TUGAS DESAIN PABRIK KIMIA – TK184803

PRELIMINARY PLANT DESIGN OF METHANOL FROM NATURAL GAS RESERVES

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SURABAYA
2020

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Laporan Tugas Desain Pabrik Kimia dengan Judul :

“Methanol from Natural Gas Reserves”

Diajukan untuk memenuhi salah satu syarat memperoleh gelar Sarjana Teknik pada Program Studi S-1 Departemen Teknik Kimia Institut Teknologi Sepuluh Nopember Surabaya.

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SUMMARY

Clean Natural Gas as raw material for upstream petrochemical industry is still one of the main choices for industry players. Basically it has constituent elements in the form of C, H, O, N, S, P.According to the presentation from the Ministry of Energy and Mineral Resources (ESDM) Indonesia's natural gas resource potential is quite large, spread in 19 provinces including South Sumatra province. Natural gas reserves in Indonesia are also estimated to reach 37 billion tons for the next 50 years.

The availability of natural gas which is increasingly depleting, encourages the development of process technologies that enable syngas raw materials. Natural Gas can be used as synthetic gas (syngas) through the gasification process. Syngas is obtained from the reaction of natural gas synthesis through a series of processes breaking down CH4 into CO, CO2, and H2. Syngas can be used for various things, one of which is to be used as methanol. So that the most possible raw material to replace natural gas is Natural Gas.

The development of industry in Indonesia, especially the chemical industry from year to year has increased the quality and quantity of both industries that produce finished materials and industries that produce semi-finished materials. The development of the chemical industry that produces these chemical products is very important because it can reduce Indonesia's dependence on foreign industries which will ultimately be able to reduce foreign exchange spending to import these goods, including one of which is methanol.

*Methyl alcohol*also known as methanol (CH3OH) is an upstream petrochemical industrial product derived from natural gas. This chemical compound is made through a catalyst synthesis reaction at low pressure which involves the parisal oxidation process of natural gas. At room temperature, methanol has the properties of clear liquid and clear, volatile and flammable, is a

polar solvent, soluble in water, ethyl alcohol, and ether. In addition it is toxic if inhaled and can cause blindness. Physical properties possessed by methanol include boiling point 64.70C and specific gravity 0.7866 g / mL. In general, methanol is used by various industries, such as plywood, textiles, plastics, synthetic resins, pharmaceuticals, insecticides, and others. Methanol is also used as a solvent, a coolant, and an adhesive raw material.

The use of methanol in Indonesia is dominated by two sectors, namely the formaldehyde industry by 25% and the MTO (Methanol to Olefin) and MTP (Methanol to Propylene) sectors by 22% while the rest is used in other industries such as acetic acid raw material, MTBE, gasoline / fuel, and so on ... The Methanol Industry is one of the prospective chemical industries in Indonesia. The need for methanol in Indonesia until 2018 is only met by one methanol producer, namely PT Kaltim Methanol Industri (KMI) with a capacity of 460,000 MTPY which produces AA grade pure methanol (minimum purity of 99.85%). As many as 70% of methanol products produced by KMI are export commodities, while the remaining 30% is used to meet methanol needs in Indonesia.

Therefore it is planned to establish a methanol plant that operates continuously 24 hours for 330 days per year with a production capacity of 460,000 tons / year with the need for raw materials for Natural Gas of 292,482 tons / year. This Natural Gas resource was taken from South Sumatra Island due to several factors that were used as a reference. The source of natural gas in South Sumatra to be processed comes from the Sakakemang Block, South Sumatra.

This methanol plant is planned to be established in the Tanjung Api-Api SEZ, South Sumatra starting in 2021 and completed in 2023. The process of making methanol from natural gas is divided into two main process sections that is:

- **Gasification** : The process of forming syngas from Natural Gas.

- **Synthesis of methanol** : The process of reacting CO and CO2 with H2O to form methanol.

Based on economic analysis, the factory's rate of return (IRR) is 37.33% at an annual interest rate of 11%, and an inflation rate of 3.5% per year. Whereas the payback period (POT) is 4 years 10 months and the break-even point (BEP) is 22.09% through a linear method. The life span of the plant is 10 years and the construction period is 2 years. To produce 460,000 tons / year of methanol, a total annual production cost (TPC) is required Rp 2,744,467,858,374.99 with a total investment cost (TCI) of Rp 4,660,384,242.75 and total sales of Rp 5,939,767,743,36565 by looking at aspects of the economic and technical analysis assessment, the methanol plant from Natural Gas is feasible to be established.

FOREWORD

Praise and gratitude to Allah SWT for all His mercy and grace that has been given to us, so we can complete our thesis proposal report with the title:

“ METHANOL FROM NATURAL GAS RESERVES”

This thesis proposal report became one of the points that should be done by the students of Chemical Engineering FTI-ITS to obtain undergraduate degree. This thesis proposal report is based on the literature of books and journals, as well as the application of knowledge that we have obtained during a period of lectures and especially in the Mass and Heat Transfer Laboratory, Department of Chemical Engineering FTI-ITS.

We as the author realized when in preparing this essay we can not do well without the help of various parties. Therefore, on this occasion we would like to thank:

1. Allah SWT
2. Parents and family for all the support, love, prayers once all sacrifice for us in educating and raising us
3. Dr.Ir. Susianto DEA as Head of the Laboratory Laboratory Mass and Heat Transfer, Department of Chemical Engineering FTI-ITS
4. Prof Dr. Ir. Ali Altway, MS as the main supervisor and Mr. Fadlilatul Taufany, S.T, Ph.D as our second supervisor.
5. Mr. Ir. Soejoto Gondosurohardjo who has guided and facilitated our research tasks.
6. Mrs. Dr.Eng Widiyastuti, S.T., M.T as head of the Department of Chemical Engineering FTI-ITS
7. Mr. Kusdianto , S.T., M.Sc.Eng as the coordinator of the Final and Final Department of Chemical Engineering FTI-ITS
8. Mr Dr.Ir. Susianto DEA and Mrs Dr. Yeni Rahmawati, M.T as our examiners
9. The entire faculty and staff in the Department of Chemical Engineering FTI-ITS

10. Our friends K-56 who have accompanied our days while in the Department of Chemical Engineering, FTI-ITS
11. Fellow at the Laboratory of Mass and Heat Transfer
12. All parties were unable to mention who have helped in the process of this thesis proposal report.

We also realize that there are still many things to be fixed on the thesis proposal that we have made, therefore we are looking forward to their criticisms and suggestions from various parties for the smooth running of our thesis kedepannya. As conclusion, we hope that eventually our Final Thesis this can be beneficial to many parties.

Surabaya, July 30th, 2020

Authors

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Chapter I
INTRODUCTION

I.1 Background

Methanol synthesis is a widely studied process and the product, methanol, is commonly utilized as a fuel, a solvent, an energy storage medium, and a feedstock in the industry. Especially in energy-related conversion, methanol is assumed to play an important role in the future. To meet the demands of the future, that makes requirements for process development to achieve sustainable and energy-efficient ways to produce methanol a necessity. Known as the “high-pressure” process, the first commercial methanol synthesis plant was opened in 1923 in Germany by Badische Anilin–& Soda Fabrik (BASF). The applied catalyst was based on Cu/ZnO. The so-called “low-pressure” process replaced the old one in the 1960s, when the ability to produce sulfur-free synthesis gas made it possible to use the more active copper-based catalysts. This process was operated in the lower pressures and temperatures, namely 60–80 bar and 250–280°C. Since then, the markets have been dominated by two low-pressure processes: ICI (Imperial Chemical Industries) and Lurgi-processes. Even though methanol synthesis has been a commercial process since 1923, there is still discussion about the reaction mechanism and kinetic modeling of the system. However, many studies assume that the methanol is produced primarily through CO_2 hydrogenation.

The modern methanol production technologies involved two major catalytic processes; the production of synthesis gas followed by upgrading the synthesis gas (i.e. syngas) into methanol. The principal raw materials for synthesis gas production are natural gas, methane gas from associated petroleum, shale gas, coal and biomass. Synthesis from fossil sources can be successfully achieved by the reforming technologies (i.e. dry and steam reforming) whereas pyrolysis and gasification processes for the biomass-based production option. The synthesis gas preparation and compression typically accounts for about 60% of the investment, and almost all energy is consumed in this process section.

Up to this day, the development of methanol production plant could be derived and varied to more than 25 products ranging from chemical use to fuel use as proven in Figure I.1. The so-called energy for the future, has so many advantages up to the level of where scientists, engineers, designers, doctors, even designers count in for the further and larger development and application of methanol in the future.

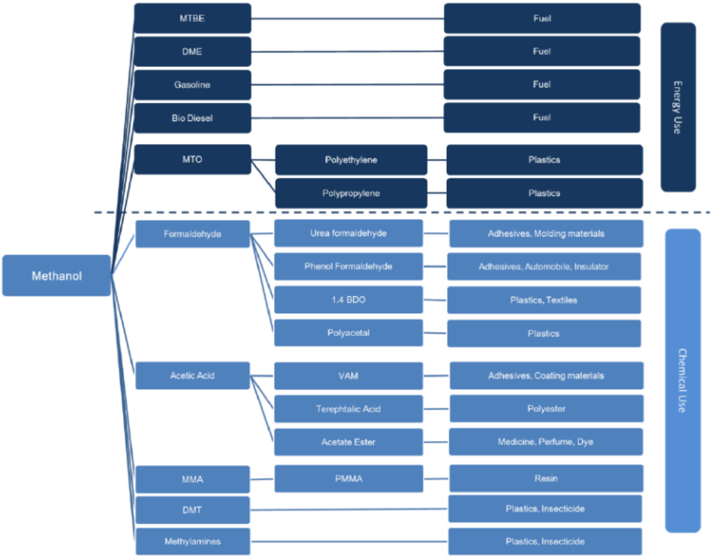


Figure I. 1 Options Available for Methanol Derivations (Sojitz Corporation,2017)

Large amounts of energy must be consumed to preserve humanity's actual living standards. Oil, natural gas and coal, still constitute our principal energy sources and offer the raw materials for producing a large variety of derivatives. Unfortunately, these resources are limited and not renewable on the human time scale, thus we will probably run out of coal, oil and natural gas within 3 centuries. For this reason, all feasible alternatives must be investigated to seek feasible and long- term solutions. Besides the problem of power generation, an important question is/will be how to store and efficiently use energy. The utilization of hydrogen is an option that is being discussed worldwide. However, its storage and transportation raises serious (safety) prob- lems, and no infrastructures are available. A feasible alternative could be liquid methanol. Methanol can serve as convenient energy storage medium, easily transportable fuel, solvent and building block for pro- ducing intermediates and synthetic hydrocarbons, also including polymers and single-cell proteins. Therefore, it might be and should be a key compound in the global economy of the future. Thus, the urge to construct a methanol production plant is planned to be a highly profitable and essential in terms of energy consumption, petrochemical needs, and all of its derivates need.

I.2 Industrial Development of Methanol Production

The first commercial methanol synthesis plant was opened in 1923 in Germany by Badische Anilin–& Soda Fabrik (BASF). The operational point in this process was at 320–450°C and 250–350 bar. The applied catalyst was based on Cu/ZnO. The so-called "low-pressure" process replaced the old one in the 1960s, when the ability to produce sulfur-free synthesis gas made it possible to use the more active copper-based catalysts. This process was operated in the lower pressures and temperatures, namely 60–80 bar and 250–280°C. Since then, the markets have been dominated by two low-pressure processes: ICI (Imperial Chemical Industries) and Lurgi-processes followed by other competitors such as Kellog, Synetix, MGC, and others as proven with the Figure I.2.

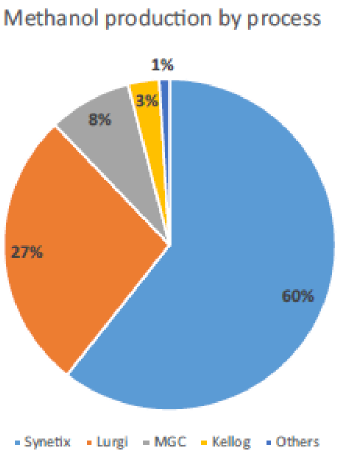


Figure I. 2 Options Available for Methanol Derivations (Matthey, 2015)

Methanol was first isolated by Robert Boyle in 1661 via wood distillation while its chemical composition was first discovered by Dumas and Peligot in 1834. At that time, production volumes were still small, e.g. 10–20 L per ton of wood treated for charcoal manufacturing. Initially, it was used for lighting, cooking and heating purposes but it was quickly replaced by more economical fuels. On the other hand, methanol was increasingly required by the chemical industry. In 1905, Sabatier proposed the first synthetic pathway for producing methanol, which implied reacting CO and H2. Based on this earlier discovery, the Badische Anilin und Soda Fabrik (BASF) patented a syngas-based methanol production process, where syngas was supplied via coal gasification. This process required a zinc/ chromium oxide catalyst as well as high temperature and pressure (300–400 °C and 250–350 atm). It was first deployed in Leuna, Germany, in 1923. The following technology developments aimed at reducing the pressure and temperature levels in order to improve process economics. Thanks to the invention of the steam reform- ing of methane, which allowed producing a purer syngas, a more active Cu/ZnO catalyst could be employed, thus decreasing the process temperature and pressure to about 300 °C and 100 atm. This significant improvement was proposed by ICI (Imperial Chemical Industries) in 1966.

Few years later, Lurgi developed a process with even lower operating pressure and temperature (230–250 °C and 40–50 atm). Subsequent developments further improved the process such that, nowadays, the production of methanol from carbon dioxide can be considered a mature technology. In fact, the actual methanol selectivity is over 99.8% with an energy efficiency of around 75%. Most of the current researches as shown in Table I.1 are the result of build-up of efforts from the earliest research up until now. The development update of methanol technology are now targeted to finding new ways to synthetize methanol using both diverse origin carbon dioxide/hydrogen feeds and different chemistries, e.g. the direct oxidation of methane and in the same time, keeping CO2 emission-free and currently developed by BASF. One of the way is to use torrefaction to reduce the emission. The fastest time to develop such technology, determined to be at least 10 years from this report comes out.

Table I. 1 The summary of the research on methanol synthesis in the chronological order

Author (Reference)	**Assumptions**	**Conditions**	**Catalyst/Other**
Natta (1955)	CO hydrogenation	High-pressure methanol synthesis High-pressure methanol synthesis	**ZnO/Cr_2O_3 catalyst**
Bakemeier et al. (1970)	The reactions of CO and		**ZnO/Cr_2O_3 catalyst**
	CO_2 noticed		
Brown et al. (1984)	The liquid-phase		**Powdered, Cu/Zn/Al**

	methanol synthesis	*T* ¼ 473–543 K,	**catalyst**
Weimer et al. (1987)	The liquid-phase	*P* ¼ 35–70 bar	**United catalyst**
	methanol synthesis	*T*¼523 K, *p*¼ 20–60 bar	
Leonov et al. (1973)	CO hydrogenation	Low-pressure methanol synthesis	**Cu-based catalyst**
Villa et al. (1985)	CO hydrogenation,		
	water-gas shift reaction		
	included	*T* ¼ 498–523	
Klier et al. (1982)	Primarily CO	K, *p* ¼ 75 bar	**CuO–ZnO catalyst and**
	hydrogenation, also CO_2	*T* ¼ 523–563 K,	**tubular, integral, fixed-**
	included in the kinetic	*P* ¼ 3–15 bar *T* ¼ 480–550 K,	**bed reactor**
	equation		
Agny and Takoudis	CO hydrogenation	*p* ¼ 15–50 bar	**Commercial Cu/ZnO/**
(1985)			**Al_2O_3 catalyst**
Graaf et al. (1988a,	CO and CO_2	*T* ¼ 453–553 K, *p* ¼ 15–51 bar	**The comparison of the**
1990),	hydrogenation and the	*T* ¼ 493–523 K,	**two- and three-phase**
Lommerts et al. (2000)	water-gas shift reaction	*p* ¼ 20–60 bar *T* ¼ 490–533	**methanol synthesis, mass**
		K,	**transport limitations**

Vanden Bussche and	CO_2 hydrogenation and	*p* ¼ 20–60 bar *T* ¼ 460–550 K, *p* ¼ 30–90 bar *T* ¼ 473–523 K, *p* ¼ 10 bar *T* ¼ 473–513 K, *p*¼34–41 bar *T* ¼ 473–548 K, *p*¼3.5–5.7 MPa *T*¼493–613 K, *p*¼50–90 bar	**$Cu/ZnO/Al_2O_3$ catalyst,**
Froment (1996)	the water-gas shift		**tubular reactor**
	reaction		
Von Wedel et al. (1988)	CO hydrogenation		**The gradientless**
			autoclave, commercial
			catalyst
Ledakowicz et al. (1992)	CO hydrogenation		**The stirred autoclave and**
			bubble-column slurry
			reactor, commercial
			catalyst
Skrzypek et al. (1991)	CO_2 hydrogenation and		**The fixed-bed (and also**
	the water-gas shift		**differential) reactor,**
	reaction		**commercial catalyst**
Coteron and Hayhurst	CO_2 hydrogenation, the		**The continuous, tubular,**
(1994)	role of CO related to the		**differential reactor**
Šetinc and Levec (2001)	catalyst surface reactions The liquid-phase		**The slurry reactor, the**

	methanol synthesis with		**commercial catalyst**
	Langmuir-Hinshelwood		
	mechanism		
Kubota et al. (2001)	CO_2 hydrogenation		**Cu/ZnO-based catalyst**
Park et al. (2014)	CO and CO_2		**Cu/ZnO/Al_2O_3 catalyst**
	hydrogenation		**(Süd-Chemie, MegaMax**
			700)

Chapter II
DESIGN DATA BASIS

II.1 Plant Production Capacity

Looking back at the very basic fundamental of Plant Prodcuction Capactiy, there are two major things have to be considered before we get to the supply and demand forecasting and planning. They are:

1. The Raw Material(s)
2. The Main Product

In the upcoming paragraphs, we are going to explain the briefs of each bullet points mentioned.

The fundamental of Chemical Engineering is to convert something or as we called it as Raw Material that is less economic and less beneficial to something that is way far more economic and more beneficial in large scale and is called product and byproduct. Raw material is so important that the existence of raw material in certain industries affect the majority of supply and demand of the product.

In this case, the material used to produce a AA grade of Methanol is Natural Gas instead of Biomasses, Coals, and Landfill gases. The natural gas itself were taken directly from the future development of Sakakemang Block near South Palembang precisely 60 kilometers northeast in KEK-Tanjung Api-Api. Looking at the supply and demand issued by the SKK-MIGAS, the biggest gas reserve for the next 20-25 years are located in 2 locations which are South Sumatera (Sakakemang Block) and West Papua (Bintuni Bay) shown in Figure II.1. As for the plan of development of gas lifting are slightly increasing and showing a promising trend couting from 2013 up to 2018 as shown in Figure II.2

Figure II. 1 Natural Gas and Oil Reserve Data per 2018

(SKK Migas,2018)

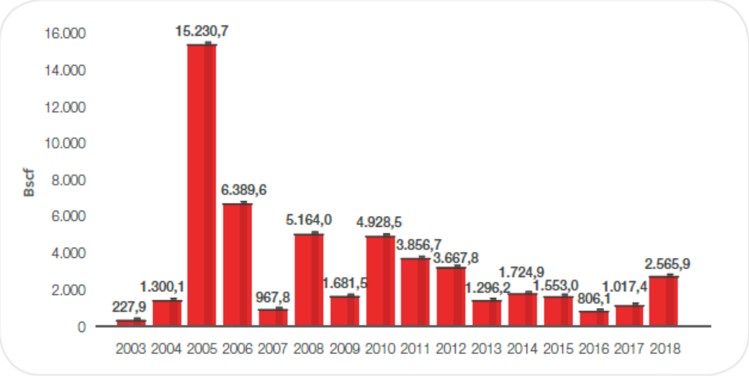


Figure II. 2 Natural Gas Plan of Development per 2018

(SKK Migas,2018)

By connecting through the dots, the raw material as per called natural gas is a recoverable resources and this is proven by the data shown in figure Figure II.3 The recoverable resources of raw material is per the most updated data by Repsol exploration team shown that South Sumatera held the biggest Natural Gas Recoverable Resources by 8.7 Tscf. This empower the decision to use natural gas as the main raw material.

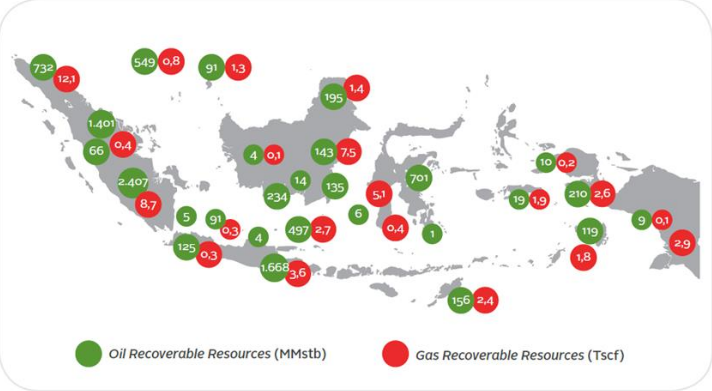


Figure II. 3 Natural Gas and Oil Recoverability Data per 2018

(SKK Migas and Repsol,2018)

Next is the main product. Current Worldwide Methanol market currently driven by china. As shown in the below Figure II-4, there will be a stable growth in methanol demand for “traditional” chemical intermediate application, such as formaldehyde, at a similar rate as the world economic growth as shown in Figure II.4 this by itself, need around more than 160,000 MT by 2021. On the other hand, MTO and energy applications are expected to lead the growth of the world methanol demand. It is estimated that the demand for MTO will reach 37 million ton per annum and this for energy application will be 22 million ton in 2021. Region-wise, the demand will be driven by China and, though the volume is relatively smaller, India and South East Asian countries will also contribute to the increase of the demand.

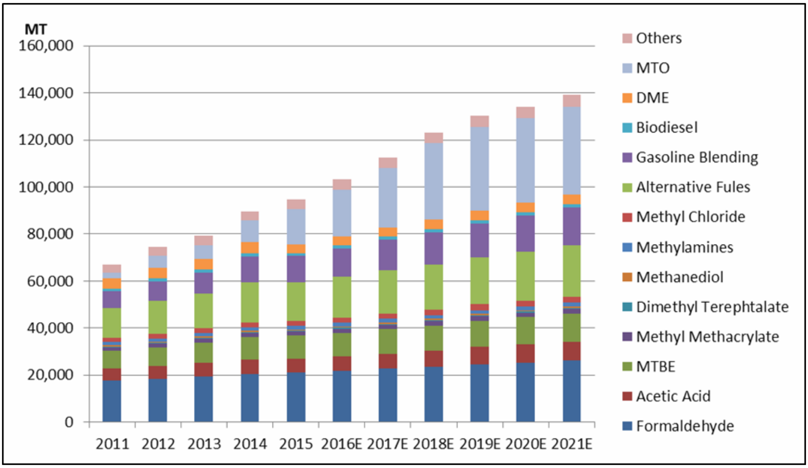


Figure II. 4 Signicficant Growth of Methanol Demand Prediction for a Variety Methanol Derivatives

(MMSA, 2019)

Based on the global market supply and demand analysis by MMSA, determined that derivates from methanol are significantly increasing from time to time as shown in Figure II.5.

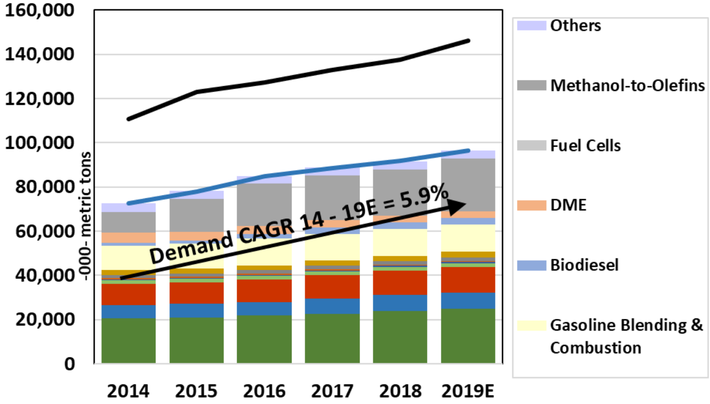


Figure II. 5 World Demand Data of Methanol Derivation (MMSA,2019)

Because of their population and economic magnitude, the methanol demand in China is extremely large, and furthermore, as abovementioned, methanol use for gasoline blending, MTO and DME are heavily concentrated in China. Therefore, the demand in China made up of a majority of the total methanol demand in the world. Almost of the demand is covered by the domestic plants, which are located inland China and producing methanol from coal. Inadequate supply in the coastal area is contributed by the import cargos, since the inland coal based methanol plants has logistic disadvantage to deliver products to the coastal area to meet their demand.

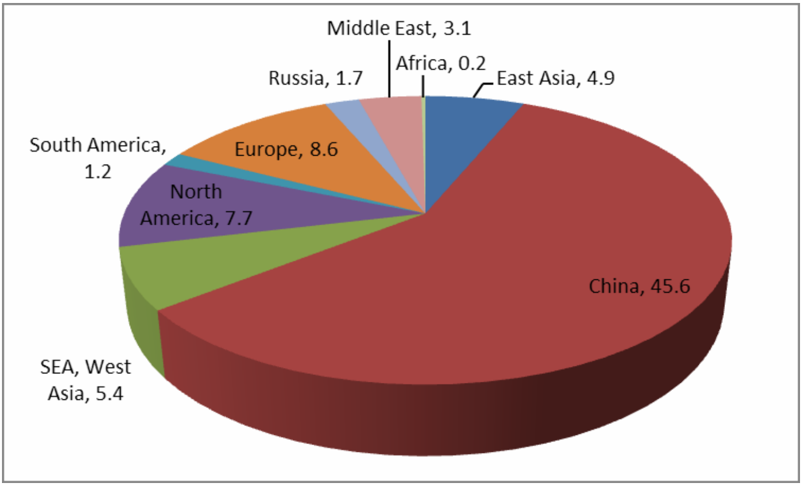


Figure II. 6 Methanol Demand by Region (MMSA,2019)

Far East Asia including China, Europe and North America are the main methanol net-import regions supplied by Middle East, South America and other regions. Import volume in Far East area of 10 million ton consists of 5 million ton by China and the rest amount by Japan, Korea and Taiwan. Although the importing volume of these 3 counties are not so much compare to the total demand in the world, those have a profound impact on the methanol trade flow as proven in Figure II.6.

Since the plant would be estimated built and running in the next 5 years, and there are is no possibility of the real data of the demand and production, factors that have been mentioned above strongly empower the decision making to estimate the needed capacity for production. Not just that, the prediction or as professionally called as “forecasting” could be empowered by the GDP Data by each nation. Methanol is one of the world’s leading economy base. Methanol fortunes remain largely tied to the performance of global economies. Prior to 2008, robust economic development in developed economies appeared, especially in service and high tech manufacturing sectors. Concurrently, developing nations, particularly the “BRIC” (Brazil, Russia, India, China) countries, saw record levels of manufacturing. This sustained expansion was essentially the result of supportive fiscal (government directed funding) and monetary (availability/cost of funds) policies in the developed countries. The supported data to empower decision making to forecast the demand could be empowered by the GDP data as below in Table II.1

Table II. 1 Worlwide GDP Data

Historic, Forecast Real GDP Growth Rates, %, 2008 - 2018E											
	2008	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**
Asia											
China	9.0	8.7	10.3	9.2	7.7	7.5	7.4	7.2	7.0	6.9	6.9
Japan	-0.2	-5.3	4.2	0.7	1.9	1.7	1.5	1.5	1.2	0.8	0.5
Taiwan	1.8	-3.5	10.1	4.2	1.8	2.7	4.1	4.0	3.8	3.7	3.6
South Korea	2.6	0.5	6.1	3.0	2.2	3.4	4.0	4.2	3.9	3.7	3.6
Malaysia	5.1	-2.4	6.8	4.8	5.2	4.9	4.8	4.7	4.6	4.5	4.5
Singapore	1.2	-2.1	14.8	4.9	1.8	3.1	5.1	4.9	4.8	4.6	4.4
Indonesia	6.1	4.6	5.9	6.5	6.4	6.7	5.9	5.7	5.4	5.2	5.1
Australia	2.2	0.9	3.1	2.1	3.0	2.7	3.1	3.3	3.1	3.0	3.0

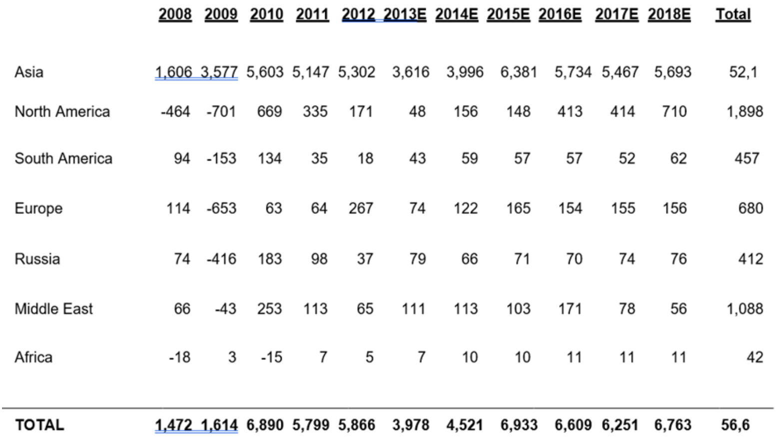
New Zealand	-0.9	-0.9	1.8	1.9	1.8	3.0	3.3	3.1	2.7	2.6	2.5
Other Southeast Asia	3.0	-2.0	7.0	3.1	5.8	4.9	5.1	4.9	4.7	4.4	4.3
India	5.3	6.5	9.1	7.6	5.8	7.1	8.6	8.4	8.1	7.8	7.8
Other South Asia	5.0	5.0	7.1	3.9	3.5	6.0	5.6	5.5	5.5	5.5	5.5
Americas											
North America	-1.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
US	1.2	-2.5	2.9	1.6	2.2	2.4	3.0	2.7	2.7	2.6	2.6
Canada	0.5	-2.5	2.5	2.4	2.2	2.1	2.6	2.5	2.4	2.4	2.4
Mexico	1.5	-6.8	5.0	3.4	4.4	3.9	3.7	3.6	3.6	3.6	3.6
South America	5.3	-0.1	7.2	4.8	2.3	3.9	4.6	4.3	4.3	4.3	4.3
Europe	0.9	-3.9	1.7	1.4	0.1	0.5	1.7	2.2	2.1	2.0	1.9
Russia	5.8	-8.0	3.7	4.3	3.7	3.7	4.0	4.0	3.9	3.9	3.9
Middle											

East											
Saudi Arabia	4.2	0.2	3.8	6.4	5.5	3.5	4.5	4.4	4.3	4.1	3.9
Iran	0.3	-3.0	2.2	4.3	2.0	0.0	1.2	2.1	3.0	3.5	3.5
Qatar	6.0	1.0	4.8	6.7	5.0	4.6	4.5	4.0	3.9	3.8	3.7
Oman	4.5	1.3	3.5	6.9	5.8	3.9	4.5	5.1	5.0	4.9	4.7
Other ME	4.0	1.6	3.4	6.3	4.7	4.7	4.5	4.3	4.2	4.1	4.1
Africa	3.0	-0.2	3.2	3.9	2.8	4.0	4.9	4.8	5.0	4.8	4.7

(Methanol Institute,2019)

Now all of the data to empower the decision is properly assembled, the next is we need supply and demand data from Asia and directly to Indonesia since the plant would not just produces methanol to Indonesia, the exports possibility is one of the main consideration. As shown in Table II.2

Table II. 2 Methanol Growth Demand Data per 2019



	2008	2009	2010	2011	2012	2013E	2014E	2015E	2016E	2017E	2018E	Total
Asia	1,606	3,577	5,603	5,147	5,302	3,616	3,996	6,381	5,734	5,467	5,693	52,1
North America	-464	-701	669	335	171	48	156	148	413	414	710	1,898
South America	94	-153	134	35	18	43	59	57	57	52	62	457
Europe	114	-653	63	64	267	74	122	165	154	155	156	680
Russia	74	-416	183	98	37	79	66	71	70	74	76	412
Middle East	66	-43	253	113	65	111	113	103	171	78	56	1,088
Africa	-18	3	-15	7	5	7	10	10	11	11	11	42
TOTAL	**1,472**	**1,614**	**6,890**	**5,799**	**5,866**	**3,978**	**4,521**	**6,933**	**6,609**	**6,251**	**6,763**	**56,6**

The table could be converted into a table to ease the readability of prediction into the Figure II.7 as shown below.

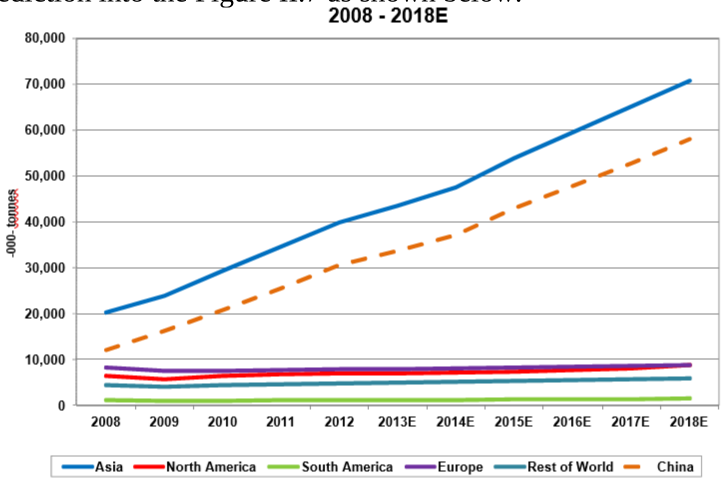


Figure II. 7 Graphic of Methanol Demand Growth per 2019

(Methanol Institute,2019)

By having the demand, now we moving to the supply section of the methanol. It is stated in MMSA, that worldwide scale of supply and demand could be as high as more than 100,000 thousand tonnes a year as shown in Figure II.9. By counting the worldwide scale, it is a hard prediction to estimate the possibility of production. Therefore, we will try to conduct the prediction by adding more region data whereas the plan would be located. In this scene, Asia-Pacific, the graph shows us that the production didn't match the total demand by as high as 15,000 thousand tonnes for Asia. This is a strong opportunity to be consider. The opportunity proven in Figure II.8 below.

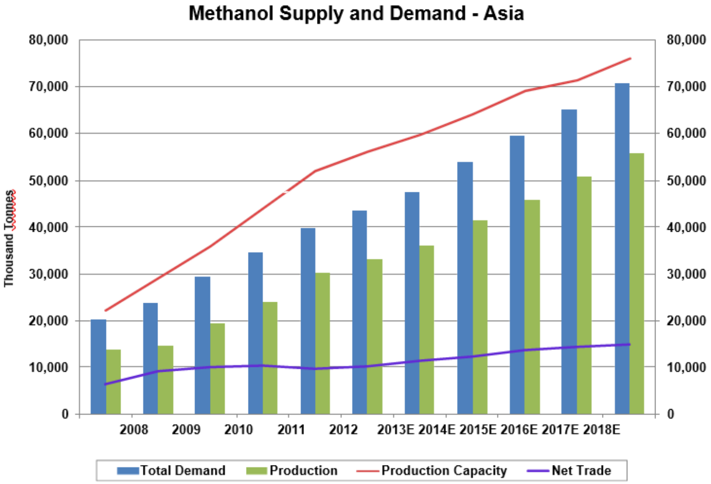


Figure II. 8 Graphic of Methanol Demand Growth per 2019 (MMSA,2019)

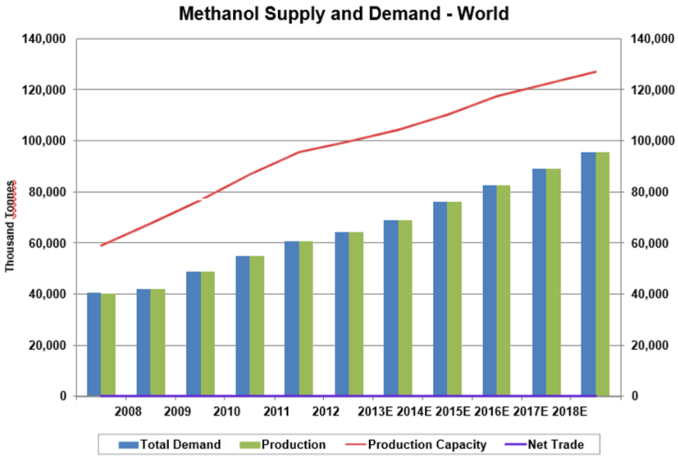


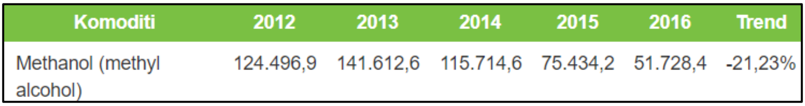
Figure II. 9 Supply and Demand Data of Methanol per 2019

(MMSA,2019)

After matching everything and connecting thorugh many dots, now we could consider a high projectile to the exact nation of the future plant location, Indonesia.

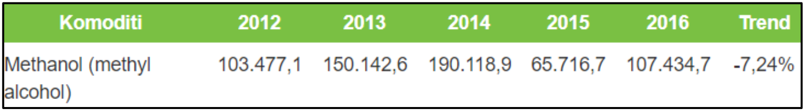
Since there are no exact Based on the supply and demand chain table below, and the arising concern directly from the Ministry of Industry in 2018, states that by the start of 2021, the demand capacity of pure methanol would arise up to 900.000 tonne/year minimum. The main consideration to conduct a market research is we need to know historical data of methanol market in Indonesia. Including export liability, import liability, production and consumption by national scale. The export and import liability could be predicted as Table II.3 and Table II.4 below.

Table II. 3 Export Liability of Methanol in Indonesia



Komoditi	2012	2013	2014	2015	2016	Trend
Methanol (methyl alcohol)	124.496,9	141.612,6	115.714,6	75.434,2	51.728,4	-21,23%

Table II. 4 Import Liability of Methanol in Indonesia

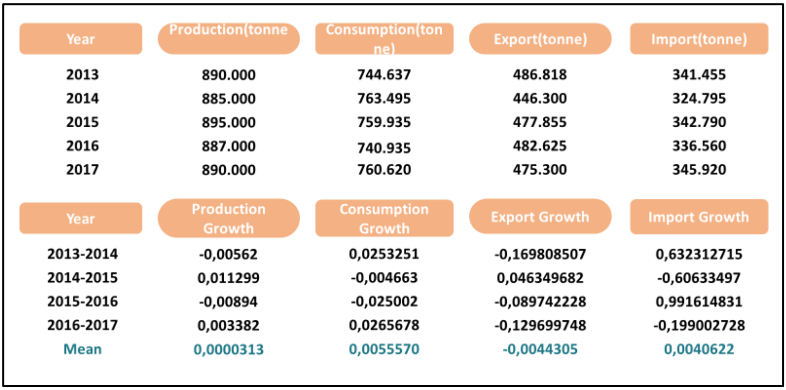


Komoditi	2012	2013	2014	2015	2016	Trend
Methanol (methyl alcohol)	103.477,1	150.142,6	190.118,9	65.716,7	107.434,7	-7,24%

(Kemenperin,2019)

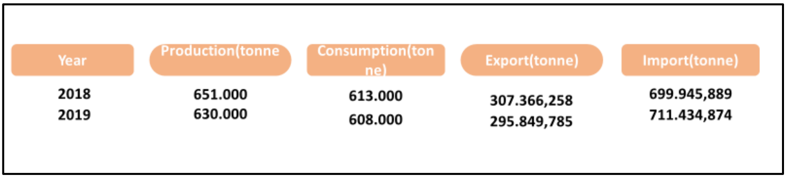
Next, we need to consider the production and consumption of Methanol. Data from Kemenperin shows that there are slightly declining of production and slightly increasing in consumption as shown in Tabel II.5, this means the opportunity to sell Methanol in Indoensia rapidly increasing. Not just that, judging from the export and import difference, there are 7% of decreasement in import and 21% in export. This is a huge opportunity for the production plant.

Table II. 5 Methanol Supply and Demand Insight in Indonesia



Year	Production(tonne	Consumption(tonne)	Export(tonne)	Import(tonne)
2013	890.000	744.637	486.818	341.455
2014	885.000	763.495	446.300	324.795
2015	895.000	759.935	477.855	342.790
2016	887.000	740.935	482.625	336.560
2017	890.000	760.620	475.300	345.920

Year	Production Growth	Consumption Growth	Export Growth	Import Growth
2013-2014	-0,00562	0,0253251	-0,169808507	0,632312715
2014-2015	0,011299	-0,004663	0,046349682	-0,60633497
2015-2016	-0,00894	-0,025002	-0,089742228	0,991614831
2016-2017	0,003382	0,0265678	-0,129699748	-0,199002728
Mean	0,0000313	0,0055570	-0,0044305	0,0040622



Year	Production(tonne	Consumption(tonne)	Export(tonne)	Import(tonne)
2018	651.000	613.000	307.366,258	699.945,889
2019	630.000	608.000	295.849,785	711.434,874

Based on the given data on Figure above, we could estimate the supply and demand by using the equation below :

$$P_{2023} = P_{2019} \text{ x } (1+i)^n$$

Known : n = 2025 – 2019 = 4

P = Production Capacity

(Peters & Timmerhaus, 1991)

For the capacity of consumption and export using the same exact equation as above, we could forecast the supply and demand for the plant production at 2022 when the plant is finally and efficiently built. For this case, we assume that there are a variety of export possibility considering the given data of supply and demand in Asia-Pacific Region with a number as high as 15,000 thousand tonnes and Worlwide with a number as high as 20,000 thousand tonnes. For this assumption, we could consider that there are zero import occurs, the methanol plant production is calculated with the mindset of to cover the import factor as highly as possible, thus, to be consider as shown in Table II.6

Table II. 6 Supply and Demand Estimation for Indonesia in 2025

Year	Export (Tonne)	Import (Tonne)	Production (Tonne)	Consumption (Tonne)
2025	305 792.34	690 315.93	615 674.94	769 079.05

Based on the given data above:
National Plant Production Capacity = (Export + Consumption) – (Import + Production)
= (305.792,34+ 769.079,05) – (615.674,94)
= 690.315,93 Tonne/Year
= 459.152,46 KTPA

II.2 Plant Location

There are various factors which affect the plant location, including both qualitative and quantitative factors like political, economic, environmental and facility criteria. Thus, there is a need for developing a systematic plant location selection process of identifying and prioritising relevant criteria and evaluating the trade-offs between all these criteria. The approach should also reduce the time in selection and develop consensus decision-making. Narasimhan (1983), Nydick and Hill (1992), and Partovi et al. (1989) suggested the use of the analytic hierarchy process (AHP) because of its inherent capability to handle the qualitative and quantitative factors. Prasanna Venkatesan and Kumanan (2012), Jahromi et al. (2012), and Sunil Kumar and Subash Babu (2011) also used the AHP in their research work for the decision-making process in various fields. AHP is easy to understand and can improve the decision-making process. It decomposes the complex decision-making problem into a simple hierarchy consists of different levels, showing the problem systematically in terms of criteria and sub-criteria. This hierarchy is called AHP model. Using this model, the pairwise comparison of the criteria and sub-criteria is done and their priority weights are found. Based on this information several plant locations can be compared effectively and the best location is selected as figured in Table II.7 below.

Table II. 7 Different Criteria and Subcriteria with references Different criteria and sub-criteria with their references

Factors affecting the plant location	*References/source*
Economic factors:	
Transportation cost	Moriarty (1980), McMillan (1965), Beckmann (1968)
Raw material cost	Schmenner (1982), Moriarty (1980), Pietlock (1992)
Cost of land	Hoover (1948), Greenhut (1956)
Cost of power	Heckman (1978), Moriarty (1980), Gold (1991)
Cost of labour	McMillan (1965), Schmenner (1982)
Income level of consumers	Carnoy (1972), Dicken and Lloyd (1978)
Environmental factors:	
Pollution and environmental constraints	Schmenner (1982), Coughlin et al. (1990)
Waste disposal	Heckman (1978), Moriarty (1980)
Climate	Spooner (1974), Moriarty (1980)
Facilities:	
Availability of labour	Carnoy (1972), Rees (1972, 1983)
Transportation facilities	McMillan (1965), Beckmann (1968)
Availability of professionals	Sharma (2004)
Availability of power	Greenhut (1956), McMillan (1965), Moriarty (1980)
Availability of infrastructure	Sharma (2004)
Availability of raw material	Greenhut (1956, 1981), McMillan (1965)
Availability of markets	Dorward (1979), Moriarty (1980), Schmenner (1982)
Size of existing market	McMillan (1965), Carnoy (1972)
Site selection	Sharma (2004)
Availability of industrial site	Greenhut (1956), Smith (1966, 1981)
Industrial relations	Sharma (2004)
Proximity to financial institutions	Sharma (2004)
Proximity to ancillary units	Schmenner (1982), Coughlin et al. (1990)
Proximity to consumers	Sharma (2004)
Proximity to suppliers	Schmenner (1982), Wheeler and Mody (1992)
Community	Spooner (1974), Moriarty (1980), Rees (1983)
Utility facilities	Walters and Wheeler (1984), Gold (1991)
Scope for expansion of site	Spooner (1974), Schmenner (1982)
Quality of life	Coughlin et al. (1990, 1991), Hudson (1983, 1988)
Global competition	Pietlock (1992), Wheeler and Mody (1992)
Union problems	Greenhut (1956), Dicken and Lloyd (1978)

Gothwal(2015), once generalized the factors that affect the decision making in the selection of plant location could be determined by taking all of the important industrial plant sector i.e Oil and Gas and Chemical. By averaging all of the numbers made by years after years of analysing and data gathering, the follow-up trend could be determined. Keep in mind that the generalization of the plant location is an empirical or we could say a general assumption/rule of thumb to make the decision easier as proven in Figure II.10.

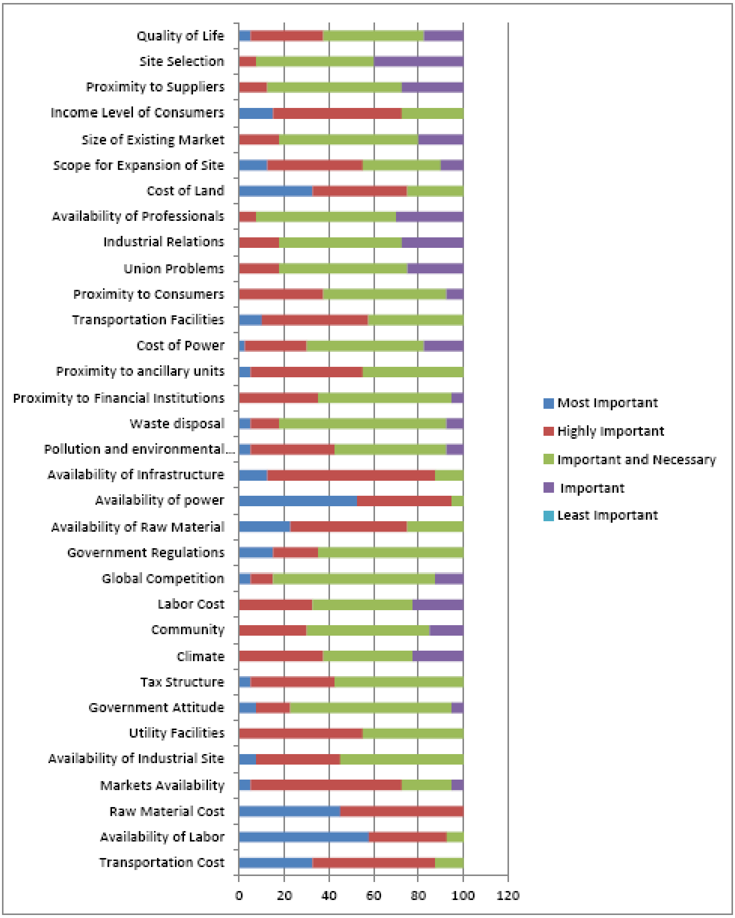


Figure II. 10 Factors affecting plant location decision

Saaty (1980) developed the AHP technique, it is one of the multi-criteria decision-making approach that decomposes a complex problem into a hierarchical order (Abdi and Labib, 2003). Partovi (1994) found it, an effective and practical approach that can consider complex and unstructured decisions. This method incorporates qualitative and quantitative criteria. AHP is an ideal method for ranking alternatives when multiple criteria and sub-criteria are present in the decision-making process. Based on the decision maker's judgements, AHP offers a methodology to rank alternative courses of action concerning the importance of the criteria and the extent to which they are met by each alternative. These judgements are expressed in terms of pairwise comparisons of items on a given level of the hierarchy with respect to their impact on the next higher level. The relative importance of one item versus another are expressed by the pairwise comparisons in meeting a goal or a criterion. Each of the pairwise comparisons represents an estimate of the ratio of the weights of the two criteria being compared. Because AHP utilises a ratio scale for human judgements, the alternatives weights reflect the relative importance of the criteria in achieving the goal of the hierarchy (Maggie and Tummala, 2001). Some benefits of AHP method:

1. The strength of the AHP method lies in its ability to structure a complex,
   multi-person, multi-attribute, and multi-period problem hierarchically (Saaty, 1980).
2. It is simple to use and understand (Chan, 2003).
3. It necessitates the construction of a hierarchy of attributes, sub attributes, alternatives and so on, which facilitates communication of the problem and recommend solutions (Yusuff and Poh Yee, 2001).
4. It provides a unique means of quantify judgemental consistency (Chan, 2003).
5. It does not greatly intuition, experience, and theoretical knowledge of the domain expert as expert system (Yusuff and Poh Yee, 2001).
6. It does not require preferential independent of its complement (i.e., the preference order of consequences, for any pair of attributes does not depend on the levels at which all other attributes are hold) as multi-attribute utility model (Chan, 2003).

The research in this pre-plant design has focused on formulating an AHP-based model to select a location for a manufacturing plant. The theory of AHP is based on the three principles (Raj et al., 2008):

1. decomposition of the decision problem
2. comparative judgement of the various elements
3. synthesis of priorities

According to Saaty (2000), there is no set of procedures for generating the levels to be included in the hierarchy. The degree of detail and complexity of the problem being analysed determines the number of the levels in a hierarchy (Zahedi, 1986). AHP allows the decision problem to be structured into a hierarchy with the objective or goal of the decision placed at the top level of the hierarchy, then criteria and sub-criteria at the intermediate levels and the decision alternatives at the last level of the hierarchy.

To develop the hierarchy, the goal is placed at the top level of hierarchy. The four strategic factors namely, political, economic, environmental factors and facilities used to achieve this goal, which forms the second level of hierarchy. The third level of the hierarchy consists of different factors defining the four strategic factors of the second level. There are three factors related to political factors, five factors related to economic factors, three factors related to environmental factors and 14 factors related to facilities. The strategic consideration factors and factors (criteria) used in these two levels can be assessed using the pairwise comparisons of elements in each level with respect to every parent element located one level above. The fourth level of the hierarchy consists of the rating scale. Here a five point rating scale of outstanding (O), good (G), average (A), fair (F), poor (P) is used and the priority weights of these five scales can be determined using the pairwise comparisons. These rating scales are used to determine the local and global priority weights. The lowest level of the hierarchy consists of the decision alternatives, i.e., different proposed locations to be evaluated to find out the best location. The AHP model is shown in Figure II.11.

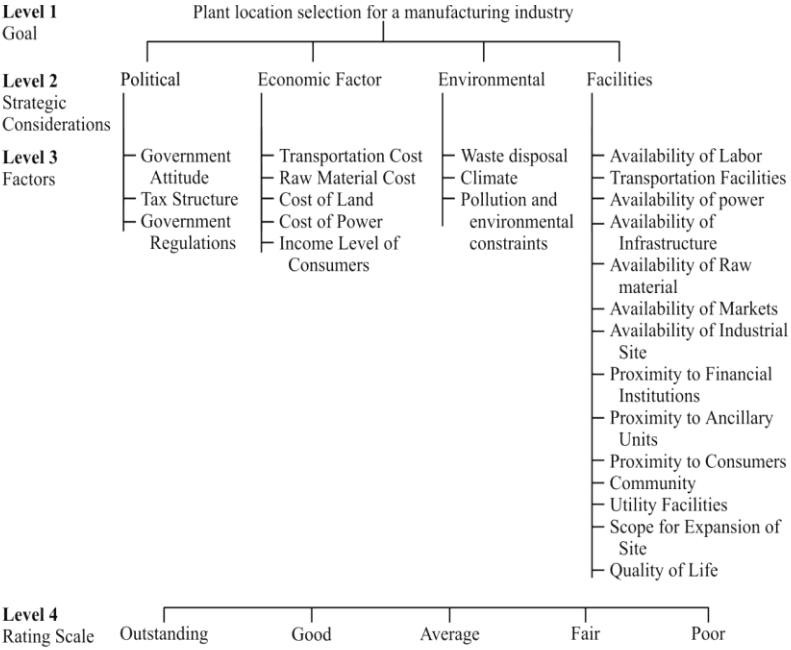


Figure II. 11 Consideration for Plant Selections

Based on the given AHP factors above, due to the limitation and duration of the study, we decide to set a limit into certain aspects, which are divided into 4 criteria and each criteria are containing each sub-criteria

1. Political
    i. Government Regulations
2. Economical
    i. Purchasing Power of Consumer
    ii. Cost of Labor
    iii. Cost of Power

    iv. Transportation Cost
    v. Raw Material Cost
    vi. Cost of Land
3. Environmental
    i. Pollution and Environmental Constraint
    ii. Climate
    iii. Waste Disposal
4. Supporting Infrastructures
    i. Availability of Labor
    ii. Transportation Facilities
    iii. Availability of Power
    iv. Availibility of Raw Material
    v. Availability of Markets
    vi. Availibility of Industrial Site

The choice (or alternative) are selected from potential natural gas reserves in Indonesia as mentioned of figure II.1, where there are 3 alternative for plant locations based on natural gas supplies from the locations such as :

1. Tanjung Api-api, South Sumatera
2. Bintuni Bay, West Papua
3. Sebadai Ulu, Natuna Regency

Where could be simplified into hierarchy diagram as the figure below

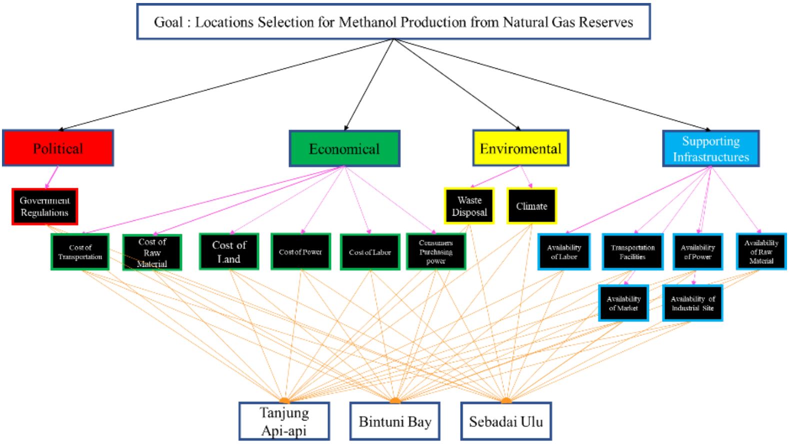


Figure II. 12 Hierarchy Diagram for AHP Methode for Selecting Plant Locations

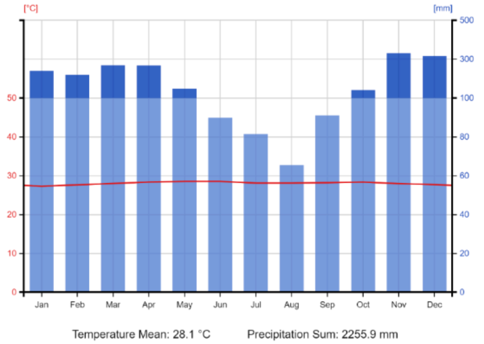
The diagram on the picture below will guide on how judging the decision based on its criteria. Each of the criteria and subcriteria will be weighed in based on available supporting data that owned by each locations. The score for each criteria will affected on the final preferred locations on our methanol plants. Below are defined the available features on alternative plant locations.

II.2.1 Description of Tanjung Api-api

The future candidate plan of development of the plant site, would be located in the area of KEK (Kawasan Ekonomi Khusus) Tanjung Api-Api, South Sumatera Province. The KEK Tanjung Api-Api has quite significant elevation ground with majority of foresty area around KEK Tanjung Api-Api. Hereby enlisted the site and climatic data per August 2019 as shown in Figure II.13.

- Humidity : 65-100%
- Average Air Temperature : 24-31°C
- Average Wind Speed : 4-9 km/hour

- Precipitation : 2255.9 nm



Month	Temp	Precip
Jan	27.3	240.4
Feb	27.6	219.7
Mar	28.0	268.0
Apr	28.4	266.5
May	28.5	147.1
Jun	28.5	89.7
Jul	28.1	81.4
Aug	28.1	65.5
Sep	28.2	91.0
Oct	28.4	140.8
Nov	28.0	330.4
Dec	27.7	315.4

Figure II. 13 Climatic Data in KEK Tanjung Api-Api

Then, the features available on Tanjung Api-api are described on the table below

Table II. 8 Features Available on Tanjung Api-api, South Sumatera

Specifications	Description
Availibility on Industrial Site	• Rubber Processing • Palm Oil • Petrochemical • Logistic
Raw Material Resources	6245.0 BSCF *(Billion Standart Cubic Feet)*
Governments Regulation	• UU no 39 Tahun 1999, about Special Economic Zone (SEZ) • PP no.51 Tahun 2014, about SEZ Tanjung Api-Api

Availibility of Labor	Up to 149,000 labors
Cost of Labor	Rp 3,091,934
Cost of Land	Rp 150,000/m2
Cost of Power	Rp 996.74/kWh
Cost of Raw Material	Rp 2,598/m3
Cost of Transportation	1.55 USD /MMScf

Tanjung Api-api as Special Economic Zone are already prepared by Indonesian goverment with supporting infrastructures such as harbor and airport which make the locations are potential to be chosen as methanols site location.

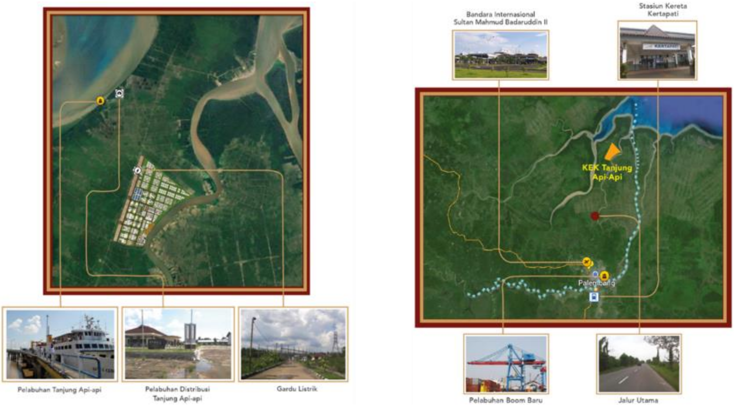


Figure II. 14 Supporting Infrastructures on Tanjung Api-api

II.2.2 Description of Bintuni Bay

Another candidate plan of development of the plant site, would be located in Bintuni Bay, West Papua Province. Hereby enlisted the site and climatic data as shown in Figure II.15

- Humidity : 83%
- Average Air Temperature : 25 ^{o}C
- Average Wind Speed : 11 km/hour
- Precipitation : 25 mm

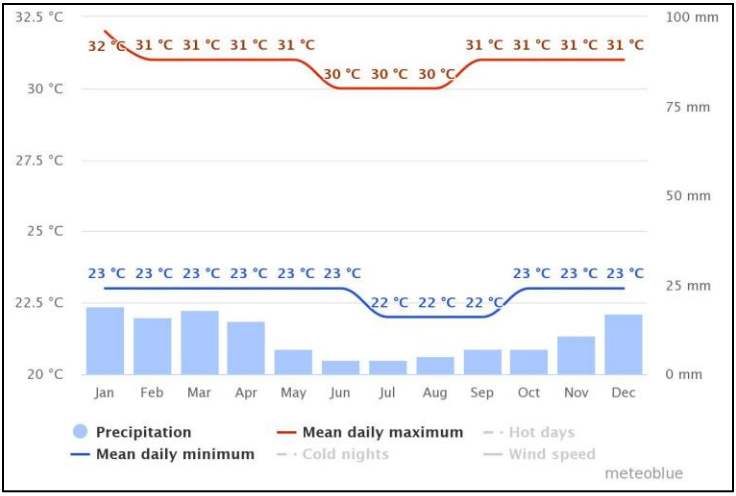


Figure II. 15 Climatic Data in Bintuni Bay, West Papua

Then, the features available on Tanjung Api-api are described on the table below

Table II. 9 Features Available on Bintuni Bay, West Papua

Specifications	Description
Availibility on Industrial Site	• Petrochemical • Fertilizer
Raw Material Resources	13982.7 BSCF *(Billion Standart Cubic Feet)*
Governments Regulation	• UU No 26 Tahun 2002, tentang pemekaran Bintuni • Pasal 29, No.2 BAB XIV, Perda Kabupaten Teluk Bintuni No.1 thn 2019 Tentang Izin Sumber Daya Alam

Availibility of Labor	Up to 20,093 labors
Cost of Labor	Rp 3,134,600
Cost of Land	Rp 150,000/m2
Cost of Power	Rp 996.74/kWh
Cost of Raw Material	Rp 2,598/m3
Cost of Transportation	1.55 USD /MMScf

II.2.3 Description of Sebadai Ulu

Another candidate plan of development of the plant site, would be located in Sebadai Ulu, Natuna Regency. Hereby enlisted the site and climatic data as shown in Figure II.16

- Humidity : 71%
- Average Air Temperature : 26 ºC
- Average Wind Speed : 11 km/hour
- Precipitation : 100 mm

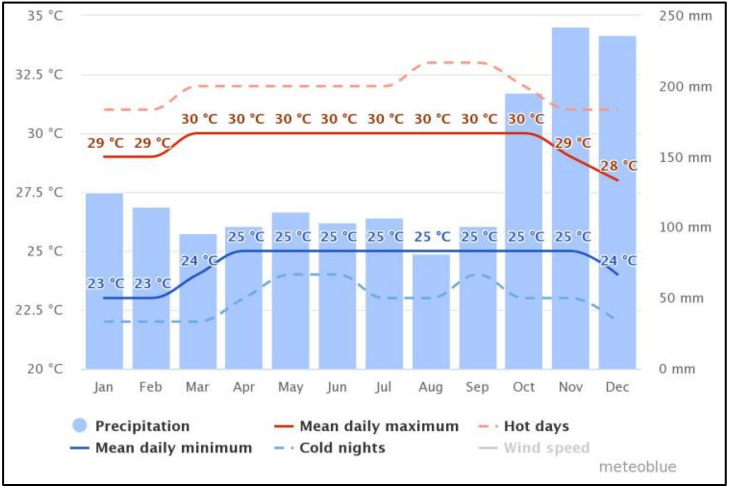


Figure II. 16 Climatic Data in Sebadai Ulu, Natuna Regency

Then, the features available on Tanjung Api-api are described on the table below

Table II. 10 Features Available on Sebadai Ulu, Natuna Regency

Specifications	Description
Availibility on Industrial Site	- Petrochemical - Fischery
Raw Material Resources	1,763.6 BSCF *(Billion Standart Cubic Feet)*
Governments Regulation	- UU No 26 Tahun 2002, tentang pemekaran Bintuni - Pasal 29, No.2 BAB XIV, Perda Kabupaten Teluk Bintuni No.1 thn 2019 Tentang Izin Sumber Daya Alam
Availibility of Labor	Up to 36,901 labors
Cost of Labor	Rp 3,005,383
Cost of Land	Rp 150,000/m2
Cost of Power	Rp 996.74/kWh
Cost of Raw Material	Rp 2,598/m3
Cost of Transportation	1.55 USD /MMScf

II.2.4 Preferred Locations for Methanol Plant Production

Section II.2.4 contain available data that will be used as basis for judging the choice using Expert Choice. The score for each criteria and sub-criteria will be defined by the writers and gained the final score for preferred locations for each available options. Here are the result of the AHP Analysis using Expert Choice as shown below

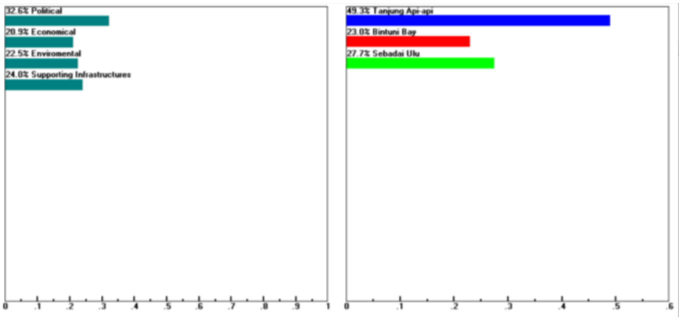


Figure II. 17 Dynamic Sensitivity for Plant Location Selections

Dynamic Sensitivity result shows that political criteria is the major critera on this analysis. This is because on the weighing scale for each criteria, the writer choose to weigh more political criteria due to the impact of regulations owned by the government can affect business because convenient regulations will ease more the plant's establishment and its long-term operating facility. The political products such as corporation tax will impact on the profit. Government can also bring a laws like National Minimum Wage (UMP) which also impacts on profit and labor rights. So, the political criteria is the most important one on this analysis.

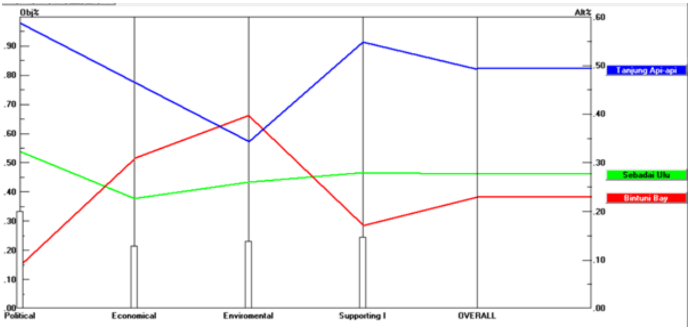
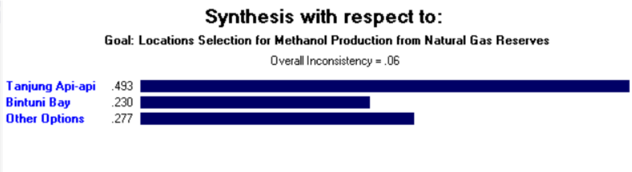


Figure II. 18 Performance Sensitivity for Plant Location Selections

The performance sensitivity result shows the distribution tendency of each criteria for each available options. The variety here are appear because of various input on weighing scale from the data on section II.2.1 until II.2.3. As an example on option Tanjung Api-api shows that supporting infrastructure as dominant criteria, otherwise on Bintuni Bay supporting infrascture are recessive criteria.



Synthesis with respect to:

Goal: Locations Selection for Methanol Production from Natural Gas Reserves

Overall Inconsistency = .06

Tanjung Api-api	.493
Bintuni Bay	.230
Other Options	.277

Figure II. 19 AHP Goal Results on Seleciting Plant Locations

Finally the results are shown on figure II.19, where Tanjung Api-api is the most preferred locations between another 2 available options with score 0.493 and overall consistency 0.06. This means the AHP analysis are valid because the overall consisteny < 0.1. Then, our plant locations will be placed at Special Economic Zone Tanjung Api-api, South Sumatera.

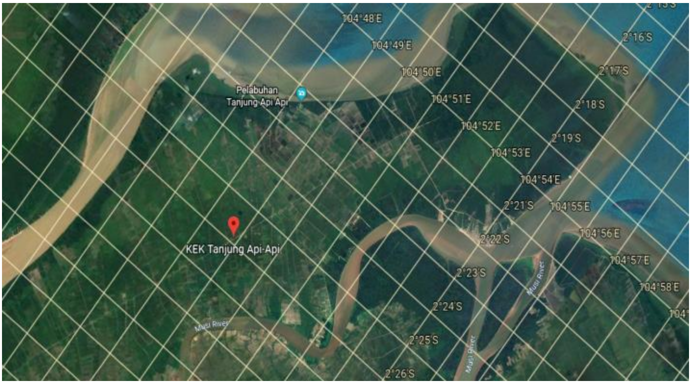


Figure II. 20 Tanjung Api-Api Area

II.3 Specification of Raw Material and Product

II.3.1 Raw Material

Natural gas shall assumed to be a clean sales gas from Sakakemang Block and ready to be convert to Methanol. The sales gas is distributed to the plant by arrangement of pipelines 60 km away from Sakakemang Block. Natural gas is becoming one of the vital components of the world's supply of energy. It is one of the cleanest, safest and most useful of all energy sources. Although we cannot say it is 100% safe, since it also emits lower levels of potentially harmful by-products into the air after burning. The composition of natural gas can vary but below is a chart displaying the typical makeup of natural gas before it is sent for refining.

The criteria of the natural gas feed to satisfy the calculated process design must be no less but not limited to the listed component below as the current specification is based on the Table II. Hereby, we assume that the Natural Gas is a clean and sales gas. According to 'Pembagunan Jaringan Gas Bumi' Report in 2018 issued by Kementrian ESDM, stated that the clean sales gas could be as clean as below:

Table II. 11 Natural Gas Projected Feed Compositions

Component	Mole Fraction
CO_2	0
H_2O	0
Nitrogen	0
Methane	0.92
Ethane	0.04

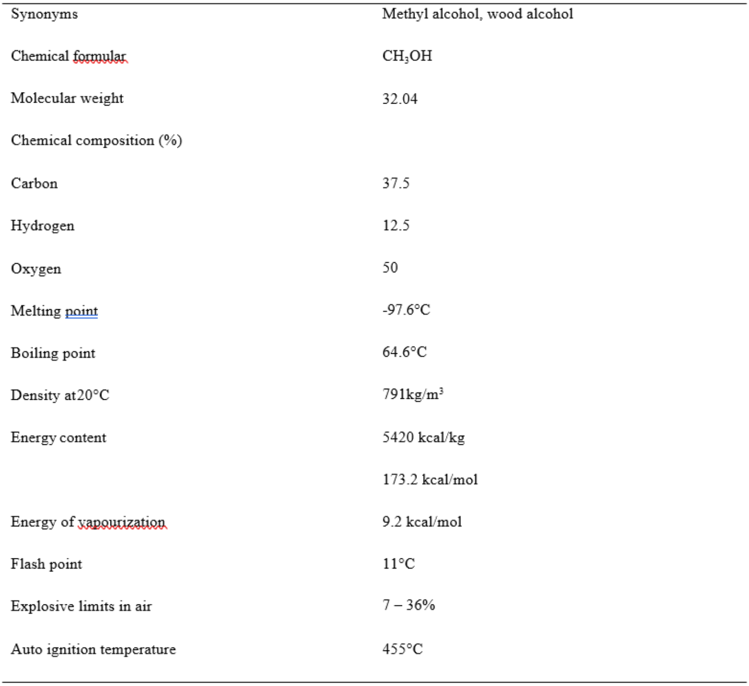
Propane	0.03
i-Butane	0
n-Butane	0.01
i-Pentane	0
n-Pentane	0
Hexane Plus*	0
Total	1.0000

(Kementrian Energi dan Sumber Daya Mineral,2018)

III.3.2 Product

Methanol also known as methyl alcohol or wood alcohol is a colourless, water-soluble liquid with mild alcoholic odour. It freezes at -97.6°C, boils at 64.6°C and a density of 791kg/m^3 at 20°C. It is polar, acid-base neutral, and generally considered non-corrosive. It is miscible with most organic solvents and is capable of dissolving many inorganic salts Methanol can be produced from a variety of sources including natural gas, coal, biomass and petroleum. Methanol also known as methyl alcohol or wood alcohol is a colourless, water-soluble liquid with mild alcoholic odour. It freezes at -97.6°C, boils at 64.6°C and a density of 791kg/m^3 at 20°C as shown in Table II. 8. It is polar, acid-base neutral, and generally considered non-corrosive. It is miscible with most organic solvents and is capable of dissolving many inorganic salts.

Table II. 12 Methanol Properties Properties



Synonyms	Methyl alcohol, wood alcohol
Chemical formular	CH_3OH
Molecular weight	32.04
Chemical composition (%)	
Carbon	37.5
Hydrogen	12.5
Oxygen	50
Melting point	-97.6°C
Boiling point	64.6°C
Density at20°C	791kg/m^3
Energy content	5420 kcal/kg
	173.2 kcal/mol
Energy of vapourization	9.2 kcal/mol
Flash point	11°C
Explosive limits in air	7 – 36%
Auto ignition temperature	455°C

(Arthur. 2010)

The target of the product shall produce a quality of AA grade methanol since that is the requirement of the globacl standard of pure methanol. Shall the specification sheet of the wanted product shown in Table II.9 below:

Table II. 13 Methanol AA Grade Specification

Num.	**Specs**	**Unit**	**Requirements**
1.	Density (20°C)	-	0,7920-0,7930
2.	Platina Color Scale	-	Max Number 5
3.	Distillation Range	-	Maximum of. 1,0°C

4.	Residue of Vapor	g/100 ml	Maximum. 0,005
5.	Odor	-	Signature Alcoholic
6.	Water Content	-	Maximum. 0,1
7.	Acidity	-	maks. 0,003
8.	Aceton Content	-	maks. 0,003
9.	Hydrocarbon	-	Lab Certified
10.	Permanganate Time	minute	min. 50
11.	Chloride Composition	ppm	Maximum of 0.1
12.	Methanol Content	-	min. 99.85

(www.pacific-int.de)

Chapter III

PROCESS SELECTION AND DESCRIPTION

III.1 EXISTING PROCESS TECHNOLOGY

To meet the demands of the future, that makes requirements for process development to achieve sustainable and energy-efficient ways to produce methanol is a one of the grand goal in the development of methanol production. The modern methanol production technologies involved two major catalytic processes; the production of synthesis gas using carbon-based raw materials followed by using the synthesis gas (i.e. syngas) to produce methanol. The principal raw materials for synthesis gas production are natural gas, methane gas from associated petroleum, shale gas, coal and biomass. Synthesis from fossil sources can be successfully achieved by the reforming technologies (i.e. dry and steam reforming) whereas pyrolysis and gasification processes for the biomass-based production option. Belows on table III.1 are listed the summary of feedstocks, process, and its catalyst for producing methanol :

Table III. 1 Feedstocks, Processes, and Catalyst for the Production of Methanol
(Source : Arthur, 2010)

Feedstocks	Process and Main Reactions	Catalyst
Syngas	$CO + 2H_2 \leftrightarrow CH_3OH$	$Cu/ZnO/Al_2O_3$
	$CO_2 + 3H_2 \leftrightarrow CH_3OH + H_2O$	$Cu/ZnO/Cr_2O_3$ $/ZnCr$
	$CH_3OH + CO \leftrightarrow HCOOCH_3$	Potassium methoxide
	$HCOOCH_3 + 2H_2 \leftrightarrow 2CH_3OH$	Cu chromite

Methane	$CH_4 + {}^{1}/_{2}\, O_2(N_2O) \leftrightarrow CH_3OH$	Metal oxides
	Bioprocessing	Enzymes

The very fist methanol technology known as the “high-pressure” process, the first commercial methanol synthesis plant was opened in 1923 in Germany by Badische Anilin–& Soda Fabrik (BASF). The operational point in this process was at 320–450°C and 250–350 bar. (Gallucci and Basile, 2007).

As the time goes by, driven by the economical factor of the production process, the high-pressured process requires so much power to produce, because of this the development reached the peak to invent the so-called “low-pressure” process replaced the old one in the 1960, when the ability to produce sulfur-free synthesis gas made it possible to use the more active copper-based catalysts. This process was operated in the lower pressures and temperatures, namely 60–80 bar and 250–280°C. (Lange, 2001).

By then, up until now, the development of the methanol production keep growing as the time goes by. Either it is the improvement of high-pressured or the enchancement of low-pressured, the sum of these two processes could be divided into 7 processes based on the use of different kind of reactors, which are the core to the process.

The existing technologies for producing methanol, can be listed as:

1. Gas-phase technologies
2. Adiabatic reactors
3. Isothermal reactors (for example, the Lurgi process)
4. Gas-phase fluidized bed converter
5. Liquid-phase technologies
6. Membrane reactors
7. One-step technologies.

Therefore, the methanol production is not just about the methanol production. As this mean, a lot could happen in the midway of the process since the production is a parallel. There are an abundance of oxygen, desalinated water/utility water, nitrogen, argon, and even power to genereate the whole plant itself using the drastic change pressure of the process itself. A couple of interesting examples of the utilization of CO_2 as a feedstock in methanol synthesis have been published. In Iceland, methanol is produced from hydrogen and carbon dioxide utilizing geothermal energy. (Halper, 2011; Olah 2013; Kauw et al., 2015). On the other hand, in Japan the pilot facility for methanol synthesis from hydrogen and carbon dioxide has been running since 2009 (Mitsui Chemicals, 2010).

Very few chemical process are carried out entirely at ambient temperature, where in real industry require process streams to heated or cooled to reach design temperature, add or remove heat of reaction, mixing, adsorption, sterilize feed, etc. Gas and liquid streams are usually being heated or cooled with indirect heat exchanger with another fluid such as : steam, refrigerant, cooling water, or hot oil. On this preliminary design, utilities are divided by 4 services that support production process, consist of :

1. Fluids for process heating or cooling
   Include of : steam, cooling water, and refrigeration system
2. Process Water
   Include of : water (for general use), and demineralized water
3. Compressed air
4. Inert-gas supplies

(Towler and Sinnot, 2012)

As for the desalinated water/utility water, there are a lot of options to describe for, but could be sum into 2 types of process which are the most advanced process up until now and set the base of water treatment development process. They are Ultrafiltration and Reverse Osmosis. The ultrafiltration using a very small filters up to 100nm and reverse osmosis using the same exact mechanism to proceed the water, the sole difference is just that the reverse osmosis process using smaller filter to prevent more impurities that ultrafiltration couldn’t filter. (Lau,2013).

For the nitrogen, oxygen, and argon are basically a main product in A.S.P (Air Separation Plant) in order to utilize the main process of methanol production. There are two different technologies, existing and emerging. The existing are the ones that achieved ‘Technology Readiness Level’ or TRL=9, that were commercially being used across various industries. While, the emerging technologies are TRL=6 and lower. Because of its avaibility, we summarize the basic two of most applicable air separation technology. (Sandeep, 2018)

Those two are pressure swing adsorption and rectification-column based process,. The progress continues up to the point of rectification-column based get upgraded to cryogenic and the rest is history, the cryogenic process could separate air up to the argon purification to purest form. This by far, are the most advanced technology in air-separation process. (Linde, 2018).

Cryogenic air separation is an old process used to produce high purity oxygen or nitrogen at high volumes. Cryogenic separation is most effective when high volumes of oxygen are required (>=102 tonnes of oxygen/day), or high pressure oxygen is required. The technology is centered on the fact that each of air’s constituent has different boiling point. The idea behind the process is to lower the temperature of the air such that nitrogen and oxygen separate based on their boiling point. In addition, liquid argon, liquid oxygen, and liquid nitrogen can be added to the product slate for stored product backup or byproduct sales at low incremental capital and power cost. For liquid oxygen, the separated cold nitrogen is used as heat transfer fluid to further cool the oxygen. The thermodynamic minimal work of oxygen separation from air is equal to 58.4 kWh/tonne of oxygen. Energy required to separate oxygen from air is 180 kJ/kg. The best presently constructed cryogenic ASU are characterized by energy consumption exceeding the thermodynamic minimum

Energy consumption of oxygen separation is an increasing function of oxygen purity. The oxygen purity decrease to 95% is usually accompanied by energy consumption reduction of at least 10%. Taking into account a further possible increase of cryogenic ASU capacity, and the efficiency increase resulting from the lower oxygen purity (95%), the expected energy demand of the separated oxygen will be about 165 kWh/tonne (540 kJ/kg). Further decrease of energy consumption is unrealistic as the capacities of ASU single trains cannot be increased much because of transport and assembly problems

Additionally, since cryogenics can produce such a high purity oxygen, the side product as nitrogen stream is usable quality. This can add significant financial benefits to a process integrated with a cryogenic air separation system. Below is table that shows single unit and largest commissioned ASU plants from different gas companies

Air Separation Unit (ASU) with Cryogenic consist of five units operation :

- Feed air compression unit
- Feed air pre-treatment unit (purification unit)
- Heat exchange and Liquefaction unit
- Cryogenic separation unit
- Product compression unit

Where air compression unit is located at the front end of this process, followed by air treatment unit, which is considered as the safe guard of the plant from the risk of hydrocarbon and water trace. Next, heat exchange and liquefaction is the core of ASU, where the heat of the feed air is transferred to liquid product and become gases, where the feed itself become liquid air. After that, following to heat exchange step, cryogenic separation, where liquid air distilled to different products as oxygen and nitrogen which are compressed in the compression unit to meet customer need.

At least there are 3 cycles of cryogenic ASU :

1. **Compression Cycle**
   Air separation process typically produce gaseous product streams at slightly above atmospheric pressure and near ambient temperature. Typically the product oxygen leaves the main heat exchanger at low pressure, ranging from 0.5 to 10 psig, and a centrifugal compressor train with relatively high inlet volumentric flow rate delivers the product at the required pressure. Many hundred of these plant operate worldwide.
2. **Pumped Liquid Cycle**
   Liquid product can be pumped from the distillation section upstream of the cryogenic heat exchangers for vaporization and warming. These product may be pumped to the required delivery pressure or to an intermediate pressure. However, since producing liquid product from the distillation system required 2-3 times power supplied than producing gaseous product, the cycle must efficiently recover the refrigeration contained in the pumped product stream. This can be accomplished by condensing an air or nitrogen feed stream at high pressure against the vaporizing product stream in the cryogenic heat exchangers, then the liquified air or nitrogen feed returns the refrigeration to distillation section.
   Pumped liquid process cycles in which the product stream is pumped to an intermediate pressure at the ASU outlet are called partial pumped liquid cycles and require additional equipment to compress the product streams to the final delivery pressure. Full or partial pumping of the product streams add another degree of freedom in optimizing the cryogenic cycle, and can eliminate or reduce the size of oxygen compressor.
3. **Low and Elevated Pressure Cycle**
   Low-pressure (LP) ASU cycles are based on compressing the feed air only to the pressure required to reject the nitrogen byproduct at atmospheric pressure. As such, feed air pressures will typically vary between 65 and 100 psia, depending on the oxygen purity and the level of energy efficiency desired. Eleveated-pressure (EP) ASU cycles produce product and

byproduct streams at well above atmospheric pressures and generally require smaller and more compact cryogenic components that can be cost effective. EP cycles typically used feed air pressure in excess of 100 psia. An EP cycle may be appropriate when all or nearly of the nitrogen byproduct will be compressed as a product stream. In addition, an EP cycle is often selected for integration of the ASU with other process units such as gas turbines

Back to the main process, the process route for the production of methanol is relatively straightforward, involving three following basic steps :

1. Production of synthesis gas (syngas)
2. Conversion of syngas into crude methanol
3. Distillation of the reactor effluent (as crude methanol) to achieve desired purity

Which are going to be explained in the subchapter below.

III.1.1 Syngas Production

Syngas is a general term used in describing a mixture of Hydrogen and Carbon Monoxide in different ratios and can be generated from any hydrocarbon feedstock, like natural gas or biomass. Direct synthesis processes (selective oxidative transformation of methane) have the major advantage of avoiding the energy intensive step of syngas production, but are technically difficult to accomplish. The main disadvantages refer to a low conversion of the feedstock and the production of undesired by-product, due to a higher reactivity of the oxidation products themselves compared to methane. The synthesis gas preparation and compression typically accounts for about 60% of the investment, and almost all energy is consumed in this process section.

Although direct processes are getting more into focus of research, industrial methanol is exclusively produced by indirect conversion routes via syngas production. In general, a synthesis route comprises a pretreatment of natural gas, a reforming unit for syngas, a gas conditioning section, a low pressure methanol synthesis unit and several distillation columns for product purification. Natural gas reforming also known as reforming of methane can be accomplished by means of an exothermic or endothermic reaction depending on the chemical process selected to perform catalytic reforming of methane. There are basically four different types of processes that can be used to carry out the reforming methane. The variety of available reforming technologies, like :

1. SMR (Steam Methane Reforming)
2. ATR (Auto Thermal Reforming)
3. DMR (Dry Methane Reforming)
4. PoX (Partial Oxidation)

III.1.1.1 Steam Methane Reforming (SMR)

This technology has been most predominantly used in which steam and methane are converted catalytically and endothermically (in the presence of a catalyst) to Hydrogen and Carbon Monoxide. The scheme of the reaction of steam reforming is shown below :

$$\mathrm{CH_4 + H_2O \rightarrow CO + 3H_2 \qquad \Delta H = 206\,\frac{kJ}{mol}}$$

$$\mathrm{CO + H_2O \rightarrow CO_2 + H_2 \qquad \Delta H = -41\,\frac{kJ}{mol}}$$

Because of the process steam reforming of methane is the reforming process that leads to the obtaining of syngas with the major H_2/CO ratio, this type of reforming process is considered ideal to obtain a gas hydrogen flow of high purity from syngas. The steam reforming of methane is an endothermic process and requires very high temperature which makes its process very expensived based on its energy needed.

The process of steam methane reforming procudes syngas $(H_2 + CO)$ with a ratio $H_2/CO = 3$. SMR is the reaction where steam and hydrocarbon such as natural gas react in the reformer at temperature of $800 - 900°C$ and moderate pressure (around 30 bar) in the presence of metal based catalyst. Most SMR units include two sections, namely a radiant and a convective section. Reforming reactions take place inside the radiant section. In the convective section, heat is recovered from the hot product gases for preheating the reactants feed and for generating superheated steam. The advantage of this process is lead to obtain syngas with the highest H_2/CO ratio, this type of reforming process is considered ideal to obtain a hydrogen gas flow of high purity from syngas.

III.1.1.2 Autothermal Reforming ATR)

Addition of oxygen to the steam reforming process is an alternative measure in obtaining lower H_2/CO ratio. Autothermic reforming is the reforming of light hydrocarbon in a mixture of steam and oxygen in the presence of catalyst. The reactor is designed with a refractory lined vessel, therefore higher temperature and pressure can be applied than in steam reforming. This ATR cannot be used alone, therefore a pre-reformer is installed downstream where typically 35 to 45% of the reforming reaction occurs. The ATR converts the remaining methane from the pre-reformer where air is used to supply the required oxygen

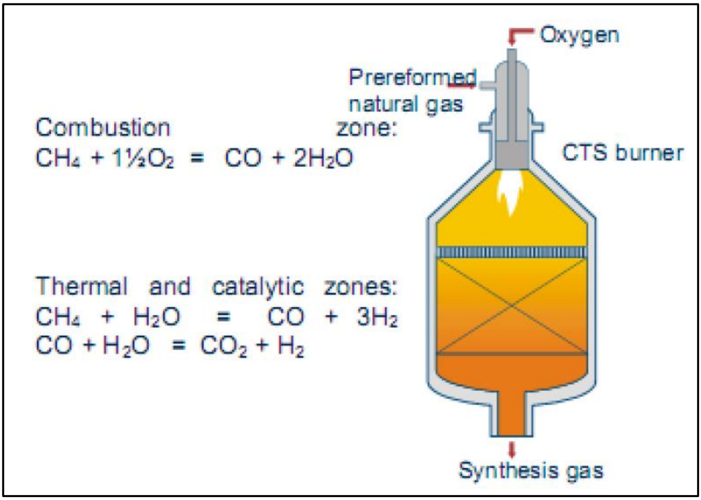


Figure III. 1 Reaction Inside Autothermal Reactor

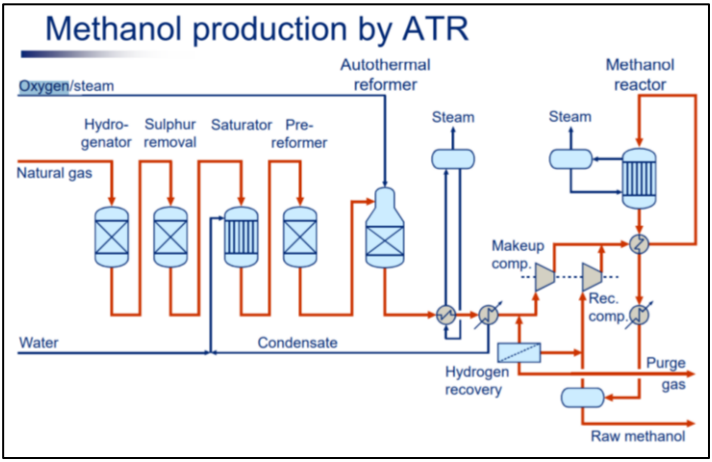


Figure III. 2 Example of Methanol Production Scheme by Haldor Topsøe

The very high temperature of Autothermal Reforming require installed catalyst that has excellent thermal stability. Haldor Topse offers 2 type of catalyst : RKA-02 and RKS-2 series. RKA-02 is a nickel based catalylst on a stabilized alumina carrier (that make this catalyst suited at extreme condition) used for top layer. RKS-2 catalyst is also nickel based catalyst, but has a carrier composed of Mg-Al spinel (that have higher melting point and thermal stability than alpha-alumina). Below are the picture of ATR catalyst produced by haldor topsoe :

Figure III. 3 Catalyst Produced by Haldor Topsoe for Autothermal Reforming

III.1.1.3 Dry Methane Reforming (DMR)

Oxidation of methane for production of syngas has gone from more oxidizing agent (oxygen) to a lesser one (steam) with pros and cons for each of the approaches. Carbon dioxide can be used as oxidizing agent via reaction called Dry Methane Reforming

$$CH_4 + CO_2 \leftrightarrow 2CO + 2H_2 \qquad \Delta Ho = 247\ kJ/mol$$

Where thermodynamic for DR are not as favorable as the ATR or SMR reactions, however proportional consumption of one mole of carbon dioxide per mole of methane invested could reduce the carbon impact which could lead to a greener consumption of natural gas (by using carbon dioxide that converted to syngas)

III.1.1.4 Partial Oxidation (PoX)

Partial oxidation is a process when methane (in natural gas) reacts with oxygen in partially oxidizing reaction to generate syngas, but also the reaction is exothermic that releases heat to drive this process.

$$\mathrm{CH_4} + \frac{1}{2}\mathrm{O_2} \rightarrow \mathrm{CO} + 2\mathrm{H_2}$$

III.1.2 Methanol Synthesis

Two main types of reactors are used today in synthesis methanol, adiabatic and isothermal, and their use differs according to working conditions. The adiabatic systems commonly include a series of fixed bed reactors, with removal of the heat, operated downstream of each reactor. These kind of reactors are characterized by low installation cost and high production capacity. However, because of adiabatic process the high equilibrium temperatures imply very low conversions for each cycle and consequently a high recycle ratio, high dilution of reagents, and high volume of catalyst.

The isothermal reactor is continuously cooled through another source, water, or gas. It is primarily a heat exchanger in which cooling is achieved by water circulation on the mantle of tube bundle. The isothermal nature of this system enables obtaining high conversions and a low volume of catalyst. However, to reach a useful reaction rate, the temperature needs to be $240°\mathrm{C} - 260°\mathrm{C}$, with a high recycle ratio. The installation costs are much higher than adiabatic system, and the size of the plants are limited because of the tube bundle. Such a scheme was developed and customized by several industries, the example one is Lurgi.

Methanol synthesis can be achieved by reduction of $CO \; or \; CO_2$, according to several reactions mechanism. For the conversion of CO_2 into methanol, direct conversion route using

CO_2 and H_2 was identified as more efficient than routes via CO. Both reactions have an exothermic nature and generare a reduction in moles number. For these reasons, low temperature and high pressure are promoted for reactions. Such conditions counter several issues in terms of reaction kinetics and a plant's fixed and operative costs. Methanol is currently produced on an industrials scale based on catalytic conversion of synthesis gas. Process are classified according to the pressure used :

1. High Pressure Process , 25-30 MPa
2. Medium Pressure Process, 10-25 MPa
3. Low Pressure Process, 5-10 MPa

The main advantages of the low pressure process are lower investment and production costs, improved operational reliability, and greater flexibility in the choice of plant size.

Another considerations for choosing methanol reactor is its heat removal performance. Because of synthesis reaction is strongly exothermic, heat removal is important step and temperature control must be important. The reactor technologies used extensively in commercial settings fall into two broad categories :

1. Multiple Bed Reactors
2. Single Bed Converters

In multiple catalyst bed reactors control the reaction temperature by separating the catalyst mass into several sections with cooling devices placed between the sections. In single bed reactors, heat is continuously removed from the reactor by its transfer to a heat removing medium so the reactor runs effectively as a heat exchanger. The advantages of these technologies are the longer lifetime of the catalyst, the higher yield due to the methanol-favouring equilibrium, and the energy recovery in the coolant. Figure below shows an example of the methanol production within an isothermal reactor. In comparison to the other two rectors shown in figure below, the gas within the isothermal reactor never reaches high temperatures resulting in the less-consumed energy by the process itself.

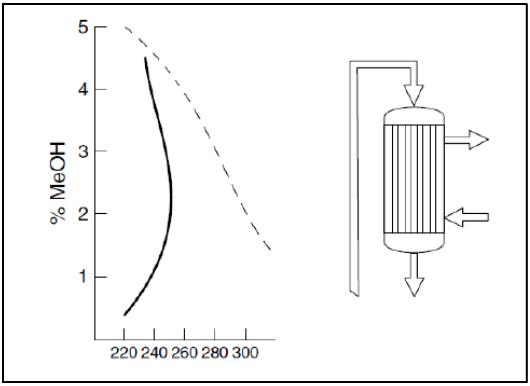


Figure III. 4 Conversion Profile (Solid Line), Equilibrium Curve (Dashed Line), and Schematic for an Isothermal Reactor with Indirect Cooling

Lurgi (as company) aimed at producing methanol in a single-train plant starting from natural gas and capable of increasing the capacity of an existing methanol plant based on steam reforming. Methanol synthesis takes place under quasi-isothermal conditions in the Lurgi water-cooled methanol reactor, in which a fixed bed catalyst is placed in tubes surrounded by boiling water. The reactor is fed by syngas produced in reforming section, mixed with recycled gas from flash separator (gas liquid separator). The recycling of unreacted gas is unavoidable because of the very low kinetic of the conversion reaction. The effluent stream is cooled down to 100℃ to separate methanol and water from the unreacted process stream. Methanol and water then separated in distillation units.

The noteworthy reactor used in the process could result in two types of low-pressured process. A LURGI (50 bar of pressure) plugged flow reactor or a ICI (100 bar of pressure) bed reactor; depending on the use of pressure before the inlet of the reactor. Due to supplied natural gas volume (100 MMSCFD approx), we define 1x100% capacity of process arrangement . Hereby, this process arrangements is applied, quoted, and calculated for LURGI low-pressured plug flow reactor. The use of single reactor unit without a spare in the Methanol Synthesis Process. This process is a highl-cost reduced in terms of CAPEX reduction, but at the same time reduce the lifetime of the reactor faster than other options, also the maintenance cost would be very high since coke deposition would be likely happen very often because of the non-stop use of maximum capacity in a year. Most of the actual process for methanol synthesis operate in gas phase and use copper-based catalyst. In particular, the most common catalyst adopeted in industry is a $CuO/ZnO/Al_2O_3$ formulation and its composition depends on the menufactur below :

Table III. 2 Composition of Methanol Synthesis Catalyst from Different Manufacturers (Source : Bozzano and Manenti, 2016)

Manufacturer	Cu (wt%)	Zn (wt%)	Al (wt%)	Other
IFP	45-70	15-35	4-20	Zr 2-18
Synetix	20-35	15-50	4-20	Mg
BASF	38.5	48.8	12.9	Rare earth oxide-5
Shell	71	24	12	
Sud Chemie	65	22	31	
DuPont	50	19	17	
United Catalyst	62	21	17	
Haldor Topsoe MK-121	>55	21-25	8-10	
Mitsubishi Gas Chemical Company	63.6	33.4	3	
Ammonia Casale	30	50	3	Cr (16)
Lonza	40	20		Zr (40)

AIST, RITE	45.2	27.1	4.5	Zr (22.6) Si (0.6)
YKK Corporation	76.3	11	12.7	

Cu is usually regarded as the active site, whereas very small amounts of Zn atoms are reduced from ZnO and decorated on Cu nanoparticles at the interface. In particular, ZnO presence along with good interface between Cu and ZnO improves copper dispersion and CO_2 adsorption. With the suggested reactor package, we suggest to use the CuZnO (with high contain of CuO) catalyst in methanol synthesis resulting in:

- Activation: 2CuI (H_2O) + ½ O2 → CuIIOCuII + H_2O
- Reaction: CuIIOCuII + CH4 → Cu2 I (CH3O-) + H+ (zeol.)
- Hydrolysis: Cu2 I (CH3O-) + H+ (zeol.) + H_2O → 2CuI (H_2O) + CH3OH

III.2 PROCESS SELECTION

With all of methanol technology production that have been explained above, we propose 4 considerations for choosing Syngas combined with Methanol Synthesis technology :

1. Financial Wise Consideration
2. Performance Wise Consideration
3. Risk, Durability, and the Fabrication Time
4. Specific Power or Energy Consumption by the Package

For detailed explanation we define our consideration for choosing methanol synthesis technology as below :

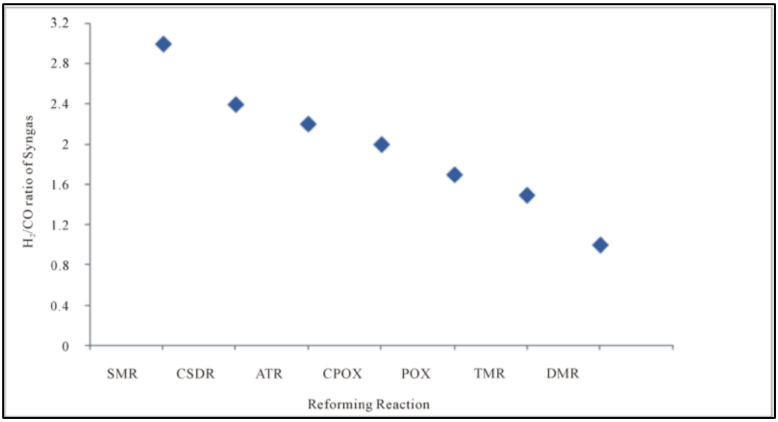


Figure III. 5 H2/CO ratio of syngas from various syngas generators

On the picture above we show various of syngas reforming technology and its H_2/CO ratio that describe ‘module’ that become requirement for syngas to produce methanol. even it is mentioned that ‘module’ should be reached 2 (minimal), but many journals said value 2-3 is acceptable. We could see in SMR produce highest value of it followed by ATR, POX, and DMR. The variety of available reforming technologies differs in construction, in supply of reforming agents and in operation conditions. The SMR technology is typically applied to plants with maximum capacity 2500 MTPD, while oxygen based technologies cover the range above.

Table III. 3 Main Reactions on Several Reforming Process

Reforming Type	Main Reactions	ΔHo
SMR	$CH_4 + H_2O \rightarrow CO + 3H_2$	206 kJ/mol
	$CO + H_2O \rightarrow CO_2 + H_2$	-41 kJ/mol
ATR	$CH_4 + 1.5O_2 \rightarrow CO + 2H_2O$	-520 kJ/mol

	$CO + H_2O \rightarrow CO_2 + H_2$	-41 kJ/mol
	$CH_4 + H_2O \rightarrow CO + 3H_2$	206 kJ/mol
DMR	$CH_4 + CO_2 \leftrightarrow 2H_2 + 2CO$	247 kJ/mol

In comparison to SMR, the use of of an ATR reduces the steam requirement and therefore results in lower mass flows and smaller equipment size of the downstream components. Additionally, the heat of reaction can be integrated to cover heat requirement of the pre-heating unit and other heat consumers.

(Blumberg et al, 2017)

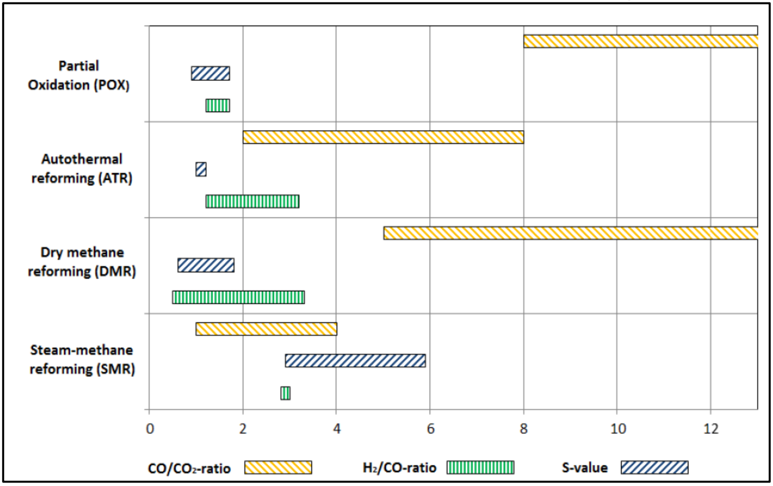


Figure III. 6 Range of the syngas composititon for the commercially applied methane reforming technologies. The composition is expressed by the three main stoichiometric modules (Source : Blumberg et al, 2017)

The bar chart above describes that in POX, ATR, and DMR produce highly reactive syngas due to high CO/CO_2 ratio, but is deficient in hydrogen content. In ATR, addition of oxygen to the steam reforming process obtain lower H_2/CO ratio. Autothermic reforming is the reforming of light hydrocarbon in a mixture of steam and oxygen in the presence of catalyst. ATR can’t be used alone, therefore a pre-reformer is installed downstream where 35-45% reforming reaction occurs. Then, the ATR converts the remaining methane from pre-reformer.

Syngas generation in a stand-alone ATR unit consist of a fired heater for preheating of feed gases and an oxygen-blown ATR reformer. In general, ATR technology is favored over conventional reforming options under the same conditions that favor of the use of oxygen-blown secondary reforming. Savings in CAPEX as well as operational expenses (OPEX) strongly recommended ATR technology as the optimal solution for large-scale methanol plants. Very recent studies suggest that CAPEX and OPEX are also lower when ATR technology is applied to medium to large-scale methanol plants (more than around 2,500 MTPD) that commonly operate with two-step reforming. Here are the comparison of typical operating parameters at 5,000 MPTD grade-AA methanol plant with ATR and two-step reforming technology at same composition and pressure in natural gas feed.

Table III. 4 Comparison of Key Operating Parameters
(Assumpstion : for a 5000 MTPD Methanol Plant)

Operating Parameter	ATR	Two-Step
S/C	0.6	1.8
Specific O2 Consumption per unti MeOH, index	100	69
Reforming and Pre-heating Duty	100	297
CO/CO2 Ratio in Syngas	5.4	3.0
H2O in Raw Product, wt%	7	12
MU Gas Compressor Size, index	100	132

Recycle Gas Compressor Size, index	100	96
Carbon Dioxide Release, index	100	143
Energy Consumption, index	100	101
ISBL Cost Index, incl A.S.U	100	110

As illustrated on tabel above, the net result of the comparison is a savings in CAPEX of 10% with similar or better total energy consumption. Other major process difference include that of S/C ratio and of oxygen consumption.

Another table show the results between syngas produced from ATR and two-step reforming viewed from its M value and compositions

Table III. 5 Syngas Composition and its Module Value from ATR and Two-Step Reforming

Component	ATR	Two-Step Reforming
H_2	64.1	68.1
N_2	1.6	1.3
Ar	0	0
CO	27.0	22.2
CO_2	5.0	7.4
CH_4	1.8	0.4
H_2O	0.5	0.6
M	**1.84**	**2.05**

In the case of two-step reforming, the ratio M of produced syngas is close to 2, whereas the ATR unit produces syngas with M<2. If ‘M’ in the syngas deviates form the requirement (M=2), then either surplus hydrogen or surplus carbon oxides has to be removed from the synthesis loop. The excess purge is typically used is typically used as fuel or as hydrogen source and recycled to the make-up gas. M is however independent of the shift equilibrium equation and should therefore not be used to describe the reactivity of syngas. For this purpose, we use CO/CO_2 ratio, where its high ratio will increase the reaction rate and the achievable per-pass conversions. In general, ATR leads to the highest CO/CO_2 ratios in syngas. With an efficient hydrogen recovery unit, the additional purge required to remove excess carbon is achieved without extra loss of hydrogen. Then, reaction efficiency remains unchanged at approximately 94-96%.

(Dahl, 2014)

Table III. 6 Summary of each Natural Gas Reforming Selection (Source : Wang and Rohr, 2002 ; Xu et Al, 2017 ;

Reforming Type	Listed Specification	Characteristic	Advantages	Disadvantages
Steam Reforming	1. Reaction Sets : $CH_4 + H_2O \rightarrow CO + 3H_2$ $CH_4 + 2H_2O \rightarrow CO_2 + 4H_2$ $CO + H_2O \rightarrow CO_2 + H_2$ 2. Operating Conditions :	Hydrogen Yield	Generally higher than 50% at $T > 600°C$ for M=1	Potential high level of carbonaceous material formation
		Heat Requirement	none	External heat transfer device is required, therefore results in system complexity and

	Temp : 800 – 1000℃ Pressure : 20-30 bar			potential higher cost
		Startup /transient	relative stable during transition operation	still needs external igniter to start up although the catalyst bed can be used for catalyst combustion tentatively. Heat transfer efficienct and higher volume makes the start-up begin slow.
Autothermal Reforming	1. Combines the SMR reactions and POX reaction to provide a nearly thermodynamicaly neutral reaction, by utilizing the heat generated in the POX to	Hydrogen Yield	about 50% concentration	lower hydrogen yield that steam reforming
		Heat Requirement	Exothermic, excess in energy that can be used for another equipment	may need startup heat, and control systems to switcj between lean burning and

	meet the heat required by the SMR 2. Combined reaction : $CH_4 + 0.5\, H_2O + 0.25\, O_2 \rightarrow CO + 2.5\, H_2$ 3. Operating Conditions : Temp :$900 - 1100℃$ Pressure : 20-30			ATR regimes
		Startup /transients	Moderate. Can be set up to fast response times by switching between catalytical P.O.X and ATR (relying in CPOX portion for the faster response time)	Transient fluctuations for load matching may be as much as 1-10 per second. Such deviations will reflect on efficiency level if we are switching between CPOX or ATR for responding to transients

Catalytical Partial Oxidation	1. Reaction Set : $CH_4 + 1/2\ O_2 \rightarrow CO + 2H_2$ 2. Operating Conditions : Temp : $700 - 900$℃ Pressure : 20-30 bar	Hydrogen Yield	none	Relatively low yield be tuned by improving catalyst and convert some CO back to H2
		Heat Requirement	No external heat required. The system is exothermic	The heat generated from the reaction needs to removed or utilized in the system
		Startup/Transients	Startup is fast. Transient test is relatively easy to control	High temperature startup/shutdowns may cause catalyst degradation

We finally choose **Autothermal Reforming** as our reforming technology, one of the reason is because its capability that can self-producing power (as electricity) by using its excess heat to support steam-powered generator. ATR technology with the analysis above establish it as the optimal choice for syngas production in large-scale methanol plants where its features are :

- High CO/CO_2 ratios in the syngas, which result in higher methanol reaction rates
- Low steam requirements, which lead to lower CAPEX and OPEX in methanol plants with capacities above 2,500 MTPD
- Low steam throughput, which enables larger single line capacity

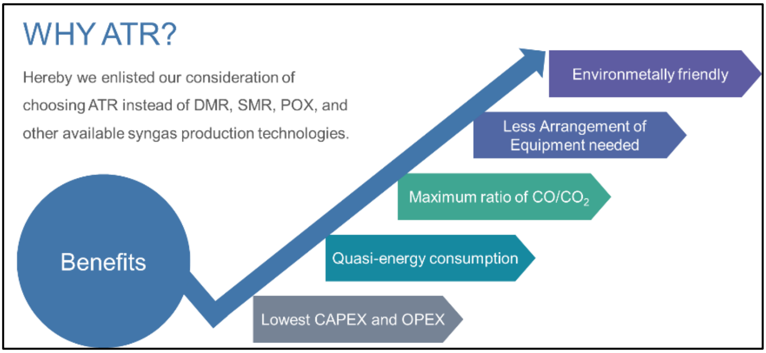


Figure III. 7 Simplified Consideration for Choosing Autothermal Reforming

Then we continue to select methanol reactor technology that described below. We first compare methanol reactor based on its technology, adiabatic and isothermal. As mentioned before, adiabatic reactor are characterized by not having an external cooling system within their reaction zone. Other type of adiabatic reactor are quench reactor, a reactor that cool the reaction gas by adding cold reaction gas throughout the reactor. Isothermal reactor is type of reactor that keep temperature constant at a low level through cooling, realizing a quasi-isothermal process. In short, the tables below describe comparison of each reactor type :

Table III. 7 Summary of each Methanol Reactors Type
(Source : Basile and Dalena, 2018)

Characteristic	Adiabatic Reactor	Quench Reactor	Isothermal Reactor
Cooling mechanism	none	cooling by addition of cold gas reaction throughout reactor	cooling the reactor, catalyst, and reaction gas. Cooling is done indirectly with GCR (gas cooling reactor) or with steam generation
Installation Cost	Low installation cost		higher than adiabatic reactor
Production Capacity	High		High
Conversion	Low conversions, due to high equilibrium temperature		High
Catalyst Load and lifespan	High		Intermediete, longer lifetime of catalyst
Temperature	270℃ and higher		240℃ − 260℃
Pressure Drop	Relatively high		Relatively small
Operatibility	Easy operation		More complicated than adiabatic reactor
Application	Mostly applied for smaller case industry		Mostly applied for higher production case industry

Table III. 8 Consideration of each Methanol Reactor Selection

Reactor Technology Manufacturers	**Listed Specifications**	**Pros**	**Constraints**
Johnson Matthey/Davy Process Technology	1.Operating pressure 50 to 100 bar 2. 270^o C 3. Quench Reactor 4. Flow in axial or radial direction 5. Useful in small scale productions	1.Low operating pressure resulting in reduced CAPEX and OPEX 2. Low-temperature required, resulting in quasi-energy consumption 3. Two options of flow direction: Axial(Simple but large) or Radial(Complicated design but compact resulting in the increase of fabrication price). 4. Useful in small scale plant	1. Only applicable in small scale plant; 1500 ton/day

		5. Comes with a package along with the syngas synthesis and refining instruments.	
Casale and ICI	1. Similar to Johnson Matthey/Davy Process Technology 2. Improved quench gas redistribution, very useful in IGCC 3. Separated catalyst beds 4. Improvement by 20% in yield	1. Better performance in terms of conversion rate rather than Johnson Matthey/Davy Process Technology by 20% margin of yield 2. Less risk in terms of fouling regarding in the separated catalyst beds from the reactor, particular places to store the catalyst are	1. Very expensive and only optional available if there is any IGCC improvement 2. Improvement of the conversion rate is questionable

		provided by the manufacturers 3. IGCC options available, for future plan development in Integrated and Gasification Combined Cycle.	

Haldor-Topsoe	1. Collect-Mix-Distribute Reactor 2. Quench Reactor 3. Radial flow through catalyst bed, and axial flow to quench 4. Catalyst beds arranged plates and separated by vertical support beams 5. Increased conversion and better temperature control	1. Integrated transportation to ease the distribution of product 2. Mix of radial and and axial resulting in higher efficiency of process, simplified fabrication to reduce the fabrication time also to ease the maintenance process and increasing the durability of the reactor 3. Less risk in terms of fouling regarding in the separated catalyst	1. Complicated fabrication method resulting in the long period of time fabrication. Estimated time of fabrication and delivery is 8-10 months.

		beds from the reactor, particular places to store the catalyst are provided by the manufacturers 4. Improvement in conversion rate and percentage control by improvement in temperature control.	

Kellog, Brown, and Root	1. Series of adiabatic reactors with intermediate cooling 2. Spherical shape 3. Gas enters outside towards inner shell 4. Catalyst loaded around inner shell 5. Thin external walls possible	1. Improvement in conversion rate by adding several adiabatic reactors with integrated cooling system resulting in higher conversion percentage 2. Spherical shape is used in order to maintain and spread the existing pressure cordially and same in every point of the reactor 3. Thin external walls to ease the fabrication	1. The increase of risk regarding to the use of thin external walls in the reactor

		process, reducing the time of fabrication, and simplified transporation mechanism	

Toyo Engineering Company	1. Radial flow with concentric catalyst beds 2. Intermediate cooling 3. Boiling water cooling 4. Good Temperature Control 5. Catalyst bed 30% smaller than conventional quench reactors 6. Good heat recovery 7. Low pressure drop independent of reactor height 8. Easy scale-up simply by increasing or adding	1. Better quality product control and efficient energy consumption regarding the improvement in good temperature and good heat recovery. 2. Good heat removal regarding the use of BWR, useful to increase the conversion of the product 3. Could contain more catalyst bed regarding to the innovation	1. The most expensive of all of the option 2. Has a very complicated installation and fabrication method; estimated time of deliver and fabrication is 8-10 months

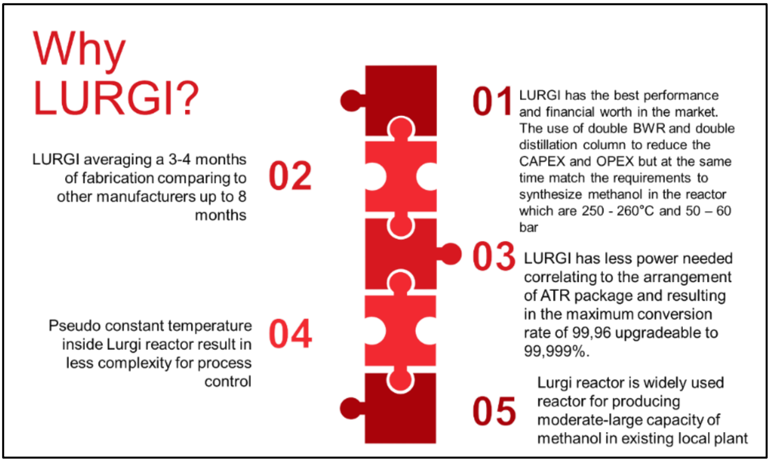
	the reactor height	in fabrication of the catalyst 4. Available option to scale up in the near future	
Linde	1. Axial flow of gas 2. Indirect cooling 3. Helical cooling inside	1. Enchancement in durability and lifetime of the reactor	1. Seldomly dynamic result of conversion rate; average conversion

	catalyst bed 4. No axial temperature variations in order to avoid stress 5. Quasi-Energy consumption	2. Quasi-energy consumption 3. Innovation in cooling integrated system, increase the conversion rate of methanol	rate is 85-99%
Lurgi	For Single Stage 1. Single indirectly cooled reactor 2. boiling water reactor (BWR) 3. Shell and tube with catalyst on the tube side For MegaMethanol 1. Two reactors: BWR and	1. Lower investment cost regarding the use of single cooled reactor 2. Currently has the best performance in terms of conversion rate and specific power used before reformer	1. Has a very strict and static set point

	a gas-cooled reactor 2. Cooling of the GCR is the unreacted syngas 3. GCR is a shell and tube with catalyst on shell side 4. Due to the second reactor, the BWR can be smaller 5. High conversion efficiency 6. Lower investment through smaller BWR 7. Suitable for large scale process		
Mistubishi Heavy Industries	1. Double-walled tube and	1. Better durability and enchancem	1. Improvement in protection of the

	shell reactor 2. Combined GCR and BWR 3. Feed-gas as coolant in the inner tube 4. Boiling water as coolant on the shellside 5. Reaction between inner and outer tube	ent in lifetime of the heat exchangers 2. Improved cooling integrated system with GCR and BWR	reactor is still highly questionable regarding to the given price and time by the manufacturer

Due to its performance, we finally choose isothermal of reactor that have higher yield than adiabatic reactor, also its low maintenance due to smaller need of catalyst load. Isothermal reactor also provide better controlling in temperature with cooling water (while adiabatic don’t have it)



Why LURGI?

01 LURGI has the best performance and financial worth in the market. The use of double BWR and double distillation column to reduce the CAPEX and OPEX but at the same time match the requirements to synthesize methanol in the reactor which are 250 - 260°C and 50 – 60 bar

02 LURGI averaging a 3-4 months of fabrication comparing to other manufacturers up to 8 months

03 LURGI has less power needed correlating to the arrangement of ATR package and resulting in the maximum conversion rate of 99,96 upgradeable to 99,999%.

04 Pseudo constant temperature inside Lurgi reactor result in less complexity for process control

05 Lurgi reactor is widely used reactor for producing moderate-large capacity of methanol in existing local plant

Figure III. 8 Simplified Consideration for Choosing Autothermal Reforming

III.3 PROCESS DESCRIPTION

As mentioned on the previous chapter, the mixture of syngas is mainly produced by Steam Reforming (SR) and Autothermal Reforming (ATR) of natural gas. The process of converting syngas into crude methanol occurs at pressure 50-100 bar and a temperature of $200 - 300℃$, where three reactions involved in methanol synthesis :

1. Hydrogenation of Carbon Monoxide
2. Hydrogenation of Carbon Dioxide
3. Water Gas Shift (WGS)

After the needed condition of reactor satisfied; $250℃$ of temperature and 50 bar of pressure, the rest of the process in the reactor is defined and concerned by the selectivity of the catalyst used(CuZnO) and the reactor licensor; Lurgi. Where 52% conversion (as minimum conversion) of Carbon Monoxide takes a part in the reaction. As of now, the outlet of the reactor Turbine Expander (TE-02) to reduce the pressure up to 20 bar and produce 4.85 MW of electricity to help power up the plant also the help minimize the use of thicker wall in the last Heat Exchanger (HE-03), Flash Drum (SP-01), and Distillation Columns (DC-01, and DC-02). After Turbine Expander (TE-02), the stream temperature reduced to 40℃ to satisfy the need of Light Gas Flash Separation Process and Purification Process. The stream then goes to the Flash Drum (SP-01) to separate the Light Gas from Raw Methanol. This was done to make crude methanol and other impurities become 2 phase as liquid and gas, where flash separation will simplify the methanol recovery by splitting light and heavy materials from the stream. The heavy stream is expected to have crude methanol and water as its dominant material. The crude methanol is purified in the distillation section to produce ASTM "AAA" grade methanol with specification : 99.85wt% MeOH, 0.1wt% water, and concentrations of higher alcohols at parts-per-million levels.

After the separation process the methanol composition in the stream is 39.15 percent and ready to be purify in the last 2 distillation columns with the condition of inlet is 40℃ and 19.98 bar. In order to gain 99.8% of Methanol, the calculation shows the need to use 5 and 15 trays sequentially. The final results of the methanol are 99.8% of purity in liquid phase with maximum capacity of 1400 tonne/day in the form of Liquid Methanol.

III.3.1 Syngas Optimization

The syngas produced from reforming process entered series of treatment process to satisfy methanol reactor specification. This process embrace Heat Exchanging (PHE-01) as preheater and Expanding (using turbine expander TE-01) sales gas to reach conditions at: 30 bar and 496.9℃. Large amount of heat is required in the steam reformer and also ATR produce heat, an advanced technology suggest that the heat from ATR can be used to supply power generator TE-01, heat needed by the steam reformer, and also supply superheated steam (water) as reactant for reforming process. This process is known as heat exchange reforming or gas-heated reforming. The major advantage of this is the reduction in investment cost by eliminating the expensive fired reformer.

There are 2 heat exchanging unit for syngas optimization, HE-01 and HE-02. The temperature of the effluent gas from autothermal reactor could reach $1095°C$ that need to be cooled down by two heat exchanging unit. The first exchanging unit (HE-01) will reduce syngas temperature to $692.4°C$ and produces steam-01 at $266.3°C$. The output syngas then mixed with recycle stream from Flash Separator (SP-01) that contain unreacted syngas from methanol reactor that still could be reacted into methanol. The mixed stream need to be compressed to 50 bar with compressor (C-02) to follow Lurgi Reactor performance. This mixed stream then compressed and cooled down again using HE-02 to reach $250°C$ and also produce steam-02 at $685.4°C$. The syngas leaving HE-02 then ready to be reacted in the reactor.

III.3.2 Methanol Synthesis

As of now, the outlet of the reactor enter Turbine Expander (TE-02) to reduce the pressure up to 20 bar and produce 4.85 MW of electricity to help power up the plant, also the help minimize the use of thicker wall in the last Heat Exchanger (HE-03), Flash Drum (SP-01), and Distillation Columns. After Turbine Expander (TE-02), the stream temperature reduced to $40°C$ to satisfy the need of Light Gas Flash Separation Process and Purification Process. The process of converting syngas into crude methanol occurs at a pressure of 50-100 bar and a temperature of $200 - 300°C$. The main reaction involved in methanol synthesis are shown below :

1. Hydrogenation of Carbon Monoxide

$$\mathrm{CO + 2H_2 \leftrightarrow CH_3OH}$$

That divided into two consecutive steps :

$$\mathrm{CO + H_2 \leftrightarrow CH_2O}$$

$$\mathrm{CH_2O + H_2 \leftrightarrow CH_3OH}$$

2. Hydrogenation of Carbon Dioxide

$$\mathrm{CO_2 + 3H_2 \leftrightarrow CH_3OH + H_2O}$$

Then, the result of Methanol reactor have 49 bar of pressure, that need to be slowed down to 20 bar using Turbine Expander (TE-02). This is to be done to minimize the use of thicker wall in the last Heat Exchanger (HE-03), Flash Drum, and Distillation Columns. The expanded effluent from turbine expander still have high temperature (183.8℃) that need to be cooled down for reaching optimum temperature, where 2 phase of liquid and gas occurs in the product. The result then entered last Heat Exchanging Unit, HE-03 to reach 40℃ to satisfy the need of Light Gas Flash Separation Process and Purification Process. At this condition, the stream contain liquid and gas phase that ready to be flashed on the next separator.

Flash Separation can be considered one of the simplest separation processes. In this process, a pressurised feed stream, which is in liquid phase, is passed through a throttling valve/nozzle or an expansion valve/ nozzle (sometimes, the feed stream may be passed through a heater before being passed through the valve/nozzle, in order to pre-heat the feed) connected to a tank or drum, which is called a “flash” tank or drum. After being passed through the valve/nozzle, the feed enters the tank/drum, whose temperature (40℃) and pressure (20 bar) is low; thus, there is a substantial pressure drop in the feed stream, causing the feed to partially vaporize.The fraction that becomes vapour goes up to and is taken off at the top of the tank/drum, while the remaining liquid part goes down to and is withdrawn at the bottom of the tank/drum. The vapour then partially recycled back to be reacted again methanol reactor because its content still have high composition of syngas which is the feed for reactor. The liquid part which have content of methanol 39.15%) will enter distillation unit to reach higher content of methanol (99.8%)

It is known that the product of plants for synthesizing methanol commonly defined as crude methanol is an aqueous solution of methanol containing by-product of synthesis reaction like water and some dissolved gases, mainly $H_2, CO_2, CH_4,$ and CO. The purification process would undergo in the importance of yielding a AA Grade of methanol (99.85% by weight) by the help of Distillation Columns, and at the same time to improve the mass flow rate which relates to the increase of tonne/day of pure Methanol by recycling the light gas from Flash Drum. Distillation separation of these constituents requires a large input of low grade heat, which is conveniently supplied by heat exchange with low pressure steam or with suitable warm gases in the synthesis plant. In general, distillation of methanol comprises the steps of :

1. Feeding a water-methanol mixture to a first distillation column, taking a product of methanol stream from an upper level, taking a product methanol stream from an upper level, taking aqueous methanol containing at least 95% weight percent of methanol as side stream and taking a predominantly water stream as bottoms
2. Feeding the aqueous methanol side stream to a second distillation column, taking a product methanol stream from an upper level and taking as bottoms a stream containing less than 60% weight of water.

Where on this case, we can clearly see the number of tray difference on DC-01, and DC-02, where DU-02 needs a high number of tray. This also will increase the load of towers operational because it is the last purification tower that produce 99.85% percent weight of methanol, which satisfies the needed specification for AA grade methanol. Besides, it will also reduce the lifespan of distillation tower due to its load dissemination.

Chapter IV
MASS AND ENERGY BALANCE

IV.1 MASS BALANCE

Mass balance calculation is one of important thing to be considered for designing process industry. Without it, we can’t measure our production capacity, raw material needed, utility, etc. We assume our calculation using steady-state condition where the accumulation term in general balance equation equals to zero and the equation (for calculation) simplifies to :

$$input - output + generation - consumption = 0$$

On this chapter, we are using kg (kilogram) as our unit standard for mass balance. Belows are mentioned basis that we used for our calculations :

Basis Calculation = 1 hour operation
Unit = kg
Production Capacity = 1400 tonne/day
= 58000 kg/h of grade AA Methanol
Raw Material Supply = 42 MMSCFD/day of Sales Gas
= 887.4 tonne/day

Operating Time = 1 year = 330 days
= 1 day = 24 hours

As Mentioned on chapter 3, methanol synthesis from sales gas were divided into 3 main process : syngas production, methanol synthesis, and methanol purification. Here, we divide the balances based on each process. Belows are the explanation :

IV.1.1 Syngas Production

Consist of equipment where sales gas are treated to produce syngas (H2 and CO), where the sales gas enter at 36974 kg/h with addition of steam (58412.9 kg/h) and pure oxygen (33432.76 kg/h) could produce mixture of syngas at 128819.75 kg/h

IV.1.1.1 Turbine Expander (TE-01)

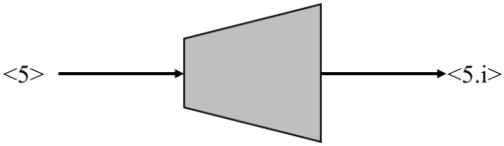


Figure IV. 1 Turbine Expander (TE-01)

Table IV. 1 Mass Balance on Turbine Expander (TE-01)

Component	**Input**		**Output**	
	Stream 5		Stream 5.i	
	%m	**Mass (kg)**	**%m**	**Mass (kg)**
Methane	0.8395	31039.25966	0.8395	31039.25966
Ethane	0.0713	2636.183096	0.0713	2636.183096
Propane	0.0628	2322.781226	0.0628	2322.781226
n-Butane	0.0264	975.8128455	0.0264	975.8128455
total	1.0000	36974.03683	1.0000	36974.03683

IV.1.1.2 Pre Heater (PHE-01)

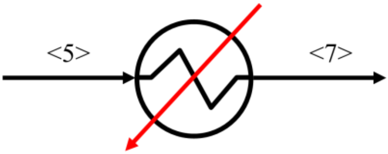


Figure IV. 2 Pre Heater (PHE-01)

Table IV. 2 Mass Balance on Pre-Heater 01

Component	**Input**		**Output**	
	Stream 5		Stream 7	
	%m	**Mass (kg)**	**%m**	**Mass (kg)**
Methane	0.8395	31039.25966	0.8395	31039.25966
Ethane	0.0713	2636.183096	0.0713	2636.183096
Propane	0.0628	2322.781226	0.0628	2322.781226
n-Butane	0.0264	975.8128455	0.0264	975.8128455
total	1.0000	36974.03683	1.0000	36974.03683

IV.1.1.3 Saturator (ST-01)

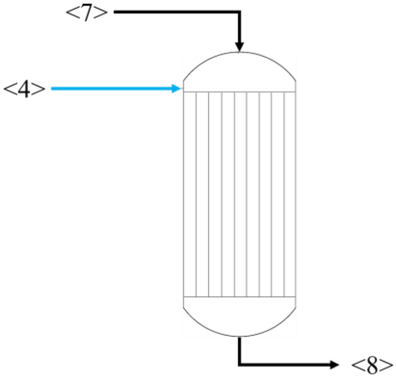


Figure IV. 3 Saturator (ST-01)

Purpose	: Mix sales gas (after being preheated) with superheated steam to maintain pressure at 30

bar and prepare reactants for reformer

Operating Condition : $P_{out} = 30\ bar\ ; T_{out} = 559.8℃$

Table IV. 3 Mass Balance on Saturator (ST-01)

Component	**Input**				**Output**	
	Stream 4		Stream 7		Stream 8	
	%m	**Mass (kg)**	**%m**	**Mass (kg)**	**%m**	**Mass(kg)**
Methane	0	0	0.8395	31039.26	0.325404	31039.26
Ethane	0	0	0.0713	2636.183	0.027637	2636.183
Propane	0	0	0.0628	2322.781	0.024351	2322.781
n-Butane	0	0	0.0264	975.8128	0.01023	975.8128
Water	1	58412.901	0.0000	0.0000	0.612378	58412.90
total	1	58412.9	1.0000	36974.04	1	95386.94
total	**95386.93818**				**95386.938**	

IV.1.1.4 Pre-Reformer (PR-01)

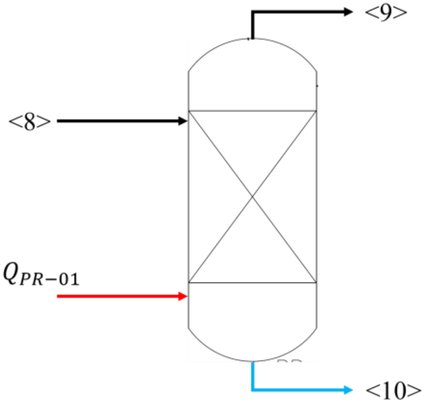


Figure IV. 4 Pre-Reformer (ST-01)

Purpose : Convert hydrocarbon from stream <6> into syngas (H2 and CO) and CO2

Operating Condition : $P_{out} = 30\ bar$; $T_{out} = 400°C$

Table IV. 4 Mass Balance on Pre-Reformer (PR-01)

Component	**Input**		**Output**	
	Stream 8		Stream 9	
	%m	**Mass (kg)**	**%m**	**Mass (kg)**
Methane	0.325404	31039.26	0.1627	15519.630
Ethane	0.027637	2636.183		
Propane	0.024351	2322.781		
n-Butane	0.01023	975.8128		
Water	0.612378	58412.9	0.3282	31305.819
CO			0.3615	34484.424
CO2			0.0631	6020.0738

H2	▨	▨	0.0845	8058.150
Total	**1**	**95386.94**	**1.0000**	**95388.10**

IV.1.1.5 Autothermal Reformer (ATR-01)

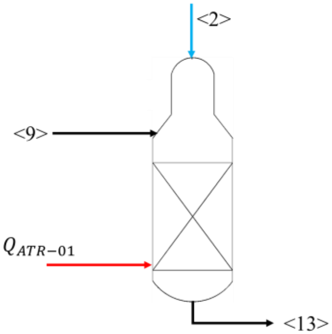


Figure IV. 5 Autothermal Reformer (ATR-01)

Purpose : Convert rest of the hydrocarbon from pre-reformer(stream <8>) into syngas using partial oxidation combined with steam methane reforming

Operating Condition : $P_{out} = 30\ bar\ ; T_{out} = 1095℃$

Table IV. 5 Mass Balance on Autothermal Reformer (ATR-01)

Component	**Input**		
	Stream 9		
	kmol	**%m**	**Mass (kg)**

Methane	967.3831	0.1627	15519.63
Ethane			
Propane			
n-Butane			
Water	1737.754	0.3282	31305.82
CO	1231.107	0.3615	34484.42
CO2	136.7897	0.0631	6020.07
H2	3997.098	0.0845	8058.15
Total	**8070.133**	**1.0000**	**95388.10**
	Stream 2		
	kmol	**%m**	**Mass (kg)**
Oxygen	1044.774	1	33432.76
TOTAL	**9114.906**		**128820.9**

Output		
Stream 13		
kmol	**%m**	**Mass (kg)**
11.4168	0.001422	183.1586
2590.0355	0.36221	46659.75
1905.7736	0.414396	53382.43

418.0897	0.142835	18400
5056.7495	0.079137	10194.41
9982.065	**1**	**128819.7519**

IV.1.2 Methanol Synthesis

Consist of equipments where syngas produced being reacted into methanol, with several treatment before to adjust reactors operating condition. At this final section, it will produce 133417.4885 kg/h of crude methanol

IV.1.2.1 Recycle Mixer (M-01)

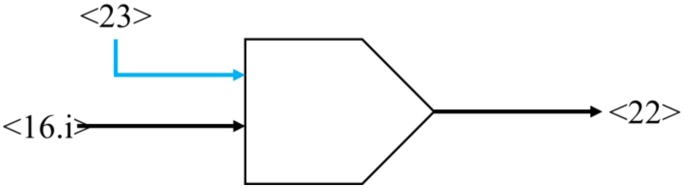


Figure IV. 6 Recycle Mixer

Purpose :	Mix syngas that have been lowed down its temperature with recycle stream from flash separator (SP-01) to raise syngas content

Operating Condition : $P_{out} = 19.91\ bar\ ; T_{out} = 733.3℃$

Table IV. 6 Mass Balance on Recycle Mixer

Component	**Input**			
	Stream 16.i		Stream 23	
	%m	**Mass (kg)**	**%m**	**Mass (kg)**
Methane	0.001422	183.1586	0.4267	1962.86
Ethane				
Propane				
n-Butane				
Water	0.36221	46659.750	0.0301	138.6666
CO	0.414396	53382.434	0.3795	1745.611
O2				0.00
H2	0.079137	10194.407	0.0000	0.122101
CH3OH			0.1636	752.3210
CO2	0.142835	18400.003	0.0000	0.0000
Total	1	128819.75	1.0000	4599.579
TOTAL	**133419.3310**			

Output	
Stream 22	
%m	**Mass(kg)**
0.016085	2146.02
	0.0000
	0.0000
	0.0000
0.350762	46798.42

0.413194	55128.04
1.52E-15	0.0000
0.07641	10194.53
0.005639	752.3210
0.137911	18400.00
1	133419.33
133419.3310	

IV.1.2.2 Methanol Reactor (R-01)

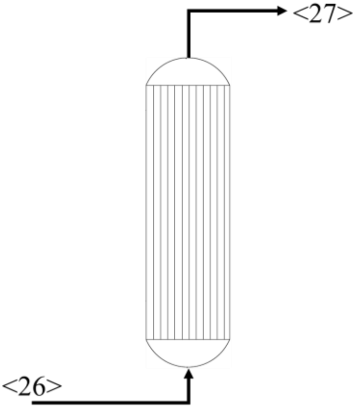


Figure IV. 7 Methanol Reactor (R-01)

Purpose :	Convert mixture of syngas into crude methanol
Operating Condition	: $P_{out} = 49\ bar\ ; T_{out} = 250°C$

Table IV. 7 Mass Balance on Methanol Reactor (R-01)

Component	**Input**
	Stream 26

	%m	Mass (kg)
Methane	0.0161	2146.02
Ethane		
Propane		
n-Butane		
Water	0.3508	46798.42
CO	0.4132	55128.04
O2		
H2	0.0764	10194.53
CH3OH	0.0056	752.32
CO2	0.1379	18400.00
Total	**1.0000**	**133419.3**

Output			
Stream 27			
kmol	**%mol**	**%m**	**Mass (kg)**
133.7674	0.021362	0.016085	2146.017
3015.822	0.481623	0.407221	54330.34
418.0897	0.066768	0.087778	11711.07
702.5352	0.112194	0.010616	1416.311

1991.572	0.318052	0.478301	63813.75
0.000	0	0	0
6261.787	1	**1**	**133417.5**

IV.1.3 Methanol Purification

Consist of 3 main equipment where syngas produced being reacted into methanol, with several treatment before to adjust reactors operating condition. At this final section, it will produce 102,000 kg/h of methanol

IV.1.3.1 Flash Separator (SP-01)

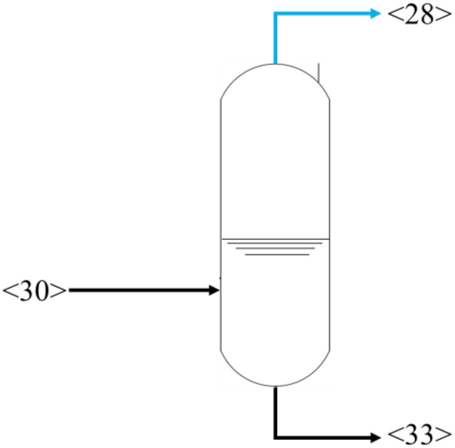


Figure IV. 8 Flash Separator (SP-01)

Purpose : Separate light gas and its liquid by flash condition. The liquid will contain methanol that will be purified with distillation unit

Operating Condition : $P_{out} = 19.96\ bar\ ; T_{out} = 100°C$

Table IV. 8 Mass Balance on Flash Separator (SP-01)

Component	**Input**			
	Stream 30			
	kmol	**%mol**	**%mass**	**Massa (kg)**
Methane	133.77	0.02	0.02	2146.0
Water	3015.8	0.4816	0.4072	54330
CO	418.09	0.0668	0.0878	11711
H2	702.54	0.1122	0.0106	1416.3
CH3OH	1991.6	0.3181	0.4783	63814
Total	**6261.8**	**1.00**	**1.00**	**133417**
Methane				
Water				
CO				
H2				
CH3OH				
Total				

Output			
Stream 28 (Top)			
kmol	**%mol**	**%mass**	**Massa (kg)**

122.48	0.097	0.1211	1964.9
2.8112	0.0022	0.0031	50.644
416.76	0.33	0.7197	11674
702.54	0.5563	0.0873	1416.3
34.756	0.0275	0.0687	1113.6
1262.9	**1.0**	**1**	**16219**
Stream 33 (Bottom)			
11.29	0.0023	0.0015	181.12
3013	0.6027	0.4631	54280
1.3339	0.0003	0.0003	37.363
4E-13	9E-17	7E-18	9E-13
1956.8	0.3915	0.535	62700
4998.9	**1.00**	**1**	**117198**

IV.1.3.2 Distillation Tower (MD-01)

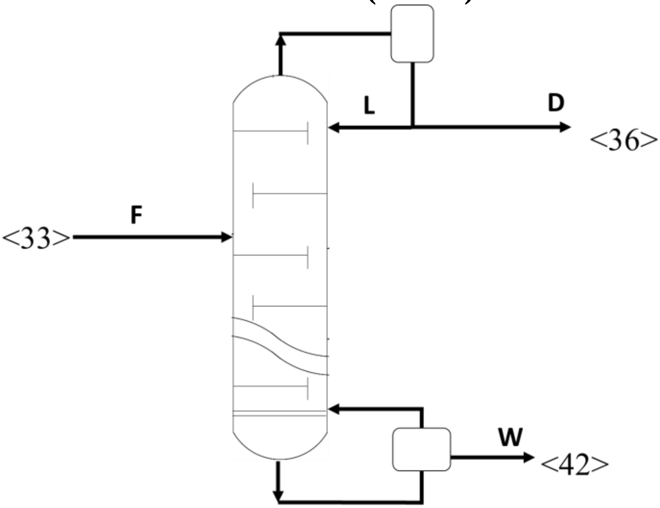


Figure IV. 9 Distillation Tower (MD-01)

Purpose : Purify methanol from its impurities until reached 95% mol of methanol

Operating Condition : $P_{cond} = 10.5\ bar$; $P_{reb} = 11\ bar$

Table IV. 9 Mass Balance on Distillation Tower (MD-01)

Component	**Input**			
	Stream 33			
	kmol	**%mol**	**%mass**	**Massa (kg)**
Methane	11.29	0.0023	0.0015	181.12323
Water	3013	0.6047	0.4631	54279.7
CO	1.3339	0.0003	0.0003	37.362563
H2	4E-13	9E-17	7E-18	8.759E-13

CH3OH	1956.8	0.3927	0.535	62700.101
Total	**4982.5**	**1**	**1**	**117198.29**
Methane				
Water				
CO				
H2				
CH3OH				
Total				
TOTAL	**4982.5**			**117198.29**

Output			
Stream 36 (Distillate)			
kmol	**%mol**	**%mass**	**Massa (kg)**
11.29	0.0054	0.0028	181.12323
150.65	0.0717	0.0418	2713.985
1.3339	0.0006	0.0006	37.362563
4E-13	2E-16	1E-17	8.759E-13
1937.2	0.9223	0.9549	62073.1
2100.5	**1**	**1**	**65005.57**

Stream 42 (Bottom)

1E-08	5E-12	5E-12	2.401E-07
2862.4	0.9932	0.988	51565.715
1E-17	5E-21	8E-21	4.105E-16
6E-95	2E-98	2E-99	1.196E-94
19.568	0.0068	0.012	627.00101
2881.9	**1**	**1**	**52192.716**
4982.5			**117198.29**

IV.1.3.3 Distillation Tower (MD-02)

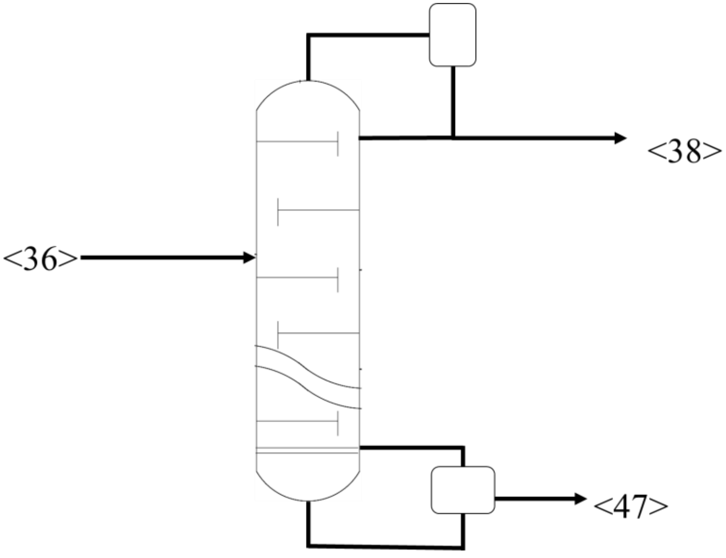


Figure IV. 10 Distillation Tower (MD-02)

Purpose : Purify methanol from its impurities until reached 95% mol of methanol

Operating Condition : $P_{cond} = 10\ bar$; $P_{reb} = 10.5\ bar$

Table IV. 10 Mass Balance on Distillation Tower (MD-02)

Component	**Input**			
	Stream 36			
	kmol	**%mol**	**%mass**	**Massa (kg)**
Methane	11.29	0.0054	0.0028	181.12323
Water	150.65	0.0717	0.0418	2713.985
CO	1.3339	0.0006	0.0006	37.362563
H2	4E-13	2E-16	1E-17	8.759E-13
CH3OH	1937.2	0.9223	0.9549	62073.1
Total	**2100.5**	**1**	**1**	**65005.57**

Methane				
Water				
CO				
H2				
CH3OH				
Total				
TOTAL	**2100.5**			**65005.57**

Output
Stream 38 (Distillate)

kmol	%mol	%mass	Massa (kg)
11.29	0.0058	0.00291	181.123231
7.5325	0.0039	0.00218	135.69925
1.3339	0.0007	0.0006	37.362563
4E-13	2E-16	1.4E-17	8.7595E-13
1933.4	0.9897	0.99432	61948.9535
1953.5	**1**	**1**	**62303.1385**

Stream 47 (Bottom)			
1E-10	8E-13	6.6E-13	1.776E-09
143.12	0.9736	0.95406	2578.28574
2E-20	1E-22	1.7E-22	4.5118E-19
2E-105	2E-107	2E-108	4.929E-105
3.8745	0.0264	0.04594	124.146199
146.99	**1**	**1**	**2702.43194**
2100.5			**65005.5705**

IV.2 ENERGY BALANCE

Energy balance calculation is one of important thing to be considered for designing process industry. Without it, we can’t measure our utilities needed. We assume our calculation using steady-state condition where the accumulation term in general balance equation equals to zero and the equation (for calculation) simplifies to :

$$Energy\ input - Energy\ output + generation - consumption = 0$$

On this chapter, we are using kJ (kilo Joule) as our unit standard for energy balance. Belows are mentioned basis that we used for our calculations :

Basis Calculation	=	1 hour operation
Unit	=	kJ
Operating Time	=	1 year = 330 days
	=	1 day = 24 hours
Reference		
T	=	25℃
P	=	1 atm

IV.2.1 Syngas Production

IV.2.1.1 Turbine Expander (TE-01)

Details of each stream :

Inlet Temperature	T5	=	30	°C
Inlet Pressure	P5	=	45.83	bar
Outlet Temperature	T5.i	=	22.46	°C
Outlet Pressure	P5.i	=	30	bar

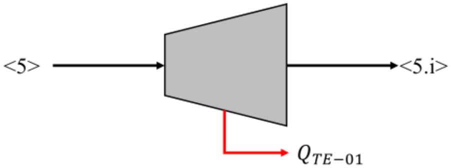


Figure IV. 11 Energy Stream for Turbine Expander (TE-01)

Table IV. 11 Energy Balance on Turbine Expander (TE-01)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 5	397692.1427	ΔH 5.i	-200742.8194
2			Q TE-01	598434.9621
	Total	397692.1427	Total	397692.1427

IV.2.1.2 Pre-Heater 01 (PHE-01)

Purpose : Raise temperature of sales gas into 496.9 ℃
Details of each stream :

Inlet Temperature	T5	=	22.46	℃
Inlet Pressure	P5	=	30	bar
Outlet Temperature	T7	=	496.9	℃
Inlet Pressure	P7	=	30	bar

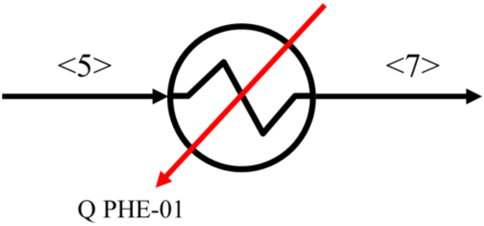


Figure IV. 12 Energy Stream for Pre-Heater 01

Table IV. 12 Energy Balance on Pre-Heater 01 (PHE-01)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 5	-200742.8194	ΔH 7	51777118.45
2	Q PHE-01	51977861.27		
	Total	51777118.45	Total	51777118.45

IV.2.1.3 Saturator (ST-01)

Inlet Temperature : T4 = 669.9 °C

: T7 = 496.9 °C

Inlet Pressure : P4 = 49.99 bar

: P7 = 30 bar

Outlet Temperature : T8 = 574.2 °C

Outlet Temperature : P8 = 30 bar

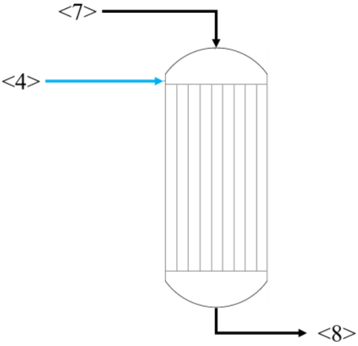


Figure IV. 13 Energy Stream for Saturator (ST-01)

Table IV. 13 Energy Balance on Saturator (ST-01)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 4	76849137.77	ΔH 8	127318934.84
2	ΔH 7	51777118.45		
	Total	128626256.22	Total	127318934.84

IV.2.1.4 Pre-Reformer (PR-01)

Inlet Temperature	T8	=	574.2	°C
Inlet Pressure	P8	=	30	bar
Outlet Temperature	T9	=	400	°C
Inlet Pressure	P9	=	30	bar

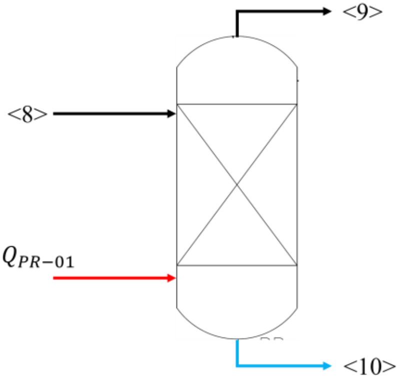


Figure IV. 14 Pre-Reformer (PR-01)

Table IV. 14 Energy Balance on Pre-Reformer (PR-01)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 8	127318934.8	ΔH 9	47564684.82
2	Q	181628044.6		
3	ΔHR	-261382294.6		
	Total	47564684.82	Total	47564684.82

IV.2.1.5 Autothermal Reformer (ATR-01)

Purpose : Converge hydrocarbon from sales gas where syngas production happened mainly on this reactor. The effluent are syngas (CO and H2) and unreacted methane

Inlet Temperature T2 = 30.0 °C
T8 = 400 °C

Inlet Pressure P2 = 30 bar

P8 = 30 bar

Outlet Temperature T13 = 1095 ℃

Outlet Pressure P13 = 30 bar

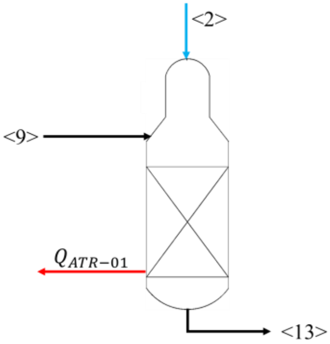


Figure IV. 15 Energy Stream for Autothermal Reformer (ATR-01)

Table IV. 15 Energy Balance on Autothermal Reformer (ATR-01)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 2	153654.19	ΔH 13	359183658.08
2	ΔH 9	99260197.59	Q ATR-01	60038935.70
3	ΔHR	319808742		
	Total	4.19E+08	Total	4.19E+08

IV.2.1.6 Heat Exchanger (HE-01)

Syngas Inlet Temperature = T16 = 1095 °C ;
Syngas Inlet Pressure = P16 = 30 bar
Syngas Outlet Temperature = T16.i = 749.1 °C ;
Syngas Outlet Pressure = P16.i = 27.97 bar
Cooling Water Inlet Temperature = T17 = 30.26 °C ;
Cooling Water Inlet Pressure = P17 = 50 bar =
Cooling Water Outlet Temperature = T20 = 317.7 °C ;
Cooling Water Outlet Pressure = P20 = 49.88 bar

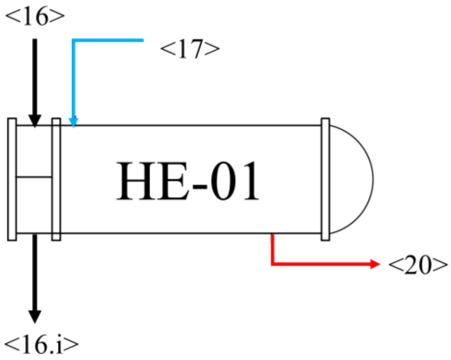


Figure IV. 16 Energy Stream for Heat Exchanger (HE-01)

Table IV. 16 Energy Balance for Heat Exchanger-01

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 16	359183658.08	ΔH 16.i	235126210.65

2	*cooling water*			
3	Δ*H 17*	2983840.69	Δ*H 20*	127041288.12
	Total	3.62E+08	Total	3.62E+08

IV.2.2 Methanol Synthesis

IV.2.2.1 Recycle Mixer (M-01)

Cooled Syngas Inlet Temperature = T16.i = 749.1 °C

Cooled Syngas Inlet Pressure = P16.i = 27.97 bar

Recycled gas Temperature = T23 = 40 °C

Recycled gas Pressure = P23 = 19.99 bar

Mixed gas Temperature = T22 = 733.8 °C

Mixed gas Pressure = P22 = 19.99 bar

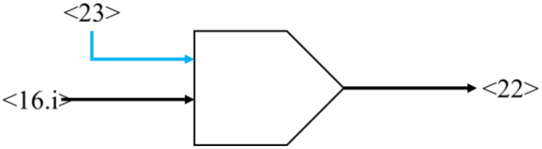


Figure IV. 17 Energy Stream for Recycle Mixer (M-01)

Table IV. 17 Energy Balance for Recycle Mixer (M-01)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 16.i	235126210.65	ΔH 22	237290471.57
2	ΔH 23	113405.29		

3				
	Total	235239615.93	Total	237290471.57

IV.2.2.2 Compressor (C-02)

Mixed Gas Inlet Temperature = T22 = 633.3 ℃

Mixed Gas Inlet Pressure = P22 = 19.99 bar

compressed gas Temperature = T24 = 917.7 ℃

Compressed gas Pressure = P24 = 50 bar

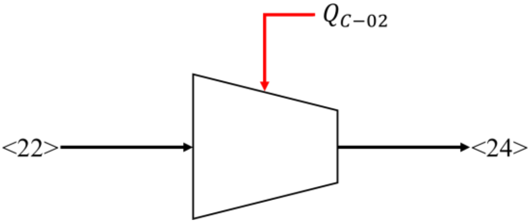


Figure IV. 18 Compressor (C-02)

Table IV. 18 Energy Balance for Compressor-02 (C-02)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 22	201492596.05	ΔH 24	304676070.84
2	Q C-01	103183474.79		
3				
	Total	304676070.84	Total	304676070.84

IV.2.2.3 Heat Exchanger (HE-02)

Syngas Inlet Temperature = T24 = 917.7 ℃ ;
Syngas Inlet Pressure = P24 = 50 bar
Syngas Outlet Temperature = T26 = 250 ℃ ;
Syngas Outlet Pressure = P26 = 49.98 bar
Cooling Water Inlet Temperature = T18 = 30.26 ℃ ;
Cooling Water Inlet Pressure = P18 = 50 bar = 49.36 atm
Cooling Water Outlet Temperature = T25 = 685.4 ℃ ;
Cooling Water Outlet Pressure = P25 = 50 bar = 49.36 atm

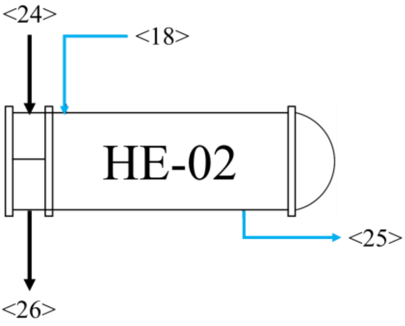


Figure IV. 19 Heat Exchanger (HE-02)

Table IV. 19 Energy Balance for Heat Exchanger-02

	Energy Input (kJ)	Energy Output (kJ)

1	ΔH 24	304676070.84	ΔH 26	71642781.78
2	*cooling water*			
3	*ΔH 18*	4420856.85	*ΔH 25*	237454145.91
	Total	309096927.69	Total	309096927.69

IV.2.2.4 Methanol Reactor (R-01)

Feed R-01 Temperature = T26 = 250 ℃

Feed R-01 Pressure = P26 = 49.98 bar

Effluent R-01 Temperature = T27 = 250 ℃

Effluent R-01 Pressure = P27 = 49 bar

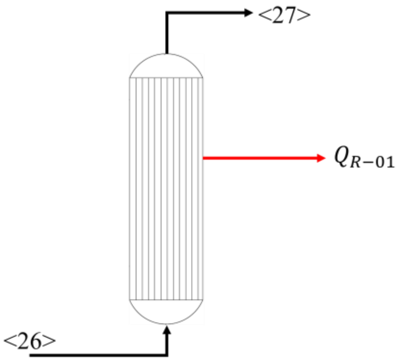


Figure IV. 20 Enery Stream of Methanol Reactor (R-01)

Table IV. 20 Energy Balance for Methanol Reactor-01

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 26	71642781.78	ΔH 27	56212806.70
2	ΔHR	161262241.8	Q	176692216.86

3				
	Total	232905023.56	Total	232905023.56

IV.2.2.5 Turbine Expander-02 (TE-02)

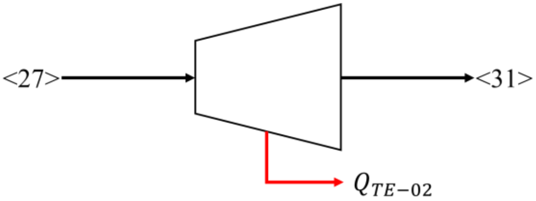


Figure IV. 21 Energy Stream for Turbine Expander (TE-02)

Table IV. 21 Energy Balance for Expander (T-01)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 27	56212806.70	ΔH 31	38676123.72
2			Q	17536682.98
3				
	Total	56212806.70	Total	56212806.70

IV.2.2.6 HEAT EXCHANGER (HE-03)

Crude methanol Inlet Temperature = T31 = 183.8 °C ;
Crude methanol Inlet Pressure = P31 = 20 bar
Crude methanol Outlet Temperature = T30 = 40 °C ;
Crude methanol Outlet Pressure = P30 = 19.98 bar
Cooling Water Inlet Temperature = T19 = 30.26 °C ;
Cooling Water Inlet Pressure = P19 = 50 bar =
Cooling Water Outlet Temperature = T32 = 181.5 °C ;

Cooling Water Outlet Pressure = P32 = 50 bar

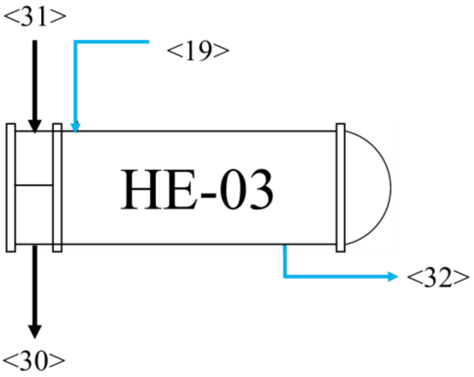


Figure IV. 22 Energy Stream of Heat Exchanger (HE-03)

Table IV. 22 Energy Balance for Heat Exchanger (HE-03)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 31	38676123.72	ΔH 30	3445390.76
2	*cooling water*			
3	Δ*H 19*	933225.42	Δ*H 32*	36163958.38
	Total	39609349.14	Total	39609349.14

IV.2.3 Methanol Purification

IV.2.3.1 FLASH SEPARATOR

Crude methanol Inlet Temperature = T30 = 40 ℃ ;

Crude methanol Inlet Pressure = P30 = 19.98 bar

Top Products Temperature = T28 = 40 ℃ ;

Top product Pressure = P28 = 19.98 bar

Bottom product Temperature = T33 = 40 °C ;

Bottom product Pressure = P33 = 19.98 bar =

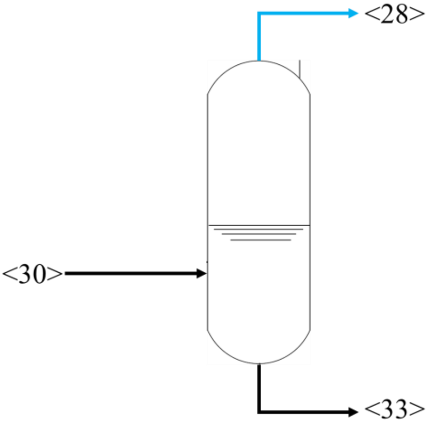


Figure IV. 23 Energy Stream for Flash Separator (SP-01)

Table IV. 23 Energy Balance for Flash Separator (SP-01)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 30	3445390.76	ΔH 28	577317.05
2			ΔH 33	2868073.709
3				
	Total	3445390.76	Total	3445390.76

IV.2.3.2 DISTILLATION TOWER (MD-01)

Feed Inlet Temperature = T33 = 40.31 °C ;

Feed Inlet Pressure = P33 = 11 bar

Distillate Temperature = T36 = -123.70 °C ;

Distillate Pressure = P36 = 10.5 bar = 1

Bottom product Temperature = T42 = 182.30 °C ;

Bottom product Pressure = P42 = 11 bar =

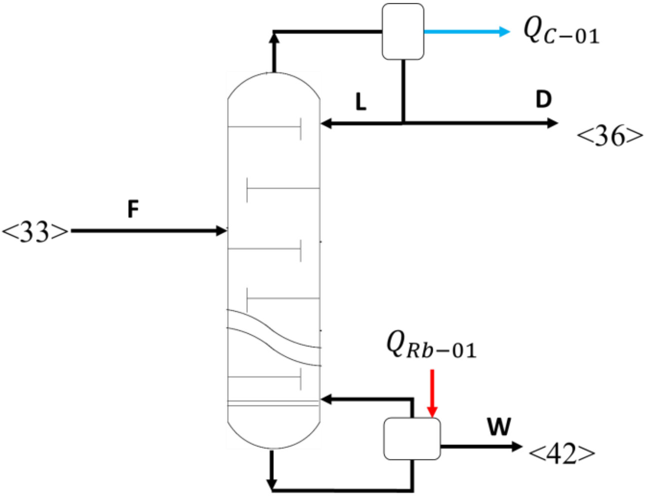


Figure IV. 24 Distillation Tower (MD-01)

Table IV. 24 Energy Balance for Distillation Tower (MD-01)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 33	2927780.86	ΔH 36	-11966092.62
2			ΔH 42	15585268.35
3	QRB-01	21297466.10	QC-01	20606071.22
4				

	Total	24225246.96	Total	24225246.96

IV.2.3.3 DISTILLATION TOWER (MD-02)

Feed Inlet Temperature = T36 = -124 °C ;

Feed Inlet Pressure = P36 = 10.5 bar

Distillate Temperature = T38 = 89.85 °C ;

Distillate Pressure = P38 = 10 bar = 9.8

Bottom product Temperature = T47 = 159.10 °C ;

Bottom product Pressure = P47 = 10.5 bar = 10

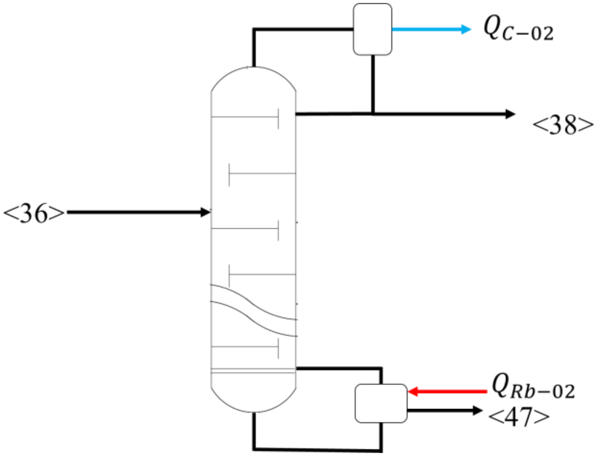


Figure IV. 25 Distillation Tower (MD-02)

Table IV. 25 Energy Balance for Distillation Tower (MD-02)

	Energy Input (kJ)		Energy Output (kJ)	
1	ΔH 36	-11966092.62	ΔH 38	6017894.46
2			ΔH 47	681869.3284
3	QRB-01	22189904.51	QC-01	3524048.10
4				
	Total	10223811.89	Total	10223811.89

Chapter V
PROCESS EQUIPMENTS AND PRICELIST

The specifications of the equipment used in the Natural Gas Methanol Plant are as follows:

V.1 Oxygen Compressor (C-01)

Table V. 1 Specification of Compressor (C-01)

Specification		Information				
Code	:	C-01				
Type	:	Reciprocating Compressor				
Purpose	:	Menaikkan tekanan gas oksigen sebagai feed untuk ATR				
Number of stage	:	1 stage				
Material	:	Carbon Steel				
Operating Condition	:	Psuction =	15 bar	Tsuction	=	25 ℃
		Pdischarge =	30.0 bar	Tdischarge	=	30.0 ℃
Capacity	:	33433	kg/jam			
Efficiency	:	79%				
Power	:	1834	hp			
Unit Price	:	$ 131199				

V.2 Sales Gas Turbine Expander (TE-01)

Table V. 2 Specification of Turbine Expander (TE-01)

Specifications	**Information**
Code	TE-01
Type	Expander
Purpose	Depressurize stream from sales gas from 45.83 bar to 30 bar

Material	Carbon Steel							
Operating Condition	Psuction	=	664.9	bar	Tsuction	=	30.0	℃
	Pdischarg	=	22	bar	Tdischarg	=	22.5	℃
Capacity (kg/h)	36974.03683							
Efficiency	72%							
Power (kW)	4.86							
Unit Price	$109224							

V.3 Saturator (ST-01)

Table V. 3 Specification of Saturator (ST-01)

Specification of Units :

Purpose : Mix syngas with superheated steam and maintain pressure at 30 bar

Type : Tangki silinder, tutup atas berbentuk *standard dished head* dan tutup bawah berbentuk flat bottom

Shell

Materials = Carbon Steel SA-240 grade M

Diameter = 50.00 ft = 15.24 m

Height = 30.47 ft = 9.29 m

Tinggi tutup atas = 298.02 inch = 7.57 m

Tinggi tutup bawah = 30.50 inch = 0.77 m

Tebal tutup atas = 0.63 inch = 0.02 m

Tebal tutup bawah = 13.00 inch = 0.33 m

Jumlah = 1 buah

Unit Price = $890943

V.4 Pre Reformer (PR-01)

Table V. 4 Specification of Pre Reformer (PR-01)

Specification		**Information**
Code	:	PR-01
Function	:	Converge hydrocarbon from sales gas where syngas production happened mainly on this reactor
Capacity	:	95387 kg/jam
Material	:	SA 283 grade C (Carbon steel)
Type	:	Reaktor fixed bed dengan pemanas furnace
Specification furnace		
Height	:	29.7 ft
Length	:	39.82 ft
Width	:	38.09 ft
Number of tube	:	130 buah
Length of tube	:	36.09 ft
Outside diameter tube	:	4.5 in
Distance of center to center tube	:	8.1 in
Specification Catalyst		
Main Component	:	Cu-ZnO-Ni
Carrier	:	Keramik
Shape	:	Rasching Ring
True density	:	3460 kg/m^3
Catalyst Diameter	:	16 mm = 0.016 m
Description	:	*Gray 6-holed domed cylinder*
UnitPrice	:	$166033

V.5 Autothermal Reformer (ATR-01)

Table V. 5 Specification of Autothermal Reformer (ATR-01)

Specification	Information
Code :	ATR-01
Purpose :	Oksidasi dari gas alam untuk menjadi CO,CO2 dan H2
Capacity :	128820.9 kg/jam
Material :	SA 283 Grade C (Carbon steel)
Type :	Fixed bed reaktor dengan tutup atas : *conical deshed head* dan tutup bawah *standard dished head*
Unit :	1 buah
Pressure drop :	0.207 atm
Dimension	
Bejana	
Volume tangki : 5	ft^3
ID : 129.2 in	ts = 2 1/4 in
OD : 133.7 in	Hs = 264 in
Tutup Atas	
Jenis :	*conical dished head*
Tebal : 2 1/4 in	
ha : 38.65 in	
Tutup Bawah	
Jenis : *standard dished head*	Jacket
Tebal : 2 1/4 in	ID = 132 in
hb : 23.32 in	OD = 133.7 in

sf : 1.5 in

Tinggi Total Tangki : 327.5 in

Unit Price $341009

V.6 Pre Heater (PHE-01)

Table V. 6 Specification of Pre Heater (PHE-01)

Specification Information

Code : E-211

Type : Shell and Tubes, 4-8 exchanger

Purpose : Memanaskan feed gas alam sebelum masuk steam reformer

Units : 1 buah Heat Exchanger 4-8 disusun paralel

Capacity : 9244 kg/jam

Size

Shell : ID = 39 in ; Baffle Spacing = 39 in

Tube : OD = 0.75 in ;ID = 0.482 in

Pt = 1 in (square)

Panjang = 12 ft,

Jumlah tube, Nt = 982

BWG = 10

Materials construction

Shell : Carbon steel

Tube : Carbon steel

DP allowance : 2 psia

Dirt factor (Rd) : 0.01141jft2oF/Btu

Unit Price $95666

V.7 Water Pump (P-01)

Table V. 7 Specification of Water Pump (P-01)

Specification		Information				
Code	:	P-01				
Type	:	Centrifugal Pump				
	:	casing dan wear ring	:	cast iron		
	:	impeller dan wear ring	:	cast iron		
	:	shaft	:	carbon steel		
Unit	:	1	unit			
Purpose	:	To Pump Cooling Water Supply				
Name of fluid	:	Water				
Design Spec.						
T desain	:	161.6	^oC			
P desain						
Suction pressure	:	10	bar	=	145.04	psi
Discharge pressure	:	50	bar	=	725.19	psi
Capacity	:	590	gpm			
Viscosity of fluid	:	0.7972	cp			
Differential head	:	10.9	ft			
Power motor	:	1.5	hp			
Unit Price	:	$68770				

V.8 Heat Exchanger 01 (HE-01)

Table V. 8 Specification of Heat Exchanger-01 (HE-01)

Specification		**Information**				
Code	:	HE-01				
Type	:	*Shell and Tubes, 1-2 exchangers*				
Function	:	Conditioning ATR outlets				
total	:	1	heat exchangers 1-2 are arranged in parallel			
Capacity	:	53840	kg / hour			
Size						
The shell	:	ID =	39	in; Baffle Spacing =	39	in
Tube	:	OD =	0.75	in; ID =	.652	in
		Pt =	1	in	*(square)*	
		Length =	12	ft,	Number of tubes, Nt =	968
		BWG		=	18	
Construction material						
The shell	:	*Carbon steel*				
Tube	:	*Carbon steel*				
DP allowance	:	2	psia			
Dirt factor (Rd)	:	0.01635	jft2oF / Btu			
Unit price	:	$ 196253				

V.9 Syngas Compressor (C-02)

Table V. 9 Specification of Syngas Compressor (C-02)

Specification		Information
Kode	:	G-314
Type	:	Reciprocating Compressor
Fungsi	:	Menaikkan tekanan gas alam agar sesuai dengan kondisi operasi reaktor methanol
Jumlah stage	:	1 stage
Bahan	:	Carbon Steel
Kondisi operasi	:	Psuction = 19.98 bar Tsuction = 633
		Pdischarge = 50.0 bar Tdischarge = 917.7
Kapasitas	:	133419 kg/jam
Efisiensi	:	79%
Power	:	54438 hp
Unit Price	:	

V.10 Heat Exchanger-02 (HE-02)

Table V. 10 Specification of Heat Exchanger 02 (HE-02)

Specification		Information
Code	:	HE-02
Type	:	*Shell and Tubes, 4-8 exchangers*
Function	:	Heat up the syngas feed before entering Lurgi Reactor
total	:	1 Heat Exchanger 4-8 arranged in parallel
Capacity	:	10095 kg / hour
Size		
The shell	:	ID = 39 in; Baffle Spacing = 39 in
Tube	:	OD = 0.75 in; ID = 0.482 in Pt = 1 in *(square)*

		Length =	12	ft,	Number of tubes, Nt =	982
		BWG	=	10		
Construction material						
The shell	:	*Carbon steel*				
Tube	:	*Carbon steel*				
DAllowance	:	2	psia			
Dirt factor (Rd)	:	0.01141	jft2oF / Btu			
Unit price	:	$ 196253				

V.11 Lurgi Methanol Reactor (R-01)

Table V. 11 Specification of Methanol Reactor (R-01)

Specification		Information					
Code	:	R-01					
Function	:	Watch CO and CO_2 with H_2 to produce CH_3OH products					
Capacity	:	133419	kg / hour				
material	:	SA 212 Grade B (Carbon Steel)					
Type	:	Fixed Bed Multitubular Reactor					
Size							
Total reactor height, HT	:	289	in				
The total volume of the reactor, Vtot	:	2912.3	ft3	=	5186,9	gallon	
Side shell							
The ID	:	155.2	in	Thick	=	0.31	in
OD	:	157.6	in	High	=	233	in
Top Close							

thick, tha : .1875 in
tall, ha : 26.6 in

Lid down
thick, thb : .1875 in
height, hb : 26.6 in
Tube side
OD : 1.25 in
The ID : 1.15 in
BWG : 18 in
Pt : 1.5625 in triangular
Nt : 18644
L. : 16 ft

Unit price : $ 696013

V.12 Turbine Expander (TE-02)

Table V. 12 Specification of Turbine Expander 02 (TE-02)

Specification		Information
Code	:	TE-02
Type	:	Expanders
Function	:	Reduces syngas pressure from the reactor
material	:	*Carbon Steel*
Operating conditions	: :	Psuction: 4900 kPa Tsuction = 250 ˚C Pdischarge: 2000 kPa

		Tdischarge = 184 ˚C
Capacity (kg / hour)	:	13341,750
Efficiency	:	0.720
Power	:	30.15 kW
Unit price	:	$ 109224

V.13 Heat Exchanger 03 (HE-03)

Table V. 13 Specification of Heat Exchanger 03 (HE-03)

Specification		**Information**
Code	:	HE-03
Type	:	*Shell and Tubes, 4-8 exchangers*
Function	:	Conditioning the Outlet After the Reactor
total	:	12 pieces of Heat Exchanger 4-8
Capacity	:	78315 kg / hour
Size		
	:	ID = 39 in; Baffle Spacing = 39 in
	:	OD = 0.75 in; ID = 0.652 in
Shell Tube	:	Pt = 1 in *(triangular)*
	:	Length = 12 ft, Number of tubes, Nt = 982
	:	BWG = 18
Construction material		
The shell	:	*Carbon steel*
Tube	:	*Carbon steel*

DAllowance	:	2	psia
Dirt factor (Rd)	:	0.02230	jft2oF / Btu
Unit price	:	$ 196253	

V.14 Flash Separator (FS-01)

Table V. 14 Specification of Flash Separator (FS-01)

Specification		Information				
Code	:	FS-01				
Function	:	Separate Heavy and Light Components from Methanol Reactor Effluent				
Construction Materials	:	*Carbon Steel*				
Pressure	:	20	bar			
Temperature	:	40	° C			
Diameter	:	104.06	in			
H / D ratio	:	1.39				
Total Height	:	3,683	m	=	145	in
Residence Time	:	5	minute			
Thick Top Lid	:	.188	in			
Bottom Lid Shape	:	*Eliptical dished head*				
Thick Bottom Lid	:	0.188	in			
Unit price	:	$683440				

V.15 Distillation Column 01 (DC-01)

Table V. 15 Specification of Distillation Column 01 (DC-01)

Specification		Information	
material	:	SA 353	
Tray Diameter	:	3.00	ft

	:	36	in	
Layout	:	*Cross flow*		
Spacing Tray	:	0.4	ft	ft
	:	5	in	in
Column Height	:	57.8		ft
	:	693,6685		in
L / D	:	0.8000		
n	:	3.5		
Thick Shell	:	0.2		in
Unit price	:	$480409		

V.16 Distillation Column 02 (DC-02)

Table V. 16 Specification of Distillation Column 01 (DC-01)

Specification		**Information**	
material	:	SA 353	
Tray Diameter	:	3	ft
	:	36	in
Layout	:	*Cross flow*	
Spacing Tray	:	0.5	ft
	:	6	in
Column Height	:	71.9	ft
	:	862,838	in
L / D	:	0.8000	
n	:	3.5	
Thick Shell	:	0.2	in
Unit price	:	$389553	

V.17 Methanol Reboiler on DC 01 (RB-01)

Table V. 17 Specification of Methanol Reboiler 01 (RB-01)

Tool Specifications:					
Name	Distillation Reboiler				
Code	RB-01				
Function	Heats the bottom product flow of the distillation column				
Type	*Reboiler Kettle*				
material	*Carbon steel*				
Temperature entered	Steam	=	182.23		°C
	To reboiler	=	124.1		°C
Temperature is out	Steam	=	182.23		°C
	Product	=	127.4		°C
The shell	The ID	=	39		in
	Baffle	=	39		in
	Passes	=	1		
Tube	OD	=	¾		in
	The ID	=	0.62		in
	BWG	=	16		
	Pitch	=	1	Triangular in2	
	Long	=	16	ft	
	total	=	1176		
	Passes	=	2		
	ΔP	=	0.522		psi
Rd	-0.002	(hr) (ft2) (oF) / (Btu)			
Area	9070.0	ft2			
total	1	fruit			

Unit price	:	$ 224242

V.18 Methanol Reboiler of DC 02 (RB-02)

Table V. 18 Specification of Methanol Reboiler 02 (RB-02)

Tool Specifications:				
Tool's name	Distillation Reboiler			
Tool Code	RB-02			
Function	Heats the bottom product flow of the distillation column			
Type	*Reboiler Kettle*			
material	*Carbon steel*			
Temperature entered	Steam	=	159.1	°C
	To reboiler	=	124.1	°C
Temperature is out	Steam	=	159.1	°C
	Product	=	127.4	°C
The shell	The ID	=	39	in
	Baffle	=	39	in
	Passes	=	1	
Tube	OD	=	3/4	in
	The ID	=	0.62	in
	BWG	=	16	
	Pitch	=	1	Triangular in2
	Long	=	16	ft
	total	=	1176	
	Passes	=	2	
	ΔP	=	0.011	psi

Rd	0.023	(hr) (ft2) (oF) / (Btu)
Area	768.4	ft2
total	1	fruit
Unit price	:	$ 224242

V.19 Methanol Condenser of DC 01 (COND-01)

Table V. 19 Specification of Methanol Condenser 01 (COND-01)

Tool Specifications:				
Tool's name	Methanol Condenser			
Tool Code	COND-01			
Function	Condensing Top Products from Distillation Column DC-01			
Type	*Vertical Condenser*			
material	*Carbon steel*			
Temperature entered	Cooling water	=	30	°C
	To condenser	=	87.6	°C
Temperature is out	Cooling water	=	40	°C
	Product	=	54.8	°C
The shell	The ID	=	10	in
	Baffle	=	10	in
	Passes	=	1	
	ΔP	=	.193	Psi
Tube	OD	=	3/4	in
	The ID	=	0.62	in
	BWG	=	16	
	Pitch	=	1	Triangular in2

	Long	=	12	ft
	total	=	52	
	Passes	=	2	
	ΔP	=	0.160	psi
Rd	1,201		(hr) (ft2) (oF) / (Btu)	
Area	122.5		ft2	
total	1		unit	
Unit price	:		$162578	

V.20 Methanol Condenser of DC 02 (COND-02)

Table V. 20 Specification of Methanol Condenser 02 (COND-02)

Tool Specifications:				
Tool's name	CO2 Methanol Condenser			
Tool Code	E-222			
Function	Condensing Top Products from Distillation Column DC-02			
Type	*Vertical Condenser*			
material	*Carbon steel*			
Temperature entered	Cooling water	=	30	°C
	To condenser	=	87.6	°C
Temperature is out	Cooling water	=	40	°C
	Product	=	54.8	°C
The shell	The ID	=	10	in
	Baffle	=	10	in
	Passes	=	1	
	ΔP	=	0.24155	Psi
Tube	OD	=	3/4	in
	The ID	=	0.62	in

BWG = 16

Triangular

Pitch = 1 in2

Long = 12 ft

total = 52

Passes = 2

ΔP = 0.160 psi

Rd 1,072 (hr) (ft2) (oF) / (Btu)

Area 122.5 ft2

total 1 fruit

Unit price : $ 162578

V.21 Accumulator of DC 01 (AC-01)

Table V. 21 Specification of Accumulator 01 (AC-01)

Specification		**Information**
Code	:	AC-01
Unit	:	1
Function	:	A place to hold the condensation liquid
Capacity	:	255 ft3
Material	:	Carbon steel SA-283 Grade C
Size		
The ID	:	64 in
OD	:	66 in
Shell Height (Ls)	:	130 ft
Shell Thickness (ts)	:	1,0 in
Cover Thickness (th)	:	1,0 in

Head type	:	Standard Dished Head
Total height of the vessel	:	155.50 in

V.22 Accumulator of DC 02 (AC-02)

Table V. 22 Specification of Accumulator 02 (AC-02)

Specification		Information
Code	:	ACC-02
Unit	:	1
Function	:	A place to hold the condensation liquid
Capacity	:	453 ft3
Material	:	Carbon steel SA-283 Grade C
Size		
The ID	:	200,6 in
OD	:	204 in
Shell Height (Ls)	:	158 ft
Shell Thickness (ts)	:	1.7 in
Cover Thickness (th)	:	1.5 in
Head type	:	Standard Dished Head
Total height of the vessel	:	229.65 in

V.23 Distillation Feed Pump (P-02)

Table V. 23 Specification of Feed Pump (P-02)

Specification		Information
Code	:	P-02

Type	:	Centrifugal Pump				
	:	casing and wear ring		:	cast iron	
	:	impeller and wear ring		:	cast iron	
	:	shaft		:	carbon steel	
total	:	1	fruit			
Function	:	Pump the product to be distilled				
Fluid name	:	Methanol				
Design conditions						
T design	:	161,6			ºC	
P design						
Suction pressure	:	10,48	bar	=	152	psi
Discharge pressure	:	13	bar	=	188.55	psi
Capacity	:	308	gpm			
Fluid viscosity	:	0.1717	cp			
Differential head	:	10.9	ft			
Motor power	:	1.5	hp			
Unit price	:		$ 68770			

V.24 Methanol Storage Tank (T-01)

Table V. 24 Specification of Methanol Storage Tank

Tool Specifications:		
Function	:	As a storage area of 99.5% methanol
Shape	:	Cylindrical tank, top lid shaped standard dished head and bottom lid flat bottom shape

The shell

Construction material	=	Carbon Steel SA-240 grade M				
Diameter	=	50.00	ft	=	15.24	m
High	=	30.47	ft	=	9.29	m
High top lid	=	298.02	inch	=	7.57	m
Bottom lid height	=	30.50	inch	=	.77	m
Thick top lid	=	0.63	inch	=	0.02	m
Thick bottom lid	=	13.00	inch	=	0.33	m
Unit Price	=	$ 318515				

Chapter VI
ECONOMIC ANALYSIS

VI.1 Human Resource Management

VI.1.1 Form of Corporate Body

The body form of the Methanol Plant from Coal was chosen by a Limited Liability Company (PT). A Limited Liability Company is an alliance that runs a company with a business capital divided into shares, in which each partner (also called a company) takes part in one or more shares. This was chosen because of the following considerations:

1. Company capital can be more easily obtained from the sale of shares and loans.
2. The capital owner is the shareholder while the executor is the board of commissioners.
3. Responsibility of shareholders is limited, because everything related to the smooth running of production is handled by the company leader.
4. Shareholder wealth is separate from the company's wealth, so shareholder wealth does not determine the company's capital.

VI.1.2 Company Organizational Structure

Based on the chosen organizational system, i.e. lines and staff, a organizational structure chart is drawn up. The amount of each part is based on the amount of labor needed by the company. Where a company with a capacity of 2000 tons / day with two main stages of the process, it requires a workforce of 90 hours / day-process.

(Timmerhauss, 1991)

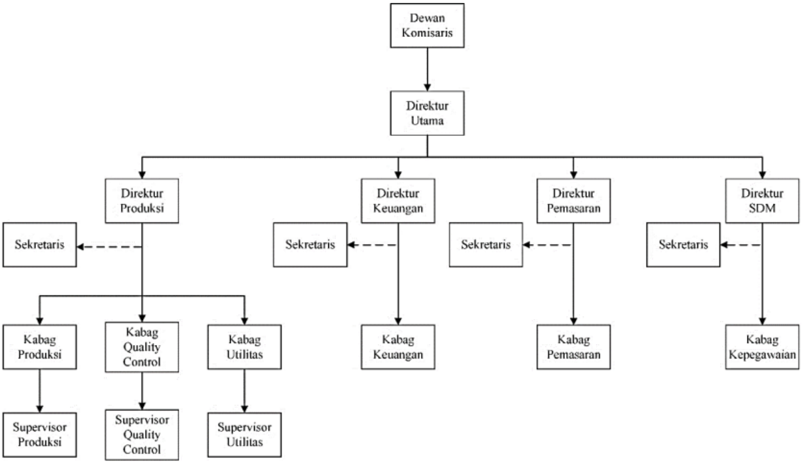


Figure VI. 1 Organizational structure

The chart above is the company's organizational structure of the Coal Methanol Plant. The following is an explanation of the division of labor within the company organization:

1. **Board of Commissioners**

The Board of Commissioners acts as a representative of the shareholders. Commissioners are appointed according to the provisions in the agreement and can be dismissed at any time if they act not in accordance with the articles of association or the interests of the shareholders of the company.

Board of Commissioners' Duties:

- Supervise the director and try to make the director's actions not detrimental to the company.
- Establish company policies.
- Conduct evaluation / supervision of the results obtained by the company.
- Give advice to the director if the director wants to make changes in the company.

2. **President Director**

The president director is the management in the company and is the highest leader and the main person in charge of the company as a whole.

The duties of the chief director:

- Establish company strategies, formulate plans and how to implement them.
- Establish the organizational system adopted and determine the division of labor, duties and responsibilities within the company to achieve the goals set.
- Establish proper coordination of all sections.
- Give instructions to subordinates to carry out their respective duties.
- Accountable to the board of commissioners, all the implementation of the company's budget and income.
- Determine financial policies.

In addition to the above tasks, the president director is also entitled to represent the company legally and directly in all matters and events related to the company's interests.

3. Director of production

The director of production and maintenance is tasked with assisting the main director in carrying out his duties, especially those related to factory production operations, factory construction and the quality of raw materials used, as well as the products produced. In this case, a director of production and maintenance is assisted by a supervisor who handles the fields of process and quality control, and oversees supervisors in their respective sections.

The duties of the production director are:

- Assist the main director in planning as well as in reviewing the main policies in the field of factory production operations in terms of production, plant construction and the quality of raw materials and products produced.
- Determine factory operating policies in order to obtain maximum results.

- Establish proper coordination of the production department.
- Give instructions to subordinates to carry out their respective duties.
- Responsible directly to the president director.

4. Director of Finance

The finance director is responsible for assisting the president in carrying out his duties related to financial matters and bookkeeping of the company. In this case, the finance director is assisted by the head of the fund management division in charge of employees in his field.

The duties of the finance director are:

- Assist the managing director in planning as well as in reviewing key financial and bookkeeping company policies.
- Determine the factory's financial policy in order to obtain maximum benefit.
- Establish proper coordination of the financial department.
- Give instructions to subordinates to carry out their respective duties.
- Responsible directly to the president director.

5. Marketing Director

The marketing director is in charge of assisting the main director in carrying out his duties related to marketing. In this case the marketing director is assisted by the head of the department in charge of the sales sector, and supervises several employees in their fields.

The duties of the marketing director are:

- Assist the managing director in planning as well as in reviewing key policies in the field of marketing.
- Determine marketing policies in order to obtain maximum results.
- Establish proper coordination of the marketing department.

- Give instructions to subordinates to carry out their respective duties.
- Responsible directly to the president director.

6. **HR Director**

The HR Director is responsible for assisting the main director in carrying out his duties related to staffing. In this case, the HR director is assisted by a head of staffing who oversees several employees in their fields.

- Assist the managing director in planning as well as in reviewing key policies in the field of staffing, facilities for employees, improving employee quality, service to the community and employees as well as factory safety.

The duties of the HR director are:

- Establish proper coordination of the personnel department.
- Give instructions to subordinates to carry out their respective duties.
- Responsible directly to the president director.

7. **Head of Production**

The duties of the head of production are:

- Responsible directly to the director of production.
- Ensuring that production runs smoothly.
- Ensuring that production is in accordance with existing equipment specifications.
- Ensuring that the production department controls the production process.

8. **Head of Quality Control**

The tasks of the head of quality control are:

- Responsible directly to the director of production
- Ensure that the product complies with the established specifications

- Ensure the entire product testing or analysis process has been carried out

9. Head of Utilities

The duties of the head of the utility are:

- Responsible directly to the director of production
- Maintaining equipment in utilities and for WWTP (Wastewater Treatment Plants)

10. Head of Sales

The duties of the head of sales are:

- Ensuring that the results of production can be distributed and distributed appropriately so that the selling price is affordable and gets the optimum profit.
- Gathering the facts then classifying it and evaluating it.
- Coordinate with sales employees.
- Responsible directly to the marketing director.

11. Head of Fund Management Section

The duties of the head of fund management are:

- Making contact with sellers of raw materials and preparing purchase orders. To prepare for a purchase, it must be specified:
  - Goods purchased
  - Amount purchased
  - Time of purchase
  - Place of purchase
  - Terms of delivery of goods to be purchased
- Gathering the facts then classifying it and evaluating it.
- Coordinate with employees in the area of fund management.
- Responsible directly to the finance director.

12. Head of Personnel Section

The duties of the head of staffing are:

- Take care of research and training of employees and students who will do practical work.
- Take care of employee welfare including salaries, benefits and recruitment of new employees.
- Take care of facilities for employees, improve employee quality, service to the community and employees and plant safety.
- Served to provide assistance to the HR director in employment matters, including: recruitment, selection, placement, termination of employment and wage issues.
- Responsible directly to the HR director.

13. Utility Supervisor

The duties of the utility supervisor are:

- Supervise and arrange forklift and heavy equipment support to all parts.
- Coordinate the employees of the utility which consists of supervisors and operators who work directly in the field.
- Maintain the availability of system / method / method / standard that can be measured to improve and evaluate the system / method / procedure for the maintenance and repair of utility machines as well as how to operate them by managing and monitoring the maintenance and repair schedule.
- Responsible directly to the head of the utility.

14. Process Supervisor

The process supervisor has the duty to ensure that the production process is carried out with effective and efficient techniques and makes it easier for employees to obtain products with low cost, high quality and competitive prices in the shortest possible time.

The task of the process supervisor is:

- Gathering the facts then classifying it and evaluating it.

- Coordinating the employees of the production department consisting of supervisors and operators who work directly in the field.
- Responsible directly to the head of production.

15. Quality Control Supervisor

The supervisor quality control tasks include:

- Controlling product quality, researching, and developing better and more economical use of raw materials and production.
- Analyzing process raw materials and product analysis both chemically and physically.
- Gathering the facts then classifying it and evaluating it.
- Responsible directly to the head of quality control.

VI.1.3 Breakdown of the Number of Workers

The Coal Methanol Plant has a capacity of 460000 tons / year or 1391 tons / day. Based on this capacity and the type of processes that occur at the factory, it is found that this factory requires labor hours of 90 labor hours per day for each stage of the process. In addition, this plant generally has two main stages, namely the gasification stage and the methanol synthesis stage. So that when multiplied by the stages of the existing process, then in one day this factory requires 600 hours of labor. With 3 work shifts for 8 hours, 32 workers are needed as operators (Timmerhauss, 1991)

VI.1.4 Employee Status and Salary Giving

Employee salary systems are distinguished according to employee status, level of education, size of responsibilities / position, and expertise and years of service. Based on their status, factory employees can be classified into 3 groups as follows:

a. Permanent employees

Permanent employees are employees who are appointed and dismissed with a decree (SK) directors and get a monthly salary based on position, expertise and years of service.

b. Daily employee
   Employees who are appointed and dismissed by the directors without a decree from the directors and receive daily wages paid every weekend.

c. Wholesale Workers
   Wholesale workers are workers who are needed by the factory if needed at certain times, for example: shut down power, loading and unloading of raw materials. Wholesale workers receive piece rates for certain jobs.

Table VI. 1 Employee Salary Calculation

No	Position	Salary / month (IDR)	total	Total (Rp)
1	board of Commissioners	30,000,000	3	90,000,000
2	President Director	25,000,000	1	25,000,000
3	director of production	20,000,000	1	20,000,000
4	Director of Finance	20,000,000	1	20,000,000
5	Marketing Director	20,000,000	1	20,000,000
6	HR Director	20,000,000	1	20,000,000
7	Secretary	5,000,000	4	20,000,000
8	Head of Division			
	a. Head of Production	12,500,000	1	12,500,000
	b. Head of Quality Control	12,500,000	1	12,500,000
	c. Head of Utilities	12,500,000	1	12,500,000
	d. Head of Marketing	12,500,000	1	12,500,000
	e. Head of Finance	12,500,000	1	12,500,000
	f. Head of Personnel	12,500,000	1	12,500,000

9	Supervisor			
	a. The utility	10,000,000	2	20,000,000
	b. Process	10,000,000	4	40,000,000
	c. Quality Control	10,000,000	2	20,000,000
10	Operator			
	a. Maintenance	4,500,000	8	36,000,000
	b. The utility	4,500,000	12	54,000,000
	c. Process	4,500,000	32	144,000,000
	d. Quality Control	4,500,000	8	36,000,000
11	Employee			
	a. Doctor	10,000,000	5	50,000,000
	b. Nurse	4,000,000	5	20,000,000
	c. Sales	3,800,000	5	19,000,000
	d. Bookkeeping	3,800,000	5	19,000,000
	e. Fund Management	3,800,000	8	30,400,000
	f. Staffing	3,800,000	5	19,000,000
	g. Education and training	3,800,000	8	30,400,000
12	Security	2,500,000	12	30,000,000
13	Driver	2,500,000	8	20,000,000
14	The errand boy / gardener	2,500,000	8	20,000,000
TOTAL			**155**	**897,800,000**

For operator and temporary employees, a shift time system is applied. This system consists of three shifts, namely morning, evening and night, with the aim of having more rest time for the employee. Where the system of division of working hours is as follows:

Table VI. 2 Distribution of Employee Work Shifts

Shift	Working hours

Shift Morning	8:00 to 16:00
Afternoon Shift	16:00 to 00:00
Shift Night	00.00-08.00

VI.2 Economic Analysis

Economic analysis is calculated to be able to find out whether a planned plant is worth building or not. In the pre-design of the Methanol Plant an evaluation or feasibility study and investment appraisal were carried out. Factors that need to be reviewed to decide whether or not this plant is established are:

1. Rate of Return on Capital (Internal Rate of Return / IRR)
2. Minimum Payback Time (POT)
3. Break Even Point / BEP

VI.2.1 Rate of Return on Capital (Internal Rate of Return / IRR)

From the results of calculations on Appendix D, the price i = 18.73% was obtained. The price i obtained is greater than the price i for loan interest which is 11% per year. With the price of i = 18.73% obtained from the calculation shows that the factory is feasible to be established.

VI.2.2 Payback Time (POT)

From calculations done on Appendix D it is found that the minimum payback period is 4 years 10 months. This shows that the factory is feasible to be built because the POT obtained is smaller than the estimated age of the factory.

VI.2.3 Break Even Point (BEP)

Break-even analysis is used to determine the size of production capacity where the total production cost is equal to the sales results. Fixed costs (FC) and variable costsel (VC), semi-variable costs (SVC) and total costs are not affected by production capacity. From calculations performed on Appendix D it was found that the Breakeven Point (BEP) = 22.09%.

Chapter VII
CONCLUSION

Based on the results explained in the previous chapters, the following conclusions can be concluded:

1. Operation planning : continuous, 24 hours / day, 330 days / year
2. Methanol production capacity : 460,000 tons / year
3. Natural Gas raw material :292.482 tons/year
4. Factory age : 10 years
5. Construction period : 2 years
6. Economic analysis :
   Total Capital Investment = Rp 4,660,384,242,750
   - *Internal Rate of Return* = 37.33%
   - *Pay Out Time* = 4 years 10 months
   - *Break Even Point* = 22.03%

Based on the results of the economic analysis, it is seen that the IRR of 37.33% is above the bank loan interest of 11%. The payback period (POT) in the fourth year of the fifth year is less than the payback period determined by the lender which is 10 years. Under these conditions, the methanol plant from natural gas is feasible to be established.

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