Report of Cooperative Education

Stripper Design for Wastewater Treatment Unit of a Purified Terephthalic Acid Plant

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ABSTRACT

Wastewater treatment unit of a purified terephthalic acid (PTA) plant consists of anaerobic and aerobic systems. The pH of wastewater must be adjusted to almost neutral (pH =7) by using 50% w/w NaOH solution for high efficiency of anaerobic bacteria in the anaerobic system. The cost of NaOH solution per year is rather high. Basically, methane and carbon dioxide gases are generated by methanogenesis reaction in the anaerobic system. CO_2 dissolved in water subsequently forms carbonic acid resulting in lower pH of the circulating stream to the anaerobic system. The purpose of this project is, therefore, to design a stripping unit for the removal of dissolved CO_2. The circulating stream of wastewater with higher pH value can potentially reduce the amount of NaOH solution. Properly designed stripping unit is a preferred method to remove dissolved CO_2. The stripping unit consists of a packed tower, a centrifugal pump, and a centrifugal blower. The packed tower is filled 50-mm plastic Pall rings with 3.95 m of packing height. The packed tower operates at the volumetric flow rate of the gas about 50 times that of the volumetric flow rate of circulating stream. The packed tower can remove dissolved CO_2 from 168.37 mg/L to 0.58 mg/L, resulting in the pH will increase from 6.99 to 9.45. The packed tower diameter is equal to 5.65 m for stripped wastewater flow rates at 955 m^3/h. The project gives 46.28% IRR, which a worthy investment. Because the minimum acceptable rate of return (MARR) is 20%as follows the company criteria.

Keywords: Anaerobic, Dissolved carbon dioxide, Carbonic acid, Stripping, Packed tower

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CHAPTER I

INTRODUCTION

1.1 Background

Siam Mitsui PTA Company Limited (SMPC) manufactures and sells purified terephthalic acid (PTA) for use in fiber, polyethylene terephthalate (PET), and film applications. SMPC operates as a joint venture between Mitsui Chemicals, Inc. and SCG Chemicals Co., Ltd. The PTA plant consists of two major units: crude terephthalic acid (CTA) unit and PTA unit. CTA is produced by the catalytic liquid phase air oxidation of para-xylene in the CTA unit. The produced CTA is converted into PTA by the PTA unit (Purification). A large amount of wastewater is produced by the production units which contain high chemical oxygen demand (COD) and low pH.

Wastewater treatment unit in a PTA plant consists of two systems: anaerobic system and aerobic system. The anaerobic system used for removal of COD in wastewater before the aerobic system. The most important process control parameter in the anaerobic system is pH because of the optimum pH range for good performance and stability. The cost of NaOH solution per year is rather high.as shown in Fig. 1.1.

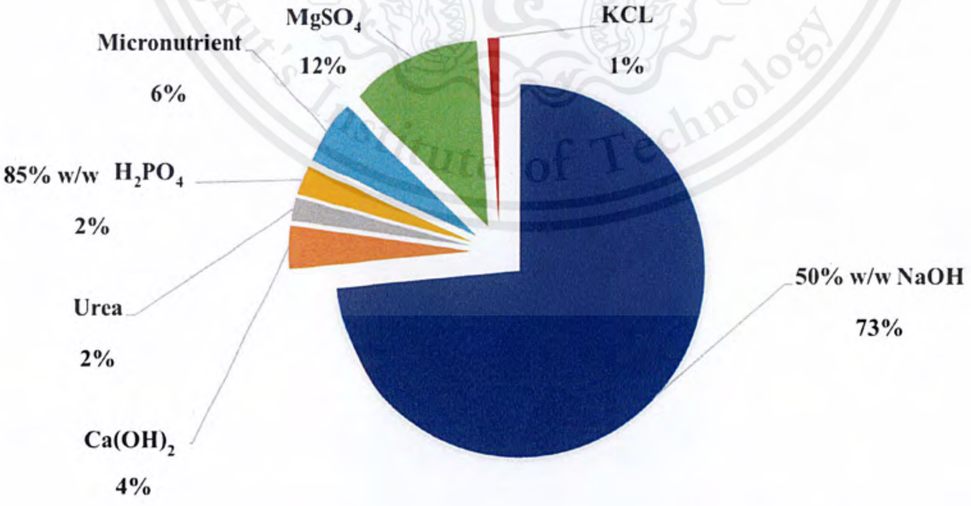


Figure 1.1 Operating Cost of the Chemicals Usage in Wastewater Treatment Unit

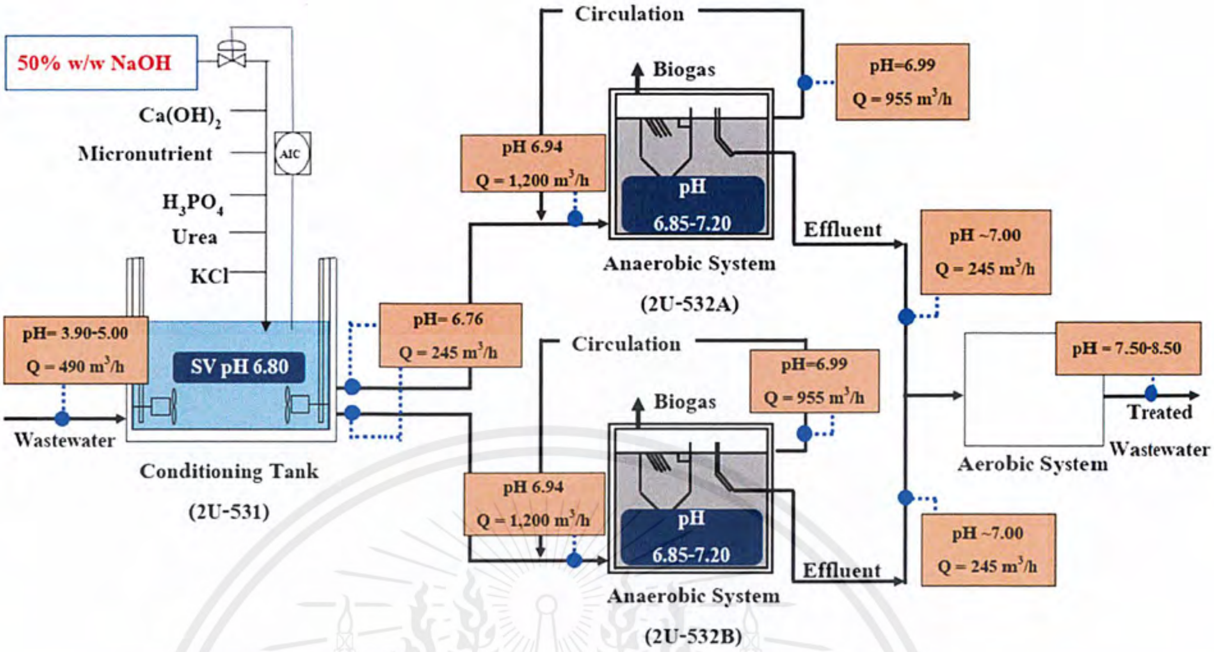


Figure 1.2 Wastewater Treatment Unit of a PTA Plant

Fig. 1.2 shows that the wastewater stream after adjusted (pH 6.76) will be mixed with a circulating stream from anaerobic reactor (pH 6.99), the pH of the adjusted stream will increase from 6.76 to 6.94. So, if the pH of the circulating stream from anaerobic reactor increased, the pH of the adjusted stream will increase. To maintain the pH of the anaerobic reactor inlet stream, the set point of the conditioning tank must be decreased.

The removal of COD is accomplished by the conversion of organics into methane (CH_4) and carbon dioxide (CO_2) in the anaerobic system, which CO_2 is a relatively soluble gas. CO_2 dissolves in water undergoes chemical reactions, forming a carbonic acid (H_2CO_3). The pH of the wastewater can be increased by CO_2 removal.

1.2 Objective

To design stripper for the reduction of NaOH solution usage in wastewater treatment unit

1.3 Scopes of Work

1.3.1 Study of NaOH solution usage and wastewater treatment unit in a PTA plant

1.3.2 Study of theory and literature review

1.3.3 Study of dissolved carbon dioxide in water

1.3.4 Design of stripper unit and equipment

1.3.5 Evaluate of cost saving from reduction of NaOH solution usage and investment

1.4 Expected Outputs

1.4.1 Reduction of sodium hydroxide cost in wastewater treatment unit

1.4.2 Possibility of installing stripper unit

CHAPTER II

LITERATURE REVIEW

2.1 Anaerobic System [Metcalf & Eddy, 2014]

Anaerobic system is biological stabilization process operated in the absence of free dissolved oxygen and oxidized compounds in which the biodegradable matter in primary and secondary sludge is converted to methane, carbon dioxide, and other end products.

2.1.1 Steps of Organic Decompositions

Three basic steps are involved in the overall anaerobic oxidation of a waste: (1) hydrolysis, (2) acidogenesis (also known as fermentation or anaerobic oxidation), and (3) methanogenesis. The three steps are shown in Fig. 2.1.

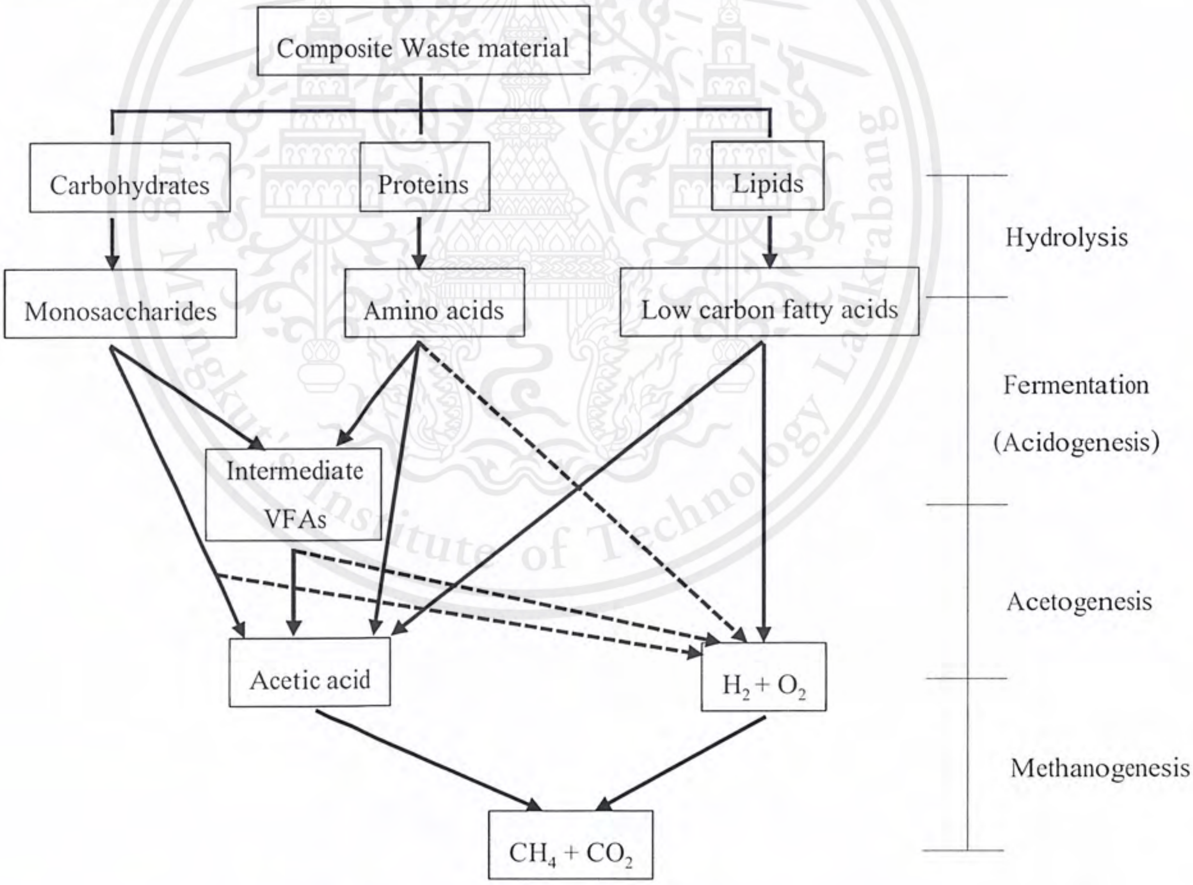


Figure 2.1 Fate of Biodegradable COD in Anaerobic Processing of Waste Solids

(Adapted from Metcalf & Eddy, 2014)

1) Hydrolysis

The first basic step, in which particulate material is converted to soluble compounds that can then be hydrolyzed further to simple monomers (monosaccharides, amino acids, and low carbon fatty acids) that are used by bacteria that perform fermentation, is termed hydrolysis. Hydrolysis is carried out with extracellular enzymes produced by variety of facultative and obligate anaerobes.

2) Acidogenesis

The second basic step, which is done by bacteria, is acidogenesis (also known as fermentation) and results in the production of volatile fatty acids (VFAs) such as propionic, butyric, valeric acid, and etc., CO_2, and H_2 as shown in Fig. 2.1. Example, conversion of glucose to propionic acid and ethanol.

$$C_6H_{12}O_6 + 2H_2 \longrightarrow 2CH_3CH_2COOH + 2H_2O$$

$$C_6H_{12}O_6 \longrightarrow 2CH_3CH_2OH + 2CO_2$$

3) Acetogenesis

Acetogenesis refers to furthers fermentation by bacteria to convert intermediate product of acidogenesis, occurs for some of VFAs produced from acidogenesis to also produce acetate, CO_2, and H_2. Thus, the final products of fermentation are acetate, CO_2, and H_2, which are precursors of methane formation. Example, propionate converted to acetate and glucose converted to acetic acid or acetate.

$$CH_3CH_2COO^- + 3H_2O \longrightarrow H_3COO^- + H^+ + HCO_3^- + 3H_2$$

$$C_6H_{12}O_6 + 2H_2O \longrightarrow 2CH_3COOH + 2CO_2 + 4H_2$$

4) Methanogenesis

The third basic step is methanogenesis, which carried out by a group of Archea organisms known as methanogens. Two groups of methanogenic organisms are involved in methanogenesis. One group, acetoclastic methanogens or acetate utilizing methanogen, convert acetate to methane and CO_2.

$$CH_3COOH \longrightarrow CH_4 + CO_2$$

The second group, hydrogenotrophic methanogens or H_2 utilizing methanogens, use hydrogen as the electron donor and CO_2 as the electron acceptor to produce methane.

$$CO_2 + 4H_2 \longrightarrow CH_4 + H_2O$$

2.1.2 Environmental Factors [สันทัด, 2549]

1) Temperature

A temperature that most affect bacteria living in wastewater. Temperature has great influences on reaction rates and microbial growth rates. There are three ranges for process operation to produce methane as follows:

Phychrophilic range	5-15 °C
Mesophilic range	30-40 °C
Thermophilic range	50-60 °C

Operation in the mesophilic range is generally used in a tropical region such Thailand.

2) pH

The most important process control parameter is pH. The optimum pH range for all methanogenic bacteria is between 6 and 8 but the optimum pH for the group as a close to 7. The pH of anaerobic systems for good performance and stability is in the range of 6.5-7.5. pH will affect to microbial activity as follows:

- *Hydrolytic Bacteria,* it will good performance in neutral wastewater, which ability to resist lower pH than 4.5.
- *Acidogenic Bacteria,* it is similar to the Hydrolytic bacteria. This type of bacteria such as Acidogenic bacteria will good performance in acidic wastewater (pH 4.0-6.0).

- *Methanogenic Bacteria,* it will good performance in closely neutral wastewater (pH 6.5-7.8). pH higher or lower of this range that can affect activity efficiency.

3) Alkalinity

Alkalinity is the bacterium's ability to resist changes in pH caused by the addition of acids, that is, wastewater fed into the anaerobic reactor, the organics are converted to organic acids and the pH will decrease. The lower pH effects of acid forming bacteria or methanogenic bacteria. The wastewater alkalinity in the reactor should be within 1,000-3,000 milligrams per litters range.

4) Volatile Fatty Acids

VFAs is the precursor for methane production, which produced from acid forming bacteria activity. So, the growth rate of acid forming bacteria related to methanogenic bacteria. The high accumulate of VFAs in wastewater that affects alkalinity and pH increasing. VFAs in wastewater should be lower than 2,000 milligrams per litters.

5) Nutrients

Though anaerobic processes produce less sludge and thus require less nitrogen and phosphorus for biomass growth, many industrial wastewaters may lack sufficient nutrients. Thus, the addition of nitrogen and/or phosphorus may be needed.

6) Toxic Substances

Proper waste analysis and treat ability studies are needed to assure that a chronic or serious transient toxicity does not exist for wastewater treated by anaerobic processes. At the same time, the presence of a toxic substance does not mean the process cannot function. Some toxic compounds inhibit anaerobic methanogenic reaction rates, but with a high biomass inventory and low enough loading, the process can be sustained. Toxic and inhibitory inorganic and organic compounds of concern anaerobic processes are presented in Table 2-1. Acclimation to toxic concentrations has also been shown but it may be necessary to apply pretreatment steps to prevent toxicity problems in the anaerobic degradation process

Table 2.1 Toxic and Inhibitory Inorganic and Organic Compounds and Concentrations Harmful to Methanogens in Anaerobic Processes [สันทัด, 2549]

Toxic substance	**Acceptable maximum concentration (mg/L)**	**Toxic substance**	**Acceptable maximum concentration (mg/L)**
Cu	1.0	Na^+	3,500
Zn	5.0	K^+	2,500
Cr^{6+}	5.0	Ca^{2+}	2,500
Chloride	15,000	Mg^{2+}	1,000
Cr^{3+}	2,000	Averylonitrite	5.0
Total chromium	5.0	Benzene	50
Ni	2.0	CCl_4	10
Cd	0.02	Chloroform	0.1
S^{2-}	100	Pentachlorophenol	0.4
SO_4^{2-}	500	Cyanide	1.0
Ammonia	1,500		

2.1.3 Types of Anaerobic Reactors [Metcalf & Eddy, 2014]

1) Low Loaded Anaerobic Lagoon System (ANL)

Generally, unmixed reactor system employing suspended/ flocculating anaerobic biomass and settled anaerobic solids with a hydraulic retention time (HRT) of 20 to 25 days and average solid retention times (SRT) of 50 to 100 days. Can handle a wider range of wastes including solids and soluble wastewaters. Designed for a total COD loading of less than 2 kg/m^3-d. The ANL schematics are shown in Fig. 2.2 .

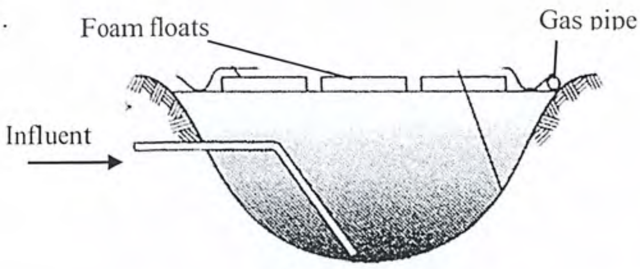


Figure 2.2 Schematics of Low Loaded Anaerobic Lagoon System [Metcalf & Eddy, 2014]

2) Up-flow Anaerobic Sludge Blanket (UASB)

An up-flow reactor with a bottom sludge bed and dense, granular anaerobic biomass with good mixing provided by the up-flow velocity and biogas generation. The reactor effective anaerobic sludge concentration may be in the range of 35 to 40 kg/m^2. A gas liquid-solid separator at the top separates granular solids from the effluent and collects biogas. The sludge blanket has SRTs in excess of 30 d with HRTs in the range of 4 to 8 h. Designed for a COD loading of 5 to 20 kg/m3 d. Up flow velocities can vary from 1 to 6 m/h and reactor height of 5 to 20 m have been used. The UASB schematics are shown in Fig. 2.3

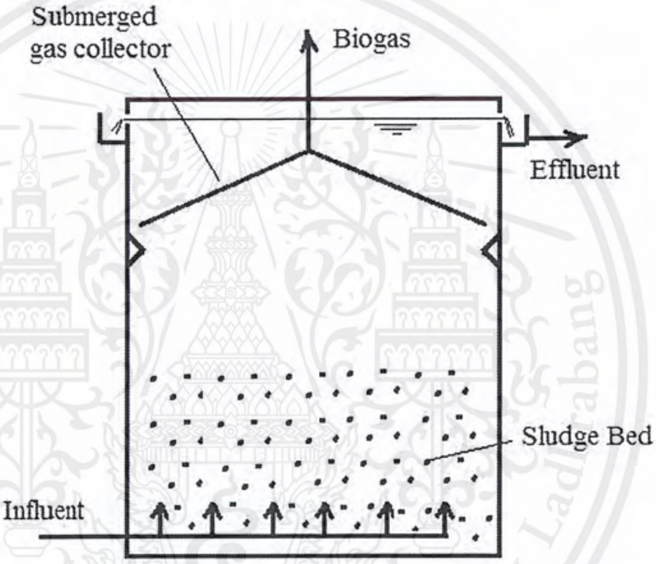


Figure 2.3 Schematics of Up-flow Anaerobic Sludge Blanket [Metcalf & Eddy, 2014]

3) Expanded Granular Sludge Blanket (EGSB)

The EGSB is a commonly used modification of the UASB by employing a higher up-flow velocity, a greater height so diameter ratio, and recirculation of effluent. Up-flow velocities may be the range of 4 to 10 m/h and reactor heights up to 25 m have been used. The higher velocity provides a more efficient reactor for treatment of soluble substrates by improving mixing, reducing dead volume, and increasing diffusion rates from the bulk liquid to the granular biofilm. Was originally developed to treat low strength wastes but has been used for high strength as well and at low temperatures of 10°C. Organic loading rates as high as 35 kg/m have been used. The EGSB schematics are shown in Fig. 2.4

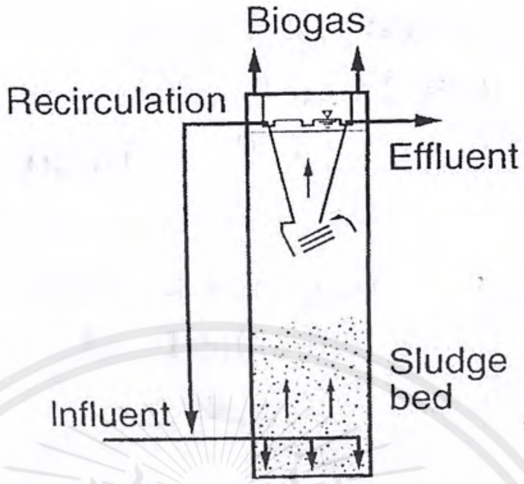


Figure 2.4 Schematics of Expanded Granular Sludge Blanket [Metcalf & Eddy, 2014]

4) Fluidized Bed (FB)

These systems are based on the development of a dense anaerobic biomass on small size (0.10 to o 30 mm) inert particles of fine sand, basalt, pumice, or plastic. The particles are kept in suspension and mixed by a high upward velocity. The higher velocities lead to what is called a fluidized bed with 25 to 300 percent bed expansion and the expanded fluidized bed refers to operation at lower velocity with 15 to 25 percent bed expansion. These reactors are applicable for soluble wastes or easily degraded small particulates, such as whey. Up-flow velocities may in the range 20 to 40 kg/m^3 were used. The FB schematics are shown in Fig. 2.5

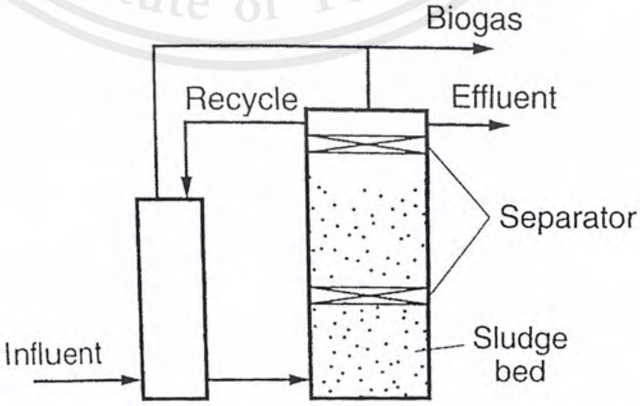


Figure 2.5 Schematics of Fluidized Bed [Metcalf & Eddy, 2014]

5) Anaerobic Contact Process (ACP)

A completely mixed reactor system employing suspended anaerobic biomass, a mixing/ flocculator degassing chamber, liquid-solids separation, and solids recycle so SRT is longer than the HRT. Designed for a COD loading of in the range of 2 to 5 kg/m^3-d. The ACP schematics are shown in Fig. 2.6 .

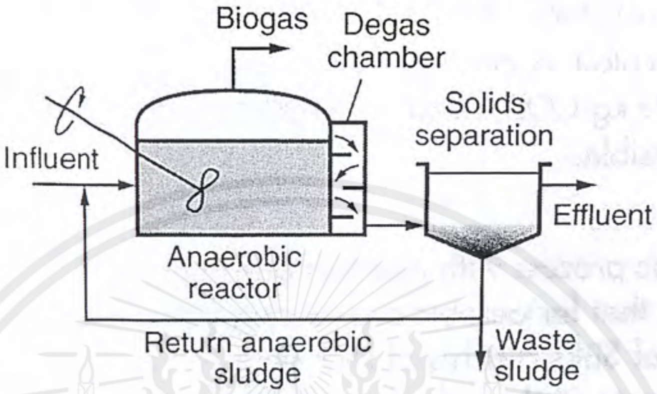


Figure 2.6 Schematics of Anaerobic Contact Process [Metcalf & Eddy, 2014]

6) Anaerobic Filter (AF)

An anaerobic filer (ANFI) system is an unmixed reactor system employing fixed film anaerobic biomass attached to supporting media, so that a large anaerobic biomass and long SRT can be maintained to allow treatment at HRTs in the range of 1 to 3 day and designed for a COD loading of 5 to 20 kg/m^3 -d. It is available in up-flow (ANFU) and down-flow (ANFD) configurations. The AF schematics are shown in Fig. 2.7 .

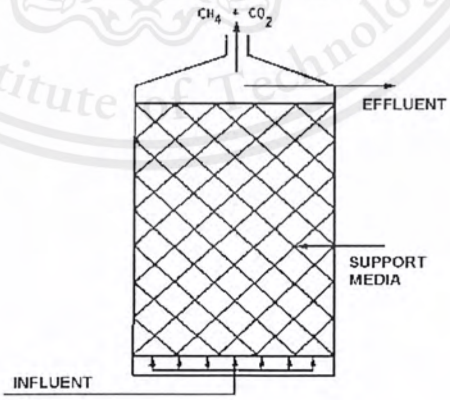


Figure 2.7 Schematics of Anaerobic Filter [Metcalf & Eddy, 2014]

7) Anaerobic Baffled Reactor (ABR)

Baffles are used to direct the flow of wastewater in an up-flow mode through a series of up-flow anaerobic sludge blanket reactors. The sludge in the reactor rises and falls with gas production and flow, but moves through the reactor at a slow rate. Reactor volatile solids concentrations vary from 2 to 10 percent. Systems have been operated with HRTs in the range of 6 to 24 h and SRTs in excess of 30 days. Designed for a COD loading of 5 to 10 kg/m^3-d. The main limitations of the ABR process are that many studies have been limited to laboratory and pilot scale treatment units. The ABR schematics are shown in Fig. 2.8.

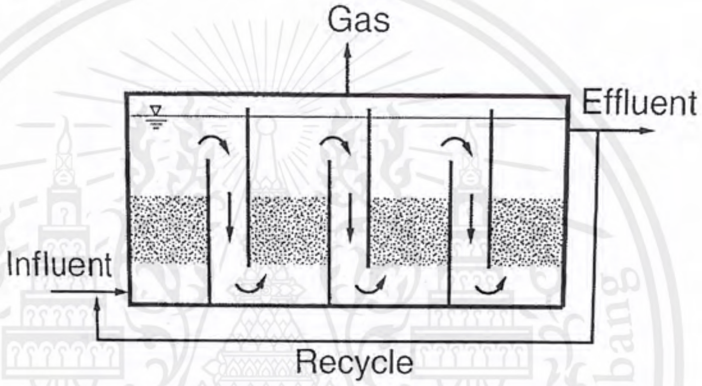


Figure 2.8 Schematics of Anaerobic Baffled Reactor [Metcalf & Eddy, 2014]

8) Anaerobic Sequencing Batch Reactor (AnSBR)

A mixed suspended growth anaerobic process with reaction and liquid-solids separation in some vessel, much like that for aerobic sequencing batch reactors (SBRs). The operation of SBRs consists of four steps: (1) feed, (2) react, (3) settle, and (4) decant/effluent withdrawal. The settling velocity of the sludge during the settlement period before decanting the effluent is critical. Settling times used are about 30 minutes. After sufficient operating time, a dense granulated sludge develops that improves the liquid-solids separation At HRTs from 6 to 24 hours, the SRT may range from 50 to 200 days, respectively. At 25°C, 92 to 98 percent COD removal was achieved at volumetric organic loadings of 1.2 to 2.4 kg COD/m^3-d. At 5°C, COD removal ranged from 85 to 75 percent for COD loadings from 0.9 to 24 kg/m^3-d, respectively. The AnSBR schematics are shown in Fig. 2.9 .

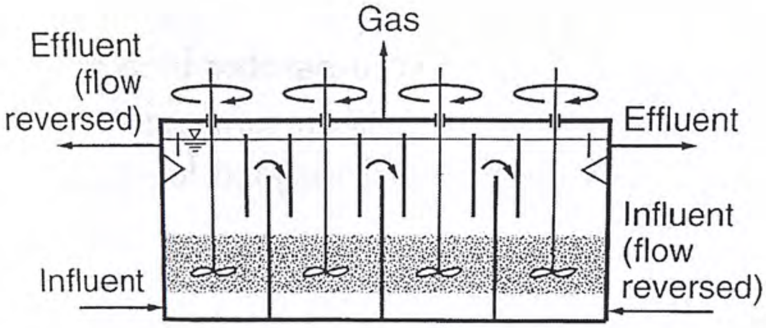


Figure 2.9 Schematics of Anaerobic Sequencing Batch Reactor [Metcalf & Eddy, 2014]

2.2 Carbonate System

2.2.1 Dissolved Carbon Dioxide in Water

Carbon dioxide (CO_2) is a gas that is relatively soluble in water; it is much more soluble than oxygen and nitrogen. When dissolved, carbon dioxide is in acid–base equilibrium with the total carbonate system, which means that its dissolved concentration is affected by the pH. Therefore, all-natural waters in contact with the atmosphere or with an inorganic carbon substrate, such as limestone, will contain carbon dioxide.

Table 2.2 Henry's Law Constants (Gases in Water at 25°C) [Sander, 2015]

Substances	**Units**			
	$\frac{L\cdot atm}{mole}$	$\frac{mole}{L\cdot atm}$	$atm\left(\frac{mol\ H_2O}{mol\ gas}\right)$	**dimensionless**
O_2	770	$1.3x10^{-3}$	$4.3x10^4$	0.032
H_2	1300	$7.8x10^{-4}$	$7.1x10^4$	0.019
CO_2	29	$3.5x10^{-2}$	$1.6x10^3$	0.830
N_2	1600	$6.1x10^{-4}$	$9.1x10^4$	0.015
He	2700	$3.7x10^{-4}$	$1.5x10^5$	0.0091
Ne	2200	$4.5x10^{-4}$	$1.2x10^5$	0.011
Ar	710	$1.41x10^{-3}$	$4.0x10^4$	0.034
CO	1100	$9.5x10^{-4}$	$5.8x10^4$	0.023

When water is contacted with the atmosphere, the equilibrium concentration of carbon dioxide in the water is proportional to the amount of carbon dioxide in the air, according to Henry's Law constant, H (see Table 2.2). Air contains a mole fraction of about 0.00032 mole of carbon dioxide per mole of air, which is equivalent to a partial pressure of 0.00032 atm. According to Henry's law, water in contact with the atmosphere at 25 °C has an equilibrium concentration of about 0.5 mg/L (0.5 ppm). That is,

$$x_{CO_2} = \frac{P_{CO_2}}{H} \tag{2.1}$$

$$= 0.00032 \ \text{atm} \times \frac{\text{mol CO}_2}{1600 \ \text{atm} \cdot \text{mol H}_2\text{O}}$$

$$\times \frac{55.6 \ \text{mol H}_2\text{O}}{\text{L}} \times \frac{44 \ \text{g}}{\text{mol CO}_2} \times \frac{10^3 \ \text{mg}}{1 \ \text{g}}$$

$$x_{CO_2} = 0.5 \ \text{mg / L CO}_2$$

2.2.2 Carbonate Chemistry [Moran, 2010]

There are two points to note about the dynamics of CO_2 in water that are important when considering degassing. The first is that, unlike other gases important to aquacultures such as nitrogen (N_2) and oxygen (O_2), when CO_2 dissolves in water it undergoes chemical reactions, forming a number of inorganic carbon species and H^+. A generalized description of the carbon species present in water is as follows:

$$CO_2\,(g) \rightleftharpoons CO_2\,(aq)$$

$$CO_2\,(aq) + H_2O\,(l) \rightleftharpoons H_2CO_3\,(aq)$$

$$H_2CO_3\,(aq) \rightleftharpoons H^+\,(aq) + HCO_3^-\,(aq)$$

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

Carbon dioxide in the gas phase is dissolved in water to form carbon dioxide in the aqueous phase, CO_2 (aq). And then, CO_2 (aq) reacts reversibly with water to form carbonic acid (H_2CO_3). Because carbonic acid is a diprotic acid it will dissociate in two steps. Thus, there are two dissociation constants, first of which is for the dissociation into the bicarbonate (also called hydrogen carbonate) ion, HCO_3^-(see Eq. 2.3). The second constant is for the dissociation of the bicarbonate ion into the carbonate ion, CO_3^{2-}(see Eq. 2.4).

Table 2.3 Carbonate Equilibrium Constants as a Function of Temperature [Metcalf & Eddy, 2014]

Temperature, °C	**Equilibrium constant**	
	$K_1 \times 10^7$	**$K_2 \times 10^{11}$**
5	3.020	2.754
10	3.467	3.236
15	3.802	3.715
20	4.169	4.169
25	4.467	4.477
30	4.677	5.129
40	5.012	6.026

The concentration of the unionized aqueous species (H_2CO_3 (aq) and CO_2 (aq)) are often summed into a single hypothetical species (H_2CO_3*) because of the low hydration of CO_2 (aq), meaning H_2CO_3* effectively represents the concentration of dissolved CO_2. The degree to which CO_2 dissolves in water is determined by the solubility constant (K_0)

$$K_0 = \frac{[H_2CO_3^*]}{[CO_2 (g)]} \tag{2.2}$$

Which is dependent on the temperature and ionic composition of the water. The concentration of HCO_3^- and CO_3^{2-} at equilibrium can be calculated from the dissociation constants K_1 and K_2 (as shown in Table 2.3)

$$K_1 = \frac{[H^+][HCO_3^-]}{[H_2CO_3^*]} \tag{2.3}$$

$$K_2 = \frac{[H^+][CO_3^{2-}]}{[HCO_3^-]} \tag{2.4}$$

As with K_0, the dissociation constants are dependent on the temperature and ionic composition of the water. When considering the dynamics of the carbonate system during CO_2 stripping, an important aspect is a relative proportion, or ionization fraction (α), of the inorganic carbon into the different species.

While K_0, K_1, and K_2 are solely dependent on water temperature and ionic composition, the ionization fraction also depends on H^+ concentration (Stumm and Morgan, 1996), such that

$$\text{a)}\ [H_2CO_3^*] = [C_T]\cdot\alpha_0 \quad \text{b)}\ [HCO_3^-] = [C_T]\cdot\alpha_1 \quad \text{c)}\ [CO_3^{2-}] = [C_T]\cdot\alpha_2 \tag{2.5}$$

$$\alpha_0 = \left(1 + \frac{K_1}{[H^+]} + \frac{K_1K_2}{[H^+]^2}\right)^{-1} \tag{2.6}$$

$$\alpha_1 = \left(\frac{[H^+]}{K_1} + 1 + \frac{K_2}{[H^+]}\right)^{-1} \tag{2.7}$$

$$\alpha_2 = \left(\frac{[H^+]^2}{K_1K_2} + \frac{[H^+]}{K_2} + 1\right)^{-1} \tag{2.8}$$

where C_T represents the total inorganic carbon content of the water (see Eq. 2.9).

$$C_T = [H_2CO_3^*] + [HCO_3^-] + [CO_3^{2-}] \tag{2.9}$$

In a system where the alkalinity remains a fixed property of the water (such as inside a CO_2 stripper), the H^+ concentration is proportional to the C_T concentration. This means that as CO_2 is stripped of water, the C_T and H^+ concentration decreases, and the ionization fraction changes. Fig. 2.10 illustrates how the ionization fraction of carbon species changes with ionic composition and C_T

concentration at a particular temperature and alkalinity. Two noteworthy patterns in Fig. 2.10 are that (1) the ionization fraction of H_2CO_3* decreases as the C_T concentration (and H^+) decreases, and (2) for any given C_T concentration, more C_T is present as H_2CO_3* in seawater compared to saline water of the same salinity, which in turn has more H_2CO_3* than freshwater.

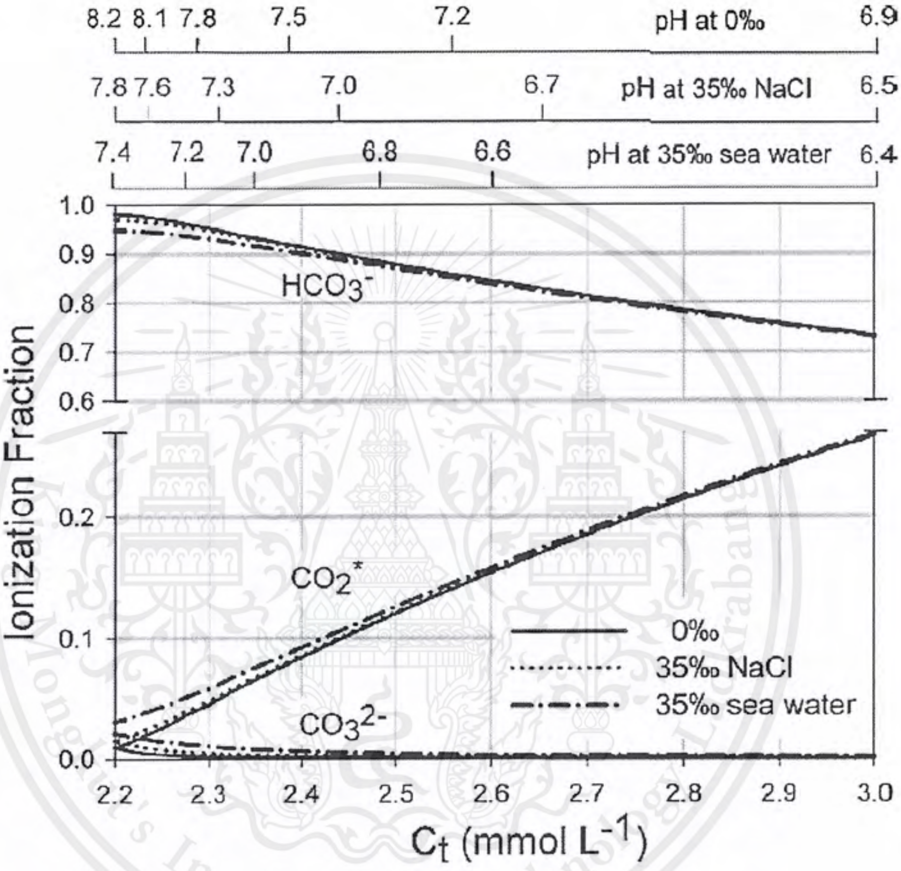


Figure 2.10 Ionization Fractions of Inorganic Carbon Species in Freshwater (0%), Seawater (35%) and Saline Water (35% NaCl) at 15 °C and a Total Alkalinity of 2.2 mmol L^{-1} [Moran, 2010]

The second important point to consider with respect to the dynamics of CO_2 in water is that the dehydroxylation of HCO_3^- to CO_2 is a slow process in terms of chemical equilibria, with a half-life of approximately 22 seconds. Such a slow process is somewhat unique in comparison to other chemical equilibria encountered in aquaculture engineering. The slow dehydroxylation rate means that as a body of water passes through a stripping unit (which often only takes a few seconds), there is effectively no replacement of the stripped CO_2 fraction from the pool of other carbon species. The CO_2 fraction is not replenished until the water re-establishes equilibrium approximately 1min after leaving the stripping unit. This complicates the calculation of CO_2 stripping rates from a practical viewpoint.

2.2.3 Alkalinity [มั่นสิน, 1994]

Alkalinity is defined as the ability of water to adjust acidic water to neutral. The alkalinity of water was measured by titration with a standard acid solution.

Total alkalinity can be calculated as follows:

$$\text{Alkalinity} = (HCO_3^-) + (CO_3^{2-}) + (OH^-) - (H^+) \tag{2.10}$$

$$\text{Or} \quad = [HCO_3^-] + [CO_3^{2-}] + [OH^-] - [H^+] \tag{2.11}$$

The difference in Eq. 2.10 and Eq. 2.11 is concentration unit. Parameters in Eq. 2.10 are expressed in equivalent weight per liter such as milligram per liter as calcium carbonate (mg/L as $CaCO_3$), equivalent per liter (eq/L) and etc. parameters in Eq. 2.11 are expressed in mole per liter (mol/L).

Water temperature does not influence alkalinity, meaning water alkalinity is constant at any temperature. In contrast, a solubility of carbonate compounds depends on temperature, such the raising temperature $CaCO_3$ precipitation will increase. Therefore, both of CO_3^- and alkalinity will decrease too. So, alkalinity depends on $CaCO_3$ precipitation or other carbonate compounds. Alkalinity does not change if does not $CaCO_3$ precipitation or $CaCO_3$ dissolution.

2.3 CO_2 Stripping

2.3.1 Theory of Gas Transfer

Proper design and operation of aeration and air stripping devices require a fundamental understanding of equilibrium partitioning of chemicals between air and water as well as an understanding of the mass transfer rate across the air-water interface is required.

Equilibrium [Rousseau, 1987]

The most important physical property data required for the design of strippers are gas-liquid equilibria. Since equilibrium represents the limiting condition for any gas-liquid contact, such data are needed to define the maximum lean solution purity attainable in strippers. Equilibrium data also are needed to establish the mass transfer driving force, which can be defined simply as the difference between the actual and equilibrium conditions at any point in a contactor. Equilibrium data are presented in a variety of ways. Frequently, the solubility of gases in liquids in which they are sparingly soluble are given in terms of Henry's Law constant, H. Henry's Law states simply that the solubility of a gas in a liquid is directly proportional to its partial pressure in the gas phase; that is,

$$p_A = H \cdot x_A \tag{2.12}$$

where p_A is the partial pressure of A in the gas phase, and x_A is the concentration of A in the liquid phase.

The constant H has the units of pressure per composition. Typical units are atmospheres for p_A and mole fraction for x_A. H is dependent on temperature but relatively independent of system pressure at moderate pressure levels for systems where Henry's Law applies. Henry's Law constants for many gases in water are given in Table 2.2. The partial pressure of component A can be correlated with Raoult's law:

$$p_A = P \cdot y_A \tag{2.13}$$

where P is the total pressure of gas phase, and y_A is the mole fraction of A in the gas phase.

Mass Transfer (Two-film Theory) [Metcalf & Eddy, 2014]

The driving force for mass transfer between one phase and another derives from the displacement of the system from equilibrium. Fig. 2.11 displays the situation where mass is being transferred from the water to the air also called stripping. The contaminant concentration in the water is high relative to the equilibrium concentration between the air and water. The tendency to achieve equilibrium is sufficient to cause diffusion of the aqueous-phase contaminant molecules from the bulk solution at some concentration, C, to the air-water interface, where the aqueous phase concentration is C_i. Because C is larger than C_i, the difference between them provides the aqueous-phase driving force for stripping. Similarly, the contaminant concentration in the air at the air-water interface, P_i, is larger than the contaminant concentration in the bulk air, P, and diffusion causes the molecules to migrate from the air-water interface to the bulk air. The difference between P_i and P is the driving force for stripping in the gas phase.

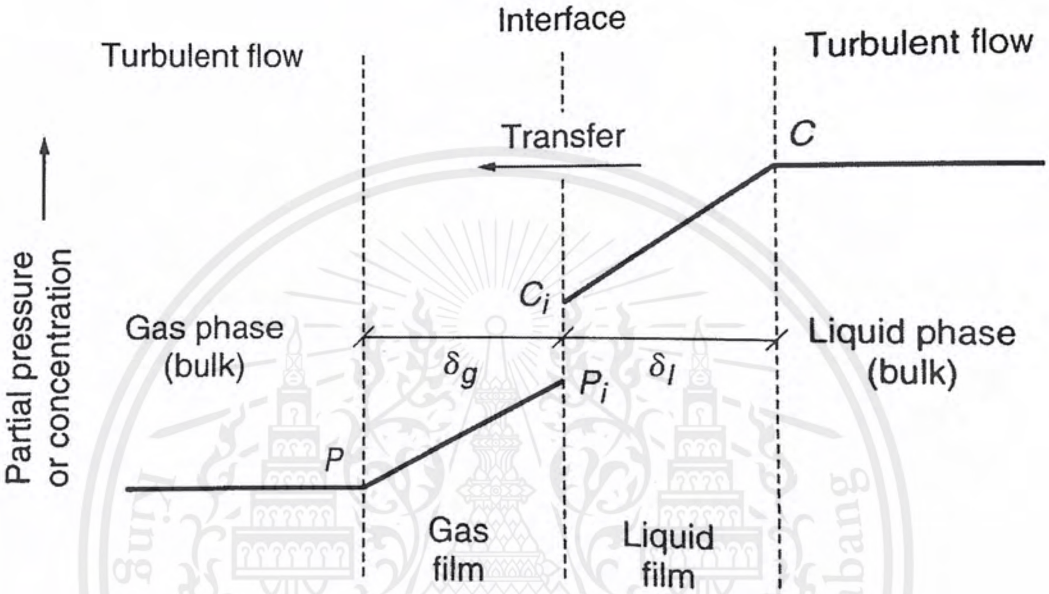


Figure 2.11 Schematic for Two-film Theory of Gas Transfer for Stripping [Metcalf & Eddy, 2014]

Steady-State Mass Transfer [Metcalf & Eddy, 2014]

Under steady-state conditions, the rate of mass transfer of a gas through the gas film must be equal to the rate transfer through the liquid film. Using Fick's first law, the mass flux for each phase for stripping is written as follows:

$$r = k_G\left(P_G - P_i\right) = k_L\left(C_i - C_L\right) \tag{2.14}$$

where r is the rate of mass transferred per unit area per unit time, k_G is the gas film mass transfer coefficient, P is the partial pressure of constituent A in the bulk of gas phase, P_i is the partial pressure of constituent A at the interface in equilibrium with concentration C_i of constituent A in liquid, k_L is the liquid film mass transfer coefficient, C_i is the concentration of constituent A at the interface in equilibrium with partial pressure P_i of constituent A in the gas phase, and C is the concentration of constituent A in the bulk liquid phase. If the terms (P_G-P_i) and (C_i-C_L) are divided by their respective film thickness values (δ_G and δ_L), the driving force can be expressed in terms of unit thickness. Thus, the degree of mass transfer can be enhanced by reducing the thickness of the film, depending on which is the controlling film.

Overall Mass Transfer Coefficients [Metcalf & Eddy, 2014]

Because it is difficult to measure the values of k_G and k_L at the interface, it is common to use overall coefficients K_G and K_L, depending on whether the resistance to mass transfer is on the gas or liquid assumed that side. If it is assumed that essentially all of the resistance to mass transfer is caused by the liquid film, then the rate of mass transfer can be defined as follows in terms of the overall liquid mass transfer coefficient:

$$r = K_L (C_S - C_L) \tag{2.15}$$

where r is the rate of mass transferred per unit area per unit time, K_L is the overall liquid mass transfer coefficient, C_S is the concentration of constituent A at the interface in equilibrium with the partial pressure of constituent A in the bulk gas phase, and C_L is the concentration of constituent A in the bulk liquid phase.

If the two expressions are given by Eqs. 2.15 and 2.14 above are equated, the following relationship can be derived from the overall liquid mass transfer coefficient and the gas and liquid film coefficients:

$$r = K_L\left(C_S - C_L\right) = k_G\left(P_G - P_i\right) = k_L\left(C_i - C_L\right) \tag{2.16}$$

Because it was assumed that essentially all of the resistance to mass transfer is caused by the liquid film, the following relationships, based on Henry's law, must apply at the interface:

$$P_G = HC_S \text{ and } P_i = HC_i$$

It will now be noted that the overall driving force (C_S-C_L) in Eq. 2.16 can be written as

$$\left(C_S - C_L\right) = \left(P_G - P_i\right) + \left(C_i - C_L\right) \tag{2.17}$$

Substituting for P_G and P_i in Eq. 2.16 and combining Eq. 2.16 and 2.17, the following relationship is obtained where the liquid film controls the mass transfer:

$$\frac{r}{K_L} = \frac{r}{k_L} + \frac{r}{Hk_G} \text{ or } \frac{1}{K_L} = \frac{1}{k_L} + \frac{1}{Hk_G} \tag{2.18}$$

In similar manner it can be shown that the following relationship holds if the transfer of mass is controlled by the gas film:

$$\frac{1}{K_G} = \frac{H}{k_L} + \frac{1}{k_G} \tag{2.19}$$

2.3.2 Selection of Equipment [Rousseau, 1987]

1) Tower Selection

A great variety of equipment designs have been developed for the performance of absorption and stripping operations. The three most commonly used are tray columns, packed columns, and spray contactors, although numerous other types are used for special applications.

Tray columns (also called plate columns) are particularly well suited for large installations; clean, noncorrosive, non-foaming liquids; and low-to-medium liquid flow rate applications. They also are preferred when internal cooling is required in the column. With appropriate tray design, cooling coils can be installed on individual trays, or alternatively, the liquid can be removed from the column at one tray, cooled, and returned to another tray. Tray columns also are advantageous for separations that require a large number of transfer units because they are not subject to channeling of vapor and liquid streams which can cause problems in tall packed columns. The efficiency of tray columns will lose due to flooding, priming, coning, weeping, and dumping. *Flooding*, this effect can be well understood as simple as liquid filling up from the bottom of the column to the top and exhausted out from the top inlet of the column. Downcomer and space between the trays are completely filled up by the liquid than the tower is said to be flooded, due to high pressure drop due to increased flow rates of the streams. *Priming* is an exaggerated condition of liquid entrainment. Due to high gas velocity, liquid from the bottom trays is carried away along with the vapor to the top trays. *Coning* occurs due to low liquids flow velocities when compared to gas which results in pushing of the liquid away from the tray openings. *Weeping* is due to at low gas velocity which is not equal to liquid flow velocity, and the liquid is not enough resisted to hold on the tray pass from the downcomers, the complete liquid will flow through the openings in the tray itself. *Dumping* similarly to severe weeping, but no liquid reaches the downspouts. Complete liquid drops down by the tray opening only.

Packed columns are gaining favor for a wide range of applications because of the development of new packings which provide higher capacity and better performance as well as the development of better design techniques for packed columns. The packed column listing in Table 2.4 refers to random packing, which is the type most commonly used in large absorption columns. However, it should be noted that some of the newer very high efficiency and very low pressure drop packings involve packing elements that are stacked individually within the column in a special ordered arrangement.

Packed columns are preferable to tray columns for small installations, corrosive service, liquids with a tendency to foam, very high liquid-to-gas ratios, and low pressure drop applications. In addition, packed columns offer greater flexibility because the packing can be changed with relative ease to modify column operating characteristics, if necessary. Commercially available random packings, which offer better overall performance than the older Raschig rings and Berl saddles, include Pall rings and Intalox saddles. Examples of very high efficiency, structurally ordered packing elements include Sulzer packing and Goodloe packing.

Spray contactors are used almost entirely for applications where pressure drop is critical, such as flue gas scrubbing. Generally, they are not suitable for applications requiring several stages of contact or a close approach to equilibrium. In spray contactors, the contact area for mass transfer is related directly to the volume of liquid sprayed into the unit. It is therefore common practice to recycle the absorbent to increase mass transfer. Types of equipment classified as spray contactors include countercurrent spray columns, venturi scrubbers, cyclone scrubbers, showerdeck towers, and spray dryers.

Table 2.4 Selection Guideline for Stripping Tower

	Tray Towers	**Packed Towers**	**Spray Towers**
Pressure Drop	High gas pressure drops	High gas pressure drops but lower than tray towers	High liquid pressure drops
Liquid Flowrate	Low-medium	High	High
Fouling	Fouling results in loss of efficiency, and increased pressure drop	Fouling results in loss of efficiency, and increased pressure drop	Fouling of nozzle may reduce treatment efficiency and increased pressure drop
Efficiency	Medium due to flooding, priming, coning, weeping, and dumping may occur	High due to high mass transfer	Low due to low mass transfer and low residence time
Capital Cost (high liquid flow rate)	High	High	Medium
Annual O&M Costs (high liquid flow rate)	Medium	Medium	High

2) Packing Material Selection

A packed tower is selected. Thus, packing material must be selected. Packing material is the heart of the tower. It provides the surface over which the scrubbing liquid flows, presenting a large area for mass transfer to occur. Packing material represents the largest material cost of the packed tower. Pictured in Fig. 2.12 are some of the more commonly used packings. These materials were originally made of stoneware, porcelain, or metal, but presently, a large majority are made of high-density thermoplastics (polyethylene and polypropylene).

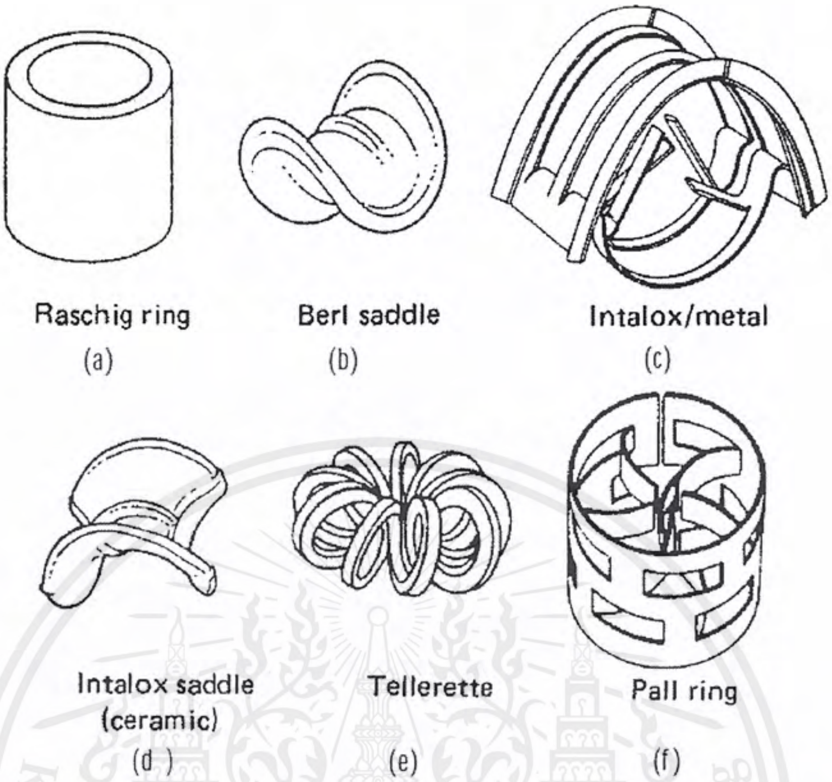


Figure 2.12 Common Random Packing Materials [Perry, 1997]

Specific packing selected for an industrial application depends on the nature of the contaminants, geometric mode of contact, size of the absorber, and scrubbing objectives. The following factors provide a general guide for selecting packing materials (MacDonald 1977):

1) *Cost* - Generally, plastic packing is less expensive than metal packing, with ceramic packing being the most expensive. Packing costs are expressed in dollars per cubic meter ($\$/m^3$).

2) *Low Pressure Drops* - Pressure drop is a function of the volume of void space in a tower when filled with packing: generally, the larger the packing size for a given bed size, the smaller the pressure drop becomes.

3) *Corrosion Resistance* - Ceramic or porcelain packings are commonly used in a very corrosive atmosphere.

4) *Large Specific Area* - A large surface area per cubic foot of packing, m^2/m^3, is desirable for mass transfer.

5) *Structural Strength* - Packing must be strong enough to withstand normal loads during installation, service, physical handling, and thermal fluctuations. Ceramic packing may crack under sudden temperature changes.

6) *Weight* - Heavier packing may require additional support materials or heavier tower construction. Plastics have a big advantage in this area since they are much lighter than either ceramic or metal packings.

7) *Design Flexibility* - The efficiency of a scrubber changes as the liquid and gas flow rates vary. Packing material must be able to handle the process changes without substantially affecting removal efficiency.

8) *Arrangement* - Packing material may be arranged in an absorber in one of two ways. The packing may be dumped into the column randomly or, in certain cases, systematically stacked, as bricks are laid atop each other.

Packing Size [Sinnott, 2005]

In general, the largest size of packing that is suitable for the size of the column should be used, up to 50 mm. Small sizes are appreciably more expensive than the larger sizes. Above 50 mm the lower cost per cubic meter does not normally compensate for the lower mass transfer efficiency. Use of too large a size in a small column can cause poor liquid distribution.

Recommended size ranges are:

Column diameter	Use packing size
<0.3 m (1 ft)	<25 mm (1 in.)
0.3 to 0.9 m (1 to 3 ft)	25 to 38 mm (1 to 1.5 in.)
>0.9 m	50 to 75 mm (2 to 3 in.)

Table 2.5 Characteristics of Common Random Packing Materials [Adapted from Perry, 1987]

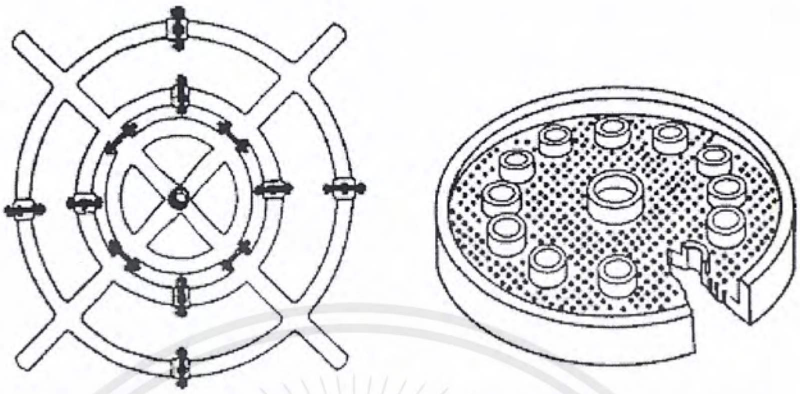
Name	**Material**	**Nominal Size (mm)**	**Bed Weight, kg/m^3**	**Area, m^2/m^3**	**% Voids**	**Packing Factor F_p, m^{-1}**
Rascig Rings	C	6	960	710	62	-
		13	880	370	64	1900
		25	670	190	74	587
		50	660	92	74	213
		75	590	62	75	121
	M	19	1500	245	80	984
		25	1140	185	86	472
		50	590	95	92	187
		75	400	66	95	105
Berl Saddles	C	6	900	900	60	-
		13	865	465	62	790
		25	720	250	68	360
		38	640	150	71	215
		50	625	105	72	150
Intalox Metal	M	25	352	230	97	134
		40	237	154	97	79
		50	150	98	98	59
		70	130	56	98	39
Intalox Saddles	C	6	864	984	65	302
		13	736	623	71	-
		25	672	256	73	302
		50	608	118	76	131
		75	576	92	79	72
Tellerettes	P	25	112	180	87	-

Name	**Material**	**Nominal Size (mm)**	**Bed Weight, kg/m^3**	**Area, m^2/m^3**	**% Voids**	**Packing Factor F_p, m^{-1}**
		50	59	125	93	-
Pall Rings	M	16	-	-	92	256
		25	480	205	94	183
		38	415	130	95	131
		50	385	115	96	89
		90	270	92	97	59
	P	25	80	206	90	180
		50	61	102	92	85
		90	53	85	92	56
	C	25	620	220	75	351
		38	540	165	78	180
		50	555	120	78	141

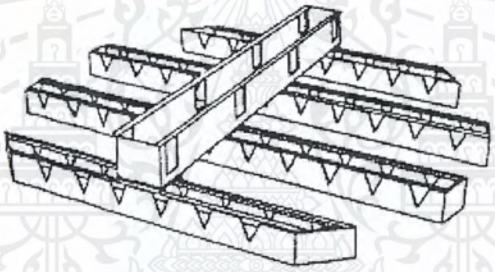
Note: M = metal, carbon steel. Other metals available. P = plastic, polypropylene. Other plastics available. C = ceramic.

3) Liquid Distributors

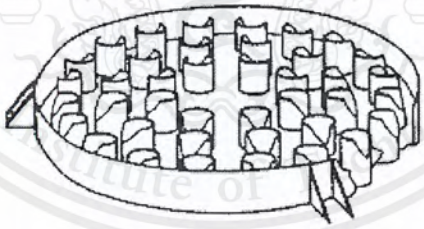
Several distributor types are available: trough, orifice-rise, and perforated pipe. Examples of these types are shown in Fig.2.13 The trough distributor provides good distribution under widely varying flow rates of gas and liquid. The liquid may flow through simple V-notch weirs, or it may flow through tubes that extend from the troughs to near the upper level of the packing. Some deposition of solids can be accommodated. In the weir design, the liquid is introduced into a trough with holes at the top. The liquid fills to the top and spills over onto the packing or another trough for redistribution. These weir designs have the advantage of being open and not plugged easily. The perforated-tube provides good liquid distribution patterns. However, the holes are subject to plugging if any particles or contaminants are in the liquid. Packed towers, designed with spray nozzles to distribute liquid, operate better with a few large nozzles than with many small nozzles. Large nozzles are less susceptible to plugging. Small nozzles that produce a finer spray are not needed in a packed tower because pollutant collection occurs on the wetted packing and not by the liquid droplets.



Perforated pipe distributor Orifice - type distributor



Trough - type distributor



Weir - riser distributor

Figure 2.13 Typical Liquid Distributors [Perry, 1997]

4) Pump Selection [Perry, 1997]

When selecting pumps for any service, it is necessary to know the liquid to be handled, the total dynamic head, the suction and discharge heads, and, in most cases, the temperature, viscosity, vapor pressure, and specific gravity. The following factors provide a general guide for selecting pumps:

- *Range of Operation* Because of the wide variety of pump types and the number of factors which determine the selection of any one type for a specific installation, the designer must first eliminate all but those types of reasonable possibility. Since the range of operation is always an important consideration, Fig. 2. 14 should be of assistance.

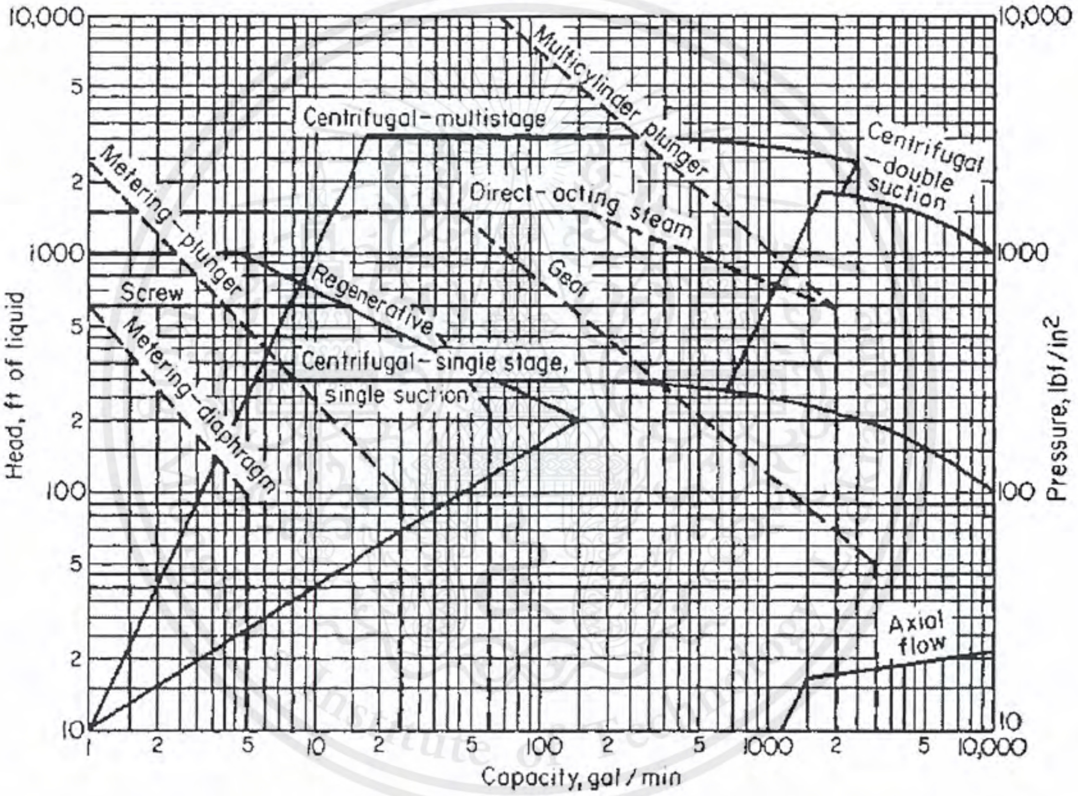


Figure 2.14 Pump Coverage Chart based on Normal Ranges

Solid lines: use left ordinate, head scale. Broken lines: use right ordinate, pressure scale.

To convert gallons per minute to cubic meters per hour, multiply by 0.2271. [Perry, 1997]

- *Pump Materials of Construction* In the chemical industry, the selection of pump materials of construction is dictated by considerations of corrosion, erosion, personnel safety, and liquid contamination. The experience of pump manufacturers is often valuable in selecting materials. See section on materials.

- *Presence of Solids* When a pump is required to pump a liquid containing suspended solids, there are unique requirements which must be considered. Adequate clear-liquid hydraulic performance and the use of carefully selected materials of construction may not be all that is required for satisfactory pump selection. Dimensions of all internal passages are critical. Pockets and dead spots, areas where solids can accumulate, must be avoided. Close internal clearances are undesirable because of abrasion. Flushing connections for continuous or intermittent use should be provided. For installations in which suspended solids must be handled with a minimum of solids breakage or degradation, such as pumps feeding filter presses, special attention is required; either a low-shear positive displacement pump or a recessed-impeller centrifugal pump may be called for.

Centrifugal Pumps The centrifugal pump is the type most widely used in the chemical industry for transferring liquids of all types—raw materials, materials in manufacture, and finished products—as well as for general services of water supply, boiler feed, condenser circulation, condensate return, etc. These pumps are available through a vast range of sizes, in capacities from 0.5 m^3/h to 2×10^4 m^3/h, and for discharge heads (pressures) from a few meters to approximately 48 MPa. The size and type best suited to a particular application can be determined only by an engineering study of the problem. The primary advantages of a centrifugal pump are simplicity, low first cost, uniform (non-pulsating) flow, small floor space, low maintenance expense, quiet operation, and adaptability for use with a motor or a turbine drive. A centrifugal pump, in its simplest form, consists of an impeller rotating within a casing. The impeller consists of a number of blades either open or shrouded, mounted on a shaft that projects outside the casing. The centrifugal pump is shown in Fig 2.15

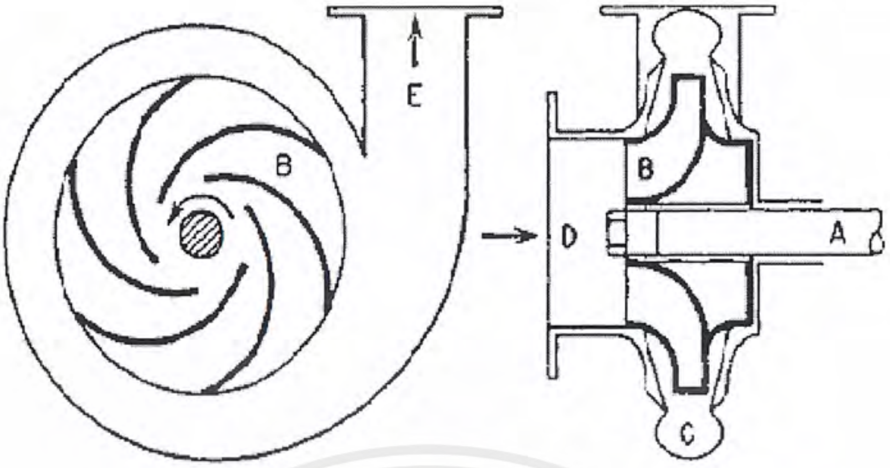


Figure 2.15 Schematics of Simple Centrifugal Pump [Perry, 1997]

Propeller and Turbine Pumps These pumps are essentially very-high-capacity low-head units. Normally they are designed for flows in excess of 450 m^3/h (2000 gal/min) against heads of 15 m (50 ft) or less. They are used to great advantage in closed-loop circulation systems in which the pump casing becomes merely an elbow in the line. A common installation is for calandria circulation.

Positive- displacement Pumps Whereas the total dynamic head developed by a centrifugal, mixed flow, or axial-flow pump is uniquely determined for any given flow by the speed at which it rotates, positive-displacement pumps and those which approach positive displacement will ideally produce whatever head is impressed upon them by the system restrictions to flow. Positive-displacement pumps may be of either the reciprocating or the rotary type. In all positive-displacement pumps, a cavity or cavities are alternately filled and emptied of the pumped fluid by the action of the pump.

Jet Pumps Jet pumps are a class of liquid-handling device that makes use of the momentum of one fluid to move another. Ejectors and injectors are the two types of jet pumps of interest to chemical engineers. The ejector, also called the siphon, exhauster, or eductor, is designed for use in operations in which the head pumped against is low and is less than the head of the fluid used for pumping

Electromagnetic Pumps The necessity of circulating liquid-metal heat-transfer media in nuclear-reactor systems has led to the development of electromagnetic pumps. All electromagnetic pumps utilize the motor principle: a conductor in a magnetic field, carrying a current which flow at right angles to the direction of the field, has a force exerted on it, the force being mutually perpendicular to both the field and the current. In all electromagnetic pumps, the fluid is the conductor. This force, suitably directed in the fluid, manifests itself as a pressure if the fluid is suitably contained. The field and current can be produced in a number of different ways and the force utilized variously. Electromagnetic pumps are available for flow rates up to 2.271 × 103 m^3/h (10,000 gal/min), and pressures up to 2 MPa (300 lbf/in^2) are practical. Performance characteristics resemble those of centrifugal pumps.

5) Blower Selection

Blower selection depends on the volume flow rate, pressure, type of material handled, space limitations, and efficiency. Centrifugal Blowers These blowers have air or gases entering in the axial direction and being discharged in the radial direction. These blowers have 3 types of blades, radial or straight blades, forward curved blades, and backward curved blades (as shown in Fig. 2.16 -2.18).

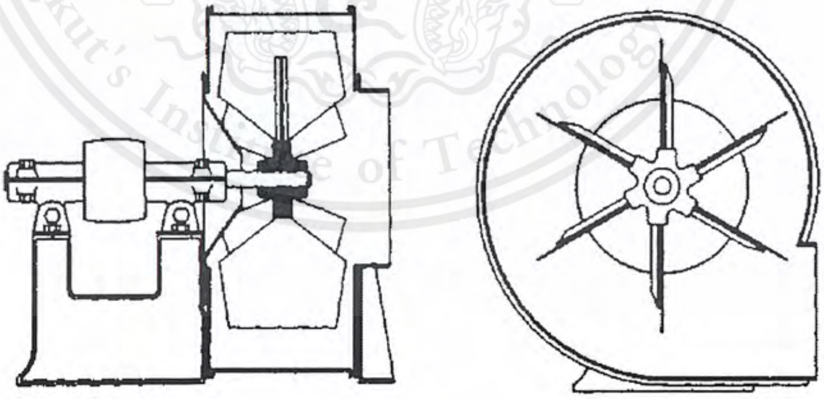


Figure 2.16 Schematics of Straight-blade, or Steel-plate, Fan [Perry, 1997]

Radial blade blowers as seen in Fig. 10-68 are usually used in large diameter or high-temperature applications. The blades being radial in direction have very low stresses as compared to the backward or forward curve blades. The rotors have anywhere between 4 to 12 blades and usually operate at low speeds. These fans are used in exhaust work especially for gases at high temperature and with suspensions in the flow stream.

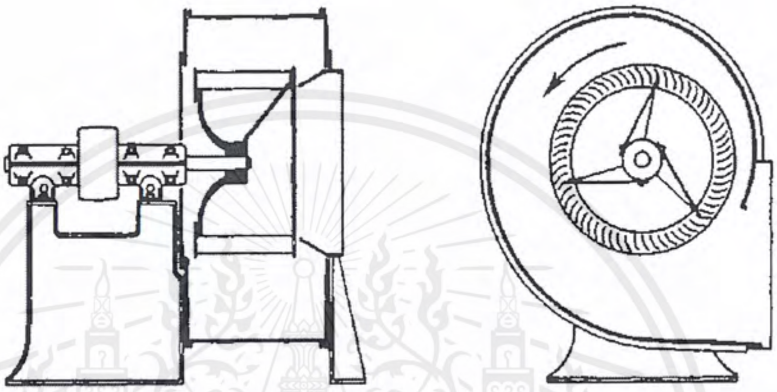


Figure 2.17 Schematics of Forward-curved Blade, or "Scirocco"-type, Fan [Perry, 1997]

Forward-Curved Blade Blowers, these blowers discharge the gas at a very high velocity. The pressure supplied by this blower is lower than that produced in the other two blade characteristics. The number of blades in such a rotor can be large up to 50 blades and the speed is high usually 3600–1800 rpm in 60-cycle countries and 3000–1500 rpm in 50-cycle countries.

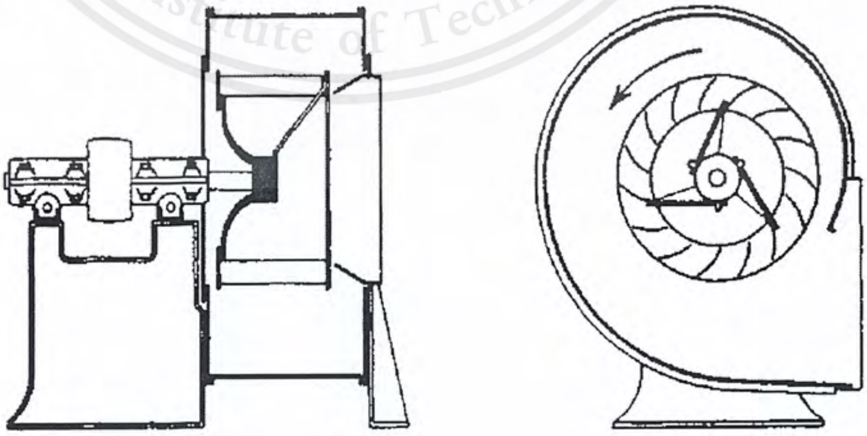


Figure 2.18 Schematics of Backward-curved Blade Fan [Perry, 1997]

Backward-Curved Blade Blowers These blowers are used when a higher discharge pressure is needed. It is used over a wide range of applications. Both the forward and backward curved blades do have much higher stresses than the radial bladed blower.

2.3.3 Tower Design

1) Material Balance

A countercurrent stripper tower is illustrated in Fig. 2.19 . The tower may contain either trays or packing. For purposes of this preliminary discussion, it is assumed that a single solute is being stripped. Since the molar flow rates of the liquid and gas streams, L_M and G_M, respectively, vary over the length of the tower due to the gain of material by the liquid and loss from the gas, it is convenient to base the material balance on flow parameters that can be considered constant. The flow rates used are usually the solute-free liquid, designated L'_M, and the solute-free (inert) gas, G'_M; however, other flow parameters may be used. The Kremser and Souders and Brown design equations, for example, are based on the lean solvent rate and the rich gas rate. Use of the solute-free flow rates requires a redefinition of concentration terms from mole fractions x and y to mole ratios X and Y, where the mole ratios are defined by the equations

$$X = \frac{x}{1-x} \tag{2.20}$$

And

$$Y = \frac{y}{1-y} \tag{2.21}$$

A material balance around the top portion of the tower then becomes

$$G'_M\left(Y - Y_2\right) = L'_M\left(X - X_2\right) \tag{2.22}$$

where G'_M is the solute-free gas flow rate, kmol/h.m^2, L'_M is the solute-free liquid flow rate, , kmol/h.m^2, x and y are mole fraction solute in liquid and gas phases, respectively, and X and Y are mole ratio of solute in liquid and gas phases, respectively.

Rearranging Eq. 2.22 gives

$$Y = \frac{L'_M}{G'_M}(X - X_2) + Y_2 \tag{2.23}$$

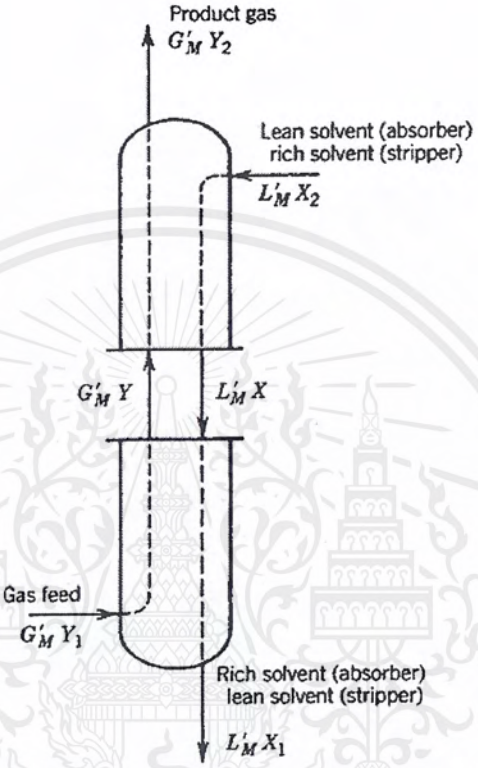


Figure 2.19 Material Balance Diagram for Countercurrent Contactor [Rousseau, 1987]

Which is a straight line with a slope of L'_M/G'_M. This is referred to as the operating line. It can be plotted readily on rectangular coordinate paper by locating the X, Y values corresponding to the top and bottom of the column, that is, X_2, Y_2 (top) and X_1, Y_1, (bottom) as shown in Fig.2.20 . Frequently, gases and liquids involved in gas absorption operations are quite dilute. When this is the case, Y is approximately equal to y and X is approximately equal to x. Also, the total molar mass flow rates L_M and G_M are approximately equal to the corresponding solute-free flow rates L'_M and G'_M so that the following approximate equation can be used:

$$y = \frac{L_M}{G_M}(x - x_2) + y_2 \tag{2.24}$$

This greatly simplifies subsequent calculations since vapor- liquid equilibrium relationships usually are given in terms of mole fractions rather than mole ratios. The equilibrium curve may be plotted on the same chart as the operating line using X and Y coordinates as shown in Fig. 2.20 (or mole fractions for the case of a dilute system). Typically, in the design of an absorber, the equilibrium curve is known and can be plotted. Also known are the flow rate of the feed gas G'_M; the mole ratio of solute in the feed gas Y_1; the mole ratio of solute (if any) in the lean solvent X_2; and the required mole ratio of solute in the product gas Y_2. Usually, it is desired to estimate the flow rate of liquid required to accomplish the separation and ultimately the dimensions of the tower.

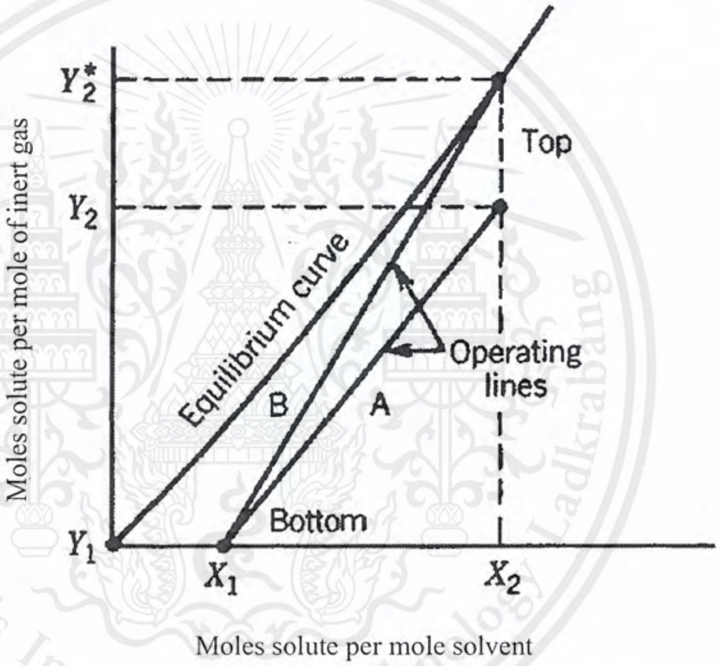


Figure 2.20 Operating Line-equilibrium Curve Diagram of Stripping g [Rousseau, 1987]

2) Determination of Height of Stripping Tower Packing [Metcalf & Eddy, 2014]

The purpose of the following analysis is to illustrate how the height of a stripping tower packing is determined, based on an analysis of the mass transfer occurring within the tower. A mass balance performed on the liquid phase within the stripping tower shown on Fig. 2.21 is as follows.

Simplified word statement

Accumulation = inflow - outflow + generation

$$\frac{\partial C}{\partial t}\Delta V = LC\Big|_{z} - LC\Big|_{z+\Delta z} + r_V \Delta V \tag{2.25}$$

where $\frac{\partial C}{\partial t}$ = change in concentration of constituent C with time (g/m^3.s)

ΔV = differential volume (m^3)

Δz = differential height (m)

L = liquid volumetric flowrate (m^3 /s)

C = concentration of constituent C (g/m^3)

r_V = rate of mass transfer of constituent C per unit volume per unit time (g/m^3.s)

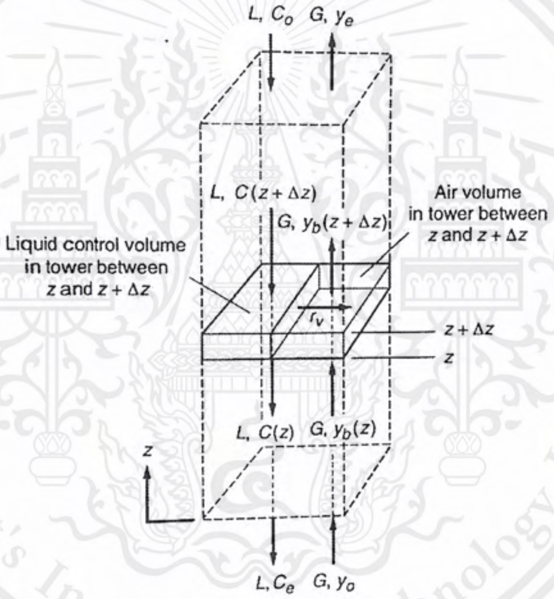


Figure 2.21 Definition Sketch for the Analysis of Mass Transfer within a Stripping Tower, Note: the packing material is not shown [Metcalf& Eddy, 2014]

Substituting area times, the differential height ($A\Delta z$) for the differential volume (ΔV) and writing the differential form for the term $LC|_{Z+\Delta Z}$ in Eq. 2.25 results in the following expression:

$$\frac{\partial C}{\partial t}A\Delta z = LC - L\left(c + \frac{\Delta C}{\Delta z}\Delta z\right) + r_V A\Delta z \tag{2.26}$$

Simplyfying Eq. 2.26 and taking the limit as Δz approaches zero yields

$$\frac{\partial C}{\partial t} = -\frac{L}{C}\frac{\partial C}{\partial z} + r_V \tag{2.27}$$

The rate of mass transfer:

$$r_V = K_L a\,(C_b - C_S) \tag{2.28}$$

where r_V = rate of mass transfer of constituent C per unit volume per unit time ($g/m^3.s$),

$K_L a$ = volumetric mass transfer coefficient which depends on water quality characteristics and temperature (1/s),

C_b = concentration of constituent C in the bulk liquid phase (g/m^3)

C_S = concentration of constituent C in liquid in equilibrium with gas as given by Henry's law (g/m^3)

Assuming steady-state conditions within the tower $\left(\frac{\partial C}{\partial t} = 0\right)$ and substituting for r_V, Eq.2.27 can be written as:

$$\frac{dC}{dz} = \frac{K_L aA}{L}\left(C_b - C_S\right) \tag{2.29}$$

The height of the tower can be obtained by integrating the above expression:

$$\int_0^z dz = \frac{L}{K_L aA}\int_{C_e}^{C_0} \frac{dC_b}{(C_b - C_S)} \tag{2.30}$$

To integrate the right-hand side of the above equation, a relationship must be found between C_b and C_S because C_S is changing continuously throughout the height of the tower.

From Henry's law, the value of C_S, is given by:

$$C_S = \frac{P_T}{H}y \tag{2.31}$$

Substituting a modified from Eq. 2.24 for y in Eq. 2.31yields:

$$C_S = \frac{P_T}{H} \times \frac{L}{G}\left(C_b - C_e\right) \qquad (2.32)$$

If Eq. 2.32 is substituted into Eq. 2.30 and the resulting expression is integrated, the following expression is obtained:

$$Z = \frac{L}{K_L aA}\left(\frac{C_0 - C_e}{C_0 - C_e - C_0^*}\right)\ln\left(\frac{C_0 - C_0^*}{C_e}\right) \qquad (2.33)$$

Where Z is the height of stripping tower packing, m and

$$C_0^* = \frac{P_T}{H} \times \frac{L}{G}\left(C_0 - C_e\right) \qquad (2.34)$$

Design Equation for Stripping Tower

Equations that can be used to determine the height of a stripping tower are as follows:

$$Z = HTU \times NTU \qquad (2.35)$$

Where Z = height of stripping tower packing material (m),

HTU = height of a transfer unit (m)

NTU = number of transfer units

The height of a transfer unit is defined as:

$$HTU = \frac{L}{K_L aA} \qquad (2.36)$$

Where L = liquid volumetric flow rate (m^3/s)

K_La = volumetric mass transfer coefficient (1/s)

A =cross-sectional area of tower (m^2)

The number of transfer units is defined as:

$$NTU = \left[\frac{C_0 - C_e}{C_0 - C_e - C_0^*}\right] \ln\left[\frac{C_0 - C_e^*}{C_e}\right] \quad (2.37)$$

Design of Stripping Towers

A stripping tower consists of a tower (usually circular cross-section), a support plate for the packing material, a distribution system for the liquid to be stripped, located above the packing material, and an air supply located at the bottom of the stripping tower (see Fig. 2.22).

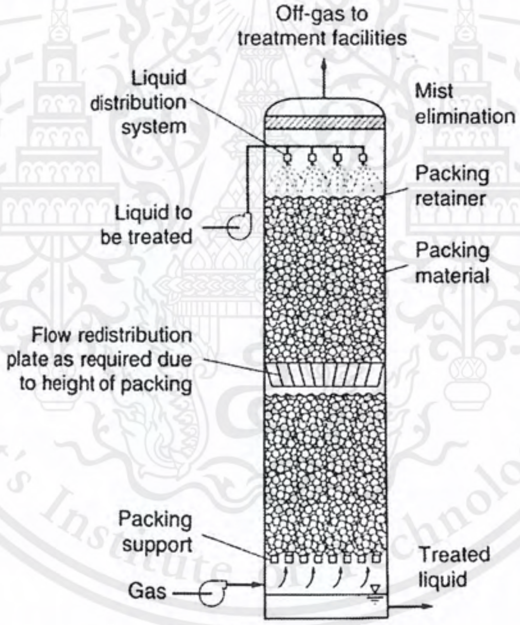


Figure 2.22 Schematic of Packed Stripping Tower [Metcalf& Eddy, 2014]

The cross-sectional area will depend on the pressure drop through the packing. The head loss through the packing is determined using a generalized gas pressure drop relationships such as plotted on Fig 6. The pressure drop is expressed in Newton per square meter per meter of depth ($(N/m^2)/m$) (i.e. 50 – 100 $(N/m^2)/m$) [Edzwald, 2011]. The unit for the x and y axis are as follows:

X axis:

$$x = \frac{L'}{G'}\left(\frac{\rho_G}{\rho_L - \rho_L}\right)^{1/2} \approx \frac{L'}{G'}\left(\frac{\rho_G}{\rho_L}\right)^{1/2} \tag{2.38}$$

Y axis:

$$y = \frac{(G')^2 (C_f)(\mu_L)^{0.1}}{(\rho_G)(\rho_L - \rho_G)} \tag{2.39}$$

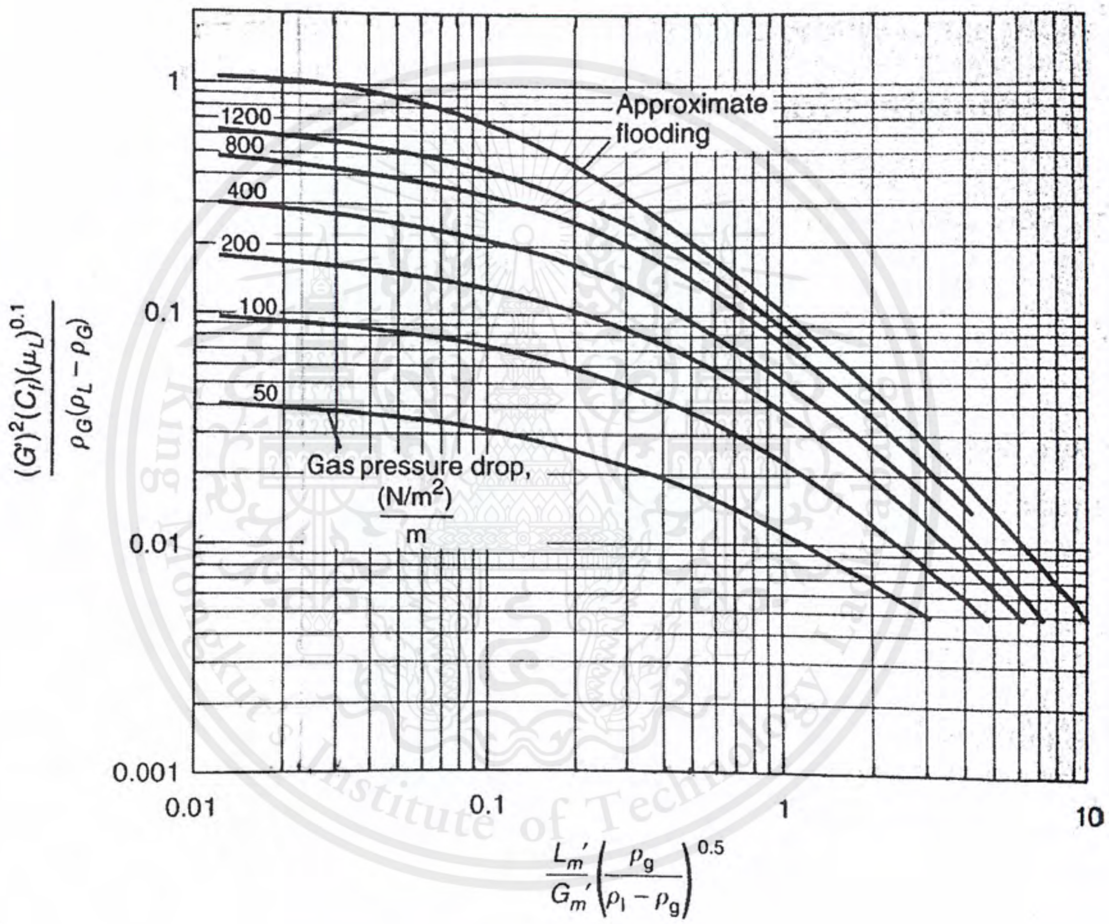


Figure 2.23 Generalized Pressure Drop Curves for Packed Stripping Towers

[Metcalf& Eddy, 2014]

Which can be written as follows:

$$G' = \left[\frac{(\text{value from y axis})(\rho_G)(\rho_L - \rho_G)}{(C_f)(\mu_L)^{0.1}} \right]^{1/2} \tag{2.40}$$

Where L’ = liquid loading rate (kg/m^2.s)

G’ = gas loading rate (kg/m^2.s)

ρ_G = density of gas (kg/m^3)

ρ_L = density of liquid (kg/m^3)

C_f = packing factor for packing material (1/m)

μ_L = viscosity of liquid (kg/m.s)

2.4 Investment Cost Estimation [Seider et al., 2009]

The purchase cost of processing equipment is generally obtained from charts, equations, or quotes from vendors. However, costs are not static. Because of inflation, they generally increase with time. Thus, charts and equations apply to a particular date, usually month and year, or to an average for a particular year. Quotes from vendors are often applicable only for a month or two. An estimate of the purchase cost at a later date is obtained by multiplying the cost from an earlier date by the ratio of a cost index, I, at that later date to a base cost index, I_{base}, that corresponds to the date that applies to the purchase cost:

$$\text{Cost} = \text{Base Cost}\left(\frac{I}{I_{base}}\right) \tag{2.41}$$

The Chemical Engineering (CE) Plant Cost Index is published in each monthly issue of the magazine Chemical Engineering, with I = 569.8 for 2017 and I_{base} = 500 for 2006.

2.4.1 Packed Tower

Packings for towers are classified as dumped (random) or structured. When packings are used, the total purchase cost of the packed tower given by the following equation:

$$C_P = F_M C_V + C_{PL} + V_p C_{PK} + C_{DR} \tag{2.42}$$

where F_M is the material factor for the vessel, F_M= 1.0 for carbon steel, 1.2 for low- alloy steel, and 1.7 for stainless steel 304, C_V is the f.o.b. purchase cost for vertical tower, V_P is the volume of the packing in cubic feet, C_{PK} is the installed cost of the packing in dollars per cubic foot (as shown in Table 2.6), and C_{DR} is the installed cost of high performance liquid distributors and redistributors required for obtaining satisfactory performance with packings.

C_V is the f.o.b. purchase cost for vertical tower including nozzles, manholes, and supports, based on the weight in pounds of the shell and the two heads, W. the correlating equations is

Tower for W = 9,000 to 2,500,000 lb:

$$C_V = \exp\{7.2756 + 0.18255[\ln(W)] + 0.02297[\ln(W)]^2\} \tag{2.43}$$

The weight, W, in the cost correlations for a pressure vessel or tower depends on the wall thicknesses of the shell and the two heads. Although the thickness of the heads may be required to be somewhat thicker than the shell, particularly at high pressures, it is sufficient for cost-estimation purposes to assume head thicknesses equal to the shell thickness, t_S, at low pressure t_s equal ½ inches. Then, with 2:1 elliptical heads, the weight of the shell and the two heads is approximately:

$$W = \pi (D_i + t_S)(L + 0.8D_i)t_S\rho \tag{2.44}$$

where the term in L accounts for the cylinder, the term in 0.8 Di accounts for the two heads, and ρ is the density of the carbon steel, which can be taken as 490 lb/ft^3.

C_{PL} from Eq.2.42 is the platforms and ladders depend on the vessel inside diameter, D_i, in feet and, for a vertical vessel, on the tangent-to-tangent length of the shell, L, in feet, and is given by:

$$C_{PL} = 361.8(D_i)^{0.73960}(L)^{0.70684} \qquad (2.45)$$

for inside diameter (D_i) between 3 and 21 ft and length of the shell (L) between 12 and 40.

Table 2.6 Installed costs of dumped packing ($/ft^3) [Adapted from Seider et al., 2009]

Types	**Material**	**Size**			
		1 in.	**1.5 in.**	**2 in.**	**3 in.**
Berl Saddles	Ceramic	48	37	28	-
Raschig Rings	Carbon Steel	54	40	34	26
	Stainless Steel	180	139	110	63
	Ceramic	26	21	19	15
Intalox Saddles	Ceramic	34	28	24	19
	Polypropylene	37	-	23	12
Pall Rings	Carbon Steel	49	37	32	-
	Stainless Steel	168	129	110	-
	Polypropylene	37	26	21	16

2.4.2 Pump and Electric Motor

Centrifugal Pumps

Much purchase-cost data for centrifugal pumps, of the most common radial type, have been published. The cost most often includes the pump, a base plate, and a direct-drive coupling. There is no general agreement on the equipment size factor to be used for correlating purchase costs. Most common are (1) brake horsepower and (2) the product of the capacity and the head (or pressure increase). Their size factor, S, which recognizes the fact that a given centrifugal pump can operate over a range of flow rate and head combinations, is

$$S = Q(H)^{0.5} \tag{2.46}$$

Where Q = flow rate through the pump (gpm)

H = pump head (ft), pressure rise/liquid density

The pump purchase cost is correlated with maximum values of S that the pump can handle. That can be calculated by following equation, which is valid from S = 400 to S =100,000:

$$C_B = \exp\left\{9.7171 - 0.6019\left[\ln(S)\right] + 0.0519\left[\ln(S)\right]^2\right\} \tag{2.47}$$

For other types of centrifugal pumps and other materials of construction, the f.o.b. purchase cost is given by:

$$C_P = F_T F_M C_B \tag{2.48}$$

Where F_M = material factor, F_M = 2.00 for stainless steel

F_T = pump-type factor, F_T = 2.00 for single stage pump at flow rate 250-5,000 gpm

Electric Motors

A centrifugal pump is usually driven by an electric motor whose cost is added to the pump cost from Eq. 2.48. The size parameter for the motor is its power consumption, P_C, which is determined from the theoretical horsepower of the pump, P_T, its fractional efficiency, η_P, and the fractional efficiency of the electric motor, η_M, by the equation:

$$P_C = \frac{P_T}{\eta_P \eta_M} = \frac{P_B}{\eta_M} = \frac{QH\rho}{33{,}000\eta_P \eta_M} \tag{2.49}$$

Where, as previously, Q is the flow rate through the pump (gpm), H is the pump head of fluid flowing (ft), and P_B is the pump brake horsepower, with ρ equal to the liquid density (lbs/gal).

Which η_P as a function of volumetric flow rate and η_M as a function of pump brake horsepower can be written as;

$$\eta_P = -0.316 + 0.24015(\ln Q) - 0.01199(\ln Q)^2 \tag{2.50}$$

for Q in the range of 50 to 5,000 gpm;

$$\eta_M = 0.80 + 0.0319(\ln P_B) - 0.00182(\ln P_B)^2 \tag{2.51}$$

for P_B in the range of 1 to 1,500 Hp.

The f.o.b. purchase cost of an electric motor can be calculated by following equation;

$$C_B = \exp\left\{\begin{array}{l} 5.8259 + 0.1341\left[\ln\left(P_C\right)\right] + 0.05325\left[\ln\left(P_C\right)\right]^2 \\ + 0.028628\left[\ln\left(P_C\right)\right]^3 - 0.0035549\left[\ln\left(P_C\right)\right]^4 \end{array}\right\} \tag{2.52}$$

which applies over the range of 1 to 700 Hp. For other motor speeds and type enclosures, the f.o.b. purchase cost is given by:

$$C_P = F_T C_B \tag{2.53}$$

where F_T is a motor-type factor applicable within a range of electric motor power consumption, P_C, from 1 to 1,500 Hp.

2.4.3 Blower

Most blowers, with compression ratios up to 2, are of the multistage centrifugal (often called turboblower) type or the rotary positive-displacement type. Axial-flow units can also be used, but must be multistaged. The centrifugal blower achieves highest discharge pressures, with axial flow provides the highest flow rates. Typical mechanical efficiencies, η_B, are 70–80% for centrifugal blowers and 50–70% for straight-lobe blowers. The centrifugal blower delivers a smooth flow rate, but the straight-lobe units deliver a somewhat pulsing flow. Both types of blowers have found a wide range of applications, but centrifugal blowers are more common in chemical processing plants and are widely used to supply air to strippers, dryers, and combustion devices. The equipment size factor for a blower is the brake horsepower, P_B, which is computed from the inlet volumetric flow rate, Q_I, in cubic feet per minute and pressures in lb_f/in^2 at the inlet, P_I, and outlet, P_O, by the following equation, which assumes the ideal gas law and a constant specific heat ratio, k:

$$P_B = 0.00436\left(\frac{k}{k-1}\right)\frac{Q_I P_I}{\eta_B}\left[\left(\frac{P_o}{P_I}\right)^{\frac{k-1}{k}} - 1\right] \tag{2.54}$$

For other materials of construction, the f.o.b. purchase cost is given by:

$$C_P = F_M C_B \tag{2.55}$$

where F_M for stainless steel is 2.5.

The base blower purchase cost given by the following equation, which P_C in horsepower:

Centrifugal (turbo) blower (valid from P_C = 5 t0 1,000 Hp):

$$C_B = \exp\left\{6.8929 + 0.7900\left[\ln\left(P_C\right)\right]\right\} \tag{2.56}$$

2.4.4 Installation Cost

Installation cost of equipment can be calculated by multiplier for installation costs of process equipment multiply by total equipment cost. The multipliers for installation costs of process equipment as shown in appendix TableA.2. The equation for the installation cost calculation as follows;

$$\text{Installation cost} = (\text{total equipment cost}) \times (\text{multiplier}) \tag{2.58}$$

CHAPTER III

RESEARCH METHODOLOGY

A large amount of sodium hydroxide solution is used in PTA plant, which it causes of production cost increases. So, the reduction of sodium hydroxide solution is the way to production cost saving. The steps of study as follows:

3.1 Wastewater Treatment Unit Study

3.1.1 Operating Cost of Wastewater Treatment Unit

3.1.2 Wastewater Treatment Unit

3.1.3 Anaerobic System

3.2 Dissolved Carbon Dioxide

3.2.1 Carbonate System

3.2.2 Dissolved Carbon Dioxide Measurement

Dissolved carbon dioxide can be measured by alkalinity measurement. Then use the relationship between carbonate system and alkalinity to find dissolved carbon dioxide [descripted in Section 2.2 – 2.3].

3.2.3 Dissolved carbon dioxide removal

1) Dissolved Carbon Dioxide Removal Methods

2) Experiment

The simple CO_2 stripping can be done by Air was blown into 500 mL anaerobic digestion effluent in a 1,000 mL beaker at a flow rate of 1 L/min. The pH variations with time were monitored every 30 seconds until 10 minutes then monitored every 1 minute until 60 minutes. This experiment to confirm that it can be pH increased.

3.3 Carbon Dioxide Stripper Design

3.3.1 Stripper Towers Selection

Packed towers, tray towers, and spray towers are generally used in industrial stripping operations. Although packed towers are used more often in air pollution control, both have their special area of usefulness. Selection guideline for stripping tower given in Table 2.4.

3.3.2 Packed Towers Design

A generalized analysis procedure is as follows:

1) Packing Materials Selection

Specific packing selected for an industrial application depends on the nature of the contaminants, geometric mode of contact, size of the absorber, and scrubbing objectives. Criteria for packing material selection given on Section 2.3.2.

2) Pressure Drop (ΔP) Selection

Select acceptable pressure drop (typically a function of the packing material).

3) Packed Tower Diameter

a) Determine the cross-sectional area of the tower, based on the allowable pressure drop.

$$A = \frac{Q\rho_L}{L'} \tag{3.1}$$

Q = liquid volumetric flow rate, m^3/s

L'= superficial mass velocity of the liquid, kg/m^2s, which can be determined from the data presented in Fig. 2.23 in Chapter II.

b) Determine the diameter of the tower using the follow equation;

$$D = \sqrt{\frac{4}{\pi}A} \tag{3.2}$$

4) Packed Height

The height of a packed tower refers to the depth of packing material needed to accomplish the required removal efficiency. The more difficult the separation, the larger the packing height required. Determining the proper height of packing important since it affects both the rate and the efficiency of stripping. A number of theoretical equations are used to predict the required packing height. These equations are based on diffusion principles. The total packing height (Z) can be calculated by following equation;

$$Z = \mathrm{HTU} \times \mathrm{NTU} \tag{3.3}$$

Where HTU = height of a transfer unit

NTU = number of transfer units

a) Height of a Transfer Unit

The film mass-transfer coefficient k_G and k_L and effective wetted area of packing (a_W) are calculated using the correlation developed by Onda et al., which can be used to HTU.

$$a_W = a_t \left\{ 1 - \exp\left[-1.45 \left(\frac{\sigma_C}{\sigma_L}\right)^{0.75} \left(\frac{L}{a_t \mu_L}\right)^{0.1} \left(\frac{L^2 a_t}{\rho_L^2 g}\right)^{-0.05} \left(\frac{L^2}{\rho_L \sigma_L a_t}\right)^{0.2} \right] \right\} \tag{3.4}$$

k_G and k_L are calculated from the following correlations:

$$k_G = 5.23 \left(a_t D_G\right) \left(\frac{G}{a_t \mu_G}\right)^{0.7} \left(\frac{\mu_G}{\rho_G D_G}\right)^{1/3} \left(a_t d_p\right)^{-2} \tag{3.5}$$

$$k_L = 0.0051 \left(\frac{L}{a_W \mu_L}\right)^{2/3} \left(\frac{\mu_L}{\rho_L D_L}\right)^{-0.5} \left(a_t d_p\right)^{0.4} \left(\frac{\rho_L}{\mu_L g}\right)^{1/3} \tag{3.6}$$

Where k_G = gas film mass transfer coefficient, kmol/ m^2 atm

k_L = liquid film mass transfer coefficient, kmol/ m^2 atm

a_t = total packing surface area per packed-bed volume, m^2/m^3

d_p = packing size, m

L,G = superficial mass velocity of liquid and gas phase, respectively, kg/m s

μ_G, μ_L = viscosity in the gas and the liquid phase, respectively, kg/m s

ρ_G, ρ_L = density in the gas and liquid phase, respectively, kg/m^3

σ_L = water surface tension, N/m

σ_C = packing material surface tension, 61 dyne/cm for ceramic packing 75 dyne/cm for steel packing, and 33 dyne/cm for plastic packing

D_G, D_L = diffusivity in the gas and liquid phase, respectively, m^2 /s

g = acceleration due to Earth's gravity, 9.81 m/s^2

The Volumetric Mass Transfer Coefficient (K_La)

The volumetric mass transfer coefficient can be calculated by following equation:

$$\frac{1}{K_L a_W} = \frac{1}{k_L a_W} + \frac{1}{Hk_G a_W} \tag{3.7}$$

The height of transfer unit for the liquid phase is defined as:

$$HTU = \frac{Q}{K_L aA} \tag{3.8}$$

b) Number of Transfer Units

The number of transfer units is a function of the constituent and solubility. The number of transfer units is defined as:

$$NTU = \left[\frac{C_0 - C_e}{C_0 - C_e - C_0^*}\right] \ln\left[\frac{C_0 - C_e^*}{C_e}\right] \tag{3.9}$$

Where C_0 = concentration of constituent at inlet

C_e = concentration of constituent at outlet

C_e^* = concentration of constituent in liquid that is in equilibrium with the gas leaving the tower

3.4 Investment

3.4.1 Investment Cost

The purchase cost of processing equipment is generally obtained from charts, equations, or quotes from vendors. However, costs are not static. The construction costs can be evaluated from;

1) Textbook [Seider et al., 2009]
2) Aspen Plus simulation
3) Vendor

In this thesis, the investment cost can be determined from Aspen Plus simulation program.

3.4.2 Operating Cost

The operating costs of stripping unit include electricity from pump and blower, operating labor, and maintenance costs. In this thesis, only electricity is considered for the operating cost.

3.4.3 Economic Cost Benefit Analysis

An internal rate of return (IRR) in one of the considerations of economic cost benefit. The IRR can be calculated from the SMPC's IRR Calculation Template for 2018.

CHAPTER IV

RESULTS AND DISCUUSION

4.1 Dissolved Carbon Dioxide

Carbon dioxide dissolved in water to form unionized aqueous species and ionized aqueous species. Unionized aqueous species are carbon dioxide in an aqueous phase ($CO_2\,(aq)$) and carbonic acid (H_2CO_3) are often summed into a single hypothetical species ($H_2CO_3^*$) also called total dissolved carbon dioxide because of the low hydration of $CO_2\,(aq)$, meaning H_2CO_3* effectively represents the concentration of dissolved CO_2. Ionized aqueous species are bicarbonate and carbonate, which related to alkalinity in water. From Eq. 2.5 through Eq. 2.11 can be derived given following equation;

$$C_T = \frac{[\mathrm{Alk.}] - \frac{K_W}{[H^+]} + [H^+]}{\alpha_1 + 2\alpha_2} \tag{4.1}$$

From Eq. 4.1, the total inorganic carbon (C_T) depends on alkalinity concentration, equilibrium constants, and pH in the wastewater. Equilibrium constants depend on temperature. To determine the total inorganic carbon, the alkalinity, pH and the temperature will be measured. The average alkalinity, pH, and temperature of the sampling from anaerobic circulation stream are 1,743 mg/L as $CaCO_3$, 6.99, and 30.4 respectively (shown in Appendix A). Substituting the variables into Eq. 4.1 yields C_T equal 0.0424 eq/L (shown in Appendix B.1). Combine Eq. 2.5a and Eq. 2.6, substitute the variables to yields $[H_2CO_3^*]$ equal 168.37 mg/L (shown in Appendix B.1). If the air used in CO_2 stripping, this method will be had limitation. The limitation of the total dissolved carbon dioxide removal is controlled by gas-liquid equilibrium relationship. Air contains a mole fraction of about 0.00032 mole of carbon dioxide per mole of air, which is equivalent to a partial pressure of 0.00032 atm. According to Henry's law, water in contact with the atmosphere at 25 °C has an equilibrium concentration of about 0.58 mg/L (0.58 ppm). Therefore, the wastewater after CO_2 removed must have the total dissolved carbon dioxide.

4.2 Cost Saving

The total dissolved carbon dioxide concentration from the wastewater can be removed is 167.8 mg/L or 3.82×10^{-3} mol/L. From the chemical reaction between carbon dioxide and sodium hydroxide, suggested that 1 mole of CO_2 reacted with 1 mole of NaOH. Therefore the total NaOH used to react with carbon dioxide is equal 3.82×10^{-3} mol/L or 0.1528 g/L. The 50% NaOH used to wastewater adjustment, so the total NaOH used is 0.3056 g/L.

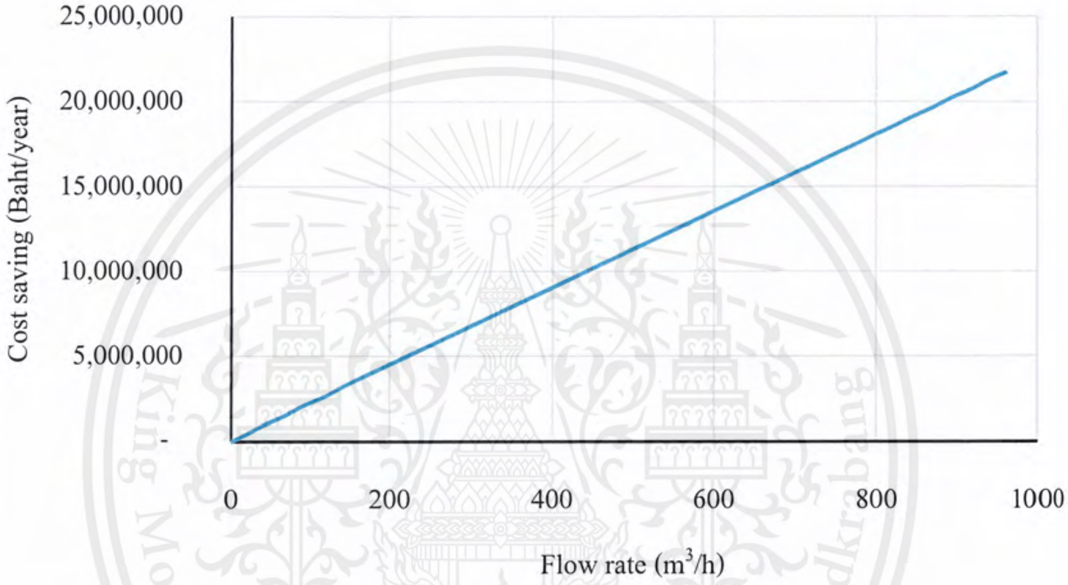


Figure 4.1 Cost Saving from Reduction of NaOH with Various Stripped Wastewater Flow Rates

Fig. 4.1 shows that the cost saving from the reduction of NaOH will be increased with the increase of the stripped wastewater flow rate. At the maximum flow rate gives the highest cost saving, which is 21,603,172.66 Baht per year.

4.3 Stripper Unit and Equipment Selection

Stripper unit for CO_2 removal consists of a tower, a pump, and a blower. Types of the towers are tray tower, packed tower, and spray tower. They are commonly custom manufactured to gas stripping. The packed tower is selected for CO_2 removal in this project. Because the packed tower is superior to other types. Criteria for selecting a tower include pressure drop, liquid flow rate, fouling, efficiency, capital cost, and annual operating and maintenance costs. For the pressure drop, high gas pressure drop causes high power consumption of the blower, in contrast, high liquid pressure drop causes high power consumption of the pump. The packed tower has high gas pressure drop but lower than tray tower. For the liquid flow rate, packed and spray towers can be handled at high liquid flow rate more than tray tower. Therefore, tray tower can be handled at same high liquid flow rate when increases the size of the tower, and increases the capital cost of the tower. For the fouling, if the tray tower and the packed tower internals are exposed to light and the wastewater contains sufficient organic matter to sustain microbial growth, tray and packed towers may be subjected by biological fouling due to bio-growth or algae formation. The spray has a possibility of nozzle fouling, resulting in increasing liquid pressure drop. The fouling can be prevented by filtration of an influent stream before CO_2 removal. For the efficiency, the packed tower gives high efficiency due to high mass transfer than tray tower and spray tower. Flooding, priming, coning, weeping, and dumping may occur tower that is causes of low efficiency in tray tower. The spray tower gives low efficiency due to low mass transfer and low residence time, which the efficiency can be increased by internals packing usage and height of tower increasing. For the capital cost, tray and packed towers are expensive than the spray tower because the towers have internal packing. The capital cost of spray will be increased if the internal packing is used. For the annual operating and maintenance costs, the spray tower is expensive than tray and packed towers because of the high-power consumption of pump due to high liquid pressure drop.

The packed tower consists of packing material, which is Pall rings. Commonly packing materials (Raschig ring, Berl saddle, metal Intalox, Intalox saddle, Tellerette, and Pall ring) are compared. Pall rings provide a large specific area more than other types. When the material for packing material structures are considered, ceramic is cheaper than other type but heavier than other types, plastic is priced close to the ceramic but lighter than either ceramic or metal packings, metal is expensive than either ceramic or plastic packing. Sizes of the packing material are the major consideration because it related to the cost of packing material and liquid distribution. The recommended size of column diameter than 0.9 meters is 50 millimeters. Therefore, 50 mm plastic Pall ring is selected for the packed tower.

Centrifugal pump is selected for the stripper unit because this pump is available through a vast range of sizes, in capacities from 0.5 m^3/h to 2×10^4 m^3/h, and for discharge heads (pressures) from a few meters to approximately 48 MPa. The primary advantages of a centrifugal pump are simplicity, low first cost, uniform (non-pulsating) flow, small floor space, low maintenance expense, quiet operation, and adaptability for use with a motor or a turbine drive. Centrifugal blower is selected for the stripper unit because this blower discharges the gas at a very high velocity.

4.4 Stripper Unit and Equipment Design

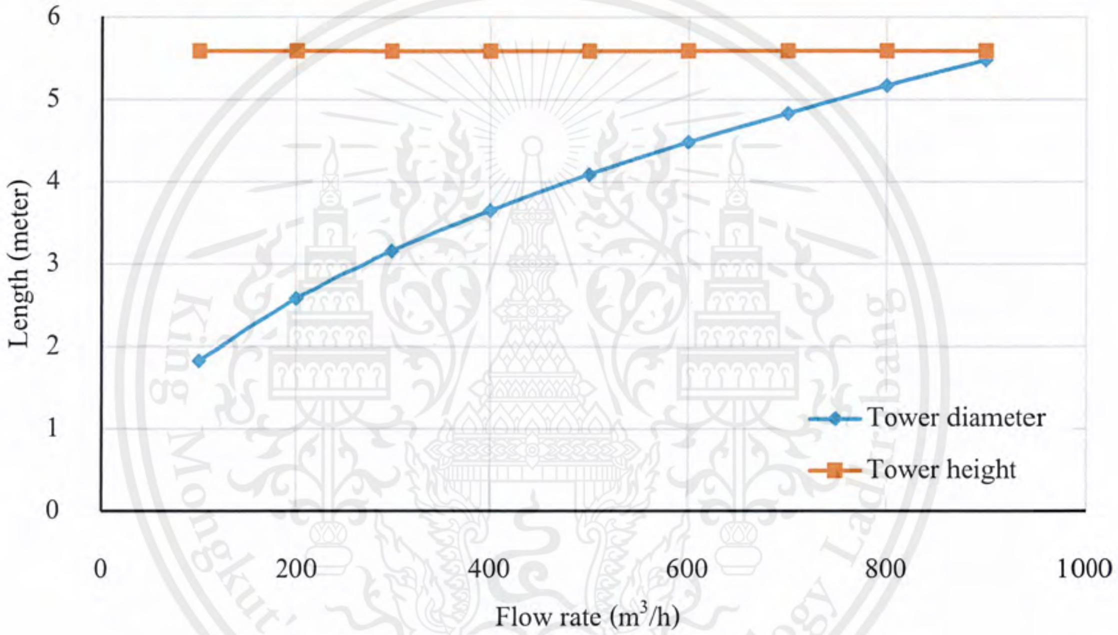


Figure 4.2 Tower Height and Diameter with Various Stripped Wastewater Flow Rates

Fig.4.2 shows that the tower diameter will be increased with the increasing of stripped wastewater flow rate because to maintain the residence time, the tower diameter will be increased. In contrast, the total height of packed tower stripper independent with the stripped wastewater flow rate but depends on the concentration of the contaminant in influent stream.

The height of stripper tower packing material depends on two factors are the height of a transfer unit and the number of transfer units. The height of a transfer unit (HTU) is a measure of the separation efficiency of the particular packings for a particular separation process. As such, it incorporates the mass transfer coefficient. The more efficient the mass transfer (i.e. larger mass transfer coefficient), the smaller the value of HTU. The values of HTU can be estimated from empirical. From Eq. 2.36;

$$\mathrm{HTU} = \frac{L}{K_L aA}$$

Suggested that, the HTU depends on the volumetric mass transfer coefficient (K_La).

The number of transfer units (NTU) is the one factor of the height of stripper tower packing material. NTU required is a measure of the difficulty of the separation. A single transfer unit gives the change of composition of one of the phases equal to the average driving force producing the change. The NTU is similar to the number of theoretical trays required for tray tower. Hence, a larger number of transfer units will be required for a very high purity of the effluent. That shown in Eq. 2.37;

$$\mathrm{NTU} = \left[\frac{C_0 - C_e}{C_0 - C_e - C_0^*}\right] \ln\left[\frac{C_0 - C_e^*}{C_e}\right]$$

Therefore, the height of stripper tower packing material for CO_2 removal in the wastewater is equal to 3.95 meters.

4.5 Investment

The investment cost of the project costs of equipment cost and installation cost. The equipment cost depends on the size of the equipment, which the larger equipment has more expensive. The installation cost depends on the size of the equipment, which the larger equipment has more expensive. The investment cost as shown in Fig. 4.3.

The Internal rate of return (IRR) refers to capital budgeting measuring the profitability of potential investments as investors choose to make decisions, it is considered that the project is a worthy investment, which SCG criteria for IRR more than 20%. Fig. 4.3 shows that the IRR depends on the stripped wastewater flow rate. The increasing of the stripped wastewater rate, IRR will be increased because of the increasing of the stripped wastewater flow rate will give high cost saving from the reduction of NaOH usage. At the stripped wastewater flow rate lower than 180 m^3/h, %IRR lower than 20% in this range is not worth investing. In contrast, the stripped wastewater rate higher than 180 m^3/h, %IRR higher than 20% in this range is worth investing follows SCG criteria.

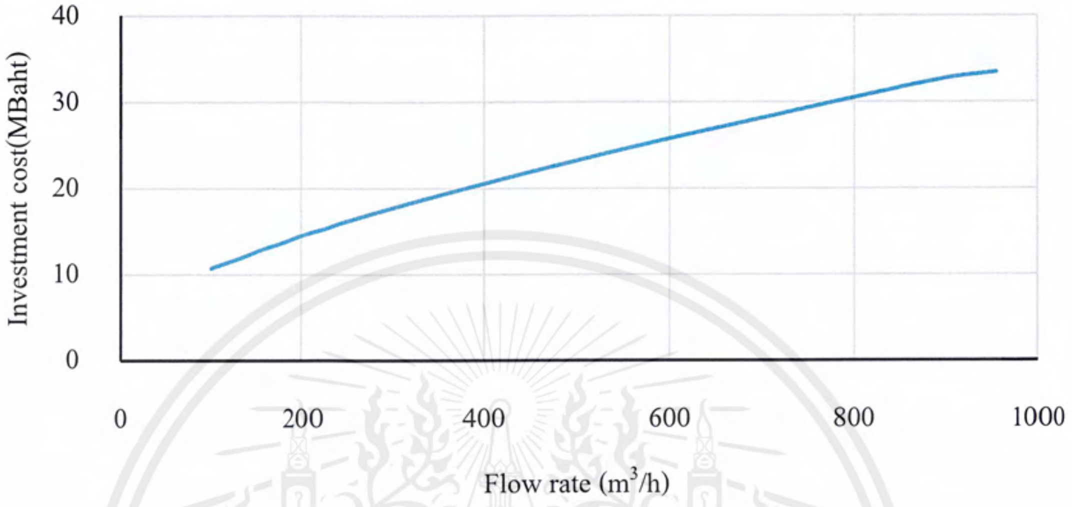


Figure 4.3 Investment Cost with Various Stripped Wastewater Flow Rates

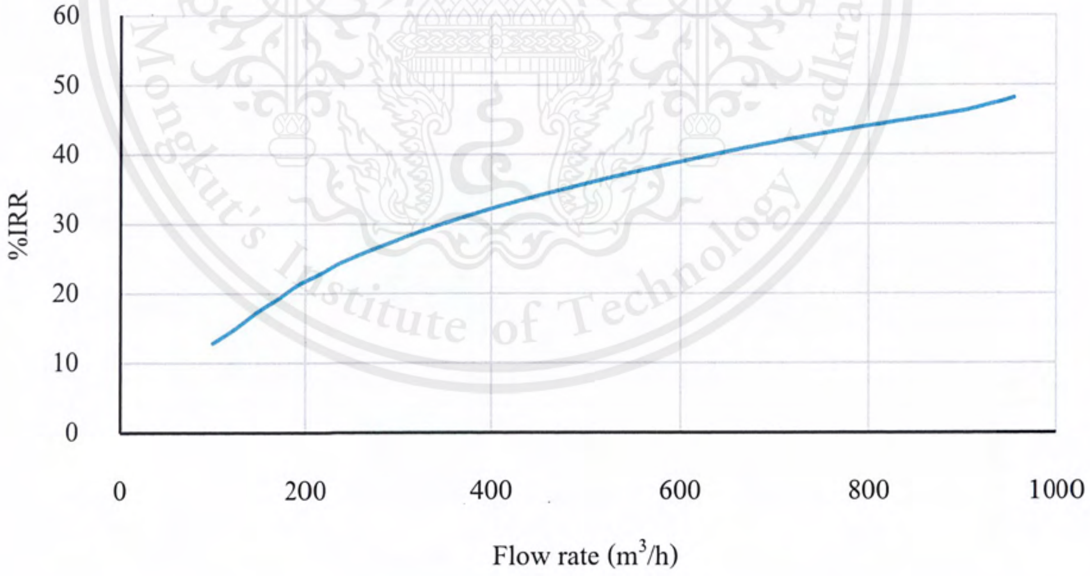


Figure 4.4 IRR Percentage with Various Stripped Wastewater Flow Rates

CHAPTER V

CONCLUSIONS

5.1 Conclusions

The cost of NaOH solution per year is rather high. The pH of the circulating wastewater stream from anaerobic reactor can be increased by CO_2 stripping. Because of the anaerobic reactor generates CO_2 which causes the acidic wastewater. The CO_2 stripping unit consists of a packed tower, a centrifugal pump, and a centrifugal blower. The packed tower is filled by 50 mm plastic Pall rings with 3.95 m of packing height. The total height of the packed tower is 5.95 m. The packed tower operates at a gas flow rate of 50 times of the stripped wastewater flow rate. The packed tower can remove dissolved CO_2 from 168.37 mg/L to 0.58 mg/L, resulting in the pH will increase from 6.99 to 9.45. The packed tower diameter is equal to 5.65 m for stripped wastewater flow rates at 955 m^3/h. The project gives 46.28% IRR, which a worthy investment. Because the minimum acceptable rate of return (MARR) is 20% as follows the company criteria.

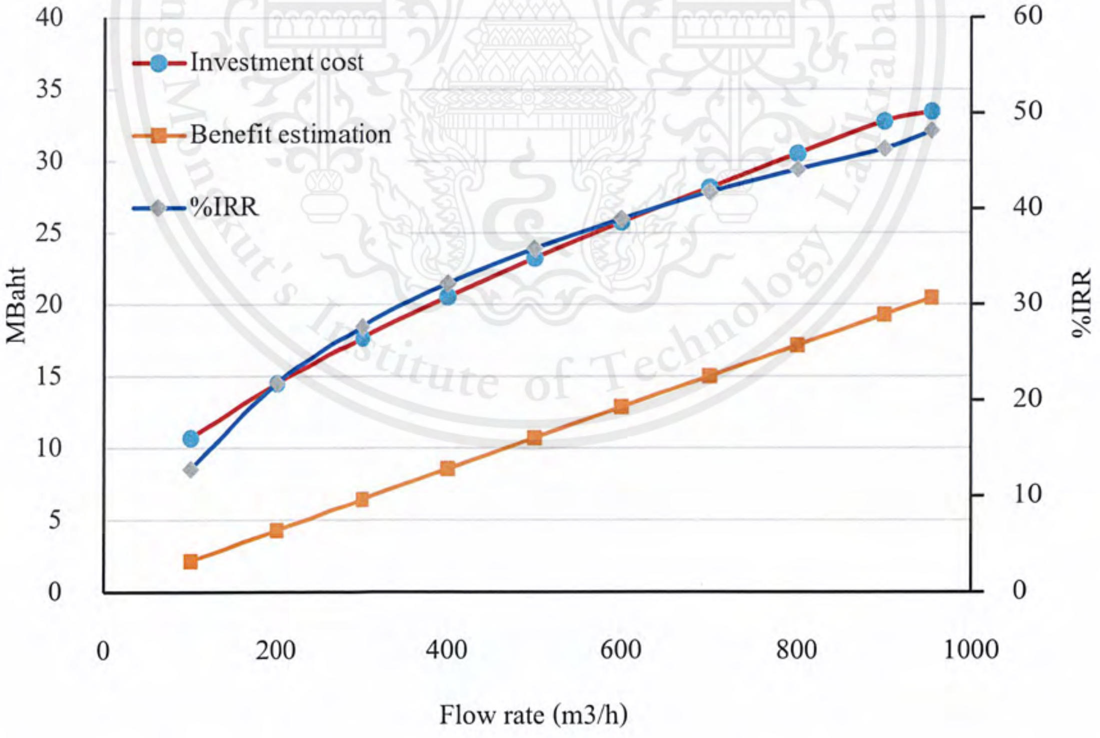


Figure 5.1 Cost Saving, IRR and Investment Cost with Various Stripped Wastewater Flow Rates

5.2 Suggestions

1) Dissolved carbon dioxide concentration can be completely removed by using pure nitrogen gas, which can be NaOH reduction increased. If pure nitrogen gas used, the operating cost will be increased.

2) If the air used for carbon dioxide stripping, dissolved oxygen must be concerned because it affects to bacteria in the anaerobic reactor, which oxidation reduction potential (ORP) is used indicator, the negative of the value is preferred for anaerobic.

3) Cost saving will be increased if the pH of the wastewater increased more than 8.3. Because at this range bicarbonate ion will dissociate to carbonate ion, then carbonate ion will be reacted with NaOH. Therefore, the cost saving from the reduction of NaOH will increase.

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APPENDIX

APPENDIX A

DATA

Table A.1 Sampling Results of Anaerobic Circulating Stream

No.	Sampling items				
	Temperature (^oC)	**pH**	**Alkalinity (mg/L as $CaCO_3$)**	**COD (mg/L)**	**TSS (mg/L)**
1	28.8	6.93	923	320	14,040
2	32.4	7.06	2,045	510	16,340
3	31.2	6.88	2,053	590	16,380
4	24.7	7.08	1,890	530	13,680
5	31.7	7.05	1,896	600	17,020
6	29.6	6.99	1,769	640	15,660
7	29.7	6.96	1,747	450	13,060
8	31.1	6.95	1,645	570	12,620
9	34	7	1,715	540	12,220
Avg.	30.4	6.99	1,743	528	14,558

Table A.2 Multipliers for Installed Costs of Process Equipment

[adapted from Couper, J.R. et al., 2012]

Equipment	Multiplier	Equipment	Multiplier
Agitators, CS	1.3	Plate, SS	1.5
SS	1.2	Plate, CS	1.7
Air heater, all types	1.5	Shell and tube, S/SS	1.9
Beaters	1.4	Shell and tube, CS/Monal	1.8
Blenders	1.3	Instruments, all types	2.5
Blowers	1.4	Pumps, centrifugal, CS	2.8
Boilers	1.5	Centrifugal, SS	2.0
Centrifuges, CS	1.3	All others, SS	1.4
SS	1.2	All others, CS	1.6
Chimneys and stacks	1.2	Reactor kettles, CS	1.9
Column, distillation, CS	3.0	Kettles, glass lined	2.1
Distillation, SS	2.1	Kettles, CS	1.9
Compressors, motor driven	1.3	Reactors, multitubular, SS	1.6
Steam or gas driven	1.5	Multitubular, Copper	1.8
Conveyors and elevators	1.4	Multitubular, CS	2.2
Crushers, classifiers and mills	1.3	Sum of equipment costs, SS	1.8
Crystallizers	1.9	Sum of equipment costs, CS	2.0
Cyclones	1.4	Tanks, process, SS	1.8
Dryers, spray and air	1.6	Tanks, process, aluminium	2.0
Other	1.4	Storage, SS	1.5
Ejectors	1.7	Storage, CS	2.3
Fans	1.4	Field erected, SS	1.2
Gas holders	1.3	Field erected, CS	1.4
Heat exchangers, air cooled, CS	2.5	Turbines	1.5
Coil in shell, SS	1.7	Vessels, pressure, SS	1.7

Note: CS= carbon steel, SS= stainless steel

APPENDIX B

EXAMMPLE OF CALCULATION

B.1 Dissolved Carbon Dioxide Concentration

Dissolved carbon dioxide concentration can be calculated from the relationship between alkalinity and carbonate system in water. The calculation can be expressed as following:

1) Data Input

Temperature	30.4	^{o}C
pH	6.99	
K_1@ 30.4 ^{o}C	$4.69\text{x}10^{-7}$	
K_2@ 30.4 ^{o}C	$5.17\text{x}10^{-11}$	
K_W@ 30.4 ^{o}C	$1.55\text{x}10^{-14}$	
Alkalinity	1,743	mg/L as $CaCO_3$

2) Steps of Calculation

a) Total inorganic carbon

Substituting Eq. 2a and 2b in Eq. 2.11, then modify to simple form yields;

$$C_T = \frac{[\text{Alk.}] - \frac{K_W}{[H^+]} + [H^+]}{\alpha_1 + 2\alpha_2} \qquad \text{(B1.1)}$$

b) Ionization fraction (α_1 and α_2) calculation

To determine $\alpha 1$. From Eq. 2.7;

$$\alpha_1 = \left(\frac{[H^+]}{K_1} + 1 + \frac{K_2}{[H^+]} \right)^{-1}$$

Substitute variable in to equation;

$$\alpha_1 = \left(\frac{10^{-6.99}}{4.69 \times 10^{-7}} + 1 + \frac{5.17 \times 10^{-11}}{10^{-6.99}} \right)^{-1}$$

$$\alpha_1 = 0.8201$$

To determine α_2. From Eq. 2.8;

$$\alpha_2 = \left(\frac{[H^+]^2}{K_1K_2} + \frac{[H^+]}{K_2} + 1 \right)^{-1}$$

Substitute variable in to equation;

$$\alpha_2 = \left(\frac{\left(10^{-6.99}\right)^2}{\left(4.69\times 10^{-7}\right)\left(5.17\times 10^{-11}\right)} + \frac{10^{-6.99}}{5.17\times 10^{-11}} + 1 \right)^{-1}$$

$$\alpha_2 = 0.0005$$

c) Substituting of known variable in Eq. B1.1 yields;

$$C_T = \frac{\frac{1{,}743}{50{,}000} - \frac{1.55\times 10^{-14}}{10^{-6.99}} + 10^{-6.99}}{0.8201 + 2(0.0005)}$$

$$C_T = 0.0424 \text{ mg/L}$$

d) Determine the dissolved carbon dioxide concentration

Combine Eq. 2.5a and 2.6 yields;

$$[H_2CO_3^*] = [C_T]\cdot\left(1 + \frac{K_1}{[H^+]} + \frac{K_1K_2}{[H^+]^2} \right)^{-1} \qquad \text{(B1.2)}$$

Substituting of variable in Eq. B1.1 yields;

$$[H_2CO_3^*] = 0.0424\cdot\left(1 + \frac{4.69\times 10^{-7}}{10^{-6.99}} + \frac{\left(4.69\times 10^{-7}\right)\left(5.17\times 10^{-11}\right)}{\left(10^{-6.99}\right)^2} \right)^{-1}$$

$$[H_2CO_3^*] = 7.59\times 10^{-3} \text{ eq/L}$$

$$= 7.59 \times 10^{-3}\ \mathrm{eq/L} \times 22{,}000\ \mathrm{mg\ of\ CO_2/eq}$$

$$= 168.37\ \mathrm{mg/L}$$

B.2 Stripper Unit and Equipment Design

Stripper unit consists of tower, pump, and blower. The tower is packed tower, which used to strip carbon dioxide out of the wastewater. The packed tower consists of packing material, liquid distributor, and liquid redistributor. Design of stripper unit and equipment are following;

1) Packed Tower Design

a) Tower Diameter

Recommended gas to water volumetric ratio (G/L) for CO_2 stripping is 50 m^3/m^3 (มั่นสิน, 2537)

i) Calculate water to gas loading rate (L'/G')

$$\frac{G}{L} = 50\frac{m^3}{m^3}$$

$$\frac{L}{G} = 0.02\frac{m^3\ \mathrm{water}}{m^3\ \mathrm{air}}$$

$$\frac{L'}{G'} = \frac{L}{G} \times \frac{\rho_L}{\rho_G} = 0.02\frac{m^3\ \mathrm{water}}{m^3\ \mathrm{air}} \times \frac{m^3\ \mathrm{air}}{1.16\ \mathrm{kg}} \times \frac{995.4\ \mathrm{kg}}{m^3\ \mathrm{water}}$$

$$= 17.16\frac{\mathrm{kg}}{\mathrm{kg}}$$

ii) Determine the flow parameter from equation;

$$\frac{L'}{G'}\left(\frac{\rho_G}{\rho_L - \rho_G}\right)^{1/2} \approx \frac{L'}{G'}\left(\frac{\rho_G}{\rho_L}\right)^{1/2} = 17.16\ \mathrm{kg/kg} \times \left(\frac{1.16\ \mathrm{kg/m^3}}{995.4\ \mathrm{kg/m^3}}\right) = 0.5858$$

iii) For an abscissa value of 0.5858 and a pressure drop of 100 (N/m^2)/m, the ordinate value from Fig. 2.23 is 0.03.

iv) Using an ordinate value 0.03, determine the loading rate using Eq. 2.40;

$$G' = \left[\frac{(\text{value from y axis})(\rho_G)(\rho_L - \rho_G)}{(C_f)(\mu_L)^{0.1}} \right]^{1/2}$$

$$G' = \left[\frac{(0.03)(1.16)(995.4 - 1.16)}{(187)\left(7.87 \times 10^{-4}\right)^{0.1}} \right]^{1/2}$$

$$G' = 0.6149 \ \ \mathrm{kg/m^2 \cdot s}$$

So, $$L' = G' \times 17.16 = 10.55 \ \ \mathrm{kg/m^2 \cdot s}$$

v) Calculate the cross-sectional area of the tower

$$A = \frac{Q\rho_L}{L'}$$

Where Q = liquid flow rate, m^3/h

ρ_L = liquid density, kg/m^3

L’ = liquid loading rate, kg/m^{2}s

$$A = \frac{100 \ \ \mathrm{m^3/h} \times 995.4 \ \ \mathrm{kg/m^3}}{10.55 \ \ \mathrm{kg/m^2 \cdot s}} \times \frac{1 \ \ \mathrm{hr}}{3600 \ \ \mathrm{s}} = 2.62 \ \ \mathrm{m^2}$$

vi) Calculate the tower diameter

$$D = \sqrt{\frac{4}{\pi} \times A} = \sqrt{\frac{4}{\pi} \times 2.62} = 1.83 \ \ \mathrm{m}$$

b) Determine Tower Height

i) Calculate the specific surface area are available for mass transfer, a_W.

Assumption $a=a_W$, from Onda's correlation

$$a_W = a_t\left\{1-\exp\left[-1.45\left(\frac{\sigma_C}{\sigma_L}\right)^{0.75}\left(\frac{L'}{a_t\cdot\mu_L}\right)^{0.1}\left(\frac{L'^2\cdot a_t}{\rho_L^2\cdot g}\right)^{-0.05}\left(\frac{L'^2}{\rho_L\cdot\sigma_L\cdot a_t}\right)^{0.2}\right]\right\} \quad (B2.1)$$

Where a_t = packing surface area, m^2/m^3

σ_C = packing surface tension, N/m

σ_L = liquid surface tension, N/m

g = gravitational acceleration, m/s^2

μ_L = liquid viscosity, kg/m.s

$$a_W = 102\left\{1-\exp\left[\begin{array}{l}-1.45\left(\dfrac{0.033}{0.0711}\right)^{0.75}\times\left(\dfrac{10.55}{102\times 7.87\times 10^{-4}}\right)^{0.1}\times\left(\dfrac{10.55^2\times 102}{995.4^2\times 9.81}\right)^{-0.05}\\ \times\left(\dfrac{10.55^2}{995.4\times 0.0711\times 102}\right)^{0.2}\end{array}\right]\right\}$$

$$a_W = 56.54 \ \ m^2/m^3$$

ii) Calculate the liquid phase mass transfer coefficient, k_L

$$k_L = 0.0051\left(\frac{L'}{a_W\cdot\mu_L}\right)^{2/3}\left(\frac{\mu_L}{\rho_L\cdot D_L}\right)^{-0.5}\left(a_t\cdot d_p\right)^{0.4}\left(\frac{\rho_L}{\mu_L\cdot g}\right)^{-1/3} \quad (B2.2)$$

Where D_L = liquid diffusivity, $1.96\text{x}10^{-9}$ m^2/s [Summerfelt, 2000]

d_p = packing diameter, m

$$k_L = 0.0051\left(\frac{10.55}{56.54\times 7.87\times 10^{-4}}\right)^{2/3}\times\left(\frac{7.87\times 10^{-4}}{995.4\times 1.96\times 10^{-9}}\right)^{-0.5}$$

$$\times(102\times 0.05)^{0.4}\times\left(\frac{995.4}{7.87\times 10^{-4}\times 9.81}\right)^{-1/3}$$

$$k_L = 3.7\times 10^{-4}\ \text{m/s}$$

iii) Calculate the gas phase mass transfer coefficient, k_g

$$k_g = 5.23\left(a_t\cdot D_g\right)\left(\frac{G'}{a_t\cdot\mu_g}\right)^{0.7}\left(\frac{\mu_g}{\rho_g\cdot D_g}\right)^{1/3}\left(a_t\cdot d_p\right)^{-2} \qquad \text{(B2.3)}$$

Where D_g = gas diffusivity, $1.38\text{x}10^{-5}$ m^2/s [Summerfelt, 2000]

$$k_g = 5.23\times\left(102\times 1.38\times 10^{-5}\right)\times\left(\frac{0.6149}{102\times 1.86\times 10^{-5}}\right)\times(102\times 0.05)^{-2}$$

$$k_g = 1.7\times 10^{-2}\ \text{m/s}$$

iv) Calculate the volumetric mass transfer coefficient, K_La

From Eq.2.18

$$\frac{1}{K_La} = \frac{1}{k_La} + \frac{1}{Hk_Ga}$$

Substituting variables in above equation

$$\frac{1}{K_La} = \frac{1}{\left(3.7\times 10^{-4}\times 53.52\right)} + \frac{1}{\left(1.7\times 10^{-2}\times 56.54\times 0.8317\right)}$$

$$K_La = 1.53\times 10^{-2}\ \text{m/s}$$

v) Calculate tower height

- Calculate the concentration of CO_2 in liquid equilibrium gas existing.

$$C_0^* = \frac{P_T}{H} \times \frac{L}{G}\left(C_0 - C_e\right)$$

$$C_0^* = \frac{1}{1626} \times 50(168.37 - 0.58)$$

$$C_0^* = 2.06 \times 10^{-3}\ \mathrm{mg/L}$$

- Tower Height

$$Z = \mathrm{HTU} \times \mathrm{NTU}$$

From Eq. 2.36 and 2.37 are substituted in above equation yields;

$$Z = \frac{Q}{K_L aA}\left[\frac{C_0 - C_e}{C_0 - C_e - C_0^*}\right]\ln\left[\frac{C_0 - C_0^*}{C_e}\right]$$

$$Z = \frac{100}{1.53 \times 10^{-2} \times 2.62}\left[\frac{168.37 - 0.58}{168.37 - 0.58 - 2.06 \times 10^{-3}}\right]\ln\left[\frac{168.37 - 2.06 \times 10^{-3}}{0.58}\right]$$

$$Z = 3.95\ \mathrm{m}$$

The total height of the packed tower stripper is computed by packing height and an additional 2 meters is added for top and bottom of the packed tower stripper.

B.3 Cost Saving

$$CO_2(aq) + H_2O(l) \longrightarrow H_2CO_3(aq) \tag{1}$$

$$H_2CO_3(aq) + NaOH(aq) \longrightarrow NaHCO_3(aq) + H_2O(l) \tag{2}$$

$$CO_2(aq) + NaOH(aq) \longrightarrow NaHCO_3(aq) \tag{1+2}$$

From the chemical reaction above, suggested that 1 mole of carbon dioxide reacted with 1 mole of sodium hydroxide to yield 1 mole of sodium bicarbonate. The assumption of the reaction is sodium bicarbonate not dissociate to carbonate because the liquid solution has a pH lower than 8.3. In the stripper tower, the volumetric flow rate of wastewater can be operated equal to 100 m3/h and CO_2 can be stripped out of wastewater, which that was $3.82\text{x}10^{-3}$ mol/L of CO_2. So, it reacted with $3.82\text{x}10^{-3}$ mol/L of NaOH or 0.1528 g/L. In the wastewater treatment unit, 50% (w/w) NaOH is used, which the price in 2018 is 8.37 baht per kilogram. The total 50% (w/w) NaOH used in the reaction with CO_2 is equal to 0.3056 g/L. The wastewater treatment unit is operated throughout 365 days/year. So, the cost saving from reduction of NaOH is the following;

$$\text{Cost saving} = \frac{0.3056\ \text{g}}{\text{L}} \times \frac{1\ \text{kg}}{1000\ \text{g}} \times \frac{100\ \text{m}^3}{\text{h}} \times \frac{1000\ \text{L}}{1\text{m}^3} \times \frac{24\ \text{h}}{\text{day}} \times \frac{365\ \text{day}}{\text{year}} \times \frac{8.45\ \text{Baht}}{\text{kg}}$$

$$\text{Cost saving} = 2{,}262{,}112\ \ \text{Baht/year@flow rate}\ 100\ \text{m}^3/\text{h}$$

B.4 Operating Power Requirements [Edzwald, J., 2011]

Pump Power Requirements

Pump power requirements can be calculated from following equation;

$$\text{Power} = \frac{\rho_L \times Q \times H \times g}{1000 \times \text{Eff}} \tag{B4.1}$$

Where ρ_L = liquid density, kg/m^3

Q = volumetric flow rate, m^3/h

H = vertical distance from the pump to the liquid distributor at the top of the tower, m

g = acceleration due to Earth's gravity, m/s^2

Eff= pump efficiency

Substituting variable into above equation yields;

$$\text{Power} = \frac{995.4 \times 0.028 \times 5.95 \times 9.81}{1000 \times 0.85}$$

$$\text{Power} = 1.9 \ \text{kW}$$

Blower Power Requirements

The term P_{in} refers to the pressure at the bottom of the tower, which is the inlet air stream. P_{in} is calculated from;

$$P_{in} = P_{ambient} + \left[\left(\frac{\Delta P}{L}\right) \times L\right] + \Delta P_{losses} \tag{B4.2}$$

Where $P_{ambient}$ = pressure at atmosphere equal 101,325 N/m^2

$\Delta P/L$ = pressure drop of the packed tower, N/m^2/m

ΔP_{losses} is determined by [Hand et al., 1986];

$$\Delta P_{losses} = \left(\frac{Q}{A}\right)^2 \times k_p \tag{B4.3}$$

Where Q = volumetric air flow rate, m^3/s

A = tower cross-sectional area, m^2

k_p = constant equal to 275 NS^S/m^4

Substituting variables into Eq. B4.3 yields;

$$\Delta P_{losses} = \left(\frac{1.39}{1.31}\right)^2 \times 275$$

$$\Delta P_{losses} = 76.76 \ \ N/m^2$$

Then, substituting variables into Eq. B4.2 yields;

$$P_{in} = 101325 + \left[(100) \times 5.95\right] + 76.76$$

$$P_{in} = 101996.8 \ \ N/m^2$$

Therefore, blower power requirements can be calculated from following equation;

$$\text{Power} = \left(\frac{G_m \times R \times T}{1000 \times n_a \times \text{Eff}}\right)\left[\left(\frac{P_{in}}{P_{out}}\right)^{n_a} - 1\right] \quad \text{(B4.4)}$$

Where G_m = mass flow rate of air, kg/s

R = universal gas constant, 286.7

T = air temperature, K

n_a = constant used in determining blower brake power equal 0.283

Eff = blower efficiency

P_{in} = inlet air pressure in the packed tower, N/m^2

P_{out} = outlet air pressure in the packed tower, N/m^2

Substituting variables into Eq. B4.4 yields;

$$\text{Power} = \left(\frac{1.61 \times 286.7 \times 303.25}{1000 \times 0.283 \times 0.4}\right)\left[\left(\frac{101{,}996.8}{101{,}325}\right)^{0.283} - 1\right]$$

$$\text{Power} = 2.32 \quad \text{kW}$$

B.5 Investment Cost Estimation

The investment cost of the project is the summation of each equipment cost and installation cost. The detail of investment cost for each equipment and installation cost are descripted in Section 2.4.

1) Packed Tower

i) Determine Weight of the Shell (W)

From Eq. 2.44

$$W = \pi\left(D_i + t_s\right)\left(L + 0.8D_i\right)t_s\rho$$

where Di = inside diameter of tower, ft

t_s = shell thickness, ft

L = total height of tower, ft

ρ = density of the carbon steel, which can be taken as 490 lb/ft^3.

Substituting variation into above equation yields;

$$W = \pi(6 + 0.042)\left(22.796 + 0.8(6)\right) \times 6 \times 490$$

$$W = 10{,}693.5 \quad \text{lb}$$

ii) Determine the Purchase Cost for Vertical Tower (C_V)

Substitute the weight of the shell (W) into Eq. 2.43 yields;

$$C_V = \exp\left\{7.2756 + 0.18255\left[\ln\left(10{,}693.5\right)\right] + 0.02297\left[\ln\left(10{,}693.5\right)\right]^2\right\}$$

$$C_V = 56{,}738.93 \quad \text{US\$}$$

iii) Calculate Platform and Ladders Cost (C_{PL})

From Eq. 2.46;

$$C_{PL} = 361.8\left(D_i\right)^{0.73960}\left(L\right)^{0.70684}$$

Substituting variables into above equation yields;

$$C_{PL} = 361.8(6)^{0.73960}(22.796)^{0.70684}$$

$$C_{PL} = 12{,}413.85 \quad \text{US\$}$$

iv) Calculate Packing Material Cost

- Volume of Packing Material

$$V_P = \frac{\pi}{4}\left(D_i\right)^2(Z)$$

$$V_P = \frac{\pi}{4}(6)^2(12.956)$$

$$V_P = 366.43 \quad \text{ft}^3$$

- Installed cost of dumped packing

2-in. polypropylene Pall rings are used in the packed column. From Table 2.6, installed cost of the dumped packing is 21 $\$/\text{ft}^3$.

v) Total purchase Cost of the Packed Tower

From Eq. 2.42;

$$C_P = F_M C_V + C_{PL} + V_p C_{PK} + C_{DR}$$

Substituting variables into above equation yields;

$$C_P = (1 \times 56{,}738.93) + 12{,}413.85 + (366.43 \times 21)$$

$$C_P = 76{,}847.8 \quad \text{US\$}$$

In currently, Total purchase cost of the packed tower from Eq. 2.41;

$$\text{Cost} = \text{Base Cost}\left(\frac{I}{I_{\text{base}}}\right)$$

Substituting variables into above equation;

$$\text{Cost} = 76{,}847.8\left(\frac{569.8}{500}\right)$$

$$\text{Cost} = 87{,}575.75 \quad \text{US\$}$$

Note: not include liquid distributor and redistributor cost

2) Pump

a) Sizing Factor (S)

From Eq. 2.47;

$$S = Q(H)^{0.5}$$

Where Q = flow rate through the pump (gpm)

H = pump head (ft), pressure rise/liquid density

Substituting variables into above equation yields;

$$S = 440 \times (12.96)^{0.5}$$

$$S = 1{,}943.78 \quad \text{ft}$$

b) Pump Purchase Cost

The pump purchase cost can be calculated from Eq. 2.48;

$$C_B = \exp\left\{9.7171 - 0.6019\left[\ln(S)\right] + 0.0519\left[\ln(S)\right]^2\right\}$$

Substituting variable into above equation yields;

$$C_B = \exp\left\{9.7171 - 0.6019\left[\ln(1{,}943.78)\right] + 0.0519\left[\ln(1{,}943.78)\right]^2\right\}$$

$$C_B = 3{,}412.78 \quad US\$$$

c) The f.o.b. Purchase Cost

The f.o.b. purchase cost can be calculated from Eq. 2.49;

$$C_P = F_T F_M C_B$$

Where F_M = material factor, F_M = 1.35 for cast steel

F_T = pump-type factor, F_T = 2.00 for single stage pump at flow rate 250-5,000 gpm

Substituting variables into an equation yields;

$$C_P = (1.35)(2)(3{,}412.78)$$

$$C_P = 9{,}214.5 \text{ US\$}$$

d) In currently, the f.o.b. purchase cost can be calculated from Eq. 2.41;

$$\text{Cost} = \text{Base Cost}\left(\frac{I}{I_{base}}\right)$$

Substituting variables into an equation yields;

$$\text{Cost} = 9{,}214.5\left(\frac{569.8}{500}\right)$$

$$\text{Cost} = 10{,}500.85 \quad US\$$$

3) Electric Motor

From the appendix B.4, power consumption of the pump is calculated. The power consumption is used for the purchase cost of electric motor determination. The purchase cost of electric motor can be calculated from Eq.2.53;

$$C_B = \exp\left\{\begin{matrix} 5.8259 + 0.1341\left[\ln\left(P_C\right)\right] + 0.05325\left[\ln\left(P_C\right)\right]^2 \\ + 0.028628\left[\ln\left(P_C\right)\right]^3 - 0.0035549\left[\ln\left(P_C\right)\right]^4 \end{matrix}\right\}$$

Substituting a power consumption variable into equation yields;

$$C_B = \exp\left\{\begin{matrix} 5.8259 + 0.1341\left[\ln(2.54)\right] + 0.05325\left[\ln(2.54)\right]^2 \\ + 0.028628\left[\ln(2.54)\right]^3 - 0.0035549\left[\ln(2.54)\right]^4 \end{matrix}\right\}$$

$$C_B = 410.83 \quad \text{US\$}$$

The f.o.b. purchase cost can be calculated from Eq. 2.54;

$$C_P = F_T C_B$$

Where $F_T = 1$ for open, drip-proof enclosure motor type, substituting yields;

$$C_P = 1 \times 410.83$$

$$C_P = 410.83$$

In currently, the f.o.b. purchase cost can be calculated from Eq. 2.41 yields;

$$\text{Cost} = 410.83\left(\frac{569.8}{500}\right)$$

$$\text{Cost} = 468.18 \quad \text{US\$}$$

4) Blower

From the appendix B. 4, power consumption of the blower is calculated. The power consumption is used for the purchase cost of the blower. The purchase of the blower can be calculated from Eq.2.57;

$$C_B = \exp\left\{6.8929 + 0.7900\left[\ln\left(P_C\right)\right]\right\}$$

Substituting the power consumption of the blower into equation yields;

$$C_B = \exp\left\{6.8929 + 0.7900\left[\ln\left(3.1\right)\right]\right\}$$

$$C_B = 2{,}409.96 \quad \text{US\$}$$

The f.o.b. purchase cost can be calculated from Eq. 2.56;

$$C_P = 2.5 \times 2{,}409.96$$

$$C_P = 6{,}024.89 \quad \text{US\$}$$

Where $F_M = 2.5$ for stainless steel of centrifugal blower

In currently, the f.o.b. purchase cost can be calculated from Eq. 2.41 yields;

$$\text{Cost} = 6{,}024.89 \times \left(\frac{569.8}{500}\right)$$

$$\text{Cost} = 6{,}865.96 \quad \text{US\$}$$

5) Installation Cost

a) Packed Tower

Installation cost of the packed tower can be calculated from Eq. 2.58, which multiplier for installation costs of the packed tower is 2.1 yields;

$$\text{Installation cost} = 2.1\text{x } 87{,}575.75$$

$$\text{Installation cost} = 183{,}909.08 \text{ US\$}$$

b) Pump

Installation cost of the pump can be calculated from Eq. 2.58. The equipment cos of the pump includes pump and electric motor costs, which multiplier for installation costs of the centrifugal pump is 2.8 yields;

Installation cost = 2.8x 10,969.02

Installation cost = 42,998.56 US$

c) Blower

Installation cost of the blower can be calculated from Eq. 2.58, which multiplier for installation costs of the blower is 1.4 yields;

Installation cost = 1.4x 6,865.96

Installation cost = 9,612.34 US$

APPENDIX C

RESULT DATA

C.1 Diameter of the Packed Tower

Table C.1 Diameter of the Packed Tower with Various Stripped Wastewater Flow Rates

Flow Rate (m^3/h)	Diameter (m)
100	1.83
200	2.58
300	3.16
400	3.65
500	4.09
600	4.48
700	4.83
800	5.17
900	5.48
955	5.65

C.2 Cost Saving

Table C.2 Cost Saving from Reduction of NaOH Usage with Various Stripped Wastewater Flow Rates

Flow Rate (m^3/h)	Cost saving (Baht)
100	2,262,112.32
200	4,524,224.64
300	6,786,336.96
400	9,048,449.28
500	11,310,561.60
600	13,572,673.92
700	15,834,786.24
800	18,096,898.56
900	20,359,010.88
955	21,603,172.66

C.3 Power Requirements

Table C.3 Power Requirements of Pump and Blower with Various Stripped Wastewater Flow Rates

Flow Rate (m^3/h)	Electricity (kW)		Electricity Bill (Baht/year)	
	Blower	Pump	Blower	Pump
100	2.315355	1.898726	65,918.15	54,056.71
200	4.637296	3.797451	132,023.80	108,113.40
300	6.955783	5.696177	198,031.10	162,170.10
400	9.273036	7.594902	264,003.30	216,226.90
500	11.57936	9.493628	329,664.40	270,283.60
600	13.89576	11.39235	395,612.30	324,340.30
700	16.22549	13.29108	461,939.60	378,397.00
800	18.53269	15.18980	527,625.70	432,453.70
900	20.85558	17.08853	593,758.30	486,510.40
955	22.12104	18.13283	629,786.00	516,241.60

C.4 Investment Cost

Table C.4 Summary of the Investment Cost with Various Stripped Wastewater Flow Rates

Flow Rate (m^3/h)	Equipment Cost (Baht)			Installation Cost (Baht)			Total Investment Cost (Baht)
	Pump and Motor	Blower	Tower	Pump and Motor	Blower	Tower	
100	356,493.20	223,143.82	2,845,336.04	998,180.95	312,401.35	5,975,205.68	10,710,761.03
200	417,359.35	386,265.78	3,866,253.14	1,168,606.19	540,772.09	8,119,131.60	14,498,388.15
300	469,380.36	532,096.51	4,714,067.15	1,314,265.02	744,935.11	9,899,541.02	17,674,285.17
400	516,153.48	667,793.77	5,474,919.87	1,445,229.73	934,911.28	11,497,331.73	20,536,339.87
500	559,418.52	795,880.54	6,194,844.95	1,566,371.85	1,114,232.76	13,009,174.39	23,239,923.01
600	600,144.35	919,208.51	6,862,907.45	1,680,404.18	1,286,891.91	14,412,105.65	25,761,662.05
700	638,931.14	1,038,946.68	7,486,895.97	1,789,007.18	1,454,525.35	15,722,481.54	28,130,787.86
800	676,179.40	1,154,006.99	8,115,512.38	1,893,302.33	1,615,609.79	17,042,576.01	30,497,186.91
900	712,171.37	1,266,842.41	8,708,152.80	1,994,079.82	1,773,579.37	18,287,120.88	32,741,946.64
955	731,508.78	1,327,190.98	8,864,129.66	2,048,224.57	1,858,067.37	18,614,672.28	33,443,793.63

Note: Convert from US$ with exchange rate 32.5 baht/US$

BIOGRAPHY

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