{Mw_{air}} + \frac{y_{in}D_{CO_2}}{Mw_{CO_2}}\right] \qquad \text{C.6}$$

$$V_{in} = V_{out} = 800\ \text{lit/h} \times \left[\frac{(1\text{-}0.25)(1.1649\ \text{g/l})}{29\ \text{g/mol}} + \frac{(0.25)(1.7475\ \text{g/l})}{44\ \text{g/mol}}\right] = 32.04\ \text{moles/h}$$

Moles flow rate of gas was substituted into moles balance equation. Mole fraction at the outlet in gas phase was

$$y_{out} = \frac{(15\ \text{lit/h})(0.00138\ \text{M}) - (32.04\ \text{moles/h})(0.25)}{(32.04\ \text{moles/h})} = 0.24935$$

The packed height is a multiple of overall height of a gas transfer unit (H_{OG}) and overall number of gas transfer units (N_{OG}).

$$z_T = H_{OG}N_{OG} \quad \text{C-7}$$

whereas

$$H_{OG} = \frac{V}{K_y aS} \quad \text{and} \quad N_{OG} = \int_{y_{out}}^{y_{in}} \frac{dy}{y-y^*} \quad \text{C-8}$$

Then,

$$z_T = \frac{V}{K_y aS}\,\frac{(y_{in}-y_{out})}{(y-y^*)_M} \quad \text{C-9}$$

where

$$(y-y^*)_M = \frac{(y_{in}-y^*_{in}) - (y_{out}-y^*_{out})}{\ln[(y_{in}-y^*_{in})/(y_{out}-y^*_{out})]} \quad \text{C-10}$$

$$y^*_{in} = H_A x_{out} \text{ and } y^*_{out} = H_A x_{in} \quad \text{C-11}$$

A common way to define Henry volatility is by dividing the partial pressure by the aqueous-phase concentration:

$$K_H^{pc} \overset{\text{def}}{=} \frac{p}{C_a} = \frac{1}{H^{cp}} \quad \text{C-12}$$

Here, C_a is the concentration of a species in the aqueous phase, and p is the partial pressure of that species in the gas phase under equilibrium conditions. However, Henry solubility defined via concentration (H_{cp}) normally based on $H^{\ominus}$ at the reference temperature $T^{\ominus}$ = 298.15 K. The temperature dependence of equilibrium constants can generally be described with the van’t Hoff equation. It also applies to Henry’s law constants:

$$H(T) = H^{\ominus} \times \exp\left(\frac{d \ln H}{d(1/T)}\left(\frac{1}{T} - \frac{1}{T^{\ominus}}\right)\right) \quad \text{C-13}$$

Table C.2 Henry's law constants for water as solvent[9]

Substance Formula [CAS registry number]	H^{cp} (at $T^{\ominus}$) $\left[\frac{\text{mol}}{\text{m}^3\text{Pa}}\right]$	$\frac{d \ln H}{d\,(1/T)}$ [K]	Reference
Carbon dioxide CO_2 [124-38-9]	3.4×10^{-4}	2,400	Edwards et al. (1978)
	3.3×10^{-4}	2,400	Sander et al. (2011)

At operated temperature = 303.15 K (30°C)

$$\text{H(T=303.15 K)} = 3.3\times10^{-4} \times \exp\left(2400\left(\frac{1}{303.15} - \frac{1}{298.15}\right)\right) \frac{\text{mol}}{\text{m}^3\text{Pa}}$$

$$\text{H(T=303.15 K)} = 2.89\times10^{-4}\,\frac{\text{mol}}{\text{m}^3\text{Pa}}$$

$$K_H^{pc} = 3460.21\ \frac{\text{m}^3\text{Pa}}{\text{mol}} = 34.15\ \frac{\text{lit}_{\text{sol}}\ \text{atm}}{\text{mol}}$$

Hence, mole fraction at the outlet and outlet in equilibrium with the mole fraction x for gas phase was

$$y^*_{in} = 0.0406 \text{ and } y^*_{out} = 0$$

Mole fraction were substituted into equation C-10

$$(y\text{-}y^*)_M = \frac{(0.25000\text{-}0.04060)\ \text{-}\ (0.24935\text{-}0)}{\ln\left[(0.25000\text{-}0.04060)/(0.24935\text{-}0)\right]} = 0.22880$$

The overall mass transfer coefficient obtained from equation C-9.

$$K_y = \frac{32.04 \text{ moles/h}}{(1\text{ m})(1007\text{ m}^2/\text{m}^3)(0.00145\text{ m}^2)} \times \frac{(0.2500\text{-}0.24935)}{0.2288}$$

$$K_y = 0.0619 \text{ moles/m}^2\cdot\text{h}$$

Overall height of a gas transfer unit (H_{OG}) and overall number of gas transfer units (N_{OG}) were summarized as

$$H_{OG} = \frac{32.04 \text{ moles/h}}{(0.0619 \text{ moles/m}^2\cdot\text{h})(1007 \text{ m}^2/\text{m}^3)(0.00145 \text{ m}^2)} = 354 \text{ m}$$

$$N_{OG} = \frac{(0.2500\text{-}0.24935)}{0.2288} = 2.84\times10^{-3}$$

In fact that packed height is 1 m.

$$z_T = (354 \text{ m})(2.84\times10^{-3})$$

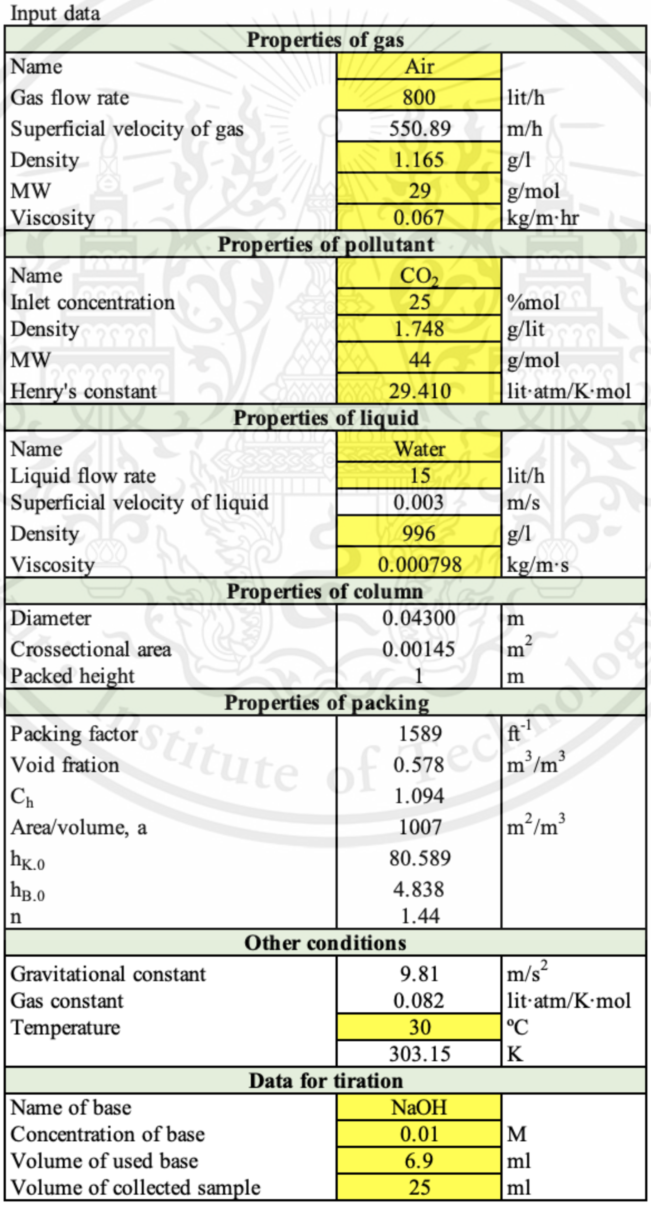
$$z_T = 1 \text{ m}$$

APPENDIX D
EXCEL TEMPLATE FOR GAS ABSORPTION LABORATORY

Excel template is created for using of gas absorption unit within the laboratory of Chemical Engineering Department.

D.1 Input data

Properties of gas and liquid are required and filled into the yellow blank as shown in Figure D.1.



Input data

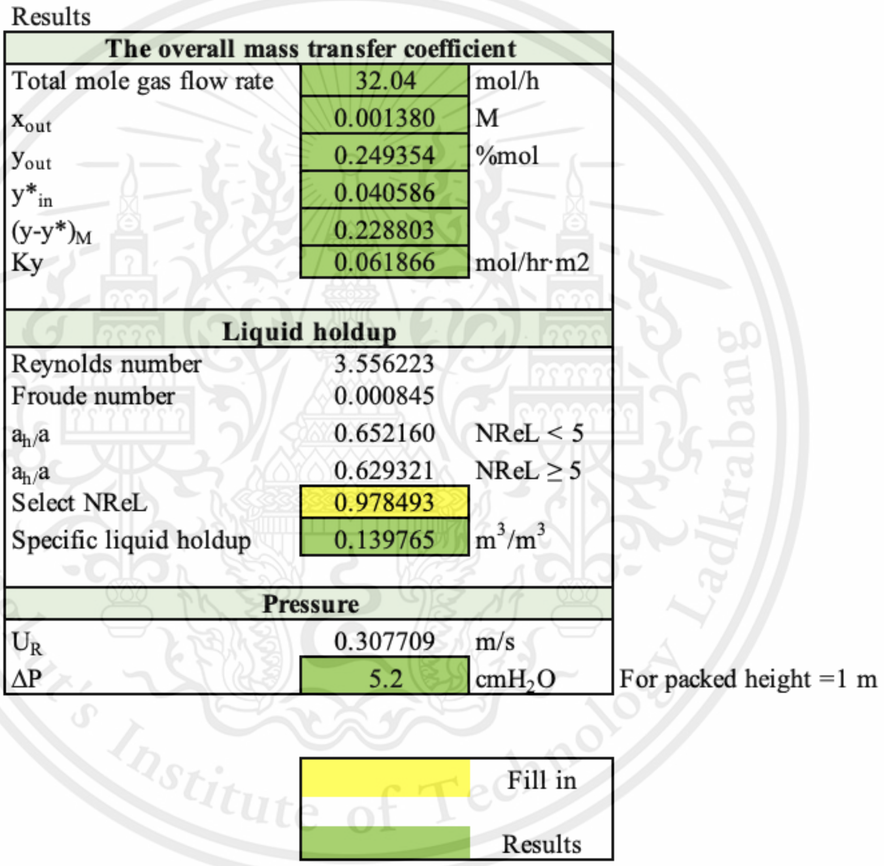
Properties of gas		
Name	Air	
Gas flow rate	800	lit/h
Superficial velocity of gas	550.89	m/h
Density	1.165	g/l
MW	29	g/mol
Viscosity	0.067	kg/m·hr
Properties of pollutant		
Name	CO_2	
Inlet concentration	25	%mol
Density	1.748	g/lit
MW	44	g/mol
Henry's constant	29.410	lit·atm/K·mol
Properties of liquid		
Name	Water	
Liquid flow rate	15	lit/h
Superficial velocity of liquid	0.003	m/s
Density	996	g/l
Viscosity	0.000798	kg/m·s
Properties of column		
Diameter	0.04300	m
Crossectional area	0.00145	m^2
Packed height	1	m
Properties of packing		
Packing factor	1589	ft^{-1}
Void fration	0.578	m^3/m^3
C_h	1.094	
Area/volume, a	1007	m^2/m^3
$h_{K.0}$	80.589	
$h_{B.0}$	4.838	
n	1.44	
Other conditions		
Gravitational constant	9.81	m/s^2
Gas constant	0.082	lit·atm/K·mol
Temperature	30	°C
	303.15	K
Data for tiration		
Name of base	NaOH	
Concentration of base	0.01	M
Volume of used base	6.9	ml
Volume of collected sample	25	ml

Figure D.1 Input data for excel template

Including titration data, the quantity of collected sample, used base and concentration are filled. The specific data such as properties of packing and absorption column are provided to match the data of those absorption laboratory.

D.2 Output data

The output data shown in Figure D.2 are including the overall mass transfer coefficient, specific liquid holdup and pressure drop. Note that, this template is internal used for gas absorption unit within the laboratory of Chemical Engineering Department.



Results

The overall mass transfer coefficient		
Total mole gas flow rate	32.04	mol/h
x_{out}	0.001380	M
y_{out}	0.249354	%mol
y^*_{in}	0.040586	
$(y-y^*)_M$	0.228803	
Ky	0.061866	mol/hr·m2

Liquid holdup		
Reynolds number	3.556223	
Froude number	0.000845	
a_h/a	0.652160	NReL < 5
a_h/a	0.629321	NReL ≥ 5
Select NReL	0.978493	
Specific liquid holdup	0.139765	m^3/m^3

Pressure		
U_R	0.307709	m/s
ΔP	5.2	cmH_2O

For packed height =1 m

🟨	Fill in
🟩	Results

Figure D.2 Output data obtained from excel template

BIBLIOGRAHPY

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