Contents

CHAPTER 1

INTRODUCTION

1.1   Process Background

The main objective of preparing this design project is to propose the most suitable plant design of producing 100,000 MMSCFD of liquefied petroleum gas (LPG). There are several processes of producing LPG and it will be further discuss in this report.

1.1.1   History of Liquefied Petroleum Gas (LPG)

LPG was first identified as a significant component of petroleum in 1910. The person who discover the LPG was Dr. Walter O. Snelling, a chemist and explosive expert in U.S. Bureau of Mines. He discovered that a large part of liquid gasoline was actually composed of LPG, including propane, butane, and other hydrocarbons. This happens when he realized that the liquid methanol was evaporated at a rapid rate. Snelling soon realized that the LPG could be used for lighting, metal cutting, and cooking. So, created a ways to liquefy the LPG during the refining of natural gasoline. He started a company along with others cooperation called American Gasol Company, the first commercial marketer of LPG. Snelling managed to produce relatively pure propane by 1911 and, in 1913, his LPG technology was awarded a U.S. patent.

Later that year, the patent was bought by Frank Philips, the founder of the Conoco Philips oil company. Information concerning practical use of LPG dates to 1918, when the fuel was utilized for brazing lamps and metal-cutting blowtorches. However, commercial production did not begin until the 1920's. LPG sales in the US topped 223 thousand gallons in 1922, while within the next 3 years the figure grew to 400 thousand gallons. In 1928, LPG was first used as motor fuel (in a truck) and the first LPG refrigerator was made. In 1929, the level of sales of the fuel was as much as 10 million gallons in the US. By 1912, propane gas was cooking food in the home. The first car powered by propane ran in 1913. By 1915 people were using propane in torches to cut through metal. LPG has been used as a transportation fuel, mainly in heavy trucks and forklift vehicles, around the world for more than 60 years.

LPG processing involves separation and collection of the gas from its petroleum base. LPG is isolated from the petrochemical mixtures in one of two ways -- by separation from natural gas or by the refining of crude oil. Both processes begin by drilling oil wells. The gas/oil mixture is piped out of the well and into a gas trap, which separates the stream into crude oil and "wet" gas, which contains natural gasoline, LPG and natural gas. The heavier crude oil sinks to the bottom of the trap and is then pumped into an oil storage tank for refining. Crude oil undergoes a variety of refining processes, including catalytic cracking, crude distillation, and others. One of the refined products is LPG. The "wet" gas, off the top of the gas trap, is processed to separate the gasoline (petrol) from the natural gas and LPG. The natural gas, which is mostly methane, is piped to towns and cities for distribution by gas utility companies. The petrol is shipped to service stations. The LPG also enters the distribution pipeline, where it eventually finds its way to end users, including for home and commercial LPG users all around Australia and the world.

1.1.2 Overview of LPG production

Liquefied Petroleum Gas (LPG) is a byproduct of natural gas extraction and crude oil refining. It is considered a clean and environmentally-friendly source of energy. If made available as a cooking energy fuel it could help to reduce the negative health impacts of 2.6 billion people who currently use biomass stoves that emit harmful smoke and particles due to incomplete combustion. Varieties of LPG bought and sold include mixes that are primarily propane (C_3H_8), primarily butane (C_4H_{10}) and, most commonly, mixes including both propane and butane. In the northern hemisphere winter, the mixes contain more propane, while in summer, they contain more butane. In the United States, primarily two grades of LPG are sold: commercial propane and HD-5. These specifications are published by the Gas Processors Association (GPA) and the American Society of Testing and Materials (ASTM). Propane/butane blends are also listed in these specifications.

Propylene, butylenes and various other hydrocarbons are usually also present in small concentrations. HD-5 limits the amount of propylene that can be placed in LPG to 5%, and is utilized as an autogas specification. A powerful odorant, ethanethiol, is added so that leaks can be detected easily. The internationally recognized European Standard is EN 589. In the United States, tetrahydrothiophene (thiophane) or amyl mercaptan are also approved odorants, although neither is currently being utilized.

Table 1.1: *Properties of LPG*

	LPG example	EN 589: 2008+A1:2012
Formula	x%, C_3H_8 x%, C_4H_{10}	
Molecular weight, g/mol	44-58	
Carbon/Hydrogen/Oxygen, wt%	82 - 83/17 - 18/0	
Applicable compression ratios	11 – 13	
Density, liquid at 20ºC, kg/dm³	0.5 - 0.58	
Boiling point, ºC	-42 - -0.5	
Blending vapor pressure at 20ºC, kPa	210 – 810	
LHV heating value, MJ/kg	44 – 46	
LHV heating value, MJ/l	23 – 26	
HHV, MJ/kg	48 -50	
Heat of vaporization at 20ºC, kJ/kg	358 -372	
Self-ignition temperature, ºC	365 – 470	
Ignition limits, fuel in air, vol%		
Stoichiometric air to fuel ratio	15.4	
Total dienes, mol%		≤ 0.5
Hydrogen sulphide		Negative
Total sulfur content, mg/kg		≤ 50
Odor		Unpleasant and distinctive at 20% LFL

*varying butane/propane ratio, e.g. 70% propane & 30% butane to 100% propane [IEA 1999]

1.1.3   Uses of LPG

LPG can be used in many applications in the industrial sector namely in space- and process-heating, powering industrial ovens, production of food, kilns, furnaces, production of packing material as well as in powering forklift trucks in warehouses.

- **Ceramic**

LPG is one of the best choices of energy in this particular industry. Ceramics made of clay require a high heating value in order to dry and become hard and solid. Easily controllable, LPG provides clean combustion and is therefore advantageous in the maintenance process. Burners and kilns have to be maintained less often causing less downtime, with the consequence of saving costs and increasing productivity. As a choice energy, LPG is widely used in pottery, roofing, ceramic tiles and sanitary ware.

- **Food Processing**

LPG is widely used in many food processing systems because of its clean burning properties. Bakeries and the manufacturers of biscuits, chips and chocolate are inclined to choose LPG as their preferred energy option as their products will not be exposed to the risk of contamination. In addition, this energy is also used in slaughterhouse for the cleaning of facilities and sterilisation, pork butchery in the process of cooking, drying and smoking as well as in the dairy industry for pasteurisation.

- **Textile**

The textile industry consists of three groups, namely:
- Natural textiles: cotton, wool, linen, silk,
- Artificial textiles: derived from natural products

- Synthetic textiles: 100% chemical products with a high degree of polymerisation.

The textile industry requires a number of energy-consuming processes for which LPG is found to be a suitable fuel. These processes are:

- Heating of the bath (cleaning, bleaching, dyeing)
- Drying, thread singeing and polymerisation
- Ironing

Total gaz LPG has multiple advantages in that it is clean and flexible, easy to maintain and produces a high thermal yield.

- **Printing**

The manufacture of pulp, paper and cardboard consumes a considerable amount of energy. If fuel oil is to be commonly used in manufacturing pulp, it is possible for LPG to find many applications in paper, cardboard manufacturing processes and printing. In colour printing, the paper has to go several times through the rotary press as it requires a fast ink drying process. It is therefore always better to use the decentralised heating solution rather than the centralised steam solution in these processes as it offers more advantages: modular heating, easy regulation of gas output, cleanliness and environmental care.

- **Chemicals production**

A number of products are part of the field of chemical engineering, i.e.

- Polymers
- Paint
- Varnish
- Colourings and dyes
- Wax and polish.

A number of products are destined to:

- The food industry (aromas, flavours, spice extracts)
- The pharmaceutical industry

Although chemical engineering is the biggest user of steam boilers, LPG-powered heat-exchangers are recommended in a number of different stages of the process. By using LPG, the chemical processes are enhanced thanks to good temperature regulation attributes, very high yields of energy and low maintenance procedures that are required by this type of installation.

- **Forklifts**

Good for meeting air quality regulations in the workplace and technical demands for a modern handling (rapidity, power, flexibility, economy), LPG is nowadays the best response for the fuelling of thermal-engine trucks. Lead-free and soot-free, LPG has a very low rate of carbon monoxide emissions and is therefore the chosen energy when a pollution-free environment is critical. Should you invest in new forklifts trucks, there are many reasons why you should use LPG models.

1.2   Physical and Chemical Properties of Substances Involved

Physical properties are alteration in temperature and pressure than can be measure by measurement tools. Physical properties include formula weight, refractive index, specific gravity, melting point, boiling point, solubility and others. Chemical properties describe the procedure of a chemical reaction and substances that forming a solution. For example enthalpy, entropy, heat capacity and others. It is essential to know the physical and chemical properties of the chemical used in order we can alter and manage the chemical easily and reduce the possibility of avoidable accident. In this following section, the physical and chemical properties of various chemicals are listed in the table

1.2.1 Dimethyl Ether

Table 1.2: *Properties of Dimethyl Ether (DME)*

Molecular formula	CH_3OCH_3
Molar mass, g/mol	46.07
Appearance	Colorless gas
Odor	Typical
Density, kg/cm^3	2.1146
Melting point, K	132
Boiling point, K	249
Solubility	71 g/dm^3 in water at 20 ºC
Henry's law constant, K_H	n/a
Specific heat capacity, J/K.mol	65.57
Std enthalpy of formation, $\Delta_f H^o{}_{298}$	-184.1
Std enthalpy of combustion, $\Delta_c H^o{}_{298}$	-1.4604
Auto-ignition temperature, ºC	350
Explosive limits, %	27

1.2.2 Methane

Table 1.3: *Properties of Methane*

Molecular formula	CH_4
Molar mass, g/mol	16.04
Appearance	colorless
Odor	odorless
Density, g/cm^3	0.42264
Melting point, K	90.7
Boiling point, K	111.66
Solubility	22.7 mg/L in water, soluble in methanol and dimethyl ether
Henry's law constant, K_H	14 nmol/Pa.kg
Specific heat capacity, J/K.mol	35.69
Std enthalpy of formation, $\Delta_f H^o_{298}$	186.25
Std enthalpy of combustion, $\Delta_c H^o_{298}$	-891.1 - -890.3
Auto-ignition temperature, ºC	537
Explosive limits, %	4.4 -17

1.2.3 Hydrogen

Table 1.4: Properties of Hydrogen

Molecular formula	H_2
Molar mass, g/mol	1.008
Appearance	Colorless gas
Odor	Typical
Density, g/L	0.08988
Melting point, K	13.99
Boiling point, K	20.271
Solubility	
Henry's law constant, K_H	n/a
Specific heat capacity, kJ/Kg.K	14.31
Std enthalpy of formation, $\Delta_f H^o{}_{298}$, kJ/mol	-
Std enthalpy of combustion, $\Delta_c H^o{}_{298}$, kJ/mol	-286
Auto-ignition temperature, ºC	-
Explosive limits, %	-

1.2.4 Water

Table 1.5: Properties of water

Molecular formula	H_2O
Molar mass, g/mol	16.04
Appearance	colorless
Odor	odorless
Density, g/cm^3	0.42264
Melting point, K	273.15
Boiling point, K	373.15
Solubility	yes
Henry's law constant, K_H	n/a
Specific heat capacity, J/K.mol	35.69
Std enthalpy of formation, $\Delta_f H^o_{298}$	n/a
Std enthalpy of combustion, $\Delta_c H^o_{298}$	n/a
Auto-ignition temperature, ºC	n/a
Explosive limits, %	n/a

1.3   Market Analysis

Liquefied petroleum gas or liquid petroleum gas (LPG), some country also referred it as simply propane or butane , are flammable mixtures of hydrocarbon gases used as fuel in heating appliances, cooking equipment, and vehicles. Varieties of LPG bought and sold include mixes of propane and butane. At the northern hemisphere, in winter, the mixes contain more propane, while in summer, they contain more butane. LPG is prepared by refining petroleum or "wet" natural gas.

Dimethyl ether (DME), also known as methoxymethane, is the organic compound and the simplest ether. It is colourless gas that is useful precursor to other compounds and an aerosol propellant and is being studied as a future energy option. The largest use of DME is as the feedstock for the production of the methylating agent. A mixture of DME and propane is used in an over-the-counter device to treat warts, by freezing them. A potentially major use of DME is as substitute for propane in LPG used as fuel in household and industry.

1.3.1   LPG worldwide market overview

A spike in Asian LPG prices caused the US-Asia business window to re-open after several months of closure, but market sentiment is that such high US-Asia price differentials are not expected to last for long. In that case, most of the US exports in 2015 would be contractually-committed, since exporters are understood to have hedged through 2015; these should still be economically viable.

The US is, however, not the only region that increases exports in 2015. The Arabian Gulf is forecast to grow by around 1 mm t where, in particular, it expected more LPG out of Abu Dhabi, Kuwait and Iran. There may also be higher exports out of West Africa if Angola finally gets its new LNG plant operating properly.

On the other hand, European exports are expected to decrease overall, with lower North Sea producer exports offsetting higher ex-Russia volumes. The majority of the increased global exports will go to Asia, which is forecast to see its imports rise from just over 43 mmt to 45.5 mmt. Imports into the Med and Latin America are also expected to rise.

Within Asia, it is China and India looking most likely to show high growth, in China this is highly dependent on the start-up and operation of still more new propane dehydrogenation (PDH) facilities; in India it is the development of the rural LPG market that leads the way, although the size of this Indian growth increase will depend on the government's approach to subsidies.

Thai imports may also expand, spurred on by high auto-gas demand growth, but here the withdrawal of government subsidies, if fully implemented it could have a dampening effect on retail market demand. Same thing could also apply in Indonesia.

In 2014 and 2015, past this full year, show the largest exporting region remained the Arabian Gulf (AG), accounting for an estimated 45% of all exports with the majority of LPG from the AG going to Asia (96%). Of the total 34.48 mmt of AG exports, Qatar and Abu Dhabi accounted for 10.4 and 9.6 mmt respectively. Shale developments in the US during 2014 significantly increased exports from there, with total US exports for the year estimated to be around 14 mmt, 5.2 mmt higher than in 2013. Trade between North America and Latin America increased, as well as from North America to Northern Europe, mainly at the expense of West Africa. As a result, West Africa moved more shipments east through 2014.

Diagram below show the global LPG consumption from 2000-2010. From the data we find that Asia Pacific consumption has increase rapidly, from 55,000-89,000 MT. The consumption of LPG are expected to keep increasing in the future especially in the Asia Pacific region.
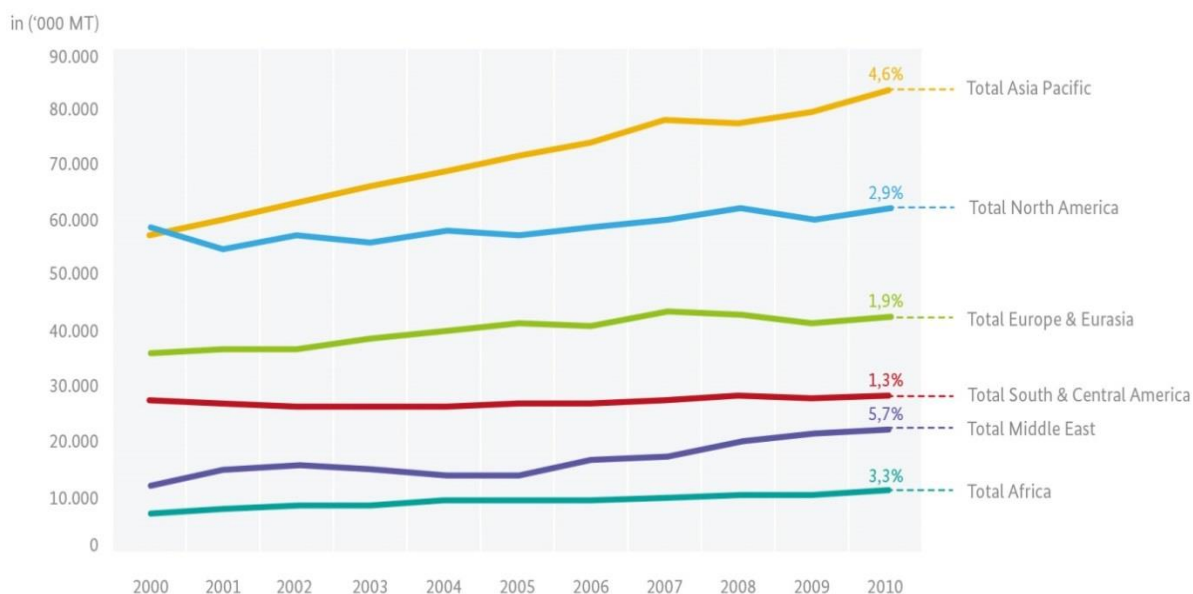
Figure 1: *Global LPG consumption from 2000-2010*

1.3.2 DME worldwide market overview

The Asia-Pacific region is the largest market for DME in the world. Jiutai Energy Group (China), China Energy Limited (Singapore) are the leading players in the Asia-Pacific market. DME is majorly used for LPG blending purpose followed by aerosol propellant. In Asia-Pacific region the capacity has gradually increased and it will grow continuously with new plant establishment for the domestic fuel supply. Asia-Pacific is a developing region. Also, being the most populated region global, the demand for fuel substitutes is the highest in this region. These factors drive the demand for fuel substitutes such as DME. Major countries which contributed the maximum shares in this market are China, Japan, South Korea, and Indonesia.

Most of the DME produced these days is an output of a two-step process in which the first one is methanol generation by using coal or natural gas as a feedstock and the second is methanol dehydration to produce DME. Coal is employed as a feedstock in Asian countries while countries such as Papua New Guinea and Saudi Arabia focus on natural gas reforming to produce methanol and then convert it to DME. Major DME manufacturers include Akzo Nobel N.V. (The Netherlands), Royal Dutch Shell Plc. (The Netherlands), The Chemours

Company (U.S.), China Energy Limited (Singapore), Mitsubishi Corporation (Japan), Ferrostal GmbH (Germany), Grillo Werke AG (Germany), Jiutai Energy Group (China), Oberon fuels (U.S.), and Zagros Petrochemical Company (Iran).

Among the various applications of DME, LPG blending represents the largest market for DME. LPG scarcity is a great concern in the countries that have less gas reserves, as they have a very limited choice of replacing LPG with natural gas. DME has emerged as a trustworthy source for such countries to minimize their LPG imports by blending a pre-described amount of DME in LPG without any change in current infrastructure. DME is also implemented as an aerosol propellant under liquefied gases propellant category. The various other applications of DME include its use in the form of chemical feedstock, solvent, refrigerant, and for welding & brazing operations.

DME has the potential to replace diesel in conventional engines with slight modification in fuel injection systems. The potential market for DME as a transport fuel is large as diesel consumption is increasing at a high pace, majorly driven by transportation sector. The global DME market is projected to witness high growth in the next five years due to its clean combustion property. The market is also growing due to the increasing demand of LPG in various countries. However there are certain challenges in using Dimethyl ether for blending with LPG. DME can corrode the rubber sealants used in LPG cylinders or tanks and hence can lead to severe accidents due to gas leakages. Only 20 % of DME can be used for blending. For higher proportions will need a huge change in infrastructure.

1.3.3 LPG local market overview

LPG is produced as a by-product of oil production (via the processing of associated gas), gas production (via the processing of non-associated gas) and refining. Therefore, LPG is a supply-driven market in which demand must adjust to match supplies. LPG production tends to increase quickly when one or more of the above industries are expanding quickly. In Malaysia, LPG is normally bottled in gas cylinders for sale to domestic, commercial and industrial consumers. LPG for cooking in homes is available either in 10kg, 12kg or 14 kg cylinders. LPG in 50kg cylinders is meant for commercial and industrial users. LPG price in Malaysia is heavily subsidized by the government for domestic consumers.

Diagram below show the LPG consumption for Malaysia from 1986 to 2010, the data being provided from The U.S. Energy Information Administration. The average value for Malaysia during that period was 34.94 thousand barrels per day with a minimum of 3.87 thousand barrels per day in 1986 and a maximum of 85.96 thousand barrels per day in 2010.

Source: TheGlobalEconomy.com, The U.S. Energy Information Administration
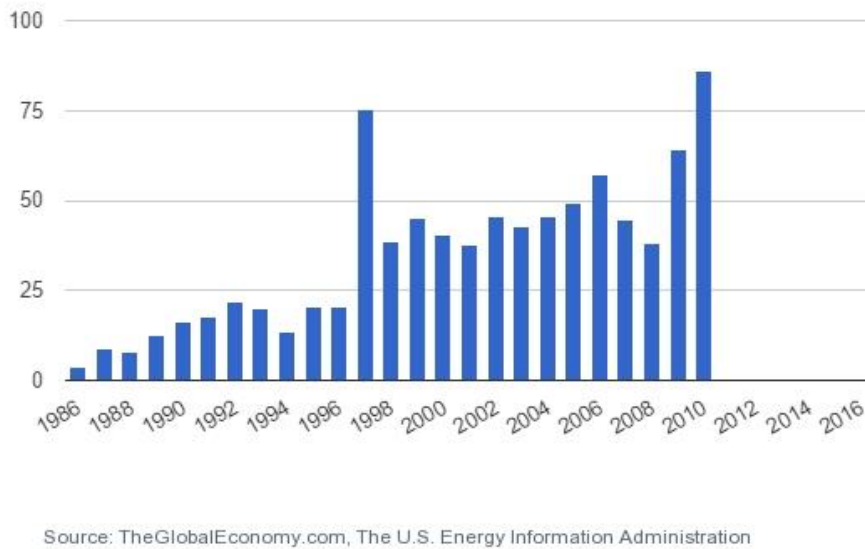
Figure 2: Malaysia LPG consumption from 1986 to 2010

1.4   Process Selection

Producing Liquefied Petroleum Gas (LPG) by using Dimethyl Ether (DME) can be achieved by two methods which are first by buying the DME from a DME producing plant or second by producing the DME ourselves.For the first method, DME will be supplied by a local or international company to meet our need. For the second method, two step process is chosen because currently there is more widespread application of the two-step process since it is relatively simple and start-up costs are relatively low. Two step process means there will be a dual catalyst system that permits both methanol synthesis and dehydration in the same process unit, with no methanol isolation and purification.

1.4.1   Pathway

There are two pathways that will be discussed in this section. The best pathway will be chosen by looking at its efficiency economically and so on.

1.4.1.1 Pathway 1: Producing LPG from DME by using Hydrogen

In this process, the raw material will be DME and hydrogen. Both of the raw material will be supplied by the supplier company. Firstly, the DME is dehydrated to generate a carbene (H2C:) by a concerted catalysis of an acidic site and a basic site, which are at spatial field inside a pore in a zeolite catalyst. Then, the carbene is polymerized to form an olefin containing propylene or a butene as a main component. After that, olefin-hydrogenation catalyst will be used to turn olefin into paraffin. Then, water, a low-boiling component such as hydrogen, and a high-boiling component is separated from the lower-paraffin-containing gas produced, as necessary, to obtain the LPG. If necessary, the gas may be pressurized and/or cooled so as to obtain the LPG. The chemical reaction will be as follows :

$$DME + H_2O \rightarrow LPG$$
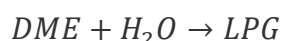
1.4.1.2   Pathway 2: Produce DME from natural gas

There are two process routes to produce DME from natural gas feedstock which are direct or indirect process. Basically, the process to produce DME from natural gas feedstock is made up of the reaction to reform raw material methane into syngas by using steam or oxygen as oxidant. Next, from syngas will produce methanol and then methanol will be dehydrated to DME. Both routes have same principle of DME production, but the different between them is the direct route makes methanol synthesis reaction and DME synthesis

reaction occur in one reactor, while the indirect route makes these reactions in separate reactor. Both routes have their own advantages that would be discussed in next section. However, in this process we prefer direct route to produce DME.
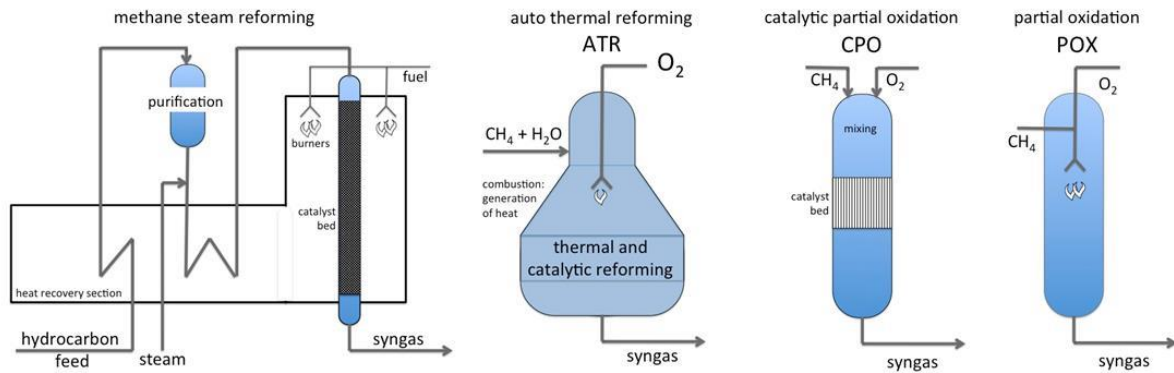
Figure 3: Reactors and process layout for syngas production from natural gas and shale gas.
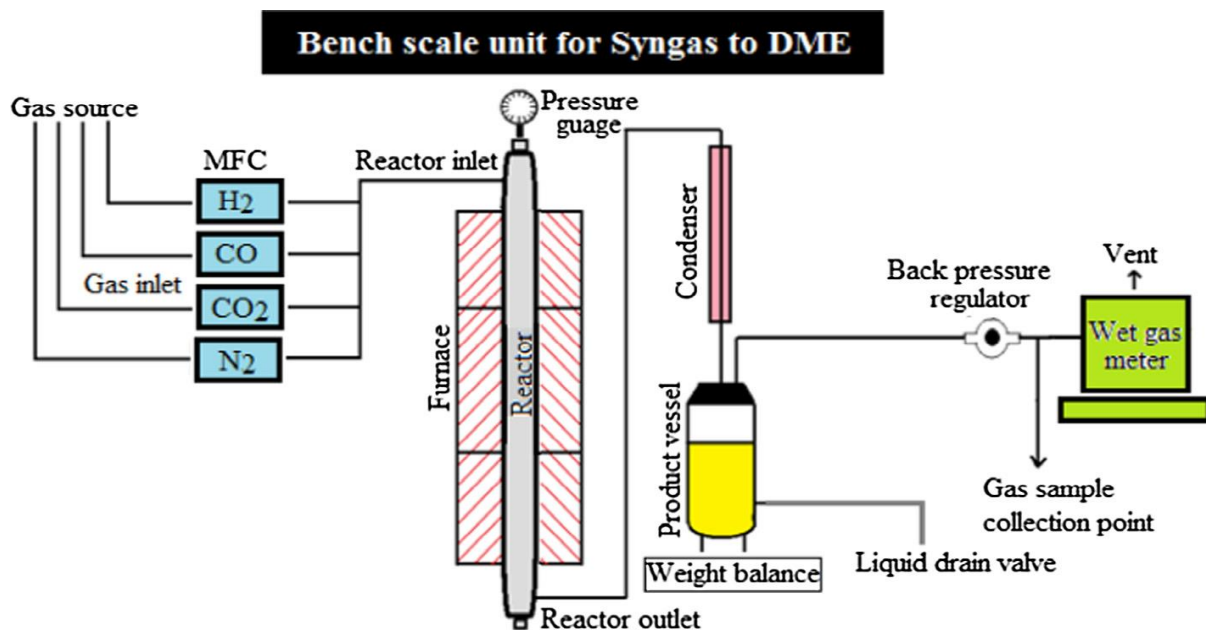
Figure 4: Schematic diagram of a tubular reactor for bench scale conversion of syngas to DME.

1.4.2   Gross Profit (GP) OR Estimated Profit (EP)

There are several ways can be taken in order to obtain an end product of LPG. But, the production of LPG also is the same like the others product in the industry which must consist of the reference or theoretical step in their production before the management along with the engineers in each company change their own raw material based on the reference step. In the calculation, we will explain thoroughly the reasons for each company to change their raw products instead of following the theoretical steps.

Note: Some assumption had been made in this calculation due to the pricing range. As the profits that need to be count is estimated one, some value than within the range is being taken. The maximum values has been taken for the product price while the minimum prices being chosen for the reactants in order to maximize the profits gain. All information regarding the prices had been retrieve from ICIS website (ICIS,2015).

1.4.2.1 Economical Pathway Equation
Calculation using pathway 1:
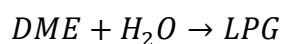
$$DME + H_2O \rightarrow LPG$$

Table 1.6: *Data of component for Pathway 1*

Compound	Molecular weight (kg/kmol)	$/kg	RM/kg
DME	46.07	0.438	1.785
H2O	18	0.02	0.07
LPG	44.01	0.8182	3.33

Compound	DME	H2O	LPG
Kmol	1	1	1
MW	46.07	18	44.01
Mass	46.07	18	44.01
kg/kg LPG	1.047	0.409	1
RM/kg LPG	1.869	0.029	3.33

Gross profit = (3.33) – (1.869+0.029)

$\qquad$ = RM 1.432 /kg LPG

Calculation using pathway 2:

1. Natural gas to syngas

$CH_4 + H_2O \rightarrow CO + 3 H_2$ $\qquad\qquad$ 1

2. Syngas to methanol

Water-gas shift reaction

$CO + H_2O \leftrightarrow CO_2 + H_2$ $\qquad\qquad$ 2

$CO_2 + 3H_2 \leftrightarrow CH_3OH + H_2O$ $\qquad\qquad$ 3

Combination 2 and 3

$CO + 2H_2 \leftrightarrow CH_3OH$ $\qquad\qquad$ 4
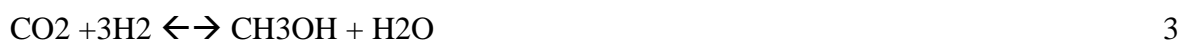
Combination 1 and 4

$CH_4 + H_2O \rightarrow CH_3OH + H_2$ $\qquad\qquad$ 5
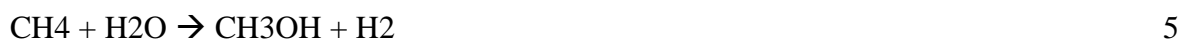
3. Methanol to DME by dehydration

$\qquad 2CH_3OH \rightarrow CH_3OCH_3 + H_2O$

Overall equation:
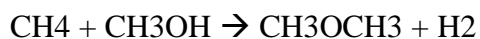
CH4 + CH3OH → CH3OCH3 + H2

Table 1.8: *Data of component of pathway 2*

Compound	Molecular weight g/mol	$/kg	RM/kg
CH4	16.04	0.09177	0.3735
CH3OH	32.04	0.255	1.038
CH3OCH3	46.07	0.438	1.785
H2	1.00	3.08	12.536

Table 1.9: *Calculation of pathway 2*

Compounds	CH4	CH3OH	CH3OCH3	H2
Kmol	1	1	1	1
Mw	16.04	32.04	46.07	1
Mass	16.04	32.04	46.07	1
Kg/kgDME	0.348	0.695	1	0.0217
RM/kgDME	0.1305	0.7214	1.78	0.272

EP Value = total product − total reactant

$$= (1.78 + 0.272) - (0.1305 + 0.7214)$$

$$= RM\ 1.20/\ kgDME$$

1.4.2.3 Comparison and Selection of Process

General description on all the alternative pathways of LPG production from DME using hydrogen have been discussed, hence in this section, all those pathways will be compared in terms of economic potential, manufacturing process in terms of the operating conditions (Briggs et.al, 1985).

NO.	Process	Parameter	Catalyst	Advantages/Disadvantages
1	Production of LPG from direct DME (by supplier)	Catalytic Yield: Temp: 450℃	1) Zeolite 2) Olefin-dehydro	• Lower the number of process unit as DME is from

		Pressure: 0.15MPa	genation catalyst	supplier • Lower the plant cost as number of process unit decrease • Simpler chemical process reaction involved
2	Produce DME from natural gas and continue with LPG production		1)Copper based catalyst	• Relatively simple • Low start-up cost

All the 2 production pathways have been thoroughly discussed and compared in terms of their advantages, disadvantages. The most important considerations are being the feasibility of the process commercially in economical and safety aspects. Conclusion, we decide to use production of LPG from direct DME (by supplier) for our plant project.

1.5   Site Location

In order to produce LPG from DME, detail study regarding on site location need to be carried out to make sure this project successful and effective. For the construction of the LPG plant we have find out the industrial area that are suitable for the construction of the plant. The location that is favourable is Teluk Kalong , Kertih or pasir gudang, Johor. The area is ready-built factories, investors have the option of moving directly into ready-built factories with preinstalled basic facilities like broadband, power and water, which reduces the time needed to get a project off the ground.

To determine suitable site location, site study need to do and it includes a few element such as land price, raw material availability, transportation, market, etc. Table 1 shows information about site study.

Table 1.10: *Site Study*

	Location	
	Eastern Gate Development Pasir Gudang, Johor	Teluk Kalong , Kertih, Terengganu
Raw Material Supply	Gas Malaysia/ Petronas Gas	Petronas Gas
Electricity supply	Tenaga Nasional Berhad (TNB)	Tenaga Nasional Berhad (TNB)
Size Of Land Vailable (Ha)	173	
Price Of Land Per Ft^2	RM 14 - 20	RM 45 - 60
Distance From Nearest Town	40 km from Kemaman	30 km from Johor Bahru and only 55 km from Singapore
Land Transportation	Lebuhraya Utara Selatan (PLUS)	Lebuhraya Pantai Timur (LPT)
Port	Pasir Gudang Port	Kemaman Port
Water Supply	Syarikat Air Johor (SAJ)	Syarikat Air Terengganu (SATU)
Electricity Tariffs Period Peak	42.70 Cent	42.70 Cent
Off-Peak	19.10 Cent	19.10 Cent
Water Rate per m^3	2.96 Cent	2.96 Cent
Worker Supply	Universiti Teknologi Malaysia (UTM) Johor Skills Development Centre (JSDEC)	Tati Universiti College, Terengganu

For long term period, the plant site can consequence capital cost, operational cost and market price. Factors that need to put into consideration when choosing the proper plant location included raw material availability and cheap, source of power and water, reasonable land price, skilled man power, and last but not least good transportation network (via land, sea and air)

Based on the detail information and consideration on table 1, our group had selected the strategic plant location at Pasir Gudang, Johor. Justification and comparison for the selection of pasir gudang johor compare to teluk kalong as below;

1.5.1 Land

Land play important role for the selection of site, because it can affect the capital investment. The land with most economical and reasonable price would be an advantage in capital cost. Based on table 1, land at pasir gudang had the lowest price per ft^2 compare to teluk kalong. When the price of land is cheapest, the fund allocation to buy the land can be minimized. The land characteristics (geological factors) at the proposed plant site should be examined carefully. Sufficient suitable land must be available for the proposed plant for future expansion. Besides that, there is still huge land available at Pasir Gudang compare to Teluk Kalong. This advantage can be used to expand our plant in the future. A suitable land should be ideally flat, well drained and have suitable load-bearing characteristics such as for piling or other special foundation.

Figure 5: *Pasir Gudang Industrial Area*

Figure 6: *Eastern Gate Development Pasir Gudang, Johor*

1.5.2   Market

The market that required LPG in bulk quantities should located close to the plant where the production and transportation require a lot of cost. This consideration will be less important for low volume production. Based on our local demand, there is huge market demand for LPG as Johor is a developed state under ISKANDAR MALAYSIA and Pasir Gudang also is an industrial area.

LPG is a product that be used in domestic and industrial sector that provide large heating value. LPG is widely used in households for heating appliances such as stoves, ovens, water heaters and dryers. It is an ideal fuel for cooking as it provides instant and consistent heat which can be controlled easily. While for industries, LPG use for gas engines in forklifts, industrial boilers, and etc because it's reliable and steady.

1.5.3   Roads

Lebuhraya Utara Selatan (plus) only take 2 hours to travelled from Pasir Gudang to nearest port which is pasir gudang port in which located 184km. Pasir Gudang also located to almost the same distance to the port klanf which took 2 hours and half if travelling through the highway. This highway makes it convenient for promoting economic activities in major districts in johor. By have this accommodation, our company can reduce the time to for the product and raw material transportation.

1.5.4   Raw material

The source of raw material is one of the most important factors influencing the selection of the plant site. The closer the site to the raw materials suppliers, the cost of transportation will be cheaper. In earlier stages, we assume that the raw material, DME to be imported from Indonesia. On the other hand, for future expansion the plant may produce their own DME from natural gas. In advantage, Pasir Gudang had been supply with natural gas

from Petronas Gas or Gas Malaysia through gas transmission line as shown in figure 1. The time to receive the raw material can be reduced as the pipeline provide continuos supply of natural gas.
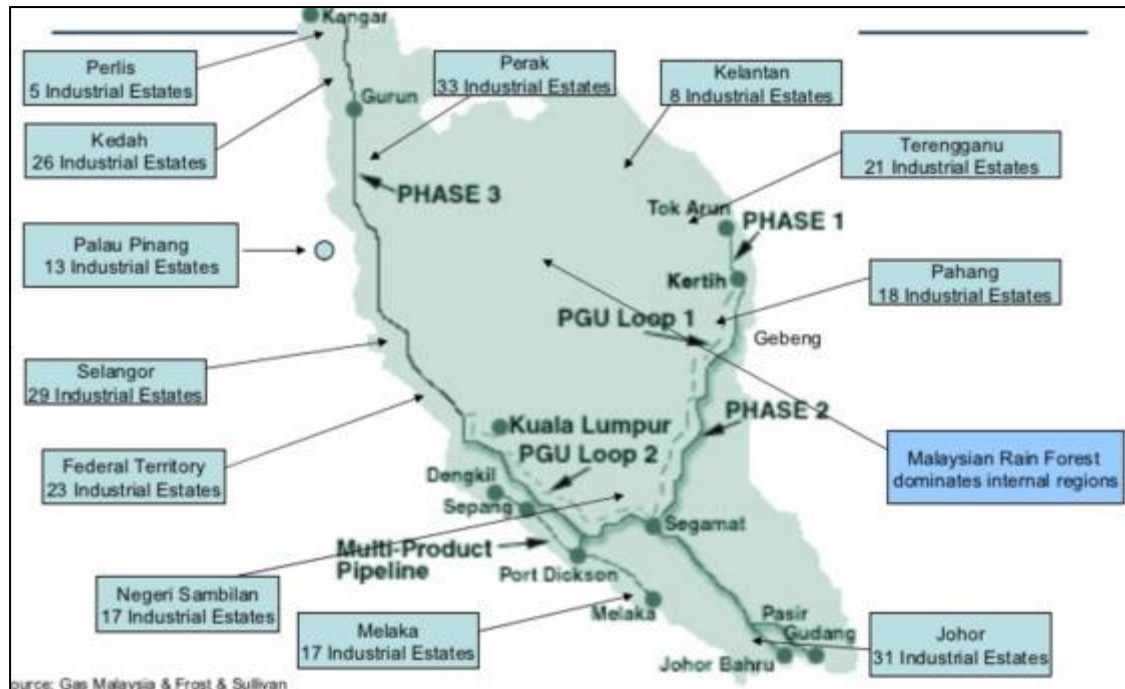
Figure 7: *Natural Gas Tranmission Pipeline Peninsular Malaysia*

1.5.5   Utilities (water and electric power supply)

Water and electric power supply utilities are very important to make sure the operation of plant effectively without any disruption, hence production of LPG going well. Large quantities of water for cooling and general processes use and the plant must be located near a source of water of suitable quality. The water may be drawn from a river, from wells, or purchased from a local authority. Stable electrical power supply will be needed inorder for the plant to run smoothly. The plant that requires large quantities of power to operate need to be located close to a cheap source of power. With the supply of water from Syarikat Air Johor (SAJ) and electricity from Tenaga Nasional Berhad (TNB), our plant can be smoothly run for the whole schedule for the year.

1.5.6 Transportation

Transportation is on of the important factor for the site selection in order to buy the raw material and to deliver product. If practicable, a site should be selected that is close to at least two major forms of trans port such as road, rail or a sea port. Air transport is convenient and efficient for the movement personnel and essential equipment and supplies. While, road and sea transport is being increasingly used, and is suitable for long- distance transport of bulk chemicals. The facilities at Johor such as Senai International Airport and Pasir Gudang Port make it conductive for our plant and investment since the increase petrol fuel and toll will affect the profit for the transportation.

PROCESS MODELLING AND FLOWSHEETING

2.1 Introduction

In this chapter, we will mainly discussed about the process which includes the chemical reaction, process unit and process condition. The modelling and flowsheeting is constructed with a proper sequence to convert the raw material to desired product, liquefied petroleum gas (LPG). The raw material for the process would be Dimethyl Ether (DME) and hydrogen.

The plant is to produce a total amount of 100,000 metric ton LPG per annum. The operating day for this plant is 335 days per year and 24 hours a day. The LPG to be produced should have more than 60% of propane purity, less than 40% butane purity and less than 2% of other hydrocarbon purity in order to be sale in the market.

In this LPG plant design, there are several main synthesis and process would be involve which are :

- Mixing section
- Hydrocarbon synthesis
- Separation and Purification of product

2.2   Preliminary Process Synthesis Step

2.2.1   Eliminate Differences in Molecular Type

There are several LPG synthesis methods but those synthesis methods have different raw material and yield different composition as well as the product composition. These are the synthesis methods that have been discussed in Chapter 1:

Pathway 1 = Production of LPG from direct DME (by supplier)
Pathway 2 = Produce DME from natural gas and continue with LPG production

As been discussed in Chapter 1, the raw material of pathway 2 is natural gas which is to produce DME which then are used to poduce LPG while in pathway 2 the raw material is DME which are supplied by the supplier. Eventhough, the gross profit for pathway 2 is mostly desirable compared to Pathway 1, we will be choosing pathway 1 because of its inherent safer (simplification of process and minimization of the number of process unit) and cheaper than pathway 2.

2.2.2 Distribute the Chemicals

Overall Chemical Reaction of pathway 1 :

$$3000 H_2 + 100 DME$$
$$\rightarrow 2.5 CH_4 + 2.0 + 52.5 C_3H_8 + 32.8 C_4H_{10} + 8.3 C_5H_{12} + 0.9 C_6H_{14} + 0.3 CO$$

Overall Chemical Reaction of pathway 2 :

$2CH_4 + O_2 \rightarrow CH_3OCH_3 \text{ (DME)} + H_2O$ (1)
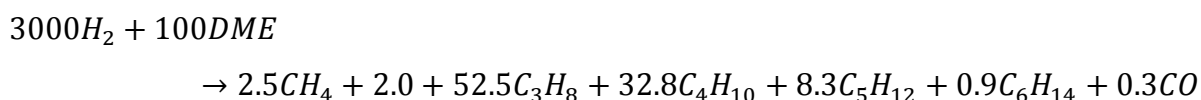
$328.7H_2 + 100DME \rightarrow 3.114CH_4 + 11.871H_6 + 24.909C_3H_8 + 38.433C_4H_{10} + 12.454C_5H_{12} + 6.4C_6H_{14} + 2.797CO$ (2)
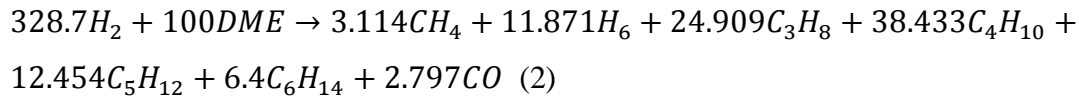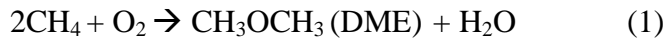
Conversion of reactant DME for producing LPG for both process are 100%.

2.2.3 Eliminate Differences in Composition

According to the third synthesis step, there are some product stream components that needed to be separate by certain separation equipments.

The separation above is based on the heuristic for determining favorable sequences:

1. Remove final product as a distillate one by one.

2. Sequence separation in order of decreasing relative volatility so that the most difficult splits are made in absence of the other components.

3. A sequence separation point to remove is based on the greatest molar percentage of the components in the feed.

4. Remove thermally the unstable component in the sequence

2.2.4 Eliminate Differences in Temperature, Pressure
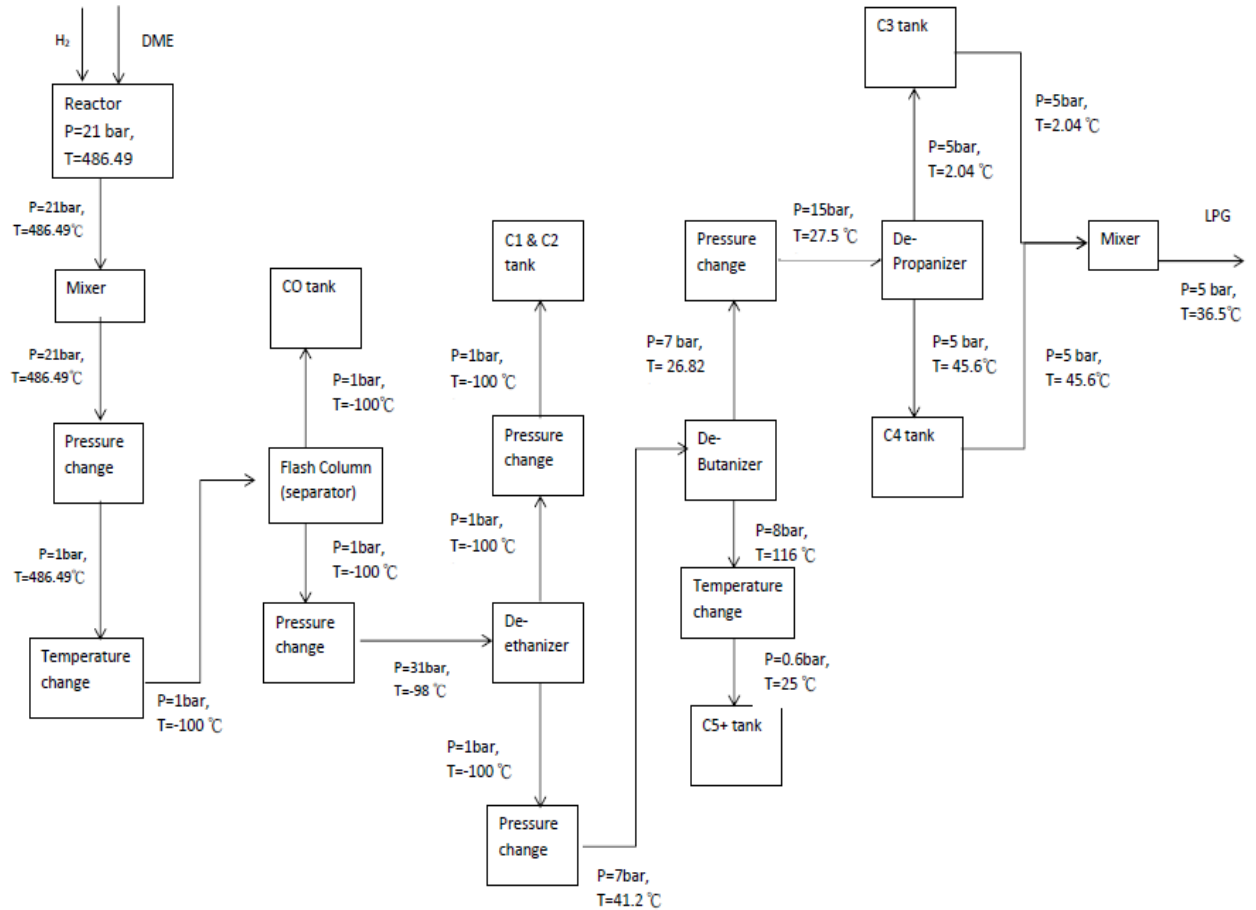
Figure 2.1 : *Process flow diagram*

From the process flow diagram above, we can see that, to eliminate the difference between temperature, pressure and phase, we are using temperature changer which is heat exchanger and for pressure, we are using compressor and pump to change the pressure.

2.3   Process Description

A process that can convert dimethyl ether (DME) to gasoline-range hydrocarbons or to lower olefins over zeolite catalysts (ZSM-5) has been developed with the aid of hydrogen with ratio H_2/DME is 2. The conversion of 100% dimethyl ether to hydrocarbons consists of the olefin formation reaction, ethylene and propylene in particular, and the olefin conversion reaction to form aromatics and paraffin by cyclization and hydrogen transfer reactions respectively as shown below:

$$CH_3OCH_3 \rightarrow C_2\text{-}C_4 \text{ Olefins} \rightarrow \text{Aromatics+ Paraffin} \rightarrow \text{LPG}$$
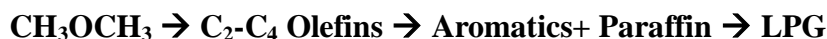
Meanwhile, after the reaction process, further cooling using cooler (E-100) will be needed in order to liquefy the hydrocarbons at -100 ℃. The product yields from the reactor also consist of carbon monoxide which needed to be removed first before entering the conventional distillation column. Thus, another separator (V-100) will be placed just before the conventional distillation columns in order to separate the hydrogen and also pump (P-100) are necessary because the pressure of the hydrocarbons will be increased after the cooling process. The process consists of mainly three distillation columns. In the distillation part where hydrocarbons will be split according to their thermodynamics properties. Initial feeds from both the separation units from the well and from the dehydration unit are mixed before entering the De-Ethanizer. In order to meet the operating pressure of De-Ethanizer, mixed feed (1 bar, -100 ^{o}C) is expanded through pump (P-100) to 31 bar.

In the De-Ethanizer column, methane and ethane are expected to be separated and flow through the top of the column. Since there is no requirement to liquefy methane and ethane, especially in small amounts, these components will be kept in vapor phase and additional condenser is unnecessary. The boiling point of mixed methane and ethane, in this case at pressure 18 bar, is -52.55^{o}C, which means that refrigerant is required to condense the gas. For economic reasons, refrigerant will not be used in this process.

In this De-Ethanizer, 100% methane and ethane in mixed feed can be separated and leaves the top of the column at 31 bar and 4.39 ^{o}C. The heavier hydrocarbons other than ethane leave the column as bottom product in liquid phase at 31.35 bar and 103.67 ^{o}C. Next it is sent to debutanizer column.

Since the bottom product of De-Ethanizer composed of propane (55.12 % of mass fraction) and butane (34 % of mass fraction), De-Butanizer is used instead of De-Propanizer. So that smaller columns will be used for next extraction. This is not only for economic reason, but also for efficiency of the separation as it is easier to extract the product this way and less duty will be required for reboiler E-RB2.

Before entering De-Butanizer column, the De-Ethanizer bottom product is expanded from 31.35 bar to 7 bar using several pressure relieve valve (RV-100,101,102) and fed into De-Butanizer as liquid feed. This feed is fractionated into mixed propane and butane as overhead product and heavier hydrocarbons (C_5+) as bottom product. Then the overhead product is totally condensed in the condenser by heat exchange with cooling water, and condensate is collected in reflux drum. The cooling water is sea water with temperature 10°C. The reflux drum should be used in order to prevent cavitation on the pump due to vapor phase. The De-Butanizer reflux and product pump P-101 serve the dual purpose of supplying reflux to the column.

About 100% propane and 99% butane can be recovered from the feed at the overhead column product. This stream leaves the column at 7 bar and 36.36 ^{o}C and is sent to De-Propanizer column to separate propane and butane. Meanwhile, the bottom product composed of pentane and heavier hydrocarbons will be stored as natural gasoline. Since the bottom product has high temperature (111.12^{o}C), it will be cooled by heat exchanger E-CL-100 before being sent to the storage tank.

Propane and butane stream is expanded from 7 bar to 15 bar and enters the De-Propanizer as mixed-phase feed. The De-Propanizer separates propane as overhead product and butane as bottom product. Sea water is used for cooling. Condensate is collected in De-Propanizer column reflux drum D-102. De-Propanizer reboiler heat is supplied by reboiler E-RB3. Reboiler circulation is aided by reboiler circulation pump.

It is about 99.9% of propane can be recovered as top product and 99.9% of butane is recovered as bottom product. The butane product is field grade butane which is composed of

41% i-butane and 56% n-butane. After De-Propanizer, the product (LPG fractions) will be mixed according to LPG standards so that it can be commercialize.

2.3.1 Reactor
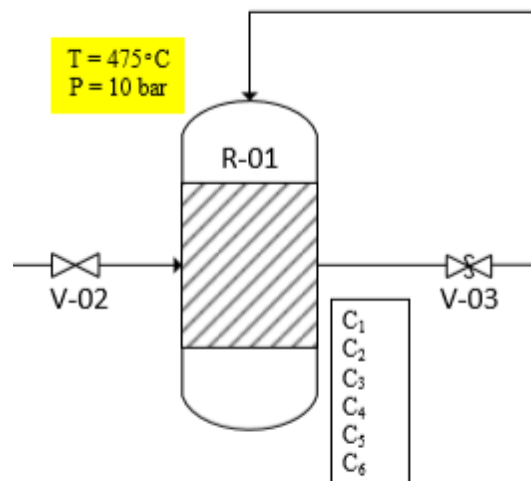
Figure 2.2: *Schematic diagram of Fixed Bed Reactor*

A DME Conversion Experimental Process Unit has been designed and set up which includes a fixed bed reactor. This type of reactor move uniformly and the production of by-product is reduced.

(R-01) Operating condition:

- Operating Temperature : 475 ^{o}C
- Operating Pressure : 10 bar
- Overall Conversion : 100%
- Catalyst : 0.5% Pd-β-243

2.3.2 Conventional column
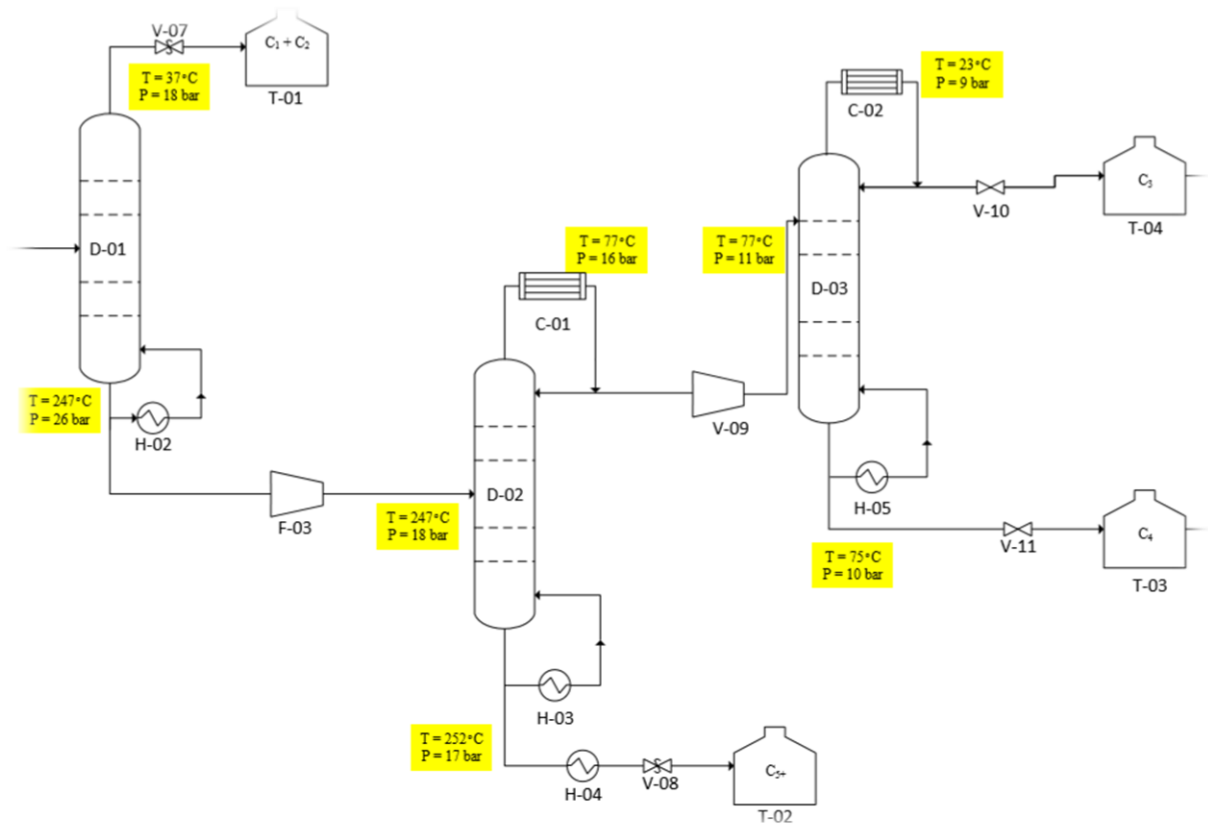
Figure 2.3*: Schematic diagram of Conventional Column (D-01), (D-02) & (D-03)*

Deethanizer

After selected the feed conditioning process, next step is defining the process in the deethanizer column as well as converging the column. In order to minimize the cost, it is avoided to use refrigerant. Therefore, there will be no condenser that is used at the overhead column stream as shown in Figure 3. Methane and ethane as top products will be separated in gas phase. By deleting the condenser there will be no reflux stream at the top of column.

First step to do after defined stream connection is selecting the operating pressure at the top and the bottom of the column. As initial guess, 26.9 bar pressure was used based on reference. Since the feed pressure is 310 bar, expansion valve is added in order to meet the operating pressure of the column. Afterwards select the parameter which will be used as column specification to converge the deethanizer column. In this case, since condenser is not used, there will be only one column specification needs to be defined to converge the column. Selecting operating pressure at the top and the bottom of the column is very important to separate the methane and ethane. The lower the pressure the more the vapour phase. The operating pressure has to be selected so that most methane and ethane flow to overhead column and keep propane and heavier hydrocarbons as the bottom product. The best result is to select operating pressure 18 bar at the top stage and 26 bar at the bottom stage. Next step is selecting the column specification. As mentioned previously, there is only one column specification in modelling the deethanizer column. In this case, the bottom product flow rate is selected since it is the recommended and default specification for the column. Other specification can also be selected, for example component fraction at the top or bottom stage, but it would be very difficult to converge the column. It depends on the composition of the feed. In this case, the composition of methane and ethane in the feed is quite less, so that it is very difficult to remain all the propane and heavier hydrocarbons in the bottom stage.

Practically, some trials have to be done to find the best result. As well as selecting operating pressure and column specification, defining the number of stages is also quite important. The more the number of stages means the better the separation process. In the other hand, more number of stages means higher column is required and the cost will be more expensive. Therefore, the number of stages has to be selected the optimum one considering both separation efficiency and cost. As initial guess, the number of stages is defined based on reference.

In order to find the optimum number of stages, the temperature profile of each tray would help. As well as the column converged by using initial guess of number of stages, it can be reduced to the optimum one by considering the temperature profile of each tray. At the point where the column is no longer converged, the previous number is selected as the optimum one. In this case, the optimum number of stages for deethanizer is 18 stages. After converged the column and find the optimum one based on the separation efficiency and economic consideration, the simulation is continued to modelling the debutanizer column.

Debutanizer

As mentioned previously in process description, for economic reason, debutanizer will be used after deethanizer instead of depropanizer, so that the next separation will use smaller column. Debutanizer is used to separate butane and lighter hydrocarbons at the top and heavier hydrocarbons at the bottom. Basically, the steps in modelling the debutanizer column are almost the same like modelling deetanizer. The differences are in selecting the column specifications. Since both condenser and reboiler are used, there are 2 column specifications to be defined in order to converge the column. In this case, components recovery of the top product is selected as column specification. As it is expected to separate propane and butane at the top and heavier hydrocarbons at the bottom, so that components recovery of the propane and butane as the top product are selected.

Even though most propane and butane can be recovered as the top product, there is one specification should be considered, that is reboiler duty. It is possible to have 99.99% of propane and butane at the top, but more reboiler duty will be required. Thus, the fraction of components recovery should be selected by considering minimum reboiler duty in order to minimize the cost for the steam.

As well as modelling deethanizer, selecting operating pressure and number of stages are also done in modelling debutanizer. Initial guess value for both specification are also based on reference then have some trial to find the optimum value.

Depropanizer

Depropanizer is used after debutanizer column in order to separate propane and butane. The modelling steps are almost the same with debutanizer. Both condenser and reboiler are used as shown. In this column, propane liquid is produced as overhead product and butane as bottom product. Modelling the depropanizer column is simpler since it only has 2 compositions in the feed. Components recovery is also used as column specification. There is no significant reboiler duty difference in changing the recovery fraction value. Selecting

operating pressure and number of stages are also done using initial guess value from reference. Some trials still have to be done to find the optimum value.

Reboiler and Condenser

Reboiler and condenser are parts of the columns. Reboiler is used to supply heat into the column, so that the lighter hydrocarbons will be vaporized and go up to the top stage. Steam is used to heat up a part of bottom product and recycle it into the column. Meanwhile, condenser is used to condense the overhead vapor so the liquid product will be produced which partially recycled into the column as reflux.

In simulation, as part of a distillation column, both reboiler and condenser are defined based on requirement in order to meet product specifications.

Recirculation Pumps

Recirculation pump is used to recycle the reflux and heat supply into the column. It is not shown in the simulation but still need to be calculated for the cost estimation.

2.3.3 Mixer
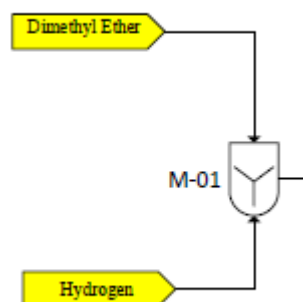
Figure 2.4: *Schematic diagram of Mixer (M-01) & (M-02)*

The dimethyl Ether and hydrogen are mixed well in the M-01 before they entering reactor to ensure that all the reactor is perfectly mix each other. The mixer helps to enhance the reaction in the reactor. While M-02, it function to mixed the final product out from the conventional column, which is propane and butane. From the mixing process, it produced well-mixed quality of LPG.

2.3.4 Compressor
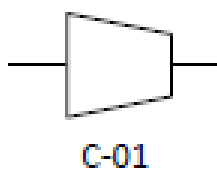
Figure 2.5: *Schematic diagram of Compressor*

This compressor is use to increase the pressure of hydrocarbon mixture from R-01.The pressure increase from 21 bar to 26 bar, the operational pressure of De-ethanizer.

2.3.5 Heat exchanger
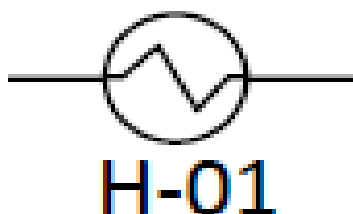
Figure 2.6: *Schematic diagram of Heat Exchanger*

Heat exchanger is used to transfer heat into or from the stream to the process fluid. In most chemical reaction process heat transfer is required to heat up or cool down the stream of fluid based on the process unit conditions. In this plant we used heat exchanger shell and tube type which is cooler.

- Cooler (HX-01): this cooler helps to decrease the temperature of stream from R-01 before feed to D-01. The temperature was decreased from 329 degree C to 247 degree C.
- Cooler (HX-02): This cooler help to reduce the temperature of stream from D-02 to be stored in T-02.
- Cooler (HX-03): This cooler helps to reduce temperature of stream from D-02 to feed to P-03. The temperature reduce from 252 degree C to 77 degree C
- Cooler (HX-04): This cooler helps to decrease the stream from D-03 to P-05. The temperature reduced from 75 degree C to 23 degree C.

2.3.6   Pump

This is isothermal pump increasing the pressure of stream by moving the fluids. Pump is identical to compressor but differ only in the phase involves. Pump is dealing only with liquid phase while compressor is only for gas phase.
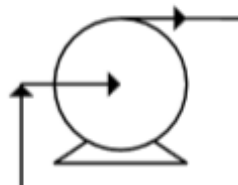
Figure 2.7*: Schematic diagram of Pump*

- P-01: Helps to pump the fluid in liquid phase from D-01 to E-01.
- P-02: Helps to pump the fluid in liquid phase from D-02 to E-02
- P-04: Helps to pump fluid from D-03 to E-03.

2.3.7 Pressure Relive Valve

The pressure relief valve (PRV) is a type of valve used to control or limit the pressure in a system or vessel which can build up for a process upset, instrument or equipment failure, or fire. The pressure is relieved by allowing the pressurised fluid to flow from an auxiliary passage out of the system. The relief valve is designed or set to open at a predetermined set pressure to protect pressure vessels and other equipment from being subjected to pressures that exceed their design limits.

CHAPTER 3

MANUAL MASS AND ENERGY BALANCE

3.1   Introduction

Material and energy balance seems to be an important factor in dealing with the process synthesis and flow sheeting. It should be done first before the simulation being carried out because the result of simulation is only used as a checking material to the manual material and energy balance.

3.2   Manual Material Balance

Procedure for Material Balance for the entire calculation was briefly describe here. For a given description of a process, the value of several process variables, and a list of quantities to be determined:

1. Draw a flowchart, and fill in all given variable values. Choose a basis of a calculation an amount or flow rate of one of the process streams. If one such quantity given, it is usually most convenient to use it as the basis of the calculation; all subsequently calculated quantities will then correctly scaled. If several amounts of flow rate are known, always use them as the basis. If no amount or flow rate is specified, assume an amount of a stream with a known composition and write the assumed quantity on the flowchart. In our proposed manual calculation, product – MEG is used as basis at 100,000 metric ton per annum. The scale up factor was multiply to it and other calculated quantities to achieve the production rate.

2. Label unknown stream variables on the chart, particularly mass or molar flow rates and mass or mole fractions of stream components. Try to incorporate given relationships between unknown quantities in labelling.

3. Do the problem bookkeeping. Count unknowns and relations among them. If the numbers are equal, lay out the problems solution.

4. Convert known stream volumes or volumetric flow rate to mass or molar quantities, using tabulated densities or gas laws.

5. If a mixed mass and mole units for a stream are given (e.g., a total mass flow rate and component mole fractions or conversely), convert all quantities to one basis or the other methods.

6. If any information is given in the problem statement that is not been used in labelling the flowchart, translate it into equations in the variables defined in step 3.

7. Write material balance equations. If no reaction occurs and there are N species present, write at most N balances, either on each species or on total mass or involve the fewest unknowns are written first. At this point you should have as many equation as unknown; if not, either the problems relationship between the variables has been forgotten. Solve the equations derived in step 7 and 8 for the unknown quantities to be determined. When the value of an unknown has been calculated, write it on the flowchart immediately and then substitute this value in any equations in which the unknown appeared.

8. If a stream quantity or flow rate was given in the problem statement and another value Q was either taken as a basis or calculated for this stream, scale the process by the ratio Qg/Qc to obtain the final results. Where Qg is the targeted flow rate; Qc is the basis flow rate used in calculation

The general equation used in the whole manual calculation is:

Input + Generation – Output – Consumption = Accumulation

Where,

Input = total mass enters through system boundary.

Generation = total mass-produced within the system.

Output = total mass leaves though system boundary.

Consumption = total mass consumed within the system.

Accumulation = total mass build up within the system.

The basis for assumption for our material balance is:

1. No leakage in the pipes and vessels in the system.
2. Catalyst used in reactor does not contribute in mass.
3. All the components in the system behave as ideal condition.
4. The total input of any substance to a compressor, pump, mixer or heat exchanger is assumed equal to the total output of the substance where no reaction occurs.
5. All calculations are done in kmol/hr.

3.3   Manual Energy Balance

The procedure for doing the manual energy balance can get from Elementary Principles Of Chemical Processes, page 312-315. The calculation is based the first law of thermodynamics where kinetic energy differences, potential energy differences and work done on the system by its surroundings are zero. So, the law become

$$Q = \Delta H = \Sigma n_o \hat{H}_o - \Sigma n_i \hat{H}_i$$

with reference conditions: the elements that constitute the reactants and products at 25°C and the non-reactive molecular species at any convenient temperature. Each enthalpy calculation is difference depends on the phase for that stream. If the phase is in vapour form, specific enthalpy is calculated by If the stream is in the liquid-vapour mixture form, the specific enthalpy can be calculated by using the following equation:

$$\Delta H_v\,(T,y) = \Delta H_f + \Delta H_T = \sum_k y_k\, H_{f,k}(T_1) + \sum_k y_k \int_{T_1}^{T_2} Cp_k\,(T)dT$$

If the phase is in liquid form, the specific enthalpy can be calculated by using:

$$\Delta H_v\,(T,y) = \Delta H_f + \Delta H_T = \sum_k x_k \left(H_{f,k}(T_1) + \int_{T_o}^{T} Cp_k\,(\tau)d\tau - \Delta H_{vap}^k(T) \right)$$

Where:

$$\Delta H_{vap}^k(T) = \Delta H_{vap}^k(T_b)\,(\frac{(T_c^k - T)}{(T_c^k - T_b)})^{0.38}$$

If the stream in liquid-vapour mixture forms, the specific enthalpy can be calculated by using the following equation:

$$\Delta H(T, z) = \phi \Delta H_v(T, y) + (1 - \phi)\Delta H_L(T, x)$$

Here, some assumptions have been made also. They are:

1. The process system is an open system.
2. Adiabatic process where no heat loss to the environment.
3. All the liquid and gas heat capacities are constant at certain temperature and the value of heat capacities from simulator can be taken directly by setting the simulator.
4. All the enthalpies calculated are taken at 25ºC at 1 atm as reference whereas for non-reactive molecular species, the reference state is taken at any convenient condition.

3.4   Summary of Mass and Energy Balance of each equipment

Included in this sub-chapter are descriptions of the calculation method, diagrams and summary of stream table involving all equipment. All calculation steps are shown in APPENDIX A

3.4.1   Reactor

Calculation of reactor mass balance is done based on the method developed extent of reaction (Felder and Rousseau, 2005). This method is used to determine amount of main product (Hydrocarbon). From that calculation, heat duty of the reactor is calculated using heat of formation method. A heating jacket is added to maintain the temperature of the reactor since the reaction is exothermic.
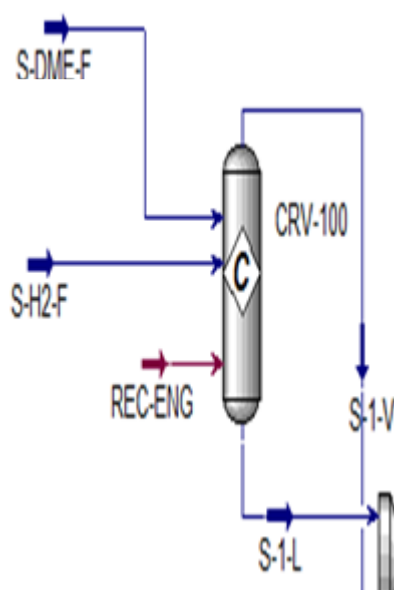
Figure 1*: Schematic diagram of Fixed Bed Reactor (CRV-100)*

Table 1: Stream table for Reactor CRV-100

Stream Tag	S-DME-F	S-H2-F	S-1-V	S-1-L
Flowrate (kmol/hr)	14.0915	4.6065778	11.782724	12.782724
Mole Fraction				
DME	33.25467	-	-	-
H2	-	66.74533	-	-
C1	-	-	0.0252	0.0154
C2	-	-	0.0201	0.0152
C3	-	-	0.5287	0.4822
i-C4	-	-	0.1903	0.2080
n-C4	-	-	0.1400	0.1553
C5	-	-	0.0835	0.1082
C6	-	-	0.0091	0.0139
CO	-	-	0.0030	0.0017
Temperature (°C)	486.49			
Pressure (bar)	21			
Heat Duty (kJ/hr)	1807931.02			

3.4.2 Separator (Flash Column)

Calculation of flash column follows closely the recommended Clifford-Rice method where vapor-liquid equilibrium of the components mixture is calculated. After that, the next step is to calculate the volatility of each components and this will correlate with the overall mass balance of flash column where feed equals to addition of vapor flowrate and liquid flowrate. Since this is an ideal system, there is no heat duty for both columns.
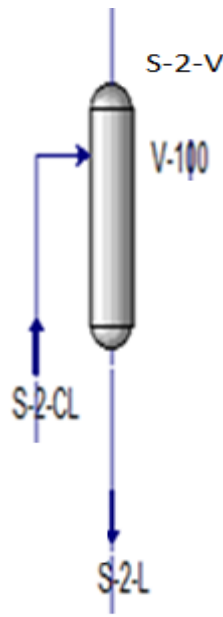
Figure 2: *Schematic diagram of separator V-100*

Table 4: Stream table for Separator V-100

Stream Tag	S-2-CL	S-2-L	S-2-V
Flowrate (kmol/hr)	13.86	13.809944	0.050094
Mole Fraction			
C1	0.0250	0.0248	0.0869
C2	0.0207	0.0208	0.0000
C3	0.5266	0.5285	0.0000
i-C4	0.1398	0.1403	0.0000
n-C4	0.1899	0.1906	0.0000
C5	0.0852	0.0855	0.0000
C6	0.0095	0.0095	0.0000
CO	0.0033	0.0000	0.9129
Temperature (°C)	-100		
Pressure (bar)	1		
Heat Duty (kJ/hr)	1245350.7	-2053557	-11214.1

3.4.3   Conventional column

The calculation of material and energy balance from conventional column is following McCABE-THIELE METHOD from book transport process and separation principle by Christie John Geankopolis, fourth edition. The separation is vapour-liquid separation process that involving countercurrent multiple components. So that, the volatility value of each component is determined. The light and heavy key component also need to be identified. The feed flowrate (F), distillate flowrate (D) and raffinate flowrate (R) can be determined via material balance. Therefore, some assumption are needed such as purity of top stream, basis for feed flowrate, the operating temperature and pressure of column.

Then, the heat duty of condenser and reboiler are devised after the top product purity are known and minimum reflux determined. The number of trays can be calculated from constructing operating and equilibrium line graph.
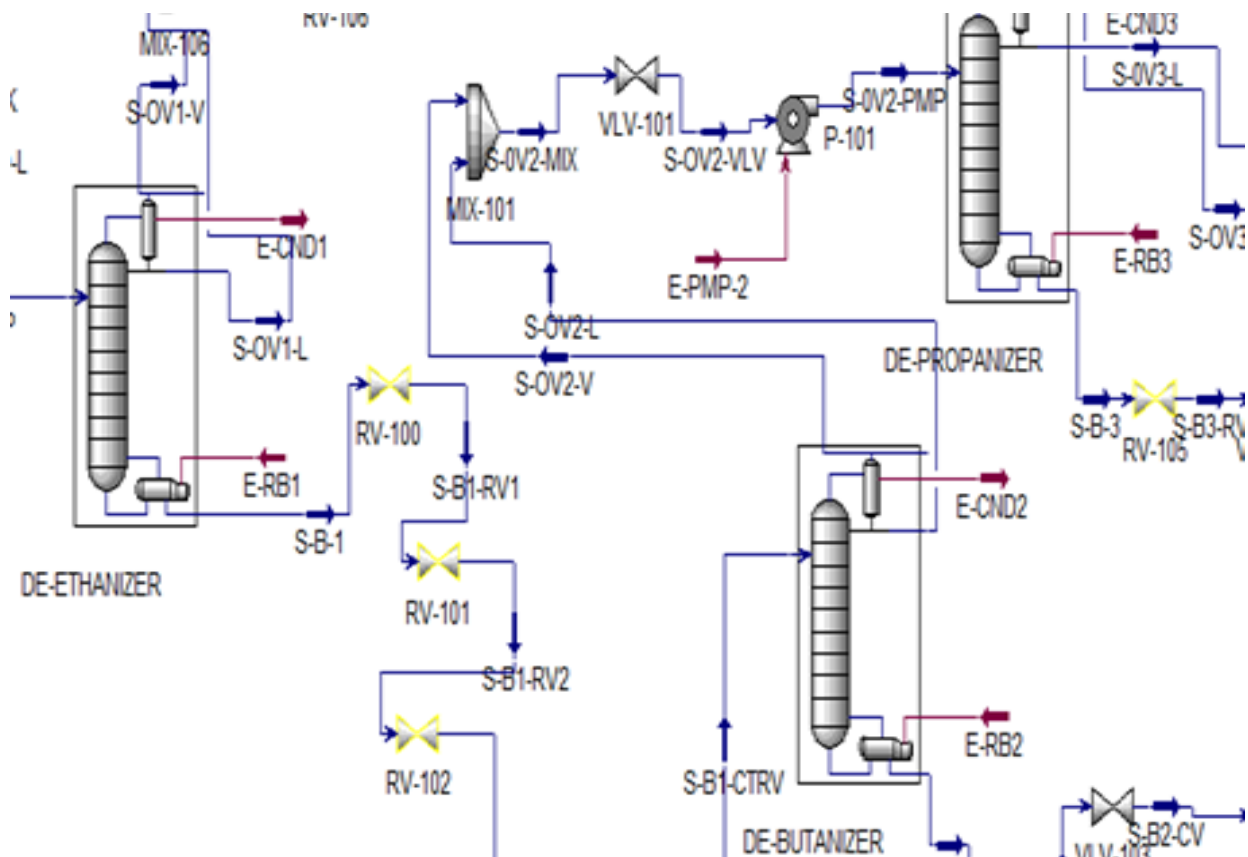
Figure 3: Schematic diagram of Conventional Column (De-Ethenizer), (De-Butanizer) & (De-Propanizer)

Table 5: Stream table for Conventional column De-ethenizer

Stream tag	Feed	Distillate	Bottom
Flowrate (kmol/hr)	13.8099	0.6984	13.1115
Mole fraction			
C1	0.3421	0.3421	0.0000
C2 (L)	0.2869	0.2840	0.0029
C3 (H)	7.2987	0.0723	7.2264
I-C4	1.9376	0.0000	1.9376
N-C4	2.6320	0.0000	2.6320
C5	1.1809	0.0000	1.1809
C6	0.1317	0.0000	0.1317
CO	0.1317	0.0000	0.0000
Temperature c	84.0000	4.7800	104.3900
Pressure (bar)	31.0000	31.0000	31.3500
Heat duty (kj/hr)	4713.9385	-1210963.5547	2033327.9783

Table 6: Stream table for Conventional column De-butanizer

Stream tag	Feed	Distillate	Bottom
Flowrate (kmol/hr)	13.1115	11.7827	1.3288
Mole fraction			
C1	0.0000	0.0000	0.0000
C2 (L)	0.0029	0.0029	0.0000
C3 (H)	7.2264	7.2264	0.0000

I-C4	1.9376	1.9363	0.0013
N-C4	2.6320	2.6054	0.0266
C5	1.1809	0.0117	1.1692
C6	0.1317	0.0000	0.1317
CO	0.0000	0.0000	0.0000
Temperature c	58.7400	36.3600	111.1200
Pressure (atm)	7.0000	7.0000	8.0000
Heat duty (kj/hr)	2033327.9783	-907899.4816	695109.7535

Table 7: Stream table for Conventional column De-Propanizer

Stream tag	Feed	Distillate	Bottom
Flowrate (kmol/hr)	11.7827	7.1761	4.6066
Mole fraction			
C1	0.0000	0.0000	0.0000
C2 (L)	0.0029	0.0029	0.0000
C3 (H)	7.2264	7.1534	0.0730
I-C4	1.9363	0.0192	1.9171
N-C4	2.6054	0.0007	2.6048
C5	0.0117	0.0000	0.0117
C6	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000
Temperature c	36.3600	28.6500	77.8900
Pressure (atm)	15.0000	11.0000	12.0000
Heat duty (kj/hr)	1666.4227	-538237.6692	585009.9769

3.4.4 Pump

Pump is an equipment which increases pressure of a fluid stream. Mass balance for pump is straight forward as the cooler and heater before it, where mass in is equal to mass out. However when heat exchanger has constant pressure to calculate the change in temperature, a pump operates at different conditions such as constant, temperature, adiabatic and other assumptions. Since the correlations between enthalpy, volume, temperature and pressure of a fluid can be quantified (Smith et al., 2005), this pump operates at constant volume and isothermal. Thus calculating the pump's work are based on these conditions.
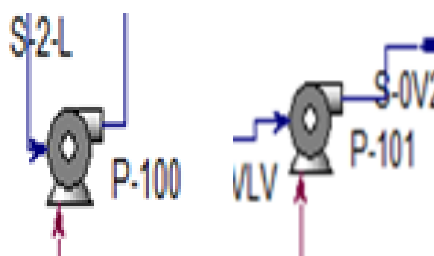
Figure 4: Schematic diagram of pump (P-100) & (P-101)

Table 8: Stream table of P-100 & P-101

P-100			P-101		
Stream Tag	S-2-L	S-2-PMP	Stream Tag	S-2-L	S-2-PMP
Flowrate (kmol/hr)	13.809944	13.809944	Flowrate (kmol/hr)	11.7827	11.7827
Mole Fraction			Mole Fraction		
C1	0.0248	0.0248	C1	0.0000	0.0000
C2	0.0208	0.0208	C2	0.0002	0.0002
C3	0.5285	0.5285	C3	0.6133	0.6133
i-C4	0.1403	0.1403	i-C4	0.1643	0.1643
n-C4	0.1906	0.1906	n-C4	0.2211	0.2211
C5	0.0855	0.0855	C5	0.0010	0.0010
C6	0.0095	0.0095	C6	0.0000	0.0000
CO	0.0000	0.0000	CO	0.0000	0.0000
Temperature (°C)	-100		Temperature (°C)	36.3600	37.8500
Pressure (bar)	1	31.35	Pressure (bar)	7.0000	15.0000
Heat Duty (kJ/hr)	4713.938533		Heat Duty (kJ/hr)	1666.4227	

3.4.5 Heat Exchanger

The description for mass balance calculation of heater is the same as cooler with the exception of a positive heat duty.
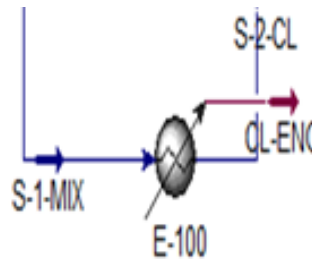
Figure 5: *Schematic diagram of Heat Exchanger E-100*

Table 9: Stream table for **Heat Exchanger E-100**

Stream Tag	S-1-MIX	S-2-CL
Flowrate (kmol/hr)	13.86	13.86
Mole Fraction		
C1	0.0250	0.0250
C2	0.0207	0.0207
C3	0.5266	0.5266
i-C4	0.1398	0.1398
n-C4	0.1899	0.1899
C5	0.0852	0.0852
C6	0.0095	0.0095
CO	0.0033	0.0033
Temperature (°C)	486.49	-100
Pressure (bar)	21	1
Heat Duty (kJ/hr)	1245350.746	

3.4.6 Mixer

To calculate mass balance for mixer are relatively straight forward since the reaction takes place inside the reactor.
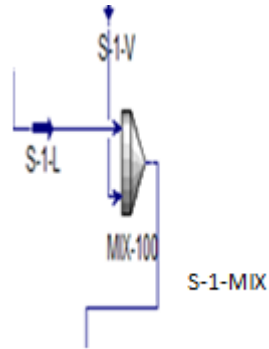
Figure 6: Schematic diagram of Mixer (MIX-100)

Table 10: Stream table for MIX-100

Stream Tag	S-1-V	S-1-L	S-1-MIX
Flowrate (kmol/hr)			
Mole Fraction			
C1	0.0252	0.0154	0.0250
C2	0.0201	0.0152	0.0207
C3	0.5287	0.4822	0.5266
i-C4	0.1903	0.2080	0.1398
n-C4	0.1400	0.1553	0.1899
C5	0.0835	0.1082	0.0852
C6	0.0091	0.0139	0.0095
CO	0.0030	0.0017	0.0033
Temperature (°C)	486.4921665		
Pressure (bar)	21		
Heat Duty (kJ/hr)	-771733.9	0	-771733.871

Figure 7: Schematic diagram of Mixer (MIX-106)

Table 11: Stream table for MIX-106

Stream Tag	S-0V1-V	S-0V1-L	S-0V1-MIX
Flowrate (kmol/hr)	2.54E-04	0.6981609	0.698414912
Mole Fraction			
C1	0.8569	0.3563	0.4899
C2	0.1039	0.3663	0.4066
C3	0.0165	0.2745	0.1035
i-C4	0.0000	0.0000	0.0000
n-C4	0.0000	0.0000	0.0000
C5	0.0000	0.0000	0.0000
C6	0.0000	0.0000	0.0000
CO	0.0227	0.0029	0.0000
Temperature (°C)	4.78		
Pressure (bar)	31		
Heat Duty (kJ/hr)	-20.61513	-77982.49	-78003.1025
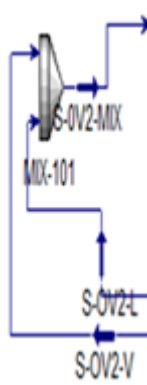
Figure 8: Schematic diagram of Mixer (MIX-101)

Table 12: Stream table for MIX-101

Stream Tag	S-0V2-V	S-0V2-L	S-0V2-MIX
Flowrate (kmol/hr)	1.809E-07	11.7827	11.7827
Mole Fraction			
C1	0.0000	0.0000	0.0250
C2	0.0001	0.0002	0.0207
C3	0.8098	0.6133	0.5266
i-C4	0.0974	0.1643	0.1398
n-C4	0.0927	0.2211	0.1899
C5	0.0000	0.0010	0.0852
C6	0.0000	0.0000	0.0095
CO	0.0000	0.0000	0.0033
Temperature (°C)	36.36		
Pressure (bar)	7		
Heat Duty (kJ/hr)	-0.01988	-1553201	-1553201.09
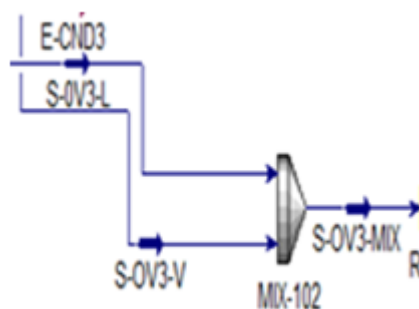
Figure 9: Schematic diagram of Mixer (MIX-102)

Table 13: Stream table for MIX-102

Stream Tag	S-0V3-V	S-0V3-L	S-0V3-MIX
Flowrate (kmol/hr)	2.026E-05	7.1761262	7.176146467
Mole Fraction			
C1	0.0000	0.0000	0.0000
C2	0.0001	0.0000	0.0004
C3	0.9983	0.9961	0.9968
i-C4	0.0003	0.0008	0.0027
n-C4	0.0013	0.0030	0.0001
C5	0.0000	0.0000	0.0000
C6	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000
Temperature (°C)	28.65		
Pressure (bar)	11		
Heat Duty (kJ/hr)	178.18439	-859967	-859788.784
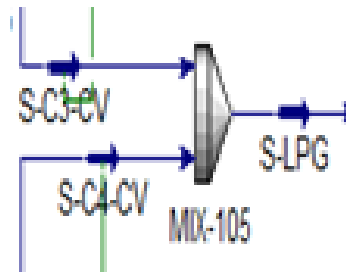
Figure 10: Schematic diagram of Mixer (MIX-105)

Table 14: Stream table for MIX-105

Stream Tag	S-C3-CV	S-C4-CV	S-LPG
Flowrate (kmol/hr)	7.1761465	4.6065778	11.78272424
Mole Fraction			
C1	0.0000	0.0000	0.0000
C2	0.0000	0.0000	0.0002
C3	0.9961	0.0001	0.6133
i-C4	0.0008	0.5782	0.2211
n-C4	0.0030	0.4217	0.1643
C5	0.0000	0.0000	0.0000
C6	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000
Temperature (°C)	28.65	77.89	77.89
Pressure (bar)	11	12	
Heat Duty (kJ/hr)	1805500.4	0	1805500.445
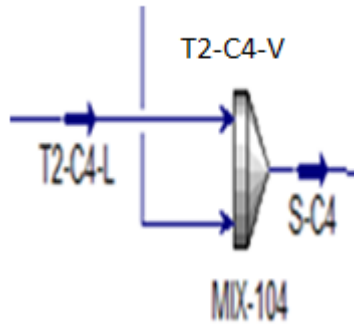
Figure 11: Schematic diagram of Mixer (MIX-104)

Table 15: Stream table for MIX-104

Stream Table	T2-C4-V	T2-C4-L	S-C4
Flowrate (kmol/hr)			
Mole Fraction			
C1	0.0252	0.0154	0.0250
C2	0.0201	0.0152	0.0207
C3	0.5287	0.4822	0.5266
i-C4	0.1903	0.2080	0.1398
n-C4	0.1400	0.1553	0.1899
C5	0.0835	0.1082	0.0852
C6	0.0091	0.0139	0.0095
CO	0.0030	0.0017	0.0033
Temperature (°C)	486.4921665		
Pressure (bar)	21		
Heat Duty (kJ/hr)	1805500.4	0	1805500.445

Figure 12: Schematic diagram of Mixer (MIX-100)

Table 16: Stream table for MIX-100

Stream Tag	T1-C3-V	T1-C3-L	S-C3
Flowrate (kmol/hr)			
Mole Fraction			
C1	0.0252	0.0154	0.0250
C2	0.0201	0.0152	0.0207
C3	0.5287	0.4822	0.5266
i-C4	0.1903	0.2080	0.1398
n-C4	0.1400	0.1553	0.1899
C5	0.0835	0.1082	0.0852
C6	0.0091	0.0139	0.0095
CO	0.0030	0.0017	0.0033
Temperature (°C)	486.4921665		
Pressure (bar)	21		
Heat Duty (kJ/hr)	1805500.4	0	1805500.445

3.5   Materials balance comparison between Manual Calculation and Aspen HYSYS Simulation Software.

      Process simulation has been done on this LPG Plant assuming steady state condition by using Aspen HYSYS Simulation Software. This software is a comprehensive process modelling tool used by the world's leading oil and gas producers, refineries, and engineering companies for process simulation and process optimization in design and operations of a plant. In this section we will compare the material and energy balances between manual calculation and by using simulation. The comparison is evaluated by using the following formula :

$$Error = \frac{|Simulated\ value - Manual\ value|}{Simulated\ value}\ x100\%$$

      The simulation allows us to easily do steady state and dynamic process modelling in process design which is to determine the process conditions required to produce the desired product, the process equipment design which is to size the equipment required to produce the desired product, the process design optimization which is to determine the optimum configuration of equipment and maximize energy recovery, and the process optimization which is to determine changes to the current operating conditions that can either reduce operating costs or Increase production.

3.5.1   Material Balances Comparison between Manual Calculation and Aspen Simulation.

Table 17: Material Balances Comparison between Manual Calculation and Aspen Simulation.

Stream	Total Manual	Total Aspen	Error (%)
S-DME-F	14.0915	14.1000	0%
S-H2-F	28.2830	26.8730	-5%
S-1-MIX	13.8600	13.9576	1%
S-2-CL	13.8600	13.9576	1%
S-2-V	0.0501	0.1228	59%
S-2-L	13.8099	13.8348	0%

S-2-MP	13.8099	13.8348	0%
S-0V1-MIX	0.6984	0.7622	8%
S-B-1	13.1115	13.0727	0%
S-B1-CRTV	13.1115	13.0727	0%
S-B-2	1.3288	1.2933	-3%
S-0V2-MIX	11.7827	11.7794	0%
S-0V2-VLV	11.7827	11.7794	0%
S-0V2-PMP	11.7827	11.7794	0%
S-0V3-MIX	7.1761	7.1960	0%
S-B-3	4.6066	4.5834	-1%
S-C3-CV	7.1761	-	
S-C4-CV	4.6066	-	
S-LPG	11.7827	-	

3.5.2 Energy Balances Comparison between Manual Calculation and Aspen Simulation

Table 18: Energy Balances Comparison between Manual Calculation and Aspen Simulation

Major Equipment	Heat Duty Manual, Q	Heat Duty Aspen, Q	Error (%)
Reactor (CRV-100)	1805500.445	2.10E+06	14%
Heat Exchanger	1.81E+06	1293036.866	-40%
Condenser (E-CND1) (E-CND2) (E-CND3)			
	-1210963.555	-1557104183	100%
	-907899.4816	-834392.9577	-9%
	-538237.6692	-550761.9683	2%
Rebolier (E-RB1) (E-RB2) (E-RB3)			
	2033327.978	1557434006	100%
	695109.7535	697640.908	0%
	585009.9769	593308.0119	1%
Pump P-100 P-101			
	4713.938533	4195.548755	-12%
	1666.422691	1209.370593	-38%

3.6 Justification

There are several justifications about the percentage error occurring between manual calculation and simulation such as the error between manual and HYSYS due to the reference of each other are different. For manual calculation, the reference based on 25℃, 1 atm. Besides that, the thermodynamic formula for such as specific heat for manual and HYSYS are different. Furthermore, for manual calculation, the temperature and pressure was set to be same for input and output. Meanwhile, in HYSYS, the temperature and pressure was set in inlet stream only.

CHAPTER 4

PROCESS ENERGY INTEGRATION, OPTIMIZATION & WASTE MANAGEMENT

4.1   Process energy integration

This chapter will cover the several algorithmic approaches to produce a network of process integration that satisfy the heating, cooling, and energy consumption of a process. At the beginning, the minimum usages of heating and cooling utilities are calculated when exchanging heat between hot and cold streams occur in order to determine the minimum energy requirement (MER). The temperature-interval method (TI) was proposed to achieve this objective of energy optimization, thus minimize the operating cost. The comparison between the utility consumption from the base case process flow heat exchanger and after the determination of heat integration was done before proceed to design the heat exchanger network (Seider et.al, 2010).

This will then be followed by the design of the heat exchanger network using the Pinch Technology Method. The Pinch Technology include deciding the position of heat exchanger in a network together with the assumption on overall heat transfer coefficient and it was developed as a systematic tool to design heat recovery network for maximum energy recovery. The procedures are based upon fundamental thermodynamics and a set of experienced-based heuristic for heat exchanger network synthesis is quick and easy to apply.

Finally, the number of heat exchanger can be reduce since the heat and power integration seeks to utilize the energy in high-temperature streams that need to be cooled and to heat up the cold streams.

4.1.1 Pinch Technology

Pinch Technology has provided a deep understanding of the overall process scenario. In Pinch Technology, the application of temperature-interval method that developed by Linhoff and Flower following by the work from Hohmann was done. A good initialization for the heat exchanger network design should be to apply this method to the hot and cold stream and utilize systematic procedure to determine the minimum utility requirement to all possible HEN given the heating and cooling requirement for process stream and the minimum approach temperature in heat exchanger ΔTmin (Seider et.al, 2010)

The first step in TI method is to adjust the source and target temperature using ΔTmin. The temperature of the hot streams reduced to by ΔTmin while the temperature of cold stream remains unchanged. The adjustment will bring both streams to common frame of reference. Therefore when energy balance involve hot and cold stream conducted, the calculation for heat transfer perform with at least a driving force of ΔTmin.

Then, the adjusted temperature were rearranged in rank-order, starting with the To the highest temperature. These orders are used to create a cascade of temperature intervals within the energy balance that are carried out. Each interval will shows enthalpy difference between the energy to be remove from the hot stream and the energy to be consume by the cold stream in that interval.

Initially no energy is assumed to enter the first interval from a hot utility as the heat duty. Hence the enthalpy is available for the next interval as residual. This will continue to the next interval until the final residue is obtained which become the cold duties.

The pinch design targets are;

a) Minimum heating requirement (usually the stream rate)

b) Minimum cooling requirement (usually the cooling water rate)

c) Minimum number of units

d) Minimum heat exchanger network

Firstly, identification of the hot and cold streams in the plant before the heat integration is carried out. All the streams in the plant must be calculated to obtain the maximum energy recovery for the plant. The value of heat flow rate, FCp obtained from energy balance summary where it is equal to $\Delta H/\Delta T$. There are several benefits of HEN have been identified are;

a) Energy saving (15% - 90%)

b) Capital saving (up to 25%)

c) Improved flexibility

d) New design and retrofit

4.1.2 Rule of Thumb for Pinch Technology Method

ΔTmin varies the relative heat recovery status. A simple criterion to develop a pinch technology design is as below;

a) Divide the problem at the pinch into separate problems that is the hot utility and cold utility.

b) Design for separate problems is started at the pinch and moving away. The streams are conveniently matched to the heat transfer streams.

c) Temperatures feasibility requires constraints on the CP value to be satisfied for matches between streams at the pinch. To avoid infeasible form (violate the minimum

temperature difference at the heat exchangers) of the matcher, CPH (hot stream) should be smaller than or equals to CPC (cold stream) above the pinch while CPH should be more than or equals to CPC below the pinch. However, the CP inequalities only apply when both ends are match to the pinch point. Minimize the number of the units.

d) Stream splitting is necessary to provide a flexible matching design.

4.1.3   Stream Table Extraction

Pinch analysis begins with the process of screening plant flow sheet for a set of hot and cold streams having potential for heat interchange. These stream data can be obtained directly from the simulation results. The extracted stream (which excluding reboiler that involved the use of utilities) are tabulated in the following table. The value of heat flow rate, CP is obtained from energy balance summary.

Table 4.1: Summary of Hot and Cold Streams (Tabulated from Simulation Results)

Heat Exchanger	Stream	T_{supply} (°C)	T_{target} (°C)	F_{Cp} (kW/K)
HX1	H1	486.5	-100	0.754873533
HX2	H2	116.89	25	0.04260705
HX3	C1	25	486.5	115.748725

4.1.4   Stream Pairing

From Problem Table and Heat Cascaded Diagram in Appendix B, the heat utility, cold utility and no pinch temperature can be obtained.

The Pinch Temperature, Tpinch hot = 35∘C

The Pinch Temperature, Tpinch cold = 25∘C

The minimum heating requirement, QHmin = 53073.7244 Kw

The minimum cooling requirement, QCmin = 102.334 kW

The stream pairing target is to achieve the minimum heating value, Qcmin = 102.334 kW and minimum cooling value, Qhmin = 53073.7244 kW with minimum number of heat exchanger.

For the above pinch design, the following rules must be fulfilled:

Only can pair stream where the cold stream FCp is greater than hot stream FCp, CpC ≥ CpH
 Start with the hot stream with biggest FCp

 Maximize the heat exchange load and tick off the streams that enthalpy balance is made.

Fill in the rest of the exchanger in the same manner, checking for temperature feasibility for heat exchange

 Calculate the QHmin

Above pinch can only have heater; cooler is prohibited in the above pinch region

The temperature difference of the hot and cold stream must be greater or equal to $\Delta Tmin = 10 \circ C$

Criteria to be considered in designing below pinch:

a) Follow Cph bigger than Cpc for pinch match
b) Start with the cold stream with biggest FCp
c) Maximize the heat exchange load and tick off the stream that is balanced with heat
d) Fill in the rest, checking for temperature feasibility for heat exchange.

4.1.5 Maximum Energy Recovery

$Maximum\ Energy\ Recovered = Total\ of\ cold\ stream\ enthalpy - minimum\ heating\quad requirement$

$Maximum\ Energy\ Recovered = Total\ of\ hot\ stream\ enthalpy - minimum\ cooling\ requirement$

Table 4.2: *Summary of the Heat Equipment Duties before and after MER*

Stream	Before MER	After MER
Hot	0KW	53073.7224 KW
cool	52971.3904 kW	102.334 KW

Energy saving (hot utilities)
% heating reduction = (53408.0366 KW - 53073.7244 KW) / 53408.0366 KW
$= 0.63\ \%$

Energy saving (cold utilities)

% cooling reduction = (446.64 KW – 102.334 KW)/ 446.64 KW

= 77.09 %

4.1.6 Summary for Heat Exchanger Network

T pinch hot = 35∘C

T pinch cold = 25∘C

QCmin = 102.334 kW

QHmin = 53073.7244 kW

Saving on QCmin = 77.09%

Saving on QHmin = 0.63 %

4.2 Process Optimization

Optimization is a process to maximize profit or minimize energy utilisation. In order to achieve this target, various method can be employed by the management and technical side. Besides that, optimization also is carried out to determine the optimum operating condition and physical design that could achieve best process performance of each unit operation without compromising all the constraint that are pre-determined.

4.2.1 Calculation Method of Optimization

In this plant we will optimized reactor equipment by determining the optimal reactor duty, used for the reaction. The main reasons to develop optimal reactor duty are:

1. Energy saving and
2. Reducing heating cost.

The decision variable being selected as below:

Objective Function = Minimize Reactor Duty = Min. Q (kJ) = Cp (Steam) (ΔT_1) + Cp (Feed) (ΔT_2)

Variable;   Product Fraction

Constraint;   Reactor operating Temperature;   475 ^{o}C < $T_{reactor}$< 0 ^{o}C

Table 4.3: Optimization of Reactor

Temperature, ^{o}C	Product Conversion	Reactor Duty, kJ/h
486.4893345	0.859069596	2100000
93.984965	0.863504024	1300000
83.26839222	0.874115191	1200000
68.89833112	0.775784018	1100000
21.75674261	0.4676652	1000000
-37.97626333	0.136044916	900000
-105.7586847	0.003707928	800000
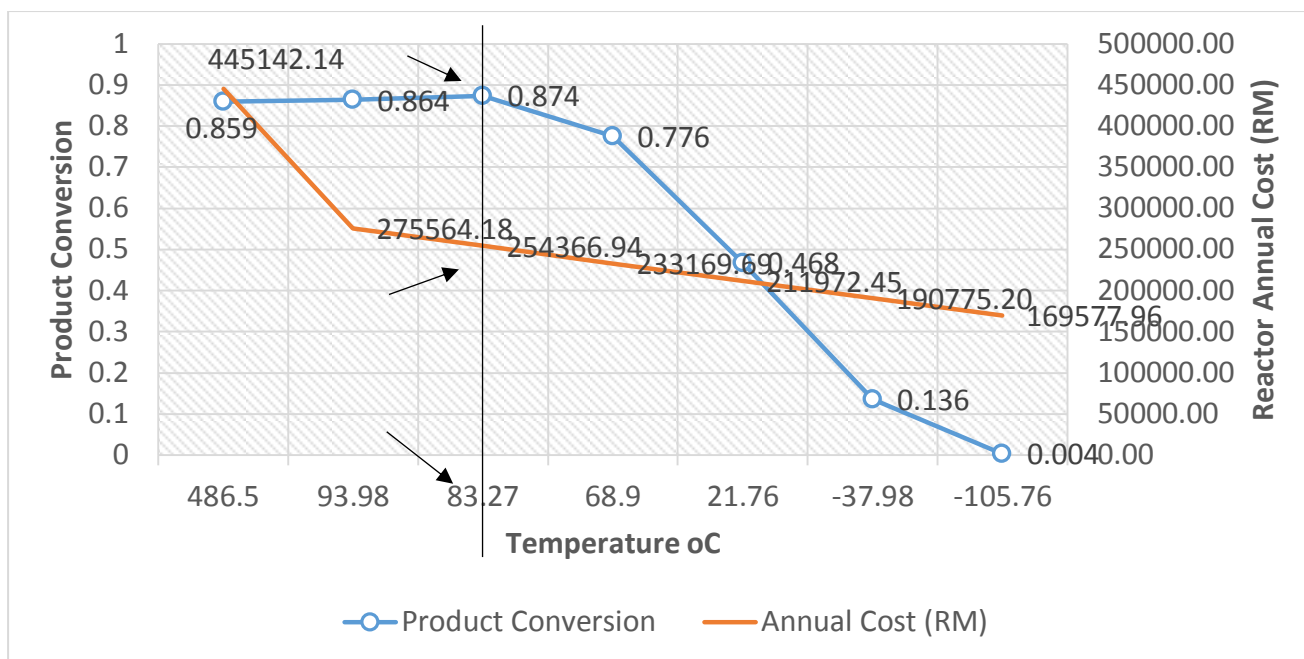
TEMPERATURE	PRODUCT CONVERSION	ANNUAL COST (RM)
83.27	0.874	250,000

Figure 4.1: Optimization of Reactor CRV-100

From figure 4.1, the optimum condition obtain at 7°C with product fraction 0.393 with reactor duty 995000 kJ/h. Optimizing of CRV 100 saved of the reactor duty.

$$\frac{2100000-995000}{21000000} \text{ x } 100= 52\%$$

4.3 Waste treatment and management.

Waste treatment refers to the activities required to ensure that waste has the least practicable impact on the environment. In many countries various form of waste treatment are required by law. In this plant, we react dimethyl ether with hydrogen and will produce carbon monoxide. Since this gas is toxic, it has to be treated before release to the atmosphere. Thus, we need to find the standard for industries emitting gaseous and air emission in Malaysia to propose the suitable treatment. Other than treating, carbon monoxide also can be reuse for other industries such as hot iron oxides and metallic ores industry.

4.3.1 Gaseous emission.

Emission standards are requirements that set specific limits to the amount of pollutant that can be released into the environment and maximum amount of a specific pollutant allowed to be discharged into the atmosphere from a single fixed or mobile source. According to "Environmental requirement: A Guide for Investors" set by Department of Environment:

1. For industries emitting gaseous and air emission, they are required to comply with the following air emission on standards for the control of air pollution and gaseous emissions :
    a. Stack gas emission standards from the Environmental Quality (Clean Air) Regulations 1978 *(Appendix J1)*
    b. Recommended Malaysian Air Quality Guidelines (Ambient Standards) *(Appendix J2)*
2. All industrial projects subject to EIA shall be designed and operated using Best Available Techniques (BAT) in achieving a high and acceptable level of protection for the environment.

STACK GAS EMISSION STANDARDS
[Extract from Environment Quality (Clean Air) Regulations 1978]

Pollution	Emission Sources		Standards
1. Dark Smoke*			
	(1.1)	Solid Fuel Equipment to Facilities	Ringlemann Chart No.2
	(1.2)	Equipment using other types of fuel	Ringlemann Chart No.1
2. Dust	(2.1)	Facilities used for the heating of metal other than Cold Blast Foundry Cupola	0.2 gm/Nm^3
	(2.2)	Facilities discharging dust containing asbestos and free silica	0.12 gm/Nm^3
	(2.3)	Portland Cement Manufacturing:	
		(2.3.1) Kiln	0.2 gm/Nm^3
		(2.3.2) Clinker, cooler, grinder, others	0.1 gm/Nm^3
	(2.4)	Asphalt concrete/bituminous mixing plant:	
		(2.4.1) Stationary Plant	0.3 gm/Nm^3
		(2.4.2) Mobile Plant	0.4 gm/Nm^3
	(2.5)	Other source	0.4 gm/Nm^3
3. Metal and Metallic Compound			
3.1 Mercury	Industry		0.01 gm/Nm^3
3.2 Cadmium	Industry		0.015 gm/Nm^3
3.3 Lead	Industry		0.025 gm/Nm^3
3.4 Antimony	Industry		0.025 gm/Nm^3
3.5 Arsenic	Industry		0.025 gm/Nm^3
3.6 Zinc	Industry		0.1 gm/Nm^3
3.7 Copper	Industry		0.1 gm/Nm^3

RECOMMENDED MALAYSIAN AIR QUALITY GUIDELINES
(Ambient Standards) (at 25^{0}Celsius and 101.13 kPa)

Pollutant and Method	Averaging Time	Malaysia Guidelines (ppm)	Malaysia Guidelines (ug/m_3)
Ozone AS 2524	1 Hour	0.10	200
	8 Hour	0.06	120
Carbon # Monoxide AS2695	1 Hour	30	35
	8 Hour	9	10
Nitrogen Dioxide AS 2447	1 Hour	0.17	320
Sulfur Dioxide AS 2523	10 Minute	0.19	500
	1 Hour	0.13	350
	24 Hour	0.04	105
Particles TSP AS 2724.3	24 Hour		260
	1 Year		90
PM$_{10}$ AS 2724.6	24 Hour		150
	1 Year		50
Lead AS 2800	3 Month		1.5

Recommended Malaysian Secondary Guidelines

Pollutant and Method	Averaging Time	Malaysia Guidelines (mg/m2/day)
Dustfall AS 2724.1	1 year	133

#mg/m^3

4.3.2 Carbon monoxide

Incomplete oxidation of carbon results in the production of carbon monoxide. Anthropogenic sources (those associated with the activities of human beings) include motor vehicles, fossil fuel burning for electricity and heat, industrial processes, solid waste disposal, and miscellaneous burning of such things as leaves and brush. Approximately 600-1250 Tg of CO are released by these sources (IPCC, 1995). Motor vehicles account for more than 60 percent of the emission.

No significant change in the global atmospheric CO level has been observed over the past 20 years. Yet the worldwide anthropogenic contribution of combustion sources has doubled over the same time period. Because there is no apparent change in the account for the missing CO. the two most probable are:

1. Reaction with hydroxyl radicals to form carbon dioxide

2. Removal by soil microorganism

The emissions of carbon monoxide from our plant have been calculated, and we find that our emissions are lower than the DOE standard, 9ppm/hr. Therefore, the carbon monoxide emits from our plant no need to be treated.
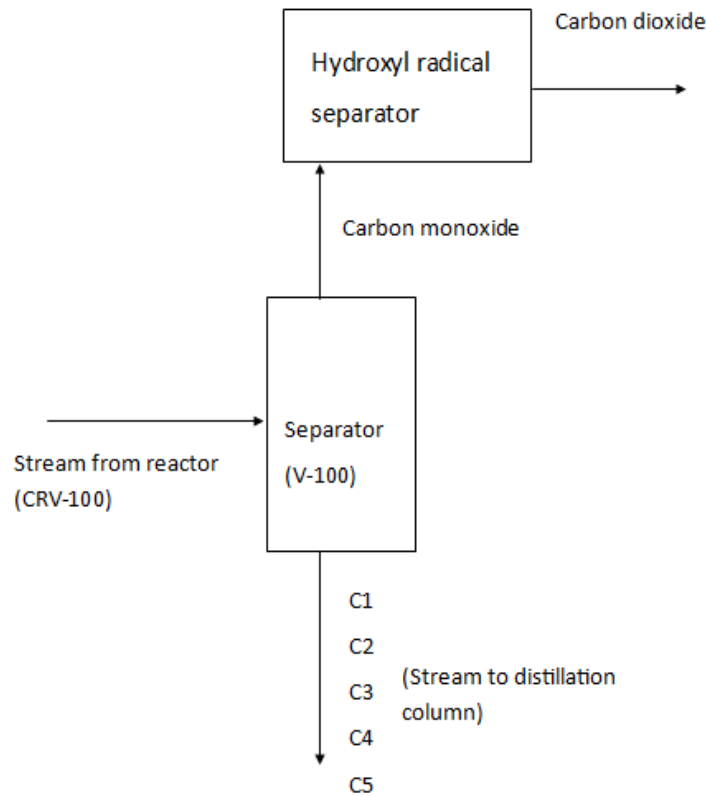
Figure 4.2: Hydroxyl radical separator for treating carbon monoxide

4.4   Water treatment.

Water treatment is any process that makes water more acceptable for a specific end-use. Water treatment will removes contaminant or reduces their concentration so that the water becomes fit to the end user. Two of the main processes of industrial water treatment are boiler water treatment and cooling water treatment. Water that use for boiler and cooling tower need to be treated to reduce the scaling and corrosion. Scaling in the boiler will reduce the heat production efficiency.

CHAPTER 5

EQUIPMENT SIZING AND MECHANICAL DESIGN

5.1   Introduction

Equipment sizing is crucial part in plant design. It affects profitability of a plant where the choice of material used and size of the units determine feasibility of a plant. For this purpose, the calculations are based on the throughput into the equipment and its corresponding operating parameters. The results will be needed for the cost correlations to estimate the equipment cost. In the following section, equipment sizing for all units in our LPG plant will be conducted.

In this LPG production plant, it involves many major types of equipment in the set-up of the plant. Generally, the units comprises of reactor, distillation column, flash column, mixer, pump, heat exchanger and compressor. All the calculated data are tabulated in Appendix C.

In this following section also shall discuss the general design procedure. In addition, the assembly drawings showing mechanical engineering design of each process vessels and its corresponding equipment specification sheet are also presented in the Appendix C, the detailed calculation for mechanical engineering design of equipment is provided in Appendix C.

In this plant design, equipment sizing and costing will be referred to *Product and Process design Principles.*

Equipment	Quantity
Heat Exchanger	2
Reactor	1
Flash Column	1
Distillation Column	3
Mixer	2
Pump	2

5.1.1 Flash Distillation

One of the simplest separation processes commonly employed is flash distillation. In this process, part of a liquid feed stream vaporizes in a flash chamber or part of a vapor feed condenses, and the vapor and liquid in equilibrium with each other are separated. The more volatile component will be more concentrated in the vapor. Usually a large degree of separation is not achieved.

The vapor travels upward at a design velocity, which minimizes the entrainment of any liquid droplets in the vapor as it exits the top of the vessel. The ways to calculate flash tank size includes in Appendix C. The sizing and costing results summarized in Table below.

Table 5.1: Sizing summary for flash column

Equipment Specification Sheet	
Item no.	V-100
Operating Condition	
Temperature (℃)	-100
Pressure (atm)	1
Specification	
Design type	Vertical vessel
Material of construction	Stainless steel
Vessel	
Diameter (m)	0.3003

Height (m)	1.2012
Volume (m^3)	0.5234

5.1.2 Pump

Pump selection is made on the flow rate and head required, together with other process considerations, such as corrosion or the presence of solids in the fluid. Pump generally can be divided into two different types, which are dynamics pumps and positive displacement pump. Positive displacement itself can be separate to two other types known as reciprocating pump diaphragm pump and centrifugal pump. The centrifugal pumps are the major types that used in the chemical plant nowadays. Centrifugal pumps are used so extensively and for such a wide variety of services that need for standardization of dimensions and operating characteristic has long been evident.

A general guide to the selection, installation and operation of pumps for the processes industries is given by Davidson and von Bertele (1999) and Jandiel (2000). The selection of the pump cannot be separated from the design of the complete piping system. The total head required will be the sum of the dynamic head due to friction losses in the piping, fittings, valves and process equipment, and any static head due to differences in elevation. The pressure drop required across a control valve will be a function of the valve design. Sufficient pressure drop must be allowed for when sizing the pump to ensure that the control valve operates satisfactorily over the full range of flow required. If possible, the control valve and pump should be sized together, as a unit, to ensure that the optimum size is selected for both

Table 5.2 : *Sizing summary for pump*

equipment specification sheet		
Item no	PUMP P-100	PUMP P-101
OPERATING CONDITION		
Temperature (°C)	-98	26

Pressure Inlet (atm)	1	6.9
Pressure Outlet (atm)	30.9	14.8
SPECIFICATIONS		
Driver type	Motor	Motor
Pump Efficiency, %	75	75
Capacity [m3/h]	1.0368	1.1338
Molar Flow Rate (kmole/hr)	13.835	11.779
Material of Construction	Carbon steel	Carbon steel
Shaft Power	1.5090	0.3280

5.1.3 Heat Exchanger

5.1.3.1 Sizing

Sizing The heat exchanger really applies to all types of equipment in which heat is exchanged but is often used specially to denote equipment in which heat is exchanged between two process streams. Exchangers in which a process fluid is heated or cooled by a plant service stream are referred to as heaters and coolers. If the process stream is vaporized the exchanger is called a vaporized if the stream is essentially completely vaporized; a reboiler and condenser if associated with a distillation column. The general equation for heat transfer across a surface is:

$= U\ A\ \ddot{A}Tm$

where,

Q = heat transferred per unit mean temperature difference, °C

U = overall heat transfer coefficient, W/m2.°C

A = transfer area, m2

ÄTm = mean temperature difference, °C

The principal types of heat exchanger used in chemical process and allied industries are:

a) Double pipe heat exchanger: the simplest type, used for cooling and heating.

b) Shell and tube exchangers: used for all applications.

c) Plate and frame exchangers (plate heat exchangers): used for heating and cooling.

d) Air cooler: cooling and condensers.

e) Direct contact: cooling and quenching.

Since shell and tube exchanger is the most commonly used of heat transfer equipment used in chemical and allied industries, it was chosen as the exchanger type in the design, the advantages of this type are:

a) The configuration gives a large surface area in a small volume.

b) Good mechanical layout: a good shape for pressure operation

c) Uses well-established fabrication techniques.

d) Can be constructed from a wide range of materials.

e) Easily cleaned.

f) Well-established design procedures.

The designing procedures are based on the Kern's and Bell's method (Sinnott, 1996). The design produce is as below:

a) Define the duty: heat transfer rate, fluids flow rate and temperature.

b) Collect the fluid physical required: density, viscosity and thermal conductivity.

c) Decide the type of exchanger to be used.

d) Select the trial value for the exchanger to be used.

e) Calculate the mean temperature different, ΔT_{lm}.

f) Calculate the area required.

g) Decide the exchanger layout.

h) Calculate the individual coefficients.

i) Calculate the overall coefficient and compare with the trial value.

If the calculated value differs from the estimate value, substitute the calculated value and return to step (6)

Below is the summary table for sizing of heat exchanger:

85

Table5.3: Sizing summary for heat exchanger E-100.

EQUIPMENT SPECIFICATION SHEET			
Identification : Heater			
Code : E-100			
Function : To heat up the feed stream to reactor			
Design specification		Tube side	
Type of heat exchanger	Shell and tube (split ring floating head)	Inlet temperature (^{O}C)	25
		Outlet temperature (^{O}C)	486.5
		Shell side	
	1 shell : 2 tube passes	Inlet temperature (^{O}C)	500
		Outlet temperature (^{O}C)	353.23
Material of construction	Stainless steal		
Tube arrangement	Triangle		
Operating Condition			
Heat transfer area		9.87 m^{2}	
Heat duty		349.82 kW	

Table 5.4: Sizing summary for heat exchanger E-101.

EQUIPMENT SPECIFICATION SHEET			
Identification : Cooler			
Code : E-101			
Function : To heat up the feed stream to reactor			
Design specification		Tube side	
Type of heat exchanger	Shell and tube (split ring floating head)	Inlet temperature (^{O}C)	464.87
		Outlet temperature (^{O}C)	-100
		Shell side	
	1 shell : 2 tube passes	Inlet temperature (^{O}C)	-101.1
		Outlet temperature (^{O}C)	292.4
Material of construction	Stainless steal		
Tube arrangement	Triangle		
Operating Condition			
Heat transfer area		15.26 m^{2}	
Heat duty		345.9 kW	

Table 5.5: Sizing summary for heat exchanger E-102.

EQUIPMENT SPECIFICATION SHEET				
Identification : Cooler				
Code : E-102				
Function : To heat up the feed stream to reactor				
Design specification		Tube side		
Type of heat exchanger	Shell and tube (split ring floating head)	Inlet temperature (^{O}C)	116.89	
		Outlet temperature (^{O}C)	25	
		Shell side		
	1 shell : 2 tube passes	Inlet temperature (^{O}C)	5	
Material of construction	Stainless steal	Outlet temperature (^{O}C)	20.3	
Tube arrangement	Triangle			
Operating Condition				
Heat transfer area		3.19 m^2		
Heat duty		1 kW		
Heat transfer coefficient		241.65 W/m^2.^{o}C		

5.1.4 Reactor

Reactors are the main component of a chemical plant where the reactions take place to convert the raw material to desired products. There are various parameters need to be take into account when designing the reactors, namely type of reaction, concentration, temperature, pressure, phase and catalyst used. There are also few types of reactors such as the batch reactor, continuous stirred tank reactor and plug flow reactor. For the production of LPG from DME, PFR was used to ensure high conversion of reactants.

In its most basis form a plug flow reactor consists of a cylindrical tube filled with piping system. Reactants flow through the pipe and reacted with other reactants and are converted into products. They may be regarded as the workhorse of the chemical industry with respect to the number of reactors employed and the economic value of the materials produced.

The catalyst constituting in the fixed bed will generally be employed in one of the following configurations multiple horizontal pipe. Other constraints are including pressure drop or adequate fluid distribution. In additional to the shell and tube configuration some of the possibilities for heat transfer to or from plug flow reactors include the use of internal heat exchanger, annular cooling spaces or cooling thimbles and circulation of a portion of the reacting gases through an external heat exchanger.

Conditions of Reactor CRV- 100

For our plant, the type of reactor that we choose is plug flow reactor.

The characteristics and assumptions for CRV-100 are stated as below:

1. Multi-phase reaction
2. Isothermal and adiabatic operation
3. Material of construction: Stainless steel (Biegler et al., 1992)
4. Steady state condition

Table 5.6: Sizing summary for reactor CRV-100

EQUIPMENT SPECIFICATION SHEET	
Item NO	Reactor CRV-100
Identification	Plug Flow Reactor (PFR)
Function	Production of hydrocarbon from DME
OPERATING CONDITION	
Temperature ($^\circ$C)	475
Pressure (bar)	2.1
Flowrates	706.21 kg/hr
Amount of catalyst, W	1.36 kg/pipe
SPECIFICATIONS	
Design type	Horizontal cylinder
Material of construction	Stainless steel : Austenitic (ATI 3161Ti)
No. Of unit	10
Residence time	1.94 hr

VESSEL EACH REACTOR	
Diameter,m	3.52
Length, m	14.08
Volume, m^3	137.14
No. of pipes	7618

5.1.5 Distillation Column

Distillation is the most common method of separating homogenous mixtures. The separation of liquid mixtures by distillation depends of the differences in volatility between the components. The greater the volatility of a component compared to other component, the easier it is to be separated. Vapour flows up to the top column and the liquid flows counter currently down to the bottom of the column. The vapour is brought into intimate contact with the liquid on every plate.

The design of distillation column is based on the typical design procedures as stated in the *Chapter 19 of product and Process Design Principles, by Seider et al (2010).* Each of the design variable is set and calculated from the design formula and based on the recommended values. By checking the key performance factors, the design parameter have been revised or otherwise determined. Some designs parameter are obtained from the simulation.

5.1.5.1   Plate Spacing

Plate spacing is the most important part for determining the overall height of the column. Plates spacing ranges from 0.15 m to 1 m are normally used. The spacing chosen depends on the column diameter and operating conditions. For columns above 1 m diameter, plate spacing 0.3 to 0.6 m will normally be used and 0.45 m can be taken as initial. This will be revised as necessary.

5.1.5.2   Column Diameter

The principle factor in determining the column diameter is the vapour flow rate. The column diameter can be calculated by calculating the top and the bottom net area at its maximum volumetric flow rate. The velocity is normally between 70 to 90% of what which could cause flooding

5.1.5.3   Height of Column

The height of column in the distillation column is calculated by knowing the number of actual stages. Minimum number of stages is calculated from Fenske Equation. The height of the column can be calculated by multiplying the number of the actual stages with tray spacing value.

Table 7,8 and 9   shows the calculation summary of distillation column. For further information, please refer to Appendix C.

Table 5.7: Sizing summary for de-ethanizer

EQUIPMENT SPECIFICATION SHEET	
Item no	D-100 (de-ethanizer)
Function	To seperate ethane from the hydrocarbon
OPERATING CONDITION	
Temperature,℃	-98
Pressure, bar	31
No.of tray	15
Thickness,inch	2.41
Weight	83906.71 lb
SPECIFICATION	
Design type	Vertical fabrication vessel
Material of construction	Stainless steel
VESSEL	
Diameter, inch	113.38
Height, m	14.4

Table 5.8: *Sizing summary for de-butanizer*

EQUIPMENT SPECIFICATION SHEET	
Item no	D-101 (de-butanize)
Function	To seperate butane from the hydrocarbon
OPERATING CONDITION	
Temperature,℃	104.39
Pressure, bar	7
No.of tray	15
Thickness,inch	0.66
Weight	91190.94 lb
SPECIFICATION	
Design type	Vertical fabrication vessel
Material of construction	Stainless steel
VESSEL	
Diameter, inch	118.11
Height, m	15.0

Table 5.9: *Sizing summary for de-propanizer*

EQUIPMENT SPECIFICATION SHEET	
Item no	D-102 (de-propanizer)
Function	To seperate propane from butane
OPERATING CONDITION	
Temperature,℃	37.85
Pressure, bar	15
No.of tray	15
Thickness,inch	1.03
Weight	91190.94 lb
SPECIFICATION	
Design type	Vertical fabrication vessel
Material of construction	Stainless steel
VESSEL	
Diameter, inch	118.11
Height, m	15

CHAPTER 6

PLANT COSTING

6.1   Introduction

After designing and sizing equipment in Chapter 5, the equipment and plant costing is an essential in chemical plant designed. It involves monetary investment that needs to be poured into the plant to access the economy performance of the plant. In the end of Chapter 1, it provides the rough estimation of economic potential in process selection. However, it different from that, this economy analysis is more precise and detail as it includes depreciation, costing and sizing for equipment and so on. All the calculation for this chapter is in Appendix E, a pre-calculation for constructing cash flow diagram and the discounted cash flow.

6.2   Equipment Costing

Costing for a plant is very important factor that should be done in good manner because it's determined the whole plant profitability beside ensure that the plant run perfectly without any losses and problems during the operation. In LPG production plant, it involves a total of 11 units in the set-up of the plant. Generally, the units comprises of reactor, distillation column, pump, heat exchanger and tank.

Table 6.1: Equipment costing

EQUIPMENT	NAME	COST (RM)
Reactor	CRV-100	RM 2,851,990.64
Distillation column	D-100	RM 6,137,043.41
	D-101	RM 7,469,667.30
	D-102	RM 7,354,678.41
Pump	P-100	RM 675,307.84
	P-101	RM 283,407.20
Heat Exchanger	E-100	RM 222,956.78
	E-101	RM 52,892.27
	E-102	RM 276,211.85
Flash Column	V-100	RM 42,169.99
LPG tank		RM 3,349,500.68
Total Bare Module Cost (C$_{TBM}$)		**RM 28,715,826.37**

6.3   Grass-Roots Capital (GRC)

Grass-roots capital plays a pioneering role in attracting private capital into the industry and plant. This is for the purpose to determine the GRC which is a contribution of bare module cost, contingency and fees as well as auxiliary facilities. According to Ulrich (1984), the contingency and fees is 18% of bare module cost whereas the auxiliary facilities hold 30% of the bare module cost.

Table 6.2: Grass-roots capital calculation

Contingency	$C_C = 0.15C_{TBM}$	RM 4,307,373.96
Fees	$C_F = 0.15C_{TBM}$	RM 4,307,373.96
Total Module Cost	$C_C + C_F + C_{TBM} = C_{BM}$	RM 37,330,574.28
Auxiliary Building	$C_{AB} = 0.04C_{TBM}$	RM 1,148,633
Site Development	$C_{SD} = 0.05C_{TBM}$	RM 1,435,791.32
Offsite Facilities	$C_{OS} = 0.21C_{TBM}$	RM 6,030,323.54
Total Auxiliary Facilities	$0.30\ C_{TBM}$	RM 861,474.79
Grass-roots Capital (GRC)	**GRC = C_{BM} + Auxiliary Cost**	**RM 38,192,049.07**

6.4   Fixed Investment Cost (FIC) and Total Investment Cost (TIC)

Fixed Capital Investment (FCI) defined as financially immobile capital cost. This can be simply analogised as the money, once spent, cannot be quickly converted back into cash or some other asset. This includes direct costs, indirect cost and the GRC. Meanwhile for Total Capital can be represented by the following question.

Total Capital Investment (TCI)   = Fixed Capital Investment (FCI) +

Working Capital + Start up

Working capital is refer to what must be invested to get the plant into productive operation which is money invested before there is a product to sell. Generally, most of the chemical plant used an initial working capital is 112% of the FCI and start-up cost is 8% out of FCI (Ulrich, 1984).

Table 6.3: Total fixed capital and total capital investment

Event		Cost
Onsite		
Purchased Equipment Installation	30% GRC	RM 11,457,614.72
Instrumentation and Control (installed)	15% GRC	RM 5,728,807.36
Piping (installed)	30% GRC	RM 11,457,614.72
Electrical and Material (installed)	8% GRC	RM 3,055,363.93
Offsite		
Building	10 % GRC	RM 3,819,204.91
Yard Improvements	1% GRC	RM 381,920.49
Service Facilities	5% GRC	RM 1,909,602.45
Land	2% GRC	RM 763,840.98
Total	**(T1)**	**RM 38,573,969.56**
Engineering and supervision	5% GRC	RM 1,909,602.45
Construction Expenses	8% GRC	RM 3,055,363.93
Contractor's Fee	1.5% GRC	RM 572,880.74
Contingency	10% GRC	RM 3,819,204.91
Total	**(T2)**	**RM 9,357,052.03**
Total (T)	**T = T1 + T2**	**RM 47,931,021.59**
Fixed Capital Investment (FCI)	FCI = GRC + T	RM 86,123,070.66
Working Capital	12% FCI	RM 10,334,768.48
Start up	8% FCI	RM 6,889,845.65
Total Capital Investment (TCI)		**RM 103,347,684.80**

6.5   Estimation of Operating Labour Cost

For estimation labour cost, it will consider for 330 working days per year with an assumption:

- Average total working period for single operators is 48 weeks/year.
- 3 weeks of vacation are off and sick leave for each operator.

Table 6.4: Operating labour estimation cost

Equipment	Categories	Number of unit	Operators per unit per shift	Operators per shift
Reactor	PFR reactors	1	0.50	0.50
Distillation column	Produce hydrocarbon from DME • Separate ethane • Separate butane • Separate propane	3	0.30	0.90
Pump	Increase pressure	2	0.0	0.00
Heat exchanger	Cooling and heating	3	0.10	0.30
Storage tank	Stored LPG	1	0.00	0.00
Waste	Wastewater treatment	1	2.00	2.00
Total				**3.70**

From the table above:

Assume that there are 3 shifts per day, thus operator required

Operator required per day $\quad$ = 3.70 x 3

$\qquad\qquad\qquad\qquad\quad$ = 11.1

$\qquad\qquad\qquad\qquad\quad$ = 11 people

Average monthly payment for the employee is RM 2,000.00

Thus the labour cost $\qquad$ = 11 x RM 2,000.00 per month

$\qquad\qquad\qquad\qquad\quad$ = RM 22,000.00 per month

Labour cost $\quad$ = RM 264,000.00 per year

6.6 Estimation of Utilities Cost

Utilities involved in this plant are electricity and water. Assuming the plant will operate for 330 working days.

Table 6.5: Total utilities

Equipment	Unit ID	Utility cost per year (RM/yr)
Cooler	E-102	312.00
Heat Exchanger	E-100	108,207.84
	E-100	107,920.80
Pump	P-100	351.22
	P-101	76.34
Material		Utility cost per year (RM/yr)
Water		9,948.47
Total Utilities		**226,816.67**

6.7   Material Costing and Product Sales

Table 6.6: Raw material costing

Price for Raw Material:	
DME (feed)	RM 0.5974 /kg
Hydrogen (feed)	RM 6.48/ kg
Raw Material Usage:	
DME (feed)	649.20 kg/hr @ 5,141,664 kg/year
Hydrogen (feed)	57.02 kg/hr @ 451,598.4 kg/year
Total Annual Cost for Feed	**RM 5,997,987.71**

Table 6.7: Product sales

Product price:	
LPG	RM1.872 /kg
Product Amount:	
LPG	1,219,830,025.491kg/year
Annual sales for LPG	RM 2,283,521,807.72/year
Total Annual Sales	**RM** 2,283,521,807.72**/year**

6.8   Manufacturing Cost

Table 6.8: Manufacturing cost summary

Fix Capital Investment (FCI)	RM 86,123,070.66
Working Capital	RM 10,334,768.48
Start Up Cost	RM 6,889,845.65
Total Capital Investment (TCI)	RM 103,347,684.80

Manufacturing Expenses	Cost (RM/year)	Cost per production (RM/kg)
Direct Production Cost		
Raw Material	5,997,987.71	7.07
Catalyst (0.1% raw material)	5,998.00	48.83
Utilities		
Hot and cold utilities	226,816.67	0.04
Miscellaneous		
Maintenance and repair (2% FCI)	1,722,461.41	0.07
Operating supplies (0.5% of maintenance & repair)	8,612.31	0.00
Operating labor	264,000.00	0.01
Direct supervision & clerical labor (10% of operating labor)	26,400.00	0.00
Laboratory charges (10% of operating labor)	26,400.00	0.00
Patents and royalties (0.01% FCI)	8,612.31	0.00
Indirect production cost		
Local taxes (1% FCI)	861,230.71	0.03
Insurance (1% FCI)	861,230.71	0.03
Plant overhead (50% of operating labor)	132,000.00	0.00
Total Manufacturing Expenses, A_{ME}	**10,616,494.83**	**56.08**

General Expenses		Cost (RM/yr.)	Cost per Production (RM/kg)
Administration Cost	25% of plant overhead	33,000.00	0.00
Distribution & Selling Experiences	8% of FCI	6,889,845.65	0.27
Research & Development	3% of FCI	2,583,692.12	0.10
Total General Expenses, A_{GE}		**9,506,537.77**	**0.37**

$$A_{PC} = A_{ME} + A_{GE}$$

	Cost (RM/yr.)	Cost per Production (RM/kg)
Total Production Cost, A_{PC}	**20,123,032.60**	**56.45**
Depreciation, A_{BD}	8,612,307.06	0.33
Total General Expenses, A_{TE}	**28,735,339.66**	**56.78**

CHAPTER 7

PROCESS CONTROL AND INSTRUMENTATION

7.1   Introduction

Chemical plants are physically large and complex. Process control is essential to control the dynamic state of process. Terminologically, control is to measure, decide and adjust the variable of a process to meet the requirements and certain objectives. The control system and Piping & Instrumentation Diagram are both important because of the arrangement of the processing units are integrated with each other in systematic and rational manner. During in plant operation, these must satisfy several requirements imposed by the designer, process requirements, economical and socially condition for any changes occurred that might deviate it from the set point or become unsteady.

Typically, there were seven major control objectives listed according to Marlin (2000). They are:

i.   Safety
   - To ensure the process variables are operated under safety limits
   - To obtain required temperature and pressure for specified equipment such a reactor and flash column
   - To detect any possible risk during operation and to provide automatic shutdown system or interlock. (Chapter 7 : Process safety analysis)

ii.   Environmental Protection
   - Process control is required to satisfy the various federal and state laws that specify the temperatures, concentrations and flow rates of the effluent from a plant are within certain limits.

iii.   Equipment Protection

- To maintain the operating condition without introducing large fluctuation this will spoil some of the expensive equipment.

iv.   Smooth Operation and Production Rate

- Naturally, the key variables in streams leaving the process should be maintained close to their desired values to prevent disturbances to downstream units.

- To make sure the product produced meet the plant's capacity

v.   Product Quality

- To achieve specified composition specification and purity for outlet stream in reactors, distillation column and so on

vi.   Profit

- The profit margin can be broaden by minimizing the cost of operation cost and increased the production rate

- To minimize the usage of heating and cooling utilities and then cut down the utility cost

vii.   Monitoring and Diagnosis

- It can also serves as measure to monitor and diagnose for prevention or earlier emergency action

- All these seven categories of control objectives must be achieved simultaneously; failure to do so unprofitable or, worse, dangerous plant operation. However for this polyethylene plant, there are a few objectives highlighted, they are:

- To have safer operation plants and to avoid explosion

- To maintain the operational condition of equipment

- To minimize the utilities cost

7.2   Procedures for Control System Design

The typical process control concept is to satisfy the operational objectives. This accomplished through a rational arrangement of equipment (measuring devices, valves, controllers, computers) and human interventions, which together constitute the control system and procedures. The three general classes of needs that a control system is called on to satisfy:

1. Suppressing the influence of external disturbances

2. Ensuring the stability of a chemical process

3. Optimizing the performance of a chemical process

In developing a particular control scheme, control objectives are first defined. Then, critical controlled variables are identified, whereby measured variables, manipulated variables must be decided to conceptualize the control strategies. The following procedure is used to identify and locate the control instruments in the process:

1. Determine and draw the relevant control loops such as temperature control, pressure control, composition control and flow control to confirm the steady state plant operation.

2. Identify the key process variables that to be controlled to achieve the specified product quality. Include controls loops using direct measurement of the controlled variable, where possible.

3. Identify and include those additional control loops required for safe operation, not covered in steps 1 and 2.

4. Decide and show those auxiliary instruments needed for the monitoring of the plant operation by the operators and for trouble-shooting.

5. Decide also on the alarms needed. There are two types of the control system namely feedback control and feed forward control.

6. In feedback control system, that applied where there is a measuring device detecting the output of the process and then a controlling device will compare the measured reading and the process set point, sending signal to the final control element that will manipulate the controlled variables.

7. Unlike the feedback control system, forward control configurations measures the disturbance (load) directly and take control action to eliminate its impact on the process output. Therefore, feed forward controllers have the theoretical potential for perfect control.

7.3 Typical Control Measurement

7.3.1 Level Control

Level control is important in equipments with interface existence. It is to maintain the interface level so that the system may not overflow or dry out during the process. Typically level control is incorporated in the design with the flow out from equipment

7.3.2 Pressure Control

Pressure is crucial for system consist of gas or vapour. It must be maintain at the specific value especially in reactor and distillation so that the equipment can function and reaction can happen. Besides the pressure also cannot be too high as it might cause failure to the equipment or worse cause explosion. This control method depends to the nature of process, for instance, pressure is control via manipulating the flow from or to equipment.

7.3.3 Temperature Control

Temperature control is important to ensure the product achieves the desired composition or specification. In this case, this refers to Ethylene Glycol with MW of 100 000. In addition, temperature control can also minimize the energy required by heater as well as consumed by cooler through controlling the flow of steam and

coolant. The life span of the equipment can be preserved as long as the temperature deviation is small

7.3.4 Flow Control

Flow control is obvious that it involved directly or indirectly the control of other parameters, such as the temperature control in heat exchanger and pressure control in the reactor. To provide constant flow rate to the equipment or to the following processing unit, usually control valve is installed to meet this purpose.

7.3.5 Composition Control

Composition control is important to ensure the purity of the product. This control is implemented in distillation columns. The composition control in this plant used the model predictive control method to readjust the manipulated value to the setpoint.

7.4   Process Control System Hardware

According to Stephanopoulos (1984), the hardware elements that can be established in control configuration include:

 i. Chemical Process:

 It indicates the material equipment together with the physical or chemical operations that occur there.

 ii. Sensor:

 These measuring instruments are used to measure the disturbances, the controlled output variables, or secondary output variables. It represents the behavior of the process. For instance, thermocouple, venture meter and gas chromatographs.

 iii. Transducer:

 It used to convert measurement to physical quantities (such as electrical voltage or pneumatic signal) which can be transmitted easily

 iv. Transmission line:

The line is used to carry the measurement signal from the measuring device to the controller. Sometimes it is equipped with amplifier due to weak signal coming from a far measuring device.

v.   Controller:

This is the hardware element that has "intelligence". It receives the information and decides what action should be taken.

vi.   Final Control Element:

It implements in real life the decision taken by the controller. For example control valve.

vii.   Recording Elements:

They are used to provide a visual demonstration of how a chemical process behaves.

7.5   Piping and Intrumentation Diagram (P&ID)

The process flowsheet exhibits the arrangement of the major equipment and their interconnection. All in all, it is a description of the process's nature. On the other hand, the Piping and Instrumentation Diagram shows the engineering details of the equipment, instrument, piping, valves and fittings as well as their arrangements. Sometimes it is referred as Engineering Flow Sheet or Engineering Line Diagram.

7.6 control system for major equipment

7.6.1 Reactor (CRV-100)

The control system of the plug flow reactor, CRV-100 is designed to opeate the reactor optimally and safely. The control objectives are as follows.

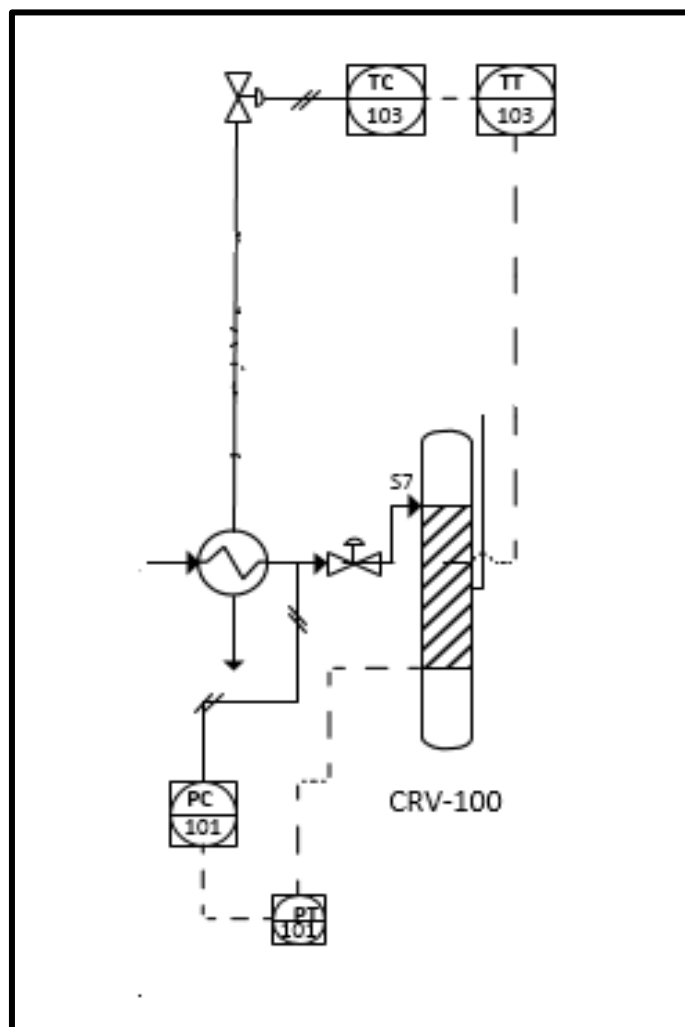Objective: To convert Dimethyl Ether (DME) into hydrocarbon

Control system:

1. To maintain the temperature at $475℃$, so that reaction can occur.
2. To control the pressure of the reactor at 2.1 bar as operating pressure

Table 7.1: Control system for reactor, CRV-100

Control Variable	Measured Variable	Manipulated Variable	Disturbance	Configuration of Controller	Set Point
Temperature of reactor	Temperature of reactor	Flowrate of cooling utilities stream	Ambient temperature	Feedback control	T=475℃
Pressure of reactor	Pressure of reactor	Inlet flow rate		Feedback control	P=2.1 bar

Alarm is set to ring when the set point is exceeded by +5% (high alarm) or -5% (low alarm) for all controllers.

Company : Ethereal Energy Sdn Bhd	**Indication :**
Title : P&ID for reactor	**Major Pipeline :** ———————
Date : 24 May 2016	**Signal Pipeline :** ———//———
Designer : Nursuhaile Ahmad	

7.6.2 Process Control Heat exchanger, cooler and heater control system

Actually not all the heat exchangers need temperature control for maintaining purpose. However for the safety reason, they are equipped with feedback control system to maintain the temperature. In addition, maximum energy recovery for heat integration can be achieved.
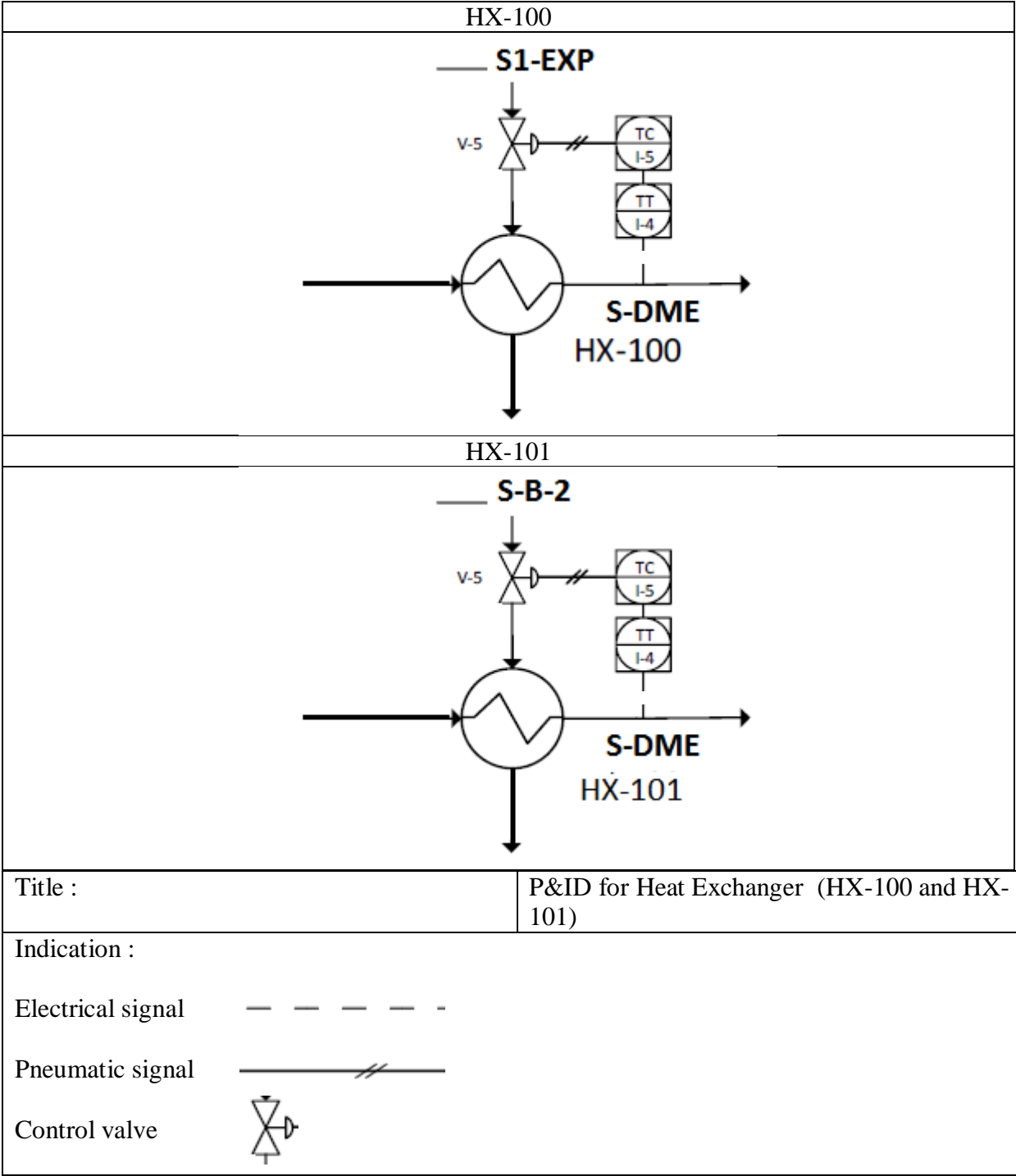
- **Heat Exchangers**

Objective: To control the temperature of stream at desired value so that the plant operates safely and can achieve maximum energy recovery

Table 7.2 : Control System for Heat Exchanger (HX-100 and HX-101)

Control Variable	Measured Variable	Manipulated Variable	Disturbance	Configuration of Controller	Set Point
HX-100					
Temp of inlet stream (S-DME)	Temp of inlet stream (S-DME)	Flow rate of process stream (S1-EXP)	Inconsistent temperature of process stream (S1-EXP) and ambient temperature	Feedback Control	486.5°C
HX-101					
Temp of inlet stream (S-DME)	Temp of inlet stream (S-DME)	Flow rate of process stream (S-B2)	Inconsistent temperature of process stream (S-B-2) and ambient temperature	Feedback Control	486.5°C

Alarm is set to ring when the set point is exceeded by +10% (high alarm) or −10% (low alarm) for all controllers.

Figure 7.1: Control System for Heat Exchanger (HX-100 and HX-101)

HX-100

HX-101

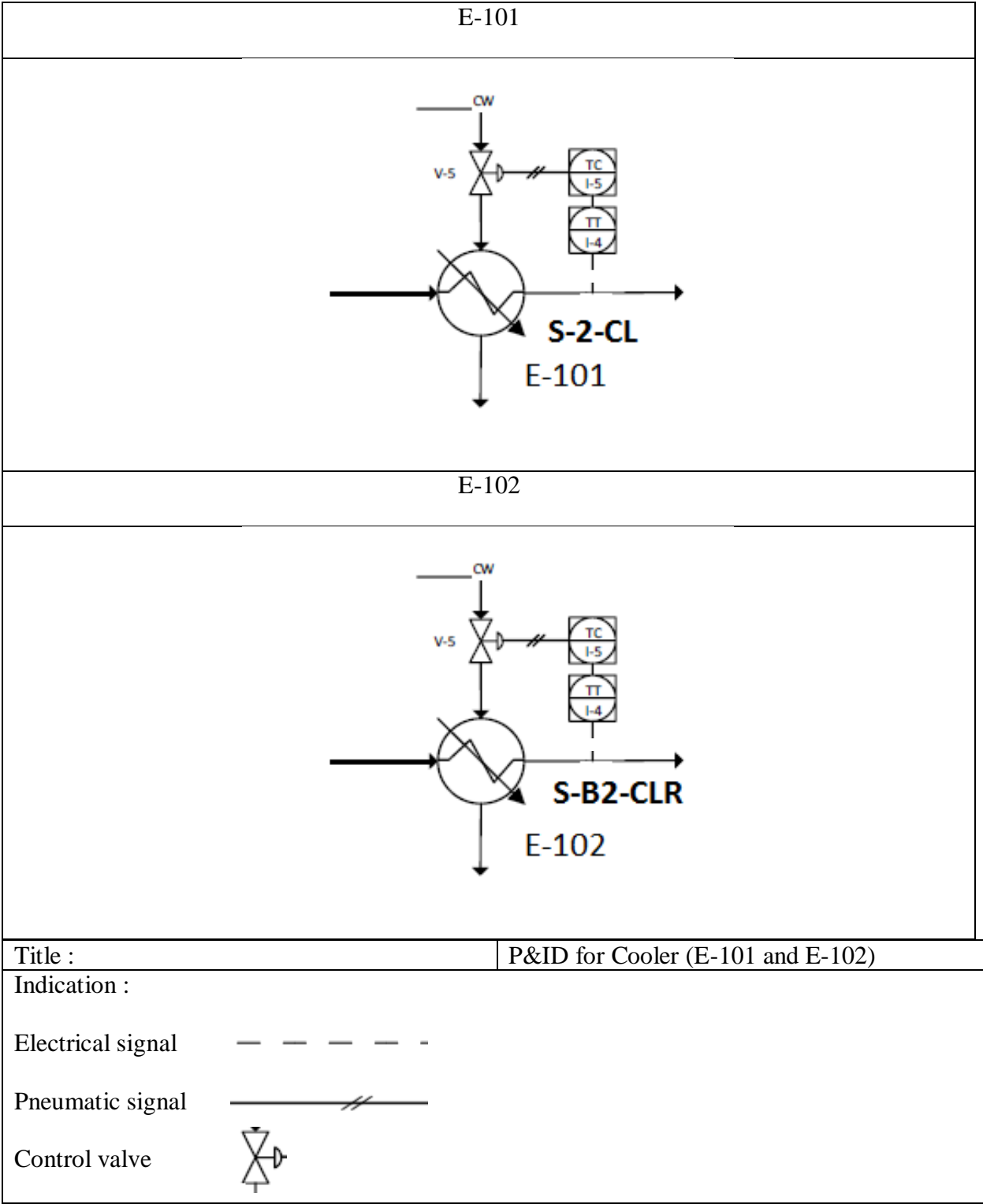| Title :	P&ID for Heat Exchanger (HX-100 and HX-101)
Indication : Electrical signal Pneumatic signal Control valve	

- **Cooler**

Objective: To control the temperature of stream at desired value so that the plant operates safely and can achieve maximum energy recovery

Table 7.3 : Control System for Cooler (E-101 and E-102)

Control Variable	Measured Variable	Manipulated Variable	Disturbance	Configuration of Controller	Set Point
E-101					
Temp of outlet stream (S-2-CL)	Temp of inlet stream (S-2-CL)	Flow rate of cooling utilities	Ambient temperature	Feedback Control	-100°C
E-102					
Temp of outlet stream (S-B2-CLR)	Temp of outlet stream (S-B2-CLR)	Flow rate of cooling utilities	Ambient temperature	Feedback Control	25°C

Figure 7.2 : Control System for Cooler (E-101 and E-102)

E-101

E-102

| Title :	P&ID for Cooler (E-101 and E-102)
Indication : Electrical signal — — — — - Pneumatic signal Control valve	

- **Heater**

Objective: To control the temperature of stream at desired value so that the plant operates safely and can achieve maximum energy recovery

Table 7.4 : Control System for Heater (E-100)

Control Variable	Measured Variable	Manipulated Variable	Disturbance	Configuration of Controller	Set Point
E-100					
Temp of inlet stream (S-DME)	Temp of inlet stream (S-DME)	Flow rate of heating utilities	Ambient temperature	Feedback Control	486.5 °C

Figure 7.3 : Control System for Heater (E-100)

P&ID for Heater (E-100)
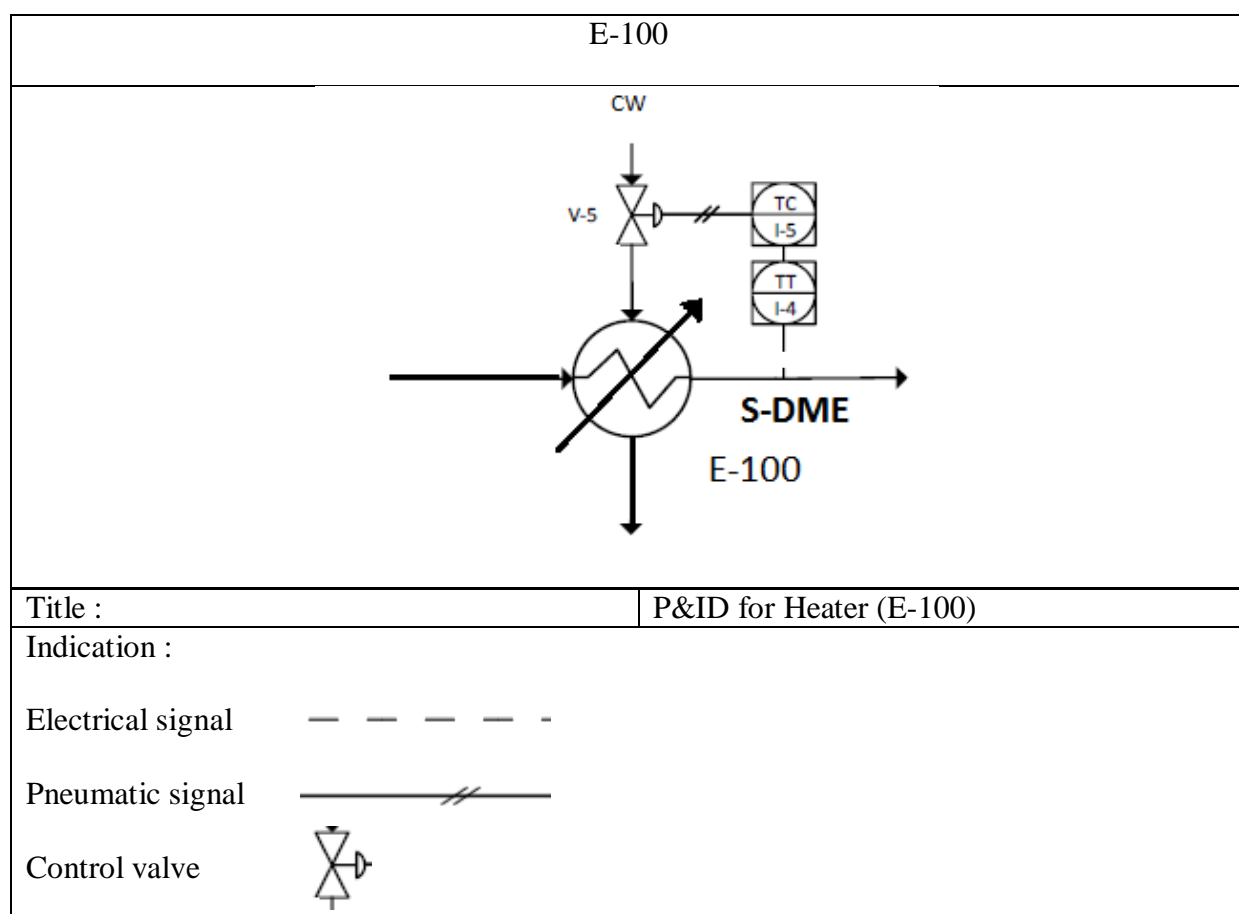
7.6.3 Distillation Column

- **Distillation Column D-100**

We will describe generally about the control objective of the distillation column. Control objectives of our distillation column are as below:
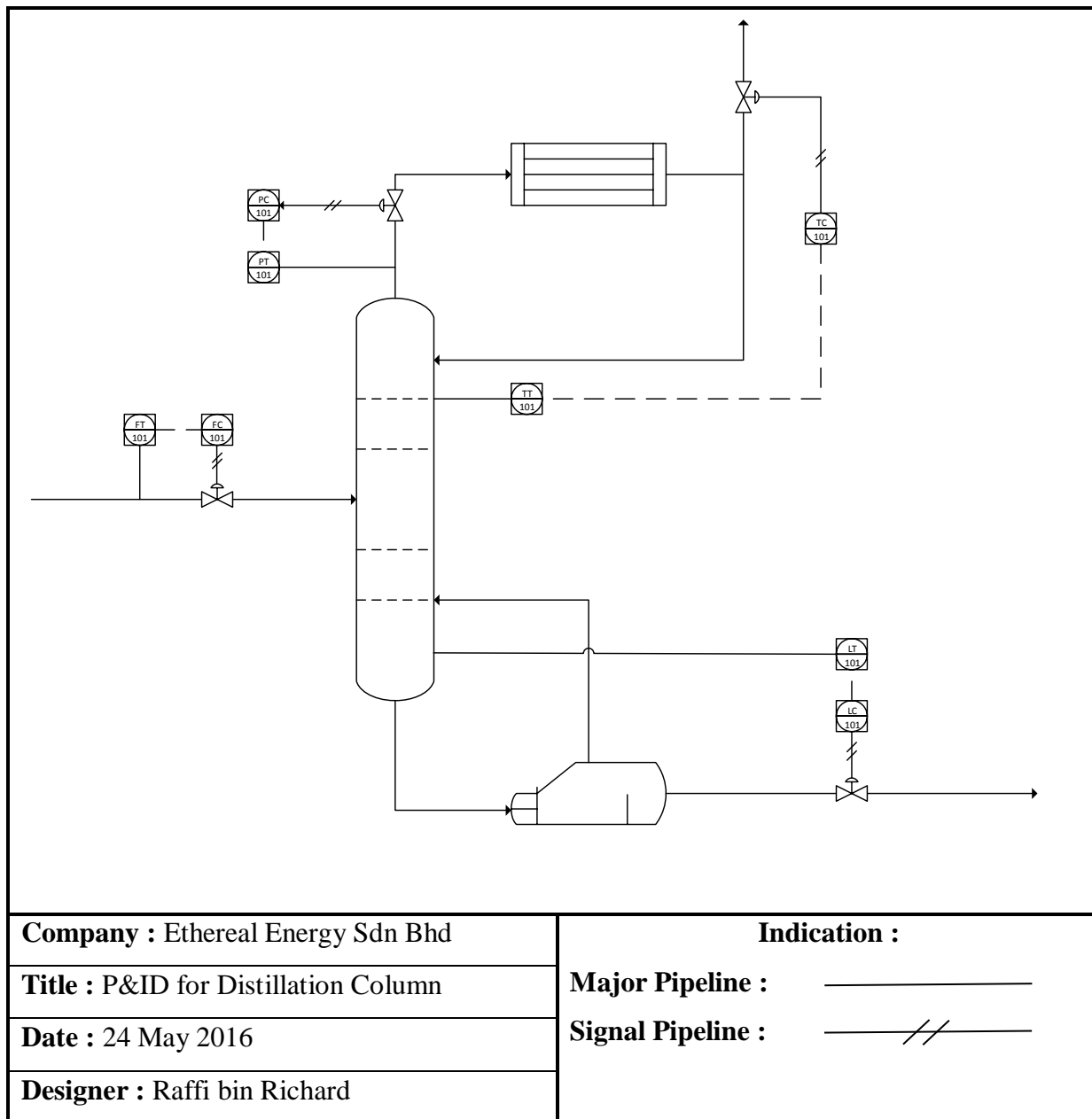
1. Maintain a constant feed flow rate to the column.
2. Control the temperature of the column to maintain it at desirable level.
3. Control the pressure of the column at desirable level.
4. Control the liquid level in distillation column from flooding or from being too low.
5. Control the composition of the distillate stream at desired level.
6.

Table 7.4: Control System of Distillation Column, D-101

Control Variables	Measured Variables	Disturbances	Manipulated Variables	Set Point
Feed to column	Flow rate of outlet of depressurizing valve	Flowrate of outlet of depressurizing valve	Feed flowrate	13.8099 kmol/hr
Level of liquid in column	Level of liquid in column	Feed flow rate, reflux ratio	Flowrate of bottom stream	Avoid flooding in the reboiler and condenser
Temperature of column	Temperature of each tray	Feed temperature, feed flowrate	Flowrate from depressurizing valve	4.78 ºC (top) 104.39 ºC (bottom) [others tray temperature between this range]
Pressure of column	Pressure of top tray	Feed pressure, feed flowrate	Flowrate of distillate	454.7 psia

Alarm is set to ring when the set point is exceeded by +10% (high alarm) 0r -10% (low alarm) for all controllers

Figure 7.4: Control System for Distillation Column, D-101

Company : Ethereal Energy Sdn Bhd	Indication :		
Title : P&ID for Distillation Column	Major Pipeline :	—————————	
Date : 24 May 2016	Signal Pipeline :	——————//——————	
Designer : Raffi bin Richard			

- **Distillation Column, D-102**

We will describe generally about the control objective of the distillation column. Control objectives of our distillation column are as below:
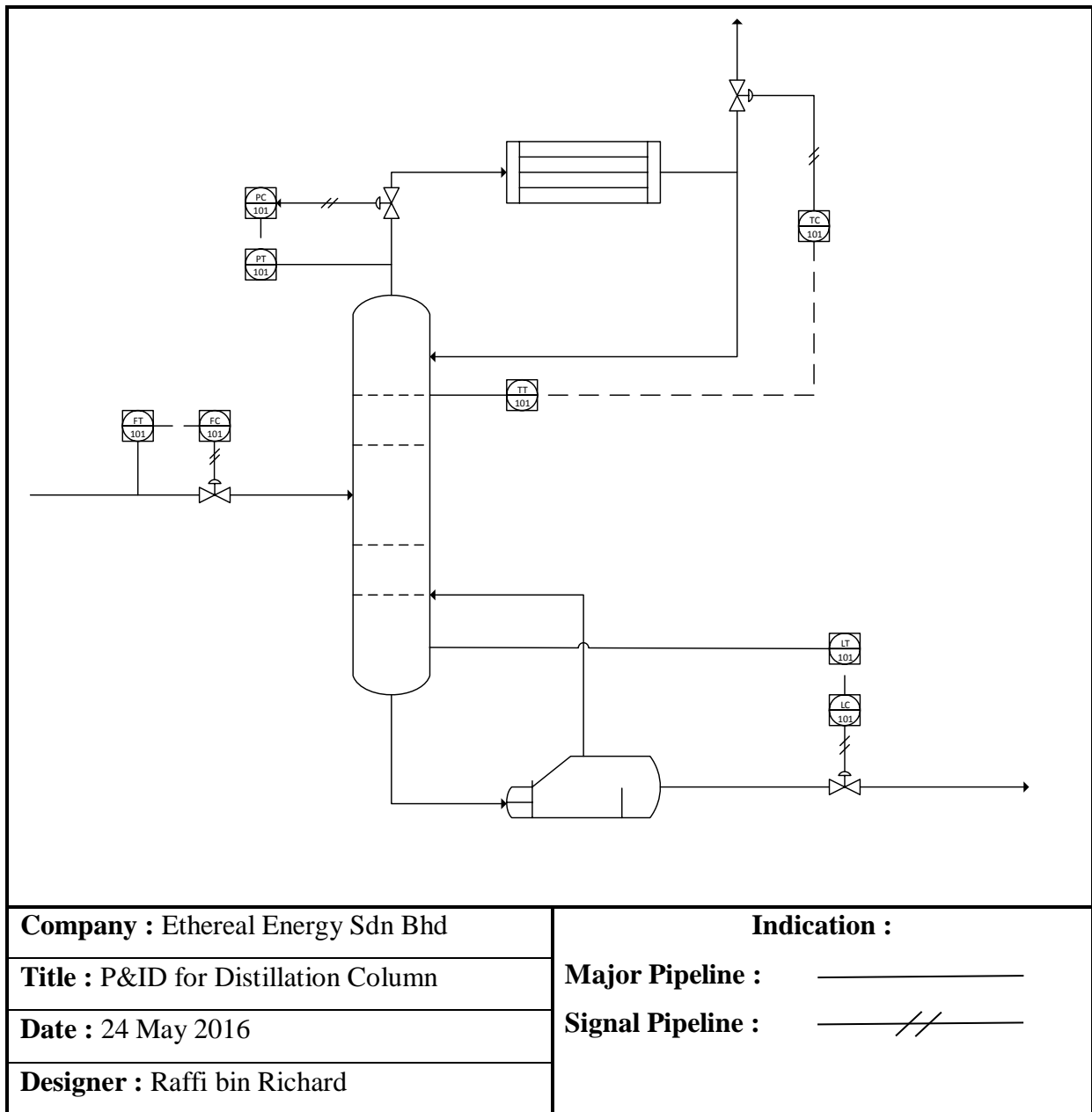
1. Maintain a constant feed flow rate to the column.
2. Control the temperature of the column to maintain it at desirable level.
3. Control the pressure of the column at desirable level.
4. Control the liquid level in distillation column from flooding or from being too low.
5. Control the composition of the distillate stream at desired level.
6.

Table 7.5: Control System of Distillation Column, D-102

Control Variables	Measured Variables	Disturbances	Manipulated Variables	Set Point
Feed to column	Flow rate of outlet of depressurizing valve	Flowrate of outlet of depressurizing valve	Feed flowrate	13.1115 kmol/hr
Level of liquid in column	Level of liquid in column	Feed flow rate, reflux ratio	Flowrate of bottom stream	Avoid flooding in the reboiler and condenser
Temperature of column	Temperature of each tray	Feed temperature, feed flowrate	Flowrate from depressurizing valve	36.36 ℃ (top) 111.12 ℃ (bottom) [others tray temperature between this range]
Pressure of column	Pressure of top tray	Feed pressure, feed flowrate	Flowrate of distillate	101.53 psia

Alarm is set to ring when the set point is exceeded by +10% (high alarm) 0r -10% (low alarm) for all controllers

Figure 7.5: Control System for Distillation Column, D-102

Company : Ethereal Energy Sdn Bhd	Indication :	
Title : P&ID for Distillation Column	Major Pipeline :	———————
Date : 24 May 2016	Signal Pipeline :	———//———
Designer : Raffi bin Richard		

- **Distillation Column, D-103**

We will describe generally about the control objective of the distillation column.
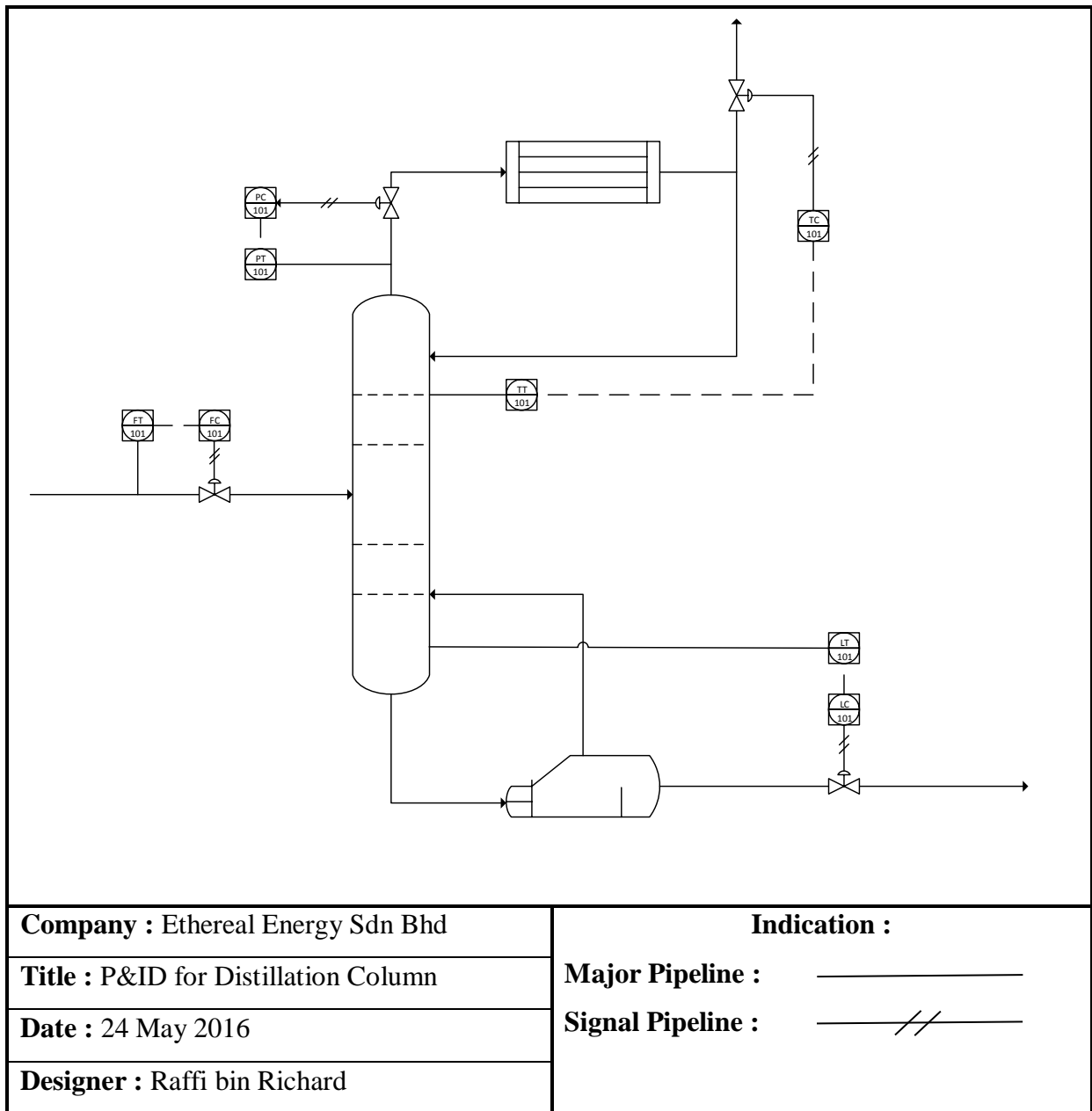Control objectives of our distillation column are as below:

1. Maintain a constant feed flow rate to the column.
2. Control the temperature of the column to maintain it at desirable level.
3. Control the pressure of the column at desirable level.
4. Control the liquid level in distillation column from flooding or from being too low.
5. Control the composition of the distillate stream at desired level.
6.

Table 7.6: Control System of Distillation Column, D-101

Control Variables	Measured Variables	Disturbances	Manipulated Variables	Set Point
Feed to column	Flow rate of outlet of depressurizing valve	Flowrate of outlet of depressurizing valve	Feed flowrate	11.7827 kmol/hr
Level of liquid in column	Level of liquid in column	Feed flow rate, reflux ratio	Flowrate of bottom stream	Avoid flooding in the reboiler and condenser
Temperature of column	Temperature of each tray	Feed temperature, feed flowrate	Flowrate from depressurizing valve	28.65 ℃ (top) 77.89 ℃ (bottom) [others tray temperature between this range]
Pressure of column	Pressure of top tray	Feed pressure, feed flowrate	Flowrate of distillate	159 psia

Alarm is set to ring when the set point is exceeded by +10% (high alarm) 0r -10% (low alarm) for all controllers

Figure 7.6: Control System for Distillation Column, D-103

Company : Ethereal Energy Sdn Bhd	Indication :	
Title : P&ID for Distillation Column	Major Pipeline :	————————
Date : 24 May 2016	Signal Pipeline :	———//———
Designer : Raffi bin Richard		

7.6.4 Flash Drum (V-100)

The control system introduce for the flash column is designed to operate column optimally and safely.

Objective: To remove carbon monoxide, CO in the stream that exist in gas phase. Two different phases, one as liquid phase and one as vapour phase will be separated in the column where the gas will go to the top while liquid at the bottom.
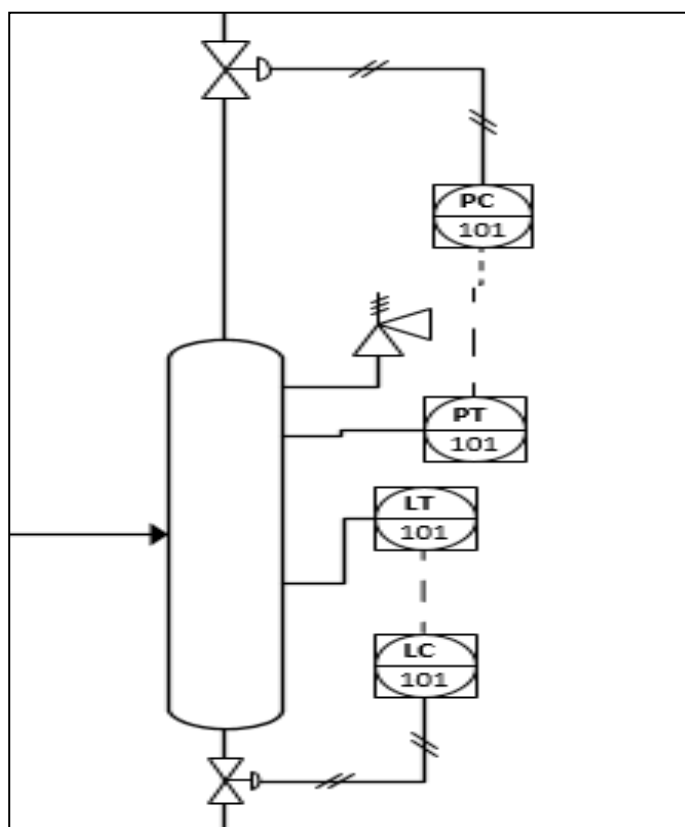
Control System:

1. To maintain the flash column pressure at 1 atm and to prevent the column from overpressure.

2. To maintain the liquid level in the flash so that it will not overflow or dry out. This is to ensure the   system interface always exist in order to make the flash column operate.

Table 7.7 : Control System for Flash Drum, V-100

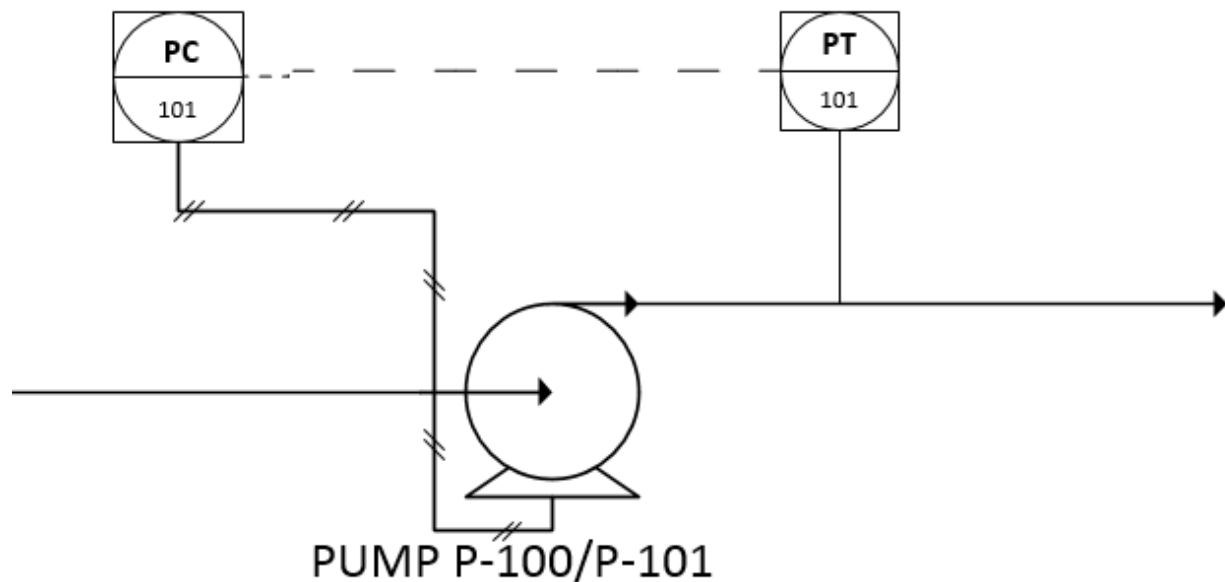| Control Variable	Measured Variable	Manipulated Variable	Disturbance	Configuration of Controller	Set Point
Pressure in flash column	Pressure in flash column	Outlet gas pressure at upper	Inconsistent pressure at the inlet stream	Feedback control	P = 1 atm
Liquid level in flash unit	Liquid level in flash unit	Outlet liquid flow rate at bottom	Inconsistent flow of inlet stream	Feedback control	Liquid do not exceed more than 1.0181 m

Alarm is set to ring when the set point is exceeded by +10% (high alarm) or −10% (low alarm) for all controllers.

Title :	P&ID for pump P-100/P-101
Indication :	
Electrical signal – – – – – –	
Pneumatic signal	
Relief valve	

7.6.5 Pump PUMP P-100/P101

The control system for a pump is designed to control the flow through the pump to get the desired pressure at the outlet stream. Pressure transmitter will indicate the pressure at the pump outlet and send the data through electrical signal to the pressure controller. The pressure controller will increase/decrease the pressure by controlling the pump efficiency.

PUMP P-100/P-101

Title :	P&ID for pump P-100/P-101
Indication : Electrical signal Pneumatic signal	

Table 7.8: Control System for Pump, P-100

Control Variable	Measured Variable	Manipulated Variable	Disturbance	Configur. of Controller	Set Point
Outlet flowrate	Outlet flowrate	Pump efficiency	• Feed pressure • Feed flowrate	Feedback	The outlet pressure is at 31 bar

Table 7.9: Control System for Pump, P-101

Control Variable	Measured Variable	Manipulated Variable	Disturbance	Configur. of Controller	Set Point
Outlet flowrate	Outlet flowrate	Pump efficiency	• Feed pressure • Feed flowrate	Feedback	The outlet pressure is at 15 bar

7.6.6 LPG Tanks

The objective:

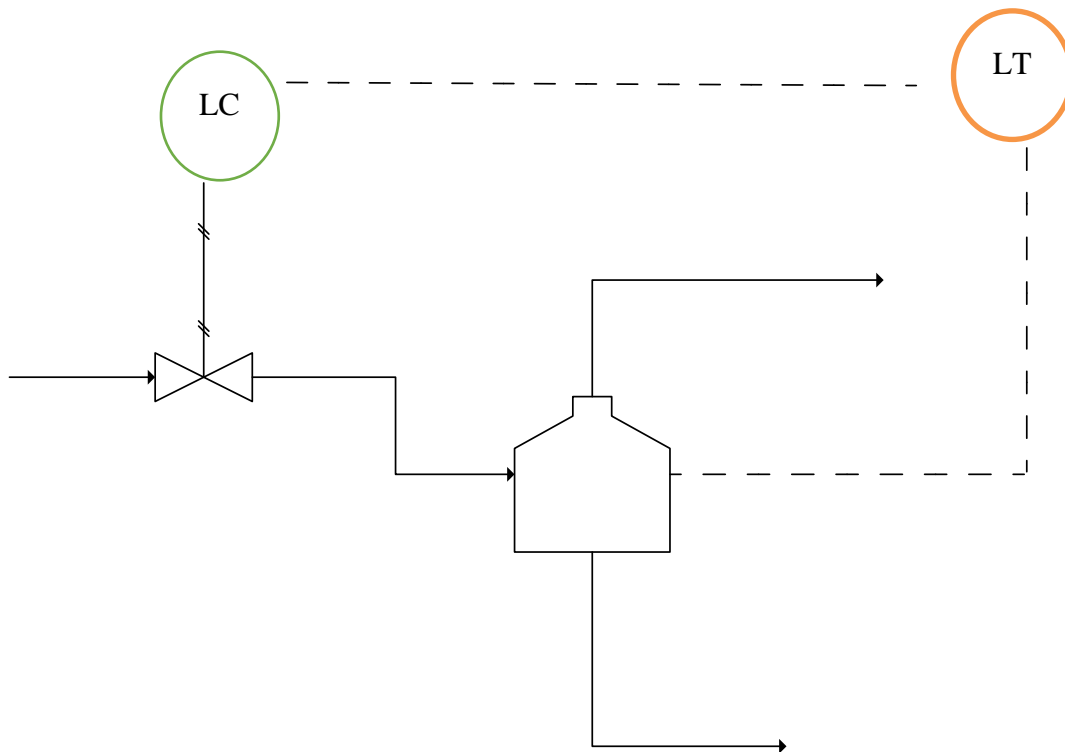1. To control level of LPG inside the tank
2. Control of flow rate of feed

Table 7.9 :control system for LPG tank

Control Variable	Manipulated Varible	Disturbance	Type Of Controller	Set Point
Flowrate of feed	Flowrate of feed	Temperature of tank	Feedback	Not sure

CHAPTER 8

PROCESS SAFETY STUDIES

8.1   Introduction

Process safety study is an important factor in designing a chemical plant to avoid major accidents in the designed plant. Furthermore, it ensures the safety of the workers, public and environment. Safety study is most effective to carry out early in a project for reducing the hazardous characteristics of a planned process. The main hazards in chemical plant are toxic and corrosive chemicals, explosions, fires and accidents common to all industrial activities. In this topic, various process safety techniques, general plant safety study, plant start-up and shutdown and the concept of hazard and risk will be discussed. Finally, hazard and operability study (HAZOP) on various unit operations will be done.

8.2   HAZOP analysis for equipment

The detail discussion on HAZOP studies will cover on some selected equipment like reactors, flash, heat exchangers, pumps, compressors, distillation columns and storage tank. When the studies are applied to this equipment, the potential hazards, which may arise from the deviations actions required to that should be taken, can be identified.

8.2.1 Reactor (R-001)

In the plant, HAZOP studies have to apply to the reactor (R-001) extensively because the process involved a reaction at high temperature and pressure. Therefore, their needs to be a proper control for this process and HAZOP are essential to identify the possible failure and step to be taken to overcome the problem when such a situation occurs. For this reaction, pressure, concentration and reactions are subject to discuss.

Table 8.1: **HAZOP Studies on Reactor**

Study Node	Process parameters	Deviation	Possible cause	Possible consequences	Action Required
Valve (feed in)	Flow	NO	1. Control valve doesn't function well 2. Valve is blocked	1. No feed comes to reactor 2. No reaction can occur	1. Check what is wrong with the valve 2. Clean the valve 3. Replace the valve if not repairable
		HIGH	1. Control valve is completely open 2. Controller doesn't function	1. Too much feed is introduced to the reactor. 2. The reactor may explode if the hydrogen feed too much	1. Install valve controller 2. Control valve manually 3. Constantly check the valve operation

		LOW	1. Leakage in the valve 2. Valve maybe be blocked partially	1. Not enough feed for the reaction to take place 2. Production quality drops	1. Check the valve 2. Clean the valve if blocked 3. Replace valve with new one 4. Install electrical valves
Reactor	Pressure	HIGH	1. Feed flow rate is too face 2. Product outflow pipe is blocked creating pressure	1. Might result in explosion 2. Leakage might happen in the reactor or pipes	1. Need to check all pipes, if blocked need cleaning 2. If feed flow rate is too fast need to install flow controllers
		LOW	1. Have leakage 2. Pump doesn't function well	1. Reaction can slow down 2. Quality of product will decrease	1. Check the pump 2. Install additional pressure controllers 3. Stop the leakage if any existing
	Temperature	HIGH	1. Heater is not working	1. Product quality affected	1. Check the heater

			well, temperature can't be controlled		2. Install temperature controller
		LOW	1. Heater is not heating well	1. Reaction rate decreases 2. Product quality decreases	1. Check the temperature controllers 2. Change heater or fix it
Reactor flow (outlet stream)	Flow	NO	1. Line blockage	1. Product loss 2. Downtime to overall process 3. Fire and explosion as the pressure built up in the reactor	1. Install flow indicator 2. Install flow control valves
		LESS	1. Line fracture 2. Valve failure	1. Leaks and spilled	1. Install flow indicator

8.2.2 Heat Exchanger, Heater and Cooler

The main objective of the HAZOP study on heat exchanger is maintaining the inlet and outlet temperature of heat exchanger equipment as best as possible to a desired temperature. For safe and satisfactory operation of the heat exchanger, certain constraints must be observed when doing HAZOP. For example, shell and tube pressure drop should be high enough to maintain effective operation. The analysis is through study of streams: inlet and outlet streams, steam, and process streams.

- Heat Exchanger Feed Flow Rate

Table 8.2: HAZOP Studies on Heat Exchanger Flow

Guide Word	Deviation	Possible Causes	Consequences	Action Required
Less	Less flow of cooling medium	• Pipe blockage • Control valve failure	• Temperature of process fluid remains constant • Effect further process	• Install High Temperature Alarm • Install filter at the coolant pipe
More	More cooling medium flow	• Failure of cooling medium valve • Failure of the coolant pump	• Temperature of process fluid decrease • Effect further process	• Install Low Temperature Alarm • Do maintainance accordingly to the valve and the pump
None	No cooling medium flow	• Failure of inlet cooling	• Process fluid temperature is not	• Install Temperature Indicator

		medium valve to open • Pipe broken • Pipe plugging	lowered accordingly • Effect further process	before and after the process fluid line • Install backup coolant line
Reverse	Reverse process fluid flow	• Failure of process fluid inlet valve • Coolant pump not functioning	• Product offset • Effect further process • Difficult to control the temperature flowing out	• Install check valve • Ensure pump is good condition regularly
Contamination	Process fluid contaminated	Leakage of tube and cooling medium goes in	Contamination of process fluid	Proper maintainance and operator alert

- Heat Exchanger Temperature

Table 8.3 : HAZOP Studies on Heat Exchanger Temperature

Guide Word	Deviation	Possible Causes	Consequences	Action Required
More	Higher temperature	• Temperature of the process fluid is too high • Fire occured near the heat	• Pressure will increase • Explosion could occur • Failure of other	• Install High Temperature Alarm • Repair damage part of the

		exchanger • Failure of heat exchanger	equipment	heat exchanger
Less	Lower temperature	• Failure of control system • No temp change at the coolant flow stream • Condensation occurs at the tube of the equipment	• Temperature increases with that condition • Failure of other equipment	• Install Low Temperatur e Alarm • Tube of the equipment should be cleaned often to remove the deposit

8.2.3 Distillation column

The studies on distillation column suggest the following objectives:

1. Product quality control to maintain either the overhead or bottom composition at a specified value
2. Material balances control to maintain its column hold-up and overhead and bottom inventories between maximum and minimum limits.
3. Constraints. The column should not flood. Pressure should be high enough to maintain effective column operation and the temperature difference in the reboiler should not exceeded critical temperature difference.

Table 8.4: HAZOP Studies on Distillation Column

Guide word	Deviation	Possible causes	Consequences	Action required
NO	No flow	- Pipe blockage - Control valve shut - Valves fail - Tube leakage and blocking - Pump failure	- Column dry out - Possible dangerous concentration - No operation	- Install low level alarm - Check maintenance procedure and schedule - Make bypass - Emergency plant shut down
LESS	Less Flow	- Pipe blockage - Control valve shut - Valves fail - Tube leakage and blocking - Pump failure	- Column dry out - Changes in product quality	- Install low level alarm - Check maintenance procedure - Make bypass - Emergency plant shut down
MORE	More flow	Control valve is fully opened	Flooding in the column	Install high level alarm

		• Increase pumping capacity • Control valves failure	• Changes in product quality • Temperature decrease • Rise in bottom liquid level	• Check maintenance procedure and schedule • Install control valves • Install flow indicator • Make bypass
HIGH	High pressure	• Vapor pressure high • Pressure indicator controller fail	• Low efficiency of separation • Rupture of column or other related equipment • Product loss	• Install high pressure alarm • Install pressure relief valve
	High Temperature	• Instrumentation failure • More steam flow • Exchanger tube failure • High steam pressure • Cooling water in condenser failure. • Reboiler control failure • Heating medium leak into process	• Separation cannot be done • Changes in product quality • Column flooding • Film boiling in column and reboiler • Phase effect • Pressure changes	• Install temperature indicator controller • Instruct operator on procedure • Attention to heat input and output control
LOW	Low pressure	• Vapour line leakage	• Low efficiency of separation • Loss of products	• Install pressure indicator
	Low	• Instrumentation	• Pressure change	• Install temperature

	temperature	failure	• Product loss	indicator
		• Less steam flow	• Changes in product quality	• Instruct operator on procedure
		• Loss of heating (steam leakage)	• Ineffective separating process	• Upgrade isolation
		• Low steam pressure and temperature	• Phase effect	• Attention to heat input and out control
		• Ineffective isolation		

8.2.4 Flash Column (V-100)

The parameters involve in this study liquid level and pressure.

1. Flash Column Liquid Level

Table 8.5: HAZOP Studies on Flash Column Liquid Level

Guide Word	Deviation	Possible Causes	Consequences	Action Required
More	Liquid level	• Control valve failure	• Flooding in column	• Install level indicator • Install high level alarm
Less	Liquid level	• Blockage in inlet stream	• No product	• Install level indicator • Install alarm

2. Flash Column Pressure

Table 8.6 : *HAZOP Studies on Flash Column Pressure*

Guide Word	Deviation	Possible Causes	Consequences	Action Required
Low	Pressure	• Input stream leakage	• Loss of products • Backward flow occur	• Install pressure indicator • Install backward check valve
High	Pressure	• High pressure in input stream • Expander malfuction	• Column rupture • Loss of product	• Install high pressure stream alarm • Install overpressure relieve valve

8.2.5 Pump P-100/P-101

HAZOP study on pump is important in order to maintain desired pressure into the distillation column. To achieve that, the pump must maintain either the pressure or flow to the desired pressure and flow.

Table 8.7 : HAZOP Studies on Pump Flow

Guide Word	Deviation	Possible Causes	Consequences	Action Required
Reverse	Flow	• Pump motor failure/not enough suction	• Contain high amount of vapour in stream	• Install backward check valve • Install backflow alarm
More of	Flow	• Pump failure/over power	• Floods in distillation column	• Install overflow alarm with automatic shutdown
No or Not	flow	• Pump failure • Stream blockage	• no flow/low pressure stream into distillation column	• Install low level alarm at the distillation column
Less of	flow	• Pump failure • Stream blockage	• Low pressure stream	• Install flow pressure alarm

8.2.6 LPG Storage Tank

HAZOP study on tank storage is crucial in order to avoid overflow and overpressure inside storage tank that would rapture the storage and flooding.

Table 8.8: HAZOP Studies on LPG storage tank

Guide Word	Deviatio	Possible cause	Consequences	Action Required
More of	Flow	Failure of control valve	Flooding inside storage tank	Install backward overflow alarm with auto-shutdown.
More of	Pressure	Failure of control valve	Tank will rapture or burst	Install pressure relief valve and install high pressure alarm

CHAPTER 9

PROFITABILITY ANALYSIS

9.1 Introduction

The economic analysis that involves the computation of the monetary investment that needs to pour into the plant should come before, because it's one of the most essential parts of the entire process design. The calculation involves the economic potential (Chapter 1) for the selected plant was just a rough estimation in which the calculation does not count in other factors such as depreciation, plant lifetime and so on the analysis will based on costing calculation from Chapter 6. Actually there are three parameters when discussing economic attractiveness: rate of return, Discounted Cash Flow Rate of Return (DCFRR) and NPV. Here, DCFRR is the main interest because it is designed to reflect the actual interest on investment and it is useful in comparing alternate possibilities.

Total Production Cost, A_{PC}	$A_{PC} = A_{ME} + A_{GE}$	20,123,032.60	56.45
Depreciation, A_{BD}	10% FCI	8,612,307.06	0.33
Total Expenses, A_{TE}	$A_{TE} = A_{PC} + A_{BD}$	**28,735,339.66**	**56.78**

Revenue from Sales		2,283,521,807.72
Net Annual Profit, A_{NP}		2,254,786,468.06
Income Tax	28% net annual profit	631,340,211.057
Net Annual Profit After Income Taxes, A_{NNP}		1,623,446,257.00
After tax Rate of Return, I	$((A_{NNP} + A_{BD})/TCI)*100$	1579.19 %

9.2 Cash Flow Analysis

The cash flow diagram cited the overall flow of the monetary. From the calculation and diagram, the payback period (PBP) can be estimated. In this plant, it is assumed that the plant will only start operate to produce ethylene glycol after 3 year of setup. This explains why the cash flow is negative for the first three years as the funder invest in the plant. It is reasonable to estimate that most of the chemical plant's life is from 15 years to 20 years. Here the operating period is 20 years with the initial 3 years of start-up operation. The TCI is funded accordingly in 1st year, 2nd year and 3rd year by the percentage of 20%, 40% and 40%. Due to start-up problem, the total annual sales only yield 80% for the first production year (4th year). But as time goes by, starting from next year (5th year) the annual sales is optimum. Table 12.1 and Table 12.2 are constructed to analyze the cash flow behavior at i = 0% and various discounted value.

Table 9.2: Undiscounted Cash Flow Analysis (Values in parentheses indicate negative cash flows

Completion of Year	Annual Capital Investment, A_i	Sales Income, A_S	Depreciation, A_{BD}	Total Expenses Less Depreciation, $A_{TE}-A_{BD}$	Cash Income, $A_{CI}=A_S+(A_{TE}-A_{BD})$	Allowances, A_A	Net Profit, $A_{NP}=A_{CI}-A_{BD}$	Federal Income Taxes, $A_{IT}=28\%A_{NP}$	Net Profit After Taxes, $A_{NNP}=A_{NP}-A_{IT}$	Cash Flow or Net Cash Income, $A_{NCL}=A_I+A_{BD}+A_A+A_{NNP}$	Cumulative Cash Flow
0											0.00
1	-20669536.96									-20669536.96	-20669536.96
2	-41339073.92									-41339073.92	-62008610.88
3	-31004305.44									-31004305.44	-93012916.32
4		1826817446	8612307.06	20123032.60	1846940478.6	0.00	1838328171.54	514731888.03	1323596283.51	1332208590.57	1239195674.25
5		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	2911742130.80
6		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	4584288587.34
7		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	6256835043.89
8		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	7929381500.44
9		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	9601927956.98
10		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	11274474413.53
11		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	12947020870.08
12		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	14619567326.63
13		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	16292113783.17
14		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	17964660239.72
15		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	19637206696.27
16		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	21309753152.82
17		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	22982299609.36
18		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	24654846065.91
19		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	26327392522.46
20		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	27999938979.00
21		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	29672485435.55
22		2283521807.72	8612307.06	20123032.60	2303644840.32	0.00	2295032533.26	642609109.31	1652423423.95	1672546456.55	31345031892.10

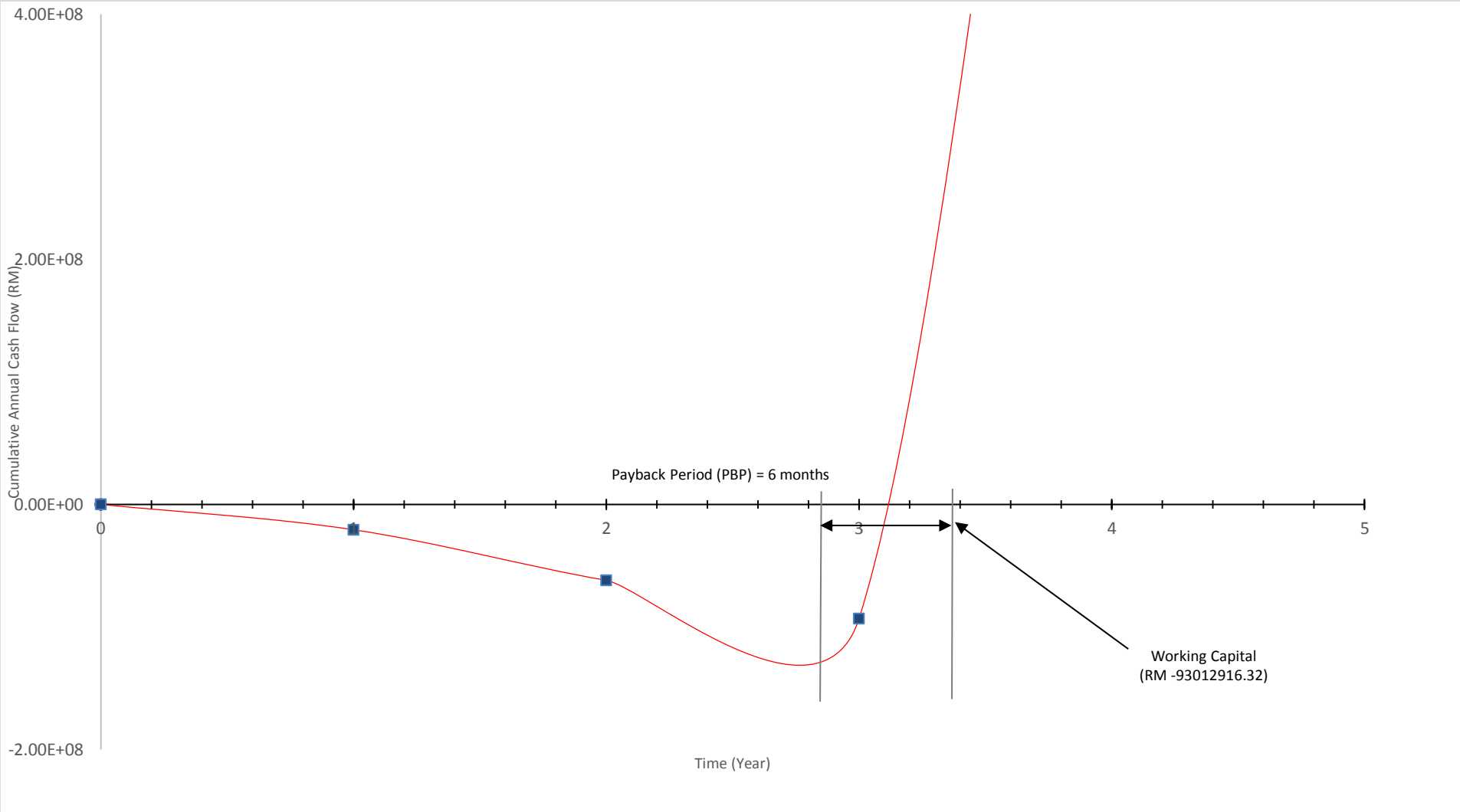| 23 | -93012916.32 | 2283521807.72 | 8612307.06 | 20123032.60 | 2303644840.32 | 0.00 | 2295032533.26 | 642609109.31 | 1652423423.95 | 1672546456.55 | 33017578348.65 |

Figure 9.1: Magnified Undiscounted Cash Flow Diagram (with i = 0%)

Table 9.3: Discounted Factor Cash Flow Analysis for 5% & 10% & 15% (Values in parentheses indicate negative cash flows)

Completion of Year	Cash Flow or Net Cash Income $A_{NCI}= AI + A_{BD} + A_A + A_{NNP}$	i= 0% Cumulative Cash Flow, $\sum A_{NCI}$	i= 5% Discount Factor, f_d	Discounted Cash Flow, $f_d A_{NCI}$	Cumulative Cash Flow, $\sum A_{NCI}$	i= 10% Discount Factor, f_d	Discounted Cash Flow, $f_d A_{NCI}$	Cumulative Cash Flow, $\sum A_{NCI}$	i= 15% Discount Factor, f_d	Discounted Cash Flow, $f_d A_{NCI}$	Cumulative Cash Flow, $\sum A_{NCI}$
0	0.00	0		0.00	0.00		0.00	0.00		0.00	0.00
1	-20669536.96	-20669536.96	0.95	-19636060.11	-19636060.11	0.91	-18809278.63	-18809278.63	0.87	-17982497.16	-17982497.16
2	-41339073.92	-62008610.88	0.91	-37618557.27	-57254617.38	0.83	-34311431.35	-53120709.99	0.76	-31417696.18	-49400193.33
3	-31004305.44	-72343379.36	0.86	-26663702.68	-83918320.06	0.75	-23253229.08	-76373939.07	0.66	-20462841.59	-69863034.92
4	1332208590.57	1301204285	0.82	1092411044.27	1008492724.21	0.68	905901841.59	829527902.52	0.57	759358896.62	689495861.70
5	1672546456.55	3004755047	0.78	1304586236.11	2313078960.32	0.62	1036978803.06	1866506705.58	0.50	836273228.27	1525769089.97
6	1672546456.55	3345092913	0.75	1254409842.41	3567488802.73	0.56	936626015.67	2803132721.25	0.43	719194976.32	2244964066.29
7	1672546456.55	3345092913	0.71	1187507984.15	4754996786.87	0.51	852998692.84	3656131414.08	0.38	635567653.49	2880531719.78
8	1672546456.55	3345092913	0.68	1137331590.45	5892328377.33	0.47	786096834.58	4442228248.66	0.33	551940330.66	3432472050.44
9	1672546456.55	3345092913	0.64	1070429732.19	6962758109.52	0.42	702469511.75	5144697760.41	0.28	468313007.83	3900785058.27
10	1672546456.55	3345092913	0.61	1020253338.49	7983011448.01	0.39	652293118.05	5796990878.46	0.25	418136614.14	4318921672.41
11	1672546456.55	3345092913	0.58	970076944.80	8953088392.81	0.35	585391259.79	6382382138.26	0.21	351234755.87	4670156428.28
12	1672546456.55	3345092913	0.56	936626015.67	9889714408.47	0.32	535214866.10	6917597004.35	0.19	317783826.74	4987940255.03
13	1672546456.55	3345092913	0.53	886449621.97	10776164030.44	0.29	485038472.40	7402635476.75	0.16	267607433.05	5255547688.07
14	1672546456.55	3345092913	0.51	852998692.84	11629162723.28	0.26	434862078.70	7837497555.45	0.14	234156503.92	5489704191.99
15	1672546456.55	3345092913	0.48	802822299.14	12431985022.43	0.24	401411149.57	8238908705.02	0.12	200705574.79	5690409766.78
16	1672546456.55	3345092913	0.46	769371370.01	13201356392.44	0.22	367960220.44	8606868925.46	0.11	183980110.22	5874389877.00
17	1672546456.55	3345092913	0.44	735920440.88	13937276833.32	0.20	334509291.31	8941378216.77	0.09	150529181.09	6024919058.08
18	1672546456.55	3345092913	0.42	702469511.75	14639746345.07	0.18	301058362.18	9242436578.95	0.08	133803716.52	6158722774.61
19	1672546456.55	3345092913	0.40	669018582.62	15308764927.69	0.16	267607433.05	9510044012.00	0.07	117078251.96	6275801026.57
20	1672546456.55	3345092913	0.38	635567653.49	15944332581.18	0.15	250881968.48	9760925980.48	0.06	100352787.39	6376153813.96

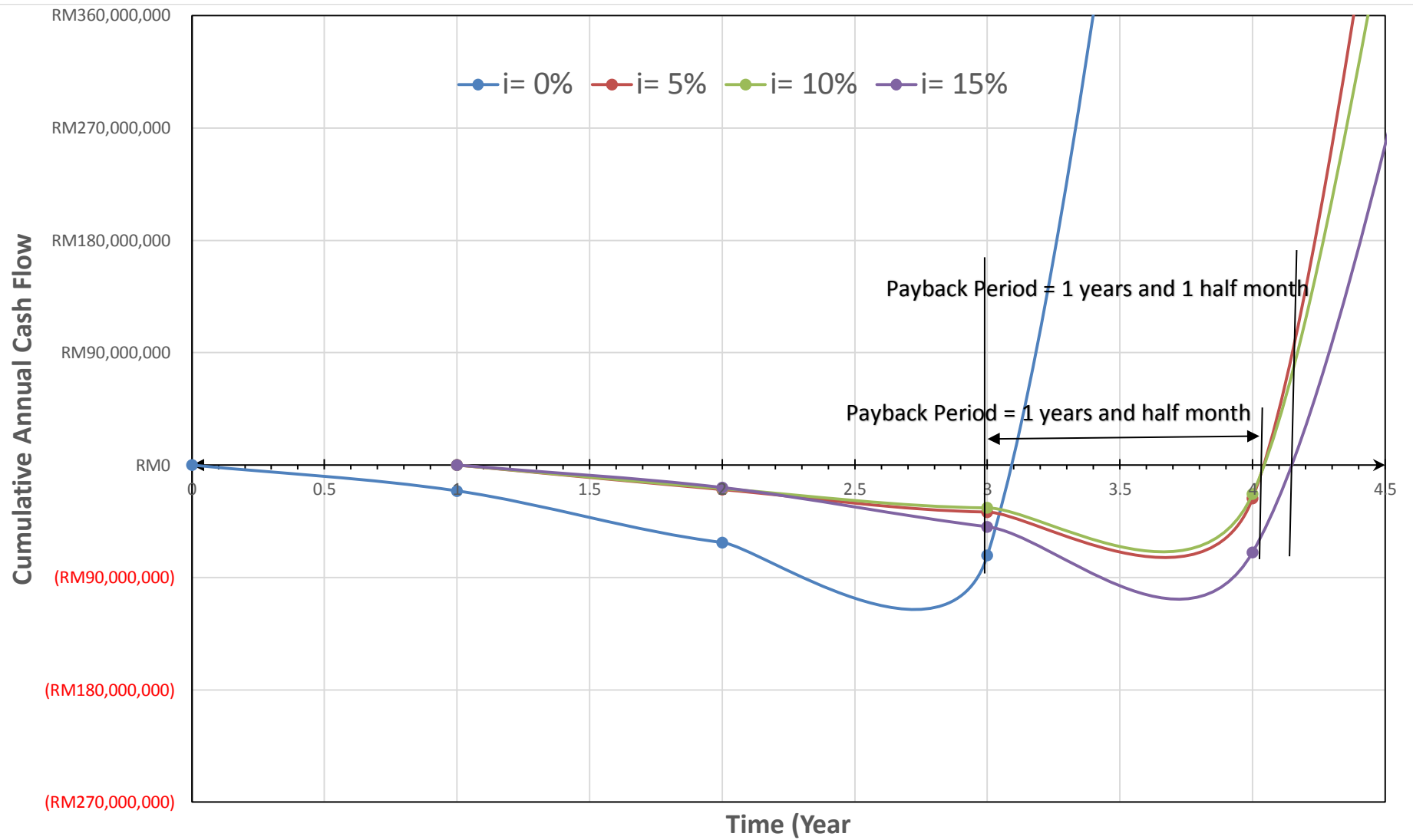| 21 | 1672546456.55 | 3345092913 | 0.36 | 602116724.36 | 16546449305.53 | 0.14 | 234156503.92 | 9995082484.40 | 0.05 | 83627322.83 | 6459781136.79 |
| 22 | 1672546456.55 | 3345092913 | 0.34 | 568665795.23 | 17115115100.76 | 0.12 | 200705574.79 | 10195788059.18 | 0.05 | 83627322.83 | 6543408459.61 |
| 23 | 1672546457 | 3345092913 | 0.33 | 551940330.7 | 17667055431 | 0.11 | 183980110.2 | 10379768169 | 0.04 | 66901858.26 | 6610310318 |

Figure 9.2: Discounted Cash Flow Diagram

From Figure 12.1, the PBP shows 1 years and half months after 3 years of start-up period. This means at i = 5%, the plant will start earning after the PBP. According to Ulrich (1984), the PBP lays between the start-up of a plant to the recovery of working capital. The discounted cash flow rate of return (DCFRR) is the point in which the rate results a Net Present Value (NPV) to zero. From Figure 12.2, within the period of 5 operating years, the NPV will be approaching zero value even having the highest i ≈ 15 %. The DCFRR point is the maximum interest rate that counted after the taxes. If the rate of interest in market is 10%, the payback period of this plant is roughly 1 years and half month after the start-up year. This means that the plant is profitable to construct even though the payback period is longer than undiscounted cash flow. This is because that the equipment will undergo regular maintenance and services. This will help to make sure that our plant can sustain for more than 15 years of operation and have a good net value profit.

9.3 Conclusion

After carried out the economic analysis to the ethylene glycol production plant, it has been meet up a few conclusions:

a) The Total Capital Investment is RM 103,347,684.80

b) The Total Expenses is RM 10,616,494.83 per year

c) The total utilities cost is RM 226,816.67 per month

d) The total operating labor costs is RM 264,000.00 per year

e) From Undiscounted Cash flow diagram, Pay Back Period is estimated at 1 years and half months after start up period at i = 5%.

f) The rate of returned obtained after the taxes is 1579.19 %.